

POWER SYSTEMS DEVELOPMENT FACILITY
TOPICAL REPORT

GASIFICATION TEST CAMPAIGN TC17

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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 (formerly Kellogg Brown & Root) Transport Gasifier, a hot gas particulate control device, advanced syngas cleanup systems, and high-pressure solids handling systems.

This report summarizes the results gasification operation with Illinois Basin bituminous coal in PSDF test campaign TC17. The test campaign was completed from October 25, 2004, to November 18, 2004. System startup and initial operation was accomplished with Powder River Basin (PRB) subbituminous coal, and then the system was transitioned to Illinois Basin coal operation. The major objective for this test was to evaluate the PSDF gasification process operational stability and performance using the Illinois Basin coal. The Transport Gasifier train was operated for 92 hours using PRB coal and for 221 hours using Illinois Basin coal.

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1.0 EXECUTIVE SUMMARY

1.1 SUMMARY

This report discusses test campaign TC17 of the gasification process at the Power Systems Development Facility (PSDF). The PSDF, located near Wilsonville, Alabama, is a flexible test facility designed to develop advanced coal-fired power system components and assess the associated integration and control issues. TC17 began on October 26, 2004, and lasted until November 18, 2004, accumulating 313 hours of on-coal operation.

1.2 TEST CAMPAIGN OBJECTIVES

The primary test objective of TC17 was to evaluate gasifier and particulate control device (PCD) performance with Illinois Basin bituminous coal using air as the gasification oxidant. The system was initially run on Powder River Basin (PRB) subbituminous coal and then transitioned to Illinois Basin coal where operation was optimized.

1.3 TEST CAMPAIGN SUMMARY

PRB coal feed began on October 26. Once the gasifier achieved steady operations, testing began on the new coal silo modifications made to prevent particle size segregation. The coal silo inserts prevented fines from accumulating in the coal feeder surge bin.

After the feeder was restarted, testing on the automatic temperature control scheme began and was successful at maintaining the desired gasifier temperature. On October 29, the piloted syngas burner (PSB) operated on PRB coal-derived syngas. The PSB combusted approximately 13,000 pph of syngas for 6 ½ hours.

Once PSB testing was complete, preparations began for Illinois Basin coal feed. These included transitioning to transport air to enhance gasification, feeding dolomite as a sorbent for sulfur capture, and increasing steam flow to the gasifier for temperature control. When Illinois Basin coal feed began, the gasifier temperatures were increased by about 30 to 50°F. Inadequate fines removal from the particulate control device (PCD) shortly after transitioning to Illinois Basin coal resulted in overfilling the PCD which led to gasification ash bridging in the lower plenum of filter elements.

Because of the limited fines removal rate and the bridging in the PCD, the coal feed rate was lowered, resulting in a lower syngas heating value. The rapidly increasing PCD baseline pressure drop (from 100 to 150 inH₂O within two days of overfilling the PCD) necessitated the removal of the bridged material. To avoid a complete shutdown, the gasifier was transitioned to combustion mode, and the bridged material was removed by on-line combustion over a 12-hour period. Once the bridged material was removed and the baseline pressure drop restored to its previous value of 100 inH₂O with Illinois Basin coal operation, the system was transitioned back into gasification mode.

The test run was briefly interrupted when a leak in the heat exchanger of the atmospheric fluidized bed combustor, needed to supply superheated steam to the gasifier, forced a shutdown. Repairs proceeded quickly, and coal feed resumed on November 10.

The final days of testing optimized the performance of the gasifier with Illinois Basin coal within the limits of fines removal system. The test run ended a few hours earlier than scheduled on November 18, when the coal feeder experienced difficulty transferring coal from the surge bin to the dispense vessel. The shutdown was smooth, and a controlled PCD dust cake combustion was performed by adding air to the PCD to remove any gasification ash dust cake from the filter elements.

1.4 TEST CAMPAIGN PERFORMANCE

Performance of the major equipment during TC17 is summarized in the following three sections:

1.4.1 Transport Gasifier

- The carbon conversion was between 94.7 and 96.7 percent for PRB testing. The carbon conversion during Illinois Basin testing was between 80.4 and 88.5 percent.
- During PRB gasification, the raw lower heating values at the exit of the gasifier were between 52 and 57 Btu/SCF, resulting in projected heating values at the turbine inlet of a commercial gasifier of between 118 and 130 Btu/SCF. The heating values were slightly lower in TC17 than in previous test runs.
- During Illinois Basin coal gasification, the raw lower heating values at the exit of the gasifier ranged from 33 to 48 Btu/SCF, resulting in projected heating values of 91 to 108 Btu/SCF.
- The raw cold gasification efficiency ranged from 51.2 to 53.7 percent during PRB operations and from 35.5 to 41.6 percent during Illinois Basin operations. The commercially projected efficiency was between 69.8 and 71.3 percent for PRB and from 49.0 to 55.9 percent for Illinois Basin.
- The hot gasification efficiency ranged from 84.2 to 85.6 percent for PRB testing and from 70.3 to 78.1 percent for Illinois Basin testing. System constraints limited the Illinois Basin coal feed rate to below 2,600 lb/hr, resulting in low syngas heating values.

1.4.2 Particulate Control Device

- Near the end of the first portion of the run, material deposited on the filter elements was removed on-line by a controlled burnoff. The material build-up was caused when the PCD overflowed due to operational error when the fines removal rate was too low. The material caused an unstable, high pressure drop across the tube sheet, but after the burnoff procedure was completed, an acceptable pressure drop was restored. All indicative instrumentation such as resistance probes and thermocouples confirmed dislodging of the material.

- The excess material apparently caused 11 filter elements to fail. However, because all of the exposed failsafes—both Pall fuses and PSDF-designed failsafes—performed well, outlet loading remained below 1 ppmw.
- The rupture disc failsafe tester was installed with the Ceramem failsafe to simulate catastrophic filter failure. Although the rupture disc burst prematurely, the test was completed, and SRI sampling later that day indicated good performance of the Ceramem failsafe.

1.4.3 Other Systems

- The PSB operated on syngas for around 6 ½ hours on October 29, and was shut down for the transition to Illinois Basin coal. It was not operated while the gasifier was operating on Illinois Basin coal due to the potential for long-chain hydrocarbon deposition in the PSB.
- The new coal silo insert improved coal feeder operations by preventing particle size segregation. The system performance was improved over previous runs.
- During the last third of the test run, the heat transfer coefficient of the primary gas cooler dropped steadily. At the same time, the primary gas cooler exit temperature increased between 60 and 120°F, while the inlet increased less than 10°F.
- Due to limitations of the fines removal rate, the Illinois Basin coal feed rate was maintained below 2,600 lb/hr. Feed rates above this value resulted in solids collecting in the PCD cone. The low coal feed rate resulted in low dry syngas heating values due to a greater proportion of the coal carbon being converted to CO₂ to provide sufficient heat.
- The Continuous Fine Ash Depressurization (CFAD) system, a Southern Company proprietary design, operated well despite the low density of the fines collected (around 6-8 lb/ft³).
- The advanced syngas cleanup system was used to test desulfurization, hydrocarbon reforming, and ammonia cracking. The desulfurization reactor reduced the H₂S content by about 97 percent. High levels of hydrocarbon and ammonia reduction were also achieved.

1.5 CONCLUSIONS

The Transport Gasifier operated for 313 on-coal gasification hours in TC17. Approximately 92 of the hours consisted of PRB coal operation with balance consisting of Illinois Basin coal operation. Throughout the run, the gasifier was able to maintain good solids circulation rates and a stable temperature profile. Due to limitations of the fines removal rate, the Illinois Basin coal feed rate was maintained below 2,600 lb/hr, resulting in low dry syngas heating values.

TC17 demonstrated a significant improvement in PCD failsafe performance following filter element failure over previous runs. For the first time following several filter element failures, good failsafe performance allowed the continuation of the run without shutting down.

Another milestone for PCD operation was the removal of bridged material by controlled combustion on-line without requiring a system shutdown.

Other systems tested included the PSB, which ran for 6 ½ hours on syngas; an ash silo insert, which decreased particle size segregation; and the syngas clean-up skid, which evaluated sulfidation, hydrocarbon cracking, and ammonia removal technologies. The run provided valuable experience that will enhance future operations.

2.0 OPERATIONS

2.1 PROJECT DESCRIPTION

The Power Systems Development Facility (PSDF), near Wilsonville, Alabama, is funded by the U.S. Department of Energy, Southern Company, and other industrial participants currently including the Electric Power Research Institute, Siemens Westinghouse Power Corporation, KBR (formally Kellogg Brown & Root), Peabody Energy, and the Lignite Energy Council. The PSDF is an engineering scale demonstration of key features of advanced coal-fired power systems designed at sufficient size to evaluate system components and assess the integration and control issues of these advanced power systems. The facility also supports clean coal technology programs to address environmental concerns associated with using fossil fuels for producing electricity, chemicals, and transportation fuels.

The KBR Transport Reactor which operates at the PSDF is a pressurized, advanced circulating fluidized bed reactor which can operate in either combustion or gasification mode. While operating in gasification mode, either air or oxygen can be used as the oxidant. The particulate-laden gas exiting the reactor is filtered by a downstream high temperature, high pressure filter vessel, the Siemens Westinghouse particulate control device (PCD). A slipstream syngas clean-up skid is also available to test various pollutant control technologies. A flow diagram of the gasifier train is shown Figure 2.1-1. As of the conclusion of TC17, the Transport Gasifier train had operated for about 5,000 hours in combustion mode and over 6,400 hours during gasification.

The Transport Gasifier, shown in Figure 2.1-2, consists of a mixing zone, a riser, a disengager, a cyclone, a standpipe, a loopseal, and a J-leg. Steam and either air or oxygen are mixed together and introduced in the lower mixing zone while the fuel, sorbent, and additional air and steam (if needed) are added in the upper mixing zone. The steam and oxidant, along with the fuel, sorbent and solids from the standpipe, are mixed together in the upper mixing zone. The upper mixing zone, located below the riser, has a slightly larger diameter than the riser. The gas and solids move up the riser before entering the disengager, which removes larger particles by gravity separation. The majority of the solids flow from the disengager into the standpipe, and the remaining solids flow, along with the syngas, to the cyclone, which removes most of the particles not collected by the disengager. At the bottom of the cyclone is a loop seal, which prevents backflow of solids. The solids collected by the disengager and cyclone are recycled back to the gasifier mixing zone through the standpipe and a J-leg. 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. Due to a lower oxygen supply pressure, the maximum operating pressure is about 180 psi during oxygen blown gasification.

For start-up purposes, a direct propane-fired burner is operated at the gasifier mixing zone. Coal and sorbent (when required for sulfur capture) are separately fed into the Transport Gasifier through lockhopper feed systems. Coal is ground to a nominal particle diameter between 250 and 400 microns. Sorbent, either limestone or dolomite, is ground to a nominal particle diameter of 10 to 100 microns.

The gas exits the Transport Gasifier cyclone and goes to the primary gas cooler and then to the PCD for final particulate clean-up. The metal or ceramic filter elements used in the PCD remove essentially all the dust from the gas stream. Shown in Figure 2.1-3, 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 solids and control the pressure drop across the tube sheet. The solids fall to the PCD hopper, are cooled in a screw cooler, and are removed through a lock hopper system or the continuous fine ash depressurization (CFAD) system, a Southern Company design for solids depressurization.

After exiting the PCD, a portion of the syngas can be directed to the piloted syngas burner (PSB), a gas turbine combustor designed to burn coal-derived syngas with a lower heating value below 100 Btu/SCF. Propane supplied to the PSB serves as a pilot for the burner as well as a supplement to the syngas fuel to maintain a stable flame. After combusting in the burner, the gas passes through the turbine before exiting the turbine stack. An associated generator supplies power to the electricity transmission grid. A small portion of the syngas, up to 100 lb/hr, can also flow to a specialized gas cleanup system downstream of the PCD. The gas cleanup system provides a means to test various pollutant control technologies, including removal of sulfur, nitrogen, and chlorine compounds.

The syngas not flowing to the PSB or cleanup system continues to the secondary gas cooler and then passes through a pressure control valve. The gas is then sent to the atmospheric syngas combustor (thermal oxidizer) which oxidizes carbon monoxide, reduced sulfur compounds (H_2S , COS , and CS_2), and reduced nitrogen compounds (NH_3 and HCN). The atmospheric syngas combustor uses propane as a supplemental fuel. The gas from the atmospheric syngas combustor goes to the heat recovery boiler, through the baghouse, and then to the stack.

The Transport Gasifier produces both fines filtered by the PCD and coarse solids extracted from the gasifier standpipe. The two solid streams are cooled using screw coolers, reduced in pressure in lock hoppers, and then combined together. The CFAD system is also available for depressurizing fines from the PCD. Before final disposal, the gasification ash is sent to an atmospheric fluid bed combustor, where reduced compounds in the ash are oxidized. The combustor provides superheated steam for the gasifier.

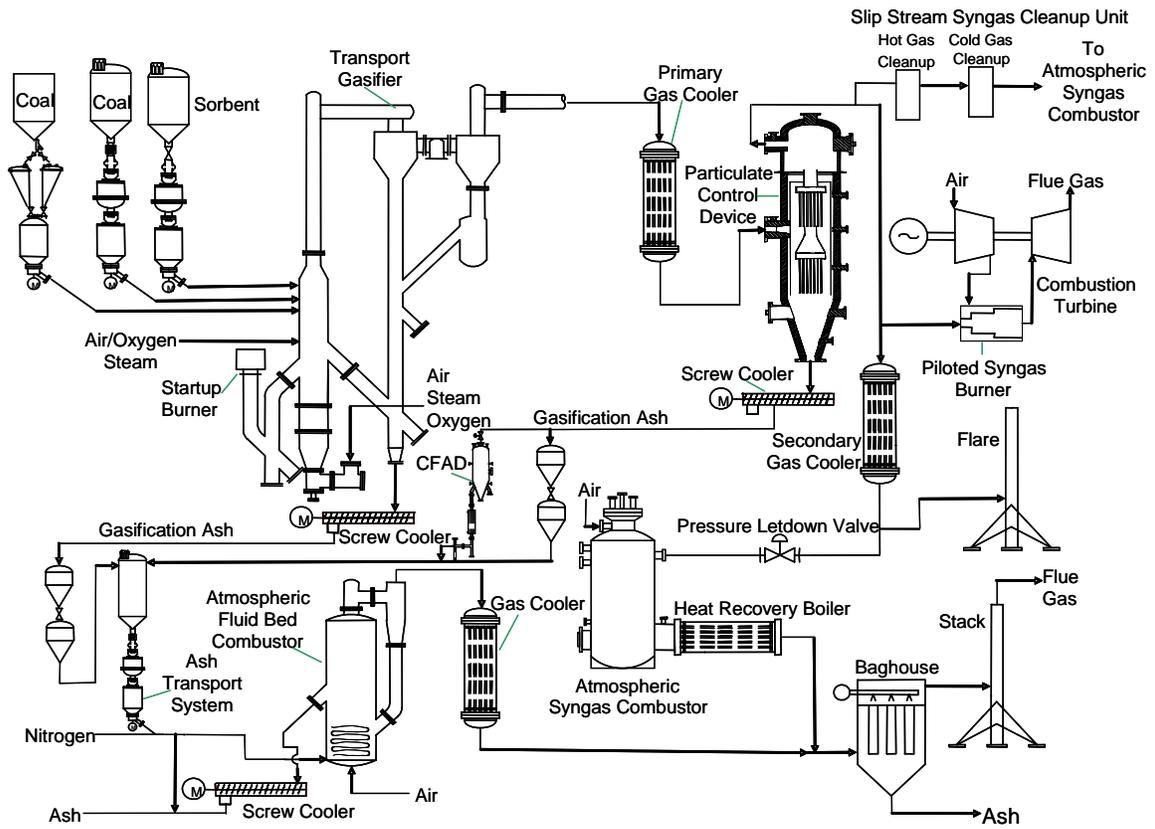


Figure 2.1-1. Flow Diagram of the Transport Gasifier Train

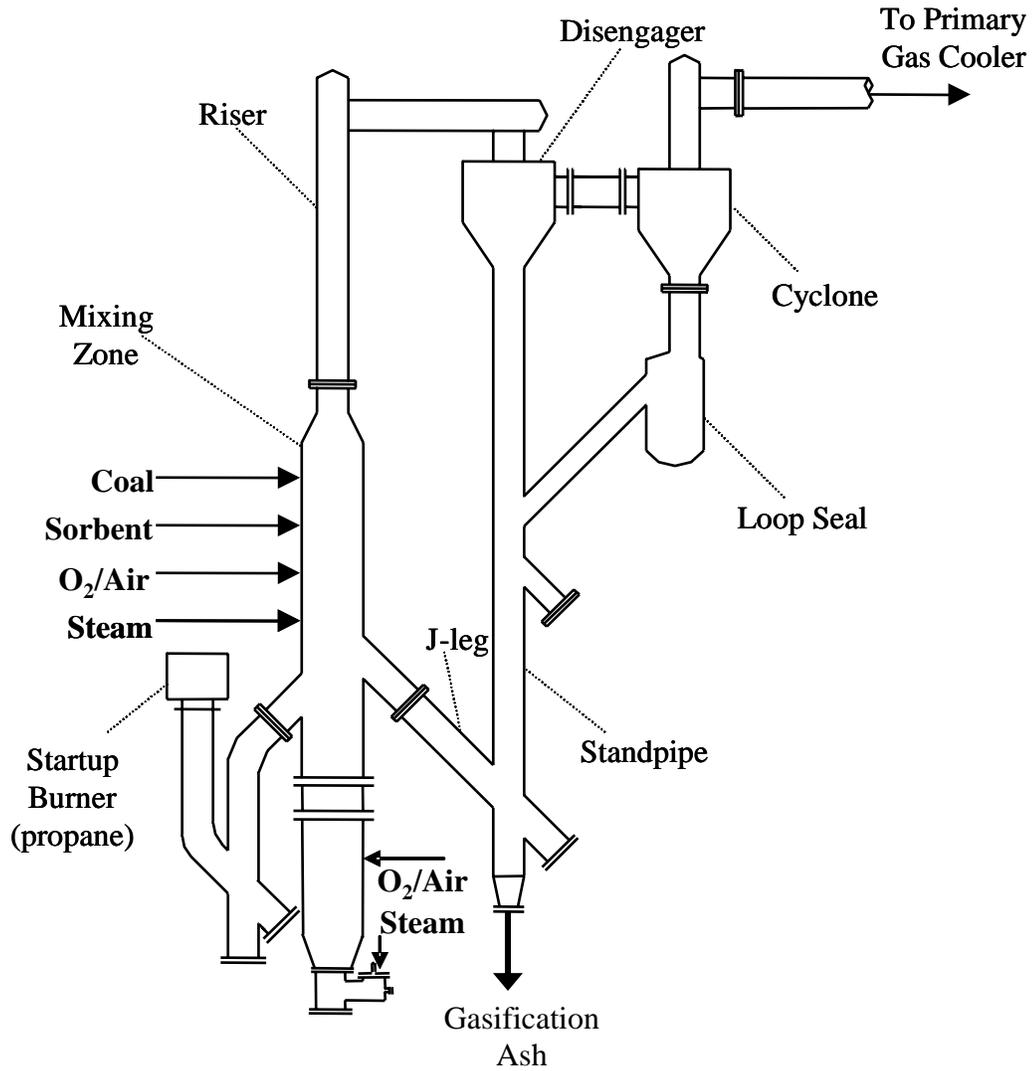


Figure 2.1-2. Transport Gasifier

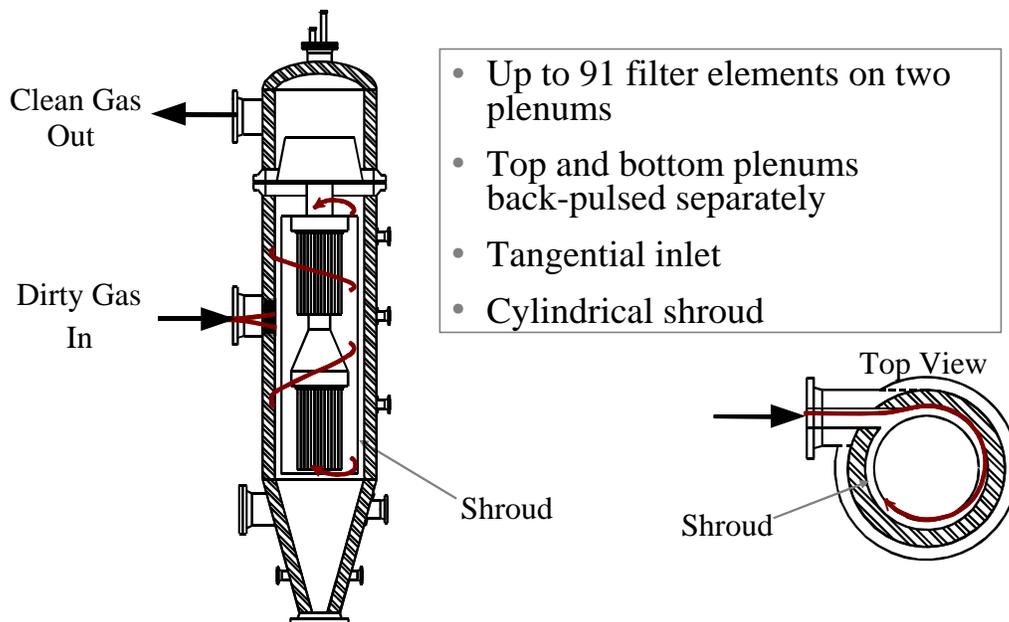


Figure 2.1-3. Particulate Control Device

2.2 DETAILED TEST CAMPAIGN OBJECTIVES

TC17 was planned as a nominally 250-hour test run with the primary objective of evaluate gasifier and PCD operations using Illinois Basin bituminous coal. Powder River Basin (PRB) subbituminous coal used as a startup fuel, the gasifier was transitioned to Illinois Basin coal feed, and the Transport Gasifier train was successfully operated for a total of 313 hours, with 221 hours of Illinois Basin coal operation

Secondary objectives included the following:

- *Operate the Piloted Syngas Burner (PSB)* – The PSB successfully operated for approximately 6.5 hours on syngas at flow rates up to 13,000 lb/hr.
- *Test Pulverized Coal Silo Insert* – The new coal silo insert was successfully tested and resulted in improvements in coal feeder operations.
- *Continue Sensor Development* – Various thermowell materials, including Hastelloy X, HR-160 and ceramic, were tested, and varying configurations were used to evaluate the effect of thermowell insertion length and position on temperature measurement. Ceramic filter elements were also used on selected pressure differential indications to evaluate their effectiveness in preventing solids plugging. Based on the installation of thermowells at different insertion lengths it was determined that a two inch insertion length beyond the refractory wall plane is sufficient for an accurate temperature measurement. The new pressure differential measurement with the ceramic filters worked fairly well. Although balancing the measurement was problematic, it correlated with other differential pressure measurements very well. The constant purge flow controllers also performed well.
- *Continue Testing and Analysis of Failsafe Performance* – The on-line failsafe tester was used to assess the performance of the CeraMem failsafe. A sample collected on the PCD outlet showed less than 0.1 ppmw, which indicated good performance of the failsafe. Failsafe performance following some unexpected filter element failures was also very effective.
- *Perform Controlled Combustion of Dust Cake* – The PCD dust cake combustion was successfully performed at system pressure and temperature. Controlled combustion was first performed to dislodge bridged material, and later, at shutdown, to remove the dustcake.
- *Continue Evaluation of Bridging* – The causes of gasification ash bridging in the PCD continued to be evaluated during the test run. Additional resistance probes were installed and evaluated to gain more understanding of bridging growth. The resistance probes worked well in detecting bridging, providing a discrete alarm signal to the distributed control system (DCS).
- *Continue Particulate Characterization and Drag Evaluation* – Gasification ash samples were collected in situ at the PCD inlet and were thoroughly characterized to examine the effects of fuel type, carbon conversion, sorbent addition, and operating conditions. The characterization included measurements of particulate loading, particle-size distribution, chemical composition, physical properties, and drag. Measurements made in the laboratory were compared to the transient drag values determined from the PCD pressure drop and particulate loadings.
- *Continue Analysis of Particulate Collection Performance* – Particle concentration at the outlet of the PCD was measured, and the collection efficiency was determined. Outlet loading sampling indicated a small amount of particle penetration of up to 0.19 ppmw during the first two days of the test run. Subsequent outlet samples indicated very high collection

efficiency. When 11 filter elements failed due to overfilling the PCD, outlet loading was still very low because of good failsafe performance.

- *Continue Evaluation of On-Line Particulate Monitor* – Output of the Dust Alert 90 particulate monitor from PCME, Inc. was evaluated to gain a better understanding of the instrument response and to detect any particle penetration through the PCD.
- *Continue Filter Element Testing*—Filtration exposure of iron aluminide filter elements was continued, and the elements with a modified weld structure were further evaluated. Further corrosion of the iron aluminide elements as seen at the conclusion of previous test campaigns was observed, although the elements maintained their mechanical integrity and high collection efficiency. The modified weld structure was determined to be inferior, and it will not be utilized in the future.
- *Commission Mini Reactor Systems* – The mini reactor systems with a slipstream processing of 1 lb/hr syngas were successfully commissioned for evaluating contaminant removal. Tests were performed for ammonia cracking and hydrocarbon reforming using various catalysts and achieving 92 to 98 percent reductions of ammonia and up to 92 percent reduction in hydrocarbons.
- *Evaluate Syngas Desulfurization Technology* – The hot gas cleanup system achieved 74 hours of desulfurization on a 12 lb/hr syngas slipstream using Sud-Chemie RVSLT-1 catalyst. The desulfurization reactor reduced the H₂S content by about 97 percent.
- *Test Hydrocarbon Cracking Unit* – The hot gas cleanup system and mini reactor achieved 67 hours of hydrocarbon cracking using a Grace Davison catalyst. Results of the testing showed reductions from 62 to nearly 100 percent for various hydrocarbons.
- *Evaluate Effects of Operating Parameters on Gas Clean-Up Performance* – The effects of different operating parameters such as space velocity, temperature, pressure and gas composition were evaluated throughout the testing.

2.3 DETAILED TEST CAMPAIGN SUMMARY

Forty four equipment modifications were completed in the outage preceding test run TC17. The most significant of these are listed below:

- A new insert was installed in the pulverized coal silo to prevent solids segregation and rat-holing.
- A new pressure differential measurement with ceramic filters was installed.
- Cooling coils were added to the Continuous Fine Ash Depressurization (CFAD) system to test cooling of gasification ash in anticipation of removing the upstream fines screw cooler following TC17.

The following is a brief operating synopsis of TC17:

October 25

The main air compressor was started, and the startup burner was used to heat the gasifier to over 1100°F in order to begin coal feed. The solids circulation rate was high and stable throughout the heating process. Before coal feed began, however, the atmospheric fluidized bed combustor fuel oil nozzles plugged, delaying the startup. The nozzles were cleared, but the atmospheric fluidized bed combustor did not warm to a high enough temperature to allow sufficient steam production for coal feed.

October 26

Once the atmospheric fluidized bed combustor was at temperature to produce sufficient quantities of superheated steam, operations attempted to start the Clyde coal feed lock hopper system, referred to as the original coal feeder. Coal feed could not be started at this time due to plugged dispense vessel vent lines. After unplugging the coal feeder vent lines, coal feed began. The coal feeder tripped on every cycle due to the dispense vessel spheri valve closing too slowly. An adjustment to the coal feeder Programmable Logic Controller (PLC) logic allowed the test run to proceed. The gasifier ran with a coal feed rate of around 2,600 lb/hr during the night. The temperatures fluctuated more than normal, possibly due to an erratic coal feed rate and a low standpipe level. The standpipe level dropped at an excessive rate.

October 27

The coal silo modifications were tested to evaluate the effectiveness in preventing particle size segregation. Although the coal feeder tripped due to a plugged conveying line (possibly due to excessive flow through the dispense vessel vent line), the tests confirmed that the modifications prevented fines from accumulating in the coal silos, and the coal transfer masses were consistent with each cycle. Sand was added to the gasifier to form pressure seals and to provide for even heat up of the gasifier. After closing both vent lines on the coal feeder dispense vessel, the unit returned to steady state conditions, and the coal feeder conveying line did not plug again. Once the gasifier temperatures stabilized, the automatic gasifier temperature controls were tested. The gasifier ran smoothly under automatic temperature control with a coal feed rate of around 3,300 lb/hr and an outlet temperature of over 1,700°F. The gasifier pressure was around 220 psig.

October 28

The tests on the coal silo modifications ended when the coal silo emptied without causing fines to accumulate in the coal feeder. The coal transfer masses were acceptable during the tests. Unfortunately, the vent line on the coal feeder lock vessel plugged due to overfilling of the lock vessel. The gasifier continued to operate at minimum coal feed and air flow rates while the vent line was unplugged. After the vent line was cleared, the test plan was resumed. The gasifier conditions returned to normal upon the increase of coal and air flow rates, and testing of the automatic gasifier temperature controls continued. A step change of the coal feed rate was made to determine how well the system responded. Sand was fed to the loop seal to control the declining bed level.

October 29

Automatic temperature control testing continued with step change increases and decreases of the air flow rate through the burner leg. During testing, the magnitude of temperature response was around 50°F. Control testing of the CFAD automation system also proceeded. The PSB ran for most of the day on propane and for about 6 ½ hours on syngas. The syngas flow rate to the PSB was around 13,000 lb/hr. To prepare for bituminous coal feed, the sorbent feeder was started with dolomite at a low rate, the coal steam shroud steam flow rate was placed in automatic control, and transport air flow through the coal feeder was initiated. The transport air made up about 75 percent of the total transport gas flow, with the remainder being nitrogen. The rupture disc failsafe tester on the particulate control device (PCD) opened prematurely. A sample collected at the PCD outlet showed <0.1 ppmw, indicating that the CeraMem failsafe successfully plugged once the rupture disc opened. Sand was added again to raise the standpipe level to maintain a sufficient solids circulation rate.

October 30

Bituminous coal was fed at a rate of around 2,500 lb/hr and then reduced to about 2,000 lb/hr. Gasifier operations were smooth during the transition. Initially, the automatic temperature controls caused some temperature swings. After tuning, the automatic controls worked well. Gasification ash began to accumulate in the PCD cone. The bituminous ash did not seem to flow well through the solids removal systems. In an effort to remove ash from the cone, the coal feed rate was minimized, and the fines removal bypass was operated. Bridging between the filter elements caused by overfilling the PCD was indicated by filter element resistance probes and thermocouples. During that time, the on-line PCD outlet particulate monitor showed a high reading, indicating a possible leak. SRI conducted outlet sampling which showed contamination by condensed material and a small amount of particulate, but did not indicate a leak. The efforts to remove the ash from the cone were successful, but the bridging between the filter elements remained. Sand was added to the gasifier.

October 31

Gasifier operations remained steady, and the steam, air, coal, and dolomite feed rates were increased slightly. The gasifier pressure was increased slightly to decrease the riser velocity. In an attempt to reduce the lower mixing zone (LMZ) temperatures, the gasifier temperature setpoint was raised by 10°F and the steam flow was increased slightly. Sand was added to the gasifier.

November 1

The bridging between the filter elements caused an abnormal increase in PCD baseline differential pressure. The gasifier was transitioned to combustion mode to use excess oxygen in an attempt to burn the bridged material off the filter elements in a slow, controlled way. After about 12 hours of combustion, the filter elements appeared free of bridging, and a stable baseline pressure drop was restored. The coal feed rate was increased to return to gasification mode. Sand was added since the standpipe level had dropped during the combustion period. The automatic temperature control scheme worked well at preserving gasifier temperatures, even during the sand addition.

November 2

The gasifier was manually tripped, and all equipment was shut down due to a water leak in the heat exchanger at the exit of the atmospheric fluidized bed combustor.

A brief outage occurred from November 2 to November 10. During this period, ruptured cooling tubes in the atmospheric fluidized bed combustor heat exchanger were repaired. While the work was in progress, the PCD was inspected, and eleven broken filter elements were found and replaced. The elements were apparently broken when the PCD was overfilled. After all repairs were complete, the test run continued.

November 10

The gasifier pressure test and PCD preheat were completed to prepare for startup. The startup burner was lit, and the gasifier heat up progressed. Coal feed began, and the gasifier temperature and pressure were steadily increased thereafter. When the gasifier temperature reached 1850°F, the coal feed rate was increased and the gasifier was transitioned from combustion mode to gasification mode.

While the gasifier was running with a maximum temperature of 1910°F, an outlet temperature of around 1850°F, and a coal feed rate of around 2,200 lb/hr, some of the PCD resistance probes began responding, which indicated bridging. The gasifier was transitioned back to combustion mode to try to dislodge the bridging.

November 11

Combustion mode continued. Coal feed was eventually stopped due to the possibility of bridging. It was later determined, based on indications from the PCD pressure drop and thermocouple responses, that the resistance probes were giving false indications. Problems with the coal feeder dispense vessel spheri valve delayed starting coal feed. After repairs were completed, coal feed resumed, heating the gasifier from 1100 to 1900°F.

November 12

When the highest gasifier temperature was around 1900°F, the gasifier was transitioned to gasification mode by increasing the coal feed rate. Sand was added to the gasifier.

November 13

The coal feed rate was increased to around 2,500 lb/hr. At a maximum gasifier temperature of 1915°F, the overall gasifier temperature was stable with less than a 5°F fluctuation. The CFAD worked well, operating at a discharge rate of approximately 500 lb/hr. The capacity of the fines removal screw cooler limited the coal feed rate.

November 14

Measurements indicated that the fines removal screw cooler was operating at its maximum capacity due to a low fines density. Gasification ash samples measured a few days earlier showed that the bulk density of the ash was as low as 11 lb/ft³. The ash density had decreased from an average value of 20.4 lb/ft³ during PRB operation to an average value of 13.9 lb/ft³ with Illinois Basin operation.

November 15

Many of the riser thermocouples began to fail, including the controlling thermocouple, which caused the control scheme to add too much air to the gasifier. The temperatures increased quickly and caused an upset in the unit. The unit recovered quickly. Sand was later added to the standpipe. The differential pressure in the loop seal downcomer steadily dropped. Even with an adequate standpipe level, the loopseal downcomer differential pressure was too low. The temperature profile still remained normal.

November 16

In the morning, the coal feed rate was reduced due to a rupture in the fines removal lock hopper system bypass line. After the line was repaired, the gasifier returned to the same operating parameters as before the rupture. The gasifier and all auxiliary systems continued to operate steadily.

November 17

The gasifier temperature setpoint was increased by 10°F to evaluate the effect of temperature on carbon conversion. The standpipe level slowly increased overnight, requiring the use of the standpipe screw cooler to remove excess material.

November 18

Since the solids circulation rates (calculated based on estimated solids/gas slip) had been decreasing, particles from the bottom of the coarse solids removal system were inspected. It was determined that an increase in particle size in the gasifier solids caused the reduction in solids circulation. Due to coal transfer difficulty, the coal feed rate was reduced. While troubleshooting coal feeder operation, the feeder tripped and could not be restarted. The system was later shut down. Dust cake combustion was performed on the PCD during shutdown to remove any gasification ash on the filter elements.

Typical operating conditions during TC17 of the Transport Gasifier and PCD are shown in Table 2.3-1. Analyses of the PRB and of the Illinois Basin coal used during TC17 are given in Table 3.1-6.

Table 2.3-1
Typical Operating Conditions for the Transport Gasifier and Particulate Control Device during TC17

Transport Gasifier	
Startup Bed Material	~ 120 μ m Sand
Startup Fuel	Coal
Fuel Type	Powder River Basin Illinois Basin
Fuel Particle Size (mmd), μ m	200 - 520
Average Fuel Feed Rate, lb/hr	1,900 - 4,000
Sorbent Type	Plum Run Dolomite
Sorbent Particle Size (mmd), μ m	100
Sorbent Feed Rate, lb/hr	0 - 370
Gasifier Temperature, $^{\circ}$ F	1725 - 1935
Mixing Zone Pressure, psig	210 - 230
Riser Gas Velocity, fps	40 - 48
Standpipe Level, inH ₂ O	90 - 160
Gasifier Outlet Gas Flow Rate, lb/hr	18,000 - 22,000
Air/coal mass ratio, lb/lb	3.0 - 5.2
Steam/coal mass ratio, lb/lb	0.3 - 1.2
Particulate Control Device	
Inlet Temperature, $^{\circ}$ F	700-975
Inlet Loading, ppmw	18,100 - 32,300
Outlet Loading, ppmw	< 0.1
Baseline Pressure Drop, inH ₂ O	60 - 90
Number of Filter Elements	72
Filter Element Type	Pall Iron Aluminide (FEAL)
Filtration Area, ft ²	204.5
Face Velocity, ft/min	3 - 4
Pulse Valve Open Time, sec	0.2
Pulse Time Trigger, min	5
Top Plenum Pulse Pressure, psi	250 above system pressure
Bottom Plenum Pulse Pressure, psi	250 - 450 above system pressure

2.4 DETAILED INSPECTIONS/CONCLUSIONS

2.4.1 Transport Gasifier

The Transport Gasifier was visually inspected after the test campaign to ascertain the condition of refractory and to look for indications of ash deposits. The riser was relatively clean near the top section with minor refractory loss around penetrations. Lower down, there were some small deposits on the walls in the top most section of the riser. In the lower portion of the riser, there were additional scattered deposits on the walls that were more numerous than those higher up. These deposits extended down to the level of the coal feed nozzle in the mixing zone.

The notch in the inlet to the primary cyclone, shown in Figure 2.4-1, had redeveloped. The notch, caused by erosion by solids entering the cyclone, results in disruption of the solids-gas flow pattern, thus reducing the cyclone solids collection efficiency. The cyclone did not show excessive wear in other areas, and there was no significant deposition.

The loop seal looked typical. The upper part of the downcomer was relatively clean with only a few small wall deposits. There was approximately an inch of deposits on the wall near the normal solids level. These extended down to the solids level in the loop seal at the time of inspection.

The walls of the disengager were clean. There was some minor cracking and spalling of the refractory.

The standpipe and mixing zone were relatively clean and in good condition. There were a few deposits spotted just above the j-leg.

The tube sheet of the primary gas cooler was inspected and it was relatively clean. About five to six tubes were either partially or fully blocked from the top as shown in Figure 2.4-2.

2.4.2 Particulate Control Device

The Particulate Control Device (PCD) was inspected during the eight day outage of the test run. A total of eleven broken filters were found on the bottom plenum. The filters were broken at or near welded junctions between the filter segments. The filters apparently broke during the time the PCD cone was overfilled with gasification ash. The cone overfilled shortly after the transition to bituminous coal due to higher ash carry over rates and poor flow characteristics of the solids. As a result, gasification ash bridged between the filter elements. There appeared to be some remaining bridged deposits between the filter holders on the bottom plenum as shown in Figure 2.4-3. This observation suggested that the gasification ash may have completely filled the bottom plenum, applying stresses on the elements that were sufficient to cause the element breakage. There was no evidence that the level of ash reached the top plenum.

An inspection was also performed after the final shutdown. The inspection revealed one broken filter element on the top plenum. The broken filter was one of several filter elements with a newly modified weld structure that were installed to evaluate the region where the iron aluminide is welded to the collar that is prone to cracking. The material had a deep crack penetrating through the filter

element. The failsafe above this element was plugged, indicating that dust had passed through the crack. There was no evidence that gasification ash bridging occurred during the second portion of the test run.

All filter elements and failsafes were removed from the PCD, pressure washed with water, thoroughly inspected and flow tested. A corrosion product was observed on the inner and outer surfaces of the filter elements which appeared to be related to syngas exposure time. As shown in Figure 2.4-3, the extent of corrosion increased with exposure time. The corrosion product was identified as an iron oxide scale, which suggested that the alumina protective layer had been penetrated, exposing the iron aluminide to attack. Flow tests revealed an increase in flow resistance with exposure time. Generally, the corroded elements exhibited a higher flow resistance, indicating partial obstruction of the pores. Corrosion was also found on some of the Pall fuses and on the newly installed resistance probes. The resistance probes were corroded at the nozzle due to inadequate purge gas flow. The corrosion caused the resistance probes to give false indications of bridging during the second portion of the test run.

The PCD hardware was visually inspected by a representative of Siemens Westinghouse Power Corporation after the run. The tube sheet welds were dye penetrant-inspected, and no relevant indications of penetration were observed. The outer ring of the tube sheet showed a considerable amount of corrosion and deterioration of the Plasite epoxy coating. This deterioration was attributed to attack of condensed acid collected adjacent to the coating. Pitting corrosion was noted on the outer surface of the outer cone. No signs of deterioration were seen on the remaining PCD components.

2.4.3 Other Systems

The Continuous Fine Ash Depressurization (CFAD) system was inspected after the test run. There were no signs of erosion at any place in the system. Inspection results of key components confirmed the viability of conceptual ideas on which the CFAD system design was based. The conveying lines showed signs of erosion even though the mean particle size of the solids was approximately 15 microns. During TC17, a newly designed bend that was based on a previous CFAD simulation was tested. Inspection of the bend showed no signs of erosion inside or downstream of the bend. Testing will continue on the newly designed bend, and its performance will be compared with other types of conventional bends that exist in various conveying lines.

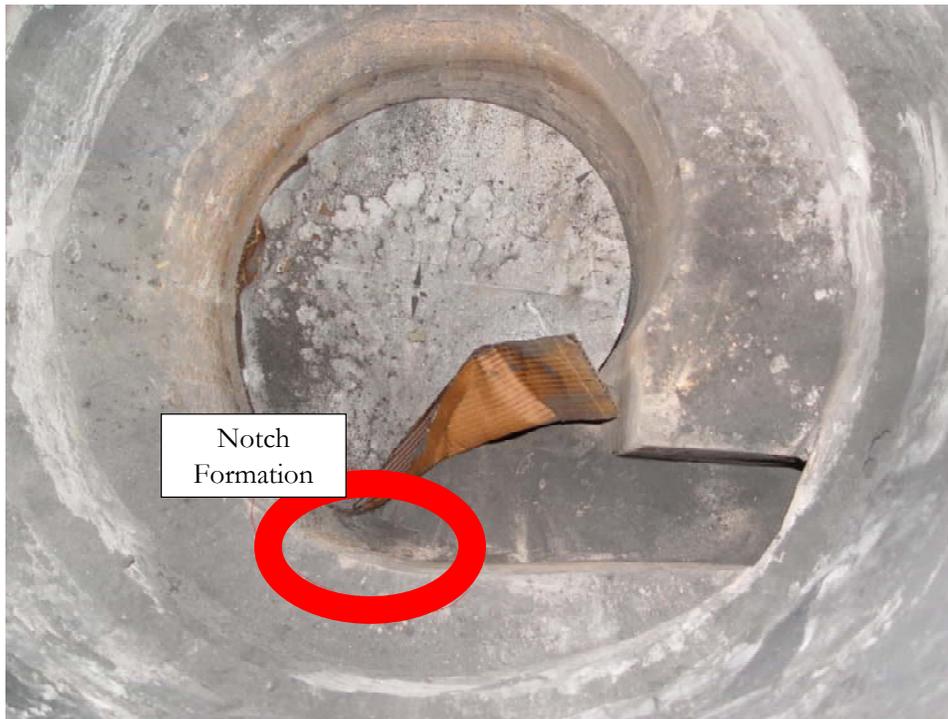


Figure 2.4-1 Notch in Primary Cyclone



Figure 2.4-2 Blocked Tubes in Primary Gas Cooler



Figure 2.4-3 Filter Holders on Bottom Plenum



Figure 2.4-4 Increase of FEAL Filter Element Corrosion with Syngas Exposure Time

3.0 PERFORMANCE

3.1 TRANSPORT GASIFIER PERFORMANCE

3.1.1 Overview

- The carbon conversion was between 94.7 and 96.7 percent for PRB testing. The carbon conversion during Illinois Basin testing was between 80.4 and 88.5 percent.
- During PRB gasification, the raw lower heating values at the exit of the gasifier were between 52 and 57 Btu/SCF, resulting in projected heating values at the turbine inlet of a commercial gasifier of between 118 and 130 Btu/SCF. The heating values were slightly lower in TC17 than in previous test runs.
- During Illinois Basin coal gasification, the raw lower heating values at the exit of the gasifier ranged from 33 to 48 Btu/SCF, resulting in projected heating values from 91 to 108 Btu/SCF.
- The raw cold gasification efficiency ranged from 51.2 to 53.7 percent during PRB operation and from 35.5 to 42.6 percent during Illinois Basin coal operation. The commercially projected efficiency was between 69.8 and 71.3 percent for PRB and from 49.0 to 55.9 percent for Illinois Basin.
- The hot gasification efficiency ranged from 84.2 to 85.6 percent for PRB testing and from 70.3 to 78.1 percent for Illinois Basin testing.
- The coal moisture averaged 21.9 percent during PRB operation and 10.0 percent during Illinois Basin operation. The PRB coal ash composition averaged 4.9 percent, the Illinois Basin coal averaged 8.0 percent ash.
- The circulating gasifier solids as sampled from the standpipe (referred to as standpipe solids) reached a steady state particle size of around 155 microns (Sauter mean diameter) during PRB operation. Due to the frequent addition of sand during Illinois Basin operation, the standpipe particle size did not reach steady state. The standpipe solids bulk density averaged 84 lb/ft³ during PRB operation and 81 lb/ft³ during Illinois Basin operation. The standpipe solids CaS content ranged from essentially zero to 0.6 percent. The standpipe carbon content was 4.7 percent or lower for the entire test run with the exception of one outlier.
- The carbon, CaS, Al₂O₃, and SiO₂ concentrations of the PCD inlet in situ samples were fairly consistent with those of the CFAD samples.
- The wet syngas molecular weight was between 26.0 and 26.5 lb/lb-mol.
- Total reduced sulfur (TRS, mostly H₂S) concentrations were between 226 and 246 ppm for PRB operation and between 523 and 672 ppm for Illinois Basin operation.

The test run had 25 periods of steady operation between October 27 and November 18. These periods are given in Table 3.1-1. The operating periods had a cumulative time of about 143 hours, about 46 percent of the total TC17 gasification time. The first seven periods occurred during PRB testing, while the remaining periods were during Illinois Basin testing. No sorbent was used with PRB operation, while dolomite sorbent was used during Illinois Basin feed.

Table 3.1-2 lists the TC17 operating conditions for steady state periods, including gasifier exit temperature, system pressure, coal feed rate, dolomite rate, air rate, syngas rate, steam rate, nitrogen rate (used for instrument purges and aeration), PCD inlet temperature, and PCD solids carryover rate. Gasifier exit temperatures were between 1,673 and 1,722°F for PRB coal and between 1,753 and 1,829°F for Illinois Basin. Since the reactivity of the Illinois Basin coal was lower than for PRB, higher exit temperatures were maintained during Illinois Basin operation to increase carbon conversion. The system pressure ranged from 214 to 230 psig.

The PCD inlet solids flow rates were interpolated from in situ sampling data at the PCD inlet (discussed below). Since only one in situ sampling occurred during PRB operation, the rates are the same for all seven PRB periods. Only limited sampling was available during the first part of Illinois Basin coal feed as well. Thus, the first four Illinois Basin periods also had the same solids loading to the PCD.

3.1.2 Gas Composition

During the test run, gasifier and syngas combustor outlet gas analyzers were continuously monitored, and the results are discussed below. Ten in situ syngas samples were taken during PCD outlet loading sampling and measured for H₂O content. Plotted on Figure 3.1-1 are the in situ H₂O concentrations versus time and H₂O content based on a correlation of temperature and the in situ data. Because the Procal 2000 online H₂O analyzer was out of service during the entire test run, the correlated value provided the only H₂O data.

The correlation used is a water gas shift thermodynamic equilibrium equation with an approach temperature of -200°F, and the measured syngas compositions of CO, CO₂, and H₂. The equation was developed from Aspen simulations. The approach temperature of -200°F was determined by a best fit of the TC17 in situ H₂O data. The correlation is:

$$K_p = \frac{[CO_2][H_2]}{[H_2O][CO]}$$

where K_p is the water gas shift equilibrium constant; [CO₂], [H₂], [H₂O], and [CO] are concentrations in mole percent, and K_p is calculated as:

$$K_p = e^{\left(\frac{.7072 \cdot 10^4}{T+T_a+460}\right) - 3.5875}$$

where T is the cyclone exit syngas temperature, and T_a is the approach temperature, both in °F.

The approach temperature is varied to get the best fit of the in situ data. The approach temperature is then used for each operating period to determine the syngas H₂O concentration using:

$$[H_2O] = \frac{100\%}{1 + [CO]_d K_p \frac{100\%}{[CO_2]_d [H_2]_d}}$$

where the subscript d indicates the as-measured dry syngas concentrations in mole percent.

The TC17 hydrogen and oxygen balances syngas were very good, which indicated that this method of estimating the H_2O concentration was satisfactory.

The H_2O concentrations calculated for the operating periods (based on the correlation) are given in Table 3.1-3. The concentrations ranged from 10.7 to 20.9 percent and varied mainly due to changes in the steam flow rate. Based on these H_2O concentrations, the estimated wet syngas compositions for the operating periods are given in Table 3.1-3 and shown on Figure 3.1-2.

The CO concentration typically ranged from 6.6 to 7.4 percent during PRB operation and from 2.9 to 5.9 percent during Illinois Basin operation. The H_2 concentration fluctuated between 7.7 and 8.4 percent during PRB operation and from 5.0 to 6.1 percent during Illinois Basin operation. During PRB operation, the CO_2 concentration ranged from 9.1 to 9.6 percent, while during Illinois Basin operation, the CO_2 content was between 8.8 and 9.6. For both fuel types, the CH_4 concentration was between 0.9 and 1.3 percent, and the C_2^+ concentration was negligible.

Also shown in Table 3.1-3 are the wet syngas molecular weights for each operating period. The syngas molecular weight and nitrogen concentration are both plotted on Figure 3.1-3, since the nitrogen content has the largest influence on the molecular weight. During TC17, the molecular weight ranged from 26.0 to 26.5 lb/lb-mol.

The TRS is the sum of the compositions of all sulfur species in the syngas, which consists mainly of H_2S and COS along with other sulfur compounds, such as CS_2 , present in small quantities. When combusted in the syngas combustor, all sulfur compounds present in the syngas are assumed to be converted to SO_2 , and the TRS is based on the syngas combustor SO_2 analyzer. The syngas TRS concentration and the sulfur removal percentage are listed in Table 3.1-4.

The TRS concentration ranged from 226 to 246 ppm during PRB operation and from 523 to 672 ppm during Illinois Basin operation. The coal analysis showed an average sulfur content of 0.28 weight percent for PRB coal and 0.86 weight percent for Illinois Basin coal. Dolomite feed occurred during the Illinois Basin portion of the test run, supplying calcium oxide to capture sulfur from the Illinois Basin fuel. Dolomite feed did not occur during PRB testing, since, in previous PRB test runs, the use of additional sorbent had little effect on syngas sulfur concentrations. As shown on Table 3.1-4, the sulfur removal was between 7.8 and 10.3 percent for the PRB portion of the testing, attributable to sulfur capture by the high coal ash calcium content of PRB. The sulfur removal during the Illinois Basin portion was between 9.4 and 30.5 percent.

3.1.3 Syngas Heating Values

Raw Syngas Heating Values. The raw syngas lower heating value (LHV) for each operating period was calculated and is listed in Table 3.1-3 and plotted on Figure 3.1-4. All raw LHVs are on a wet basis.

The lower heating value at gas industry standard conditions of 60°F and 14.7 psia was calculated using the formula:

$$LHV(Btu / SCF) = \frac{275 \times (H_2\%) + 322 \times (CO\%) + 913 \times (CH_4\%) + 1641 \times (C_2^+\%)}{100}$$

The raw LHV ranged from 52 to 57 Btu/SCF during PRB operation and ranged from 33 to 48 Btu/SCF during Illinois Basin operation. The lower volatile content of the Illinois Basin coal, the lower coal feed rates, and the higher steam flow rates necessary for heat mitigation kept the heating value lower during Illinois Basin operation.

One way to combine the effects of changes in coal feed and steam flow rates is to determine the overall percent of oxygen of all the gas that enters the gasifier. The overall percent O₂ is calculated by the following formula:

$$\text{Overall Percent O}_2 = \frac{0.21 \times \text{Air} + \text{O}_2}{\text{Air} + \text{O}_2 + (\text{Pure N}_2) + \text{Steam}} \times 100\%$$

All flows are in moles per hour. At the PSDF, a large amount of pure nitrogen is fed to the gasifier for instrument purges, coal and sand transport, and equipment purges. During air blown gasification about half of the nitrogen flowing through the gasifier comes from these pure nitrogen flows, and the remainder comes from the nitrogen in the air.

The TC17 overall percent O₂ are listed in Table 3.1-5. The values ranged from 12.1 to 12.9 percent O₂ during PRB feed and from 11.0 to 12.2 percent O₂ during Illinois Basin coal feed. The overall percent O₂ was higher during PRB coal due to lower steam flow rates.

The TC17 raw LHV data are plotted against overall percent O₂ on Figure 3.1-5. As the overall percent O₂ increases, the LHV also increases. For comparison, the curves of previous PRB data (TC06 through TC08, TC10, and TC12 through TC16), previous Hiawatha data (TC09), Falkirk lignite (TC11), and Freedom Lignite (TC13 and TC16) are included. The general trend for TC17 was slightly lower than previous PRB data. The TC17 PRB heating values were higher than for lignite at the same overall percent O₂, but lower than the Hiawatha coal. When plotted against the overall percent O₂, the Illinois Basin coal had a higher LHV than the Falkirk lignite, but lower than any of the other fuels.

Projected Syngas Heating Values. A commercial-sized Transport Gasifier will produce syngas of a higher quality than the PSDF gasifier due to:

- The use of recycle gas rather than nitrogen for aeration and PCD backpulsing.
- A lower heat loss per pound coal gasified because of a lower surface area to volume ratio.
- A smaller number of instruments and instrument purges.
- A cold gas cleanup train to remove contaminants and water from the syngas.

For details on the projected LHV calculation, see Appendix 5.

The projected LHV and adjusted syngas composition data at the gas turbine inlet for each operating period are given in Table 3.1-5. The projected LHV was between 118 and 130 Btu/SCF during PRB operation and was between 91 and 108 Btu/SCF during Illinois Basin operation.

Based on the adjusted syngas composition, the CO/CO₂ ratios were calculated from the gas data for each operating period, and are listed in Table 3.1-5. The CO/CO₂ ratio varied from 1.2 to 1.4 with PRB coal and from 0.51 to 1.0 with Illinois Basin coal. The lower ratios during Illinois Basin operation were most likely due to the lower volatile content of the fuel and the higher operating temperature.

3.1.4 Gasifier Solids Analyses

TC17 solid samples were analyzed for chemical composition and particle size and were taken from the following locations:

- Coal feeder, FD0210
- Sorbent feeder, FD0220
- Gasifier standpipe (circulating solids)
- Gasifier loop seal downcomer
- Continuous fine ash depressurization (CFAD) system, FD0540
- PCD inlet in situ

PRB coal was used for the first 92 hours of the test run, while Illinois Basin coal was used for the remaining portion of the test run. Table 3.1-6 gives the average proximate, ultimate and ash mineral analyses of the PRB and Illinois Basin coals from samples taken from the coal feeder, as well as the average calcium to sulfur (Ca/S) molar ratios of the coal. The standard deviation for these analyses is also listed. The fuel carbon and moisture contents are shown in Figure 3.1-6.

Both the H₂O and the carbon content of the PRB coal remained essentially constant, while the Illinois Basin coal moisture and carbon contents deviated slightly. The coal H₂O content averaged 21.9 percent during PRB operation and 10.0 percent during Illinois Basin operation. The H₂O content of the Illinois Basin coal was approximately 12 percent lower than the PRB coal. Note that the decrease in moisture content appeared to occur about twenty hours after the coal feeder actually transitioned to Illinois Basin coal. Since samples taken during this transition period apparently contained both PRB and Illinois coal, subsequent data averaging and range reporting excludes them.

Figure 3.1-7 shows the coal sulfur and ash as sampled from the coal feeder during TC17. The sulfur content in the PRB coal was typically between 0.24 and 0.30 weight percent, while the Illinois Basin coal contained a significantly higher percentage of sulfur of 0.63 to 0.96 weight percent. The ash content for the Illinois Basin coal averaged 8.0 weight percent, while the PRB coal averaged 4.9 weight percent.

The higher heating values (HHVs) and lower heating values of the coal are given on Figure 3.1-8 with the TC17 average values of the two types of coal given in Table 3.1-6. The coal HHV is determined using a bomb calorimeter. The calorimeter condenses all the coal combustion moisture

as liquid water. The LHV is calculated by subtracting the heat of vaporization of the coal moisture from the HHV. Since heat recovery steam generators do not recover the coal syngas moisture heat of vaporization, the LHV is a more useful measure of coal heating value. The lower heating values for the PRB coal ranged from 8,700 to 8,800 Btu/SCF during the test run. The higher rank Illinois Basin coal had a LHV ranging from 10,400 to 11,300 Btu/SCF.

Table 3.1-7 gives the chemical analysis for the sorbent as sampled from the sorbent feeder. Plum Run dolomite was the sorbent fed during the Illinois Basin portion of the test run. Dolomite is comprised of calcium carbonate (52.6 weight percent) and magnesium carbonate (42.3 weight percent). No dolomite feeding occurred during PRB operation.

The chemical compositions of the gasifier circulating solids were determined based on the chemical analysis and the following assumptions:

1. All carbon dioxide measured is from CaCO_3 , hence moles CO_2 measured = moles CaCO_3 .
2. All sulfide sulfur measured is from CaS .
3. All calcium not taken by CaS and CaCO_3 is from CaO .
4. All magnesium is from MgO .
5. Total carbon is measured, which is the sum of organic and inorganic (CO_2) carbon. The organic carbon is the total carbon minus the inorganic carbon (CO_2).
6. All iron reported as Fe_2O_3 is assumed to be present in the gasifier and PCD solids as FeO . Thermodynamically, the mild reducing conditions in the gasifier should reduce all Fe_2O_3 to FeO .
7. Inerts are the sum of the BaO , P_2O_5 , Na_2O , K_2O , and TiO_2 concentrations.

Table 3.1-8 gives the analysis of the gasifier circulating solids as sampled from the standpipe (referred to as standpipe solids). These solids circulate through the mixing zone, riser, and standpipe. Typically, the properties of these solids change slowly with time.

Figure 3.1-9 shows the standpipe solids SiO_2 , CaO , and Al_2O_3 contents versus run time. Due to low solids capture efficiency, the coarse standpipe gasification ash removal system operated infrequently. As material left the gasifier, sand was fed to maintain sufficient solids inventory. Since the sand additions were relatively small in size, the standpipe solids silica content declined from 91.2 to 70.8 percent at hour 156, prior to the outage. During the outage at hour 163, however, a large quantity of sand had to be added, and the silica content increased to 80.1 percent. After the outage, the silica content decline to 56.4 percent. Note as gasification ash replaced the startup sand, the silica content decreased and the CaO and Al_2O_3 contents increased.

The standpipe solids organic carbon content is plotted on Figure 3.1-10. The organic carbon is the total carbon in the solids minus inorganic carbon measured as CO_2 . Based on previous experience, the standpipe organic carbon content is an imprecise measurement because the value comes from a difference of two small values that are nearly equal. The standpipe organic carbon content was between zero and 4.7 weight percent, with one outlier at 11.9 weight percent. In general, the PRB standpipe solids carbon content in TC17 was typical of previous PRB test campaigns, while the Illinois Basin standpipe solids carbon content was higher than previous PRB test campaigns.

The standpipe solids CaS content ranged from essentially zero to 0.6 weight percent, with the higher values obtained during Illinois Basin coal feed. The standpipe CaCO_3 was between zero and 2.7 percent.

The loop seal solids samples analyses are given in Table 3.1-9. The solids from the loop seal are the solids that pass through the disengager with the syngas, but are captured in the cyclone. After the cyclone captures the loop seal solids, they flow back to the standpipe where they join the solids falling from the disengager. Due to problems with the sampling system, only five loop seal samples were taken during the test campaign.

Figure 3.1-11 shows the SiO_2 , CaO, and Al_2O_3 contents of the loop seal samples. The loop seal solids were similar to the standpipe solids in SiO_2 content. The first loop seal SiO_2 content available was 72.1 percent. The silica content then declined to 61.2 percent as gasification ash replaced the silica. The Al_2O_3 content increased from 8.3 to 12.7 over the same time period, while the CaO content fluctuated from 4.8 to 7.0 weight percent.

Figure 3.1-12 shows the organic carbon (total carbon minus CO_2 carbon) and CaCO_3 contents for the loop seal solids. The carbon content of the loop seal solids was between zero and 8 percent. The loop seal CaCO_3 remained constant at near 1.4 percent. The MgO, Fe_2O_3 , and other inerts contents are shown in Table 3.1-9. The loop seal solids CaS content was small, ranging from 0.07 to 0.65 percent.

The complete solids analysis as well as organic carbon content for the solids captured by PCD and sampled from the CFAD system is given in Table 3.1-10. In situ PCD inlet particulate samples were also analyzed. Figure 3.1-13 plots the organic carbon for the PCD solids sampled from the CFAD system as well as for the in situ samples. Only one in situ PCD sample was taken during PRB operation. Its carbon content was 33.8 percent. The in situ carbon contents ranged from 45.7 to 62.4 percent for the Illinois Basin portion of the test run. Higher carbon contents in the PCD solids were expected during Illinois Basin operation, since the coal was less volatile and the carbon conversion was generally lower. The carbon contents of the CFAD samples agreed fairly well with those of the in situ samples during the Illinois Basin portion of the test run.

Figure 3.1-14 and Table 3.1-10 show the amounts of SiO_2 and Al_2O_3 in the solids sampled from CFAD. The in situ solids concentrations for SiO_2 and Al_2O_3 are also plotted in Figure 3.1-14. The SiO_2 solids concentrations are a function of the efficiency of the disengager and cyclone as well as the SiO_2 concentration of the coal. The SiO_2 in the PCD solids originates from coal ash SiO_2 and sand SiO_2 . The SiO_2 PCD solids concentration during the beginning of the test run was 46.8 and dropped to 15.1 percent by the end of the run. This behavior is typical, since the SiO_2 content usually drops below 30 percent within a few days after startup as gasification ash replaces the startup sand. The major sand addition at hour 163 did not seem to affect the PCD silica content.

Since only a minimal amount of Al_2O_3 is in the start-up sand, the PCD solids Al_2O_3 content comes predominantly from the coal ash. The PCD solids Al_2O_3 concentration remained low, between 5.9 and 10.3 percent, for the entire test run.

Figure 3.1-15 shows the calcium carbonate and calcium sulfide concentrations in the PCD solids taken from CFAD and from in situ samples. The concentrations for CaO, CaS, and CaCO₃ are also listed in Table 3.1-10. The in situ sample CaS concentrations agreed fairly well with the CFAD solids samples calcium sulfide concentrations, while the in situ sample calcium carbonate contents did not agree well with the CFAD samples. The calcium carbonate concentration fluctuated from 1.2 to 4.3 percent during the test run. The calcium came from the PRB coal ash during PRB operation and from the Illinois Basin coal ash and dolomite during the Illinois Basin portion of the test run. CaS concentration for the PRB portion of the test run was between 0.3 and 0.4 weight percent. During the Illinois Basin testing, it ranged from 1.0 to 2.3 weight percent.

The PCD solids calcination is defined as:

$$\% \text{ Calcination} = \frac{\text{mol\% CaO}}{\text{mol\% CaO} + \text{mol\% CaCO}_3}$$

The PCD solids calcination data are plotted on Figure 3.1-16. The PCD solids calcination ranged from 88 to 92 percent during the PRB portion of the test run. These values are slightly higher than usual. During the Illinois Basin portion of the test run, calcination ranged from 30 to 91 percent.

The calcium sulfation is defined as:

$$\% \text{ Sulfation} = \frac{\text{mol\% CaS}}{\text{mol\% CaO} + \text{mol\% CaCO}_3 + \text{mol\% CaS}}$$

The PCD solids sulfation is plotted on Figure 3.1-16 with the PCD solids calcination. The PCD solids sulfation was minimal during the PRB portion of the test run, always less than 3 mole percent. During the higher sulfur Illinois Basin portion of the test run, the sulfation increased and ranged from 9 to 36 mole percent.

Solids Sample Comparison. The following comparison of the analysis of the standpipe, loop seal, and PCD solids demonstrates the solids compositions change throughout the process.

Figure 3.1-17 compares the organic carbon content of the standpipe, loop seal, and PCD solids samples. The loop seal and standpipe organic carbon contents were typically below 10 weight percent. The PCD solids carbon content ranged from 15 to 25 percent range during PRB operation and in the 40 to 60 percent range during Illinois Basin testing. In general, the loop seal organic carbon data fell between the standpipe and PCD solids data. These data indicate that the carbon is contained in small particles which are only partially captured by the disengager. The cyclone separates a slightly larger portion of the carbon, but the vast majority of the particles exit the gasifier and flow to the PCD.

Figure 3.1-18 compares the calcium concentration of the standpipe, the loop seal, and PCD solids samples. During the PRB testing, the calcium content was highest for the PCD solids, at approximately 10 weight percent. The standpipe calcium content slowly increased during the run as calcium accumulated in the gasifier. At the same time the PCD solids calcium content declined.

During the Illinois Basin portion of the test run, the calcium contents from all three locations were similar, indicating that the calcium was likely contained in particles of variable size. (Note that the calcium is distributed between the compounds CaO, CaCO₃, and CaS.)

The silica entering the process primarily remains in the gasifier, since the sand particle size of about 120 microns is greater than that of the standpipe solids. Figure 3.1-19 shows that the standpipe solids and loop seal had similar silica contents. The PCD solids had much lower silica content. The samples from all three locations exhibited the same trend of slowly declining silica content. As the gasifier operates, sand eventually breaks down by attrition and exits through the cyclone before being collected by the PCD. The data suggest that the sand loss through the cyclone was minimal, and the replacement of sand with bed material was a slow process. The silica inventory in the gasification loop did not appear to reach steady state.

Solids Particle Size. The TC17 Sauter mean diameter (SMD) and mass mean diameter (MMD) particle sizes of the coal sampled from the original coal feeder are plotted on Figure 3.1-20. The PRB coal SMD particle size averaged 271 microns and was fairly steady, with a standard deviation of only 30 microns. At the beginning of Illinois Basin testing, the average SMD was 190 microns with a standard deviation of 12 microns. Due to changes in coal mill operation after the outage at hour 163, the SMD averaged 251 microns, with a standard deviation of 13 microns.

Figure 3.1-21 shows both the coal feed percent below 45 microns (fines) and percent above 1,180 microns (coarse particles). During TC17, the percent fines was acceptable at below 8 percent, except for samples above 18 percent taken around hour 320.

A large amount of 1,180 micron particles increases the difference between the SMD and the MMD since the SMD is a surface area average. Therefore, the larger particles with less surface area per pound have a lesser effect on the SMD than the MMD, where the larger particles skew the MMD due to their higher weight per particle. The average percent above 1,180 microns for the PRB portion of the test run was 8.5 percent with a standard deviation of 2.9. During the first part of Illinois Basin testing, the average percent above 1,180 was only 3.3 percent with a standard deviation of 0.5. After the outage at hour 163, the average increased to 7.7 percent with a standard deviation of 1.4. The high spikes in coarse solids correspond to the high MMD particle size.

The TC17 standpipe solids particle sizes are given in Figure 3.1-22. The PRB standpipe solids particle sizes slowly increased as gasification ash replaced the startup sand, eventually achieving an average particle size of 289 microns. The average particle size of the standpipe solids was around 127 microns MMD and 152 microns SMD during PRB operation and 183 microns MMD and 207 microns SMD during Illinois Basin operation.

The percent of gasifier solids greater than 600 micron particles and the percent less than 45 microns are plotted on Figure 3.1-23. The gasifier solids had an average coarse particle (greater than 600 microns) content of less than 1.0 percent during the first 70 hours of the test run. The value fluctuated from 2.8 to 22.37 percent during the remainder of the run (which included all Illinois Basin testing). The average fines content (less than 45 microns) was less than 1.0 percent during the entire test run.

For some of the previous test campaigns, the gasifier circulating solids achieved a steady particle size, typically between 165 and 230 microns SMD as shown in Table 3.1-11. By comparison, the startup sand is around 120 microns SMD. For tests that reached steady state the standpipe particle size slowly increased asymptotically to reach the steady state value. During PRB operation, the gasifier solids appeared to reach a steady state average particle size of around 155 microns SMD. During Illinois Basin operation, the gasifier solids did not reach steady state conditions.

The particle sizes of the loop seal solids are as shown in Figure 3.1-24. Both the SMD and the MMD of the loop seal solids varied widely. The SMD ranged from 56.9 to 133 microns, while the MMD varied from 110 to 134.6 microns.

Figure 3.1-25 plots the SMD and MMD for the PCD solids sampled from the CFAD system as well as the eight in situ samples collected at the PCD inlet. Five of the eight in situ solids particle sizes agreed with the particle size of the solids collected from the CFAD system. Sand addition may have caused the disagreement with the other three samples. In order to collect a representative sample of gasification ash, most of the in situ samples were collected during periods when sand was not being added to the gasifier. The CFAD samples were often taken at times when sand had been added.

The PCD solids SMD started high at around 114 microns due to the startup conditions, but dropped quickly. The average SMD during PRB operation was 9.4 microns with a standard deviation of 2.2. The average SMD during Illinois Basin operation was 15.2 microns with a standard deviation of 2.3. The MMD was about seven microns larger than the SMD for all of the samples taken and followed the same trends as the SMD particle size. The PCD solids particle size during PRB operation was consistent with the particle size of previous PRB test campaigns as shown in Table 3.1-11. The particle size for the Illinois Basin bituminous coal was slightly higher than the PRB, but similar to that of the Alabama and Hiawatha bituminous coals.

Particle Size Comparison. Figure 3.1-26 plots the solids SMD particle sizes, including coal and standpipe, loop seal, and PCD solids. The coal fed to the gasifier ranged between 153 to 358 microns SMD. These values were usually slightly higher than the standpipe solids which ranged from 145 to 289 microns. Note that that average particle sizes of the standpipe samples taken at hours 259, 303, and 307 slightly exceeded the respective coal samples. The loop seal solids ranging from 56.9 to 134.1 were always more coarse than the PCD solids (at an average of under 14 microns), but less coarse than the standpipe solids.

Standpipe and PCD Solids Bulk Densities. The standpipe, loop seal, PCD in situ, and CFAD solids bulk densities are given in Figure 3.1-27. As previously mentioned, the standpipe solids bulk density remained constant due to the frequent sand additions that were necessary. The standpipe solids bulk density averaged 84 lb/ft³ during PRB operation and 81 lb/ft³ during Illinois Basin operation. Both values are close to that of sand. The data for the loop seal solids averaged 71 lb/ft³, but varied considerably more than the standpipe solids data, occasionally attaining higher values than the standpipe. The PCD solids had the lowest average values, 20.4 lb/ft³ during PRB testing and 13.9 lb/ft³ during Illinois Basin operation. All of the in situ PCD data points were in the general range of the CFAD sample data.

The minimum standpipe solids densities for past PSDF gasification test campaigns are shown in Table 3.1-11. Also listed are the average SMD particle size and standard deviation of the PCD solids for all previous gasification test campaigns.

3.1.5 Carbon Conversion

Carbon conversion is defined as the percent of fuel carbon that is gasified to CO, CO₂, CH₄, C₂H₆, and higher hydrocarbons versus the amount of carbon that exits the gasifier with the PCD and gasifier solids. In a typical flow sheet for integrated gasification combine cycle (IGCC), the carbon in the solids collected from the gasifier or PCD is burned in a combustor or sent to a landfill for disposal.

While carbon conversion can be calculated several different ways, the most accurate method divides the carbon content in the syngas by the total carbon exiting the gasifier (from both solid and gas streams). Table 3.1-12 gives the carbon conversions for the test periods, while Figure 3.1-28 shows the carbon conversion versus time. The carbon conversion ranged from 94.7 to 96.7 percent during PRB operation and from 80.4 to 88.5 percent during Illinois Basin testing. The carbon conversion was lower during the Illinois Basin coal operation due to the lower reactivity of the high rank fuel.

The average carbon conversions of PRB, Illinois Basin, Hiawatha, Falkirk lignite, and Freedom lignite are compared on Figure 3.1-29. All of the fuels except Illinois Basin show separate values for air and oxygen blown operation. (No oxygen blown testing has occurred using Illinois Basin coal). These data came from Test Runs TC06 through TC17, with the exception of TC14 data, since the poor performance of the solids collection systems in TC14 caused abnormally low carbon conversion. The low temperature Freedom lignite carbon conversion data are plotted separately from the high temperature Freedom lignite carbon conversion data to illustrate that significantly lower temperatures adversely affect the carbon conversion.

For all fuels with oxygen blown testing, air blown operation yielded a slightly higher carbon conversion than oxygen blown operation. The reason for the lower carbon conversion was that previously, the oxygen system at the PSDF only supplied oxygen to the lower mixing zone, causing the temperatures to be higher in that region and lower through the rest of the gasifier. The lower temperatures in the upper mixing zone and riser lowered the carbon conversion. Future tests with oxygen to the upper mixing zone should improve the carbon conversion during oxygen blown operation.

Falkirk lignite had the highest average carbon conversion of the four coals tested. PRB and Freedom lignite (at higher temperatures) had about the same average carbon conversion, with Hiawatha bituminous coal slightly lower. Illinois Basin coal had the lowest average carbon conversion for the conditions studied. Although the data in Figure 3.1-29 show general trends in carbon conversion over test runs, the values obtained are the result of operating over a small range of conditions for all fuels except PRB coal.

3.1.6 Gasification Efficiencies

Gasification efficiency is defined as the percentage of the entering energy that is converted to potentially useful syngas energy. Two types of gasification efficiencies have been defined: the cold gasification efficiency and the hot gasification efficiency. The cold gasification efficiency is the percentage of energy fed that is available to a gas turbine as syngas latent heat. The hot gasification efficiency is the percentage of total energy fed that is available to produce electricity. The total energy to produce electricity includes the syngas latent heat recovered in a gas turbine plus the sensible heat recovered in a heat recovery steam generator.

The cold gasification efficiency, Eff_c , is calculated by:

$$Eff_c = \frac{G * LHV}{G * (LHV + H_{sg}) + FP * (Q_{fp} + H_{fp}) + CP(Q_{cp} + H_{cp}) + Q_{loss}}$$

where G is the syngas rate; LHV is the syngas lower heating value; FP is the fine particle rate (from the PCD); CP is the coarse particle rate (from the standpipe); Q is the latent heat of fine particles (fp), coarse particles (cp), or syngas (sg); H is the sensible heat of fine particles (fp), coarse particles (cp), or syngas (sg); and Q_{loss} is the gasifier heat loss.

The hot gasification efficiency, Eff_H , is calculated as:

$$Eff_H = \frac{G * (LHV + H_{sg})}{G * (LHV + H_{sg}) + FP * (Q_{fp} + H_{fp}) + CP(Q_{cp} + H_{cp}) + Q_{loss}}$$

The cold gasification efficiency is plotted in Figure 3.1-30 and is listed in Table 3.1-12. Note that in previous two run reports (TC15 and TC16), the cold gasification efficiency was defined as the amount of latent heat in the syngas divided by the total amount of coal latent heat entering the gasifier in order to more closely conform to industry standards as established in ASME Performance Test Code PTC-16 for determining cold gasification efficiency. Since PTC-16 has since been withdrawn, this report (along with the TC06-TC14 reports) defines cold gasification efficiency as the syngas latent heat divided by the total amount of energy entering the gasifier, thus including the air and steam sensible heats. This practice is commonplace in the gasification industry and provides a better indication of the overall process efficiency. Thus, the values reported for test runs prior to TC15 are artificially higher than those in this report.

The cold gasification efficiencies were as high as 53.7 percent during the test run. During the PRB operating periods, the cold gasification efficiencies ranged from 53.0 to 53.7 percent. The Illinois Basin periods had cold gasification efficiencies of between 35.5 percent and 41.6 percent. Since the gasifier operated at higher temperatures during the Illinois Basin testing to improve carbon conversion, the heat losses were higher and the gasifier less efficient.

The hot gasification efficiency assumes that the sensible heat of the syngas can be recovered in a heat recovery steam generator, so the hot gasification efficiency is always higher than the cold

gasification efficiency. The hot gasification efficiency is the latent and the sensible heat of the syngas exiting the gasifier divided by the total amount of energy entering the gasifier, including the latent heat of the coal and the sensible heats of the air and steam. The hot gasification efficiency is plotted in Figure 3.1-31 and shown in Table 3.1-12. The efficiency boundary for the values found in Table 3.1-12 and Figure 3.1-31 is the gasifier itself, not including downstream equipment. The PRB hot gasification efficiencies were between 84.2 and 85.6 percent, while the Illinois Basin periods had hot gasification efficiencies between 70.3 and 78.1 percent.

The two main sources of efficiency losses are the gasifier heat loss and the latent heat of the PCD solids. The gasifier heat loss of 3.5 million Btu/hr was about 12 percent of the feed energy, while the total energy of the PCD solids was from 3 to 20 percent of the feed energy (with the higher values occurring during the periods of low coal feed rate).

A commercial gasifier will be more efficient than the PSDF gasifier due to the use of recycle gas and lower heat losses. The heat loss as a percentage of energy fed will be much smaller in a commercial size gasifier. While the Transport Gasifier does not recover the latent heat of the PCD solids, this latent heat could be recovered in a combustor. The total enthalpy of the PCD solids can be decreased by decreasing both the PCD solids carbon content (heating value) and the PCD solids rate (by improving solids collection efficiency).

Gasification efficiencies can be calculated from the commercially projected gas heating values and adjusted flow rates that were determined when calculating the projected heating value. The required adjustments for projected flow rates are described in Appendix 5, LHV Projection Calculations. The main adjustment is that the syngas rate decreases per pound of coal gasified due to the elimination of non-air nitrogen in the syngas and increased efficiencies due to no heat loss. The commercially projected cold gasification efficiencies for all of the operating periods are listed in Table 3.1-12.

The projected cold gasification efficiency, Eff_{CP} , is calculated by:

$$Eff_{CP} = \frac{LHV_p * G_p}{G_p * (LHV_p + H_{sgp}) + FP * (Q_{fp} + H_{fp}) + CP(Q_{cp} + H_{cp})}$$

where LHV_p is the projected syngas LHV; G_p is the projected syngas rate; and H_{sgp} is the projected syngas sensible heat.

The corrected efficiencies are calculated assuming an adiabatic gasifier, since zero heat loss was one of the assumptions in determining the corrected LHV in Section 3.1.3. When projected for a commercial plant, the cold gas efficiencies were from 69.8 to 71.3 percent during PRB operation, and between 53.6 and 55.9 percent during Illinois Basin operation. The average projected cold gas efficiency for PRB averaged 70.2 percent. Illinois Basin projections averaged 53.2 percent. The commercially projected efficiencies were higher than the observed cold gasification efficiencies by about 18 percent during PRB testing and 15 percent during Illinois Basin testing. The use of recycle gas and the lower heat losses in the commercial projection are the main factors that increase the efficiency.

Table 3.1-1 Selected Steady State Operating Periods

Operating Period	Start Time	End Time	Duration Hours	Average Time	Run Time Hours
TC17-1	10/27/04 17:30	10/27/04 20:45	3:15	10/27/04 19:07	29
TC17-2	10/28/04 2:00	10/28/04 5:15	3:15	10/28/04 3:37	38
TC17-3	10/28/04 5:30	10/28/04 8:30	3:00	10/28/04 7:00	41
TC17-4	10/28/04 21:00	10/29/04 1:00	4:00	10/28/04 23:00	57
TC17-5	10/29/04 5:45	10/29/04 9:15	3:30	10/29/04 7:30	65
TC17-6	10/29/04 20:00	10/29/04 23:15	3:15	10/29/04 21:37	80
TC17-7	10/30/04 1:30	10/30/04 4:15	2:45	10/30/04 2:52	85
TC17-8	10/31/04 1:30	10/31/04 9:15	7:45	10/31/04 5:22	111
TC17-9	10/31/04 12:30	10/31/04 15:15	2:45	10/31/04 13:52	120
TC17-10	10/31/04 20:30	11/1/04 0:00	3:30	10/31/04 22:15	128
TC17-11	11/2/04 5:30	11/2/04 8:00	2:30	11/2/04 6:45	161
TC17-12	11/12/04 9:30	11/12/04 13:30	4:00	11/12/04 11:30	173
TC17-13	11/12/04 16:00	11/12/04 22:30	6:30	11/12/04 19:15	180
TC17-14	11/13/04 1:00	11/13/04 5:45	4:45	11/13/04 3:22	188
TC17-15	11/13/04 14:15	11/13/04 17:45	3:30	11/13/04 16:00	201
TC17-16	11/13/04 18:15	11/14/04 9:00	14:45	11/14/04 1:37	211
TC17-17	11/14/04 11:00	11/15/04 1:30	14:30	11/14/04 18:15	227
TC17-18	11/15/04 4:30	11/15/04 8:45	4:15	11/15/04 6:37	240
TC17-19	11/15/04 15:00	11/15/04 22:00	7:00	11/15/04 18:30	251
TC17-20	11/15/04 22:00	11/16/04 5:15	7:15	11/16/04 1:37	259
TC17-21	11/16/04 8:30	11/16/04 12:00	3:30	11/16/04 10:15	267
TC17-22	11/16/04 17:00	11/17/04 3:30	10:30	11/16/04 22:15	279
TC17-23	11/17/04 3:30	11/17/04 13:45	10:15	11/17/04 8:37	290
TC17-24	11/17/04 15:45	11/18/04 0:00	8:15	11/17/04 19:52	301
TC17-25	11/18/04 8:30	11/18/04 12:30	4:00	11/18/04 10:30	315

Notes:

- TC17-1 through TC17-7 used PRB coal.
 TC17-8 through TC17-25 used Illinois Basin coal.

Table 3.1-2 Operating Conditions

Operating Periods ¹	Average Relative Hours	Cyclone Exit Zone °F	System Pressure psig	Coal Rate lb/hr	Dolomite Rate lb/hr	Air Rate lb/hr	Syngas Rate lb/hr	Steam Rate ² lb/hr	Nitrogen Rate lb/hr	PCD Inlet Temperature °F	PCD Solids Rate pph
TC17-1	29	1722	220	3,300	0	11,700	21,600	1,400	5,800	744	360
TC17-2	38	1717	220	3,400	0	11,900	22,200	1,400	5,800	740	360
TC17-3	41	1704	220	4,000	0	11,800	22,000	1,100	5,900	735	360
TC17-4	57	1673	220	3,800	0	12,000	22,500	1,300	6,000	739	360
TC17-5	65	1679	220	3,300	0	11,500	21,700	1,400	5,900	730	360
TC17-6	80	1687	220	3,600	0	12,000	22,000	1,300	6,000	735	360
TC17-7	85	1685	220	3,400	0	11,600	20,500	1,400	4,900	725	360
TC17-8	111	1753	230	2,300	200	9,900	19,100	2,300	5,200	875	410
TC17-9	120	1760	230	2,300	0	10,100	19,200	2,300	5,200	881	410
TC17-10	128	1763	230	2,500	280	10,700	20,100	2,600	5,100	911	410
TC17-11	161	1792	232	2,500	370	11,400	21,700	3,100	4,700	892	410
TC17-12	173	1802	214	2,300	220	10,700	19,000	2,000	5,200	865	610
TC17-13	180	1804	214	2,200	220	10,900	19,500	2,100	5,400	890	580
TC17-14	188	1812	214	2,200	230	11,000	19,900	2,200	5,500	908	550
TC17-15	201	1818	220	2,500	230	11,500	20,500	2,200	5,500	921	520
TC17-16	211	1818	226	2,400	230	11,500	20,300	2,200	5,400	930	500
TC17-17	227	1819	226	2,500	240	11,400	20,100	2,100	5,300	953	470
TC17-18	240	1802	226	2,600	240	11,200	20,000	2,000	5,600	957	450
TC17-19	251	1820	226	2,500	240	11,100	19,800	1,900	5,600	973	460
TC17-20	259	1820	226	2,400	230	11,200	19,900	1,900	5,700	984	480
TC17-21	267	1820	226	1,900	220	10,200	18,300	2,000	5,100	973	500
TC17-22	279	1821	226	2,500	240	11,000	19,000	1,500	5,400	970	510
TC17-23	290	1819	226	2,500	120	10,900	18,900	1,600	5,300	971	510
TC17-24	301	1829	226	2,500	120	11,100	19,100	1,600	5,300	978	490
TC17-25	315	1822	226	2,400	70	10,700	18,500	1,600	5,000	1,009	450

Notes:

1. TC17-1 to TC17-7 were Powder River Basin coal; TC17-8 to TC17-25 were Illinois Basin coal.
2. All steam rates by hydrogen balance

Table 3.1-3 Wet Gas Composition, Molecular Weight, and Heating Value

Operating Period ¹	Average Relative Hour	H ₂ O ² Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	C ₂ H ₆ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Wet Syngas LHV Btu/SCF	Dry Syngas TRS ³ ppm	Wet Syngas MW lb/mole
TC17-1	29	11.6	6.8	7.8	9.2	0.9	0.0	0.5	63.2	100.0	52	256	26.2
TC17-2	38	11.2	7.1	7.8	9.2	1.0	0.0	0.5	63.3	100.0	54	267	26.3
TC17-3	41	10.8	7.2	7.9	9.2	1.1	0.0	0.5	63.3	100.0	55	276	26.3
TC17-4	57	10.7	7.1	7.9	9.3	1.3	0.0	0.5	63.1	100.0	57	272	26.3
TC17-5	65	11.1	6.6	7.7	9.1	1.1	0.0	0.5	63.9	100.0	52	230	26.3
TC17-6	80	11.1	6.9	8.0	9.3	1.1	0.0	0.5	63.1	100.0	54	257	26.2
TC17-7	85	11.4	7.4	8.4	9.6	1.1	0.0	0.5	61.5	100.0	57	249	26.1
TC17-8	111	18.5	2.9	5.0	9.1	1.1	0.0	0.5	62.9	100.0	33	683	26.3
TC17-9	120	18.7	3.0	5.1	9.1	1.1	0.0	0.5	62.5	100.0	34	727	26.2
TC17-10	128	19.8	2.9	5.2	9.3	1.1	0.0	0.5	61.0	100.0	34	759	26.1
TC17-11	161	20.9	3.2	5.5	9.6	1.1	0.0	0.5	59.2	100.0	35	850	26.0
TC17-12	173	15.9	4.1	5.5	9.2	1.2	0.0	0.5	63.6	100.0	39	621	26.4
TC17-13	180	15.8	4.1	5.5	9.1	1.2	0.0	0.5	63.8	100.0	39	675	26.4
TC17-14	188	16.1	4.0	5.5	9.2	1.2	0.0	0.5	63.6	100.0	39	715	26.4
TC17-15	201	15.7	4.5	5.9	9.2	1.2	0.0	0.5	63.0	100.0	42	737	26.3
TC17-16	211	15.6	4.5	5.9	9.1	1.3	0.0	0.5	63.0	100.0	42	750	26.3
TC17-17	227	15.1	4.8	6.1	9.1	1.3	0.0	0.5	63.1	100.0	44	769	26.3
TC17-18	240	14.6	4.6	5.8	9.0	1.3	0.0	0.5	64.1	100.0	43	692	26.4
TC17-19	251	14.3	4.7	5.7	9.0	1.3	0.0	0.5	64.5	100.0	42	707	26.4
TC17-20	259	14.5	4.5	5.6	9.0	1.2	0.0	0.5	64.7	100.0	41	713	26.5
TC17-21	267	16.0	3.9	5.1	9.3	1.0	0.0	0.5	64.1	100.0	36	716	26.5
TC17-22	279	12.4	5.6	6.0	8.8	1.3	0.0	0.5	65.4	100.0	46	744	26.5
TC17-23	290	12.5	5.6	6.0	8.9	1.3	0.0	0.5	65.2	100.0	46	724	26.5
TC17-24	301	12.2	5.9	6.1	8.8	1.3	0.0	0.5	65.2	100.0	48	734	26.5
TC17-25	315	13.6	5.1	5.9	9.1	1.2	0.0	0.5	64.5	100.0	44	747	26.5

Notes:

1. TC17-1 to TC17-7 were Powder River Basin coal; TC17-8 to TC17-25 were Illinois Basin coal.
2. The H₂O concentration was estimated using a correlation between the in-situ samples and gasifier temperature.
3. Syngas total reduced sulfur (TRS) estimated from syngas combustor SO₂ analyzer data.

Table 3.1-4 Syngas Total Reduced Sulfur Concentration and Sulfur Removal

Operating Period ¹	Average Relative Hour	Total Reduced Sulfur ² ppm	Sulfur Removal %
TC17-1	29	226	7.8
TC17-2	38	238	8.5
TC17-3	41	246	8.7
TC17-4	57	243	9.2
TC17-5	65	205	10.3
TC17-6	80	228	8.0
TC17-7	85	220	7.9
TC17-8	111	557	15.2
TC17-9	120	591	13.4
TC17-10	128	609	10.9
TC17-11	161	672	9.4
TC17-12	173	523	30.5
TC17-13	180	569	26.7
TC17-14	188	600	23.6
TC17-15	201	621	21.9
TC17-16	211	633	21.8
TC17-17	227	653	21.3
TC17-18	240	591	22.9
TC17-19	251	606	23.6
TC17-20	259	610	24.3
TC17-21	267	601	26.4
TC17-22	279	652	25.0
TC17-23	290	634	26.0
TC17-24	301	645	24.7
TC17-25	315	645	23.6

Notes:

1. TC17-1 to TC17-7 were Powder River Basin coal; TC17-8 to TC17-25 were Illinois Basin.
2. Syngas total reduced sulfur (TRS) calculated from syngas combustor SO₂ analyzer data.

Table 3.1-5 Projected¹ Syngas Composition, Molecular Weight, and Heating Value

Operating Period ²	Average Relative Hour	H ₂ O Mole %	CO Mole %	H ₂ Mole %	CO ₂ Mole %	CH ₄ Mole %	Argon Mole %	N ₂ Mole %	Total Mole %	Syngas MW lb/mole	Syngas CO/CO ₂ Ratio	Syngas LHV ¹ Btu/SCF
TC17-1	29	0.8	15.6	17.7	12.5	2.1	0.6	50.7	100.0	25.2	1.2	118
TC17-2	38	0.8	16.4	17.8	12.5	2.4	0.6	49.6	100.0	25.1	1.3	123
TC17-3	41	0.8	16.6	17.7	12.3	2.5	0.6	49.5	100.0	25.1	1.4	125
TC17-4	57	0.8	16.7	17.5	12.4	3.1	0.6	48.9	100.0	25.1	1.3	130
TC17-5	65	0.8	16.0	18.0	12.5	2.6	0.6	49.6	100.0	25.0	1.3	125
TC17-6	80	0.8	15.9	17.5	12.3	2.5	0.6	50.6	100.0	25.1	1.3	122
TC17-7	85	0.8	16.1	17.4	12.3	2.3	0.6	50.5	100.0	25.2	1.3	121
TC17-8	111	0.8	8.1	15.7	16.0	3.3	0.7	55.5	100.0	26.1	0.5	100
TC17-9	120	0.8	8.1	15.5	15.8	3.2	0.7	55.9	100.0	26.1	0.5	98
TC17-10	128	0.8	7.5	14.7	15.9	3.1	0.7	57.4	100.0	26.4	0.5	93
TC17-11	161	0.8	7.7	14.6	16.6	2.8	0.7	56.7	100.0	26.5	0.5	91
TC17-12	173	0.8	9.3	14.5	14.2	3.0	0.7	57.5	100.0	26.2	0.7	97
TC17-13	180	0.8	9.4	14.8	14.2	3.0	0.7	57.1	100.0	26.1	0.7	99
TC17-14	188	0.8	9.2	14.6	14.4	2.9	0.7	57.3	100.0	26.2	0.6	97
TC17-15	201	0.8	9.8	14.8	14.0	2.9	0.7	56.9	100.0	26.1	0.7	99
TC17-16	211	0.8	9.9	15.0	13.9	3.0	0.7	56.8	100.0	26.0	0.7	100
TC17-17	227	0.8	10.4	15.1	13.7	3.0	0.7	56.4	100.0	25.9	0.8	102
TC17-18	240	0.8	10.3	15.1	13.6	3.2	0.7	56.3	100.0	25.9	0.8	104
TC17-19	251	0.8	10.5	14.9	13.6	3.1	0.7	56.5	100.0	26.0	0.8	103
TC17-20	259	0.8	10.3	14.8	13.7	3.0	0.7	56.8	100.0	26.0	0.7	101
TC17-21	267	0.8	9.1	14.3	14.7	2.6	0.7	57.8	100.0	26.3	0.6	93
TC17-22	279	0.8	12.0	15.1	12.5	3.1	0.7	55.9	100.0	25.7	1.0	108
TC17-23	290	0.8	12.1	15.0	12.7	3.1	0.7	55.8	100.0	25.8	1.0	108
TC17-24	301	0.8	12.4	14.9	12.4	2.9	0.7	56.0	100.0	25.8	1.0	108
TC17-25	315	0.8	11.2	14.8	13.3	2.9	0.7	56.4	100.0	26.0	0.8	103

Notes:

- Adjustments are based on the following assumptions: that only air nitrogen is in the syngas, the gasifier is adiabatic, and syngas is at the turbine inlet after the syngas cleanup processes.
- TC17-1 to TC17-7 were Powder River Basin coal; TC17-8 to TC17-25 were Illinois Basin coal.

Table 3.1-6 Coal Analysis

	Powder River Basin		Illinois Basin	
	Average Value	Standard Deviation	Average Value	Standard Deviation
Moisture, wt%	21.89	0.32	10.05	0.61
Carbon, wt%	54.83	0.67	67.22	0.81
Hydrogen, wt%	3.46	0.20	4.23	0.06
Nitrogen, wt%	0.73	0.03	1.47	0.04
Oxygen, wt%	13.91	0.40	8.19	0.81
Sulfur, wt%	0.28	0.02	0.86	0.06
Ash, wt%	4.89	0.71	7.98	0.85
Volatiles, wt%	32.65	3.53	31.26	4.30
Fixed Carbon, wt%	40.57	3.49	50.71	3.77
Higher Heating Value, Btu/lb	9,142	139	11,651	112
Lower Heating Value, Btu/lb	8,821	136	11,260	111
CaO, wt %	1.01	0.26	0.32	0.04
SiO ₂ , wt %	1.38	0.23	4.33	0.59
Al ₂ O ₃ , wt %	0.71	0.13	1.75	0.17
MgO, wt %	0.25	0.07	0.11	0.02
Fe ₂ O ₃ , wt %	0.27	0.03	0.62	0.07
Na, wt % in ash	1.26	0.09	1.06	0.10
Ca/S, mole/mole	2.21	0.79	0.20	0.01

Notes:

1. All analyses are as sampled at FD0210 coal feeder.
2. Hydrogen in coal is reported separately from hydrogen in moisture.
3. Oxygen calculated by difference.

Table 3.1-7 Sorbent Analysis

Compound	Dolomite	
	Weight %	Standard Deviation
CaCO ₃ , wt %	52.6	1.40
MgCO ₃ , wt %	42.3	0.59
CaSO ₄ , wt%	0.3	0.05
SiO ₂ , wt %	1.3	0.20
Al ₂ O ₃ , wt %	0.3	0.11
Other inerts ²	0.6	0.07
H ₂ O, wt %	0.3	0.01
Total	97.8	

Notes:

1. All samples taken from FD0220 sorbent feeder.
2. Other inerts consist of P₂O₅, Na₂O, K₂O, Fe₂O₃, and TiO₂.

Table 3.1-8 Standpipe Solids Analysis

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	FeO Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon ² Wt. %	Total Wt. %
AB16078	10/26/2004 18:00	4	91.2	3.1	0.5	2.7	0.5	0.0	1.3	0.7	0.0	99.9
AB16118	10/27/2004 18:00	28	87.9	4.0	1.0	2.8	0.3	0.0	3.1	0.8	0.1	99.9
AB16154	10/29/2004 10:00	68	85.1	5.0	1.2	3.1	0.0	0.0	4.3	0.9	0.9	100.4
AB16175	10/30/2004 10:00	92	80.6	5.2	1.2	3.2	0.3	0.2	4.3	1.5	1.4	97.9
AB16176	10/30/2004 18:00	100	77.4	5.3	1.2	2.9	0.4	0.2	4.2	1.3	2.5	95.2
AB16192	10/31/2004 10:00	116	80.5	6.6	1.6	2.9	0.5	0.1	4.5	1.8	0.3	98.9
AB16212	11/1/2004 10:00	140	72.8	8.9	2.3	3.8	0.0	0.0	7.3	3.7	0.8	99.6
AB16226	11/1/2004 18:00	148	71.2	9.9	2.5	3.9	2.5	0.4	6.2	3.8	0.0	100.3
AB16227	11/2/2004 2:00	156	70.8	9.5	2.5	3.2	2.7	0.6	6.3	4.3	0.0	100.0
AB16268	11/11/2004 2:00	169	79.8	7.7	1.7	3.6	0.5	0.1	3.8	2.0	0.3	99.6
AB16278	11/12/2004 13:00	174	80.1	7.8	1.9	2.8	0.0	0.1	4.0	2.2	1.6	100.4
AB16302	11/13/2004 18:00	203	75.6	8.6	2.8	3.1	0.3	0.1	4.6	2.7	1.0	98.8
AB16305	11/14/2004 18:00	227	62.6	13.2	4.0	3.7	0.0	0.2	5.6	3.4	3.8	96.4
AB16307	11/15/2004 10:00	243	65.3	13.3	3.6	4.3	0.0	0.2	5.0	2.8	0.9	95.4
AB16331	11/16/2004 2:00	259	63.3	13.7	4.2	4.3	0.0	0.2	5.8	3.6	0.9	95.9
AB16361	11/17/2004 2:00	283	61.1	16.3	4.4	4.3	0.3	0.3	5.9	3.6	11.2	107.4
AB16381	11/17/2004 22:00	303	57.5	17.7	4.7	4.9	0.0	0.3	6.5	4.0	4.7	100.3
AB16391	11/18/2004 10:00	315	56.4	17.3	5.0	5.4	0.0	0.3	7.2	4.2	1.6	97.2

Notes:

1. Other inerts consist of P₂O₅, Na₂O, K₂O, & TiO₂.
2. Organic carbon was below the detectible limit when 0% was measured.

Table 3.1-9 Loop Seal Solids Analysis

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	FeO Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB16213	11/1/2004 10:00	140	73.7	8.4	2.4	3.1	1.4	0.1	7.0	3.7	1.4	101.1
AB16277	11/12/2004 13:00	174	77.6	7.5	1.9	3.5	1.4	0.4	4.1	3.0	5.2	104.6
AB16311	11/14/2004 2:00	211	66.1	12.6	3.9	3.3	1.3	0.7	6.0	4.4	8.0	106.3
AB16312	11/14/2004 18:00	227	68.7	12.8	3.5	3.7	1.4	0.6	4.6	3.2	8.3	106.6
AB16328	11/15/2004 18:00	251	66.1	13.7	3.8	3.8	1.4	0.6	5.2	3.6	5.3	103.5

Notes:

1. Other inerts consist of P₂O₅, Na₂O, K₂O, & TiO₂.

Table 3.1-10 PCD Solids Analysis from CFAD Samples

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO ₂ Wt. %	Al ₂ O ₃ Wt. %	FeO Wt. %	Other Inerts ¹ Wt. %	CaCO ₃ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic C (C-CO ₂) Wt. %	Total Wt. %	HHV Btu/lb	LHV Btu/lb
AB16123	10/27/2004 22:00	32	46.8	10.3	3.1	3.6	2.1	0.3	12.7	3.3	15.4	97.7	2,419	2,393
AB16125	10/28/2004 6:00	40	41.6	10.0	3.0	3.6	2.9	0.4	11.4	3.1	22.2	98.2	3,393	3,360
AB16146	10/28/2004 22:00	56	39.7	8.7	3.0	3.2	2.7	0.4	11.7	3.1	25.5	98.1	3,798	3,759
AB16165	10/29/2004 22:00	80	39.9	9.0	3.0	3.5	2.9	0.3	11.7	3.1	25.3	98.7	3,771	3,735
AB16167	10/30/2004 6:00	88	36.5	7.7	2.5	2.9	3.0	0.3	16.0	7.2	21.8	97.8	3,189	3,159
AB16189	10/31/2004 10:00	116	29.2	6.3	2.0	2.2	1.2	1.3	3.9	3.4	49.2	98.6	7,267	7,223
AB16202	10/31/2004 22:00	128	22.5	6.3	1.9	1.9	1.3	1.0	2.4	2.6	56.5	96.4	8,556	8,508
AB16232	11/2/2004 6:00	160	26.5	6.7	2.1	2.1	1.2	1.0	7.3	5.7	44.0	96.8	6,593	6,553
AB16280	11/12/2004 14:00	175	28.3	6.8	1.9	2.3	1.8	2.0	3.9	4.1	45.9	96.9	7,180	7,146
AB16286	11/13/2004 6:00	191	24.4	6.2	1.9	2.2	1.8	1.8	3.4	3.6	51.3	96.7	7,848	7,812
AB16290	11/13/2004 22:00	207	21.3	6.2	1.8	1.8	1.8	2.0	2.9	3.7	57.1	98.6	8,707	8,667
AB16296	11/14/2004 22:00	231	18.8	6.5	2.0	1.8	1.9	2.1	2.5	3.5	58.1	97.2	8,669	8,630
AB16322	11/15/2004 18:00	251	19.2	6.2	1.9	2.0	2.2	2.2	2.3	3.7	58.8	98.5	8,764	8,726
AB16324	11/16/2004 2:00	259	18.5	6.2	2.0	1.7	2.9	2.2	2.1	3.8	57.4	96.7	8,712	8,672
AB16358	11/16/2004 14:00	271	18.7	6.5	2.1	1.9	2.8	2.1	2.7	4.1	57.3	98.2	8,547	8,510
AB16366	11/16/2004 22:00	279	15.1	5.9	1.7	1.6	1.2	2.2	2.4	3.5	62.9	96.6	9,338	9,290
AB16379	11/18/2004 2:00	307	15.1	6.0	1.7	1.7	3.2	2.3	1.5	3.6	60.5	95.7	9,083	9,041
AB16390	11/18/2004 10:00	315	15.9	6.5	1.9	1.8	4.3	2.2	1.0	3.7	60.7	98.0	9,022	8,982

Notes:

1. Other inerts consist of P₂O₅, Na₂O, K₂O, & TiO₂.

Table 3.1-11 Historical Standpipe and PCD Solids

Test Campaign	Fuel	Standpipe			PCD Fines			
		Maximum	Steady		Average	St. Dev.		
		Particle	State	Minimum	Particle	Particle	Average	St. Dev.
		Size	Part. Size	Bulk	Size	Size	Bulk	Bulk
		SMD	SMD	Density	SMD	SMD	Density	Density
microns	microns	lb/ft ³	microns	microns	lb/ft ³	lb/ft ³		
TC06	Powder River Basin	204	165	80	10.8	1.1	24	4
TC07	Powder River Basin	191	175	80	10.2	1.1	28	8
TC07	Alabama Bituminous	232	none	66	16.2	3.2	32	7
TC08	Powder River Basin	250	205	77	13.1	3.2	25	7
TC09	Hiawatha Bituminous	233	180	76	15.7	4.6	29	12
TC10	Powder River Basin	280	none	76	10.7	3.6	23	7
TC11	Falkirk Lignite	200	200	75	12.3	2.4	36	3
TC12	Powder River Basin	300	none	76	9.8	2	18	6
TC13	Powder River Basin	165	165	81	10.4	1.4	18	4
TC13	Freedom Lignite Low Sodium	230	none	56	15.3	3.9	26	6
TC13	Freedom Lignite High Sodium, High Temp.	425	none	46	30.0	32.3	39	14
TC13	Freedom Lignite High Sodium, Low Temp.	457	none	67	13.9	2.3	26	5
TC14	Powder River Basin	220	none	84	18.7	14.6	27	14
TC15	Powder River Basin	156	none	79	10.7	1.3	20	4
TC16	Powder River Basin	288	230	75	11.4	2.7	17	3
TC16	Freedom Lignite, Low Temp.	173	135	64	11.0	1.8	32	4
TC17	Powder River Basin	162	155	81	9.4	2.2	20	3
TC17	Illinois Basin Bituminous	289	none	71	15.2	2.3	14	3

Table 3.1-12 Carbon Conversion and Gasification Efficiencies

Operating Period ¹	Average Relative Hours	Carbon Conversion %	Efficiency		
			Raw		Projected ²
			Cold %	Hot %	Cold %
TC17-1	29	96.7	51.3	85.6	69.8
TC17-2	38	95.9	52.0	85.4	70.2
TC17-3	41	95.6	52.6	85.2	70.5
TC17-4	57	95.2	53.7	85.4	71.3
TC17-5	65	94.7	51.2	84.2	69.8
TC17-6	80	95.0	52.3	84.8	69.8
TC17-7	85	95.3	53.1	84.7	70.0
TC17-8	111	84.9	35.5	73.9	52.6
TC17-9	120	84.6	35.8	73.9	52.2
TC17-10	128	84.3	36.2	74.2	51.0
TC17-11	161	88.5	38.2	78.1	53.0
TC17-12	173	81.9	37.2	71.8	51.3
TC17-13	180	82.2	37.6	72.5	51.9
TC17-14	188	82.5	37.4	73.0	51.7
TC17-15	201	83.1	39.5	74.3	53.3
TC17-16	211	82.9	39.7	74.3	53.5
TC17-17	227	83.6	40.6	74.9	54.8
TC17-18	240	83.8	40.4	75.0	55.3
TC17-19	251	83.4	39.9	74.5	54.7
TC17-20	259	82.9	39.0	73.9	53.7
TC17-21	267	80.4	34.6	70.3	49.0
TC17-22	279	81.0	40.6	72.7	54.9
TC17-23	290	81.0	40.6	72.6	54.7
TC17-24	301	82.3	41.6	73.7	55.9
TC17-25	315	82.6	39.9	73.3	54.6

Notes:

1. TC17-1 to TC17-7 were Powder River Basin coal; TC17-8 to TC17-25 were Illinois Basin coal.
2. Projection assumes that all the nitrogen in the syngas is from air and that the gasifier is adiabatic.

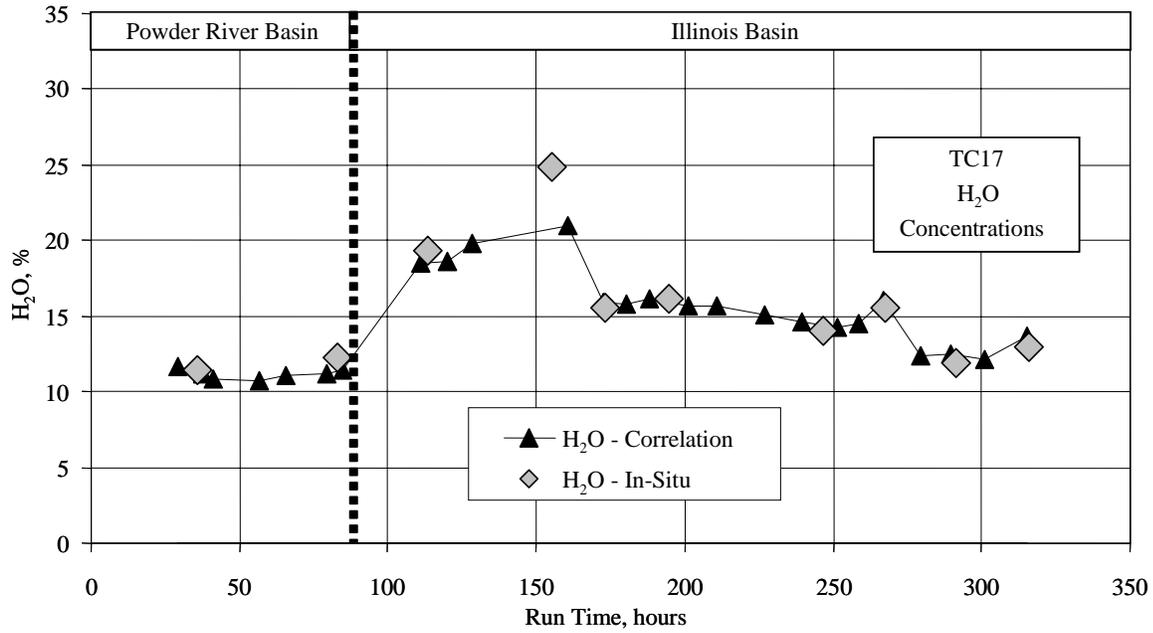


Figure 3.1-1 Syngas H₂O Concentration

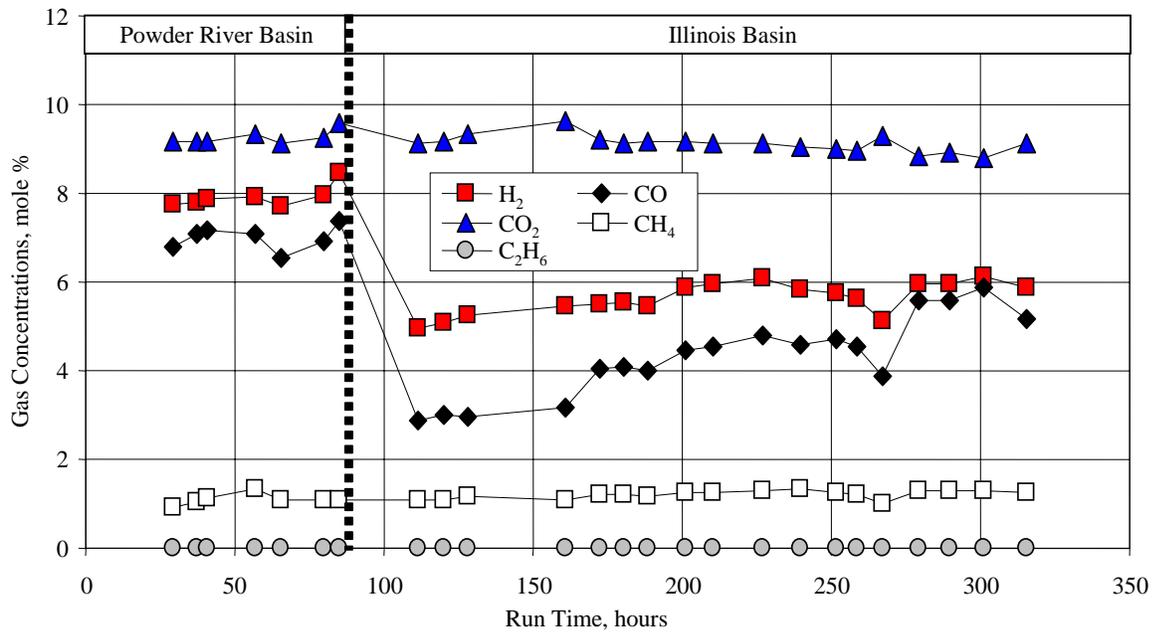


Figure 3.1-2 Wet Syngas Compositions

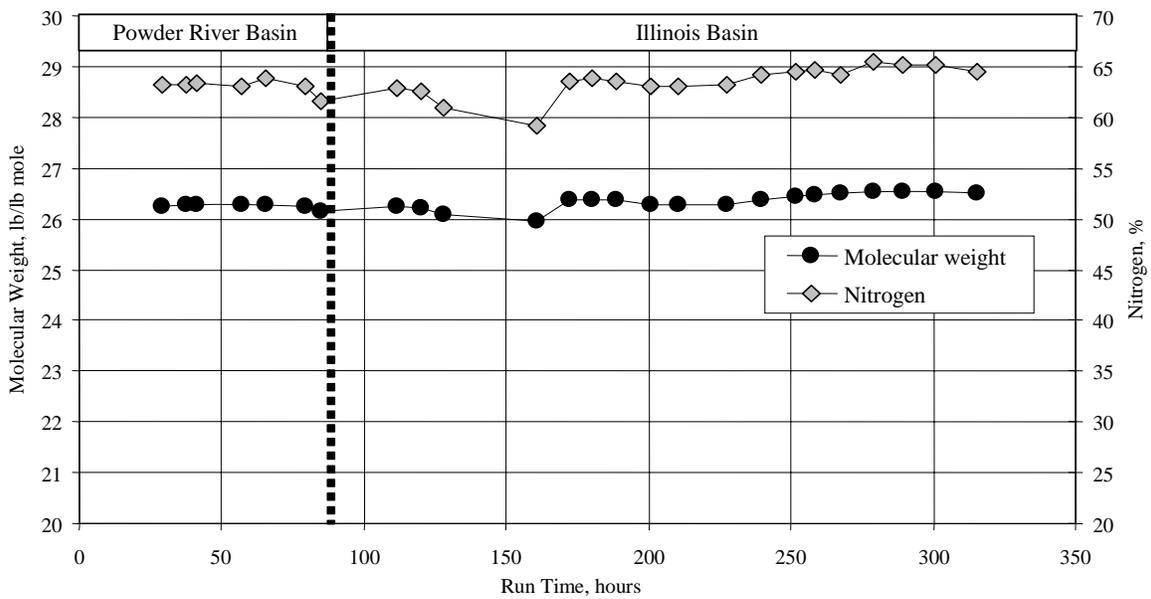


Figure 3.1-3 Wet Syngas Molecular Weight & Nitrogen Concentration

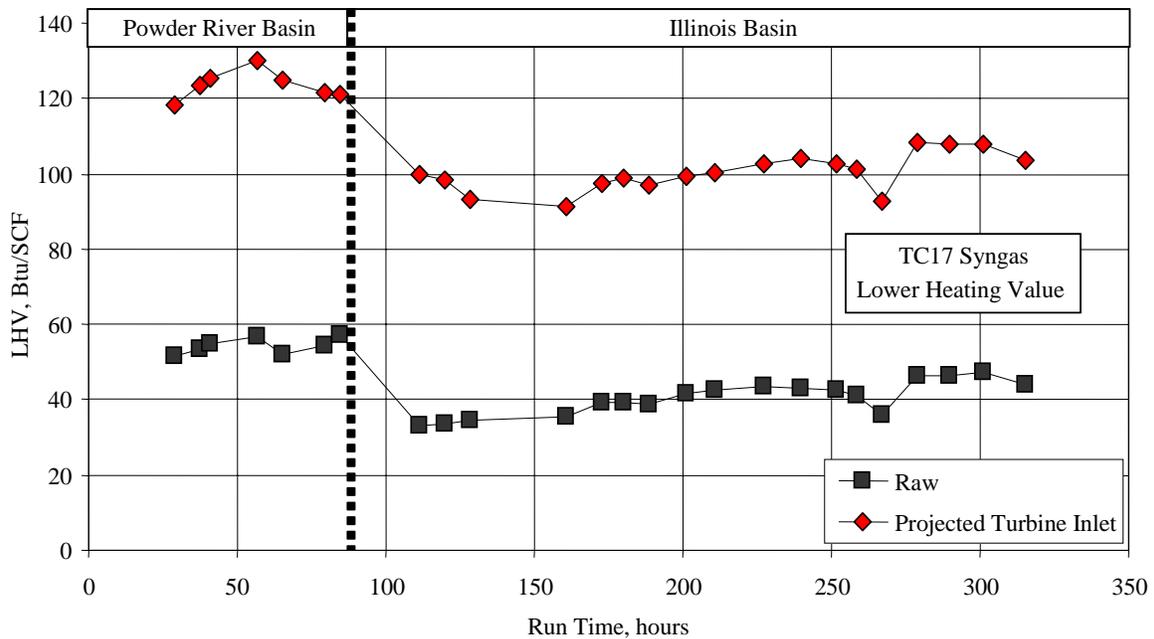


Figure 3.1-4 Syngas Lower Heating Values

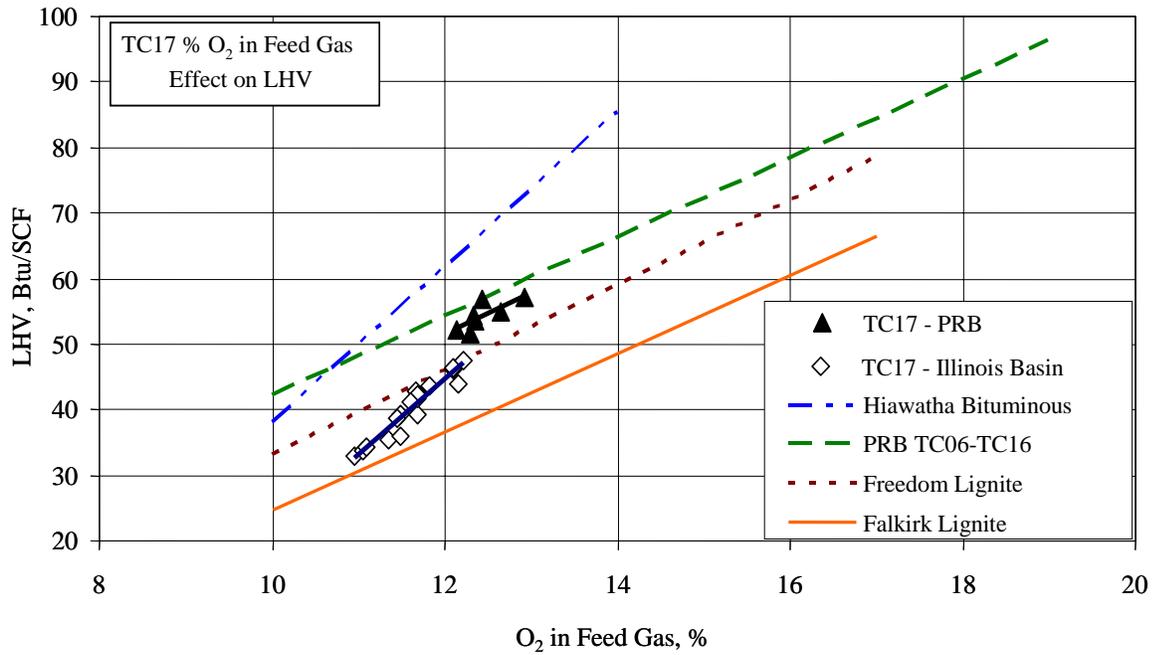


Figure 3.1-5 Raw Lower Heating Value & Overall Percent O₂ in Feed Gas

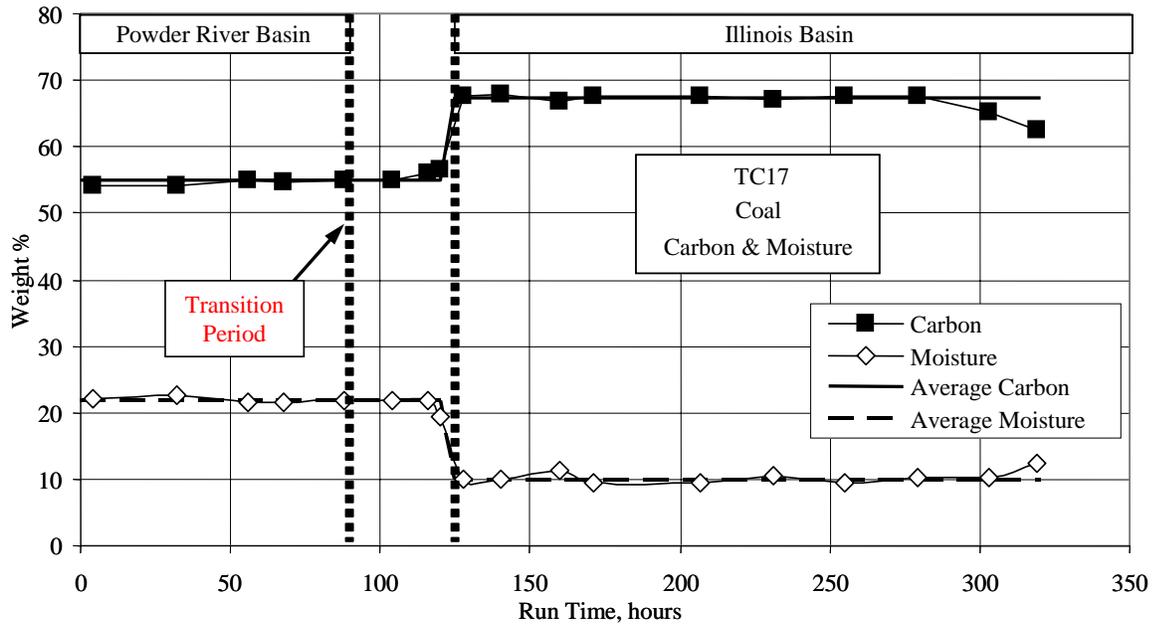


Figure 3.1-6 Coal Carbon & Moisture Content

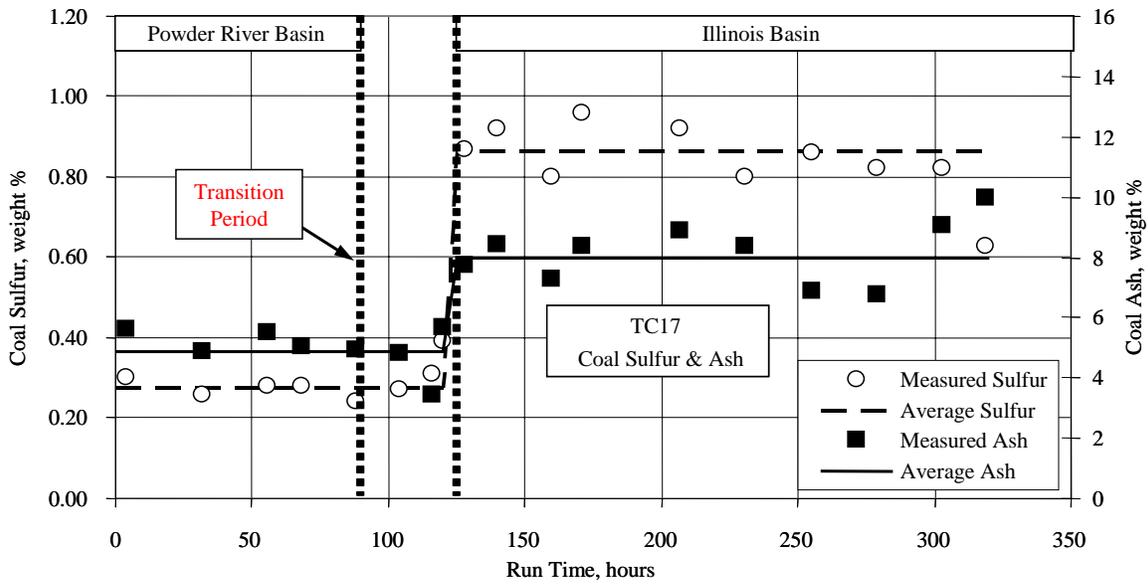


Figure 3.1-7 Coal Sulfur & Ash Content

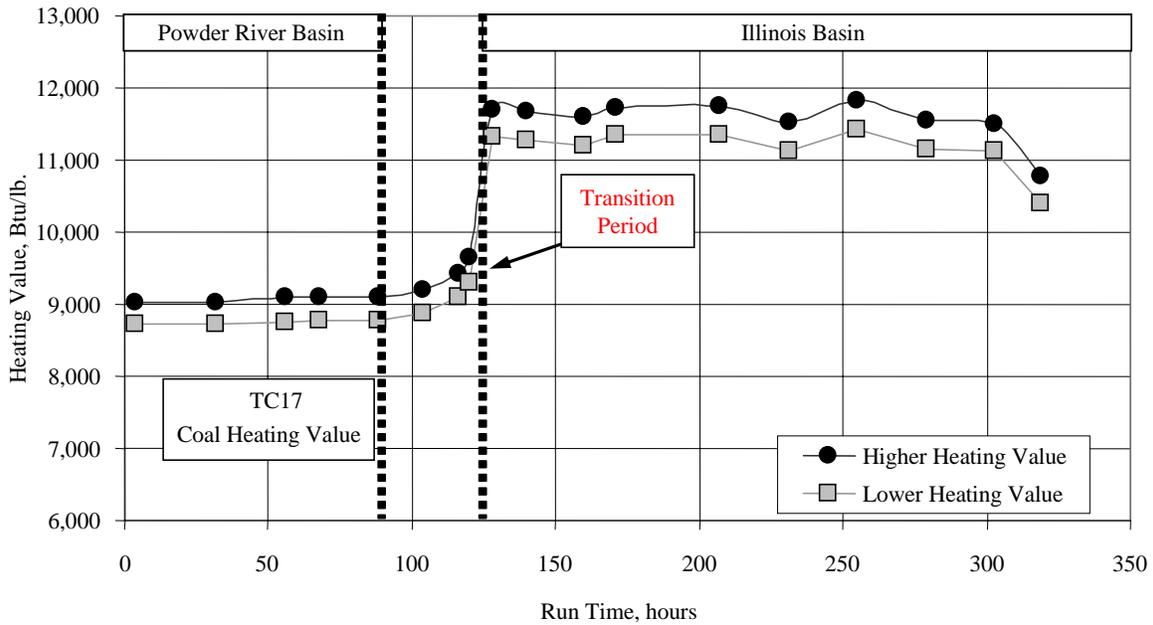


Figure 3.1-8 Coal Heating Value

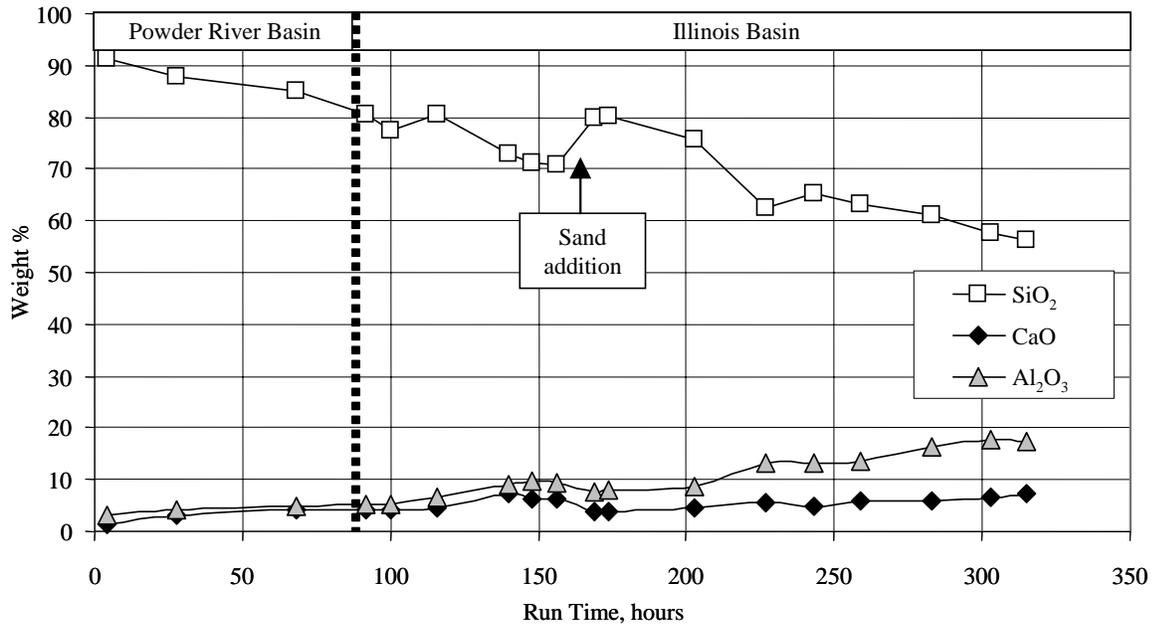


Figure 3.1-9 Standpipe Solids SiO₂, CaO, & Al₂O₃

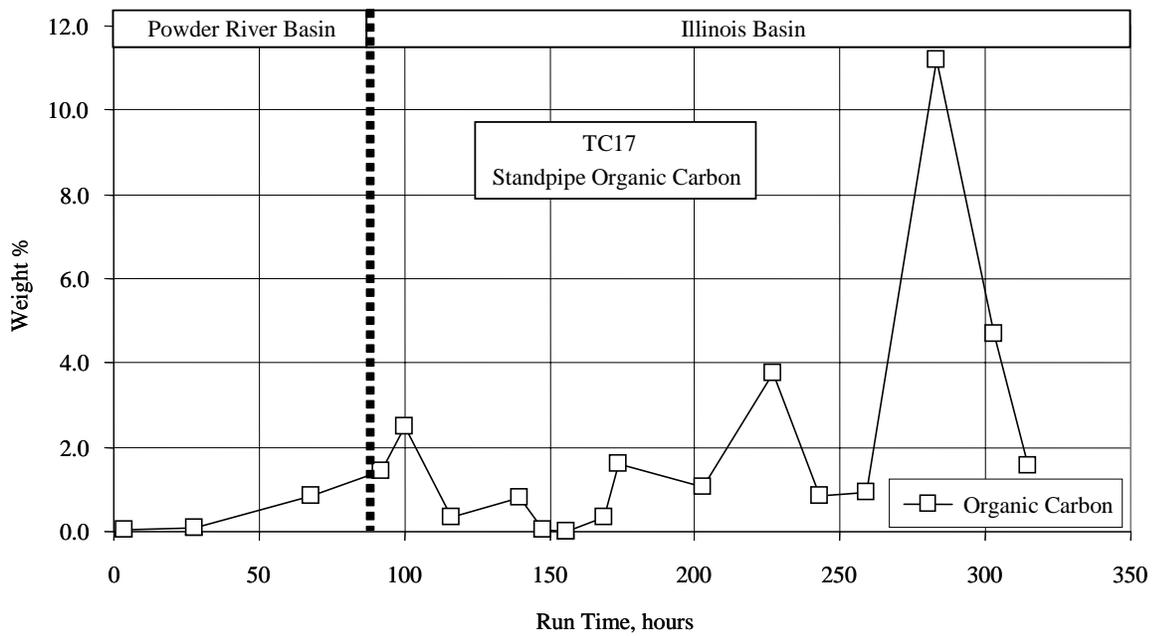


Figure 3.1-10 Standpipe Solids Organic Carbon

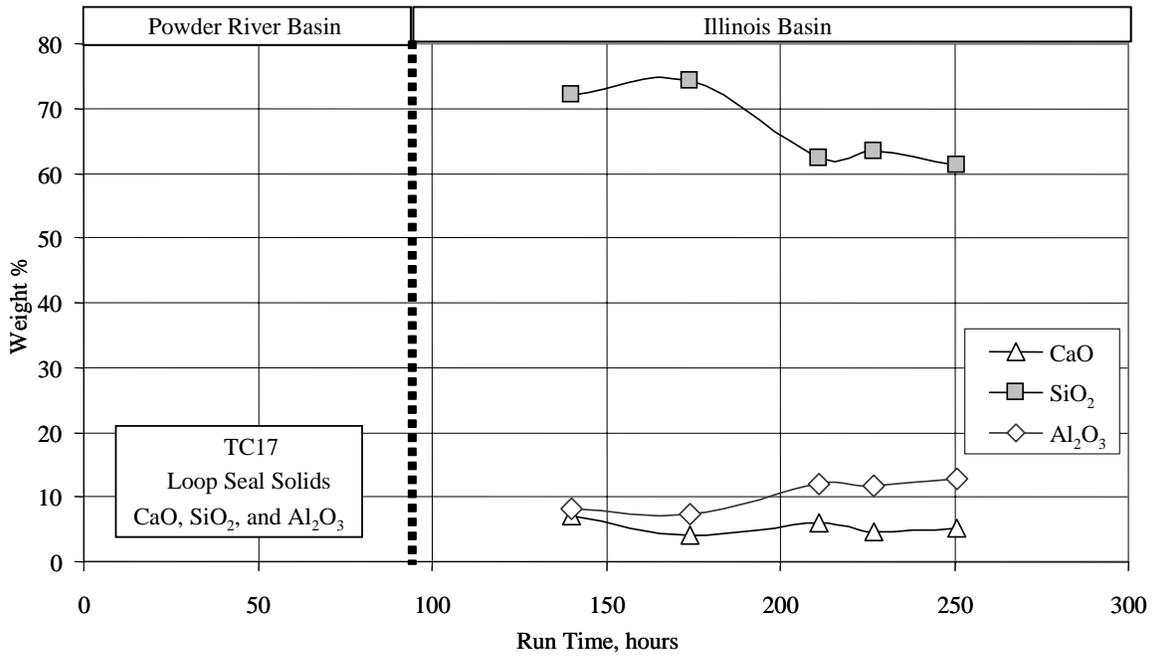


Figure 3.1-11 Loop Seal Solids SiO₂, CaO, & Al₂O₃

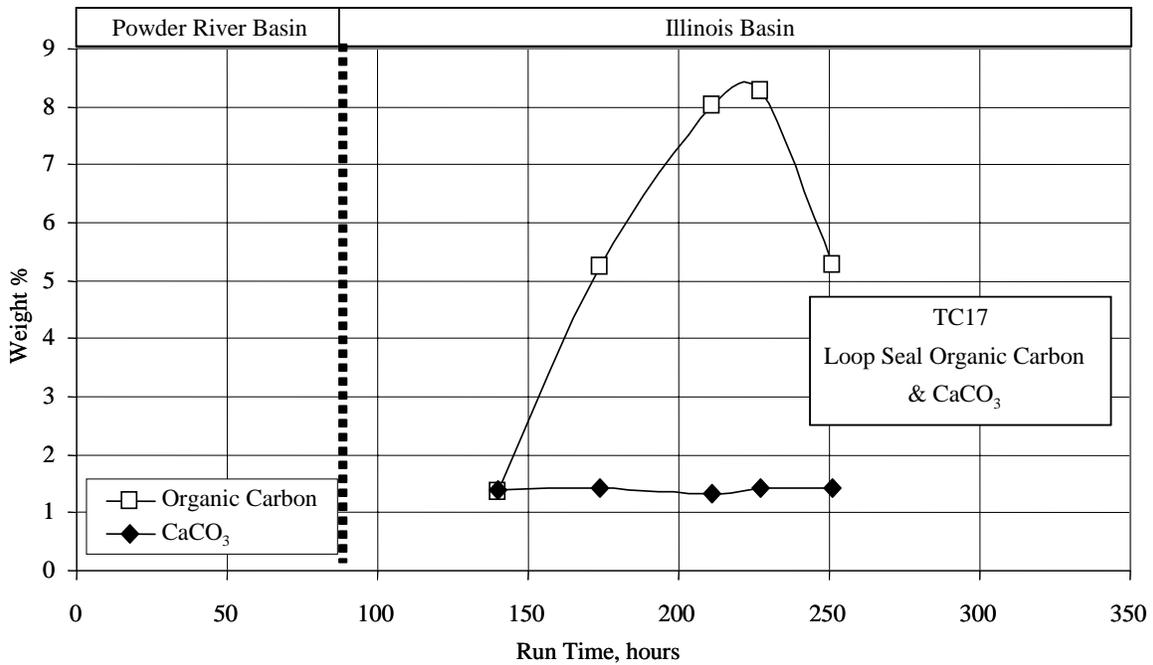


Figure 3.1-12 Loop Seal Solids Organic Carbon and Calcium Carbonate

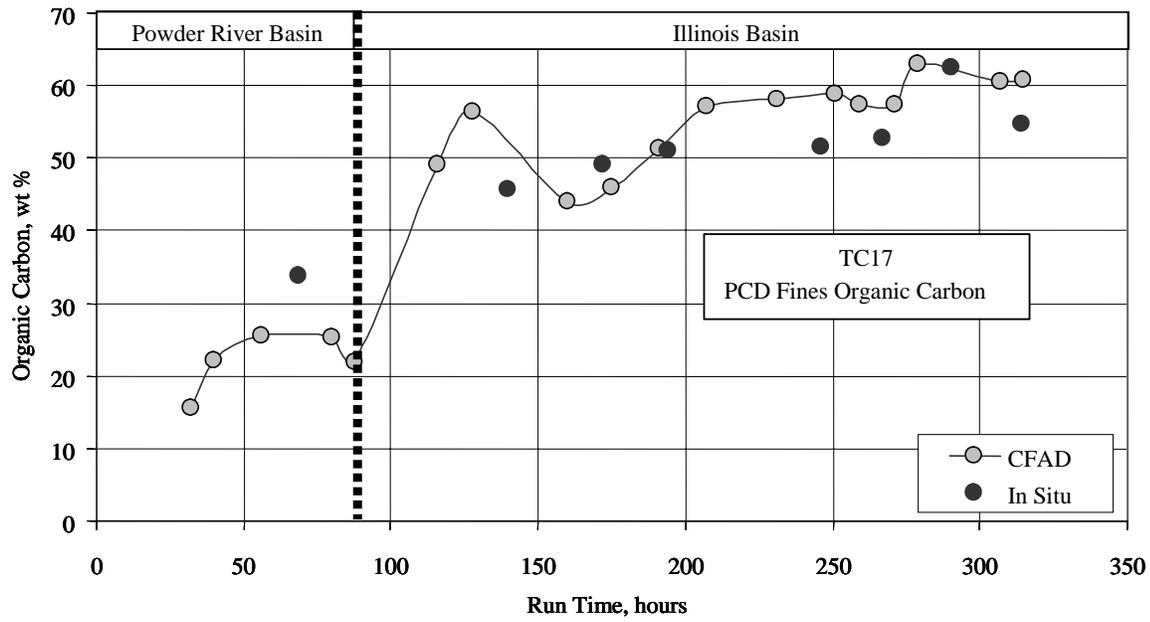


Figure 3.1-13 PCD Solids Organic Carbon

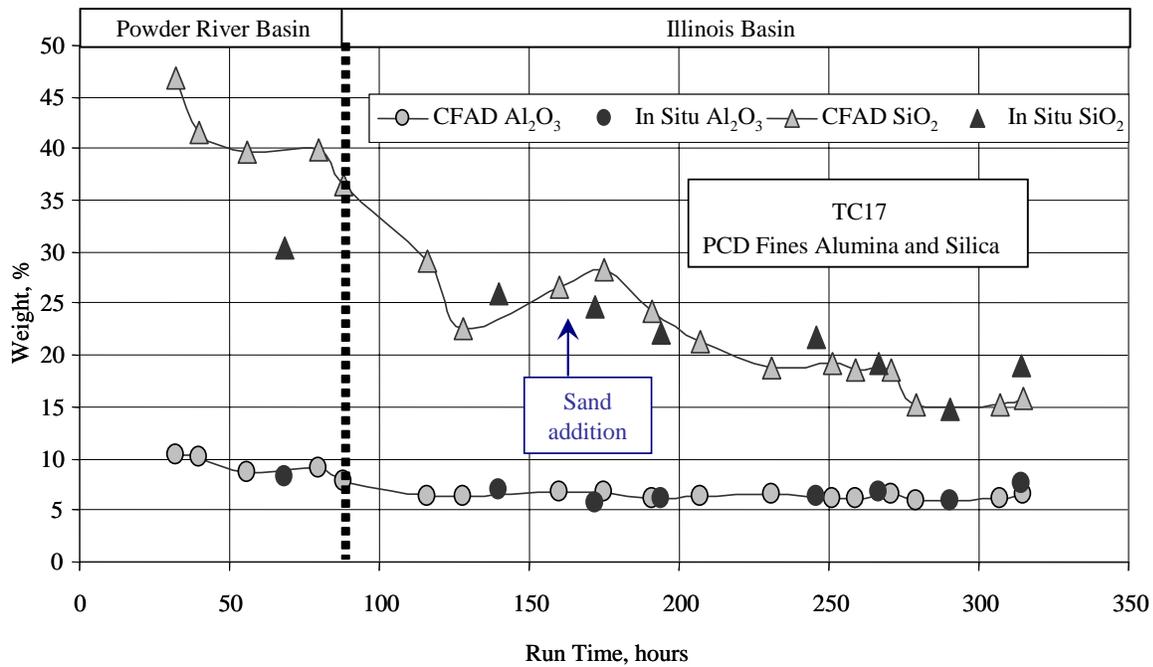


Figure 3.1-14 PCD Solids Silica & Alumina

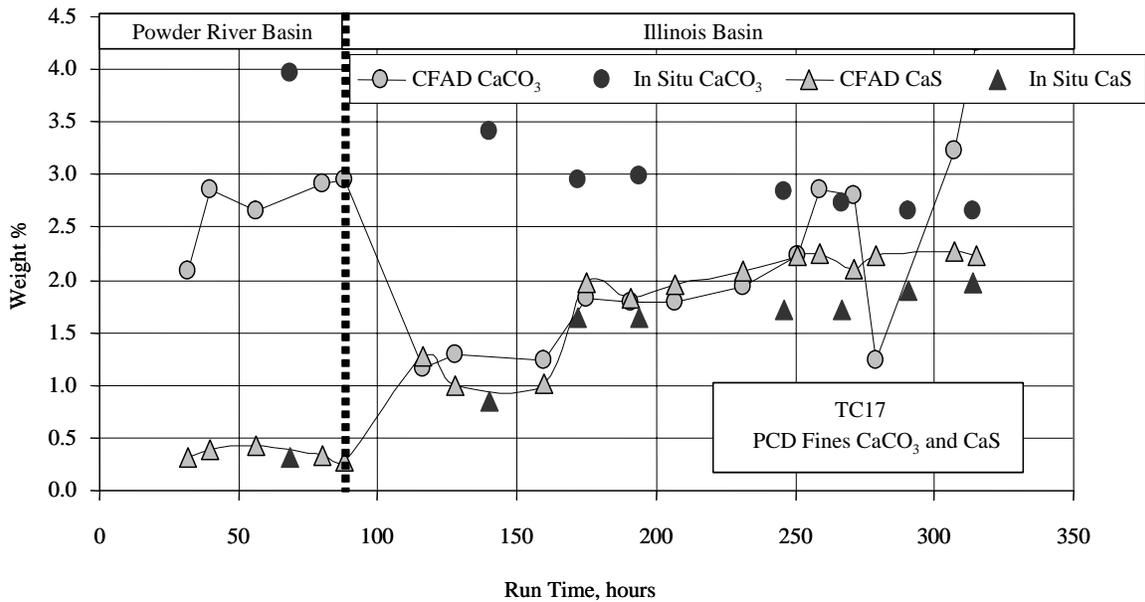


Figure 3.1-15 PCD Solids Calcium Carbonate & Calcium Sulfide

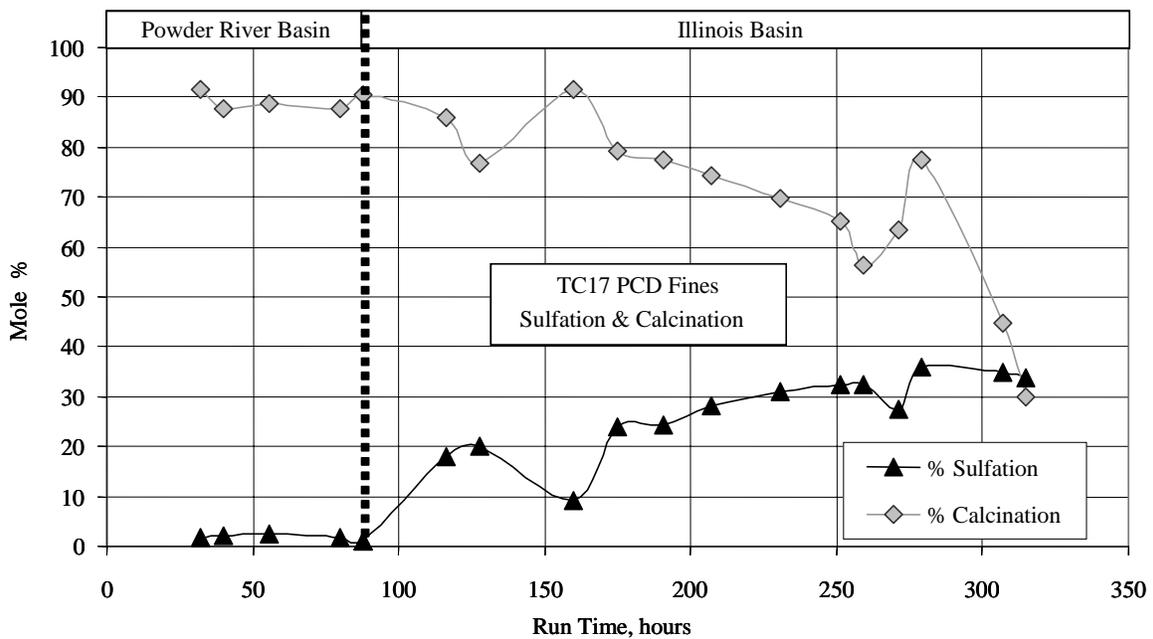


Figure 3.1-16 PCD Solids Calcination & Sulfation

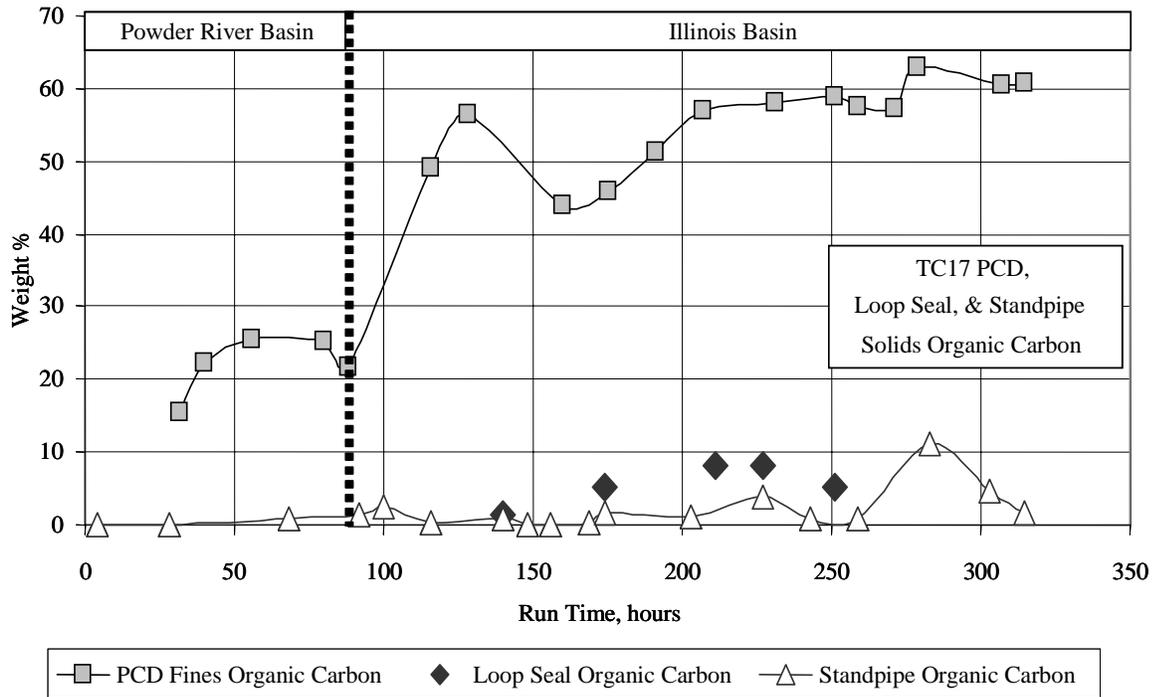


Figure 3.1-17 Standpipe, Loop Seal, & PCD Solids Organic Carbon Content

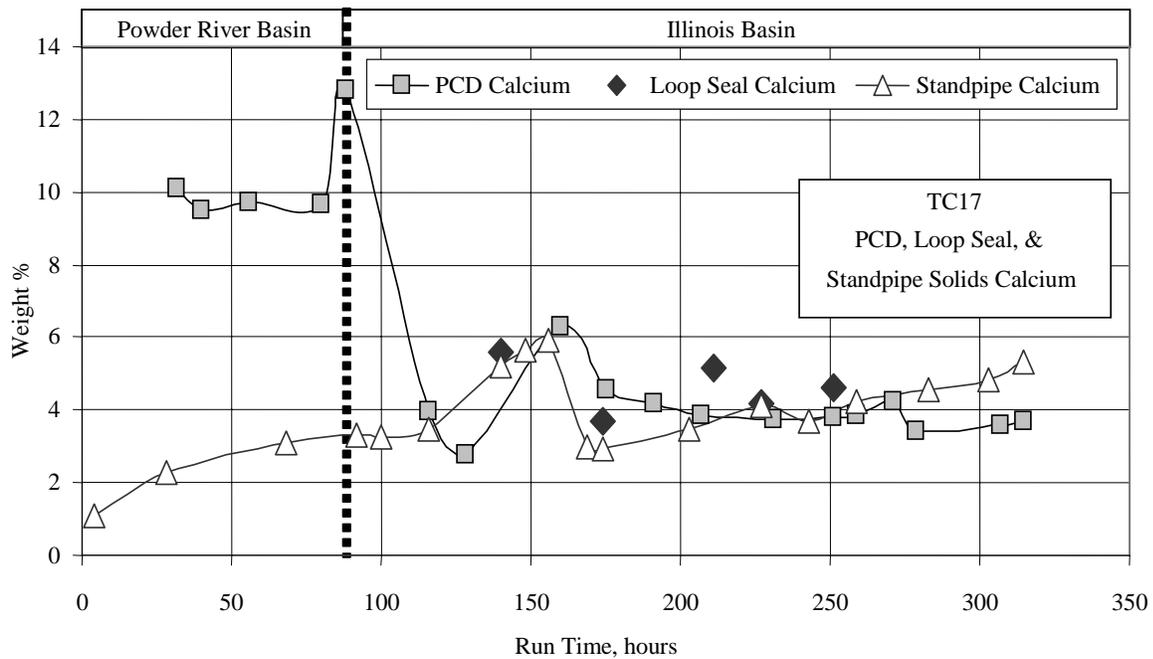


Figure 3.1-18 Standpipe, Loop Seal, & PCD Solids Calcium

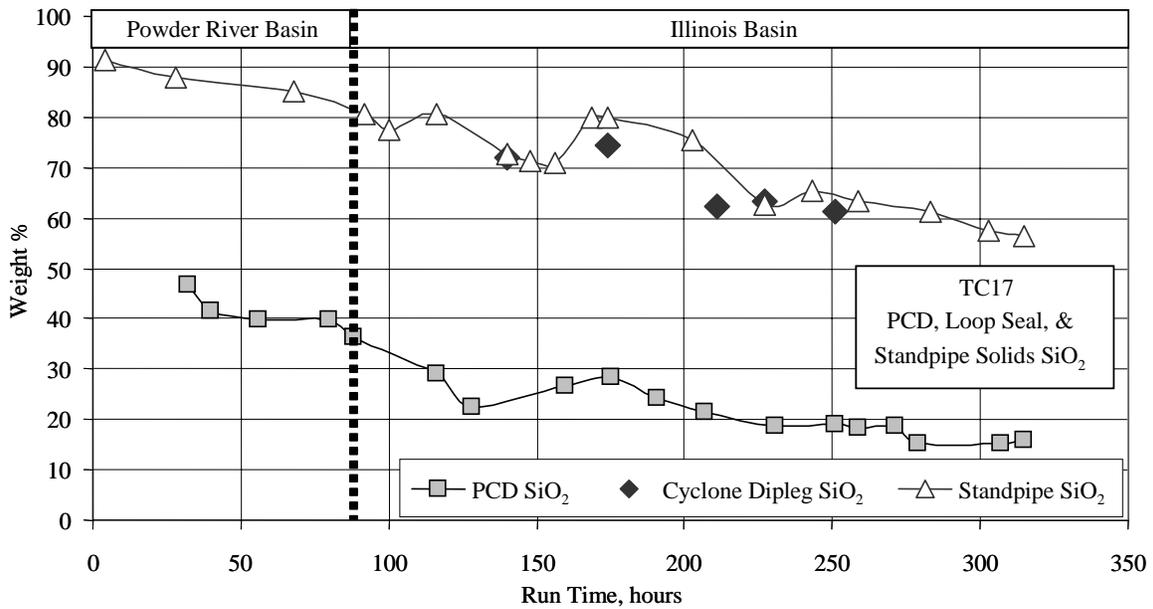


Figure 3.1-19 Standpipe, Loop Seal, & PCD Solids Silica

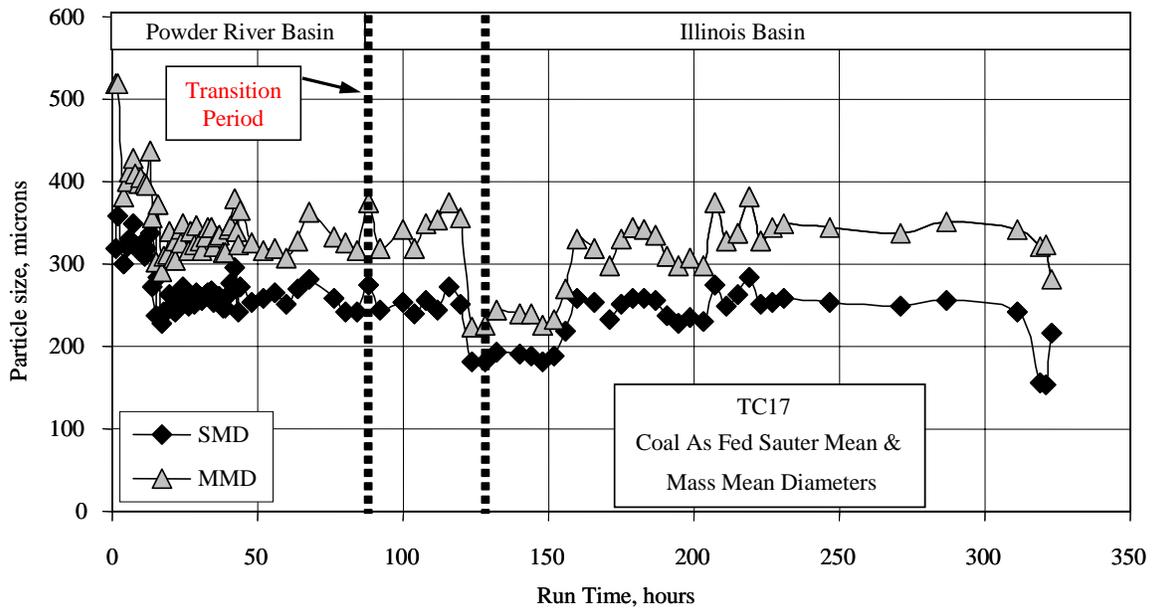


Figure 3.1-20 Coal Particle Size

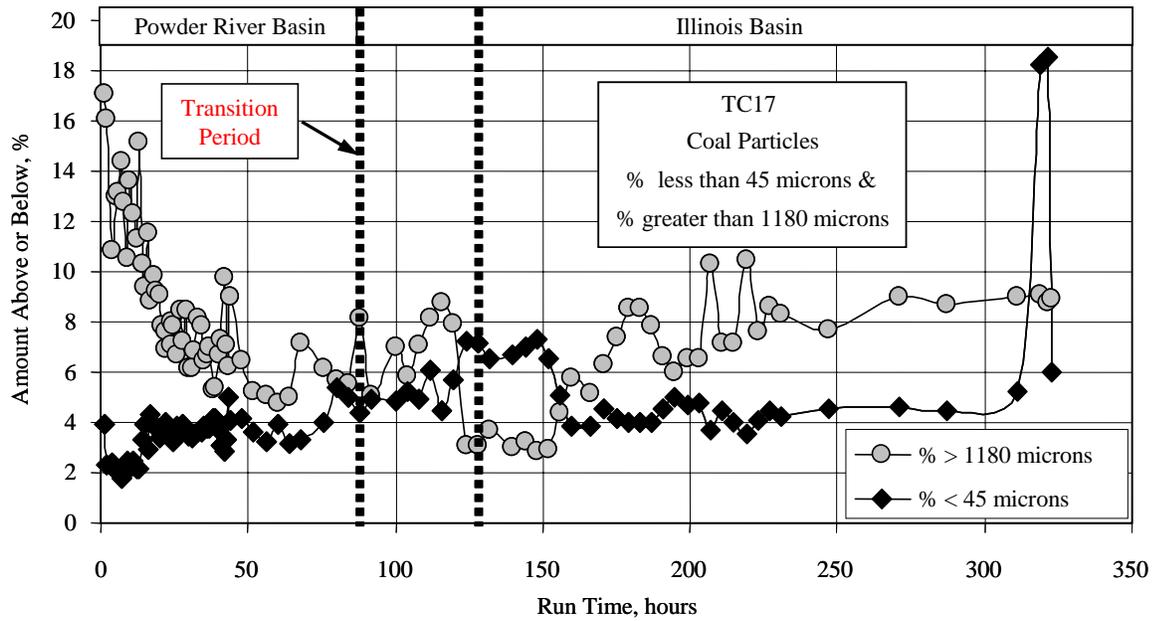


Figure 3.1-21 Percent Coal Fines & Oversize

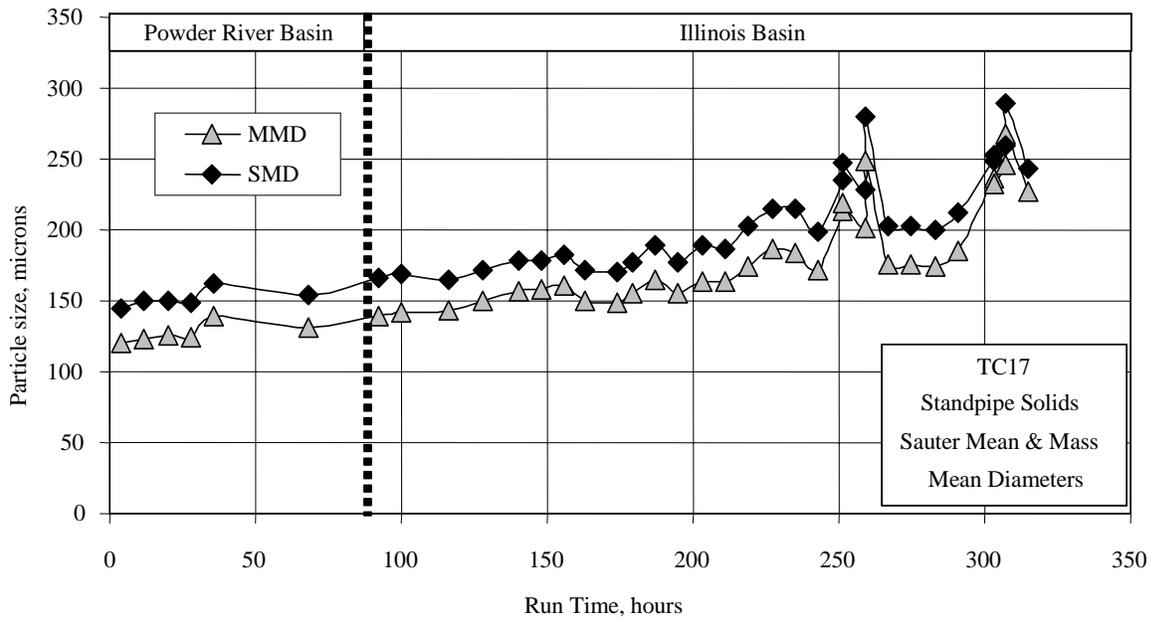


Figure 3.1-22 Standpipe Solids Particle Size

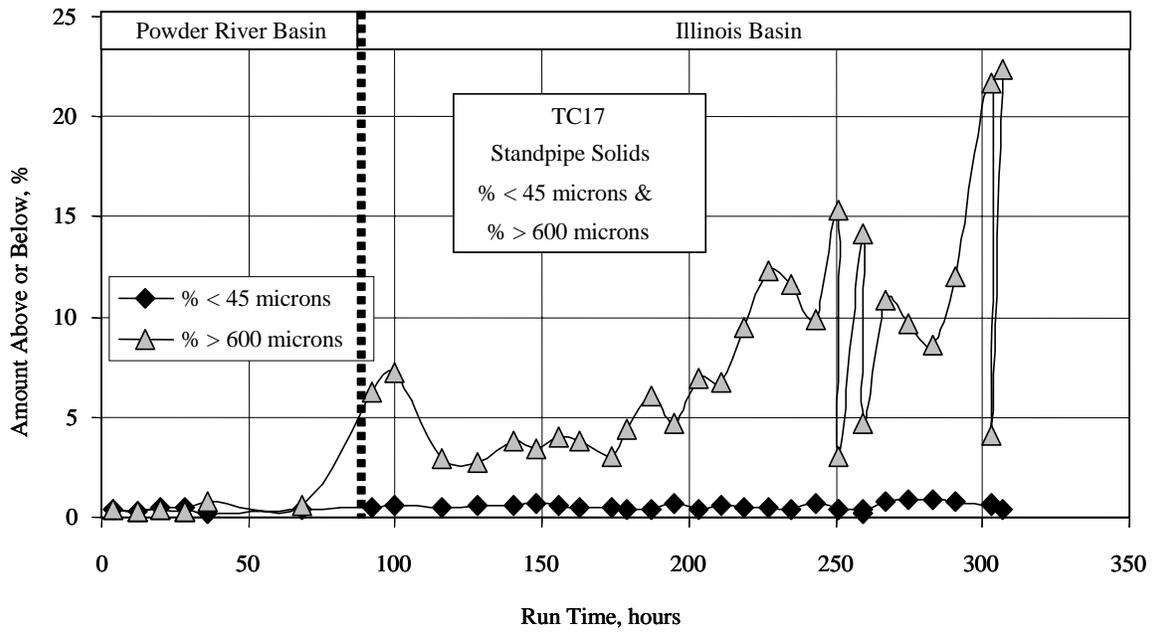


Figure 3.1-23 Standpipe Solids Fine and Coarse Particles

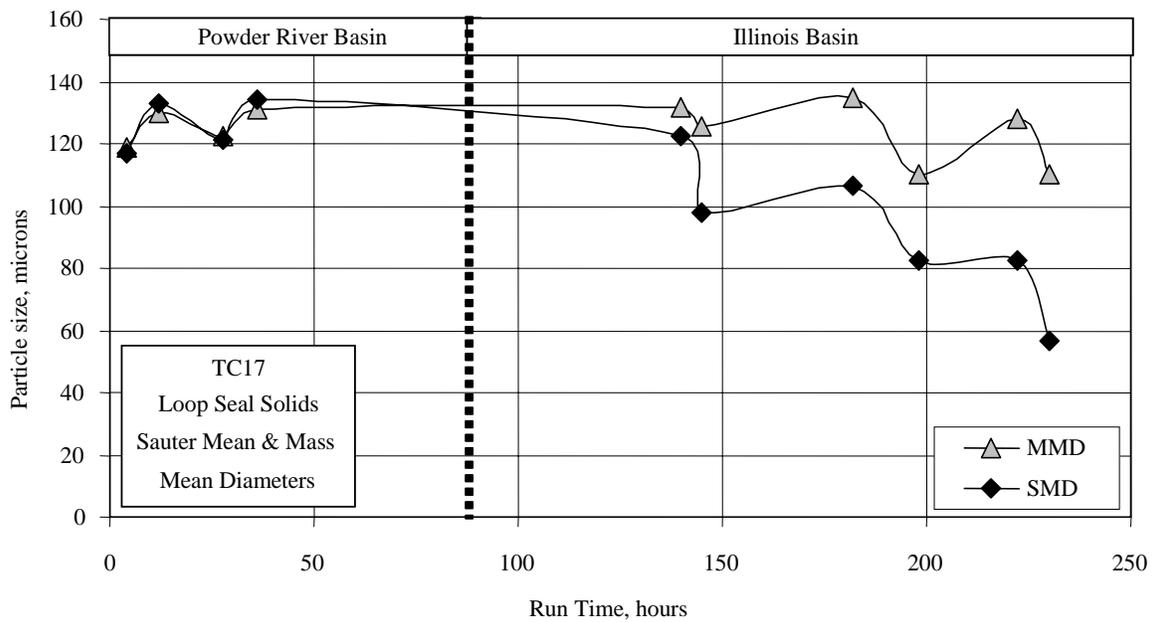


Figure 3.1-24 Loop Seal Solids Particle Sizes

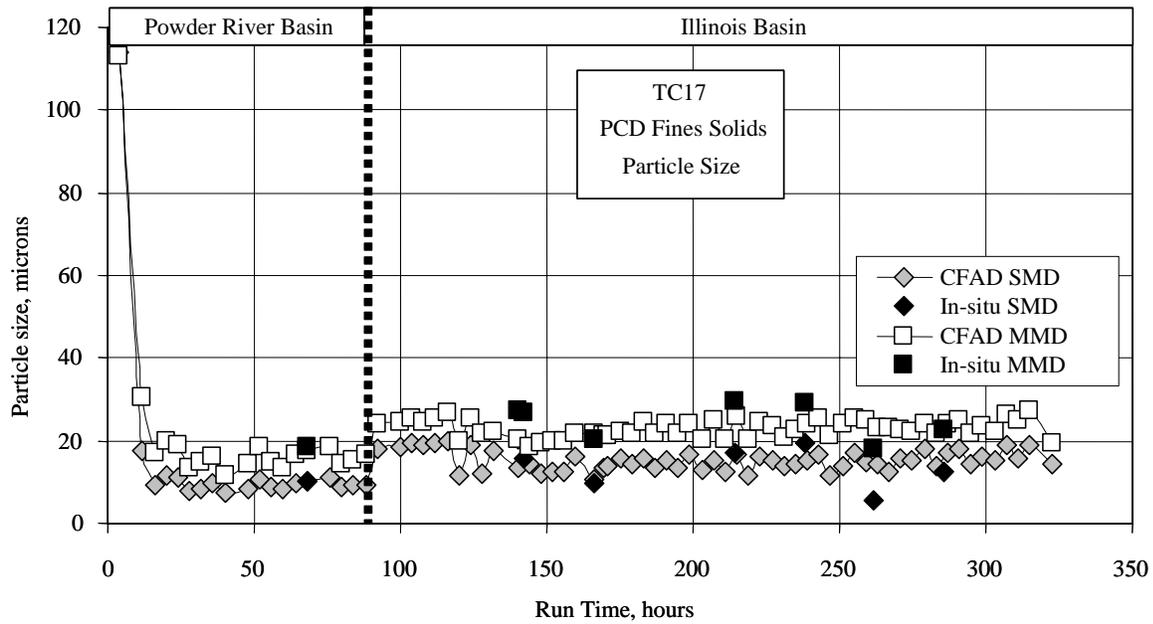


Figure 3.1-25 PCD Solids Particle Sizes

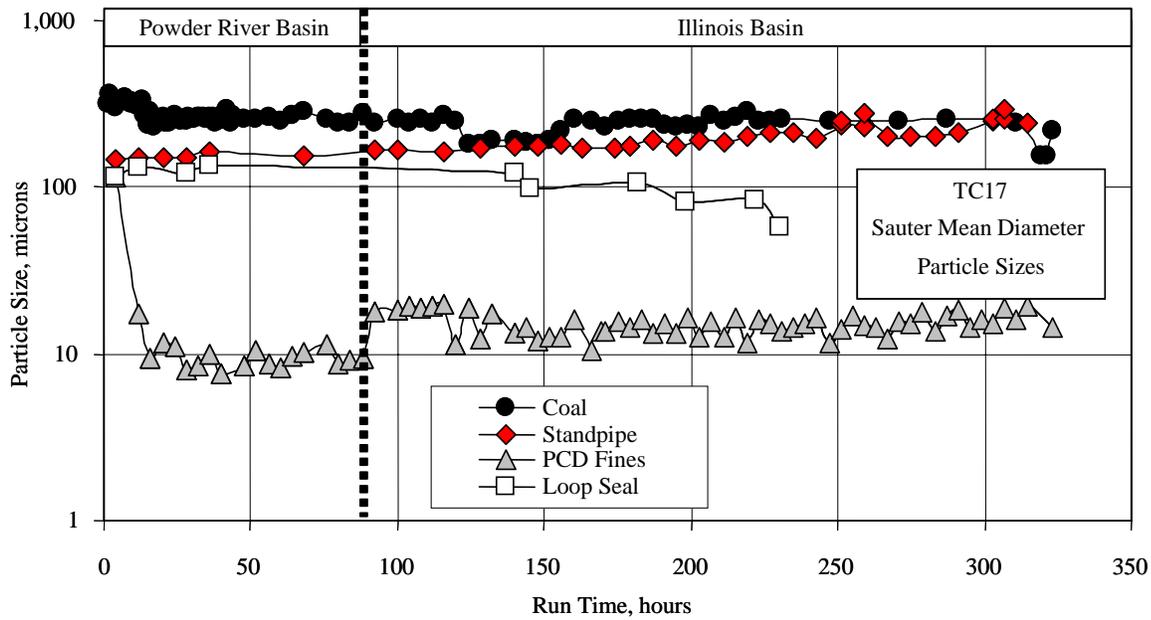


Figure 3.1-26 Coal and Standpipe, Loop Seal, & PCD Solids Particle Size Comparison

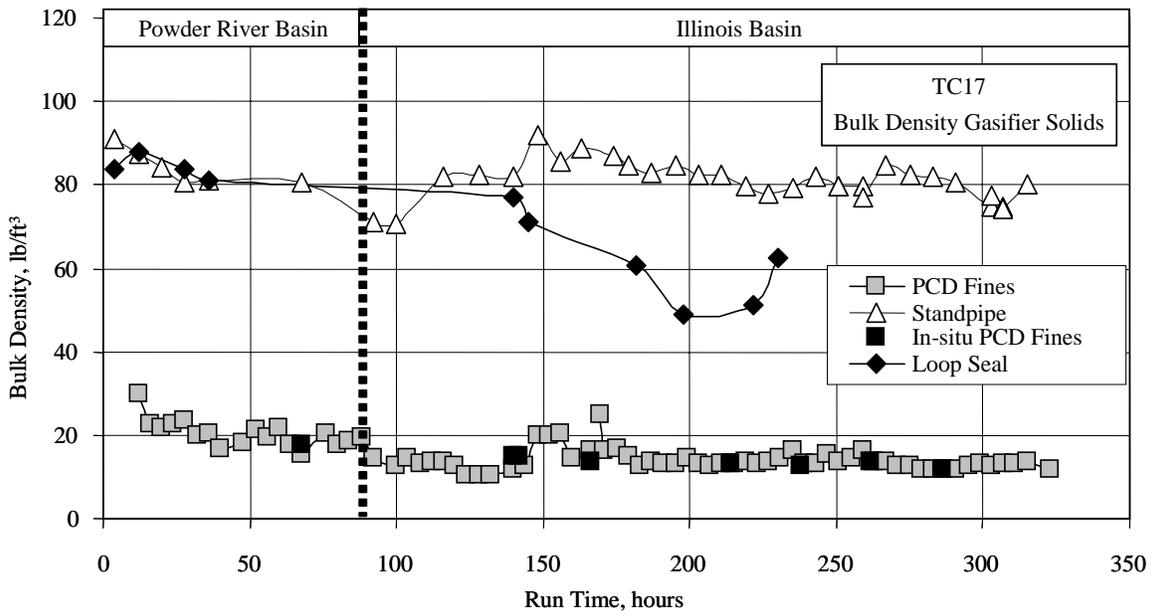


Figure 3.1-27 Standpipe, Loop Seal, and PCD Solids Bulk Density

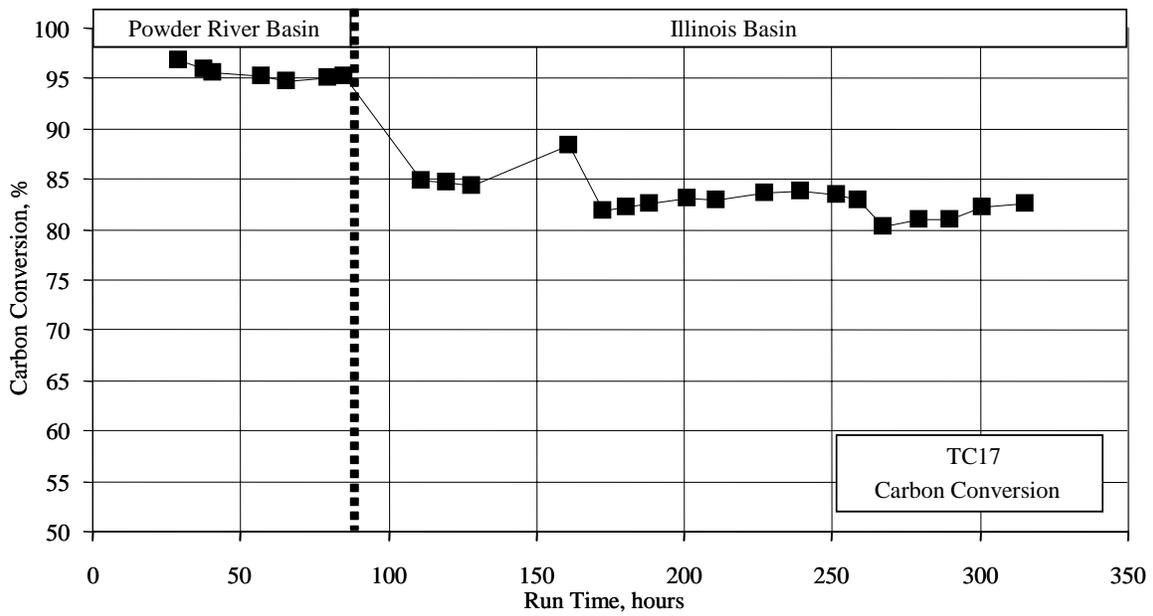


Figure 3.1-28 Carbon Conversion

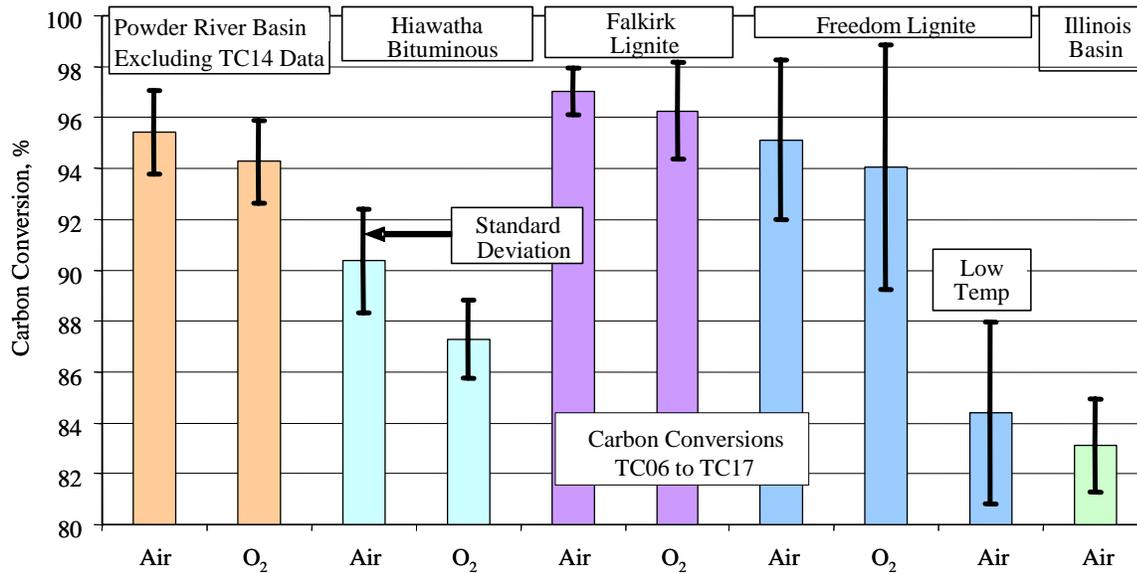


Figure 3.1-29 Carbon Conversion of Five Coals

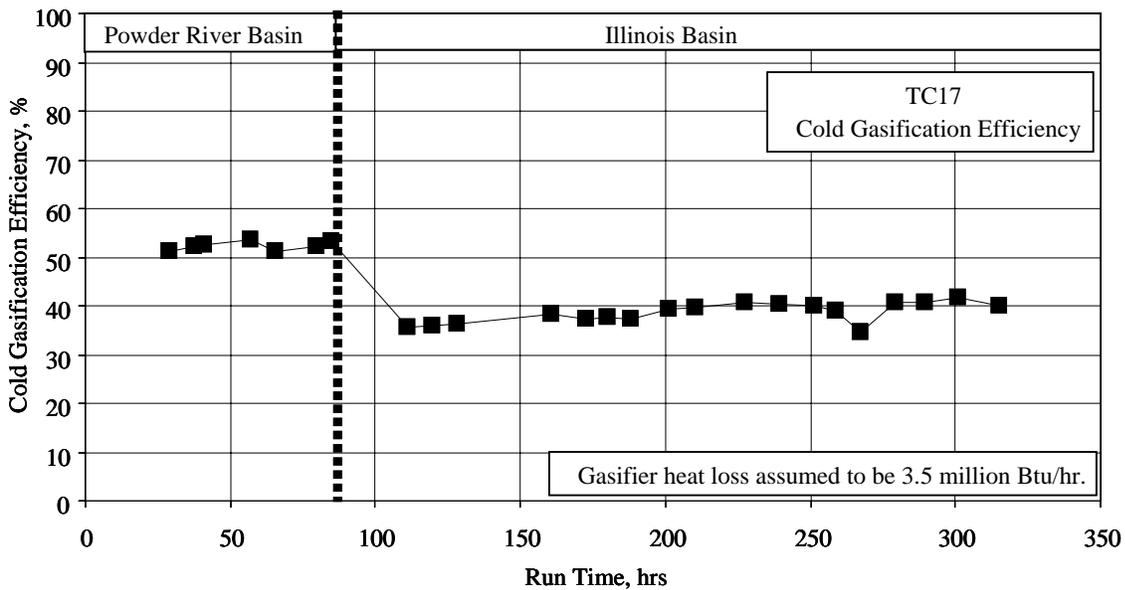


Figure 3.1-30 Cold Gasification Efficiency

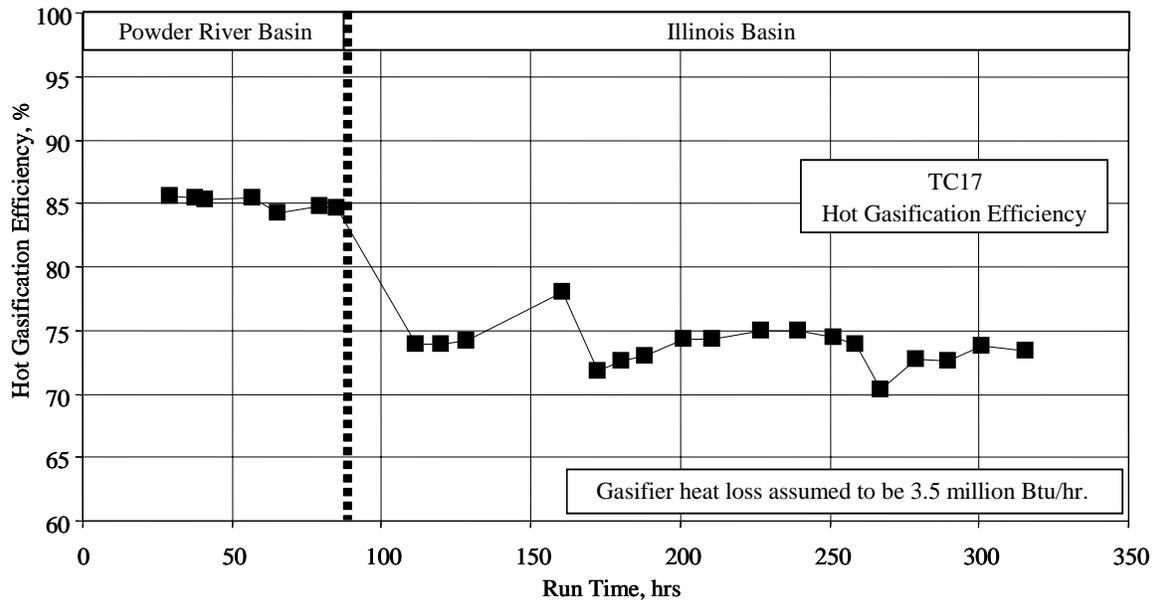


Figure 3.1-31 Hot Gasification Efficiency

3.2 PARTICULATE CONTROL DEVICE PERFORMANCE

3.2.1 Overview

The performance of the particulate control device (PCD) during the initial portion of test run was relatively stable until the transition to bituminous coal feed. Apparently, the bituminous gasification ash did not flow as easily as the PRB gasification ash and it was more difficult to remove from the PCD hopper. The gasification ash accumulated in the hopper, eventually reaching the filter elements and causing gasification ash bridging. As a result, the pressure drop across the tube sheet showed a rapidly increasing trend. The bridged gasification ash was successfully removed on-line by controlled combustion by adding diluted air to the PCD. Although the controlled combustion was restored a stable pressure drop, the system was later shutdown due to a water leak in heat exchanger in the atmospheric fluidized bed combustor. The inspection revealed eleven broken filter elements that evidently failed when the hopper was overfilled. Despite the failure of these elements, outlet loading was no more than 1 ppmw as measured by in situ sampling.

PCD performance during the second portion of the test run was relatively stable with a baseline pressure drop of 60 to 95 inH₂O. During steady-state operations, the inlet temperature was between 725 and 950°F and the face velocity was 3 to 4 ft/min. Filter element and failsafe testing was continued, and is discussed in Section 3.2.6. Bridging did not occur and operations with bituminous coal were more stable than in the first portion of the test run. After coal feed was discontinued at the conclusion of TC17, the dust cake was combusted.

3.2.2 Particle Mass Concentrations

In situ particulate sampling was performed at the PCD inlet and outlet using the in situ batch sampling systems. The in situ sampling, performed by Southern Research Institute, is described in the article “In Situ Particulate Sampling and Characterization at the Power Systems Development Facility” (Dahlin, et. al, 1998), which can be found in the technical papers section of the PSDF website, <http://psdf.southernco.com/>.

PCD Inlet Mass Loadings. Particle mass concentrations and mass rates measured at the PCD inlet are given in Table 3.2-1, and the mass rates are plotted as a function of coal feed rate in Figure 3.2-1. As shown in the plot, the rate of solids carryover to the PCD was generally higher with the Illinois Basin coal than it was with the PRB coal in previous tests. One factor that may have contributed to the higher solids carryover was the higher ash content of the bituminous coal (nominally 9 percent versus 6 percent for the PRB coal). Another factor was the higher residual carbon content of the solids, as discussed in Section 3.2.4.3. The higher carbon content suggested that carbon conversion with the bituminous coal was somewhat lower than the carbon conversions typically seen with PRB coal.

PCD Outlet Mass Loadings. Particle concentrations measured at the PCD outlet are included in Table 3.2-1 and compared to other test programs in Figure 3.2-2. Continuing the trend seen in previous test programs, a slightly elevated particulate loading was measured on the first day of testing. Unfortunately, it was not possible to track the trend in particle loadings after the first day due to contamination of the sampling filters. However, on the third day of testing, a single

measurement was obtained in the absence of any contamination, and it suggested the outlet particle loading was below the limit of resolution (< 0.1 ppmw).

After measuring < 0.1 ppmw at the PCD outlet while still on PRB coal, the transition was made to Illinois Basin coal. As the transition was being made, the rate of ash discharge from the PCD hopper was not increased to compensate for the higher ash content of the bituminous coal. As a result, the PCD hopper was accidentally overfilled. Afterwards, the on-line particulate monitor, the Dust Alert-90 from PCME, Inc. (referred to as the PCME) started to give elevated readings which suggested the presence of an increased particle loading at the PCD outlet. Two outlet sampling runs were performed to quantify the outlet loading. Contamination of the sampling filters by condensed organic compounds prevented accurate quantification of the particle loading, but subsequent microscopic examination of the filters indicated the loading was no more than 1 ppmw. The system was later shut down due to a water leak, and an inspection revealed the failure of 11 filter elements.

After repairing the leak and replacing the broken elements, the system was restarted. An initial particulate measurement was made while still on the startup burner. Due to the presence of large-particle contamination, the measured particle loading was 0.85 ppmw, but microscopic examination of the sampling filter indicated that the true loading of ash particles was less than 0.1 ppmw. The large particle contamination was made up of rust flakes and large black particles around 100 microns in diameter and appeared to be trash from the duct. The source of these particles is unknown, but the lack of fine particles on the filter indicates that this mass was not associated with a PCD leak.

Sampling runs performed after bituminous coal was fed were all contaminated with condensed organic compounds. Based on microscopic examination of the sampling filters, the particle loading appeared to gradually increase from an estimated initial loading of 0.2 to 0.3 ppmw to an estimated final loading of 0.4 to 0.6 ppmw. During this time period, the PCME particulate monitor was apparently fouled with condensed material and failed to indicate any increase in the outlet particle loading.

3.2.3 Real-Time Monitoring

The PCME particulate monitor was operational throughout TC17 but was affected by organic compound condensation contamination during most of the testing with Illinois Basin coal. The PCME indicated elevated readings shortly after the hopper was overfilled which suggested possible particulate leakage through the PCD. In order to accurately quantify the outlet loading, a particulate sampling run was performed and the PCME showed a reading of approximately 30 percent during that time. Based on prior calibration data, a 30 percent reading would correspond to a particulate loading in the range of 6 to 40 ppmw. The actual particulate loading was not quantified precisely due to contamination by condensed organic compounds, but the loading was estimated to be no more 1 ppmw based on microscopic examination of the filter. For a particulate loading of 1 ppmw, the PCME reading should have been only 1 to 5 percent. Therefore, the PCME was reading higher than expected based on the previous calibration data. The reason for the high reading is unknown, but it was possibly caused by droplets of condensed organic compounds since the sampling filter from this time period was contaminated. In addition, the PCME probe was also found to be fouled with condensed deposits when it was removed after the test run.

3.2.4 PCD Solids Analysis

Pressure drop, cleaning requirements, and bridging tendency can be influenced by changes in the characteristics of the solids collected in the PCD. Important characteristics of the solids include particle size distribution, bulk density, true density, porosity, surface area, composition, and flow resistance. The effects of these parameters must be considered in analyzing the performance of the PCD.

3.2.4.1 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.

In Situ Samples. Figure 3.2-3 shows differential mass particle size distributions measured on the PCD inlet in situ samples with both the Illinois Basin coal and PRB coal used in TC17. In the figure, $dM/d\log D$ is the slope (or differential) of the plot of cumulative mass concentration (in ppmw) versus the log of particle size (diameter). It has the same units as cumulative mass concentration, M , which is expressed as ppmw. This is because the denominator, $d\log D$, is dimensionless, since $d\log D = \log D_1 - \log D_2 = \log (D_1/D_2)$. $dM/d\log D$ is commonly used in aerosol particle measurement rather than dM because this presentation will reproduce a log-normal size distribution correctly despite measurements which have unevenly spaced intervals on the size axis. Unevenly spaced size intervals are a common problem with particle size measuring devices and will cause distortion of the resulting curve. (That is, a log-normal or "bell-shaped curve in log space" will not be bell-shaped.) Since particle distributions in nature tend to be approximately log-normal, correctly rendering this property is an important aspect of this type of data presentation. Regardless, $dM/d\log D$ versus diameter is an accurate visual representation of the distribution of mass since the area under the curve in a size interval is proportional to the mass in the interval. Significant differences can be observed in the distributions from the two coals. The bituminous coal produced fewer particles in the size range of 1 to 10 microns but produced more mass in particles larger than 10 microns.

Hopper Sample. Figure 3.2-4 compares the differential mass percentage distributions for the in situ samples and hopper sample used for the TC17 lab drag measurements. The differential mass percentage distribution used in the figure is similar to the differential mass particle size distribution, $dM/d\log D$ except it is $d(\%Mass)/d\log D$, where $d(\%Mass)$ is the percentage of cumulative mass at size interval one minus the percent mass at size interval two. This characterization is used when comparing hopper samples and other samples where an accurate total mass is not known - that is, the absolute mass in a size range is unknown, but the fraction of the total sample is known. Good agreement is seen between the size distribution from the average of the bituminous in situ samples and the hopper sample. (The size distributions measured on the seven bituminous coal in-situ samples were averaged to compare to the single bituminous coal hopper sample used for lab measurements.). Both sets of bituminous samples show fewer fine particles than the PRB dust. Therefore, it appears that the hopper sample was appropriate for use in lab drag studies. These results are discussed later in this report.

3.2.4.2 Dustcake Observations

The inspection performed after the first portion of the test run (referred to in this section as TC17A, which included steady state operating periods TC17-1 through TC17-11) revealed an extremely thin residual cake of approximately 0.01 inches thick on the bottom plenum and slightly thicker (~0.02 inches) on the top plenum. On both plenums, the top layer of the cake appeared to be black gasification ash with a white or light gray layer underneath. The lighter-colored inner layer was probably combustion ash from the dustcake burnoff; while the material in the black outer layer was presumably carried over after the burnoff. The layers were too thin to sample separately, but separate bulk samples of the dustcake were obtained from both the top and bottom plenums. A sample of the deposit between the filter holders and a sample of the deposit on the failsafe tester were also obtained. When the cake was scraped off the elements, it was noted that many of the elements themselves had turned reddish-brown in color.

The post-run inspection (following TC17B, or operating periods TC17-12 through TC17-25) also revealed a very thin residual dustcake on the top plenum of approximately 0.01 inches thick. The bottom plenum cake was somewhat thicker (~0.02 in.) and darker in color than the top plenum cake. On the bottom plenum elements, there was a lighter-colored outer layer with a black layer underneath. This suggested that the second burnoff may not have completely consumed all of the carbon in the cake, leaving a carbon-rich inner layer. On the top plenum, the layers were reversed, with a darker outer layer and lighter inner layer. The light inner layer on the top plenum was most likely combustion ash from the first burnoff. This layer was not present on the bottom plenum elements, because they were all replaced after TC17A. Differences in the outer layers suggested the second burnoff was more effective on the bottom plenum than it was on the top plenum. Despite this fact, the top plenum cake from TC17B apparently had less overall carbon due to the effect of the remaining inner-layer of combustion ash from the first burnoff.

The presence of the darker outer layer on the top plenum may seem counter-intuitive, but it is possible that this layer was so thin that the combustion was not self-supporting (i.e., any heat generated was quickly dissipated to the gas, so the cake did not self-heat). The heat dissipation could have been enhanced by greater gas flow through the top plenum due to the lesser flow resistance from the thinner cake on the top.

3.2.4.3 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, the dustcake samples, deposits between filter holders and deposits on the failsafe tester.

In Situ Samples. Tables 3.2-2 and 3.2-3 give the physical properties and chemical compositions of the in situ samples collected at the PCD inlet. As shown in the tables, the gasification ash from the Illinois Basin coal had a lower surface area than that of the PRB gasification ash. In the past, it was seen that the surface area of the PRB gasification ash generally increased with the ash carbon content. Figure 3.2-5 shows the trend of surface area versus carbon content that was previously established for the PRB gasification ash. The TC17 data for the Illinois Basin coal is also included for comparison. The TC17 data do not follow the trend that was established for the PRB gasification ash. This difference suggests that the bituminous gasification ash has less internal

porosity or that some of the internal pore structure has been blocked by adsorbed organic compounds. The contamination of the outlet sampling filters with organic compounds, which was mentioned in Section 3.2.2, suggested that the Illinois Basin coal produced more heavy organic compounds than did the PRB coal.

Hopper Sample. In addition to the in situ samples, Tables 3.2-2 and 3.2-3 include data for the composite PCD hopper sample that was used for laboratory drag measurements. In terms of both the physical properties and the chemical composition, the composite hopper sample was very similar to the in situ samples, suggesting that it was a representative sample for the laboratory drag measurements.

Dustcake and Bridged Deposit Samples. Tables 3.2-4 and 3.2-5 give the physical properties and chemical composition of the dustcake and deposit samples taken after TC17A on November 5, 2004, and after TC17B on December 8, 2004. The properties of the TC17A deposit between the filter holders closely resembled the properties of the TC17A in situ samples, which indicated that the deposit was actually part of the bridged material that filled the bottom plenum when the PCD hopper was overfilled. The dustcake samples had a finer particle size than that observed in previous tests. The dustcake samples also had lower loss on ignition (LOI) values and lower non-carbonate carbon (NCC) content, which suggested a more efficient burnout than the deposits. This difference was expected, since the gas flow through the dustcake produced more contact with oxygen.

Compared to the dustcake from TC17A, the dustcake from TC17B had lower LOI and NCC on the top plenum, but higher LOI and NCC on the bottom plenum. As mentioned previously, this difference may have resulted from the presence of an inner combustion-ash layer on the top plenum elements, which were not replaced after TC17A. For this reason, the dustcake on the bottom plenum was probably more representative of the TC17B burnoff effectiveness. Therefore, it appears that the TC17B burnoff was not as effective as the TC17A burnoff.

3.2.4.4 Dustcake Flow Resistance

Lab Drag Measurements. Since many gasifiers use filter systems to protect downstream equipment, the flow resistance (drag) of the dust cake collected on the filters is an important characteristic of the gasification ash. Gasifiers can produce dusts that have 100 times the drag of fly ashes and knowing the drag in advance is required for accurate PCD design. At the PSDF, drag measurements are made with a device developed by Southern Research Institute known by the acronym RAPTOR (Resuspended Ash Permeability Tester for Operational Research). This device works by resuspending the dust in a fluidized-bed dust generator then collecting the suspended particles on a filter while flow and pressure drop are monitored. The collection device was optimized to collect a dust cake of uniform thickness. After sufficient dust is collected the device is disassembled and the dust cake thickness is measured and the mass determined. From these parameters the normalized drag (pressure drop normalized to 1 ft/min face velocity and 1 lb/ft² areal dust loading) can be calculated. A series of cyclone particle collectors are inserted between the dust generator and the collection filter so that the drag can be determined for four different median particle sizes between 2 and 20 microns. This allows a plot of drag versus particle size and provides information on filter pressure drop expected with gasifier recycle systems of various characteristics.

Drag measurements were made on the hopper sample described previously, and the results are illustrated in Figure 3.2-6. The data from two previous runs with other bituminous coals (Alabama bituminous in TC07C and Hiawatha bituminous in TC09) are also shown for comparison, along with the data from the TC15 test run with PRB coal. Comparing only the bituminous data, the gasification ashes produced in TC17 and TC09 have very similar drag characteristics. These two particular samples also had essentially the same carbon contents (55 percent for the Illinois Basin ash in TC17 and 53 percent for the Hiawatha bituminous ash in TC09). The gasification ash from the Alabama bituminous coal had a higher drag, possibly because it had a slightly higher carbon content (60 percent). In the past, the correlation of drag with carbon content did not always apply across different coal types, and the bituminous ashes exhibited less drag than the PRB ashes.

Transient PCD Drag. During each in situ sampling run, the PCD transient drag was calculated using the measured pressure drop, gas flow, and particle concentration (see Table 3.2-6). The calculated transient drag at PCD conditions is listed under the column heading “PCD.” The corresponding normalized value of transient drag at room temperature is listed under the heading “PCD@RT” and is plotted as a function of non-carbonate carbon (NCC) content in Figure 3.2-7, along with data from recent PRB runs and data from the high-sodium lignite run in TC16. There is considerable scatter in the data due to variations in equipment configuration, process conditions, coal composition, and limestone addition. Nevertheless, the data show a definite trend toward increasing drag with increasing carbon content. The drag data obtained with the high-sodium lignite and with the Illinois Basin coal fall on lower trend lines than do the data from the PRB runs. This downward shift in the drag trend line could be related to morphological differences between the ashes produced from lignite, PRB, and bituminous coals.

Comparison of Lab Measurements with Transient Drag. Average lab and PCD drag values for all gasification test runs are summarized in Table 3.2-7 and plotted in Figure 3.2-8. The comparison shows excellent overall agreement (average difference of 6.9 percent), even though the difference is much higher for certain test programs. For TC17, the difference was about 28 percent, which seems very reasonable considering all of the uncertainties in various factors that are involved in the calculations. The plot of average lab versus PCD drag values shows that the data points are almost symmetrically scattered around the perfect agreement line. As illustrated in the graph, the perfect agreement line falls within the 95 percent confidence interval on the regression line to the data, which indicated good agreement between the data sets.

3.2.5 Filter Element Gasket Testing

Filter element gasket testing was performed on the clean side of the PCD during the test run to evaluate methods of reducing gasket leakage and to test the material exposure to syngas. These gaskets are used to attach the filter elements and failsafes to the plenum. Laboratory tests have shown gas penetration through the gaskets and that the gas penetration rate increases with loss of bolt torque. Small particles of gasification ash may penetrate and pass through the gaskets resulting in particulate emissions. As the gaskets gradually become plugged with ash particles, the particulate emissions are expected to drop. This was seen in the last four test runs where the outlet particulate loading was initially elevated (0.2 to 0.4 ppmw) and gradually decreased to less than 0.1 ppmw after the first few days of operation.

One possible method of reducing gasket leakage is to replace the fiber gaskets with a non-fibrous, non-porous gasket material. To investigate the feasibility of various potential replacement gasket materials, test fixtures containing six different types of gaskets were tested. The six gasket types were: (1) the ceramic fiber gasket that has been used in the past, (2) a Garlock 3125SS graphite with stainless steel insert, (3) a Novatec 825 graphite with no insert, (4) a Garlock 9900 graphite with binder, (5) a Thermiculite sheet gasket, and (6) a spiral-wound Flexitallic gasket with thermiculite filler. The different gasket types were bolted into identical test fixtures that simulated the sealing between the filter element and filter element holder. Each test fixture was torqued to the nominal specification of 100 in-lbs prior to the test. The fixtures were then installed on the clean side of the PCD.

After the clean syngas exposure in TC17, the fixtures were removed and gas leakage through the seals was measured. The leakage measurements were made with an applied differential pressure of 200 inH₂O in the forward gas flow direction to simulate PCD operation at a high pressure drop. Leakage measurements were also made with an applied pressure of 20 psig in the reverse direction to simulate the maximum pressure generated in the element during a backpulse. For the fiber gaskets, the forward and reverse leakage rates were 0.10 and 0.44 cfm, which were somewhat lower than the leakage rates measured in previous lab tests at similar final torque values of approximately 50 in-lbs. This anomaly was probably a result of the multiple retorquing that was performed on the test fixtures before installation. The test fixtures exposed in TC17 were retorqued six times due to delays in the installation. In the lab tests, the test fixtures were only retorqued twice.

For all of the other non-fibrous gasket types, there was essentially no measurable leakage, even though the loss of torque was similar to that measured with the fiber gaskets (~ 40 to 60 in-lbs). While the lack of leakage with the graphite and thermiculite gaskets was encouraging, the loss of torque could still present problems. The loss of torque could allow some movement of the element at each back-pulse. Since the graphite and thermiculite gaskets do not provide any cushioning between the element and the tubesheet, the back-pulsing could transmit force to the element and possibly result in long-term damage. Because of this effect, higher values of applied torque should be used with the non-fibrous gaskets as recommended by the gasket manufacturers. Unfortunately, this is not possible with the existing ¼-in. bolts, since the applied load would exceed the recommended safety allowance on the bolt tensile strength. Because of space constraints, it is not feasible to increase the bolt size. Other methods of improving torque retention and sealing without the use of higher applied torque values or higher compressive loads on the gaskets will be investigated in future test campaigns.

3.2.6 Filter Element and Failsafe Testing

Filter element testing continued with 72 iron aluminide (FEAL) elements. At the conclusion of TC17, several of the filter elements had been exposed to syngas for approximately 5,800 hours. Inspection results are discussed in Section 2.4.

Several FEAL elements with a modified weld structure were further evaluated. The FEAL filter elements are made from multiple filter media sections. The adjacent sections are welded together with solid metal adaptors, with the welding point outside of the media section. Because of the difficulty in welding the porous media to the solid metal adaptor, the weld structure has tended to be the weakest point in the filter element, with occasional filter element failures occurring at or near the

welds under unusual stresses. To strengthen the weld structure, a modified type of weld configuration was proposed by the manufacturer. The modified configuration has a ring adaptor on the outside of the media with the welding point inside the ring. The ring adaptor was thought to be a strengthener. However, it was found that the welding inside the media section was quite challenging and yielded inconsistent weld quality. As a result, several such filter elements failed at the welds during installation, modifying, and operation when elements with the previous weld structure stayed intact. The modified weld structure thus was proved inferior to the previous weld design and therefore the use of the modified design was discontinued after TC17.

Testing continued with the evaluation of Pall fuses, CeraMem failsafes, and PSDF-designed failsafes. These failsafes are designed to close off the gas outlet of individual filter elements in the event of an element failure, preventing particle laden gas from traveling to downstream equipment. A CeraMem failsafe was successfully tested with the rupture disc failsafe tester to simulate a catastrophic filter element failure. The outlet loading at the end of the test was less than 0.1 ppmw, which indicated that the failsafe performed well. The failsafes tested as well as the on-line testing device have been described previously in the paper “Hot Gas Filtration Meeting Turbine Requirements for Particulate Matter” by Garder, et al, 2005, which is available from the technical papers section of the PSDF website, <http://psdf.southernco.com/>.

Table 3.2-1 PCD Inlet and Outlet Particulate Measurements

Test Date	PCD Inlet					PCD Outlet				
	Run No.	Start Time	End Time	Particle Loading,		Run No.	Start Time	End Time	H ₂ O Vapor, vol %	Particle Loading, ppmw
				ppmw	lb/hr					
<i>PRB</i>										
10/27/04	--	--	--	--	(1)	1	10:00	13:50	11.4	0.19
10/28/04	--	--	--	--	(1)	2	8:45	9:25	--	(2)
10/29/04	1	10:30	10:45	18100	364	3	9:15	13:15	12.2	< 0.1
<i>Illinois Basin</i>										
10/30/04	--	--	--	--	--	4	17:00	18:00	19.3	~ 1 ⁽³⁾
11/1/04	2	9:45	10:00	19100	412	5	9:30	13:30	24.9	~ 1 ⁽³⁾
<i>Shutdown -- Replace Broken Filter Elements -- Continue with Illinois Basin</i>										
11/10/04	--	--	--	--	--	6	13:00	14:00	3.0	< 0.1 ⁽⁴⁾
11/12/04	3	11:00	11:15	32300	608	7	10:00	14:00	15.5	0.2-0.3 ⁽³⁾
11/13/04	4	8:46	9:01	26700	527	8	7:30	11:30	16.2	0.3-0.4 ⁽³⁾
11/15/04	5	12:40	12:55	22800	440	9	13:00	14:30	14.0	0.3-0.4 ⁽³⁾
11/16/04	6	9:45	10:00	27600	502	10	8:15	13:15	15.5	0.4-0.6 ⁽³⁾
11/17/04	7	9:15	9:30	27600	517	11	8:45	12:45	11.9	0.4-0.6 ⁽³⁾
11/18/04	8	9:00	9:15	24500	451	12	8:45	12:45	13.0	0.4-0.6 ⁽³⁾
<p>Notes:</p> <ol style="list-style-type: none"> 1. Gasifier operation unstable. No sample taken. 2. TR upset. Heavy tar contamination of the filter. Emission rate cannot be estimated. 3. Tar contamination of filter. Emission rate estimated by microscopic examination of sampling filter. 4. Operation with startup burner - no coal. Trash particles excluded from emission estimate. 										

Table 3.2-2 Physical Properties of In Situ and Hopper Samples

Sample ID	Run No.	Sample Date	Bulk Density g/cc	True Density g/cc	Uncompacted Bulk Porosity %	BET Specific Surface Area m ² /g	Mass Median Diameter μm
<i>In-Situ Samples, PRB Coal</i>							
AB16399	1	10/29/04	0.28	2.33	88.0	203	18.5
<i>In-Situ Samples, Illinois Basin</i>							
AB16400	2	11/1/04	0.24	2.34	89.7	113	27.2
AB16401	3	11/12/04	0.24	2.36	89.8	63	26.9
AB16402	4	11/13/04	0.22	2.26	90.3	87	20.2
AB16403	5	11/15/04	0.21	2.31	90.9	75	29.3
AB16404	6	11/16/04	0.20	2.16	90.7	114	29.0
AB16405	7	11/17/04	0.22	2.09	89.5	76	18.0
AB16406	8	11/18/04	0.19	2.26	91.6	74	22.4
<i>Average</i>			0.22	2.26	90.4	86	24.7
<i>Hopper Sample, Illinois Basin</i>							
AB16407	...	11/15/04	0.24	2.23	89.2	93	18.9

Table 3.2-3 Chemical Composition In Situ and Hopper Samples

Sample ID	Run No.	Sample Date	CaCO ₃ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %
<i>In-Situ Sample, PRB Coal</i>								
AB16399	1	10/29/04	3.95	0.27	10.36	33.84	51.59	38.34
<i>In-Situ Samples, Illinois Basin</i>								
AB16400	2	11/01/04	3.41	0.96	5.56	45.72	44.35	47.44
AB16401	3	11/12/04	2.95	2.01	2.33	49.06	43.66	54.00
AB16402	4	11/13/04	2.98	1.96	2.56	51.02	41.47	55.71
AB16403	5	11/15/04	2.84	2.00	2.10	51.64	41.42	55.94
AB16404	6	11/16/04	2.73	2.01	4.00	52.79	38.48	55.28
AB16405	7	11/17/04	2.66	2.25	1.65	62.38	31.06	64.77
AB16406	8	11/18/04	2.66	2.30	2.32	54.68	38.05	56.53
<i>Average</i>			<i>2.89</i>	<i>1.93</i>	<i>2.93</i>	<i>52.47</i>	<i>39.78</i>	<i>55.67</i>
<i>Composite Hopper Sample, Illinois Basin</i>								
AB16407	---	11/15/04	2.32	2.40	2.12	55.45	37.70	59.27

Table 3.2-4 Physical Properties of Residual Dustcakes and Deposits

Sample ID	Sample Date	Bulk Density g/cc	True Density g/cc	Uncompacted Bulk Porosity %	Specific Surface Area m ² /g	Mass-Median Diameter μm	Loss on Ignition Wt %
<i>TC17A Bulk Dustcake from Top Plenum</i>							
AB16253	11/05/04	0.37	2.82	86.9	25	4.0	12.0
<i>TC17A Bulk Dustcake from Bottom Plenum</i>							
AB16252	11/05/04	0.35	2.72	87.1	34	5.3	17.3
<i>TC17A Deposit Between Filter Holders</i>							
AB16250	11/05/04	0.22	2.24	90.2	84	20.3	55.4
<i>TC17A Deposit on Failsafe Tester</i>							
AB16249	11/05/04	0.32	2.58	87.6	29	15.5	31.5
<i>TC17B Bulk Dustcake from Top Plenum</i>							
AB16456	12/08/04	0.36	2.81	87.2	22	3.9	8.6
<i>TC17B Bulk Dustcake from Bottom Plenum</i>							
AB16457	12/08/04	0.31	2.41	87.1	55	6.5	24.1

Table 3.2-5 Chemical Composition of Residual Dustcakes and Deposits

Sample ID	Sample Date	CaCO ₃ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %
<i>TC17A Bulk Dustcake from Top Plenum</i>							
AB16253	11/5/04	2.07	3.14	14.12	9.85	70.83	11.96
<i>TC17A Bulk Dustcake from Bottom Plenum</i>							
AB16252	11/5/04	2.23	2.59	13.00	14.96	67.22	17.32
<i>TC17A Deposit Between Filter Holders</i>							
AB16250	11/5/04	2.30	1.38	3.73	50.39	42.20	55.39
<i>TC17A Deposit on Failsafe Tester</i>							
AB16249	11/5/04	2.41	1.35	12.31	28.25	55.68	31.48
<i>TC17B Bulk Dustcake from Top Plenum</i>							
AB16456	12/8/04	1.73	1.55	10.26	6.12	80.34	8.55
<i>TC17B Bulk Dustcake from Bottom Plenum</i>							
AB16457	12/8/04	1.95	1.05	7.94	20.83	68.23	24.08

Table 3.2-6 Transient Drag Determined from PCD Pressure Drop and from Lab Measurements

Run No.	$\Delta P/\Delta t$, inwc/min	$\Delta(AL)/\Delta t$, lb/ft ² /min	FV, ft/min	MMD, μm	NCC, %	Drag, inwc/(lb/ft ²)/(ft/min)		
						PCD	PCD@RT	Lab
PRB								
1	1.65	0.028	3.26	18.5	33.8	59	35	---
Illinois Basin								
2	7.34	0.032	3.93	27.2	45.7	231	124	15
3	1.51	0.047	3.51	26.9	49.1	32	18	16
4	1.88	0.041	3.83	20.2	51.0	46	25	23
5	1.80	0.034	3.70	29.3	51.6	53	28	14
6	1.90	0.039	3.51	29.0	52.8	49	26	14
7	1.62	0.040	3.62	18.0	62.4	41	22	26
8	1.98	0.035	3.62	22.4	54.7	57	30	20
AVG	1.78	0.039	3.63	24.3	53.6	46	25	19
<p>Note: Lab drag data calculated from linear regression to MMD. Averages for Illinois Basin do not include Run No. 2 due to unusually high $\Delta P/\Delta t$.</p>								

Nomenclature:

$\Delta P/\Delta t$ = rate of pressure drop rise during particulate sampling run, inwc/min.

$\Delta(AL)/\Delta t$ = rate of increase in areal loading during sampling run, lb/min/ft².

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.

RT = room temperature, 77°F (25°C).

Table 3.2-7 Comparison of Average Drag Values Determined from PCD Performance and from 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, %
GCT2	PRB	29.3	20.9	-33.5
GCT3	PRB	80.2	92.7	14.5
GCT4	PRB	66.4	57	-15.2
TC06	PRB	89.4	81.2	-9.6
TC07	PRB	47.7	49.8	4.3
TC08	PRB	46.5	50	7.3
TC09	Hiawatha	29.0	23.3	-21.8
TC10	PRB	44.7	57.6	25.2
TC11	Falkirk Lignite	16.1	35.9	76.2
TC12	PRB	58.0	60.8	4.7
TC13	Freedom Lignite	34.4	39.4	13.6
TC14	PRB	47.4	41.6	-13.0
TC15	PRB	54.6	76.4	33.3
TC16	PRB + Limestone	49.3	51.7	4.8
TC16	Lignite + Dolomite	25.8	41.7	47.1
TC17	IL Basin + Dolomite	24.8	18.7	-27.8
<i>Average</i>		46.5	49.9	6.9

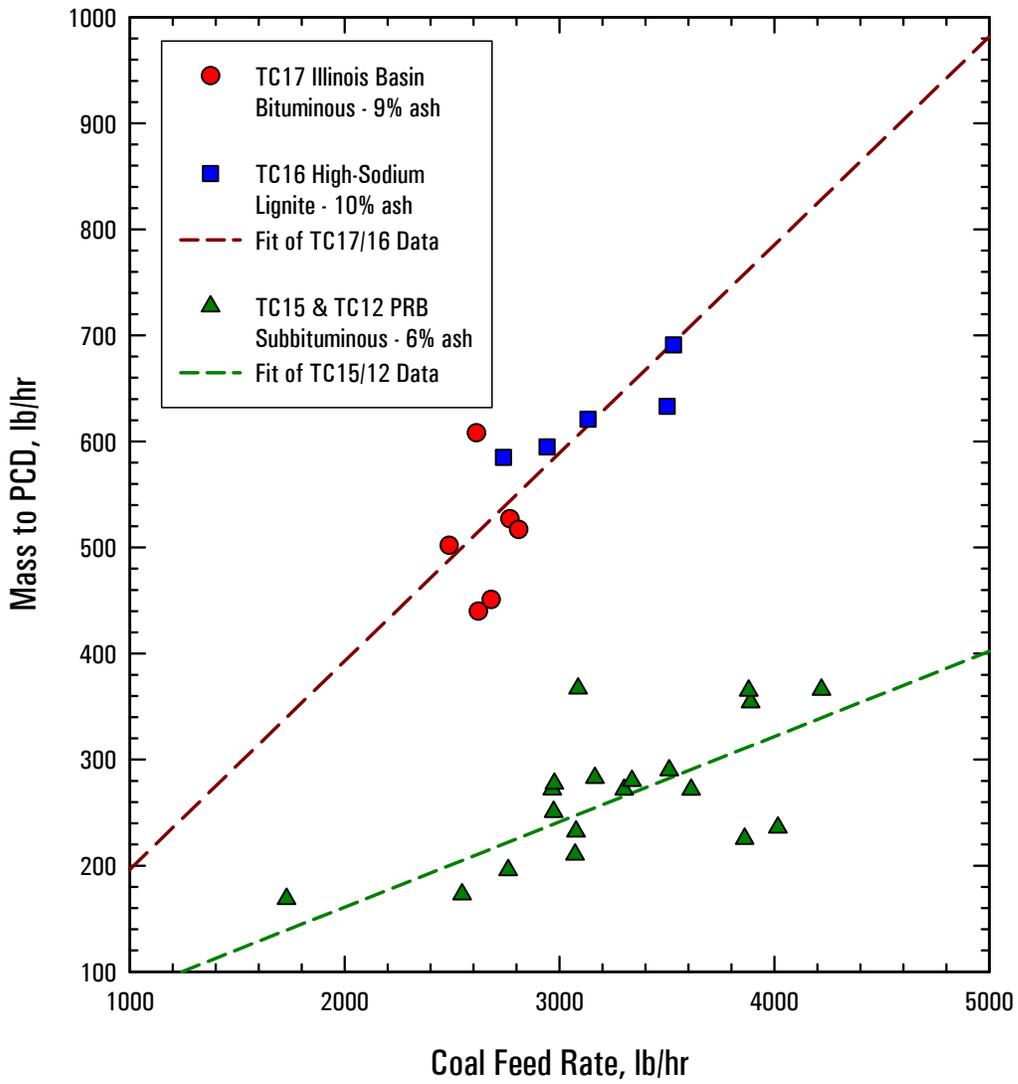


Figure 3.2-1 PCD Inlet Particle Concentration as a Function of Coal Feed Rate

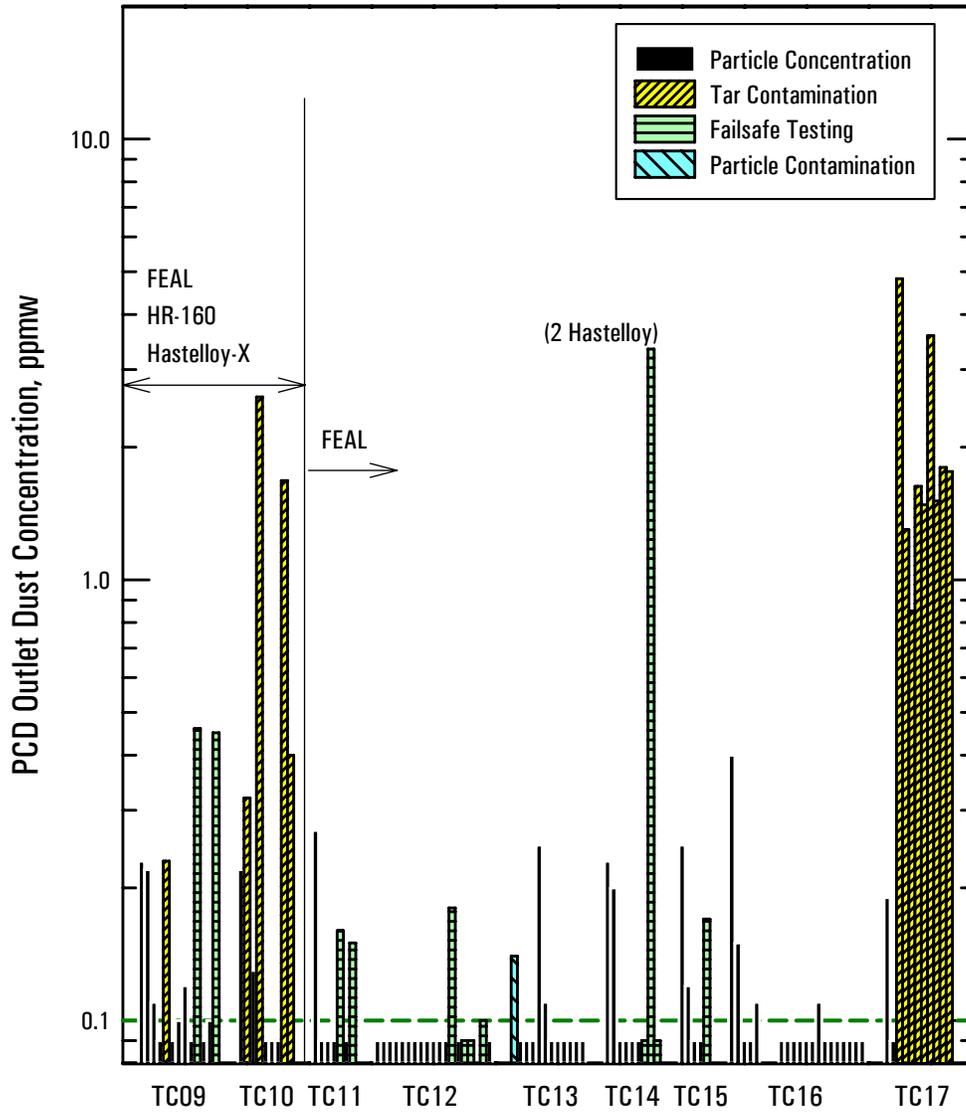


Figure 3.2-2 PCD Outlet Dust Concentration in Recent Gasification Runs

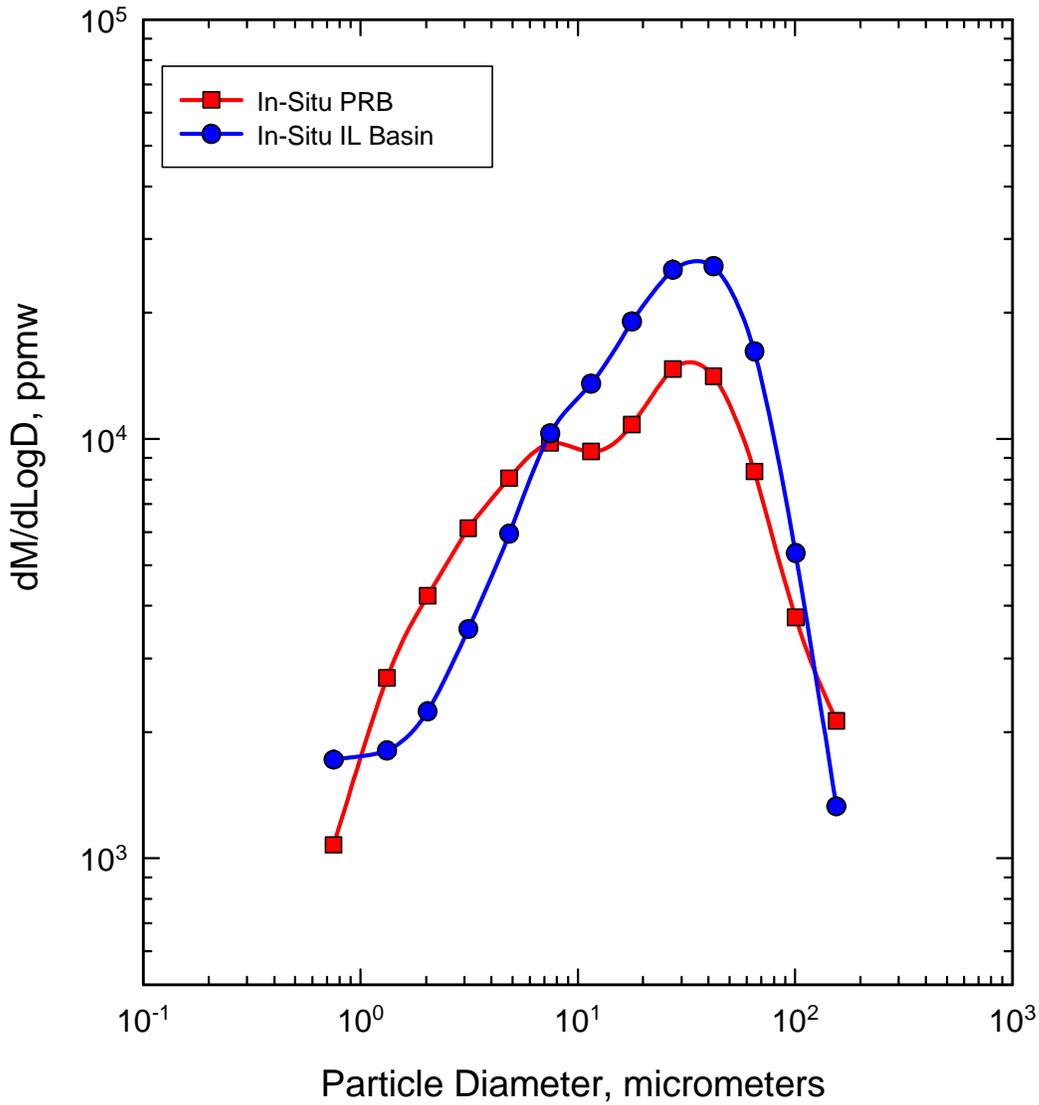


Figure 3.2-3 Comparison of Illinois Basin Coal and PRB Coal Particle Size Distributions

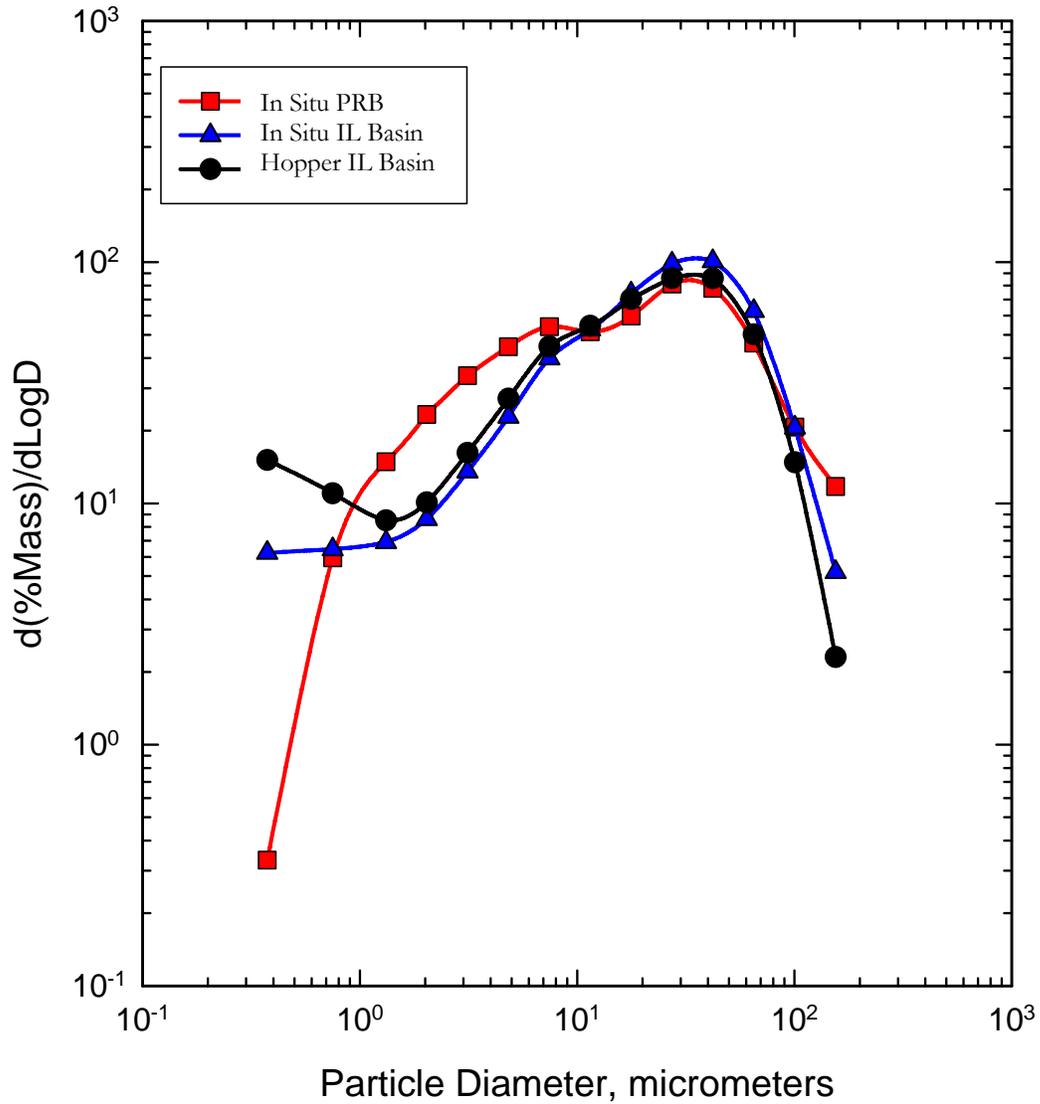


Figure 3.2-4 Comparison of Hopper and In Situ Particle Size Distributions

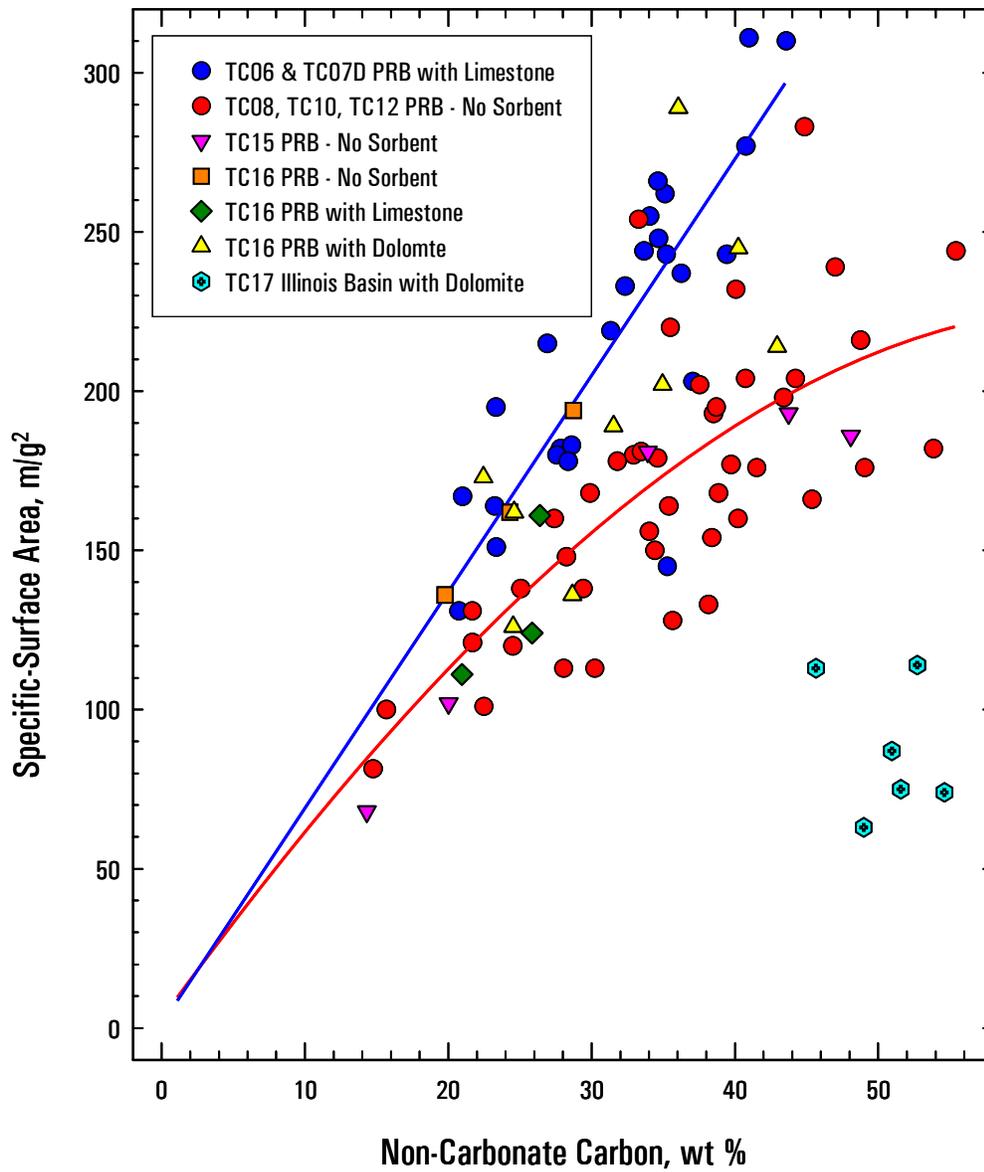


Figure 3.2-5 Specific Surface Area versus Carbon Content of In Situ Samples

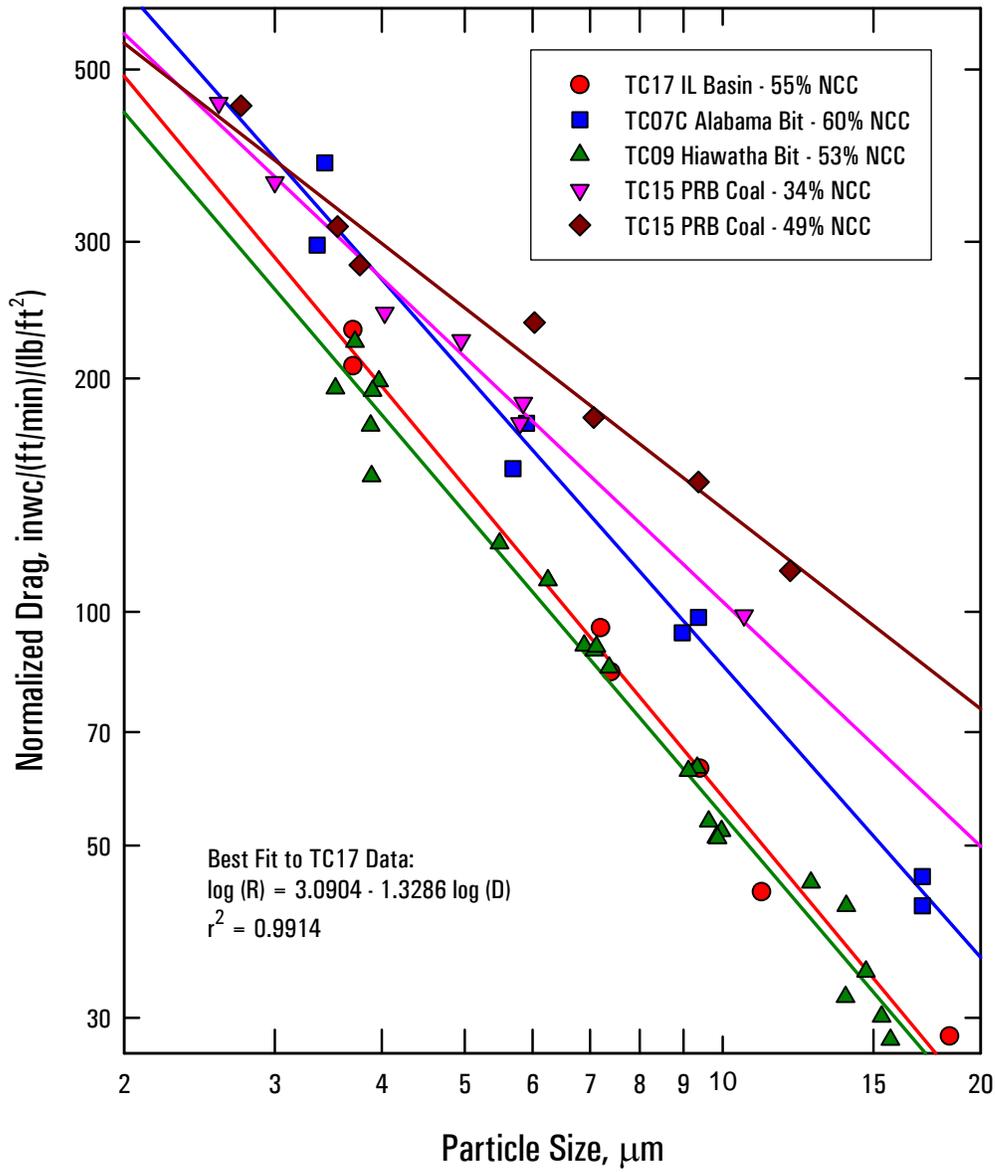


Figure 3.2-6 Lab-Measured Drag as a Function of Particle Size

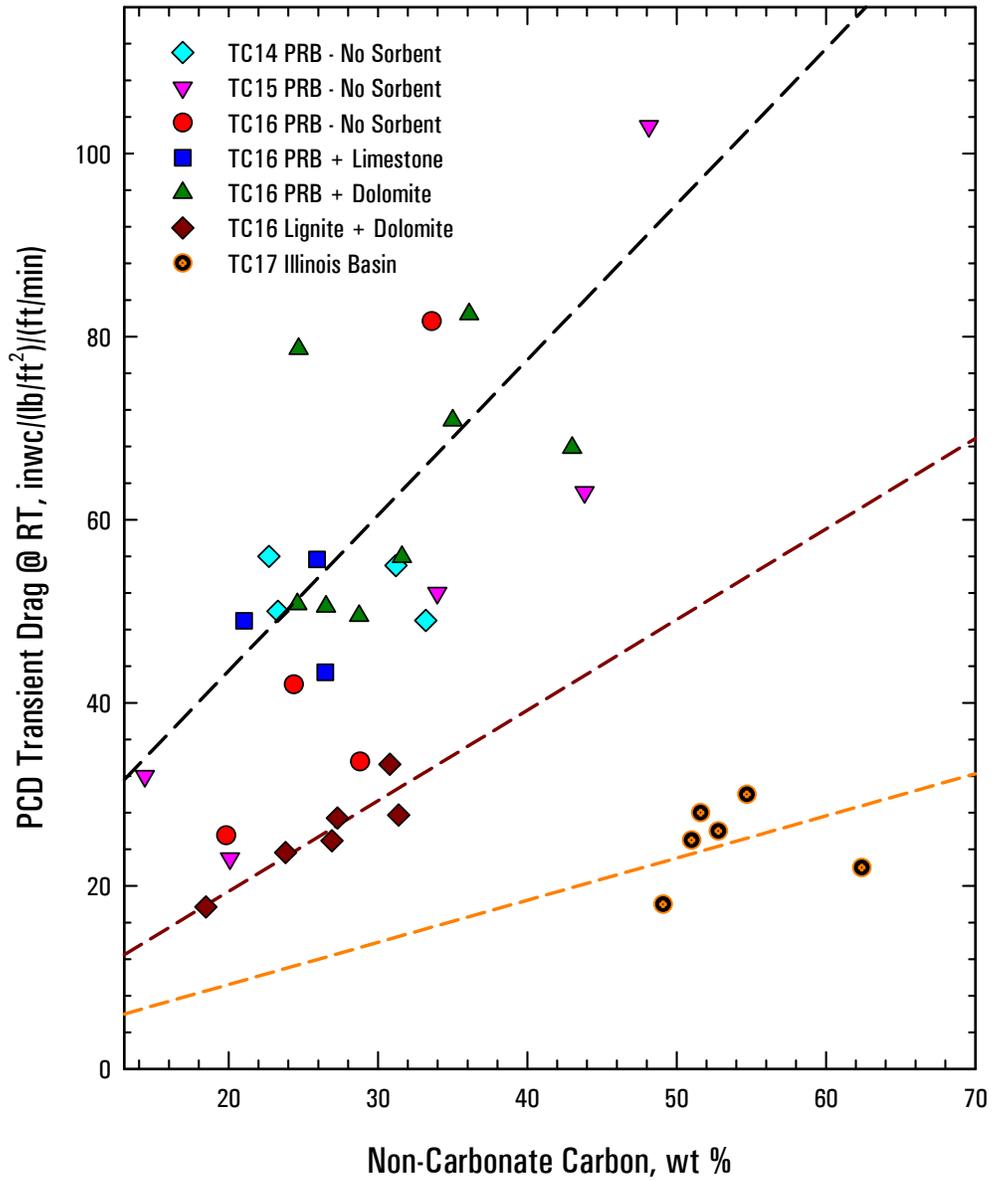


Figure 3.2-7 PCD Transient Drag versus Carbon Content of In Situ Samples.

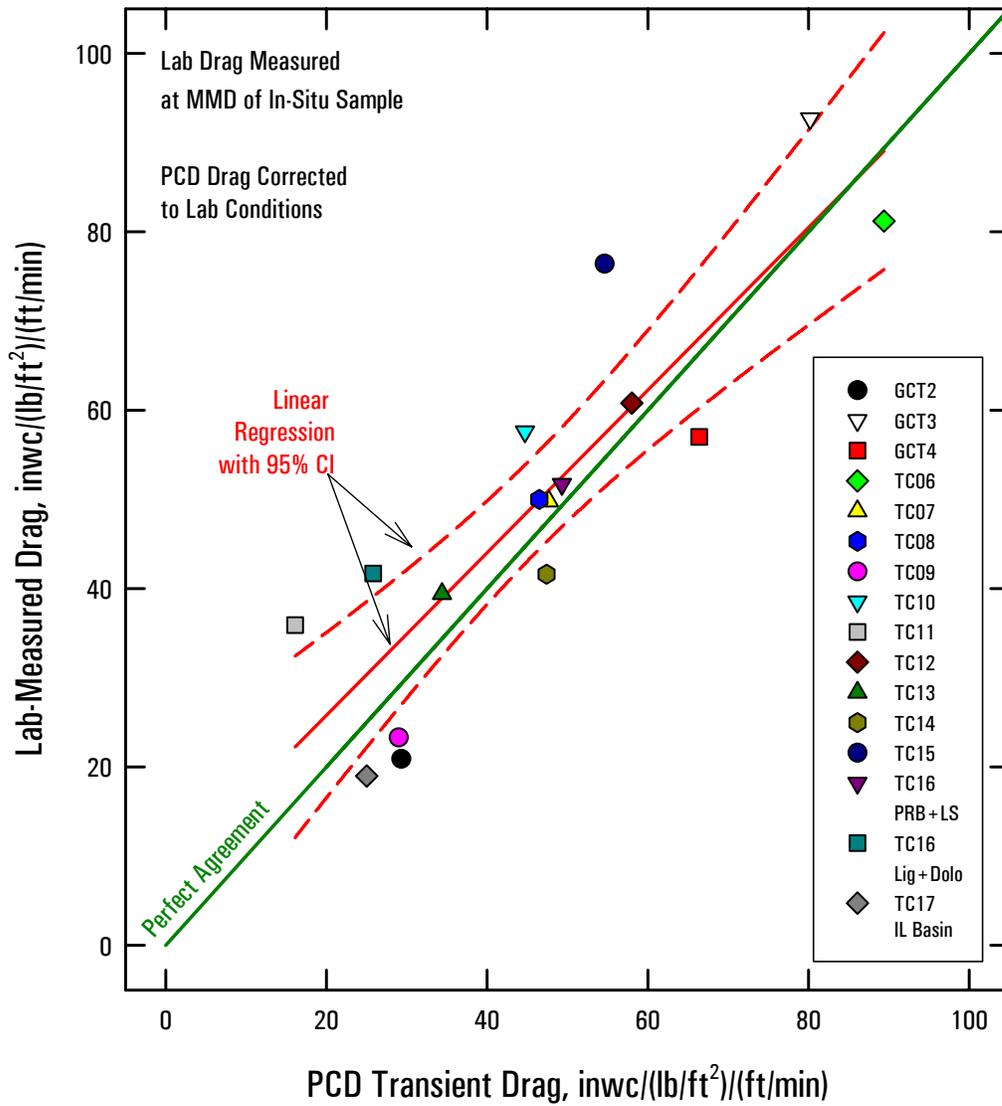


Figure 3.2-8 Comparison of PCD Transient Drag with Laboratory Measurements

3.3 PERFORMANCE OF OTHER SYSTEMS

3.3.1 Piloted Syngas Burner

On October 29, 2004, the piloted syngas burner (PSB) operated most of the day on propane and approximately 6.5 hours on syngas. The syngas flow rate to the PSB was around 13,000 lb/hr, with a dry lower heating value of approximately 70 to 75 Btu/SCF. The output from the combustion turbine was approximately 2.5 MW. The PSB was shut down prior to the transition to bituminous coal due to concerns of lower coal conversion resulting in tar deposition in the syngas inlet line to the PSB.

3.3.2 Coal Preparation and Feed Systems

The new coal silo insert was installed to prevent coal particle size segregation by reducing rat-hole flow inside the silo, thus preventing excessive fines in the coal feeder. A liner made of low-friction Tivar material was also added to the bottom of the silo hopper to promote flow. The modifications are shown in Figure 3.3-1.

Testing with the insert showed improved coal feeder operations, although in the initial test, a slight decrease in mean particle size occurred as the silo was emptied. However, no operational problems occurred, and the decrease in mean particle size as the silo was emptied was less than that observed prior to the addition of the silo insert. A second test of the system resulted in some particle segregation that led to the coal feeder problems seen at the end of the test run. The problems occurred after problems with coal mill operations which negatively affected the coal size and moisture consistency.

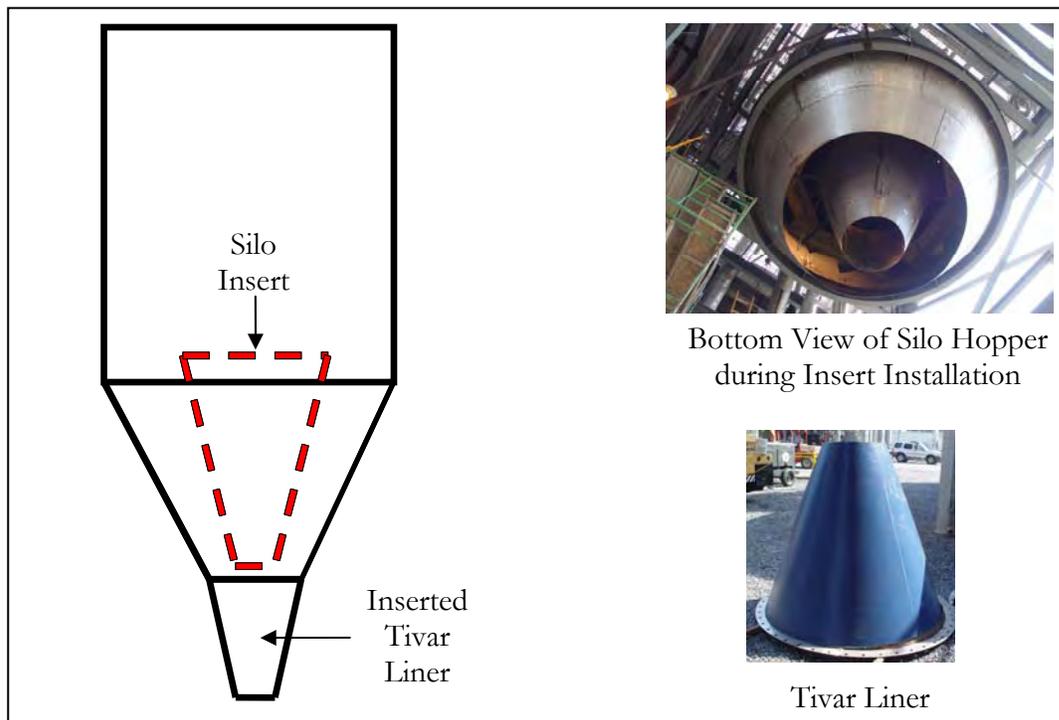


Figure 3.3.2-1 Coal Mill Silo Modifications

The original coal feeder, a Clyde lock hopper system, was the sole source of coal feed for the test run. The average coal feed rate was about 2,000 to 3,400 lb/hr.

3.3.3 Continuous Fine Ash Depressurization (CFAD) System

As in the previous test runs, the continuous fine ash depressurization system (CFAD) operated well with the fines from PRB coal operation. It also operated well with the bituminous coal fines despite the low density of the fines collected (around 6-8 lb/ft³). The system was able to discharge fines at a rate of approximately 500 lb/hr. At the conclusion of TC17, the CFAD system had operated for about 1,500 hours.

3.3.4 Sensor Development

Developmental work with gasifier temperature and pressure differential measurements was continued during TC17. To improve gasifier temperature measurement reliability, various materials and configurations were tested. The effect of thermowell insertion length was also studied. Three thermowell materials were tested in the riser, including ceramic, Stellite #1 coated Hastelloy X, and HR-160. Three of the five HR-160 thermowells failed during the first part of TC17 and were replaced. The same three failed again by the end of the test run. The other two HR-160 thermowells showed significant wear. The Hastelloy X thermowell lasted longer than the HR-160 but showed considerable wear upon inspection. The ceramic thermowell lasted the entire run and showed only about 5 percent wear upon inspection at the end of the run. The thermowell configuration utilizing a Sialon spoiler inserted at a steep angle lasted five hours more than the one without a spoiler; however, the temperature deviation was almost 100°F at times. The Sialon spoiler did not show any noticeable signs of wear, but additional testing will not be conducted due to the high temperature measurement error caused by the spoiler. Based on the different insertion lengths tested, it was determined that a 2 inch insertion beyond the refractory wall plane is sufficient for 0.75 inch thermowells. There was no difference between the measurements made at insertion lengths from 2 to 8 inches. However, the condition of the refractory around the thermowell hole is an important factor to evaluate in a new refractory installation.

Ceramic filters were installed in one gasifier pressure differential measurement to reduce purge flow requirements and to prevent plugging of nozzles. Although balancing the measurement was problematic, the measurement correlated well with other differential pressure measurements with only a slight offset. The constant purge flow controllers also performed well maintaining a stable flow at all gasifier pressures.

3.3.5 Advanced Syngas Cleanup

Advanced gas cleanup technology testing was performed using a slipstream of syngas from the Transport Gasifier during Illinois Basin coal operation. Testing consisted of fixed bed desulfurization with a sulfur sorbent from Sud-Chemie, fixed bed hydrocarbon cracking using a Fluid Catalytic Cracking (FCC) catalyst from Grace Davison, and ammonia cracking and hydrocarbon reforming using a Sud-Chemie catalyst. The testing was performed independently in parallel reactors.

The nominal properties of the Sud-Chemie sulfur sorbent (RVSLT-1) and the Grace Davison hydrocarbon cracking catalyst are shown in the Tables 3.3 -1 and 3.3-2, respectively. The nominal properties of the Sud-Chemie ammonia cracking and hydrocarbon reforming catalyst (G-117RR) are shown in Table 3.3-3. The operating parameters for fixed bed sulfidation and hydrocarbon cracking in the hot vessels (RX700A and B) using syngas are shown in Tables 3.3-4 and 3.3-5, respectively, while the operating parameters for ammonia cracking and hydrocarbon reforming in the mini reactor (RX301) using syngas are shown in Table 3.3-6.

The major accomplishments and observations are listed below:

- Sud-Chemie sulfur sorbent (RVSLT-1) was installed in the first hot vessel (RX700A) for syngas desulfurization. The desulfurization test was conducted at about 600°F and 200 psig pressure with a typical H₂S inlet concentration of 600 to 700 ppm. The H₂S concentration at the outlet was as low as 3 ppm. A sulfur loading of 30.7 weight percent was achieved in the RVSLT-1 sorbent. Breakthrough occurred at about hour 43 of the testing. The temperature and pressure profiles are shown in Figures 3.3-2, and the sulfur profile (dry basis) is shown in Figure 3.3-3.
- An FCC catalyst from Grace Davison was installed in the second hot vessel (RX700B) for syngas hydrocarbon cracking. The hydrocarbon cracking test was conducted at around 750 to 850°F and 200 psig. High reduction levels were demonstrated. Ethylene was reduced by nearly 100 percent; acenaphthene was reduced by about 87 percent; phenanthrene was reduced by about 75 percent; and naphthalene was reduced by about 62 percent. The temperature and pressure profiles are shown in Figure 3.3-4. The hydrocarbon profiles, all on a dry basis, are shown in Figures 3.3-5 through 3.3-8.
- A Sud-Chemie nickel based catalyst (G-117RR) was installed in the mini reactor, RX301, for syngas ammonia cracking and hydrocarbon reforming. The mini reactor test was conducted at temperatures from about 1650 °F to 1750 °F and pressures from 2 to 10 psig. The ammonia concentration at the inlet of RX301 was typically around 2000 ppm, and the outlet concentration was typically 40 ppm, resulting in a removal efficiency of around 98 percent. The ammonia profile is shown in Figure 3.3-9. The catalyst also reduced syngas hydrocarbon concentrations. Phenanthrene was reduced by 92 percent, naphthalene was reduced by 78 percent, and acenaphthene was reduced by 92 percent. The hydrocarbon profiles are shown in Figures 3.3 -10 to 3.3-12. All concentration profiles are on a dry basis.
- The electrical band heaters around the hot vessels, RX700A and B, worked well to control the heating and cooling during testing.
- Gas analyzers were used intermittently during the test run with the hot vessels and mini reactor. Duration of the gas sampling time was restricted due to condensation in the gas sample lines.

Table 3.3-1 Nominal Properties of Sulfur Sorbent in Hot Vessel, RX700A

Sorbent	RVSLT-1
Manufacturer	Sud-Chemie
<u>Chemical Composition</u>	<u>Weight Percent</u>
Zinc Oxide	50 - 70
Calcium Sulfate	15 - 30
Calcium Oxide	5 - 15
Nickel Oxide	-
Bentonite	5 - 15
Silica, Quartz	<5
<u>Physical Properties</u>	
Shape	Spherical
Size	3 - 4 mm
Density	60 - 85 lb/ft ³

Table 3.3-2 Nominal Properties of Hydrocarbon Cracking Catalyst in Hot Vessel, RX700B

Catalyst	FCC Catalyst
Manufacturer	Grace Davison
<u>Chemical Composition</u>	<u>Weight Percent</u>
Silica SiO ₂	40 - 80
Alumina Al ₂ O ₃	20 - 60
Rare Earths Re ₂ O ₃	0 - 10
Sulfate SO ₄	0.1 - 2.4
Sodium Oxide Na ₂ O	0 - 1.0
Titania TiO ₂	0 - 1.0
Total Volatiles	2 - 6
Quartz SiO ₂ (Max)	1.0
<u>Physical Properties</u>	
Shape	Powder
Size, D ₅₀	70 micron
Density	28.1 - 62.6 lb/ft ³

Table 3.3-3 Nominal Properties of Ammonia Cracking and Hydrocarbon Reforming Catalyst in Mini Reactor, RX301

Catalyst	G-117RR
Manufacturer	Sud-Chemie
<u>Chemical Composition</u>	<u>Weight Percent</u>
Magnesium Oxide	75 – 90
Nickel Oxide	5 - 15
Calcium Oxide	1 - 5
Aluminum Oxide	1 - 5
<u>Physical Properties</u>	
Shape	Rings
Size	3 - 4 mm
Density	55 - 75 lb/ft ³

Table 3.3-4 Operating Parameters for Fixed Bed Syngas Desulfurization in Hot Vessel, RX700A

Sorbent	RVSLT-1
Gasifier Operation	Air Blown
Coal Type	Illinois Basin
Reactor	RX700A
Reactor Size	5.187" ID x 5' Ht
Reactor Material	310 SS
Sorbent bed mass, lb	7
Sorbent bed height, in	7.5
Syngas flow rate, lb/hr	12
Pressure, psig	200
Temperature, °F	600
Space Velocity, hr ⁻¹	1820
Inlet H ₂ S, ppm	600 -- 700

Table 3.3-5 Operating Parameters for Fixed Bed Syngas Hydrocarbon Cracking in Hot Vessel, RX700B

Catalyst	FCC Catalyst
Manufacturer	Grace Davison
Gasifier Operation	Air Blown
Coal Type	Illinois Basin
Reactor	RX700B
Reactor Size	5.187" ID x 5' Ht
Reactor Material	310 SS
Catalyst bed mass, lb	5.5
Catalyst bed height, in	7.5
Syngas flow rate, lb/hr	12
Pressure, psig	200
Temperature, °F	750 -- 850
Space Velocity, hr ⁻¹	1840

Table 3.3 -6 Operating Parameters for Fixed Bed Syngas Ammonia Cracking and Hydrocarbon Reforming in Mini Reactor, RX301

Catalyst	G-117RR
Manufacturer	Sud-Chemie
Gasifier Operation	Air Blown
Coal Type	Illinois Basin
Mini Reactor	RX301
Mini Reactor Size	1.5" ID x 4' Ht
Reactor Material	310 SS
Catalyst bed mass, lb	0.25
Catalyst bed height, in	5
Syngas flow rate, lb/hr	0.5 - 1
Pressure, psig	2 - 10
Temperature, °F	1650
Space Velocity, hr ⁻¹	2,000 - 3,000
NH ₃ inlet concentration, ppm	2000
NH ₃ conversion, percent	92 - 98

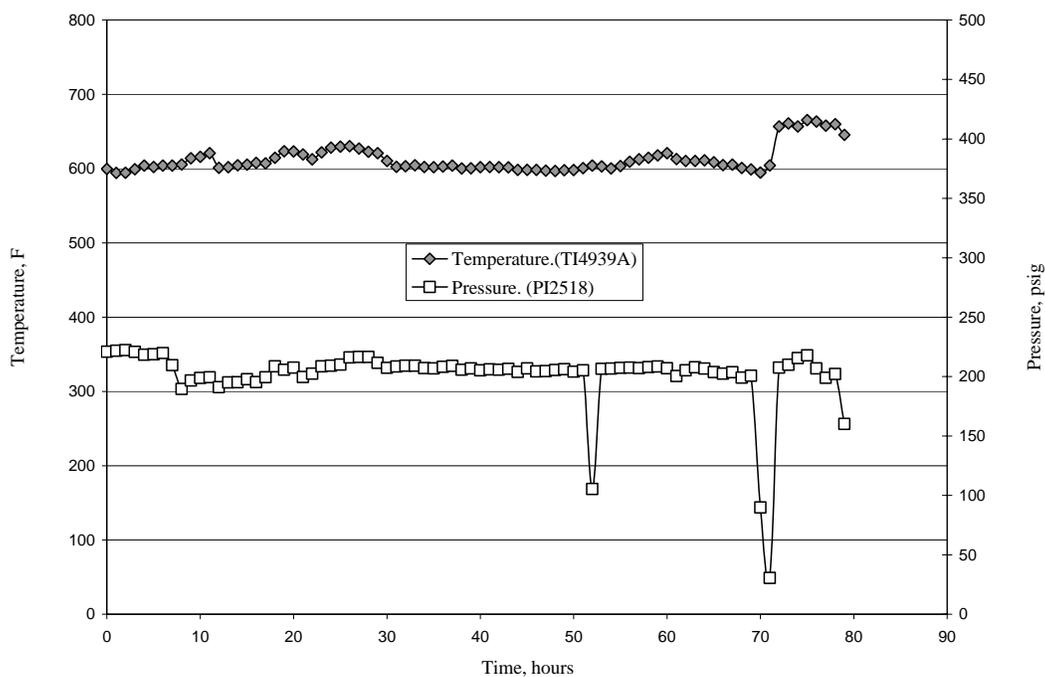


Figure 3.3-2 RX700A Temperature and Pressure Profile
RVSILT-1 Sorbent

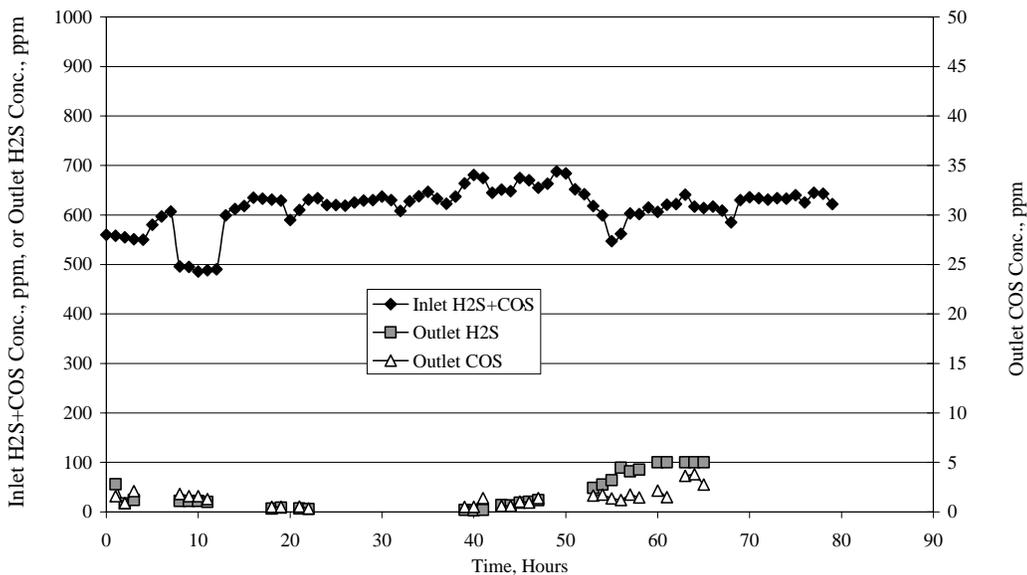


Figure 3.3-3 RX700A Sulfur Profile
RVSILT-1 Sorbent

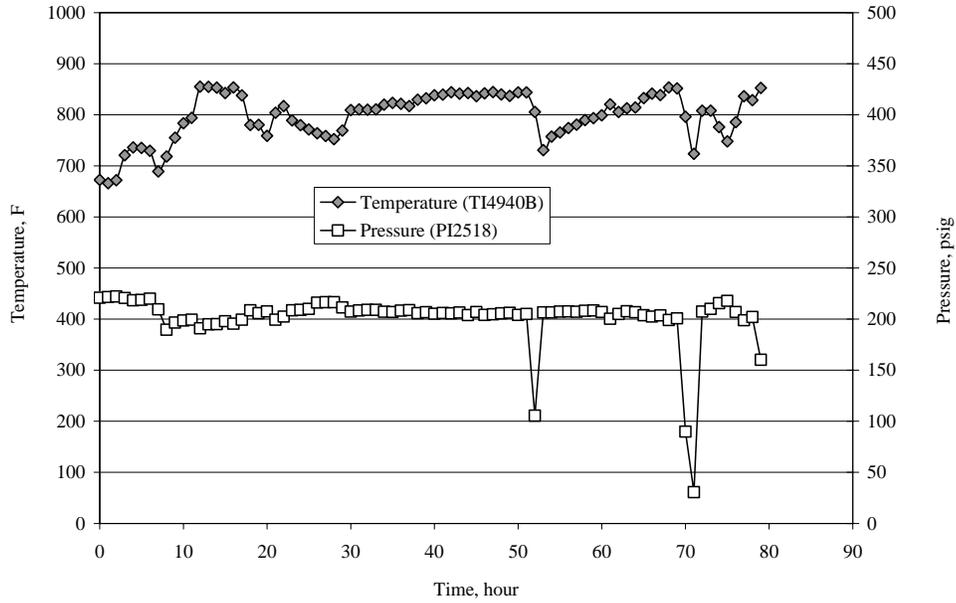


Figure 3.3-4 RX700B Temperature and Pressure Profile
FCC Catalyst

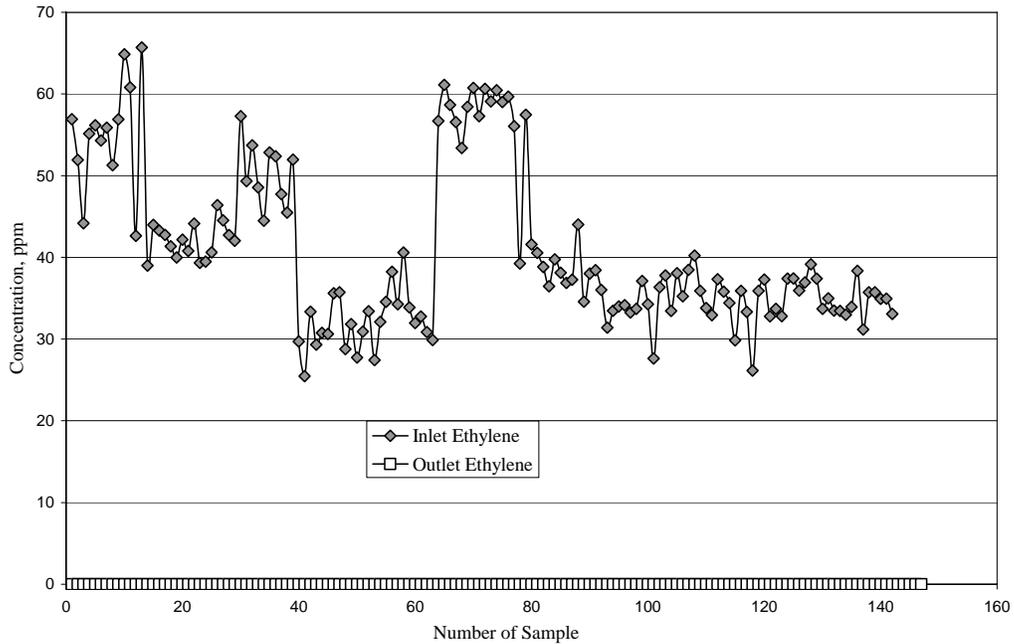


Figure 3.3-5 RX700B Ethylene Profile
FCC Catalyst

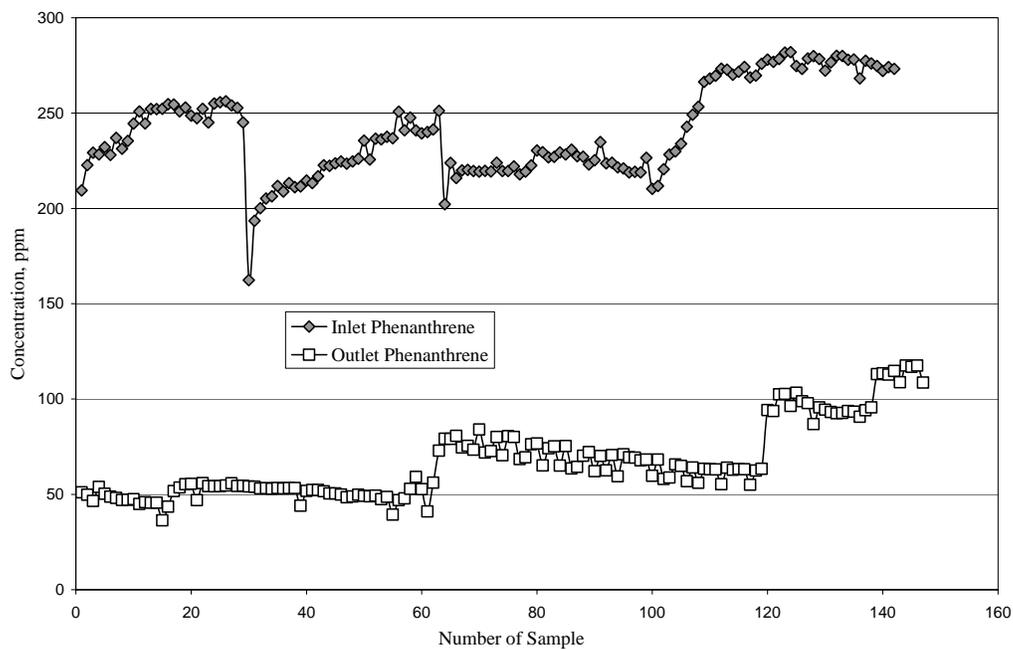


Figure 3.3-6 RX700B Phenanthrene Profile
FCC Catalyst

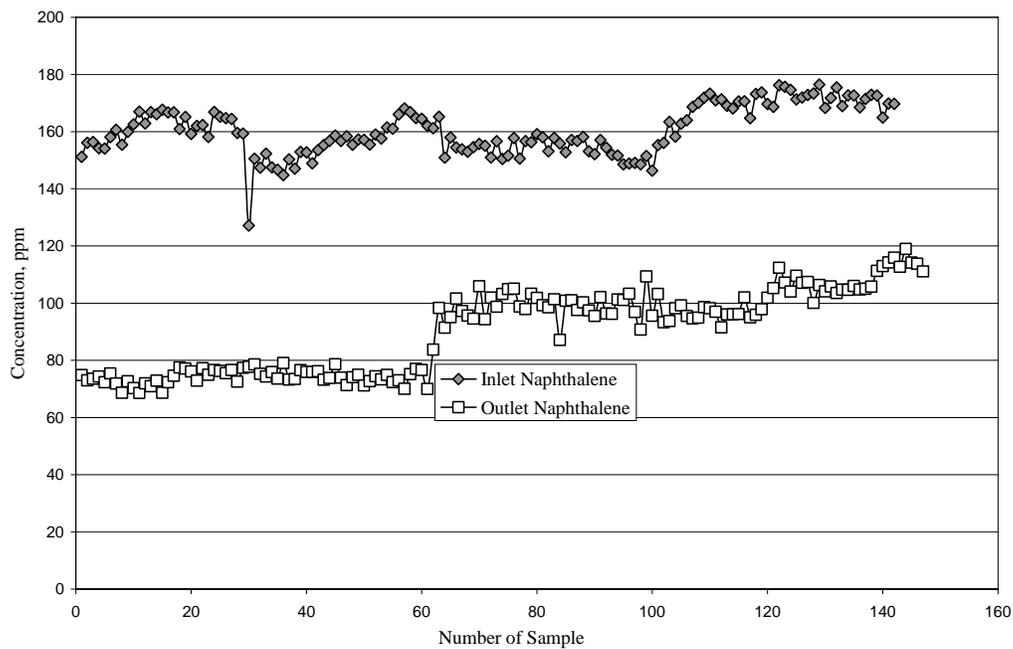


Figure 3.3-7 RX700B Naphthalene Profile
FCC Catalyst

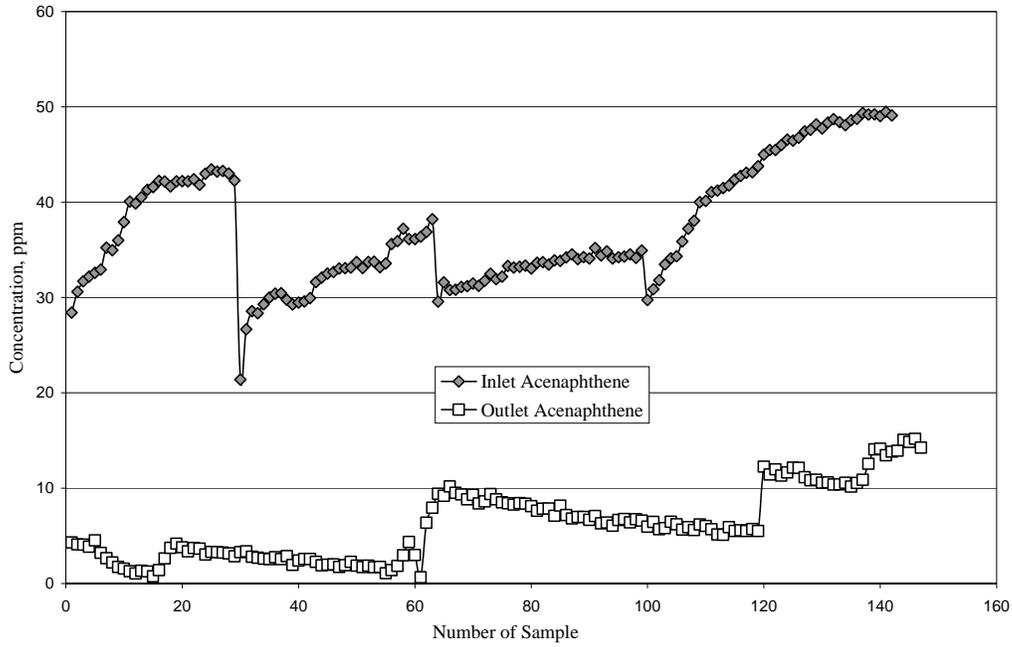


Figure 3.3-8 RX700B Acenaphthene Profile
FCC Catalyst

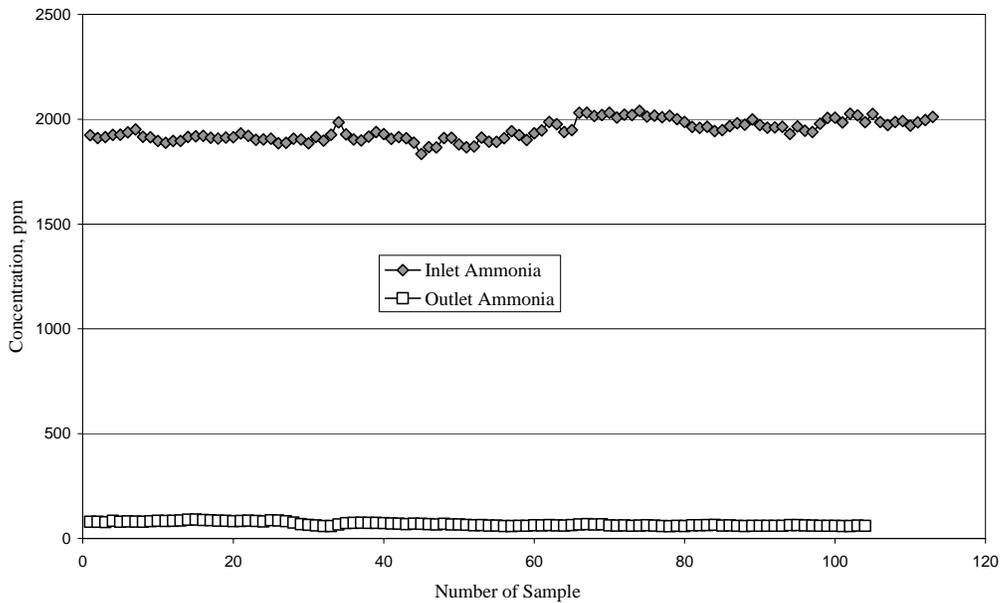


Figure 3.3-9 Mini Reactor Ammonia Profile
G-117RR Catalyst

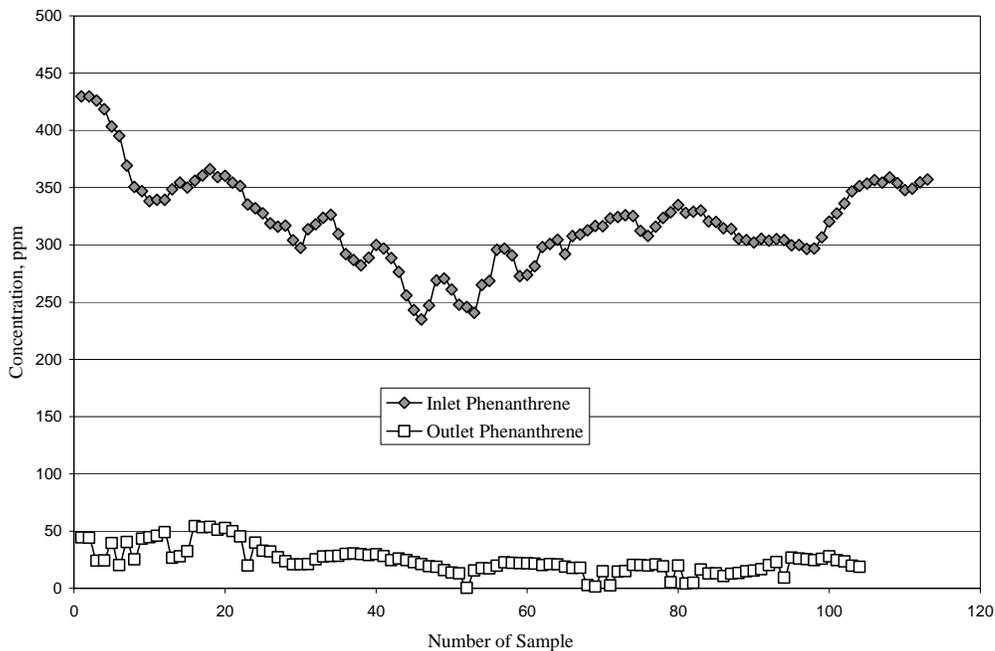


Figure 3.3-10 Mini Reactor Phenanthrene Profile
G-117RR Catalyst

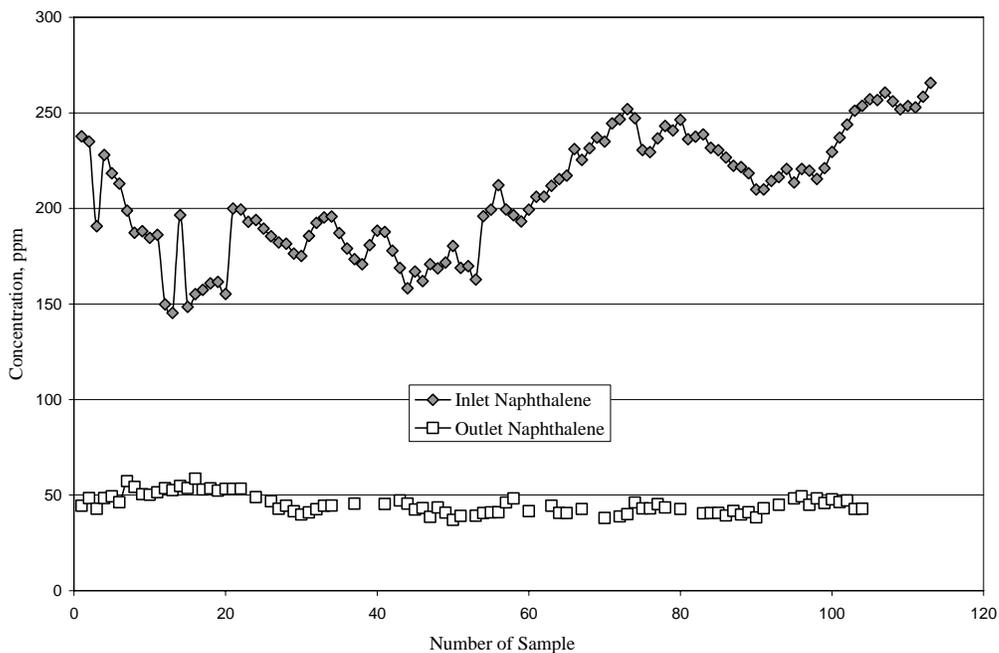


Figure 3.3-11 Mini Reactor naphthalene Profile
G-117RR Catalyst

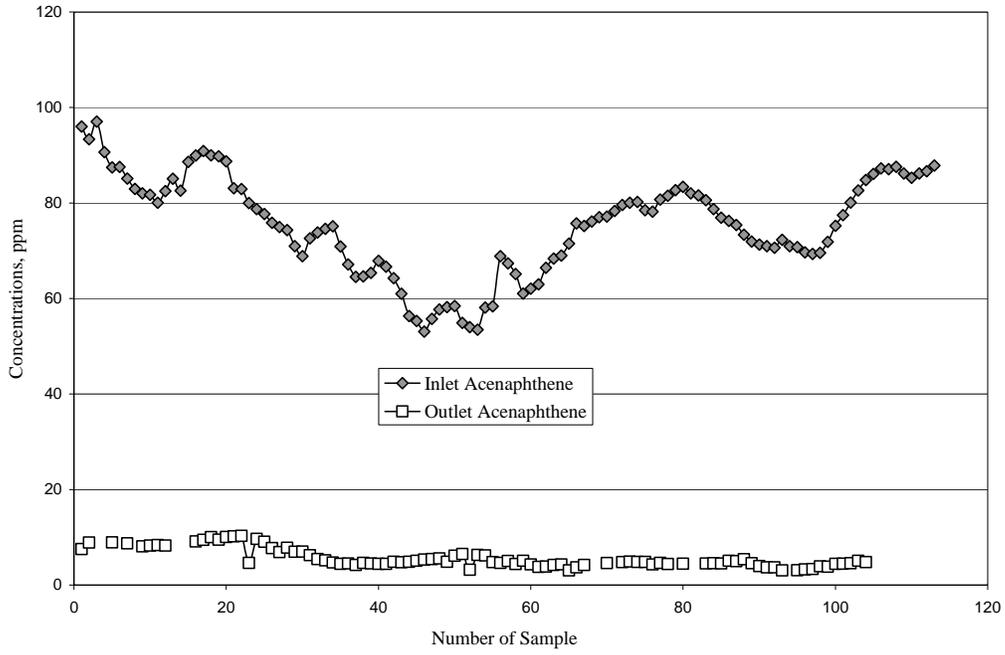


Figure 3.3-12 Mini Reactor Acenaphthene Profile
G-117RR Catalyst

4.0 CONCLUSIONS

TC17 was a successful test run accumulating 313 hours of on-coal operation, 92 of which used Powder River Basin (PRB) subbituminous coal, with the remaining 211 hours using Illinois Basin bituminous coal. Several process enhancements, such as the coal silo modifications, were tested. In addition, the pilot syngas burner operated well on PRB-derived syngas, while the bituminous portion of the test run provided the valuable information on eastern coal operation in the Transport Gasifier. Although several operational issues emerged during the bituminous portion of the test run, including low carbon conversion and high solids carryover to the PCD, the operational experience during TC17 provided valuable insight into further developing the operating envelope of the Transport Gasifier using bituminous fuels.

APPENDIX A1 OPERATION HISTORY

Conversion of the Transport Reactor train to gasification mode of operation was performed from May to September 1999. The first gasification test run, GCT1, was a 233-hour test run to commission the Transport Gasifier and to characterize the limits of operational parameter variations. GCT1 was started on September 9, 1999, with the first part completed on September 15, 1999. The second part of GCT1 was started on December 7, 1999, and completed on December 15, 1999. This test run provided the data necessary for preliminary analysis of gasifier operations and for identification of necessary modifications to improve equipment and process performance. Five different feed combinations of coal and sorbent were tested to gain a better understanding of the gasifier solids collection system efficiency.

GCT2, a 218-hour characterization test run, was started on April 10, 2000, and completed on April 27, 2000. Additional data was taken to analyze the effect of different operating conditions on gasifier performance and operability. A blend of several Powder River Basin (PRB) coals was used with Longview limestone from Alabama. In the outage following GCT2, the Transport Gasifier was modified to improve the operation and performance of the gasifier solids collection system. The most fundamental change was the addition of the loop seal underneath the primary cyclone.

GCT3 was a 184-hour characterization with the primary objective of commissioning the loop seal. A hot solids circulation test (GCT3A) was started on December 1, 2000, and completed December 15, 2000. After a one-month outage to address maintenance issues with the main air compressor, GCT3 was continued. The second part of GCT3 (GCT3B) was started on January 20, 2001, and completed on February 1, 2001. During GCT3B, a blend of several PRB coals was used with Ohio Bucyrus limestone. The loop seal performed well, allowing much higher solids circulation rates and higher syngas heating values. Also, the improved collection efficiency of the cyclone resulted in lower relative solids loading to the PCD and higher carbon conversion.

GCT4, a 242-hour characterization test run, was started on March 7, 2001, and was completed on March 30, 2001. A blend of several PRB coals with Bucyrus limestone from Ohio was used. More experience was gained with the loop seal operations, and additional data was collected to better understand gasifier performance.

TC06, a 1025-hour test campaign, was started on July 4, 2001, and completed on September 24, 2001. A blend of several PRB coals with Ohio Bucyrus limestone was used. Due to its length and stability of operation, the TC06 test run provided valuable data necessary to analyze long term gasifier operations and to identify necessary modifications to improve equipment and process performance, as well as progressing the goal of many thousands of hours of PCD filter element exposure.

TC07, a 442-hour test campaign, was started on December 11, 2001, and completed on April 5, 2002. A blend of several PRB coals and a bituminous coal from the Calumet mine in Alabama were tested with Ohio Bucyrus limestone. Due to operational difficulties with the gasifier stemming from instrumentation problems, the unit was taken offline several times.

TC08 was a 365-hour test campaign to commission the gasifier in oxygen blown mode of operation. TC08 was started on June 9, 2002, and completed on June 29, 2002. A blend of several PRB coals

were tested in air blown, enriched air, and oxygen blown modes of operation. The transition from different modes of operation was smooth, and it was demonstrated that the transition from air to oxygen could be made within 15 minutes.

TC09 was a 309-hour test campaign to characterize the gasifier and PCD operations in air and oxygen blown mode using a bituminous coal from the Sufco mine in Utah. TC09 was started on September 3, 2002, and completed on September 26, 2002.

TC10 was a 416-hour test campaign to conduct long-term tests to evaluate the gasifier and PCD operations in oxygen blown mode of operations using a blend of several PRB coals. TC10 was started on November 16, 2002, and completed on December 18, 2002. Despite problems with the coal mills, coal feeder, pressure tap nozzles and the standpipe, the gasifier did experience short periods of stability during oxygen blown operations. During these periods, the syngas quality was high. During TC10, over 609 tons of Powder River Basin subbituminous coals were gasified.

TC11 was a 192-hour test campaign to conduct short-term tests to evaluate the gasifier and PCD operations in air and oxygen blown mode of operations using Falkirk lignite from North Dakota. TC11 was started on April 7, 2003, and completed on April 18, 2003. During TC11, the lignite proved difficult to feed due to difficulties in the mill operation as a result of the high moisture content in the fuel. However, the gasifier operated well using lignite, with high circulation rates, riser densities and stable temperature profiles. Consequently, the temperature distribution in both the mixing zone and the riser was more uniform than in any previous test run, varying less than 10°F throughout the gasifier.

TC12 was a 733-hour test campaign to conduct short-term tests to evaluate the gasifier and PCD operations in air and oxygen blown mode using a blend of several PRB coals. TC12 was started on May 16, 2003, and completed on July 14, 2003. A primary focus for TC12 was the commissioning of a new gas cleanup system and operating a fuel cell on syngas derived from the Transport Gasifier. The fuel cell system and gas cleanup system both performed well during the testing.

TC13 was a 501-hour test campaign to conduct short-term tests to evaluate gasifier, PSB, and PCD operations in air blown mode of operations using a blend of several PRB coals as well as to conduct short-term tests to evaluate gasifier and PCD operations using two different types of lignite from the Freedom Mine in North Dakota. One type of lignite had a high ash sodium content, while the other types had a low ash sodium content. TC13 was started on September 30, 2003, and completed on November 2, 2003. The syngas-to-PSB testing lasted for a total of about six hours. While successful, the hydraulic system on the turbine cranking motor failed and prevented further PSB testing. The low sodium lignite testing went well, but lowering the gasifier temperature to below 1500°F was necessary to prevent ash agglomeration with the high sodium lignite.

TC14 was a 214-hour test campaign to conduct short-term tests to evaluate the gasifier, PSB, and PCD operations in air and oxygen blown mode of operations using a blend of several PRB coals. TC14 began on February 16, 2004, and ended on February 28, 2004. The syngas-to-PSB testing lasted for a total of about 17 hours at syngas flow rates up to 17,000 pph, contributing about 82% of the total energy to the PSB. The Continuous Fine Ash Depressurization unit was commissioned during TC14. The new system worked well and operated for 190 hours.

TC15 was a 200-hour test campaign to conduct short-term tests to evaluate the gasifier, PSB and PCD operations in air and oxygen blown modes of operations using a blend of several PRB coals. TC15 began on April 19, 2004, and ended on April 29, 2004. The syngas-to-PSB testing lasted for approximately 15 hours at syngas flow rates up to 17,000 pph, contributing about 86% of the total energy to the PSB. The gasifier experienced stable operations in air blown mode but less stable operations in oxygen blown mode due to poor solids circulation. A primary focus of TC15 was to commission and test modifications made to the syngas cleanup system, which proved effective in reducing the syngas sulfur content.

TC16 was an 835-hour test campaign to conduct short-term tests to evaluate gasifier, PSB, and PCD operations in air and oxygen blown modes of operations using a blend of several PRB coals as well as to conduct short-term tests to evaluate gasifier and PCD operations using lignite coal from the Freedom Mine in North Dakota. TC16 began on July 14, 2004, and ended on August 24, 2004. The syngas-to-PSB testing lasted for approximately seven hours at syngas flow rates up to 13,000 pph and the combustion turbine (CT) operated for about 20 hours. A Delphi solid oxide fuel cell operated on syngas for 118 hours during TC16. The first fuel cell stack ran for 28 hours, during which time the performance declined significantly. Another fuel cell stack was installed and fuel cell performance only degraded slightly during the first eight hours of testing, then remained steady for 82 hours. The new steam/oxygen eductor operated very well, blending the steam and oxygen and allowing oxygen addition at higher gasifier pressures.

TC17, the subject of this report, was a 313-hour test campaign to conduct short-term tests to evaluate gasifier, PSB, and PCD operations in air blown mode of operation using a blend of several PRB coals and bituminous coal from the Illinois Basin. TC17 began on October 25, 2004, and ended on November 18, 2004. The PSB testing lasted for approximately 6 ½ hours at syngas flow rates up to 13,000 pph.

Figure A1-1 gives a summary of operating test hours achieved with the Transport Reactor at the PSDF.

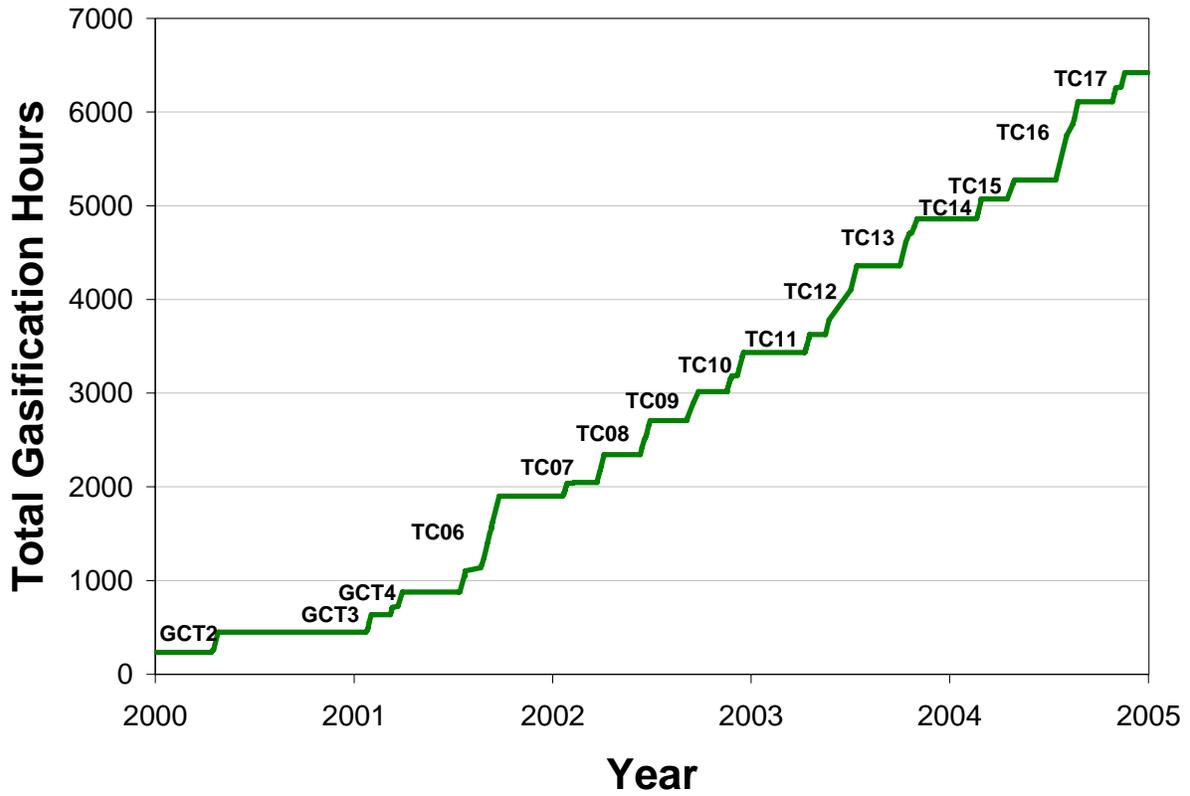


Figure A1-1 Operating Hours Summary for the Transport Gasifier Train

APPENDIX A2 EQUIPMENT LIST

Major Equipment in the Transport Gasifier Train

TAG NAME	DESCRIPTION
BR0201	Reactor Start-Up Burner
BR0401	Atmospheric Syngas Combustor (Thermal Oxidizer)
BR0452	Piloted Syngas Burner
BR0602	Atmospheric Fluidized Bed Combustor (AFBC) Start-Up Burner
CO0201	Main Air Compressor
CO0451	Turbine Air Compressor
CO0601	AFBC Air Compressor
CY0201	Primary Cyclone in the Reactor Loop
CY0207	Disengager in the Reactor Loop
CY0601	AFBC Cyclone
DR0402	Steam Drum
DY0201	Feeder System Air Dryer
FD0206	Spent Solids Screw Cooler
FD0210	Coal Feeder System
FD0220	Sorbent Feeder System
FD0502	Fines Screw Cooler
FD0510	Spent Solids Transporter System
FD0520	Fines Transporter System
FD0530	Spent Solids Feeder System
FD0602	AFBC Solids Screw Cooler
FD0610	AFBC Sorbent Feeder System
FL0301	Particulate Control Device
FL0401	Compressor Intake Filter
GN0451	Turbine Generator
GT0451	Gas Turbine
HX0202	Primary Gas Cooler
HX0204	Transport Air Cooler
HX0402	Secondary Gas Cooler
HX0405	Compressor Feed Cooler
HX0540	CFAD Collection Drum/Heat Exchanger
HX0601	AFBC Heat Recovery Exchanger
RX0201	Transport Reactor
SI0602	Spent Solids Silo
SU0601	AFBC

Major Equipment in the Balance of Plant (Page 1 of 3)

TAG NAME	DESCRIPTION
B02920	Auxiliary Boiler
B02921	Auxiliary Boiler – Superheater
CL2100	Cooling Tower
C02201A-D	Service Air Compressor A-D
C02202	Air-Cooled Service Air Compressor
C02203	High-Pressure Air Compressor
C02601A-C	Reciprocating N ₂ Compressor A-C
CRO104	Coal and Sorbent Crusher
CV0100	Crushed Feed Conveyor
CV0101	Crushed Material Conveyor
DP2301	Baghouse Bypass Damper
DP2303	Inlet Damper on Dilution Air Blower
DP2304	Outlet Damper on Dilution Air Blower
DY2201A-D	Service Air Dryer A-D
DY2202	Air-Cooled Service Air Compressor Air Dryer
DY2203	High-Pressure Air Compressor Air Dryer
FD0104	MWK Coal Transport System
FD0105	FW Coal Transport System
FD0111	MWK Coal Mill Feeder
FD0112	FW Coal Mill Feeder
FD0113	Sorbent Mill Feeder
FD0140	Coke Breeze and Bed Material Transport System
FD0154	MWK Limestone Transport System
FD0810	Ash Unloading System
FD0820	Baghouse Ash Transport System
FL0700	Baghouse
FN0700	Dilution Air Blower
H00100	Reclaim Hopper
H00105	Crushed Material Surge Hopper
H00252	Coal Surge Hopper
H00253	Sorbent Surge Hopper
HT2101	MWK Equipment Cooling Water Head Tank
HT2103	SCS Equipment Cooling Water Head Tank
HT0399	60-Ton Bridge Crane
HX2002	MWK Steam Condenser
HX2003	MWK Feed Water Heater

Major Equipment in the Balance of Plant (Page 2 of 3)

TAG NAME	DESCRIPTION
HX2004	MWK Subcooler
HX2103A	SCS Cooling Water Heat Exchanger
HX2103B	FW Cooling Water Heat Exchanger
HX2103C	MWK Cooling Water Heat Exchanger
LF0300	Propane Vaporizer
MC3001-3017	MCCs for Various Equipment
ME0700	MWK Stack
ME0814	Dry Ash Unloader for MWK Train
ML0111	MWK Coal Mill
ML0112	FW Coal Mill
ML0113	Sorbent Mill for Both Trains
PG0011	Oxygen Plant
PG2600	Nitrogen Plant
PU2000A-B	MWK Feed Water Pump A-B
PU2100A-B	Raw Water Pump A-B
PU2101A-B	Service Water Pump A-B
PU2102A-B	Cooling Tower Make-Up Pump A-B
PU2103A-D	Circulating Water Pump A-D
PU2107	SCS Cooling Water Make-Up Pump
PU2109A-B	SCS Cooling Water Pump A-B
PU2110A-B	FW Cooling Water Pump A-B
PU2111A-B	MWK Cooling Water Pump A-B
PU2300	Propane Pump
PU2301	Diesel Rolling Stock Pump
PU2302	Diesel Generator Transfer Pump
PU2303	Diesel Tank Sump Pump
PU2400	Fire Protection Jockey Pump
PU2401	Diesel Fire Water Pump #1
PU2402	Diesel Fire Water Pump #2
PU2504A-B	Waste Water Sump Pump A-B
PU2507	Coal and Limestone Storage Sump Pump
PU2700A-B	Demineralizer Forwarding Pump A-B

Major Equipment in the Balance of Plant (Page 3 of 3)

TAG NAME	DESCRIPTION
PU2920A-B	Auxiliary Boiler Feed Water Pump A-B
SB3001	125-V DC Station Battery
SB3002	UPS
SC0700	Baghouse Screw Conveyor
SG3000-3005	4160-V, 480-V Switchgear Buses
SI0101	MWK Crushed Coal Storage Silo
SI0102	FW Crushed Coal Storage Silo
SI0103	Crushed Sorbent Storage Silo
SI0111	MWK Pulverized Coal Storage Silo
SI0112	FW Pulverized Coal Storage Silo
SI0113	MWK Limestone Silo
SI0114	FW Limestone Silo
SI0810	Ash Silo
ST2601	N ₂ Storage Tube Bank
TK2000	MWK Condensate Storage Tank
TK2001	FW Condensate Tank
TK2100	Raw Water Storage Tank
TK2300A-D	Propane Storage Tank A-D
TK2301	Diesel Storage Tank
TK2401	Fire Water Tank
XF3000A	230/4.16-kV Main Power Transformer
XF3001B-5B	4160/480-V Station Service Transformer No. 1-5
XF3001G	480/120-V Miscellaneous Transformer
XF3010G	120/208 Distribution Transformer
XF3012G	UPS Isolation Transformer
VS2203	High-Pressure Air Receiver

APPENDIX A3 MASS AND ENERGY BALANCES

Material balances are useful in checking the accuracy and consistency of data as well as determining periods of operation where the data is suitable for model development and commercial plant design. Total material balances for each operating period of TC17 are given in Figure A3-1 which compare the total mass in and the total mass out. The overall material balance was good, with all of the relative differences at $\pm 10\%$ for PRB and bituminous coal. The relative difference (relative error) is defined as the Transport Gasifier feeds minus the products divided by the feeds ($\{\text{In-Out}\}/\text{In}$).

The main contributors to the material balance are the coal feed rate, air flow rate, steam flow rate, nitrogen flow rate, dolomite feed rate, sand feed rate, syngas flow rate, and gasification ash flow rate. All but two of the PRB operating periods had higher overall mass flow rates than the bituminous operating periods.

The TC17 Transport Gasifier energy balance is shown in Figure A3-2 with standard conditions chosen to be a pressure of 1.0 atmosphere and a temperature of 80°F. As shown in the figure, the TC17 energy balances were mostly within $\pm 10\%$ error with the exception of a few that fell within $\pm 20\%$ error. The energy entering the gasifier consisted of the coal, air, and steam fed to the Transport Gasifier. The nitrogen, dolomite, and sand fed to the gasifier were considered to be at standard conditions (80°F) and, hence, had zero enthalpy. The nitrogen feed actually entered the gasifier at a higher temperature than standard conditions, but compared to the other feed enthalpies, this neglected input energy is insignificant. Since the amount of solids removed from the standpipe was negligible, the energy exiting the gasifier consisted of only the syngas and PCD solids. The analysis used the lower heating value of the coal, the PCD solids, and the syngas.

The energy of the syngas was determined at the Transport Gasifier primary cyclone exit. Since the total syngas flow measurement is located downstream of the PCD, 320 pounds of nitrogen per hour that flowed to the PCD inlet and outlet particulate sampling trains was subtracted from the exit flow rate to determine the actual syngas rate from the cyclone. The sensible enthalpy of the syngas was determined by the overall gas heat capacity from the syngas compositions and by using gas heat capacities information. The syngas and PCD solids energy consists of both latent and sensible heat. The heat loss from the Transport Gasifier was estimated to be 3.5 million Btu/hr.

The TC17 carbon balance is shown in Figure A3-3. All but one of the TC17 operating periods carbon balances were within $\pm 15\%$. The carbon balance gives a measure of how accurate the TC17 carbon conversions are. The most probable sources of error in the carbon balance are the coal feed rate measurement and the syngas flow rate measurement.

The TC17 sulfur balance is shown in Figure A3-4. The PRB sulfur balances were all within $\pm 30\%$, while the Illinois Basin sulfur balances were all within $\pm 25\%$. The sulfur balance gives a measure of the accuracy of the TC17 sulfur captures. The most probable sources of error in the sulfur balance is the calculation of the syngas sulfur. The syngas sulfur is calculated by a combustion calculation based on a flue gas sulfur dioxide measurement rather than directly measured from the syngas. Typically, the higher the coal sulfur concentrations are, the better the sulfur balances are because the sulfur flow in the gasifier are larger and the concentration measurements are more accurate.

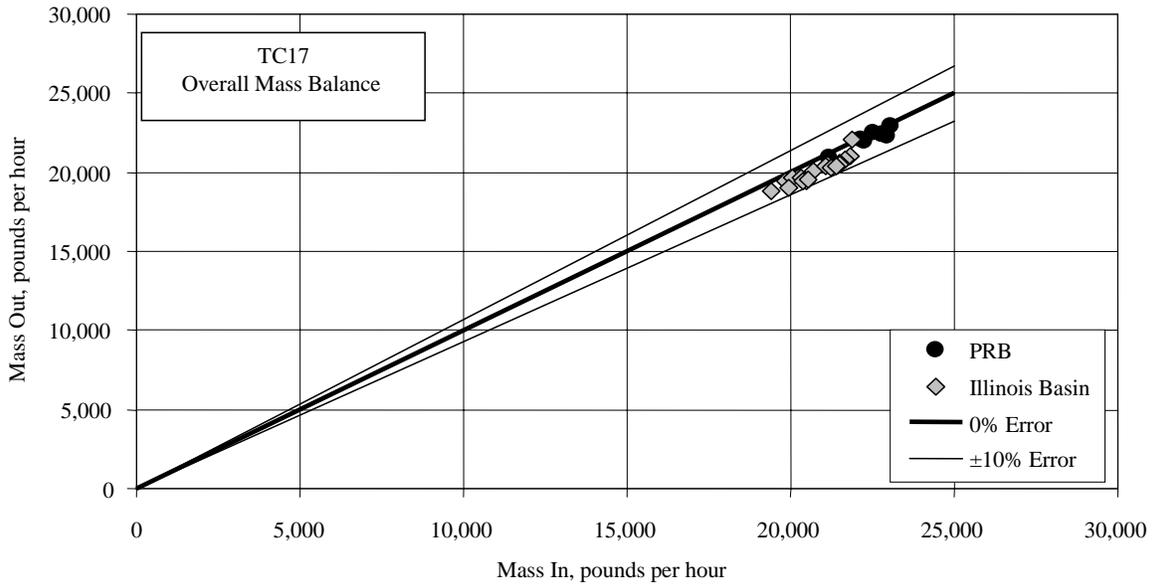


Figure A3-1 Mass Balance

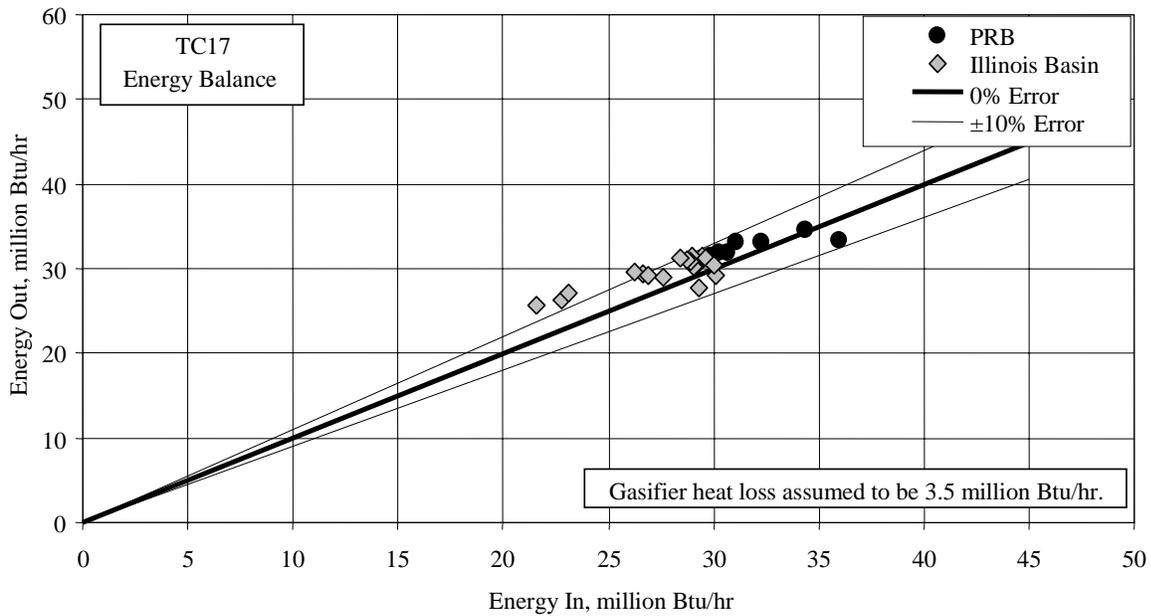


Figure A3-2 Energy Balance

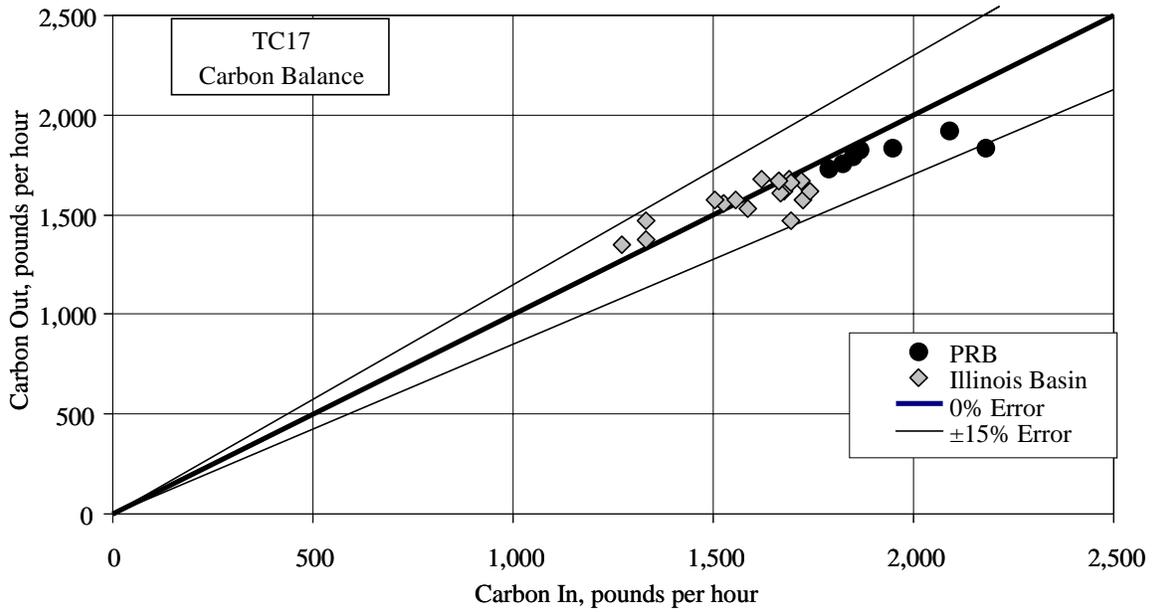


Figure A3-3 Carbon Balance

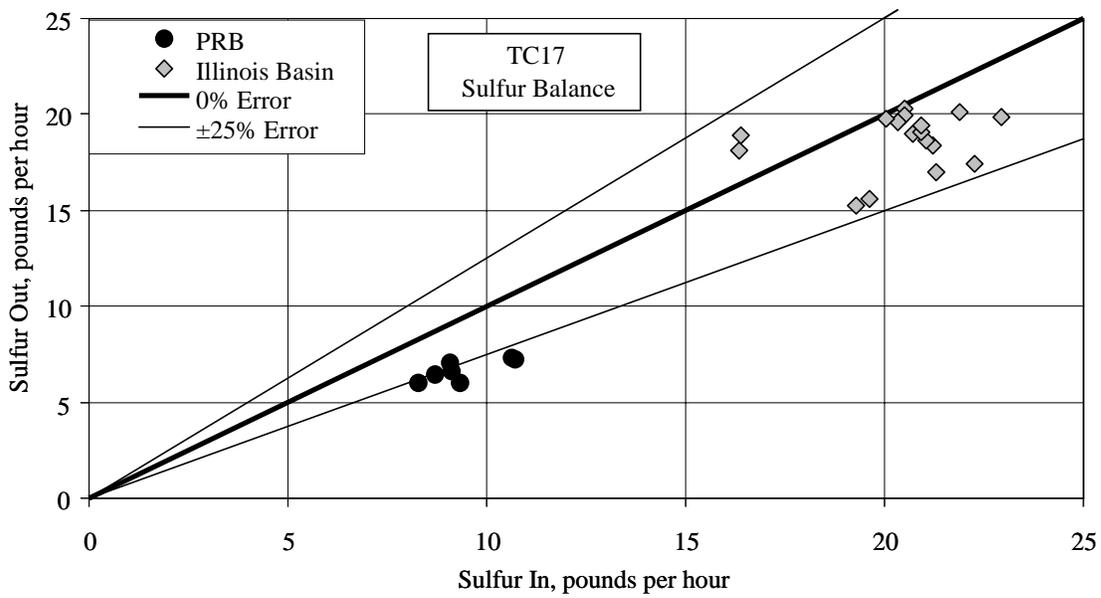


Figure A3-4 Sulfur Balance

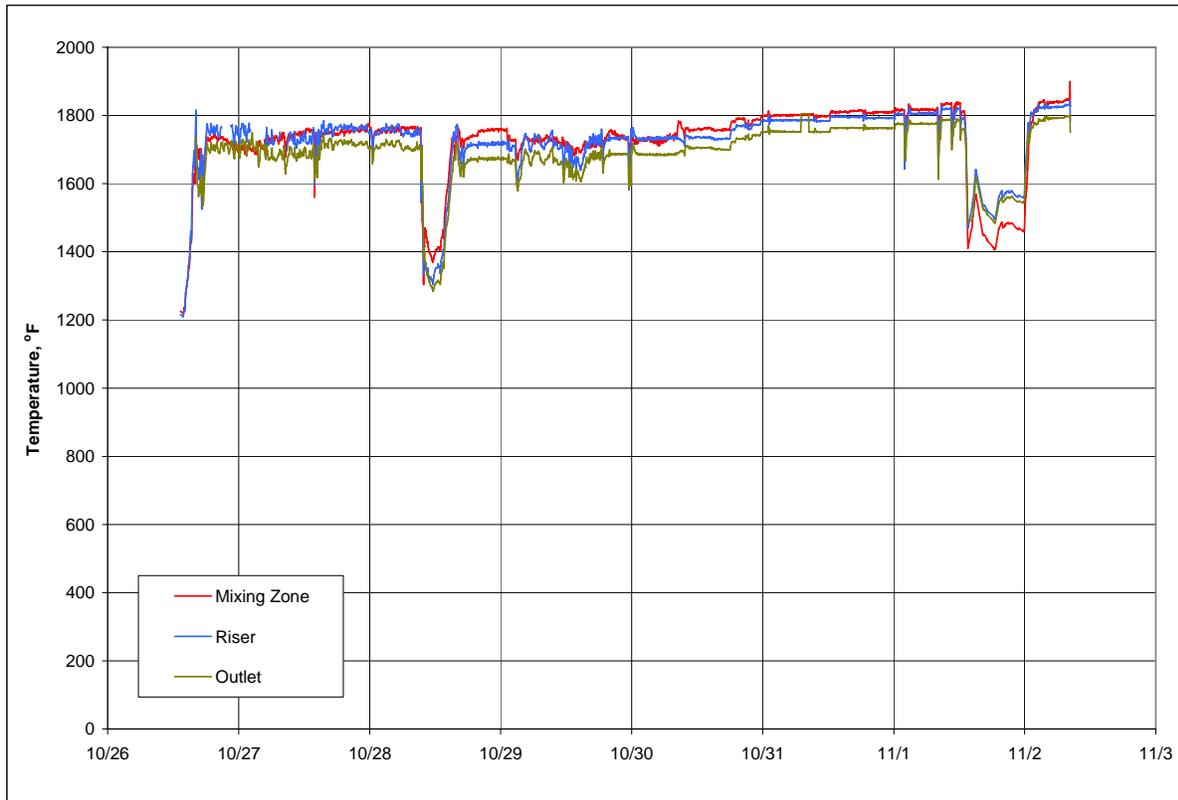


Figure A4-1 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 10/26/04 through 11/3/04

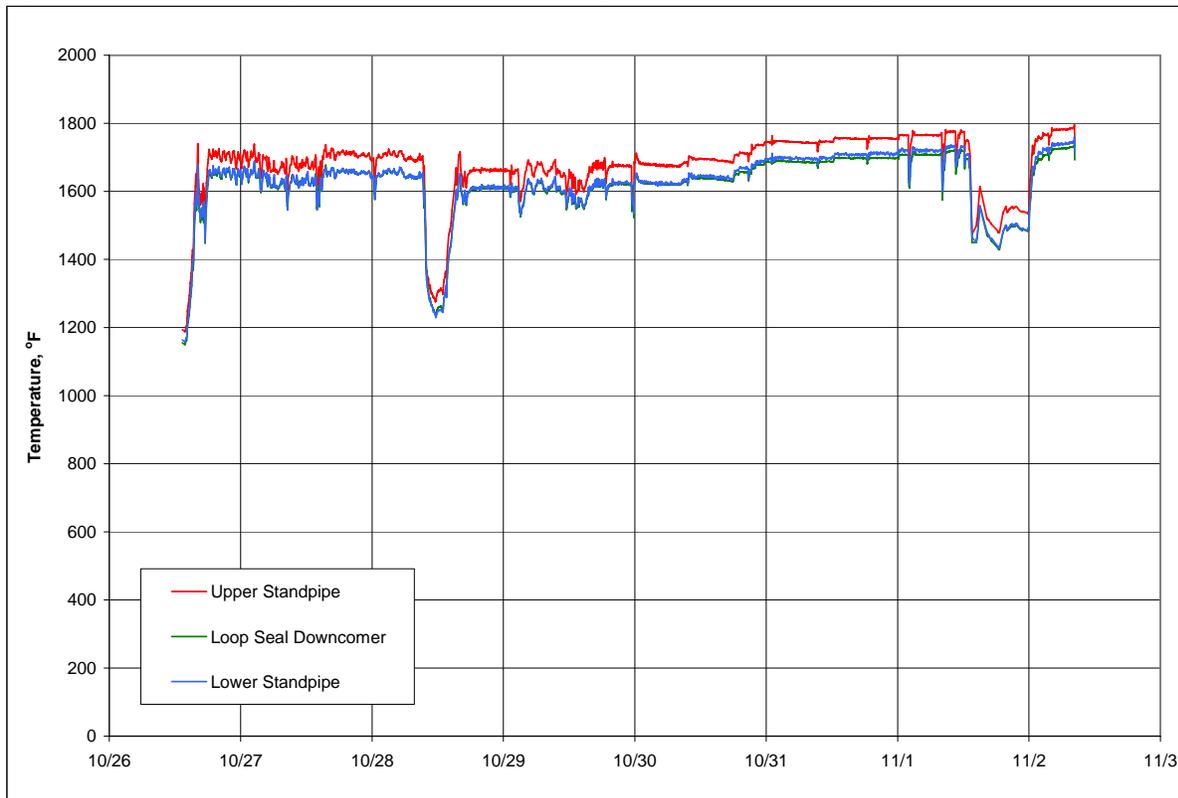


Figure A4-2 Standpipe and Loop Seal Temperatures, 10/26/04 through 11/3/04

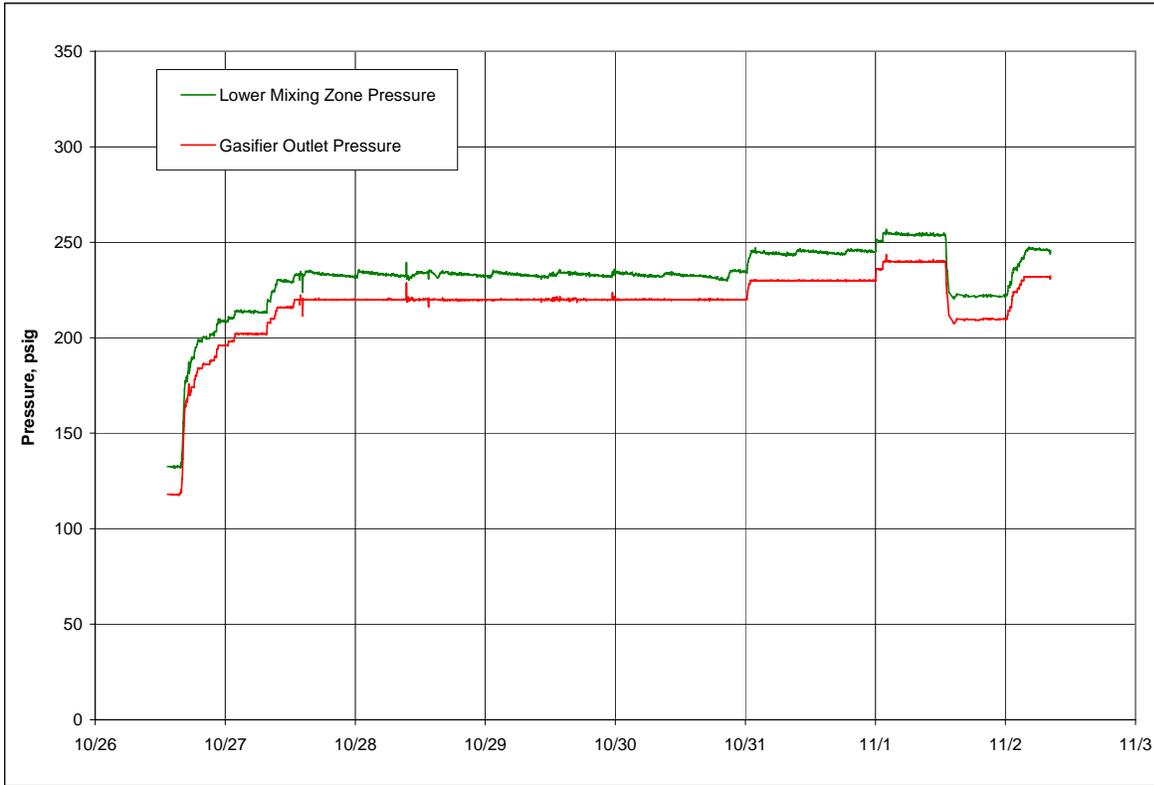


Figure A4-3 Gasifier Pressures, 10/26/04 through 11/3/04

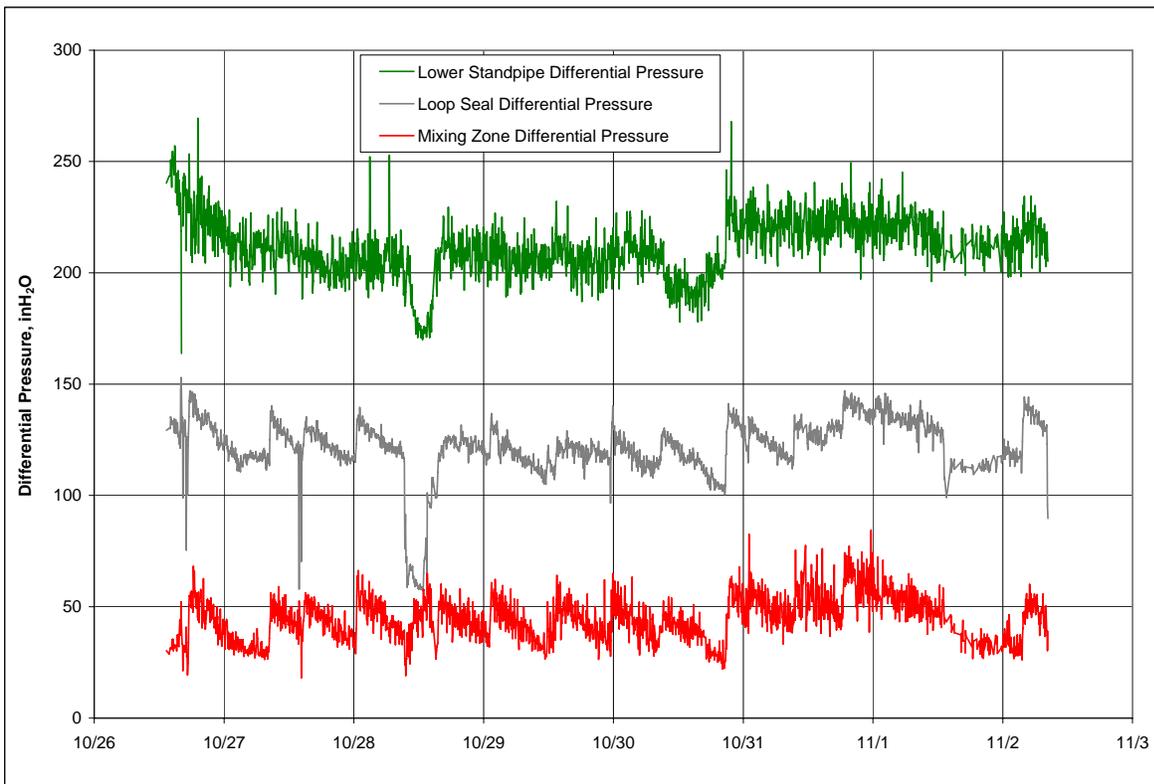


Figure A4-4 Gasifier Differential Pressures, 10/26/04 through 11/3/04

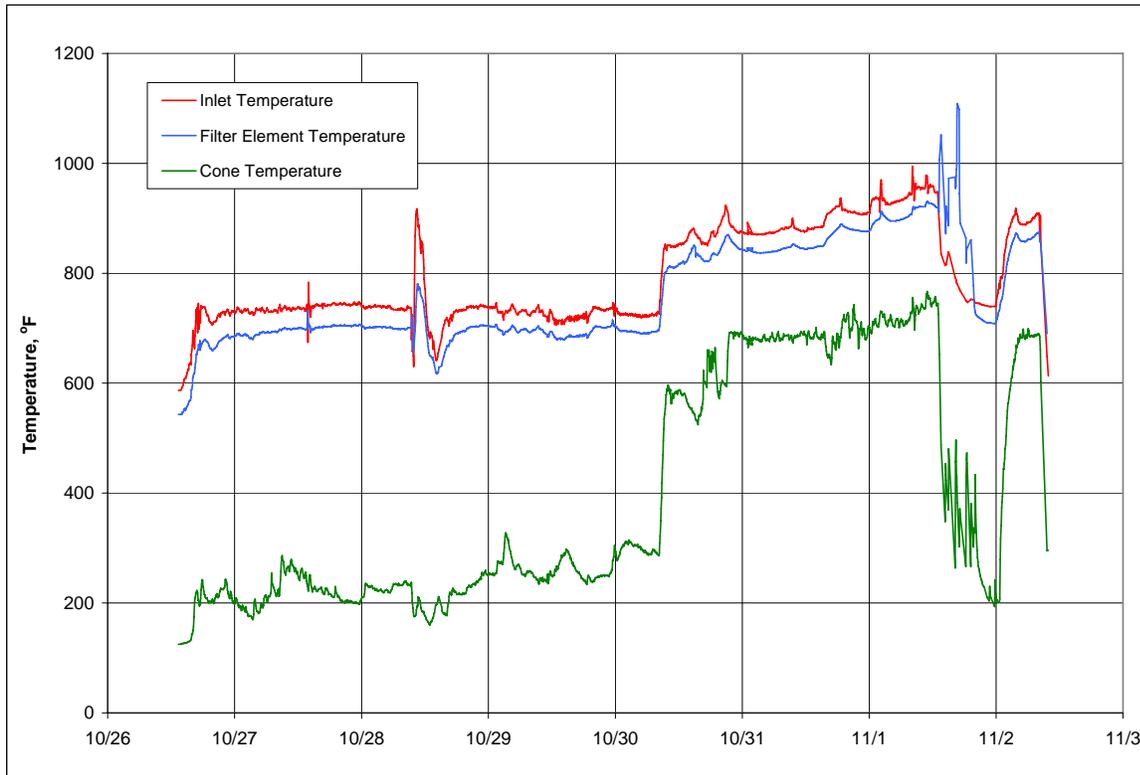


Figure A4-5 PCD Temperatures, 10/26/04 through 11/3/04

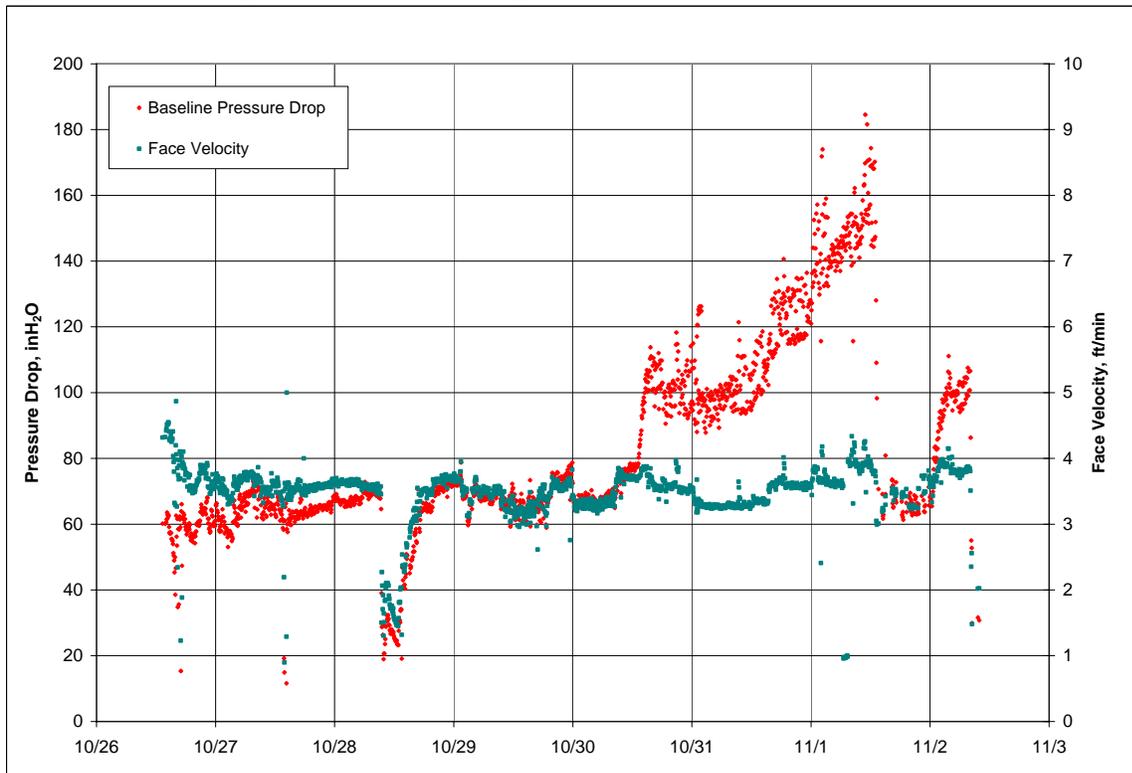


Figure A4-6 PCD Baseline Pressure Drop and Face Velocity, 10/26/04 through 11/3/04

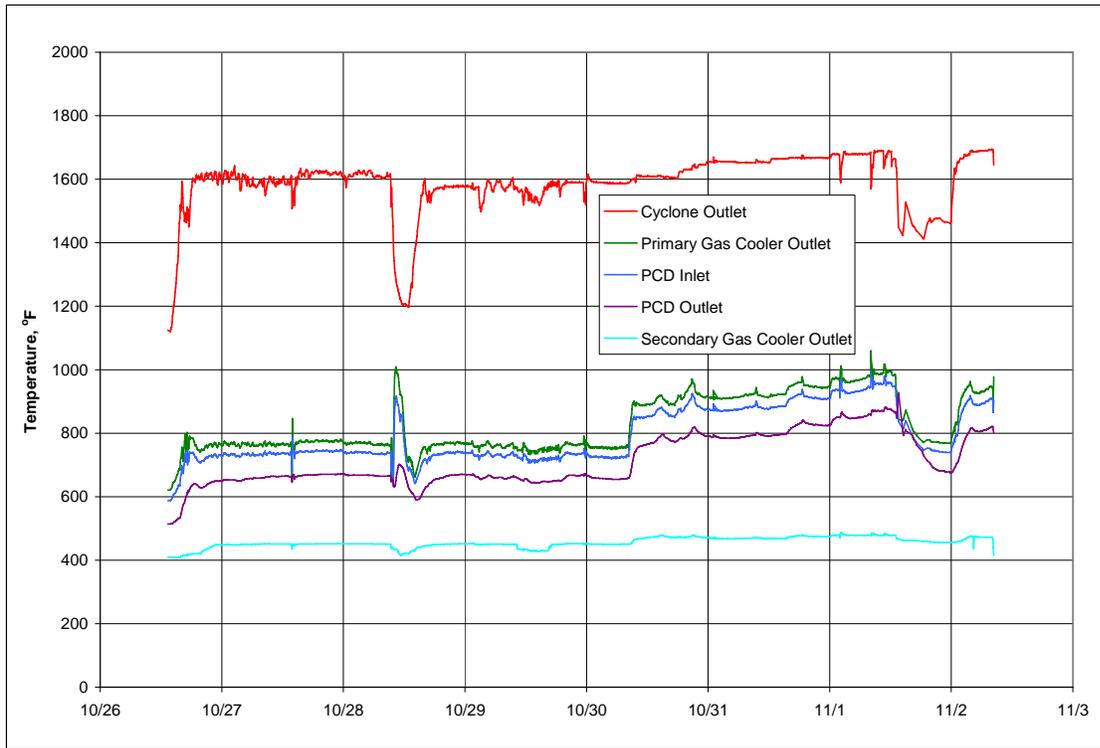


Figure A4-7 System Temperature Profile, 10/26/04 through 11/3/04

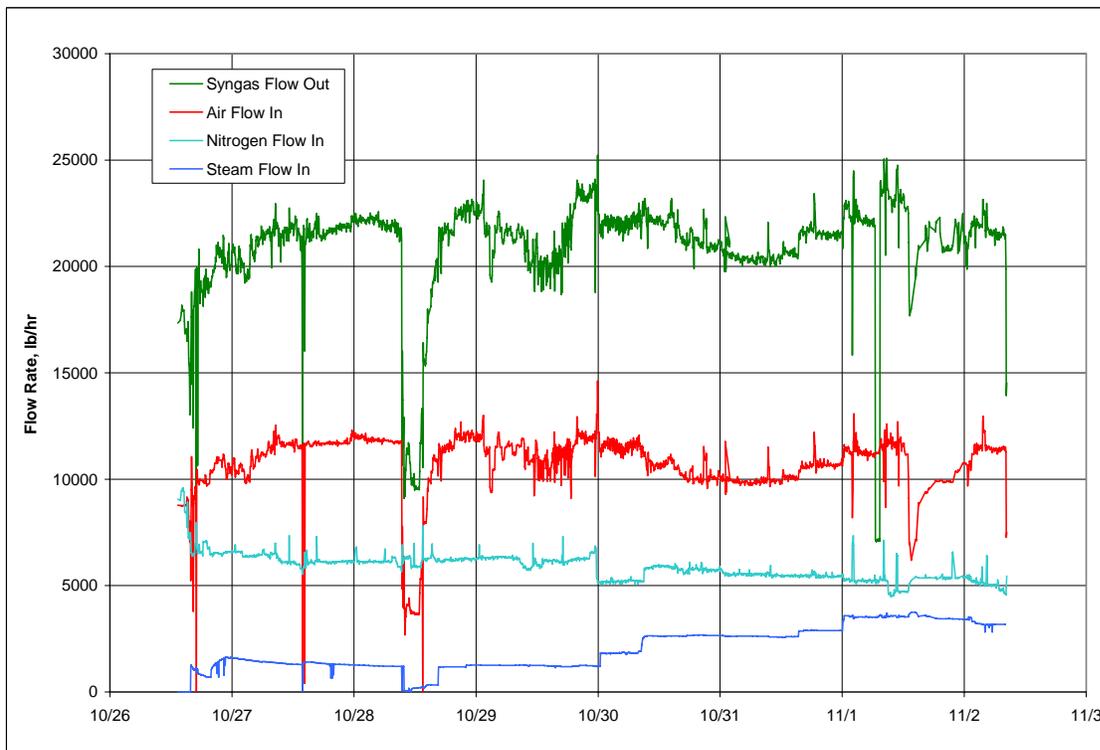


Figure A4-8 System Gas Flows, 10/26/04 through 11/3/04

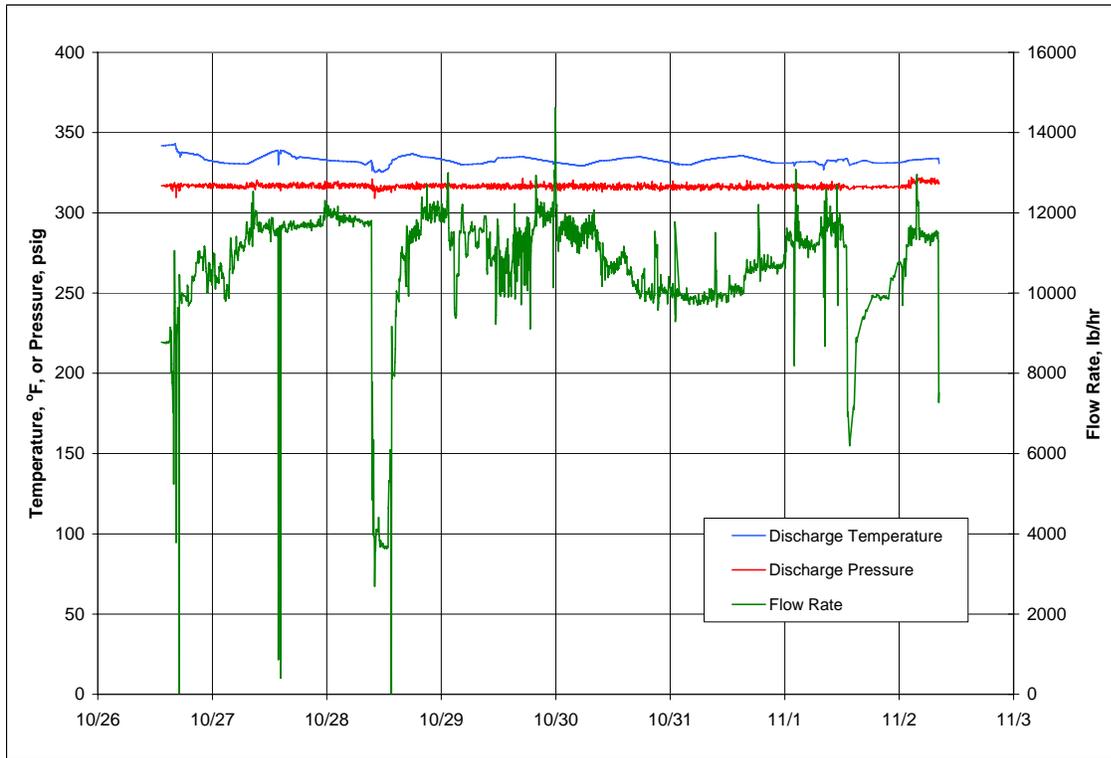


Figure A4-9 Main Air Compressor Operation, 10/26/04 through 11/3/04

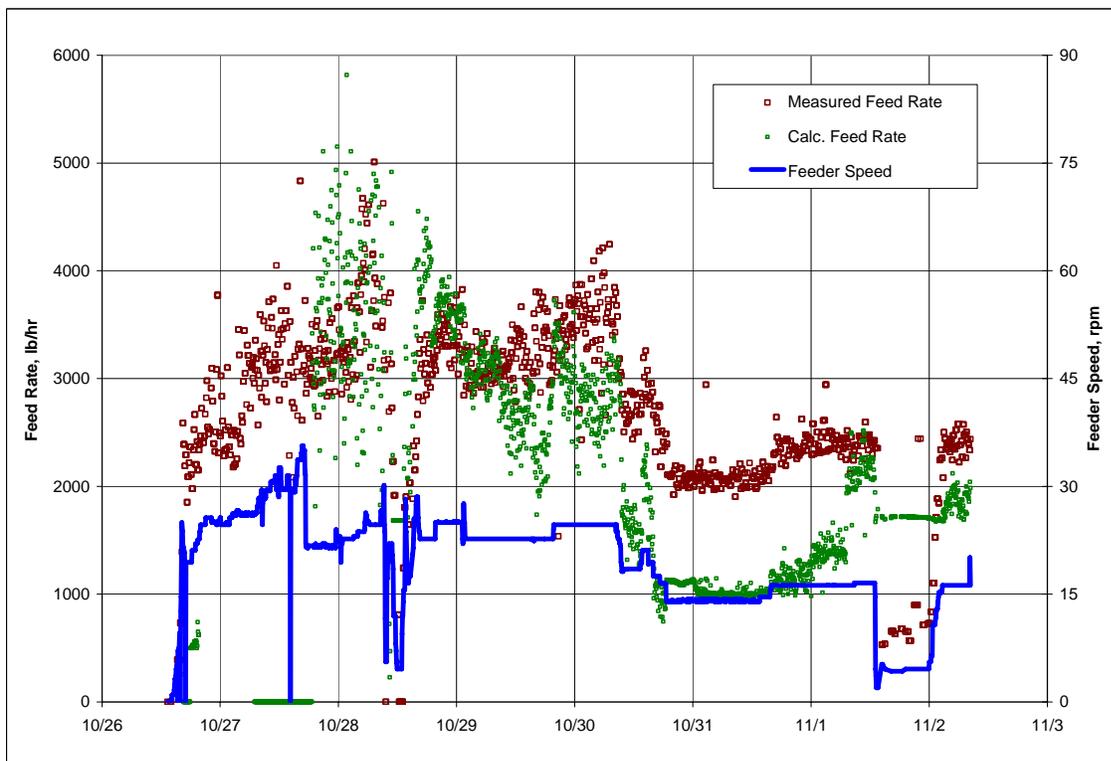


Figure A4-10 Original Coal Feeder Operation, 10/26/04 through 11/3/04

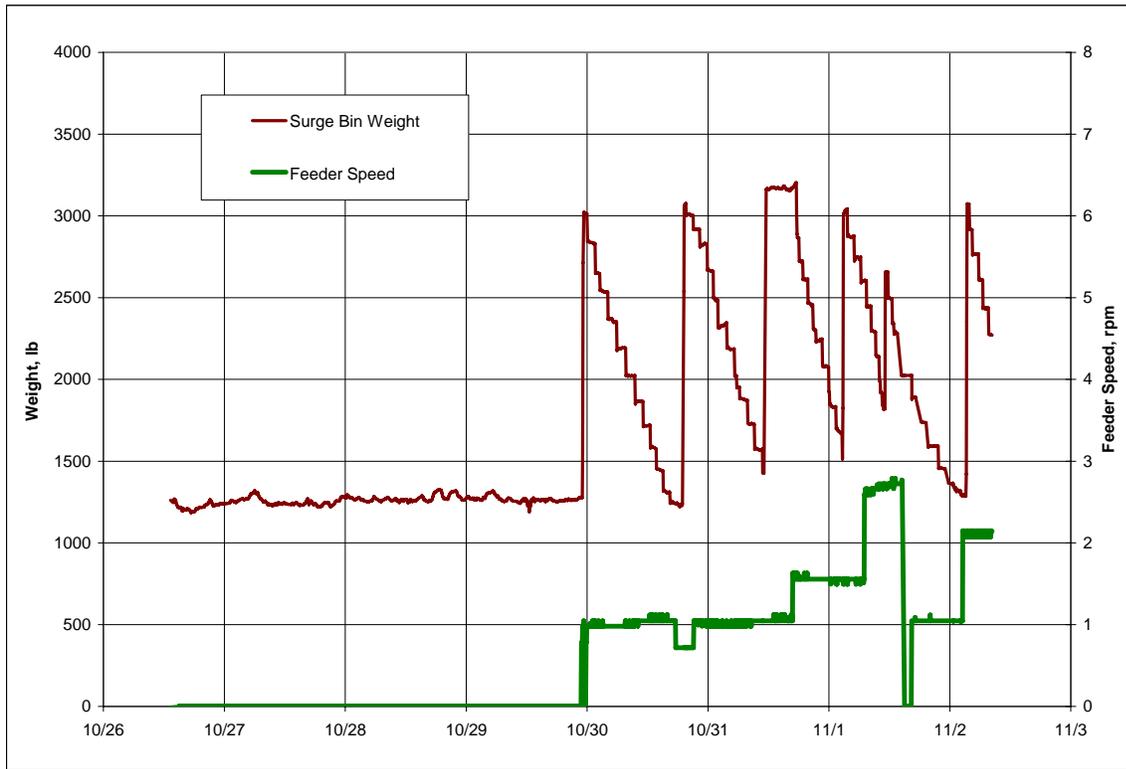


Figure A4-11 Sorbent Feeder Operation, 10/26/04 through 11/3/04

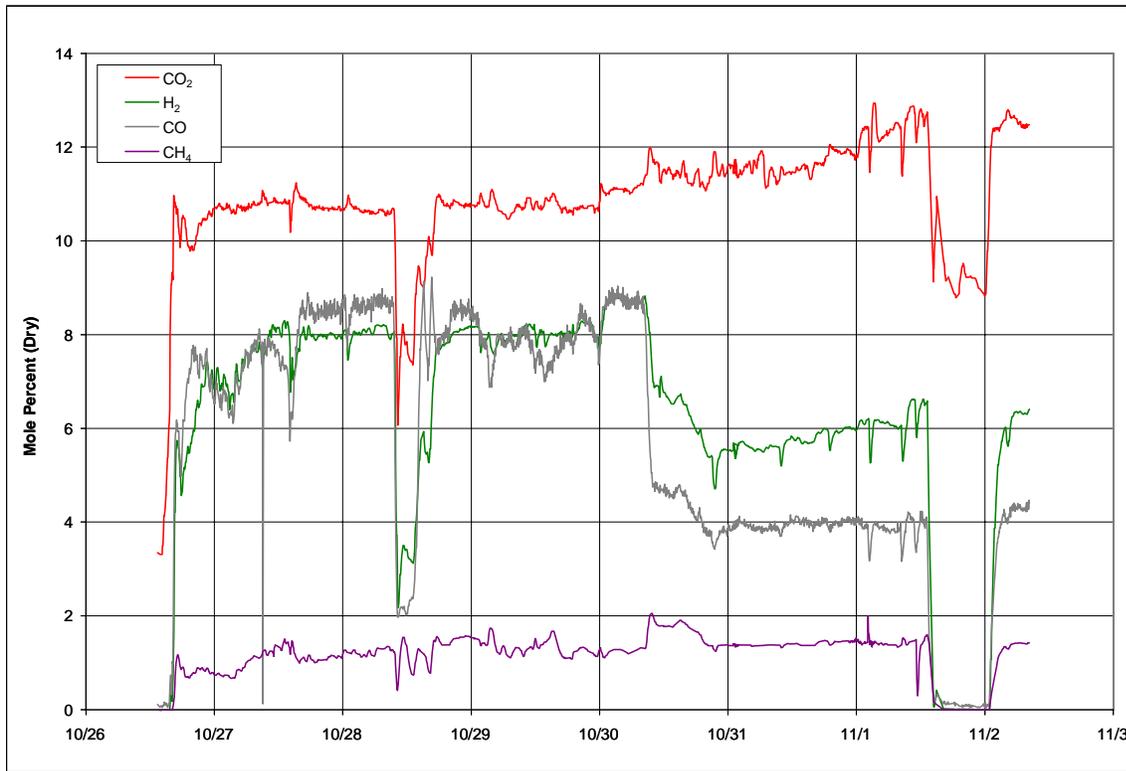


Figure A4-12 Syngas Analyzers, 10/26/04 through 11/3/04

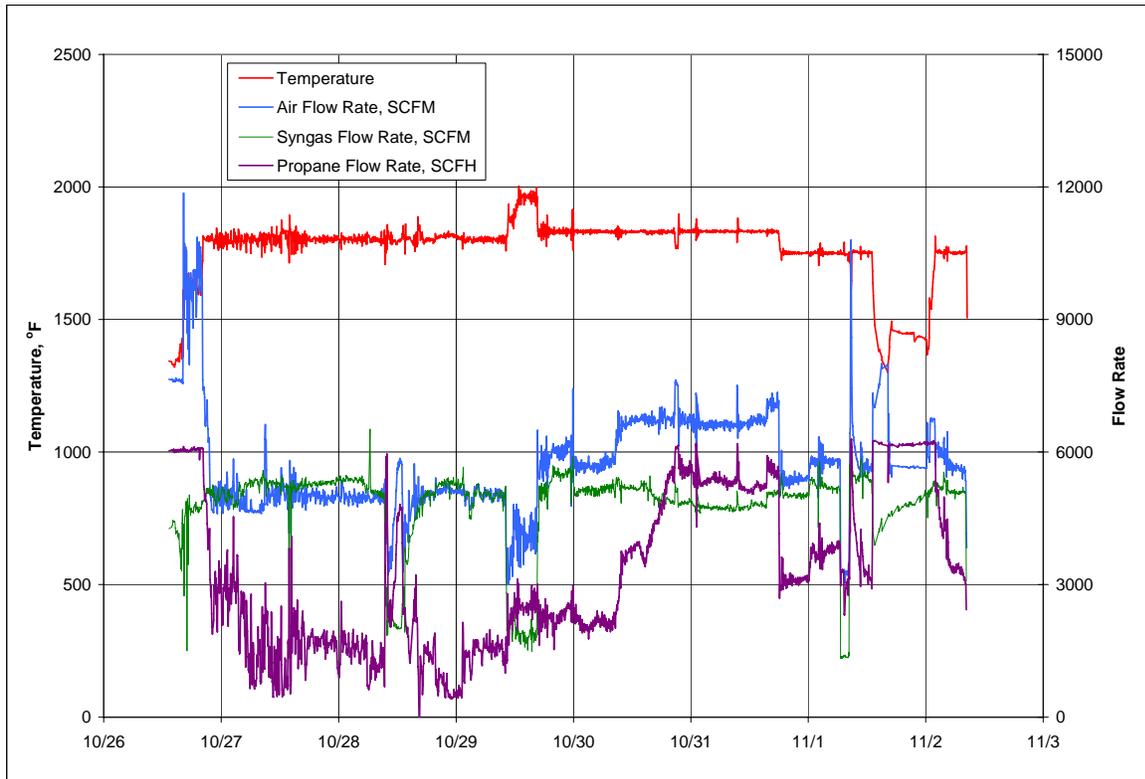


Figure A4-13 Atmospheric Syngas Combustor Operation, 10/26/04 through 11/3/04

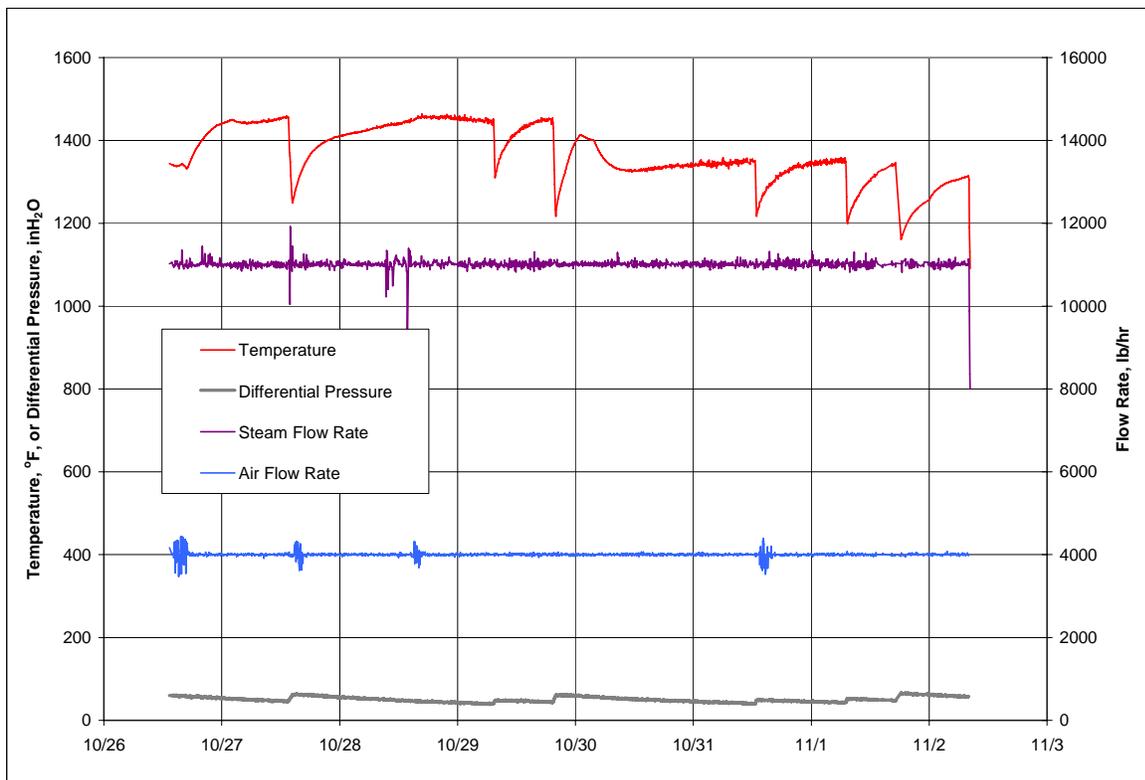


Figure A4-14 Fluidized Bed Combustor Operation, 10/26/04 through 11/3/04

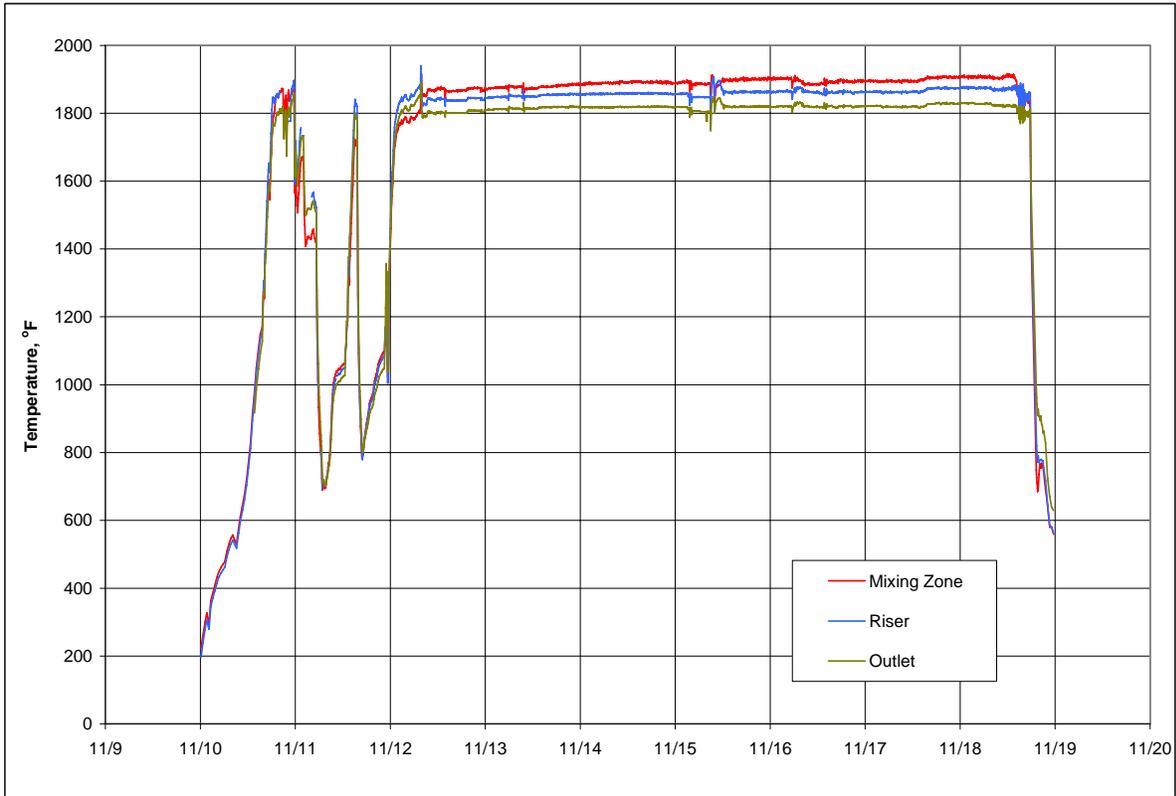


Figure A4-15 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 11/10/04 through 11/19/04

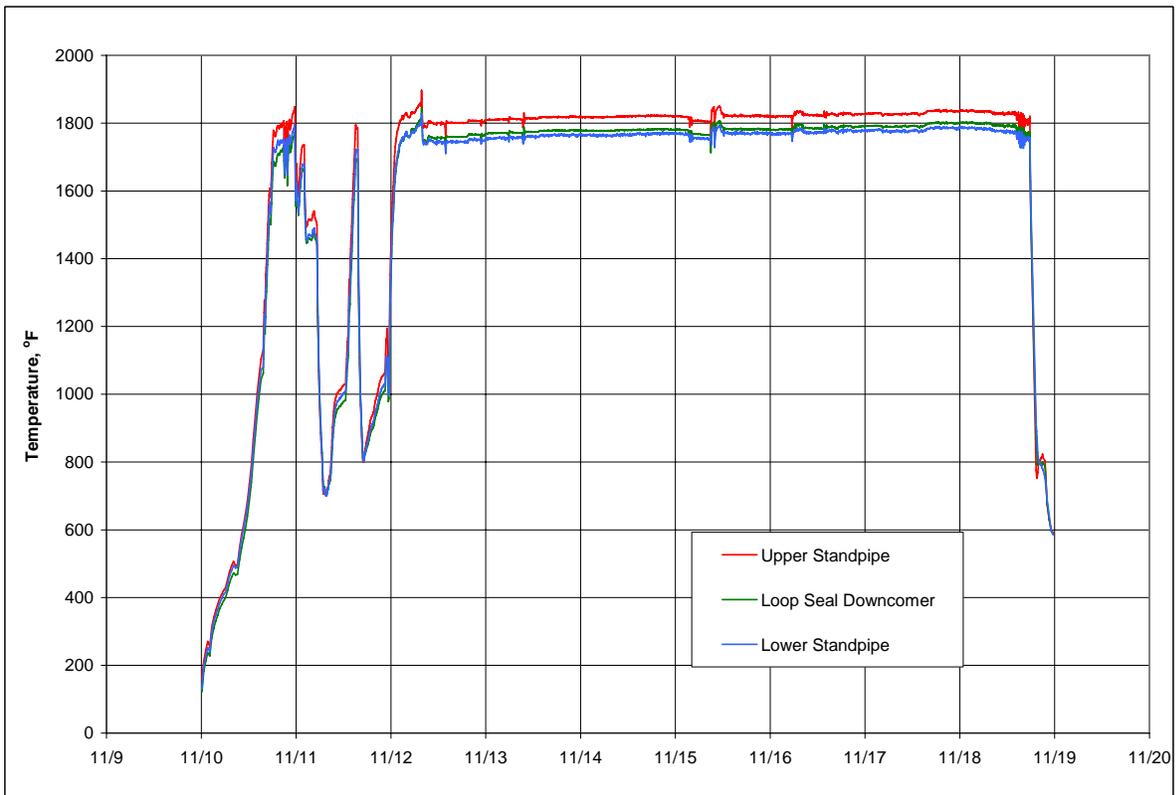


Figure A4-16 Standpipe and Loop Seal Temperatures, 11/10/04 through 11/19/04

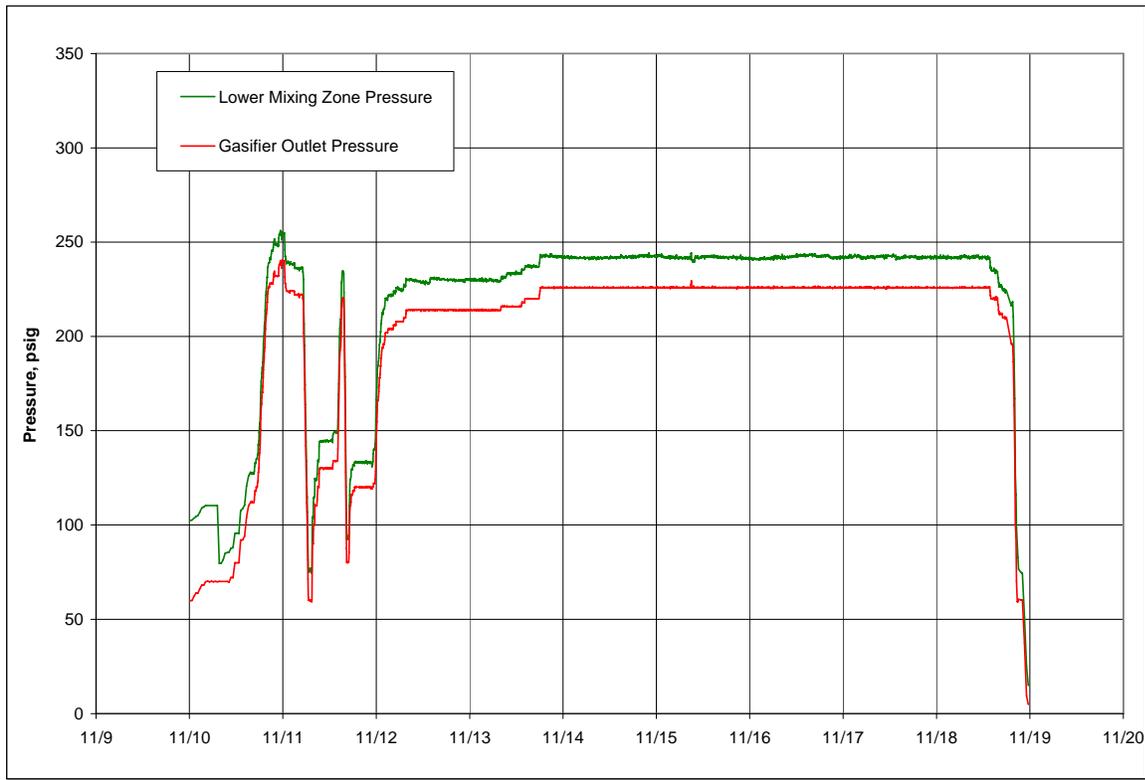


Figure A4-17 Gasifier Pressures, 11/10/04 through 11/19/04

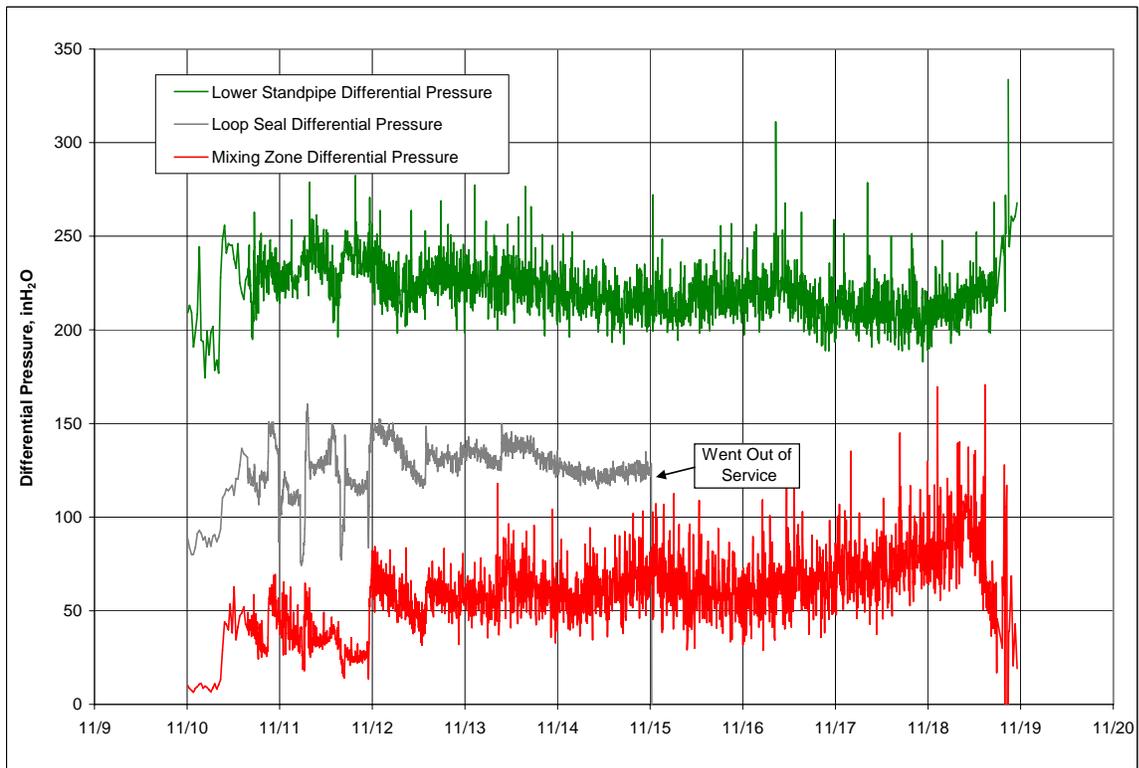


Figure A4-18 Gasifier Differential Pressures, 11/10/04 through 11/19/04

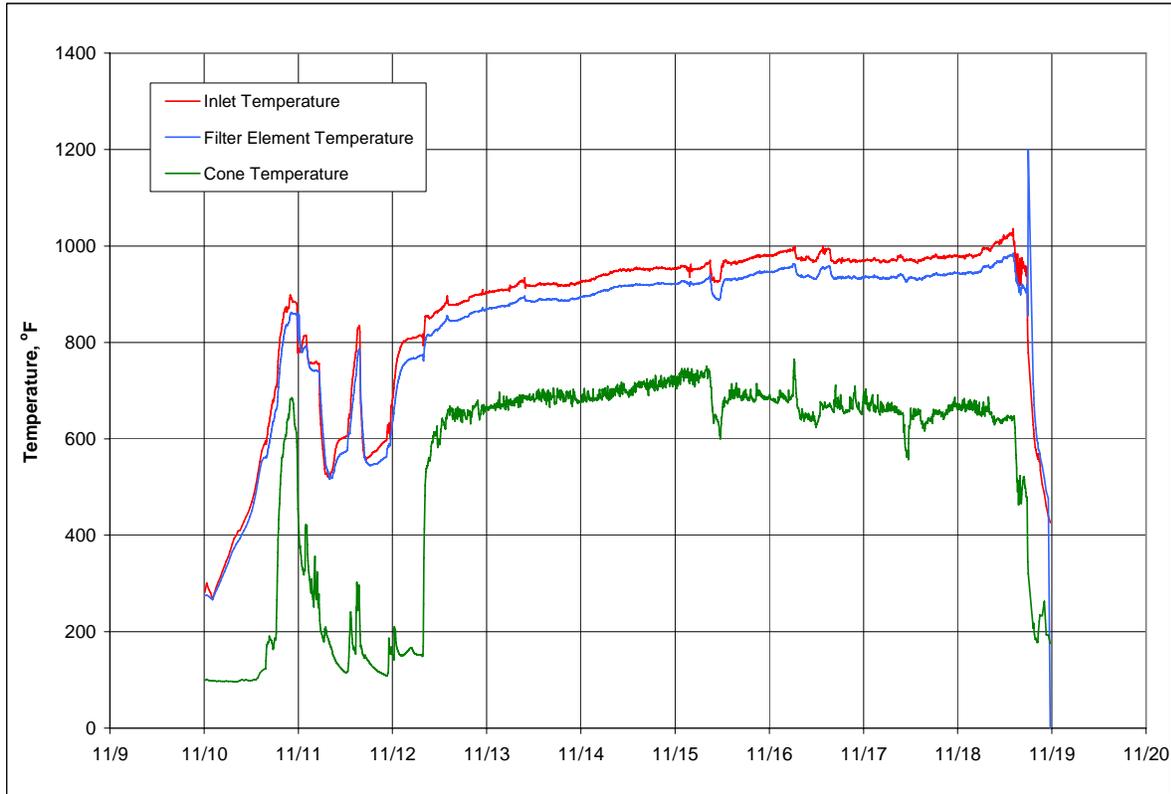


Figure A4-19 PCD Temperatures, 11/10/04 through 11/19/04

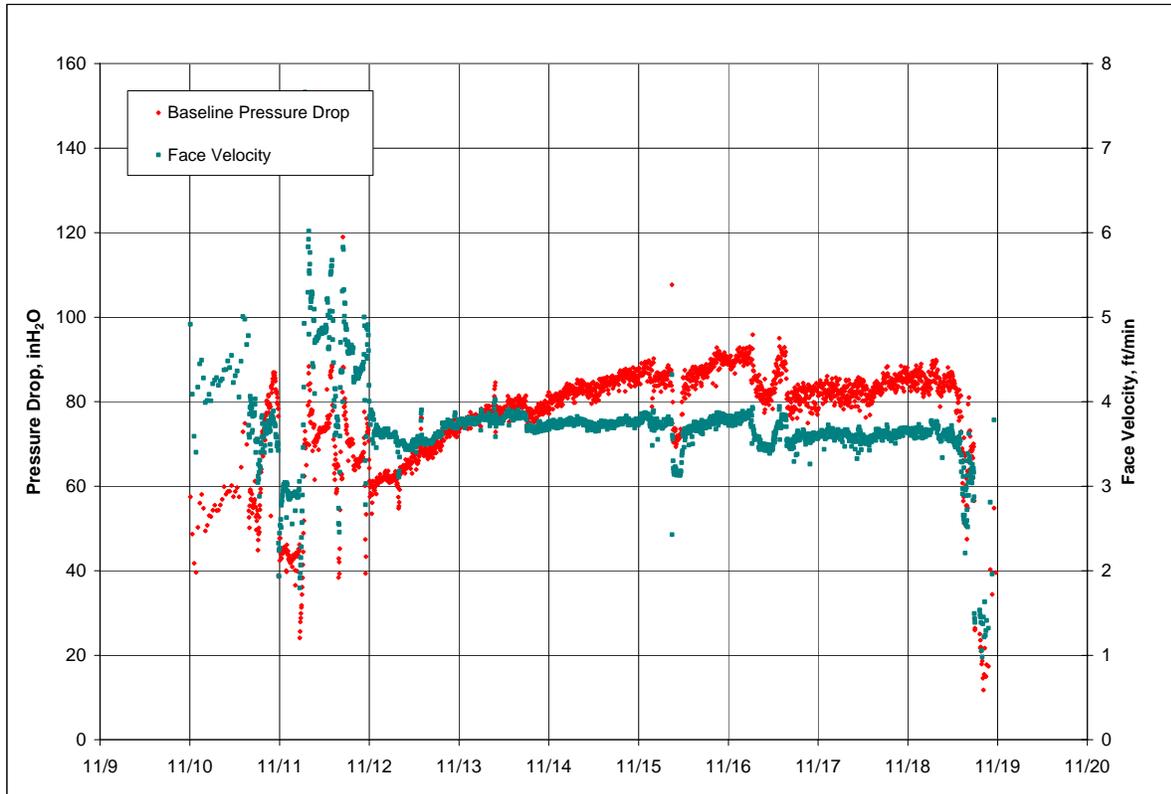


Figure A4-20 PCD Baseline Pressure Drop and Face Velocity, 11/10/04 through 11/19/04

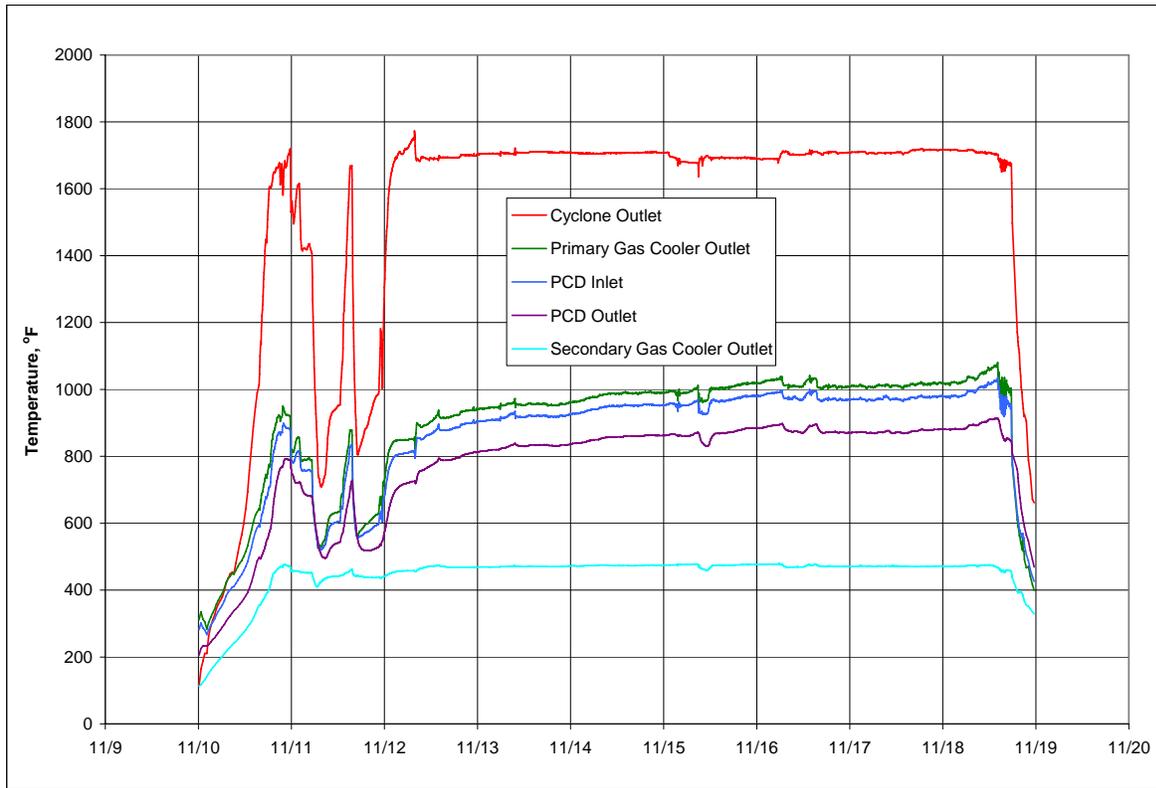


Figure A4-21 System Temperature Profile, 11/10/04 through 11/19/04

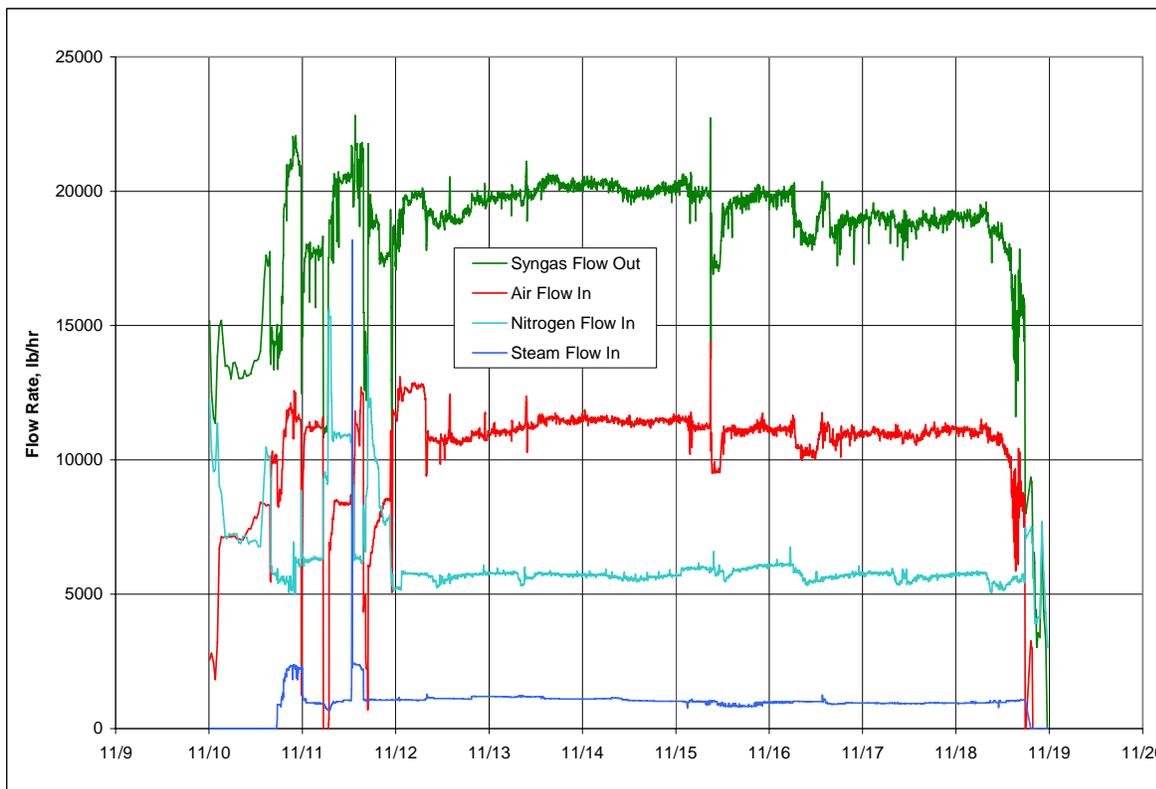


Figure A4-22 System Gas Flows, 11/10/04 through 11/19/04

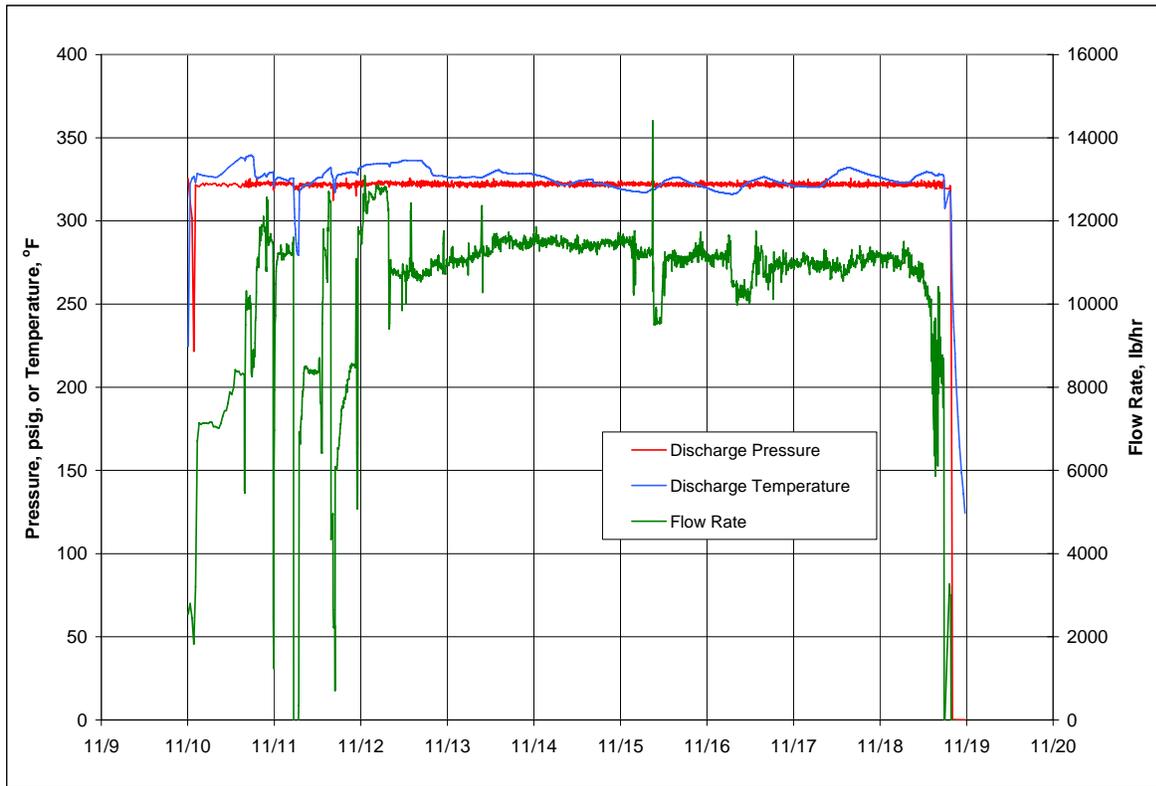


Figure A4-23 Main Air Compressor Operation, 11/10/04 through 11/19/04

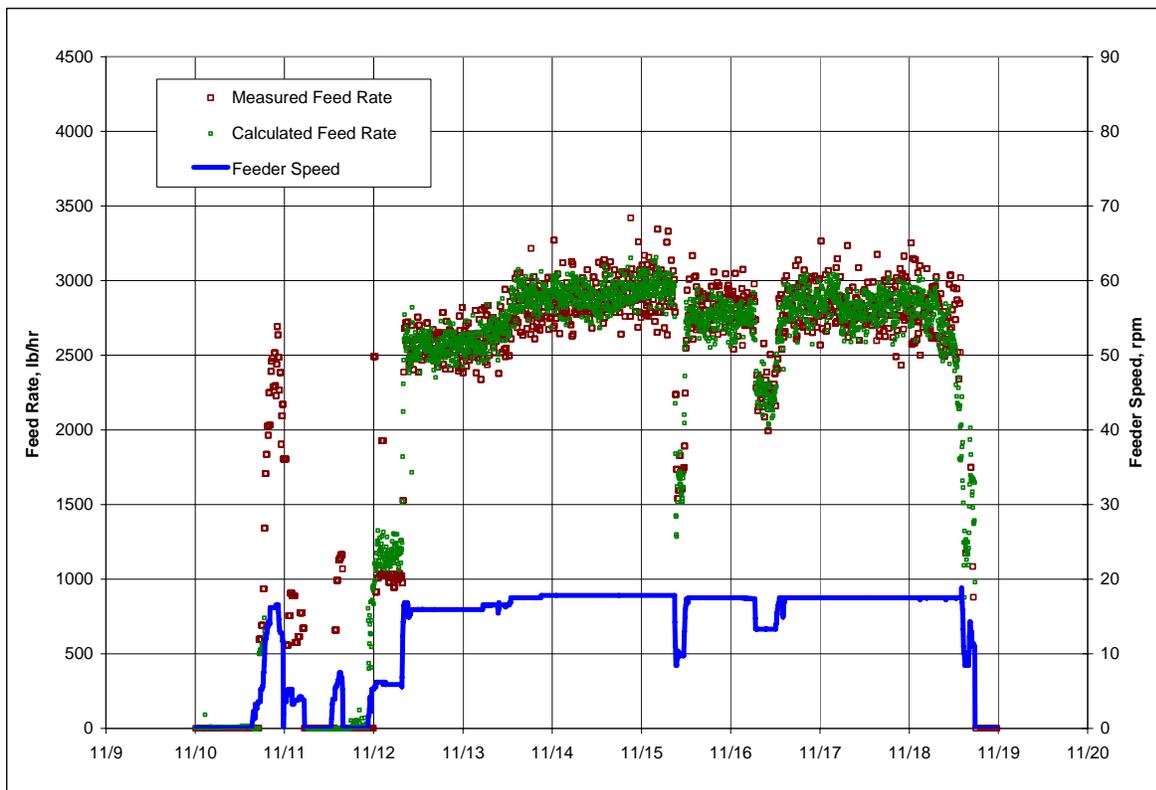


Figure A4-24 Original Coal Feeder Operation, 11/10/04 through 11/19/04

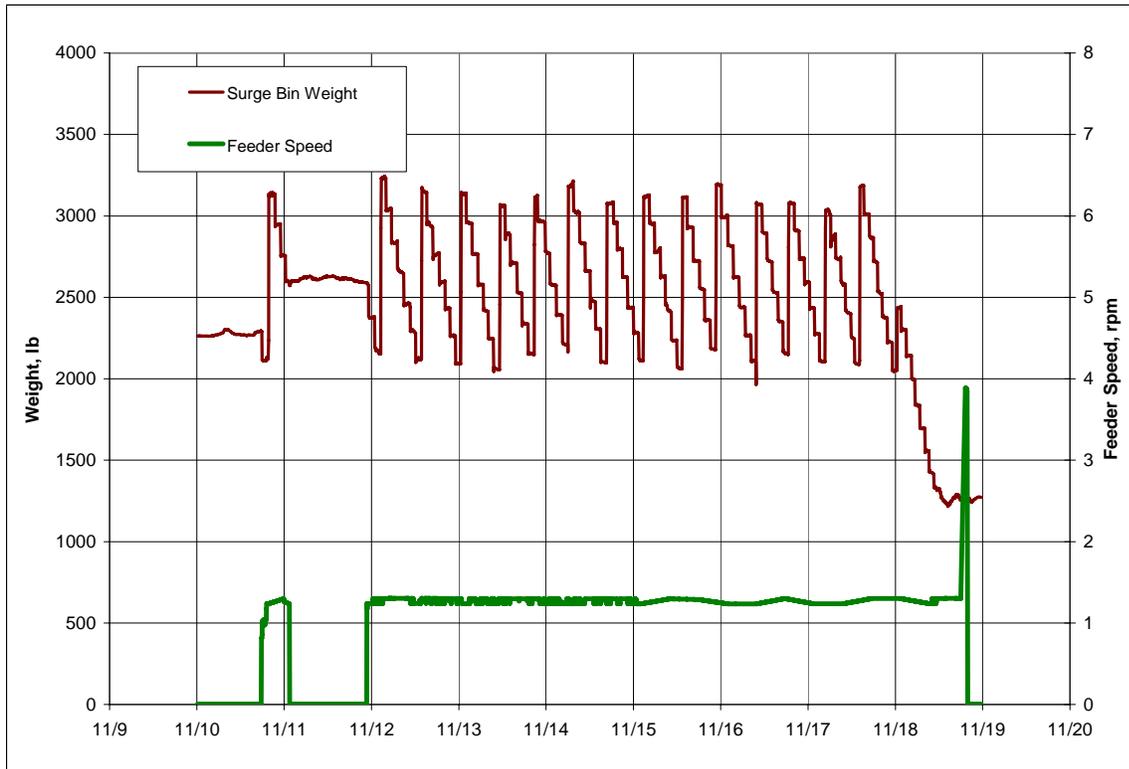


Figure A4-25 Sorbent Feeder Operation, 11/10/04 through 11/19/04

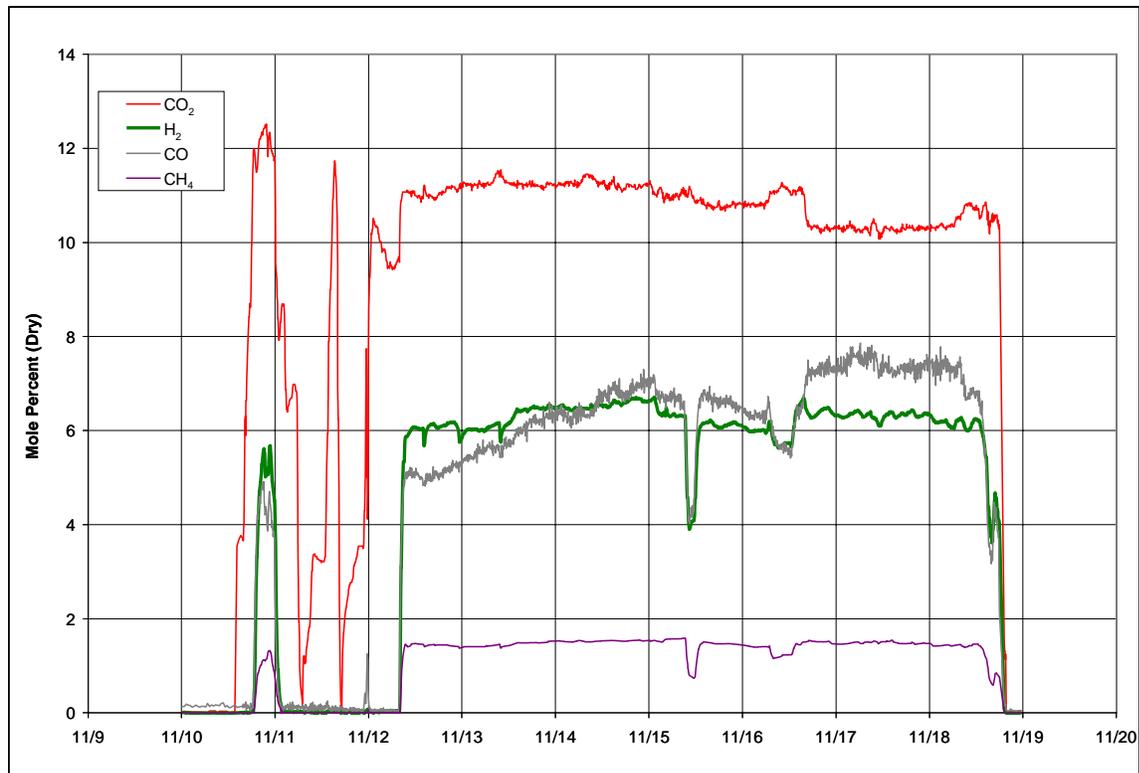


Figure A4-26 Syngas Analyzers, 11/10/04 through 11/19/04

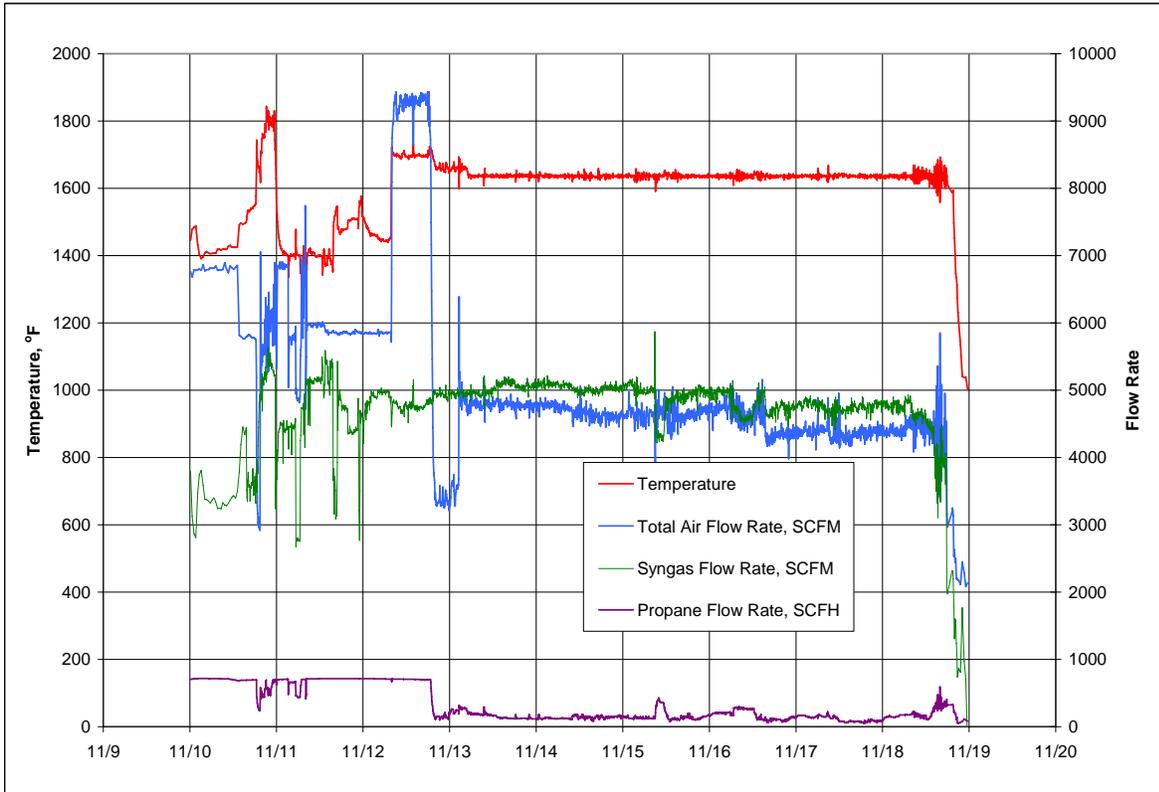


Figure A4-27 Atmospheric Syngas Combustor Operation, 11/10/04 through 11/19/04

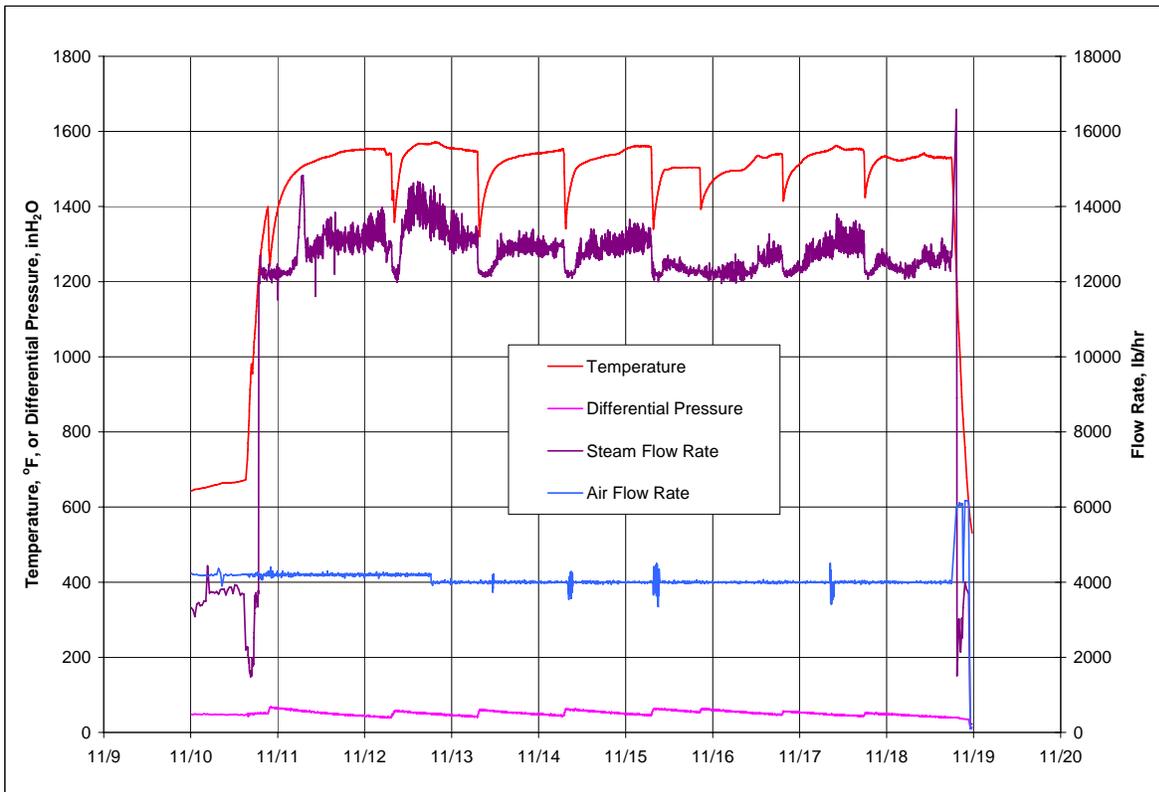


Figure A4-28 Fluidized Bed Combustor Operation, 11/10/04 through 11/19/04

APPENDIX A5 LHV PROJECTION CALCULATIONS

To project a commercial syngas LHV, the following adjustments are made to the raw syngas composition:

1. All non-air nitrogen is removed from the syngas. A commercial plant will have substantially less instrumentation than the PSDF. Because each individual instrument in a commercial plant will require the same purge flow rate as the corresponding instrument at the PSDF, the total instrument purge flow rate will be less. It is assumed that recycled syngas will be used in a commercial plant for aeration. This correction has the effect of increasing all the non-nitrogen syngas compositions and decreasing the nitrogen syngas composition. The recycle syngas flow enters the compressor after the “cold” gas cleanup system and is reheated before being used. Since the total amount of nitrogen entering the system is reduced, less coal energy will be required to heat the nitrogen, and the coal and air/oxygen feed rates will decrease accordingly. It is assumed that this coal would have been combusted to CO₂ and H₂O. Eliminating this additional coal reduces the syngas CO₂ and H₂O concentrations. The lower projected air rates for air blown mode also decrease the nitrogen content in the projected syngas, and thus decreases the syngas flow rate. The CO/CO₂ ratio will change due to the reduction in CO₂. This calculation requires an estimated recycle gas flow rate and an estimated steam aeration rate to determine the heat required to heat the recycle gas to system temperature. The recycle gas flow rate is estimated to be 2.4% of the syngas flow rate from the gasifier and is available at 235°F. The aeration steam flow rate is estimated to be 1.45% of the syngas flow rate from the gasifier and available at 660°F.
2. Small-scale pilot and demonstration units, such as the PSDF, have higher surface area to volume ratios than their scaled up commercial counterparts. Since the heat loss of a commercial plant is difficult to estimate, the projected heat loss is assumed to be zero (adiabatic). The coal, air, and oxygen rates are reduced; the syngas CO₂, H₂O, and N₂ concentrations are reduced; the CO/CO₂ ratio change. Based on energy balance data, the heat loss for the PSDF Transport gasifier is approximately 3.5 million Btu/hr.
3. The steam flow rate is adjusted. The steam to oxygen ratio will be the same for the PSDF and the commercial Transport Gasifier. Since Steps 1 and 2 reduce the amount of oxygen required, the steam flow rate will decrease correspondingly. The effect of lowering the steam rate will decrease the amount of H₂O in the syngas by the amount the steam rate was reduced. The steam rate and the H₂O content of the syngas are reduced, and hence, the LHV also changes.
4. The water gas shift is recalculated to reflect the gasifier exit temperature. Corrections #1, #2, and #3 change the water gas shift equilibrium constant without affecting the gasifier exit temperature. The commercial plant will operate at the same gasifier exit temperature as the PSDF and hence have the same water gas shift equilibrium constant. The H₂O, CO₂, CO, and H₂ concentrations are then adjusted based on the water gas shift equilibrium for the temperature of that particular operating period. The LHV could increase if H₂ and CO₂ are converted to H₂O and CO, since the

- LHV for CO is higher than for H₂. The LHV will decrease if H₂O and CO are converted to H₂ and CO₂. The LHV correction is usually small, but the change in composition is important if the syngas is used in a fuel cell or for chemical production where the H₂ concentration is a critical design parameter.
5. The commercial plant will use a cold syngas cleanup train that will drop the syngas temperature to 150°F, before being reheated prior to entering the gas turbine. At these conditions moisture will condense from the syngas and exit via a liquid stream. For the commercial design at 388 psia, the syngas water composition at the gas turbine inlet is 0.96 percent. Thus, the final step reduces the syngas moisture content to this value and adjusts the other contents accordingly.

The result of all of these corrections is the commercially projected LHV. Changes #1 and #2 both increase the oxygen blown LHV more than for the air blown LHV because 100% of the syngas nitrogen is removed in the oxygen blown projection, while only about 50% of the syngas nitrogen is removed for the air blown projection.

These calculations are an oversimplification of the gasification process. A more sophisticated model is required to precisely predict the effects of decreasing pure nitrogen and gasifier heat loss. Note that the projected syngas compositions are based on a projected coal rate, projected air rate, projected oxygen rate, projected steam rate, and a projected syngas rate.