POWER SYSTEMS DEVELOPMENT FACILITY
TOPICAL REPORT

GASIFICATION TEST CAMPAIGN TC18

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ABSTRACT

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

This report details Test Campaign TC18 of the PSDF gasification process. Test campaign TC18 began on June 23, 2005, and ended on August 22, 2005, with the gasifier train accumulating 1,342 hours of operation using Powder River Basin (PRB) subbituminous coal. Some of the testing conducted included commissioning of a new recycle syngas compressor for gasifier aeration, evaluation of PCD filter elements and failsafes, testing of gas cleanup technologies, and further evaluation of solids handling equipment. At the conclusion of TC18, the PSDF gasification process had been operated for more than 7,750 hours.

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TABLE OF CONTENTS

Section .... Page

Inside Cover
Disclaimer
Abstract
Acknowledgement

List of Tables iii
List of Figures iv

1.0 EXECUTIVE SUMMARY 1-1

1.1 Summary 1-1
1.2 Test Campaign Objectives 1-1
1.3 Test Campaign Summary 1-1
1.4 Test Campaign Performance 1-2
1.4.1 Transport Gasifier Performance 1-2
1.4.2 Particulate Control Device Performance 1-2
1.4.3 Performance of Other Systems 1-3

2.0 OPERATIONS 2.1-1

2.1 Transport Gasifier System Description 2.1-1
2.2 Detailed Test Campaign Objectives 2.2-1
2.3 Test Campaign Summary 2.3-1
2.4 Detailed Inspections/Conclusions 2.4-1
2.4.1 Transport Gasifier 2.4-1
2.4.2 Particulate Control Device 2.4-1
2.4.3 Other Systems 2.4-2
2.4.3.1 Recycle Gas Compressor 2.4-2
2.4.3.2 Continuous Fine Ash Depressurization System 2.4-2

3.0 PERFORMANCE 3.1-1

3.1 Transport Gasifier Performance 3.1-1
3.1.1 Overview 3.1-1
3.1.2 Gas Composition 3.1-2
3.1.3 Syngas Heating Values 3.1-3
3.1.4 Gasifier Solids Analyses 3.1-4
3.1.5 Carbon Conversion 3.1-9
3.1.6 Gasification Efficiencies 3.1-10

3.2 Particulate Control Device 3.2-1
3.2.1 Overview 3.2-1
3.2.2 Particle Mass Concentrations 3.2-2
3.2.3 Real Time Particle Monitoring 3.2-3
3.2.4 PCD Solids Analysis 3.2-3
3.2.4.1 Particle-Size Distributions 3.2-3
3.2.4.2 Dustcake Observations and Thickness Measurements 3.2-4
3.2.4.3 Physical Properties and Chemical Compositions 3.2-4
3.2.4.4 Dustcake Flow Resistance 3.2.5

3.3 Other Systems 3.3-1
3.3.1 Automation 3.3-1
3.3.2 Coal Feed Systems 3.3-1
3.3.3 Recycle Gas Compressor 3.3-1
3.3.4 Continuous Fine Ash Depressurization System 3.3-2
3.3.5 Advanced Syngas Cleanup 3.3-2

APPENDICES

A-1 Operation History A1-1
A-2 Equipment List A2-1
A-3 Material and Energy Balances A3-1
A-4 Operating Trends A4-1
A-5 LHV Projection Calculations A5-1
A-6 FTIR Hydrocarbon Results A6-1

LIST OF TABLES

Page

Table 2.3-1 TC18 Typical Operating Conditions for the Transport Gasifier and Particulate Control Device 2.3-12

Table 3.1-1 Operating Periods 3.1-13
Table 3.1-2 Operating Conditions 3.1-16
Table 3.1-3 Wet Gas Compositions, Molecular Weight, and Heating Value 3.1-19
Table 3.1-4 Syngas Sulfur Concentrations 3.1-22
Table 3.1-5 Projected Gas Compositions, Molecular Weight, and Heating Value 3.1-25
Table 3.1-6 Coal Analysis 3.1-28
Table 3.1-7 Standpipe Solids Analysis 3.1-29
Table 3.1-8 Loop Seal Solids Analysis 3.1-30
Table 3.1-9 PCD Solids Analysis 3.1-32
Table 3.1-10 Historical Standpipe and PCD Solids Particle Sizes and Densities 3.1-35
Table 3.1-11 Carbon Conversion and Gasification Efficiencies 3.1-36

Table 3.2-1 PCD In Situ Particulate Measurements 3.2-7
Table 3.2-2 Thickness Measurements of Residual Dustcake 3.2-9
Table 3.2-3 Physical Properties of In Situ and Hopper Samples 3.2-10
Table 3.2-4 Chemical Composition of In Situ and Hopper Samples 3.2-12
Table 3.2-5 Physical Properties of Residual Dustcake 3.2-14
Table 3.2-6 Chemical Composition of Residual Dustcake 3.2-14
Table 3.2-7 Transient Drag Determined from PCD Performance and from Lab Measurements 3.2-15
Table 3.2-8 Comparison of Average Drag Values Determined from PCD Performance and from Lab Measurements 3.2-17

Table 3.3.4-1 Nominal Properties of Sulfur Sorbent 3.3-4
Table 3.3.4-2 Nominal Properties of Ammonia Cracking and Hydrocarbon Reforming Catalysts 3.3-4
Table 3.3.4-3 Operating Parameters for Fixed Bed Syngas Desulfurization 3.3-5
Table 3.3.4-4 Operating Parameters for Fixed Bed Syngas Ammonia Cracking and Hydrocarbon Reforming 3.3-5

LIST OF FIGURES

		Page
Figure 2.1-1	Flow Diagram of the PSDF Gasification Process	2.1-3
Figure 2.1-2	KBR Transport Gasifier	2.1-4
Figure 2.1-3	Siemens Particulate Control Device	2.1-5
Figure 2.4-1	Titanium Filter Elements on Bottom Plenum	2.4-3
Figure 2.4-2	Titanium Filter Element with Spots of Molten Material	2.4-3
Figure 2.4-3	Ruptured Pall Fuse Failsafe	2.4-4
Figure 3.1-1	Syngas H_2O Concentration.	3.1-38
Figure 3.1-2	Wet Syngas Composition	3.1-38
Figure 3.1-3	Wet Syngas Molecular Weight & Nitrogen Concentration	3.1-39
Figure 3.1-4	Syngas Sulfur Concentration	3.1-39
Figure 3.1-5	Syngas Lower Heating Values	3.1-40
Figure 3.1-6	Coal Carbon & Moisture	3.1-41
Figure 3.1-7	Coal Sulfur & Ash	3.1-41
Figure 3.1-8	Coal Lower Heating Value	3.1-42
Figure 3.1-9	Standpipe Solids SiO_2, CaO, & Al_2O_3	3.1-42
Figure 3.1-10	Loop Seal Solids SiO_2, CaO, & Al_2O_3	3.1-43
Figure 3.1-11	Loop Seal Solids Organic Carbon and $CaCO_3$	3.1-43
Figure 3.1-12	PCD Solids Organic Carbon	3.1-44
Figure 3.1-13	PCD Solids Alumina & Silica	3.1-44
Figure 3.1-14	PCD Solids $CaCO_3$ & CaS	3.1-45
Figure 3.1-15	PCD Solids Sulfation & Calcination	3.1-45
Figure 3.1-16	Loop Seal & PCD Solids Organic Carbon	3.1-46
Figure 3.1-17	Standpipe, Loop Seal, & PCD Solids Calcium	3.1-46
Figure 3.1-18	Standpipe, Loop Seal, & PCD Solids Silica	3.1-47
Figure 3.1-19	Coal Sauter Mean Diameter Particle Size	3.1-47
Figure 3.1-20	Coal Mass Median Diameter Particle Size	3.1-48
Figure 3.1-21	Percent Coal Oversize	3.1-48
Figure 3.1-22	Percent Coal Fines	3.1-49
Figure 3.1-23	Standpipe Solids Particle Size	3.1-49
Figure 3.1-24	Loop Seal Solids Particle Sizes	3.1-50
Figure 3.1-25	PCD Solids Particle Sizes	3.1-50
Figure 3.1-26	Standpipe, Loop Seal, & PCD Solids Bulk Density	3.1-51
Figure 3.1-27	Carbon Conversion	3.1-51
Figure 3.1-28	Carbon Conversion of Five Coals	3.1-52
Figure 3.1-29	Cold Gasification Efficiency	3.1-52
Figure 3.1-30	Hot Gasification Efficiency	3.1-53

Figure 3.2-1 PCD Inlet Particle Concentration as a Function of Coal Feed Rate 3.2-18
Figure 3.2-2 PCME Particulate Monitor Response to a Major Leak on August 22, 2005 .... 3.2-19
Figure 3.2-3 Comparison of Particle-Size Distributions of In Situ Samples 3.2-20
Figure 3.2-4 Comparison of Hopper and In Situ Particle Size Distributions........ 3.2-21
Figure 3.2-5 Effect of Non-Carbonate Carbon Content on Specific Surface Area 3.2-22
Figure 3.2-6 Lab-Measured Drag as a Function of Particle Size 3.2-23
Figure 3.2-7 PCD Transient Drag Versus Carbon Content of In Situ Samples 3.2-24
Figure 3.2-8 Comparison of PCD Transient Drag with Laboratory Measurements.... 3.2-25
Figure 3.2-9 Rupture Disc Failsafe Test Setup........ 3.2-27

Figure 3.3.3-1 Recycle Gas Compressor 3.3-2

Figure 3.3.4-1 Temperature, Pressure and Flow Profiles, RVS-1 Sulfur Sorbent........ 3.3-6
Figure 3.3.4-2 Sulfur Profile, RVS-1 Sulfur Sorbent 3.3-6
Figure 3.3.4-3 Ammonia Profile, G-117RR Catalyst........ 3.3-7
Figure 3.3.4-4 Benzene Profile, G-117RR Catalyst 3.3-7
Figure 3.3.4-5 Ethylene Profile, G-117RR Catalyst........ 3.3-8
Figure 3.3.4-6 Phenanthrene Profile, G-117RR Catalyst........ 3.3-8
Figure 3.3.4-7 Naphthalene Profile, G-117RR Catalyst........ 3.3-9
Figure 3.3.4-8 Acenaphthene Profile, G-117RR Catalyst........ 3.3-9
Figure 3.3.4-9 Ammonia Profile, G-31 Catalyst........ 3.3-10
Figure 3.3.4-10 Benzene Profile, G-31 Catalyst 3.3-10
Figure 3.3.4-11 Ethylene Profile, G-31 Catalyst........ 3.3-11
Figure 3.3.4-12 Phenanthrene Profile, G-31 Catalyst........ 3.3-11
Figure 3.3.4-13 Naphthalene Profile, G-31 Catalyst........ 3.3-12
Figure 3.3.4-14 Acenaphthene Profile, G-31 Catalyst........ 3.3-12

APPENDICES
Figure A1-1 Operating Hours Summary for the PSDF Gasification Process........ A1-4

Figure A3-1 Mass Balance........ A3-2
Figure A3-2 Energy Balance A3-2
Figure A3-3 Carbon Balance........ A3-3

Figure A4-1 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 6/23/05 through 7/24/05 .... A4-1
Figure A4-2 Standpipe and Loop Seal Temperatures, 6/23/05 through 7/24/05 .... A4-1
Figure A4-3 Gasifier Pressures, 6/23/05 through 7/24/05 .... A4-2
Figure A4-4 Gasifier Differential Pressures, 6/23/05 through 7/24/05........ A4-2
Figure A4-5 PCD Temperatures, 6/23/05 through 7/24/05 .... A4-3
Figure A4-6 PCD Baseline Pressure Drop and Face Velocity, 6/23/05 through 7/24/05........ A4-3
Figure A4-7 System Temperature Profile, 6/23/05 through 7/24/05........ A4-4
Figure A4-8 System Gas Flows, 6/23/05 through 7/24/05........ A4-4
Figure A4-9 Main Air Compressor Operation, 6/23/05 through 7/24/05........ A4-5

Figure A4-10 Original Coal Feeder Operation, 6/23/05 through 7/24/05 .... A4-5
Figure A4-11 Developmental Coal Feeder Operation, 6/23/05 through 7/24/05 .... A4-6
Figure A4-12 Syngas Analyzers, 6/23/05 through 7/24/05 .... A4-6
Figure A4-13 Atmospheric Syngas Combustor Operation, 6/23/05 through 7/24/05 .... A4-7
Figure A4-14 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 7/24/05 through 8/23/05 .... A4-8
Figure A4-15 Standpipe and Loop Seal Temperatures, 7/24/05 through 8/23/05.......... A4-8
Figure A4-16 Gasifier Pressures, 7/24/05 through 8/23/05 .... A4-9
Figure A4-17 Gasifier Differential Pressures, 7/24/05 through 8/23/05.......................... A4-9
Figure A4-18 PCD Temperatures, 7/24/05 through 8/23/05... A4-10
Figure A4-19 PCD Baseline Pressure Drop and Face Velocity, 6/23/05 through 7/24/05 .... A4-10
Figure A4-20 System Temperature Profile, 7/24/05 through 8/23/05........................... A4-11
Figure A4-21 System Gas Flows, 7/24/05 through 8/23/05.. A4-11
Figure A4-22 Main Air Compressor Operation, 7/24/05 through 8/23/05 .... A4-12
Figure A4-23 Original Coal Feeder Operation, 7/24/05 through 8/23/05.................... A4-12
Figure A4-24 Developmental Coal Feeder Operation, 7/24/05 through 8/23/05 .... A4-13
Figure A4-25 Syngas Analyzers, 7/24/05 through 8/23/05.. A4-13
Figure A4-26 Atmospheric Syngas Combustor Operation, 7/24/05 through 8/23/05 .... A4-14

Figure A6-1 Ethane & Ethylene Concentrations A6-1
Figure A6-2 Propane & Cyclopropane Concentrations A6-2
Figure A6-3 Benzene Concentrations A6-2
Figure A6-4 Cyclone Exit Temperature & Benzene Concentration A6-3
Figure A6-5 Naphthalene & Phenanthrene Concentrations A6-3

1.0 EXECUTIVE SUMMARY

1.1 SUMMARY

Test campaign TC18 began on June 23, 2005, and lasted until August 22, 2005, accumulating 1,342 hours of on-coal operation for the PSDF gasification process. For the first time, recycled syngas was used for aeration in the Transport Gasifier. The recycle gas system operated for approximately 376 hours, resulting in higher syngas heating values and causing no disturbance in gasifier operation. Both the original and the developmental coal feeders were operated, and adjustments were made to coal mill operating parameters to optimize coal feed operations.

During the previous test run, erosion in the primary cyclone caused the gasifier to continuously lose bed material. In TC18, the repaired cyclone had a higher efficiency, which made the solids carryover rate to the particulate control device (PCD) lower, despite a high coal feed rate.

PCD filter element and failsafe testing was continued, and the PCD operated without problems until a temperature and pressure excursion occurred, which caused a filter element and failsafe failure. The continuous fine ash depressurization (CFAD) system, which was slightly modified prior to the run, continued to demonstrate reliable operation. The hot gas cleanup system was operated to evaluate desulfurization, hydrocarbon reforming, and ammonia cracking technologies.

1.2 TEST CAMPAIGN OBJECTIVES

TC18 evaluated gasifier train operations with Powder River Basin (PRB) subbituminous coal using air as the gasification oxidant. The primary test objectives were:

- Recycle Syngas System Commissioning
- Evaluating Minor Gasifier Modifications
- Evaluating CFAD Modifications
- PCD Filter Element and Failsafe Testing.

1.3 TEST CAMPAIGN SUMMARY

On June 25, coal feed began in combustion mode to heat the gasifier. Once the gasifier reached operating temperature, it was transitioned to gasification mode. On June 29, testing of the developmental feeder began. The feeder ran well for short periods of time, feeding approximately 3,000 pph of coal. Later, the developmental feeder automatic mass flow control was successfully implemented.

The developmental feeder was tested using the offline coal feed system to examine its reliability using larger coal particles, and the system seemed to handle the particle size increase well. On July 8, a test was performed by increasing the air flow to the lower mixing zone (LMZ), with the additional amount of air varying from 2,500 to 6,800 pph, causing the high temperature zone to shift.

The recycle syngas compressor was started on July 27. While operating with syngas for gasifier aeration, operations remained stable, and the gasifier temperature profile was not negatively affected. The raw syngas lower heating value increased approximately 8 to 10 percent.

From August 1 to August 9, the nitrogen supply vendor reduced its delivery capacity due to operational problems. To reduce nitrogen consumption, the coal mills were not operated, so the coal feed rate was decreased. Use of recycle syngas allowed gasifier operation to continue. During this time, the dry lower heating value rose to the highest value seen during air blown operation due to lower nitrogen purge rates and the use of recycle gas for aeration.

Testing a mixture of 50 percent air and 50 percent nitrogen to convey the coal into the gasifier rather than 100 percent nitrogen began on August 20. The transition was smooth, and the system operated reliably. The use of transport air slightly increased riser and standpipe temperatures, and this temperature effect was compensated for by reducing air flow rates through the mixing zone and lower mixing zone.

On August 21, fine coal caused poor coal transfer through the original coal feeder. Eventually, loss of coal feed led to oxygen breakthrough to the PCD. A significant temperature excursion took place, and the online particulate monitoring device showed an elevated reading, indicating a leak. On August 22, a PCD outlet sample was taken and revealed substantial solids loading at the PCD outlet, so the system was shut down.

1.4 TEST CAMPAIGN PERFORMANCE

Performance of the major equipment during TC18 is summarized in the following three sections:

1.4.1 Transport Gasifier Performance

- The TC18 carbon conversion ranged from 88 to 97 percent, typical for PRB air blown gasification.
- The raw lower heating values at the gasifier outlet ranged from 42 and 73 Btu/SCF without recycle syngas and from 55 to 74 Btu/SCF with recycle syngas. Some overlap between the two sets of data exists due to other operating parameters such as the coal feed rate. These raw lower heating values resulted in projected turbine inlet heating values between 105 and 153 Btu/SCF. The use of recycle syngas did not affect the projected values, as the projection accounts for the use of recycle syngas.
- The cold gasification efficiency ranged from 45.8 to 57.0 percent. The use of recycle syngas did not greatly affect the cold gasification efficiency. The commercially projected cold gasification efficiency was between 64.7 and 73.5 percent.

- The hot gasification efficiency ranged from 77.8 to 86.0 percent. As in the case of cold gasification efficiency, the use of recycle syngas did not greatly affect the hot gasification efficiency.
- The sulfur concentration of the syngas at the gasifier exit ranged from 410 to 462 ppm. No sulfur sorbent was fed into the gasifier during the test campaign.
- The syngas ammonia concentration ranged from 898 to 1,835 ppm.

1.4.2 PCD Performance

- Testing of iron aluminide elements continued, with several individual elements having accumulated over 7,000 hours of gasification exposure at the conclusion of TC18.
- The rupture disc failsafe tester was used with the Pall fuse failsafe to simulate catastrophic filter failure. Although the rupture disc burst prematurely, the test was completed, and outlet sampling indicated good performance of the Pall fuse.
- Outlet loading sampling indicated a small amount of particle penetration of up to 0.39 ppmw during the first two days of the run. Subsequent outlet samples indicated solids loading below the lower limit of detection of 0.1 ppmw until the final day of testing, when particle penetration resulted from a filter element and failsafe failure.
- An excursion in the PCD precipitated the shutdown because of significant particle penetration. The upset occurred when a coal feed stoppage caused oxygen breakthrough to the PCD, which caused a severe temperature excursion and a pressure swing. After this excursion, the online particulate monitor indicated particle penetration, which was later confirmed by outlet sampling. Upon inspection, the three titanium elements installed were found to be severely damaged, as was a failsafe corresponding to one of these elements.

1.4.3 Performance of Other Systems

- The recycle syngas compressor, a centrifugal design manufactured by Sundyne Corporation, performed well throughout the test run. The compressor operated on syngas for a total of 376 hours. The raw lower heating value increased approximately 8 to 10 percent as a result of the use of recycled syngas for gasification aeration.
- Both coal feeders ran well throughout the run. The original coal feeder operated for more than 1,300 hours without any major system component failures, and the spheri valves cycled over 13,000 times without failure. The developmental coal feeder was operated for over 600 hours, the longest run time to date.
- Before the test run, the fines removal screw cooler was removed, and the CFAD, a Southern Company proprietary design, system was relocated and connected directly to the PCD hopper. The system operated for over 1,300 hours and removed over 250 tons of gasification ash. CFAD successfully operated using an automated solids level controller. The CFAD system was available throughout the run and did not require maintenance.

- The hot gas cleanup unit achieved 76 hours of desulfurization using Sud-Chemie RVS-1 sulfur sorbent. The inlet H_2S level was typically around 300 ppm, while the outlet H_2S level was below the detection limit.
- Organic reforming and ammonia cracking tests were conducted in the hot gas cleanup unit for 290 hours using the previously tested Sud-Chemie nickel-based catalyst, G-117RR. These tests were also performed using a new Sud-Chemie nickel-based catalyst, G-31, for a period of 13 hours. The new G-31 catalyst showed higher conversions, and both the ammonia and benzene levels were reduced by up to 99 percent.

2.0 OPERATING SUMMARY

2.1 PSDF GASIFICATION PROCESS 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 Power Generation, 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 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 using either air or oxygen as the gasification oxidant. The particulate-laden gas exiting the gasifier is filtered by a downstream high temperature, high pressure filter vessel, the Siemens particulate control device (PCD). A slipstream syngas clean-up skid is also available to test various pollutant control technologies. A flow diagram of the gasification process is shown Figure 2.1-1. At the conclusion of TC18, the Transport Gasifier train had operated for over 7,750 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 loop seal, and a J-leg. Steam and either air or oxygen are mixed together and introduced in the lower mixing zone while the fuel and additional air and steam are added in the upper mixing zone. The steam and oxidant, along with the fuel, and circulating 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. To control the standpipe level, solids can be removed from the gasifier, cooled in a screw cooler, and reduced in pressure in a lock hopper system.

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 lock hopper 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 backpulsing, 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 and are removed through a continuous fine ash depressurization (CFAD) system, a Southern Company designed system for solids removal.

After exiting the PCD, a portion of the syngas can be directed to the piloted syngas burner (PSB), where the gas is combusted using air from the turbine compressor. The PSB is a gas turbine combustor designed to burn coal-derived syngas with a lower heating value of less than 100Btu/sct. 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 main stream of syngas is then cooled in a secondary gas cooler, which reduces the temperature to about 450°F. Some of this cooled is compressed and sent to the gasifier for aeration to aid in solids circulation. The remaining syngas is reduced to near atmospheric pressure through a pressure control valve. The gas is then sent to the atmospheric syngas combustor which oxidizes carbon monoxide, reduced sulfur compounds (H_2S, COS, and CS_2), and reduced nitrogen compounds (NH_3 and HCN). The gas from the atmospheric syngas combustor goes to a heat recovery boiler, through a baghouse, and then is discharged out a stack.

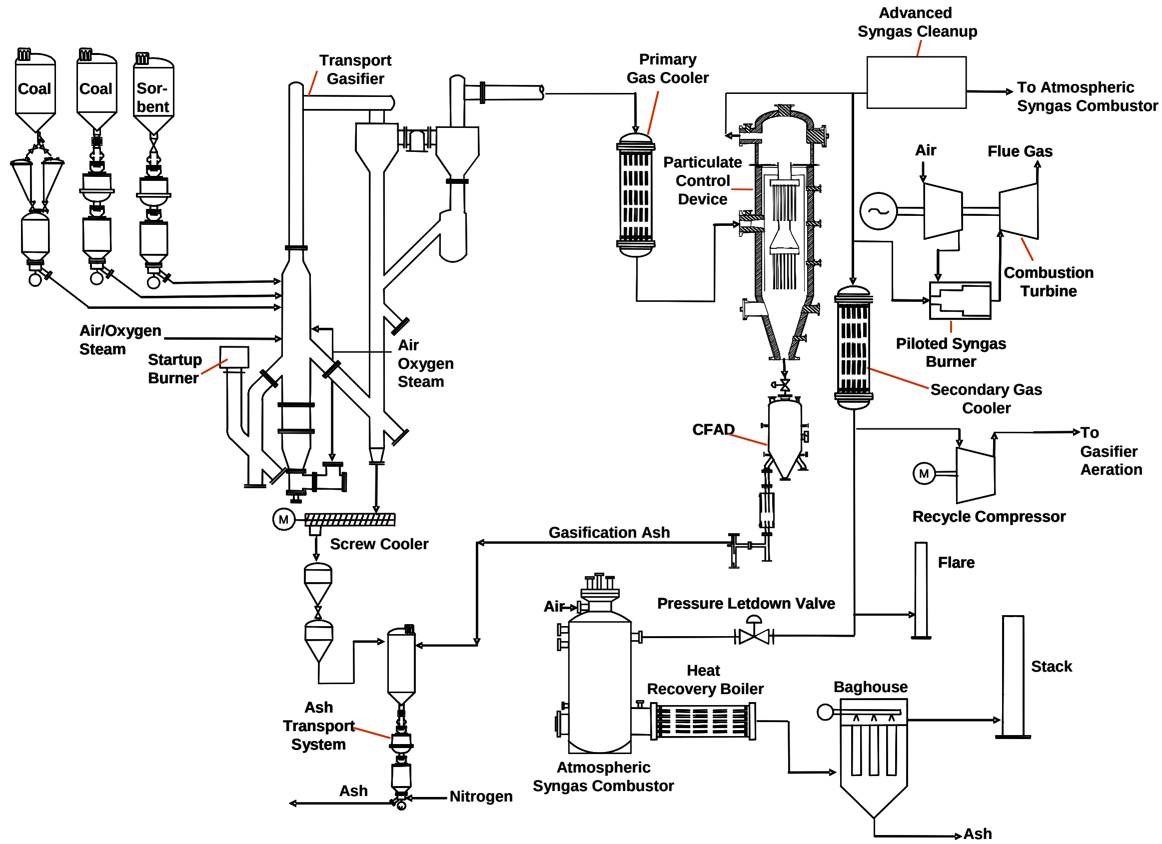

Figure 2.1-1 Flow Diagram of the PSDF Gasification Process

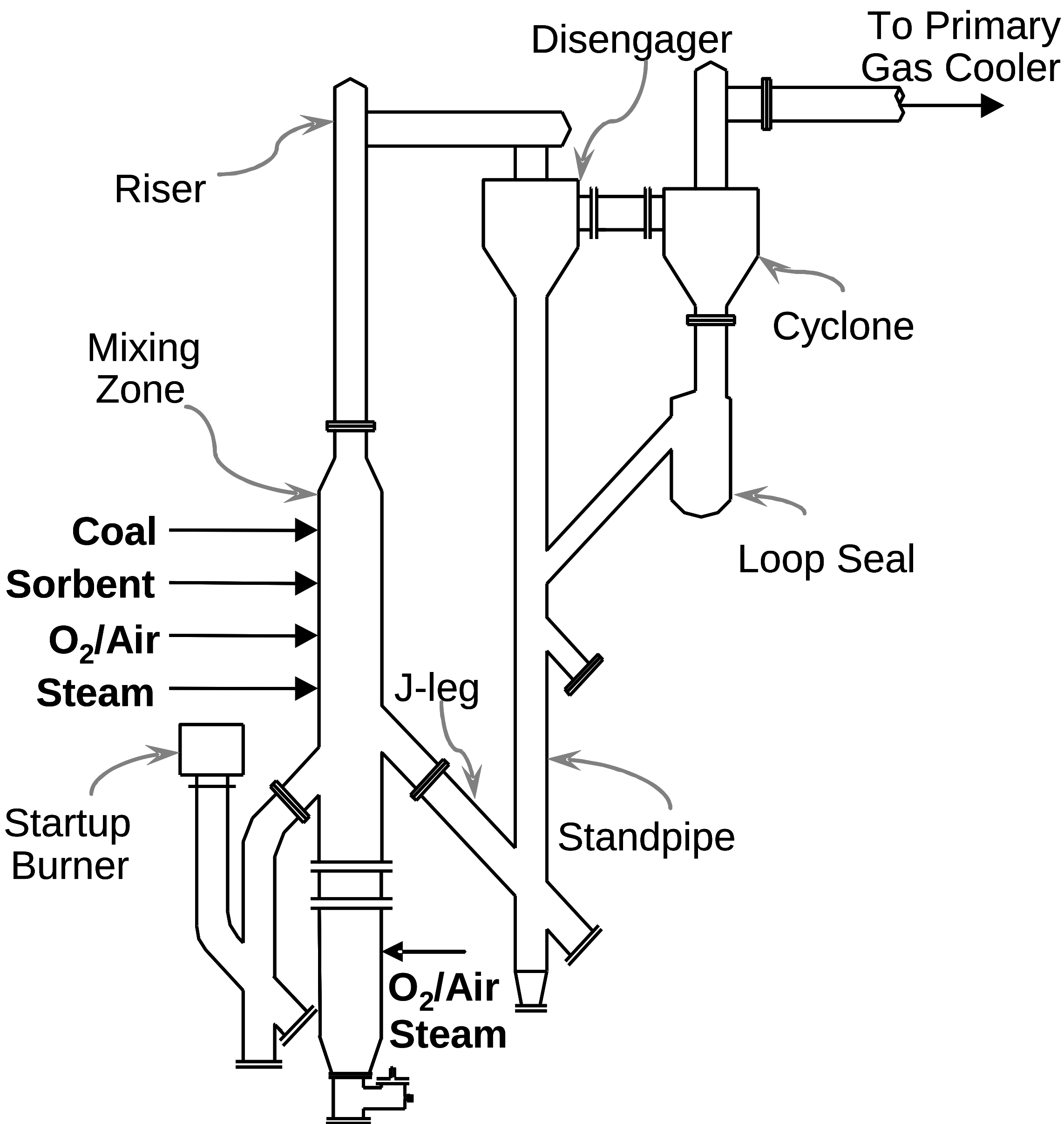

Figure 2.1-2 KBR Transport Gasifier

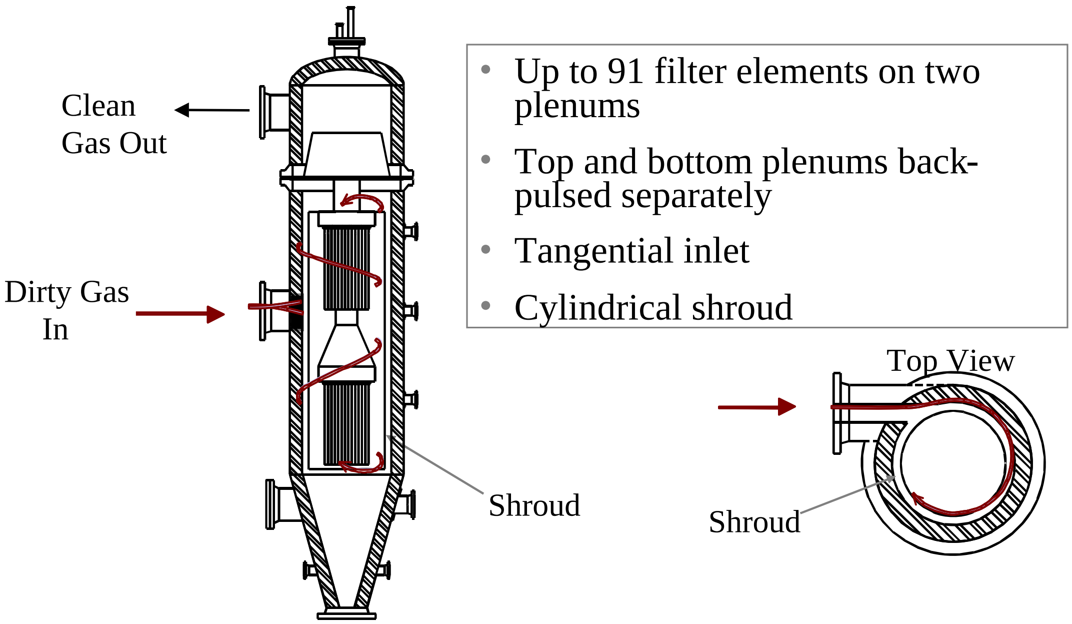

Figure 2.1-3 Siemens Particulate Control Device

2.2 DETAILED TEST CAMPAIGN OBJECTIVES

The primary test campaign objectives for TC18 are listed below.

- *Recycle Gas System Commissioning* – The recycle gas compressor was used for the first time for aeration of the gasifier loop seal, standpipe, and J-leg. Using recycled syngas for aeration in place of nitrogen increased the syngas quality, with the syngas lower heating value increasing between 8 and 10 percent when compared on an air-to-coal basis. The compressor operated on recycled syngas for a total of 376 hours. While using recycled syngas, operations of the Transport Gasifier were stable, with steady solids circulation and acceptable solids collection efficiencies. As expected, the gasifier mixing zone temperatures increased slightly. The use of recycled syngas during a nitrogen supply shortage was necessary for the continuation of the test campaign.
- *Gasifier Modifications* – The new lower mixing zone aeration system was successfully tested. The increase in air flow redistributed the solids inventory in the gasifier, resulting in an increase in standpipe level, which in turn increased the solids circulation rate.
- *Continuous Fine Ash Depressurization (CFAD) Modifications* – The CFAD system was relocated and connected directly to the bottom of the PCD. The system operated well throughout the test run as it cooled and depressurized solids collected in the PCD. The CFAD system successfully operated using an automated level controller and removed over 250 tons of solids from the PCD.
- *Filter Element and Failsafe Testing* – The performance of previously exposed iron aluminide and new titanium filter elements were evaluated during the test run. The iron aluminide filter elements performed well during upsets, and several of these elements have been exposed to syngas for over 7,000 hours. Unfortunately, the titanium filter elements failed, apparently during a severe temperature excursion in the PCD. The rupture disc failsafe tester was installed with a Pall fuse failsafe to simulate a catastrophic filter element failure and its performance was evaluated. Although the device opened prematurely during the first few days of the test run, the test was completed, and outlet sampling indicated good failsafe performance with an outlet loading below the detection limit of 0.1 ppmw.

Secondary objectives included the following:

- *Continued Sensor Development* – Sensor development continued in several different areas of the process. The new ceramic tip thermowells worked well and improved thermocouple longevity significantly compared to previous test campaigns. Nine of fourteen riser thermowells lasted the entire run compared to four of fifteen in TC17. A new Granucore coal feed measurement was installed which gave reliable feed rate indications.
- *Continued Automation Development* – Enhancements to the automatic temperature controls were successfully tested by manually forcing a low temperature in the control measurement block and verifying that the temperature controls rejected to manual. The scheme protected against erroneous readings reasonably well, until several thermocouples were lost. The gasifier temperature controls performed well at typical gasifier operating conditions, maintaining both mixing zone and gasifier exit temperatures within 10°F of their respective set points.
- *Coal Feeder Development* – The original coal feeder ran well for over 1,300 hours without any major system component failures. The particle size distribution and moisture content were

varied to continue evaluating the operating envelope. A mixture of nitrogen and air was used to transport coal in the original coal feeder. The initial transition resulted in a slight increase in upper mixing zone temperatures as expected. The developmental coal feeder also ran well with over 600 hours of operation until the system was shut down to conserve nitrogen. The system was automatically controlled by mass flow and exhibited acceptable responses to setpoint changes, but had difficulty when a high amount of fines or moisture was present in the coal.

- *Air Compliance Testing* – Annual air compliance testing was completed, verifying compliance with the PSDF air permit.
- *Superheated Steam System Commissioning* – The new electric steam superheater was successfully commissioned, and the low flow trip was verified. The superheater experienced no problems throughout the entire test run and successfully supplied superheated steam to the gasifier in place of the atmospheric fluidized bed combustor.
- *Coal Mill Modifications* – Testing continued with the anti-rathole device in the Foster Wheeler pulverized coal silo, which prevented particle size segregation. New dehumidifiers in the coal mill systems were tested, decreasing coal moisture by one to two percent by weight.
- *Modified Filter Installation Evaluation* – New filter holding mechanisms, gaskets, and self-compensating torque mechanisms for filter element and failsafe installation were implemented, resulting in improved torque retention.
- *Gas and Solids Sampling* – Gas and solids samples were taken during the test run to provide composition data. Two mixing zone gas samples, two riser gas samples, and one riser solids sample were taken. Data was provided to DOE to support its ongoing modeling effort.
- *Syngas Desulfurization Testing* – The hot gas cleanup system achieved 76 hours of desulfurization on a 12 lb/hr syngas slipstream tests using Sud-Chemie RVS-1 sulfur sorbent, resulting in outlet H_2S concentrations below the detection limit.
- *Ammonia and Organics Cracking* – Ammonia and organics cracking tests were conducted for 290 hours using the previously tested Sud-Chemie nickel-based catalyst, G-117 RR. Ammonia cracking tests were also conducted for 13 hours using the new Sud-Chemie nickel-based catalyst, G-31.

2.3 DETAILED TEST CAMPAIGN SUMMARY

The activities that occurred during the outage preceding test run TC18 included 10 equipment modifications. The most significant of these are listed below:

- A new recycle gas compressor was installed to supply recycled syngas to the Transport Gasifier to fluidize the loop seal, standpipe, and J-leg.
- An electric steam superheater was installed to supply superheated steam to the gasifier in place of the atmospheric fluidized bed combustor.
- The Continuous Fine Ash Depressurization (CFAD) system was modified to take solids directly from the PCD for cooling and depressurization.
- Modifications to the lower mixing zone (LMZ) were made to test the gasifier performance at higher air flow rates to the LMZ.

Typical operating conditions for the Transport Gasifier and the PCD can be found in Table 2.3-1. Operating trends for TC18 can be found in Appendix 4. The following is a brief summary of operations from June 23 to August 22, 2005:

June 23

System start up commenced, including sand addition to the gasifier to form pressure seals and to provide for even heat-up of the gasifier, PCD preheating, and operation of the gasifier startup burner.

June 24

Heat up continued, and additional sand was added to increase the solids level in the standpipe. It was found that the PCD rupture disc failsafe tester had opened prematurely. In situ sampling at the PCD outlet was conducted to assess the failsafe performance, and the sample indicated very little particle penetration.

June 25

The startup burner maintained the gasifier temperatures around 1200°F, and sand was added to increase the standpipe level in preparation for coal feed. After maintenance was completed on the clyde coal feed lock hopper system referred to as the original coal feeder, coal feed was started at a rate of approximately 3,600 lb/hr. The gasifier was heated from 1100°F to 1650°F in combustion mode, and then was transitioned into gasification mode. The gasifier pressure and inlet flows were gradually increased.

June 26

The system pressure was increased to reduce the riser velocity. The coal feed rate was increased and sand was added in attempt to increase bed level. Although the level increased, it did not remain constant, and the solids inventory was lost at a rate of about 5.5 in H_2O per hour of standpipe bed height expressed as differential pressure. The original coal feeder transferred smaller amounts of coal than usual, forcing operation at lower feed rates to avoid running out of coal. However, after swapping coal silos, the issue was resolved.

June 27

Sand was added periodically to maintain the standpipe level. Automatic temperature control was attempted in support of automation testing, but the main air compressor surged twice due to the fluctuating air flow demand which caused all air flow rates to be decreased to a minimum and rejected the temperature controls to manual. The loop seal and J-leg velocity controllers were tuned. All of the velocity controllers worked well in automatic. Unfortunately, a glitch in the programmable logic controller (PLC) system caused the temperature control to reject to the local automatic setpoint. After addressing operational issues, automatic temperature control was resumed.

June 28

Fluidization flow was significantly increased in the lower portion of the standpipe to change the fluidization regime. Gasifier solids loss continued, so sand was added to maintain standpipe level.

June 29

Operation of the developmental coal feeder was started, but was problematic due to plugging of the feeder discharge line. Steady operations resumed using only the original coal feeder.

June 30

Developmental coal feeder testing was resumed. Low carbon content in the circulating solids indicated a need for additional tuning of the automatic temperature control scheme.

July 1

The gasifier continued to lose bed material. Automatic temperature control enhancements were successfully completed. Developmental coal feeder testing was continued, and the feeder ran well in automatic mass flow control. The mass flow control setpoint was varied, and the system responded well. The gasifier continued to operate using both coal feeders. The bed level continued to decrease at approximately 5 inH_2O per hour.

July 2

The steam flow rate to the gasifier and aeration flow to J-leg and standpipe were reduced to decrease the inlet velocity to the disengager, increasing its collection efficiency. As a result, the standpipe level increased, and the solids carryover rate to the PCD decreased.

July 3

The gasifier operated well throughout the day in spite of some riser temperature fluctuations. The fluctuations were attributed to variations in both the main air compressor discharge pressure and coal feed rates. The main air compressor discharge pressure swung 2-3 psi, and the flow rates varied about 200 lb/hr. Combinations of these changes caused the gasifier temperature to fluctuate ±25ºF. Sand was added to the gasifier to evaluate the standpipe level stability at higher circulation rates. The standpipe level was then kept constant throughout the day. The temperature profile

July 4

The unit ran well throughout the day with a stable standpipe level.

July 5

A high solids discharge rate test was performed on the CFAD system, obtaining a discharge rate of about 12,000 lb/hr. Automatic temperature control testing was attempted, but the main air compressor discharge pressure experienced a rapid swing when the controller demanded air, so testing was ended to avoid a compressor surge. A standpipe level test was performed by reducing the coal feed rate from the original coal feeder, which had coarser coal particles, to a minimum speed setting and by increasing the developmental coal feeder output, which had smaller particles. Adjustments made to the coal feeder controls caused the developmental coal feeder to trip momentarily. The unit was restarted, but the main air compressor surged twice, causing the air flows to be decreased to a minimum, delaying return to normal operations. Once conditions stabilized, automatic temperature control was resumed.

July 6

Gasifier temperature control changed from automatic to manual when the developmental coal feeder tripped due to solids plugging. The original coal feeder remained the sole feeder. The standpipe level decreased throughout the morning but later stabilized.

July 7

The temperature controller responded well throughout the day. Several set point changes were made to the riser outlet temperature, and the automatic temperature controls ramped up to setpoint with minimal overshoot. The developmental coal feeder was tested using the offline coal feed system to evaluate its reliability using larger coal particles, and the system seemed to handle the particle size increase well.

July 8

Increased air to the lower mixing zone (LMZ) test was performed. An additional 2500 lb/hr of air was injected into the LMZ, anticipating an increase in the solids circulation rate due to the increased gas velocity from the J-leg into the mixing zone. As a result, the high temperature zone shifted from the middle part of the upper mixing zone to the lower portion of the mixing zone and the LMZ. The unit ran steady during the night, and the air to the lower mixing zone test continued. The amount of air entering the LMZ was increased from 4600 to 6800 lb/hr. The ultimate effect of this test was to slightly lower the temperatures in the LMZ and to shift the highest temperature to the lower part of the mixing zone.

July 9

The original coal feeder rate began decreasing which caused slight fluctuations in the temperature profile and a decrease in standpipe level. Later, the coal feed rate was increased and the temperature profile became more stable with the maximum riser temperature below 1800°F.

July 10

The gasifier continued to operate well with the majority of the air fed directly to the LMZ. Due to a problem with the PLC, operation was interrupted by a coal feeder trip, and the startup burner was lit to reheat the system.

July 11

After addressing issues with the coal feeder and the PLC, the system was restored. Gasification operation was resumed with both the developmental and the original feeder coal feeders in service.

July 12

Developmental coal feeder testing continued, and some additional work on tuning the feeder controls was performed. Temperature control testing also continued. The developmental feeder mass flow controller was put in manual and differential pressure control was resumed because the temperature profile continuously worsened. The profile improved after resuming differential pressure control due to a more stable coal feed rate.

July 13

The developmental coal feeder provided the coal feed source. The coal feed rate controller was tuned by varying coal feed demand. The mass flow controller remained in automatic throughout the night. While in automatic temperature control, the riser exit temperature set point was decreased to improve the overall temperature profile. The temperature controller performed well without any temperature swings.

July 14

Parametric tests were performed to support automation development. The tests varied the gasifier pressure by 6 psig, decreased air in the upper mixing zone by 7.5 percent, and adjusted the riser outlet temperature set point by 10°F. With a total feed time of 200 hours, the developmental coal feeder remained the primary feeder and provided 3,500 lb/hr. Once again, the automatic temperature control worked well with the mass coal flow controller. During the evening, the gasifier temperatures began to deviate from their desired values due to low circulation rates. Sand was added to increase the bed level and improve the temperature profile.

July 15

The gasifier ran well with the developmental coal feeder supplying the primary coal feed. Automatic temperature control testing was performed by changing the coal feed rate and the air to the lower mixing zone. A test of CFAD with a high solids withdrawal rate was performed, and the larger coal feed particle size testing was also completed. Air to the lower mixing zone was decreased to lower temperatures.

July 16

The original coal feeder was started as a backup to the developmental coal feeder. Testing of the load simulation controller continued, as well as some additional tuning on the alternate coal feeder controls. The mass flow control was not working well toward the end of the test, forcing the feeder to operate on differential pressure control for the remainder of the night.

July 17

Additional work was performed to improve the developmental coal feeder rate control. The load simulation controller was successfully tested by adjusting the set point several times. An air flow runback occurred due to an upset of the developmental coal feeder controls. All of the automation controllers rejected to manual as required, and the air flow rate controllers decreased to a minimum position. Once normal operating conditions resumed, automatic temperature control was resumed with the developmental coal feeder still in operation.

July 18

Due to operational difficulties resulting from coal with high moisture and larger particle sizes, coal feed from the developmental coal feeder was discontinued, and the feeder was lined up to the offline system. Operations returned to steady state conditions after a successful transition to the original coal feeder. The larger coal feed particle size test continued using the original coal feeder.

July 19

Tuning of the main air compressor commenced, and the compressor was returned to initial tuning parameters in order to evaluate its operation. The automatic temperature control scheme continued to operate well, but rejected to manual as expected when the selected riser temperature deviated 25°F beyond the median riser temperature.

July 20

The main air compressor work was completed. Several upsets occurred due to the compressor work. The original coal feeder experienced some problems due to fine, wet coal, so the developmental coal feeder was operated. The developmental coal feeder supplied the primary coal feed and utilized differential pressure control.

July 21

The original feeder was isolated to clear a plugged vent line, and the developmental coal feeder feed rate was increased to compensate for the loss of flow. Operations returned to steady state after the line was cleared. The PSDF lost power in mid-afternoon, which caused a system trip. All equipment was quickly restarted, and the gasifier was returned to normal operating conditions without lighting the startup burner.

July 22

The unit operated well despite a malfunctioning transformer. Electrical modifications were made and the affected systems were restarted.

July 23

Air distribution tests were started. An increase in air flow through the lower mixing zone resulted in an increase in standpipe level due to shifting inventory. The coal feed rate was reduced in an attempt to stabilize the standpipe level. Only a slight increase in the riser differential pressure occurred relative to the standpipe level. Automatic temperature control worked well.

July 24

The coal feed rate was increased to ensure that air distribution had no impact on the syngas heating value. An increase or decrease in the coal feed rate was directly proportional to an increase or decrease in the CO and H_2 content in the syngas. Increases in the coal feed rate resulted in increases in the lower mixing zone temperatures.

July 25

Steady state conditions were maintained throughout the day.

July 26

An outside engineering consulting firm performed air compliance testing. Upon completion of these tests, conditions were held steady with small increases in steam flow. The standpipe level was reduced.

July 27

Recycle gas was sent to the gasifier with the syngas compressor vendor assisting with the startup of the compressor. The vendor performed a surge testing on the compressor. Gasifier operations were stable, with no sudden changes with the introduction of syngas. The lower heating value increased 8 percent with the introduction of syngas through the fluidization nozzles.

July 28

The recycle gas compressor surge testing continued. The recycle gas compressor was later shutdown due to tuning problems with the controllers.

July 29

The gasifier operated well with a slight reduction in solids inventory. Both coal feeders continued to operate. The recycle gas compressor was started, and recycle gas was sent to the gasifier. The recycle gas compressor reliably supplied syngas for fluidization for the remainder of the night, and did not cause any abnormalities in the gasifier temperature profile.

July 30

A system trip was experienced due to a high riser temperature which resulted from a decrease in steam flow. The automatic controller attempted to compensate but reacted too slowly. The system recovered quickly, although automatic temperature control continued to be problematic.

July 31

The gasifier pressure was increased to lower the riser velocity, which caused the standpipe level to stabilize. The original coal feeder ran at a higher flow rate to allow coarse coal to accumulate in the gasifier.

August 1

The temperature controller was adjusted and placed in automatic. The recycle gas compressor was stable throughout the day, supplying approximately 1500 lb/hr of recycle syngas to the gasifier. Due to the reduced nitrogen supply from the nitrogen vendor, nitrogen flows were minimized throughout the plant. The lack of nitrogen prevented the coal mills from running. The original coal feeder was the only feed source during this time. The dry lower heating value rose to the highest seen during air-blown operations.

August 2

Nitrogen conservation measures were successful in reducing consumption below the on-site nitrogen plant production. Nitrogen levels and gasifier operations were stable throughout the day. Work continued on the steam drum level and pressure controllers. Recycle gas compressor operation was also stable throughout the day.

August 3

The developmental coal feeder discharge line plugged, which caused the gasifier to trip on a low coal feed rate and also tripped the recycle gas compressor. The original coal feeder was started to compensate for the loss in flow. Recycled syngas was restored to the gasifier later that afternoon. The standpipe level increased at a steady rate throughout the night. The recycle gas compressor continued to supply recycled syngas to aeration nozzles that would normally be supplied by nitrogen.

August 4

The unit ran steadily throughout the day.

August 5

The standpipe level remained constant. The recycle gas compressor continued to operate well. Nitrogen conservation was still a major concern.

August 6

The gasifier and PCD remained steady all day, and the standpipe level continued to slowly increase.

August 7-8

Conditions were steady while nitrogen usage remained limited.

August 9

The standpipe level continued to increase. Increasing the riser velocity (as well as increasing the coal feed rate, air flow rate, and steam flow rate) and the gasifier exit temperature kept the level steady.

August 10

Operating at a higher riser velocity began to decrease the standpipe level, which eventually leveled off. The recycle gas compressor continued to run well.

August 12

The developmental coal feeder was prevented from running the remainder of the test campaign due to a plugged conveying line and a failed rotameter.

August 13

Automation parametric testing was performed by varying the LMZ steam flow, gasifier pressure, coal feed rate, J-leg aeration, temperature control set point, varying the air flow rates through the mixing zone, lower mixing zone, and burner leg separately to quantify their effects on gasifier operations.

August 14

The recycle gas compressor was shut down in preparation for scheduled maintenance. Apparently, some of the aeration nozzles to the loop seal downcomer were partially plugged before the compressor stopped. When nitrogen replaced recycle gas through the aeration lines, the loop seal downcomer packed due to loss of flow, and most of the gasifier inventory was lost to the PCD. Approximately 1,000 lbs of solids were conveyed by the CFAD system. Clearing out the plugged nozzles and increasing other loop seal flows allowed the downcomer to unpack, but the nozzles never completely regained their typical flow rates. After about 45 minutes of acceptable solids circulation, the coal feed rate was increased and the gasifier was placed back in automatic temperature control.

August 15

The gasifier operated well, despite the modified flows to the loop seal.

August 16

The lock vessel spheri valve on the original coal feeder began to leak slowly, and the feeder tripped several times. The feeder was restarted each time, but the leak worsened to the point that the feeder would not run. With no coal feeder available, the gasifier pressure was lowered in preparation for lighting the startup burner. The startup burner was lit and the gasifier was kept warm. When the gasifier reached 1,100°F, attempts were made to start the coal feeder, but the motor required maintenance work. Sand was added to the standpipe to increase the gasifier inventory and gasifier temperatures gradually increased. Maintenance work continued on the coal feeder.

August 17

Coal feed to the gasifier resumed. Sand was added again to the gasifier to increase the standpipe level. The main air compressor surged, so slight adjustments were made to the compressor operation. The gasifier pressure, flow rates, and temperatures were increased to normal operating conditions. The coal fed rate was approximately 3,500 lb/hr, and the pressure was 200 psig. Solids circulation through the loop seal was still not as high as expected for these conditions due to only minimal aeration flow through the downcomer section. The coal feed rate periodically increased unexpectedly, causing some temperature swings throughout the gasifier. The automatic temperature controller had difficulties handling the wide fluctuations.

August 18

The standpipe level increased steadily. A main air compressor surge early in the afternoon tripped the system. The system recovered from the upset quickly and previous conditions were restored.

August 19

The recycle gas compressor was started on nitrogen and then transitioned to syngas. The dry syngas heating value increased from 66 to 77 Btu/SCF using recycle gas. The mixing zone temperatures also increased slightly during the same time. The main air compressor surged several times, tripping the unit each time. Lowering the gasifier pressure prevented further surges, and system conditions were restored. The recycle gas compressor tripped during the first surge, and was not restarted.

August 20

The coal transport gas was transitioned from 100 percent nitrogen to 50 percent air and 50 percent nitrogen. Riser temperatures increased up to 35°F. Mixing zone temperatures were generally less affected. The air flow rates through the mixing zone and lower mixing zone were reduced to balance out the temperature profile. The steam flow through the coal feed nozzle shroud was also increased by about 150 lb/hr. Operations were stable throughout the night while running with transport air.

August 21

The cyclone inlet temperature measurement failed, and since this temperature was vital to the automatic temperature control scheme, automatic control was discontinued.

The original coal feeder operated poorly due to fine material. The lock vessel vents plugged, preventing it from depressurizing. During this time, the coal feed rate began to decrease, causing oxygen breakthrough in the PCD. A severe temperature excursion occurred, and the online particulate monitoring device located downstream of the PCD indicated a leak.

August 22

PCD outlet sampling showed very high solids penetration through the PCD, so the system was shut down.

Table 2.3-1 TC18 Typical Operating Conditions for the Transport Gasifier and Particulate Control Device

Transport Gasifier	
Startup Bed Material	~120 micron Sand
Startup Fuel	Powder River Basin Coal
Fuel Type	Powder River Basin Coal
Fuel Particle Size (mmd), micron	120 – 360
Fuel Feed Rate, lb/hr	2,400 – 4,500
Sorbent Type	None
Gasifier Temperature, °F	1765 - 1830
Mixing Zone Pressure, psig	205 – 240
Riser Gas Velocity, fps	33 – 52
Standpipe Level, inH_2O	80 - 180
Gasifier Outlet Gas Flow Rate, lb/hr	15,000 – 25,000
Oxygen/coal mass ratio, lb/lb	0.65 – 1.0
Air/coal mass ratio, lb/lb	2.8 – 4.3
Steam/coal mass ratio, lb/lb	0.1 – 0.5
Particulate Control Device	
PCD Temperature, °F	660 - 750
PCD Inlet Loading, ppmw	12,300 - 23,600
PCD Outlet Loading, ppmw	<0.1
PCD Pressure Drop, inH_2O	100 - 160
Number of Filter Elements	72
Filter Element Type (number used)	Iron Aluminide (69), Titanium (3)
Filtration Area, ft^2	204.5
Face Velocity, ft/min	3 - 5
Pulse Valve Open Time, sec	0.2
Pulse Cycle Time, min	5
Pulse Pressure	250 psi 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 the refractory, and no major problems were observed. The mixing zone was relatively clean. There were a few small deposits in the area surrounding two of the coal feed nozzles. These appeared to be left from previous runs. The region below the coal feed nozzles was relatively clean. The refractory in this region showed wear.

The riser was in good condition. A few of the longer thermowells could be seen, and some showed degradation. The riser crossover was filled with loose solids. The disengager was in good condition, though the walls showed wear.

A notch was again found in the cyclone inlet, but it did not appear to be as extensive as seen previously and described in the TC17 Topical Report. The roof of the cyclone was in good condition, although there were a few deposits seen on the walls.

The loop seal downcomer had many soft deposits on the area from the wall to nearly the top of the downcomer. Some of the deposits from the walls may have fallen to the bottom of the downcomer.

Although the standpipe was in fair condition, there was one apparent agglomeration found near the middle section.

The burner leg was also inspected during the outage and it was bridged over with refractory pieces that were approximately two inches in diameter. The leg was drained of bed material after dislodging all the refractory pieces. About eight feet of refractory pipe towards the upper mixing zone could be seen from the bottom, and no deposits or major cracks were noted on the refractory.

The primary gas cooler tube sheet was covered mostly in dust, and a few of the tubes were plugged.

2.4.2 Particulate Control Device

The PCD was opened after shutdown. Inspection revealed that the three titanium filter elements were severely damaged and all iron aluminide filter elements were intact. Figure 2.4-1 shows two of the damaged titanium elements. Apparently, the porous media of the titanium elements was disintegrated. The material was delaminated with a distinct layer on both the inside and outside wall surfaces. As shown in Figure 2.4-2, some spots of molten material were found on the element surface, possibly formed from molten ash or other materials.

A Pall fuse failsafe located above the disconnected titanium filter element was ruptured, and is shown in Figure 2.4-3. Inspection of the failsafe indicated that it failed under high dynamic pressure. This broken failsafe and the corresponding failed filter element left an exposed hole in the bottom plenum, allowing particulate to pass through to the outlet. This hole could solely account for the significant particulate leaking after the thermal excursion.

All filter elements were removed from the PCD and flow tested. Flow tests were conducted using air at ambient temperature and pressure. For reinstallation, the acceptable pressure drop limit for filter elements and failsafes is 10 inH_2O at a face velocity of 3 ft/min. Of the 69 iron aluminide filter elements installed, only 4 were below 10 inH_2O at a face velocity of 3 ft/min. Sixteen of the remaining elements were above 50 inH_2O, the upper limit of measurement. The filter elements were pressure washed to decrease their flow resistance and were flow tested again. Subsequent test results revealed that 44 were acceptable, but 7 were between 11 and 18 inH_2O at a face velocity of 3 ft/min. Six filter elements with internal fuses were higher, with pressure drops between 25 and 45 inH_2O. Only seven filter elements were still above 50 inH_2O after being pressure washed.

All failsafes were removed and flow tested. Four Pall fuses were broken during removal, apparently due to weakness in the weld structure. The acceptable range for failsafes is also based on a pressure drop of 10 inH_2O at a face velocity of 3 ft/min. Failsafe flow tests revealed that 5 Pall fuses were below 10 inH_2O at a face velocity of 3 ft/min. Seventeen of the remaining failsafes were above 50 inH_2O. The failsafes were ultrasonically cleaned and flow tested again. Flow test results showed that 21 failsafes were acceptable for reinstallation after being cleaned.

2.4.3 Other Systems

2.4.3.1 Recycle Gas Compressor

The recycle gas compressor and associated piping were inspected upon the completion of the test run. Water was found in the aeration lines which had condensed in the headers. This may have occurred from the steam used to purge the line following a compressor trip. Apparently, the steam temperature had dropped below the dew point, and steam condensed in the line.

2.4.3.2 Continuous Fine Ash Depressurization (CFAD) System

The CFAD system surge vessel was also inspected. The cooling bundles were clean. There was no sign of erosion seen on the cooling coils. Some fines were found in the bottom of the vessel and were removed to avoid caking. The settled fines may have come from a funnel collapse after shutdown. Further inspection of the key components of the CFAD system showed no unusual fines buildup or erosion anywhere in the system.

The newly designed cushioning bend was inspected, and there was no visible wear or erosion inside the bend or in the outlet piping. A conventional cross and a valve located on a tee on the line to ash silo were eroded. Several diaphragm-type pressure transducers in the solids line were inspected and found to be in their original condition with no signs of erosion. In addition, there were no signs of fines packing around the transducers.

The safety-interlocked ball valve at the outlet of the CFAD system was exercised a few hundred times during operation to test its durability. The valve was inspected, and the ball surface was still in good condition with no signs of erosion.

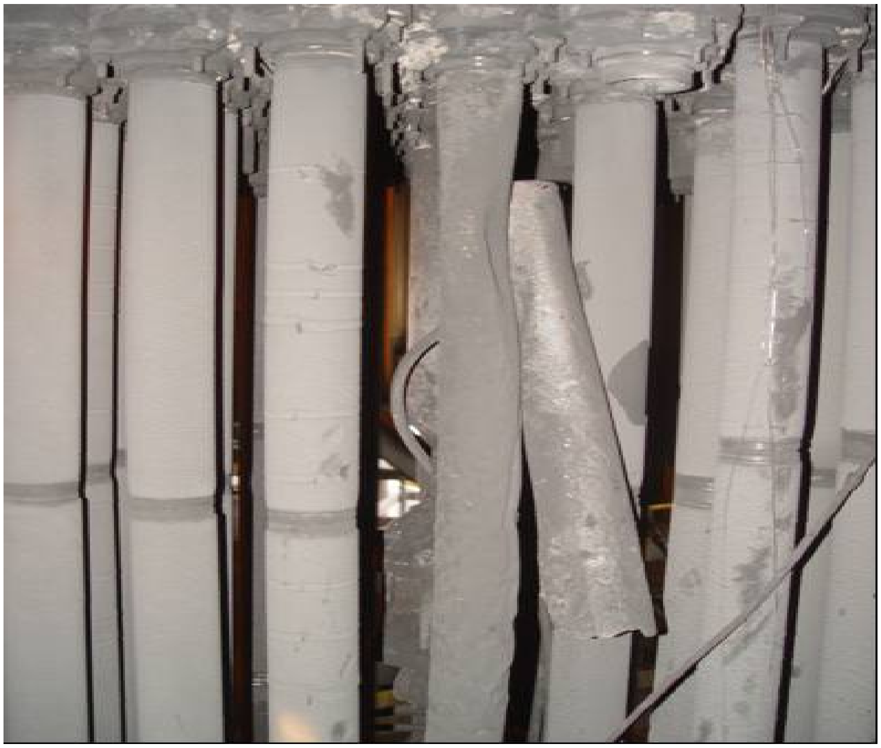

Figure 2.4-1   Titanium Filter Elements on Bottom Plenum

Figure 2.4-2   Titanium Filter Element with Spots of Molten Material

Figure 2.4-3 Ruptured Pall Fuse Failsafe

3.0 TEST CAMPAIGN PERFORMANCE

3.1 TRANSPORT GASIFIER PERFORMANCE

3.1.1 Overview

Test campaign TC18 consisted of 95 periods of steady state operation, which are given in Table 3.1-1. The operating periods had a cumulative time of about 574 hours, which was about 37 percent of the total TC18 on-coal operation time. Air was the oxidant for the entire test run. For the first time, recycle gas was available for use as aeration gas in the gasifier J-leg, loop seal, and standpipe. Operating periods TC18-58 through TC18-87 featured recycle gas use, while the remaining periods used nitrogen for gasifier aeration. No sorbent was used during TC18.

Table 3.1-2 lists the TC18 operating conditions, including coal feed rate, riser outlet temperature, system pressure, PCD inlet temperature, PCD solids rate, air rate, oxygen rate, syngas rate, steam rate, and nitrogen rate during steady state operating periods. The coal feed rate ranged from 2,400 to 4,500 lb/hr. Riser outlet temperatures were between 1,634 and 1,725°F, and the system pressure ranged from 205 to 240 psig.

3.1.2 Gas Composition

During the test run, the Transport Gasifier and syngas combustor outlet gas analyzers were continuously monitored and recorded by the Plant Information system. Thirty-eight in situ samples of syngas were taken during PCD outlet sampling and measured for moisture content. Plotted on Figure 3.1-1 are data from the syngas moisture analyzer AI475H, syngas moisture analyzer AI2530H, the in situ H_2O concentrations, and the moisture content based on a correlation of temperature and in situ data. The H_2O in situ concentrations and syngas moisture analyzer AI2530H concentrations were not necessarily measured during steady state operating periods.

The selected moisture concentrations for the operating periods are given in Table 3.1-3. Most of these values are from the moisture analyzer. Whenever the moisture analyzers appeared to be in error, as in periods TC18-1,-2, and TC18-9 through TC18-30, the moisture contents are from the correlation. The moisture content was generally between 7 and 15 percent, with the higher contents occurring during periods of high steam flow rate.

Based on these moisture concentrations, the estimated wet syngas compositions for the TC18 operating periods are given in Table 3.1-3 and shown on Figure 3.1-2. Also shown in Table 3.1-3 are the syngas molecular weights for each operating period. The concentrations of all compounds except nitrogen were generally higher during recycle gas use due to decreased nitrogen dilution. Some overlap in the data exist, since other factors such as the coal feed rate, air rate, temperature, and pressure also affect the gas composition.

Concentrations of CO, H_2, and CH_4, and C_2^+ are important in that these compounds provide the heating value for the syngas. The CO concentration ranged from 6.0 to 10.7 percent when recycle gas was not in use, except for TC18-6 when the CO content was only 5.2 percent due to a low coal feed rate. The CO content was between 9.1 and 10.7 percent during the recycle gas periods, except for TC18-79, when the CO concentration was 6.7 percent due to a high syngas moisture content which caused CO to shift to CO_2 via the water gas shift reaction. The H_2 concentration fluctuated between 5.7 and 8.3 percent during the periods without recycle gas and between 7.0 and 8.8 percent with recycle gas. The CH_4 concentration was between 0.8 and 2.0 percent without recycle gas and between 1.0 and 1.9 with recycle gas. The C_2^+ concentration was negligible for all of the test periods with or without recycle gas.

CO_2 and nitrogen are also present in the syngas, though they do not contribute to the heating value. The CO_2 concentration ranged from 7.9 to 9.7 percent without recycle gas and from 8.7 to 10.5 percent with recycle gas. The nitrogen content ranged from 59.6 and 69.3 without recycle gas and from 58.4 to 65.4 with recycle gas.

The wet syngas molecular weight and nitrogen concentration are plotted on Figure 3.1-3. The molecular weights ranged between 25.9 and 27.0 lb/lb-mol. The variation in molecular weight was mostly affected by the moisture content rather than the use of recycle gas.

The CO/CO_2 ratios were calculated from the gas data for each operating period, and are listed in Table 3.1-3. The CO/CO_2 ratio varied from 0.58 to 1.29, with the lower ratios occurring during periods of low coal feed rate.

The main sulfur species in syngas are hydrogen sulfide (H_2S) and carbonyl sulfide (COS) with other sulfur compounds, such as CS_2, present in small quantities. The total reduced sulfur, TRS, is the sum of the compositions of all sulfur species in the syngas. When the syngas is combusted, the sulfur compounds are converted to SO_2. The measured wet H_2S syngas concentration and the SO_2 concentration measured at the syngas combustor outlet are plotted on Figure 3.1-4. No SO_2 data are available for the first eight periods and TC18-38 since the SO_2 analyzer was not in service for these time periods. Also plotted on the graph is the calculated syngas TRS concentration derived from the SO_2 content at the syngas combustor outlet. The wet H_2S concentration and the syngas TRS concentration are also listed in Table 3.1-4. Again, note that the TRS data are not available for the first eight periods and TC18-38 since the SO_2 analyzer was not in service. The AI419 analyzers measure the gas composition on a dry basis, so the values from AI419J were corrected to include moisture. The H_2S analyzer AI419J was out of service from TC18-1 to TC18-35 and from TC18-88 to TC18-95. The syngas combustor SO_2 analyzer, AI476N, measures the total sulfur emissions leaving the system.

The TRS concentration ranged between 196 and 462 ppm during the test run. The use of recycle gas did not appear to increase the TRS. No sorbent feed occurred during the test run; therefore, the only calcium present in the gasifier to capture sulfur came from the PRB coal ash. In previous PRB test campaigns, the use of sorbent had little effect on syngas sulfur concentrations.

The volatile nitrogen released from the coal into the syngas is mostly in the form of ammonia. HCN is also present. When these species are combusted in the syngas combustor, they produce NOx. The ammonia concentrations are given in Table 3.1-4 as measured by ammonia analyzer AI2530Q. The ammonia concentration ranged from 1,600 to 2,400 ppm and appeared independent of recycle gas use.

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 3.1-5. All raw LHVs are on a wet basis.

The lower heating value was calculated using the formula:

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

Without recycle gas, the LHV typically ranged from 42 to 70 Btu/SCF, with one exception of 73 Btu/SCF during a period of high coal feed. The LHV ranged from 55 to 74 Btu/SCF during recycle gas use. The periods with recycle gas averaged 66 Btu/SCF, while the periods without recycle gas averaged 61 Btu/SCF.

Some overlap between the two sets of data exists due to other factors influencing the LHV, such as coal feed rate and steam flow rate. Past test runs have indicated that the most significant impact on LHV are the coal and steam feed rates. As coal rate increases, the syngas production rate increases while the aeration and instrument purge nitrogen flow rates remain constant. Therefore, the nitrogen constituent of the syngas decreases (less dilution), and the syngas LHV increases. Increasing the steam flow decreases the LHV by diluting the syngas with moisture.

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 PCD backpulsing.
- A lower heat loss per pound coal gasified because of a lower gasifier surface area to volume ratio.
- Less instrumentation and thus lower nitrogen flow used for instrument purging.
- Moisture removal from the syngas in the cold gas cleanup process.

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

The commercially projected gas turbine inlet LHV and adjusted syngas composition data for each operating period are given in Table 3.1-5. The projected LHV was between 105 and 153 Btu/SCF. The utilization of recycle gas at the PSDF does not affect the projected values, since the calculation adjusts the LHV to include recycle gas for all periods.

3.1.4 Gasifier Solids Analyses

During TC18, the solid samples were taken from the following locations:

- The coal feed systems (FD0210 and FD0200).
- The Transport Gasifier standpipe (coarse gasification ash).
- The Transport Gasifier loop seal downcomer.
- The continuous fine ash depressurization system (CFAD or FD0540).

In situ solids samples were also collected from the PCD inlet. All solid samples were analyzed for chemical composition and particle size.

Table 3.1-6 gives the average proximate, ultimate, and ash mineral analyses of the PRB coal, as well as the average coal molar calcium to sulfur ratios (Ca/S) and the standard deviation for the samples analyzed as sampled from the original coal feeder, FD0210, and the developmental coal feeder, FD0200. The fuel carbon and moisture contents are shown in Figure 3.1-6 for both feeders. The carbon content of the coal remained relatively constant during TC18, varying from 54.3 to 60.6 percent. The moisture content fluctuated between 15.2 and 21.2 percent.

Figure 3.1-7 shows the coal sulfur and ash as sampled from both coal feeders during TC18. The sulfur level remained between 0.20 and 0.37 weight percent. The ash content was typical for PRB coal, at between 4.93 and 7.24 weight percent.

The lower heating value (LHV) of the coal is given on Figure 3.1-8 with the TC18 average value given in Table 3.1-6. The LHV is not measured directly, but is calculated by subtracting the heat of vaporization of the coal moisture from the HHV. The coal HHV is determined using a bomb calorimeter. The calorimeter condenses all the coal combustion moisture as liquid water. 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 8,900 to 9,600 Btu/SCF during the test run.

During gasification, the Transport Gasifier produces both fine solids collected in the PCD and removed by the CFAD system and coarse solids removed via the standpipe screw cooler. Analyses of the chemical compositions of these solid compounds are based on the following assumptions:

1. All carbon dioxide measured is from $CaCO_3$.
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 reducing conditions in the Transport 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.

Both elemental sulfur (ultimate analysis) and ash inerts sulfur contents were measured. It is assumed that no FeS forms in the Transport Gasifier and all of the sulfur in the standpipe and PCD solids is present as CaS. Thermodynamically some FeS formation is possible, but most of the captured sulfur should be in the form of CaS due to the larger concentration of calcium than iron in the system.

Table 3.1-7 gives the TC18 standpipe solids analyses. Only three samples were chemically analyzed, since the standpipe solids sampler was out of service for the majority of the test run. The standpipe solids re-circulate through the mixing zone, riser, and standpipe. Typically, the properties of these solids change slowly with time. Unfortunately, since the only samples available are from the beginning of the test run, all three contain mostly sand, the startup material. The silica contents were all greater than 85 weight percent. Figure 3.1-9 shows the standpipe SiO_2, CaO, and Al_2O_3 contents versus run time, showing the high silica concentration. The data show a very slight decrease in silica and increase in alumina over time as is typical for a test run.

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 inaccurate measurement because the value comes from a difference of two small values that are nearly equal. The standpipe organic carbon content, shown in Table 3.1-7, was negligible for all three sample due to the high amount of sand present just after startup.

The standpipe solids CaS content was negligible for all of the standpipe samples. The calcium in the PRB coal ash captured only a minimal amount of sulfur which is consistent with previous PRB test data. The standpipe $CaCO_3$ was between 1.2 and 1.5 percent for the three samples available in TC18. About 68 percent of the calcium in these standpipe solids was calcined to CaO.

Table 3.1-8 lists the solids sample analysis for the loop seal. 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.

Figure 3.1-10 shows the CaO, SiO_2, and Al_2O_3 contents of the loop seal samples. The loop seal SiO_2 content were at 82.1 percent early in the test run and declined to 43.8 percent as coal ash replaced the silica. The silica content increased to 51.8 percent during the sand addition that took place near hour 800, and then tapered back to under 45 percent. The sand addition during the short outage at hour 1157 caused the silica content to increase again. CaO and Al_2O_3 contents both rose steadily during the test run, except during major sand additions. The CaO and Al_2O_3 content reached values as high as 20.3 and 21.4 percent, respectively. The MgO, Fe_2O_3, and other inerts contents are not plotted, but they follow the same trend as the Al_2O_3, since they change as spent solids replace the original bed material. The loop seal solids CaS content was mostly negligible for all of the samples.

Figure 3-1.11 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 is usually higher in the loop seal than in the standpipe because the cyclone captures a greater percentage of the smaller carbon particles than the disengager. In TC18, the values ranged from 0.2 to 8.9 weight percent. The loop seal $CaCO_3$ fluctuated from 0.5 to 5.2 percent.

The complete solids analysis as well as organic carbon content for the PCD solids samples is given in Table 3.1-9. In situ PCD inlet particulate solid samples were also analyzed. Figure 3.1-12 plots the organic carbon for the PCD solids sampled from the continuous fine ash depressurization system (FD0540) as well as for the in situ samples. The in situ sample data agree moderately well with the FD0540 sample data. The FD0540 sample carbon content ranged from 18.4 to 52.7 percent during the test run.

Figure 3.1-13 and Table 3.1-9 also show the amounts of SiO_2 and Al_2O_3 in the PCD solids as sampled from FD0540. The in situ solids concentrations for SiO_2 and Al_2O_3 are also plotted on Figure 3.1-13. The SiO_2 PCD solids concentrations are a function of the efficiency of the disengager and cyclone as well as the SiO_2 concentration of circulating solids in the gasifier. The SiO_2 PCD solids concentration began the test run between 34 and 47 percent before dropping to between 13 and 35 percent for the remainder of the test run. The drop in SiO_2 content a few days after startup is typical, since gasification solids replace the startup sand.

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 fairly low, between 4 and 15 percent, for the majority of the test run.

Figure 3.1-14 shows the calcium carbonate and calcium sulfide concentrations in the PCD solids as sampled from FD0540. The concentrations for CaO, CaS, and $CaCO_3$ are also listed in Table 3.1-9. Also plotted on Figure 3.1-14 are the calcium carbonate and calcium sulfide concentrations for the in situ solids samples. All of the in situ sample CaS concentrations agreed well with the FD0540 solids calcium sulfide concentrations. The in situ sample calcium carbonate contents did not agree with the FD0540 samples. The PCD solids calcium carbonate concentration fluctuated from less than 4.1 percent to 9.2 percent. Since no sorbent feed occurred during the test run, all of the PCD solids calcium came from the PRB coal ash. The PCD solids CaS concentration was 1.9 percent or less for the entire test run, indicating very low sulfur capture.

The PCD solids calcination is defined as:

$$Percent\ Calcination = \frac{mol\%CaO}{mol\%CaO + mol\%CaCO_3}$$

The PCD solids calcination data are plotted on Figure 3.1-15. The PCD solids calcination ranged from 39 to 81 percent, as was typical in previous PRB runs.

The solids sulfation is defined as:

$$Percent\ Sulfation = \frac{mol\%CaS}{mol\%CaO + mol\%CaCO_3 + mol\%CaS}$$

The PCD solids sulfation is plotted on Figure 3.1-15 with the PCD solids calcination. The PCD solids sulfation varied between 1 and 15 percent during the test run. These values indicate t minimal sulfur capture.

Solids Sample Comparison

The following comparison of the analysis of the loop seal downcomer solids and the PCD solids shows how the solids compositions change throughout the process. The comparison does not include the standpipe solids, since the standpipe sampler went out of service early in the test campaign, and only six samples were taken, three of which were chemically analyzed.

Figure 3.1-16 compares the organic carbon content of the loop seal and PCD solids samples. The PCD solids carbon content ranged from 18.4 to 52.7 percent, with high variability. The loop seal organic carbon content ranged between 0.2 and 8.9 percent. Also, the loop seal variability much lower in TC18 than in previous test runs. These data seem to indicate that the carbon is contained in small particles which are only partially captured by the cyclone.

Figure 3.1-17 compares the calcium concentration between the loop seal and the PCD solids samples. The calcium content was higher for the PCD solids at the beginning of the test run, where it ranged from The calcium content of the PCD solids typically remained in range of 6 to 11 weight percent for the entire test run. The loop seal solids calcium content began the test run at about 3 weight percent, but increased to over 15 percent as the run progressed. Whenever sand was added, the loop seal solids calcium content declined sharply. The higher loop seal solids calcium content indicated that the calcium was concentrated in the particles captured by the cyclone. (Note that the figure refers to the total calcium which is distributed between the compounds CaO, $CaCO_3$, and CaS.)

The silica entering the process primarily remains in the gasifier, since the sand particle size is greater than that of the loop seal solids. Over time, the sand attrites, eventually passing through the cyclone before being collected by the PCD, resulting in the replacement of sand with gasification ash in the gasifier bed. Figure 3.1-18 shows the silica content of the gasifier and PCD solids. During the two major sand additions in TC18 near hour 800 and hour 1,157, the silica content of the loop seal solids spiked. The loop seal solids appear to have reached a steady state value, however, around 43 percent, before the sand addition around hour 800.

Solids Particle Size

The TC18 Sauter mean diameter (SMD) particle sizes of the coal sampled from FD0210 and FD0200 are plotted on Figure 3.1-19, while the mass median diameter (MMD) particle sizes are shown on Figure 3.1-20. The PRB coal SMD particle size of the coal from FD0200 averaged 153 microns and ranged from just under 99 to 238 microns. The coal from FD0210 had an average SMD of 166 microns and ranged from 98 to 238 microns. The average MMD for the FD0200 feeder was 238 microns, with a standard deviation of 32 microns. The average MMD for the FD0210 feeder was 235 microns, with a standard deviation of 38 microns.

Figure 3.1-21 plots the coal feed percent above 1,180 microns (coarse particles), while Figure 3.1-22 plots the coal feed percent below 45 microns (fines). A large amount of 1,180 micron particles increases the difference between the SMD and the MMD, because the SMD is a surface area average. Therefore, the larger particles with less surface area per pound have a weaker 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 FD0200 was 0.5 percent with a standard deviation of 1.4, while the average percent above 1,180 microns for FD0210 was 2.0 percent with a standard deviation of 1.4. The percent above 1,180 microns remained low during the entire test run, usually less than 5 percent.

In past testing, a high fines content in the feed coal resulted in an increased number of coal feeder outages due to the packing of coal fines in the coal feed system lock vessel. These problems did not occur to a great extent in TC18. The average percent below 45 microns for FD0200 was 14.6 percent with a standard deviation of 3.6 percent, higher than typical for PRB test runs. The average percent below 45 microns for FD0210 was 11.6 percent with a standard deviation of 4.3 percent, also higher than typical for PRB test runs. The values spiked occasionally to over 20 percent. Even during these periods, the lock vessels did not experience any packing.

The TC18 standpipe solids particle sizes are given in Figure 3.1-23. Since the standpipe samples only available early in the test campaign, the sample particle sizes remained constant because not enough time elapsed for spent solids to replace the startup sand. The average particle size of the standpipe solids was around 141 microns MMD and 162 microns SMD, values close to that of sand at 150 microns. For some of the previous test campaigns, the gasifier circulating solids achieved a steady particle size, typically between 165 and 205 microns SMD as shown in Table 3.1-10. For tests that reached steady state, the standpipe particle size slowly increased asymptotically to reach the steady state value.

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 28 to 174 microns, while the MMD varied from 19 to 166 microns.

Figure 3.1-25 plots the SMD and MMD for the PCD solids sampled from the fines removal systems as well as the in situ samples collected at the PCD inlet. The in situ solids particle sizes agreed fairly well with the particle size of the solids collected from the hoppers. Sand addition is likely the reason for any disagreements. 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. A few of the hopper samples were often taken near times when sand had been added.

The PCD solids SMD started TC18 around 11 microns, and ranged from 6 to 16 microns for the remainder of the run. The MMD was about 7 microns larger than the SMD for most of the samples and followed the same trends as the SMD particle size. The TC18 PCD solids particle size was consistent with the particle size of previous PRB test campaigns as shown in Table 3.1-10.

Standpipe, Loop Seal, and PCD Solids Bulk Densities

The standpipe, loop seal, PCD in situ, and FD0540 solids bulk densities are given in Figure 3.1-26. The average bulk density of the six standpipe solids was 84 lb/ft^3, a value close to that of sand. Slightly lower, the data for the loop seal solids averaged 66 lb/ft^3. Since the disengager captured most of the sand particles, the loop seal solids varied more with the density of the coal fed. The

PCD solids had the lowest average values, around 15 lb/ft^3 for the material removed from the fines systems. All of the in situ PCD data points agreed well with the fines systems sample data.

The minimum standpipe solids densities for past PSDF gasification test campaigns are shown in Table 3.1-10. 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, CO_2, CH_4, C_2H_6, and higher hydrocarbons. The carbon conversion can be calculated by dividing the carbon content in the syngas by the total carbon exiting the gasifier (from both solid and gas streams). Table 3.1-11 gives the carbon conversions for the test periods, while Figure 3.1-27 shows the carbon conversion versus time. The carbon conversion ranged from 88 to 97 percent during the test run. The use of recycle gas did not appear to affect the carbon conversion. The carbon conversion was typically lower during periods in which the PCD solids rate was higher. During these periods, more fine carbon particles exited the gasifier without being recycled, thus lowering the carbon conversion.

The average carbon conversions of Powder River Basin, Hiawatha bituminous, Falkirk lignite, Freedom lignite, and Illinois Basin coal for both air and oxygen blown operation are compared on Figure 3.1-28. These data came from test campaigns TC06 through TC13 and TC15. The graph does not include TC14 data, since the poor performance of the solids collection systems 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. Air blown operation yielded a slightly higher carbon conversion than oxygen blown operation except for operation with PRB coal. However, the difference in carbon conversion with PRB for air blown versus oxygen blown operation was not significant.

Falkirk lignite had the highest average carbon conversion of the five coals tested. PRB and Freedom lignite had about the same average carbon conversion, while Hiawatha bituminous, followed by Illinois Basin bituminous, had the lowest average carbon conversion. 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 total 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 steam turbine.

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 cold gasification efficiency is plotted in Figure 3.1-29 and is listed in Table 3.1-11. During TC18, the values ranged from 45.8 to 57.0 percent.

The hot gasification efficiency assumes that the sensible heat of the syngas can be recovered in a heat recovery steam generator, thus the hot gasification efficiency is 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 recycled syngas and the sensible heats of the air and steam. The hot gasification efficiency is plotted in Figure 3.1-30 and shown in Table 3.1-11. The efficiency boundary for the values found in Table 3.1-11 and Figure 3.1-30 is the gasifier itself, not including any downstream equipment.

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 hot gas efficiencies were between 77.8 and 86.0 percent, with the periods of lowest efficiency occurring during the periods low coal feed rate (TC18-6, TC18-41, TC18-42).

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 10 percent of the feed energy, while the total energy of the PCD solids was from 2 to 10 percent of the feed energy (the higher numbers occurring during the periods of high PCD carryover rate).

A commercial gasifier will be more efficient than the PSDF gasifier due to lower relative nitrogen use, additional use of recycle gas, and lower heat losses. The heat loss as a percentage of energy fed will be much smaller in a commercially sized 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 3.1-11.

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. The corrected cold gas efficiencies ranged from 64.7 to 73.5 percent and averaged 69.9 percent. The commercially projected efficiencies were higher than the observed cold gasification efficiencies by about 17 percent.

Table 3.1-1 Operating Periods
(Page 1 of 3)

Operating Period[1,2]	Start Time	End Time	Duration Hours	Operating Period Average Time	Operating Period Relative Hours
TC18-1	6/27/05 5:45	6/27/05 9:45	4:00	6/27/05 7:45	31
TC18-2	6/27/05 15:15	6/27/05 22:45	7:30	6/27/05 19:00	42
TC18-3	6/28/05 5:00	6/28/05 9:15	4:15	6/28/05 7:07	54
TC18-4	6/28/05 12:00	6/28/05 16:00	4:00	6/28/05 14:00	61
TC18-5	6/30/05 2:15	6/30/05 7:15	5:00	6/30/05 4:45	100
TC18-6	7/1/05 3:15	7/1/05 7:45	4:30	7/1/05 5:30	125
TC18-7	7/1/05 8:45	7/1/05 15:30	6:45	7/1/05 12:07	131
TC18-8	7/2/05 0:15	7/2/05 5:00	4:45	7/2/05 2:37	146
TC18-9	7/2/05 19:45	7/3/05 4:30	8:45	7/3/05 0:07	167
TC18-10	7/3/05 15:30	7/3/05 19:45	4:15	7/3/05 17:37	185
TC18-11	7/3/05 20:00	7/4/05 5:45	9:45	7/4/05 0:52	192
TC18-12	7/4/05 8:15	7/4/05 12:15	4:00	7/4/05 10:15	201
TC18-13	7/4/05 15:30	7/4/05 20:15	4:45	7/4/05 17:52	209
TC18-14	7/4/05 23:00	7/5/05 11:00	12:00	7/5/05 5:00	220
TC18-15	7/5/05 11:15	7/5/05 15:30	4:15	7/5/05 13:22	228
TC18-16	7/6/05 10:15	7/6/05 17:30	7:15	7/6/05 13:52	253
TC18-17	7/6/05 17:30	7/7/05 6:15	12:45	7/6/05 23:52	263
TC18-18	7/7/05 21:00	7/8/05 4:00	7:00	7/8/05 0:30	287
TC18-19	7/8/05 7:00	7/8/05 11:00	4:00	7/8/05 9:00	296
TC18-20	7/8/05 14:30	7/8/05 22:30	8:00	7/8/05 18:30	305
TC18-21	7/9/05 12:00	7/9/05 18:15	6:15	7/9/05 15:07	326
TC18-22	7/9/05 19:15	7/10/05 1:15	6:00	7/9/05 22:15	333
TC18-23	7/10/05 1:30	7/10/05 6:15	4:45	7/10/05 3:52	339
TC18-24	7/10/05 7:15	7/10/05 13:00	5:45	7/10/05 10:07	345
TC18-25	7/10/05 15:15	7/10/05 22:30	7:15	7/10/05 18:52	354
TC18-26	7/12/05 4:30	7/12/05 8:30	4:00	7/12/05 6:30	374
TC18-27	7/12/05 20:15	7/13/05 1:45	5:30	7/12/05 23:00	391
TC18-28	7/13/05 3:30	7/13/05 10:00	6:30	7/13/05 6:45	399
TC18-29	7/13/05 18:30	7/13/05 23:00	4:30	7/13/05 20:45	413
TC18-30	7/13/05 23:00	7/14/05 3:00	4:00	7/14/05 1:00	417
TC18-31	7/14/05 3:00	7/14/05 9:30	6:30	7/14/05 6:15	422
TC18-32	7/14/05 16:00	7/14/05 20:15	4:15	7/14/05 18:07	434
TC18-33	7/15/05 17:30	7/15/05 23:45	6:15	7/15/05 20:37	461
TC18-34	7/16/05 2:30	7/16/05 7:00	4:30	7/16/05 4:45	469
TC18-35	7/16/05 23:30	7/17/05 3:45	4:15	7/17/05 1:37	490

Table 3.1-1 Operating Periods
(Page 2 of 3)

Operating Period[1,2]	Start Time	End Time	Duration Hours	Operating Period Average Time	Operating Period Relative Hours
TC18-36	7/17/05 18:45	7/18/05 0:15	5:30	7/17/05 21:30	509
TC18-37	7/18/05 16:00	7/18/05 20:15	4:15	7/18/05 18:07	530
TC18-38	7/19/05 3:15	7/19/05 7:30	4:15	7/19/05 5:22	541
TC18-39	7/19/05 16:15	7/20/05 0:30	8:15	7/19/05 20:22	556
TC18-40	7/21/05 1:15	7/21/05 5:45	4:30	7/21/05 3:30	588
TC18-41	7/21/05 22:15	7/22/05 8:15	10:00	7/22/05 3:15	611
TC18-42	7/22/05 9:30	7/22/05 15:30	6:00	7/22/05 12:30	620
TC18-43	7/22/05 21:15	7/23/05 2:30	5:15	7/22/05 23:52	632
TC18-44	7/23/05 3:45	7/23/05 8:00	4:15	7/23/05 5:52	638
TC18-45	7/23/05 9:30	7/23/05 16:30	7:00	7/23/05 13:00	645
TC18-46	7/23/05 17:15	7/24/05 0:15	7:00	7/23/05 20:45	653
TC18-47	7/24/05 0:15	7/24/05 5:45	5:30	7/24/05 3:00	659
TC18-48	7/24/05 21:15	7/25/05 4:15	7:00	7/25/05 0:45	681
TC18-49	7/25/05 4:15	7/25/05 13:45	9:30	7/25/05 9:00	689
TC18-50	7/25/05 13:45	7/26/05 0:15	10:30	7/25/05 19:00	699
TC18-51	7/26/05 0:30	7/26/05 7:00	6:30	7/26/05 3:45	708
TC18-52	7/26/05 7:00	7/26/05 18:30	11:30	7/26/05 12:45	717
TC18-53	7/26/05 21:45	7/27/05 4:00	6:15	7/27/05 0:52	729
TC18-54	7/27/05 4:45	7/27/05 10:00	5:15	7/27/05 7:22	735
TC18-55	7/28/05 0:15	7/28/05 4:30	4:15	7/28/05 2:22	754
TC18-56	7/28/05 10:45	7/28/05 15:00	4:15	7/28/05 12:52	765
TC18-57	7/29/05 4:00	7/29/05 9:30	5:30	7/29/05 6:45	783
TC18-58	7/29/05 22:45	7/30/05 3:30	4:45	7/30/05 1:07	801
TC18-59	7/30/05 20:15	7/31/05 6:30	10:15	7/31/05 1:22	825
TC18-60	7/31/05 7:30	7/31/05 12:30	5:00	7/31/05 10:00	834
TC18-61	7/31/05 22:30	8/1/05 4:45	6:15	8/1/05 1:37	850
TC18-62	8/1/05 6:45	8/1/05 12:15	5:30	8/1/05 9:30	857
TC18-63	8/1/05 13:00	8/1/05 17:30	4:30	8/1/05 15:15	863
TC18-64	8/3/05 18:30	8/4/05 1:30	7:00	8/3/05 22:00	918
TC18-65	8/4/05 2:15	8/4/05 6:15	4:00	8/4/05 4:15	924
TC18-66	8/4/05 7:15	8/4/05 16:30	9:15	8/4/05 11:52	932
TC18-67	8/4/05 17:00	8/4/05 21:30	4:30	8/4/05 19:15	939
TC18-68	8/5/05 8:30	8/5/05 12:30	4:00	8/5/05 10:30	954
TC18-69	8/5/05 13:45	8/5/05 18:00	4:15	8/5/05 15:52	960
TC18-70	8/5/05 19:00	8/5/05 23:00	4:00	8/5/05 21:00	965

Table 3.1-1 Operating Periods
(Page 3 of 3)

Operating Period[1,2]	Start Time	End Time	Duration Hours	Operating Period Average Time	Operating Period Relative Hours
TC18-71	8/6/05 7:30	8/6/05 13:45	6:15	8/6/05 10:37	979
TC18-72	8/6/05 17:45	8/6/05 22:00	4:15	8/6/05 19:52	988
TC18-73	8/6/05 22:00	8/7/05 3:15	5:15	8/7/05 0:37	993
TC18-74	8/7/05 6:15	8/7/05 10:15	4:00	8/7/05 8:15	1000
TC18-75	8/7/05 10:15	8/7/05 18:15	8:00	8/7/05 14:15	1006
TC18-76	8/7/05 20:00	8/8/05 0:00	4:00	8/7/05 22:00	1014
TC18-77	8/8/05 11:00	8/8/05 21:00	10:00	8/8/05 16:00	1032
TC18-78	8/9/05 0:15	8/9/05 5:15	5:00	8/9/05 2:45	1043
TC18-79	8/10/05 12:30	8/10/05 17:45	5:15	8/10/05 15:07	1079
TC18-80	8/11/05 1:45	8/11/05 6:45	5:00	8/11/05 4:15	1092
TC18-81	8/11/05 8:15	8/11/05 17:00	8:45	8/11/05 12:37	1101
TC18-82	8/11/05 18:30	8/11/05 23:45	5:15	8/11/05 21:07	1109
TC18-83	8/12/05 1:30	8/12/05 6:00	4:30	8/12/05 3:45	1116
TC18-84	8/12/05 8:30	8/12/05 16:15	7:45	8/12/05 12:22	1124
TC18-85	8/12/05 17:15	8/12/05 21:30	4:15	8/12/05 19:22	1131
TC18-86	8/12/05 22:00	8/13/05 2:30	4:30	8/13/05 0:15	1136
TC18-87	8/14/05 7:00	8/14/05 11:00	4:00	8/14/05 9:00	1169
TC18-88	8/17/05 20:00	8/18/05 0:30	4:30	8/17/05 22:15	1223
TC18-89	8/18/05 5:30	8/18/05 12:30	7:00	8/18/05 9:00	1234
TC18-90	8/18/05 18:15	8/19/05 1:45	7:30	8/18/05 22:00	1247
TC18-91	8/19/05 2:00	8/19/05 8:15	6:15	8/19/05 5:07	1254
TC18-92	8/19/05 8:15	8/19/05 12:45	4:30	8/19/05 10:30	1260
TC18-93	8/20/05 11:00	8/20/05 17:00	6:00	8/20/05 14:00	1287
TC18-94	8/20/05 17:30	8/21/05 3:45	10:15	8/20/05 22:37	1296
TC18-95	8/21/05 15:15	8/22/05 0:15	9:00	8/21/05 19:45	1317

Notes:

1. Recycle gas was used as gasifier aeration during TC18-58 to TC18-87.
2. Transport air was used as coal conveying gas during TC18-94 and TC18-95.

Table 3.1-2 Operating Conditions
(Page 1 of 3)

Operating Periods[3,4]	Average Relative Hours	Riser Outlet Temperature TI443 °F	Pressure PI287 psig	Coal Rate[2] lb/hr	Air Rate lb/hr	Steam Rate[1] lb/hr	Nitrogen Rate lb/hr	Recycle Rate lb/hr	Syngas Rate[5] lb/hr	PCD Inlet Temperature TI458 °F	PCD Solids Rate lb/hr
TC18-1	31	1,675	230	3,400	11,700	890	5,700	0	21,000	746	440
TC18-2	42	1,658	230	3,300	11,300	1,060	6,200	0	20,700	744	420
TC18-3	54	1,684	234	3,500	12,100	1,060	5,900	0	21,700	755	400
TC18-4	61	1,673	234	3,300	11,600	1,120	6,000	0	21,100	748	390
TC18-5	100	1,705	240	3,900	13,400	1,310	6,300	0	24,300	772	370
TC18-6	125	1,690	230	2,700	11,400	1,110	6,100	0	21,400	754	350
TC18-7	131	1,672	230	3,100	12,000	1,100	6,000	0	22,200	754	370
TC18-8	146	1,689	230	3,900	13,500	1,080	6,200	0	24,500	766	470
TC18-9	167	1,659	234	4,300	12,600	560	5,400	0	21,500	730	460
TC18-10	185	1,656	234	4,200	12,700	560	5,400	0	21,600	728	460
TC18-11	192	1,658	234	4,500	12,800	560	5,400	0	21,900	731	450
TC18-12	201	1,645	234	4,300	12,600	530	5,400	0	21,500	726	450
TC18-13	209	1,658	226	3,900	12,100	550	5,500	0	20,600	723	450
TC18-14	220	1,674	226	4,000	12,400	540	5,400	0	21,100	732	440
TC18-15	228	1,672	226	4,200	12,400	540	5,200	0	20,700	729	410
TC18-16	253	1,703	226	3,100	10,700	550	6,100	0	17,600	717	300
TC18-17	263	1,692	226	3,200	10,800	550	5,500	0	17,800	714	330
TC18-18	287	1,647	226	3,600	10,900	570	5,400	0	18,200	705	390
TC18-19	296	1,646	226	3,500	10,500	570	4,800	0	17,000	699	410
TC18-20	305	1,652	226	3,400	10,500	930	4,700	0	17,700	707	410
TC18-21	326	1,664	226	3,600	11,000	990	4,700	0	18,300	714	390
TC18-22	333	1,660	226	3,600	10,900	960	4,800	0	18,300	712	390
TC18-23	339	1,671	226	3,600	11,100	950	4,800	0	18,300	715	390
TC18-24	345	1,670	226	3,100	10,200	900	4,700	0	17,100	704	380
TC18-25	354	1,671	226	3,100	10,200	850	4,800	0	16,900	700	380
TC18-26	374	1,684	225	3,400	11,400	980	5,200	0	20,600	744	350
TC18-27	391	1,662	225	3,800	12,100	1,570	5,400	0	22,400	754	410
TC18-28	399	1,661	225	4,000	11,900	1,520	5,300	0	22,000	748	440
TC18-29	413	1,650	225	4,000	12,000	1,500	5,400	0	22,200	751	430
TC18-30	417	1,646	225	4,300	12,500	1,500	5,400	0	23,000	756	420
TC18-31	422	1,645	225	4,400	12,500	1,500	5,400	0	23,000	760	410
TC18-32	434	1,634	225	4,000	11,600	780	5,400	0	21,100	762	430
TC18-33	461	1,661	225	3,900	11,500	730	5,400	0	20,600	732	450
TC18-34	469	1,662	225	3,700	11,500	720	5,400	0	20,500	731	430
TC18-35	490	1,662	225	3,400	11,100	870	5,400	0	20,300	732	370

Table 3.1-2 Operating Conditions
(Page 2 of 3)

Operating Periods[3,4]	Average Relative Hours	Riser Outlet Temperature TI443 °F	Pressure PI287 psig	Coal Rate[2] lb/hr	Air Rate lb/hr	Steam Rate[1] lb/hr	Nitrogen Rate lb/hr	Recycle Rate lb/hr	Syngas Rate[5] lb/hr	PCD Inlet Temperature TI458 °F	PCD Solids Rate lb/hr
TC18-36	509	1,661	225	3,300	10,900	930	5,300	0	19,900	728	320
TC18-37	530	1,662	225	3,500	10,400	930	5,400	0	18,300	717	290
TC18-38	541	1,683	223	3,300	10,300	430	5,500	0	18,100	718	310
TC18-39	556	1,665	223	3,400	10,300	430	5,300	0	18,100	712	340
TC18-40	588	1,665	230	3,400	11,900	400	4,900	0	20,600	729	410
TC18-41	611	1,664	228	2,900	9,500	230	5,000	0	17,400	706	310
TC18-42	620	1,660	228	2,900	9,300	20	4,900	0	16,900	701	270
TC18-43	632	1,652	228	3,600	10,200	280	5,100	0	17,600	699	270
TC18-44	638	1,655	228	3,200	10,500	250	5,000	0	18,900	709	260
TC18-45	645	1,678	228	3,400	10,400	350	5,000	0	18,900	718	260
TC18-46	653	1,676	228	3,300	10,200	320	5,000	0	18,400	713	260
TC18-47	659	1,666	228	2,900	9,700	280	5,000	0	17,700	703	260
TC18-48	681	1,687	225	3,000	10,900	330	5,100	0	19,500	724	260
TC18-49	689	1,687	225	3,400	11,000	340	4,900	0	19,500	728	260
TC18-50	699	1,687	225	3,400	11,300	350	4,900	0	19,700	727	280
TC18-51	708	1,687	225	3,300	11,600	400	4,700	0	20,000	729	280
TC18-52	717	1,688	225	3,400	11,800	390	4,600	0	20,100	731	340
TC18-53	729	1,669	209	3,300	10,800	450	4,800	0	18,600	717	310
TC18-54	735	1,703	209	3,300	11,600	460	4,900	0	19,900	735	290
TC18-55	754	1,670	221	3,500	12,400	530	6,200	0	22,600	755	370
TC18-56	765	1,667	209	3,300	11,500	420	5,800	0	20,200	738	380
TC18-57	783	1,678	225	3,300	11,500	460	5,300	0	20,400	741	320
TC18-58	801	1,704	227	3,100	12,200	480	4,100	1,974	21,600	756	330
TC18-59	825	1,673	220	3,700	11,800	200	4,100	1,547	21,000	744	350
TC18-60	834	1,684	220	3,200	11,800	220	3,800	1,542	20,700	747	360
TC18-61	850	1,660	224	3,700	11,500	200	4,000	1,570	20,500	737	340
TC18-62	857	1,678	224	3,300	11,900	190	3,700	1,560	20,700	743	370
TC18-63	863	1,663	227	3,300	11,600	210	3,700	1,587	20,400	735	370
TC18-64	918	1,654	205	2,800	9,500	220	3,800	1,408	16,300	712	280
TC18-65	924	1,656	205	2,800	9,500	220	3,800	1,411	16,300	705	280
TC18-66	932	1,663	205	2,900	9,700	340	3,700	1,412	16,500	718	290
TC18-67	939	1,673	205	2,800	9,800	390	3,800	1,391	16,600	710	300
TC18-68	954	1,674	205	2,900	9,800	240	3,700	1,386	16,600	713	310
TC18-69	960	1,673	205	2,800	9,800	250	3,800	1,389	16,600	707	310
TC18-70	965	1,674	205	2,800	9,700	220	3,800	1,396	16,500	702	310

Table 3.1-2 Operating Conditions
(Page 3 of 3)

Operating Periods[3,4]	Average Relative Hours	Riser Outlet Temperature TI443 °F	Pressure PI287 psig	Coal Rate[2] lb/hr	Air Rate lb/hr	Steam Rate[1] lb/hr	Nitrogen Rate lb/hr	Recycle Rate lb/hr	Syngas Rate[5] lb/hr	PCD Inlet Temperature TI458 °F	PCD Solids Rate lb/hr
TC18-71	979	1,675	205	2,800	9,600	220	3,800	1,403	16,400	705	310
TC18-72	988	1,674	205	2,800	9,700	220	3,000	1,397	16,300	702	320
TC18-73	993	1,672	205	2,800	9,400	220	3,000	1,404	15,900	691	320
TC18-74	1000	1,675	205	3,000	10,200	210	3,000	1,397	17,100	709	320
TC18-75	1006	1,672	205	3,100	10,100	210	3,000	1,401	16,900	707	320
TC18-76	1014	1,671	205	2,700	9,300	210	3,000	1,401	15,800	693	320
TC18-77	1032	1,675	205	3,000	10,000	210	3,000	1,396	16,800	704	310
TC18-78	1043	1,673	205	2,800	9,300	210	3,000	1,399	15,800	687	290
TC18-79	1079	1,705	205	2,900	11,200	1,360	3,000	1,676	20,100	748	240
TC18-80	1092	1,675	205	2,400	8,900	130	3,000	1,466	15,200	682	250
TC18-81	1101	1,674	205	2,700	9,500	110	3,000	1,464	15,800	692	260
TC18-82	1109	1,681	205	2,900	10,000	370	3,100	1,492	16,900	703	300
TC18-83	1116	1,677	205	2,600	9,300	370	3,100	1,507	15,900	689	330
TC18-84	1124	1,661	205	3,400	10,600	360	3,000	1,489	17,800	706	350
TC18-85	1131	1,659	205	2,900	9,700	360	3,100	1,506	16,600	697	360
TC18-86	1136	1,656	205	2,800	9,300	370	3,100	1,512	15,900	684	360
TC18-87	1169	1,676	205	3,300	10,600	350	3,100	1,465	17,800	708	380
TC18-88	1223	1,668	220	3,500	11,400	880	6,400	0	19,800	742	370
TC18-89	1234	1,691	220	3,700	11,900	870	6,200	0	20,400	752	370
TC18-90	1247	1,681	216	3,800	11,700	400	6,400	0	20,500	747	360
TC18-91	1254	1,680	220	3,500	11,500	240	6,400	0	20,100	743	350
TC18-92	1260	1,681	220	3,600	11,600	240	6,500	0	20,000	747	350
TC18-93	1287	1,709	216	3,800	11,900	250	6,300	0	21,000	759	350
TC18-94	1296	1,725	216	3,700	11,900	250	6,200	0	20,900	761	350
TC18-95	1317	1,710	216	3,900	11,700	260	6,100	0	20,500	753	350

Notes:

1. All steam rates by steam measurements.
2. Coal rates are by weigh cells with the exception of TC18-60, TC18-62 and TC18-63 which are by carbon balance.
3. Recycle gas was used as gasifier aeration during TC18-58 to TC18-87.
4. Transport air was used as coal conveying gas during TC18-94 and TC18-95.
5. Syngas rate is from the cyclone exit.

Table 3.1-3 Wet Gas Composition, Molecular Weight, and Heating Value
(Page 1 of 3)

Operating Period[2,5]	Average Relative Hour	H_2O[3] Mole %	CO Mole %	H_2 Mole %	CO_2 Mole %	CH_4 Mole %	C_2H_6 Mole %	Argon Mole %	N_2 Mole %	Total Mole %	Syngas LHV Btu/SCF	Syngas TRS[1] ppm	CO/CO_2 Molar Ratio	Syngas MW lb/Mole
TC18-1	31	14.8	6.6	7.8	9.4	1.2	0.0	0.4	59.6	100.0	54	(4)	0.70	25.9
TC18-2	42	14.9	6.2	7.7	9.4	1.3	0.0	0.4	60.1	100.0	53	(4)	0.66	25.9
TC18-3	54	11.6	7.1	8.3	9.7	1.2	0.0	0.5	61.6	100.0	57	(4)	0.74	26.2
TC18-4	61	12.9	6.3	7.9	9.6	1.2	0.0	0.4	61.7	100.0	53	(4)	0.65	26.1
TC18-5	100	14.2	6.9	8.2	9.5	1.1	0.0	0.4	59.6	100.0	55	(4)	0.73	25.9
TC18-6	125	14.5	5.2	6.8	9.0	0.8	0.0	0.4	63.3	100.0	42	(4)	0.58	26.2
TC18-7	131	13.8	6.0	7.4	9.3	1.1	0.0	0.4	62.0	100.0	49	(4)	0.64	26.1
TC18-8	146	12.8	7.1	7.7	9.2	1.2	0.0	0.4	61.5	100.0	56	(4)	0.77	26.1
TC18-9	167	7.8	10.1	7.3	8.4	1.8	0.0	0.5	64.2	100.0	69	279	1.20	26.5
TC18-10	185	8.0	10.0	7.4	8.5	1.8	0.0	0.5	63.9	100.0	69	262	1.18	26.5
TC18-11	192	8.0	10.1	7.5	8.5	1.8	0.0	0.5	63.6	100.0	70	254	1.19	26.5
TC18-12	201	7.9	10.0	7.5	8.5	1.8	0.0	0.5	63.8	100.0	70	279	1.18	26.5
TC18-13	209	7.9	9.4	6.9	8.4	1.6	0.0	0.4	65.3	100.0	64	241	1.12	26.6
TC18-14	220	8.2	9.6	7.1	8.5	1.6	0.0	0.5	64.5	100.0	65	242	1.14	26.6
TC18-15	228	8.3	9.8	7.2	8.6	1.6	0.0	0.5	64.1	100.0	66	283	1.14	26.5
TC18-16	253	9.1	9.0	7.0	8.5	1.2	0.0	0.5	64.7	100.0	59	233	1.06	26.5
TC18-17	263	8.6	9.2	6.9	8.4	1.3	0.0	0.5	65.0	100.0	61	211	1.09	26.6
TC18-18	287	8.2	9.7	7.3	8.7	1.8	0.0	0.5	63.8	100.0	68	250	1.12	26.5
TC18-19	296	8.2	10.0	7.4	8.8	1.8	0.0	0.5	63.3	100.0	69	263	1.14	26.5
TC18-20	305	9.8	9.1	7.7	9.1	1.8	0.0	0.5	62.1	100.0	67	317	1.00	26.3
TC18-21	326	10.0	9.4	7.9	9.2	1.8	0.0	0.5	61.2	100.0	69	326	1.02	26.3
TC18-22	333	9.6	9.4	7.8	9.0	1.8	0.0	0.5	61.8	100.0	69	244	1.04	26.3
TC18-23	339	9.9	9.4	7.7	9.1	1.7	0.0	0.5	61.7	100.0	67	267	1.03	26.3
TC18-24	345	9.8	8.9	7.4	9.0	1.6	0.0	0.5	62.9	100.0	63	265	0.99	26.4
TC18-25	354	9.6	8.7	7.2	9.0	1.5	0.0	0.5	63.5	100.0	62	226	0.97	26.5
TC18-26	374	11.6	6.9	6.6	9.0	1.4	0.0	0.4	64.1	100.0	53	241	0.76	26.5
TC18-27	391	11.4	7.6	7.4	9.2	1.5	0.0	0.5	62.4	100.0	59	292	0.83	26.3
TC18-28	399	11.2	7.6	7.3	9.1	1.6	0.0	0.5	62.8	100.0	59	286	0.84	26.3
TC18-29	413	10.8	7.8	7.3	9.1	1.7	0.0	0.5	62.8	100.0	61	281	0.86	26.3
TC18-30	417	10.8	7.9	7.4	9.2	1.7	0.0	0.5	62.5	100.0	62	311	0.86	26.3
TC18-31	422	11.5	7.9	7.4	9.2	1.8	0.0	0.5	61.7	100.0	62	375	0.87	26.2
TC18-32	434	9.7	8.1	6.8	8.8	1.7	0.0	0.5	64.3	100.0	61	414	0.93	26.5
TC18-33	461	7.0	8.9	6.7	8.5	1.7	0.0	0.5	66.8	100.0	62	344	1.05	26.8
TC18-34	469	7.0	9.6	6.6	8.2	1.6	0.0	0.5	66.5	100.0	64	212	1.18	26.8
TC18-35	490	8.1	8.6	6.8	8.6	1.6	0.0	0.5	65.9	100.0	61	270	1.00	26.7

Table 3.1-3 Wet Gas Composition, Molecular Weight, and Heating Value
(Page 2 of 3)

Operating Period[2,5]	Average Relative Hour	H_2O[3] Mole %	CO Mole %	H_2 Mole %	CO_2 Mole %	CH_4 Mole %	C_2H_6 Mole %	Argon Mole %	N_2 Mole %	Total Mole %	Syngas LHV Btu/SCF	Syngas TRS[1] ppm	CO/CO_2 Molar Ratio	Syngas MW lb/Mole
TC18-36	509	8.3	8.0	6.7	8.7	1.5	0.0	0.4	66.3	100.0	58	386	0.92	26.7
TC18-37	530	8.7	8.9	7.6	9.1	1.7	0.0	0.5	63.6	100.0	64	442	0.98	26.5
TC18-38	541	8.7	8.9	7.5	8.8	1.2	0.0	0.5	64.4	100.0	61	0	1.01	26.5
TC18-39	556	8.9	8.9	7.6	9.0	1.6	0.0	0.5	63.6	100.0	64	507	0.98	26.5
TC18-40	588	8.3	8.5	6.9	8.6	1.5	0.0	0.5	65.8	100.0	60	452	0.99	26.6
TC18-41	611	7.3	8.1	5.8	8.1	1.3	0.0	0.5	68.9	100.0	54	336	1.01	26.9
TC18-42	620	6.8	8.5	5.7	7.9	1.4	0.0	0.5	69.3	100.0	56	366	1.08	27.0
TC18-43	632	7.3	10.7	7.3	8.3	2.0	0.0	0.5	63.8	100.0	73	332	1.29	26.5
TC18-44	638	7.2	9.1	6.4	8.2	1.7	0.0	0.5	66.9	100.0	62	278	1.11	26.8
TC18-45	645	8.0	8.7	6.6	8.3	1.5	0.0	0.5	66.5	100.0	59	315	1.04	26.7
TC18-46	653	7.2	9.1	6.3	8.1	1.5	0.0	0.5	67.3	100.0	61	243	1.12	26.8
TC18-47	659	7.1	8.7	6.0	8.0	1.5	0.0	0.5	68.2	100.0	58	217	1.08	26.9
TC18-48	681	7.7	9.2	6.7	8.2	1.5	0.0	0.5	66.2	100.0	62	226	1.12	26.7
TC18-49	689	7.8	9.3	6.8	8.3	1.6	0.0	0.5	65.7	100.0	63	229	1.12	26.7
TC18-50	699	7.8	9.4	6.8	8.3	1.6	0.0	0.5	65.6	100.0	64	288	1.13	26.7
TC18-51	708	8.0	9.2	6.9	8.4	1.6	0.0	0.5	65.4	100.0	63	252	1.10	26.6
TC18-52	717	7.9	9.5	7.0	8.4	1.6	0.0	0.5	65.0	100.0	65	280	1.13	26.6
TC18-53	729	8.3	9.1	7.2	8.6	1.7	0.0	0.5	64.5	100.0	65	273	1.06	26.5
TC18-54	735	8.4	8.8	6.9	8.5	1.4	0.0	0.5	65.5	100.0	60	241	1.03	26.6
TC18-55	754	8.6	7.6	6.4	8.5	1.5	0.0	0.5	66.8	100.0	56	252	0.89	26.7
TC18-56	765	8.3	8.4	6.9	8.8	1.7	0.0	0.5	65.4	100.0	62	290	0.95	26.6
TC18-57	783	8.4	8.3	6.6	8.5	1.5	0.0	0.5	66.2	100.0	58	232	0.97	26.7
TC18-58	801	8.7	9.1	7.2	8.9	1.4	0.0	0.5	64.2	100.0	62	247	1.02	26.6
TC18-59	825	8.1	9.3	7.0	8.9	1.7	0.0	0.5	64.5	100.0	65	246	1.05	26.7
TC18-60	834	8.5	9.2	7.1	9.0	1.5	0.0	0.5	64.3	100.0	63	270	1.02	26.6
TC18-61	850	8.0	9.3	7.1	9.0	1.8	0.0	0.5	64.4	100.0	65	249	1.03	26.7
TC18-62	857	8.2	9.5	7.1	9.0	1.7	0.0	0.5	64.1	100.0	65	258	1.05	26.6
TC18-63	863	7.9	9.5	7.1	9.0	1.8	0.0	0.5	64.1	100.0	67	251	1.06	26.6
TC18-64	918	8.1	9.4	7.4	9.3	1.7	0.0	0.5	63.6	100.0	67	314	1.01	26.6
TC18-65	924	8.6	9.1	7.4	9.3	1.7	0.0	0.5	63.4	100.0	65	249	0.98	26.6
TC18-66	932	8.4	9.5	7.5	9.3	1.7	0.0	0.5	63.2	100.0	67	350	1.02	26.6
TC18-67	939	8.4	9.8	7.7	9.1	1.6	0.0	0.5	63.0	100.0	67	396	1.07	26.5
TC18-68	954	8.3	9.8	7.5	9.1	1.7	0.0	0.5	63.1	100.0	68	367	1.08	26.5
TC18-69	960	8.3	9.9	7.7	9.0	1.6	0.0	0.5	63.1	100.0	67	419	1.09	26.5
TC18-70	965	8.3	9.8	7.6	9.0	1.5	0.0	0.5	63.4	100.0	66	350	1.09	26.5

Table 3.1-3 Wet Gas Composition, Molecular Weight, and Heating Value
(Page 3 of 3)

Operating Period[2,5]	Average Relative Hour	H_2O[3] Mole %	CO Mole %	H_2 Mole %	CO_2 Mole %	CH_4 Mole %	C_2H_6 Mole %	Argon Mole %	N_2 Mole %	Total Mole %	Syngas LHV Btu/SCF	Syngas TRS[1] ppm	CO/CO_2 Molar Ratio	Syngas MW lb/Mole
TC18-71	979	8.2	9.8	7.5	9.1	1.5	0.0	0.5	63.4	100.0	66	348	1.07	26.6
TC18-72	988	8.1	10.1	7.6	9.0	1.5	0.0	0.5	63.2	100.0	68	352	1.12	26.5
TC18-73	993	8.2	9.7	7.5	9.0	1.4	0.0	0.5	63.8	100.0	64	318	1.08	26.6
TC18-74	1000	8.1	10.3	7.6	9.1	1.7	0.0	0.5	62.6	100.0	70	335	1.13	26.5
TC18-75	1006	7.9	10.5	7.6	9.1	1.8	0.0	0.5	62.6	100.0	71	362	1.15	26.5
TC18-76	1014	8.2	9.7	7.4	9.0	1.5	0.0	0.5	63.7	100.0	65	350	1.07	26.6
TC18-77	1032	8.0	10.3	7.8	9.0	1.7	0.0	0.5	62.6	100.0	71	349	1.14	26.5
TC18-78	1043	8.2	9.6	7.5	9.0	1.3	0.0	0.5	63.8	100.0	64	333	1.06	26.6
TC18-79	1079	14.1	6.7	8.8	10.5	1.0	0.0	0.5	58.4	100.0	55	295	0.64	25.9
TC18-80	1092	7.9	9.3	7.0	8.7	1.1	0.0	0.5	65.4	100.0	60	250	1.07	26.7
TC18-81	1101	7.4	10.3	7.1	8.7	1.5	0.0	0.5	64.4	100.0	66	274	1.17	26.7
TC18-82	1109	8.5	9.9	7.6	9.0	1.5	0.0	0.5	63.0	100.0	67	264	1.11	26.5
TC18-83	1116	8.6	9.1	7.4	9.0	1.2	0.0	0.5	64.1	100.0	61	214	1.01	26.6
TC18-84	1124	8.1	10.7	8.0	9.1	1.9	0.0	0.5	61.6	100.0	74	239	1.18	26.4
TC18-85	1131	8.4	9.7	7.6	9.2	1.7	0.0	0.5	62.9	100.0	68	249	1.06	26.5
TC18-86	1136	8.5	9.1	7.5	9.2	1.4	0.0	0.5	63.7	100.0	63	244	0.99	26.6
TC18-87	1169	8.1	10.6	7.9	9.0	1.7	0.0	0.5	62.2	100.0	72	247	1.18	26.5
TC18-88	1223	7.8	8.7	6.5	8.3	1.5	0.0	0.5	66.6	100.0	60	270	1.05	26.7
TC18-89	1234	7.6	9.1	6.7	8.3	1.4	0.0	0.5	66.4	100.0	61	281	1.09	26.7
TC18-90	1247	7.9	8.4	6.8	8.5	1.4	0.0	0.5	66.5	100.0	59	304	0.99	26.7
TC18-91	1254	7.3	8.7	6.7	8.4	1.4	0.0	0.5	67.0	100.0	60	299	1.04	26.8
TC18-92	1260	7.2	9.0	6.7	8.2	1.5	0.0	0.5	66.9	100.0	61	307	1.09	26.7
TC18-93	1287	7.9	9.4	7.3	8.5	1.4	0.0	0.5	65.1	100.0	63	353	1.11	26.6
TC18-94	1296	7.9	9.5	7.3	8.4	1.3	0.0	0.5	65.2	100.0	62	289	1.13	26.6
TC18-95	1317	7.9	9.6	7.2	8.2	1.3	0.0	0.5	65.1	100.0	63	259	1.17	26.6

Notes:

1. Syngas total reduced sulfur (TRS) estimated from Syngas combustor SO_2 analyzer data.
2. Recycle gas was used as gasifier aeration during TC18-58 to TC18-87.
3. The H_2O concentration was estimated using a correlation between the in-situ samples and gasifier temperature for TC18-1, TC18-2 and TC18-9 through TC18-30. The analyzer data were used for all remaining periods.
4. No data available. SO_2 analyzer out of service.
5. Transport air was used as coal conveying gas during TC18-94 and TC18-95.

Table 3.1-4 Syngas Sulfur Concentrations
(Page 1 of 3)

Operating Period[2,3]	Average Relative Hour	Wet AI419J H_2S ppm	Syngas Total Reduced Sulfur[1] ppm
TC18-1	31	(4)	(5)
TC18-2	42	(4)	(5)
TC18-3	54	(4)	(5)
TC18-4	61	(4)	(5)
TC18-5	100	(4)	(5)
TC18-6	125	(4)	(5)
TC18-7	131	(4)	(5)
TC18-8	146	(4)	(5)
TC18-9	167	(4)	257
TC18-10	185	(4)	241
TC18-11	192	(4)	234
TC18-12	201	(4)	257
TC18-13	209	(4)	222
TC18-14	220	(4)	222
TC18-15	228	(4)	259
TC18-16	253	134	212
TC18-17	263	126	193
TC18-18	287	(4)	229
TC18-19	296	(4)	242
TC18-20	305	(4)	286
TC18-21	326	119	293
TC18-22	333	143	220
TC18-23	339	150	241
TC18-24	345	148	239
TC18-25	354	150	205
TC18-26	374	(4)	213
TC18-27	391	(4)	259
TC18-28	399	(4)	254
TC18-29	413	(4)	251
TC18-30	417	(4)	277
TC18-31	422	(4)	332
TC18-32	434	120	374
TC18-33	461	129	320
TC18-34	469	110	197
TC18-35	490	174	249

Table 3.1-4 Syngas Sulfur Concentrations
(Page 2 of 3)

Operating Period[2,3]	Average Relative Hour	Wet AI419J H_2S ppm	Syngas Total Reduced Sulfur[1] ppm
TC18-36	509	238	354
TC18-37	530	306	403
TC18-38	541	330	(5)
TC18-39	556	336	462
TC18-40	588	311	414
TC18-41	611	298	312
TC18-42	620	193	341
TC18-43	632	203	308
TC18-44	638	206	258
TC18-45	645	200	290
TC18-46	653	179	226
TC18-47	659	176	202
TC18-48	681	183	209
TC18-49	689	184	211
TC18-50	699	191	266
TC18-51	708	202	232
TC18-52	717	201	258
TC18-53	729	229	250
TC18-54	735	215	221
TC18-55	754	209	230
TC18-56	765	218	266
TC18-57	783	220	213
TC18-58	801	204	225
TC18-59	825	214	226
TC18-60	834	227	247
TC18-61	850	227	229
TC18-62	857	226	237
TC18-63	863	223	231
TC18-64	918	231	289
TC18-65	924	270	228
TC18-66	932	294	320
TC18-67	939	316	363
TC18-68	954	232	337
TC18-69	960	223	384
TC18-70	965	282	321

Table 3.1-4 Syngas Sulfur Concentrations
(Page 3 of 3)

Operating Period[2,3]	Average Relative Hour	Wet AI419J H_2S ppm	Syngas Total Reduced Sulfur[1] ppm
TC18-71	979	289	319
TC18-72	988	323	323
TC18-73	993	311	292
TC18-74	1000	308	308
TC18-75	1006	316	334
TC18-76	1014	331	321
TC18-77	1032	324	321
TC18-78	1043	318	306
TC18-79	1079	289	253
TC18-80	1092	242	230
TC18-81	1101	242	253
TC18-82	1109	253	242
TC18-83	1116	250	196
TC18-84	1124	141	219
TC18-85	1131	142	228
TC18-86	1136	192	223
TC18-87	1169	237	227
TC18-88	1223	(4)	249
TC18-89	1234	(4)	260
TC18-90	1247	(4)	280
TC18-91	1254	(4)	277
TC18-92	1260	(4)	285
TC18-93	1287	(4)	325
TC18-94	1296	108	266
TC18-95	1317	180	238

Notes:
1. Syngas total reduced sulfur (TRS) estimated from Syngas combustor SO_2 analyzer data.
2. Recycle gas was used as gasifier aeration during TC18-58 to TC18-87.
3. Transport Air was used as coal conveying gas during TC18-94 and TC18-95.
4. H_2S Analyzer AI419J out of service
5. SO_2 analyzer AI476N out of service

Table 3.1-5 Projected[1] Wet Syngas Composition, Molecular Weight, and Projected Heating Value
(Page 1 of 3)

Operating Period[2,3]	Average Relative Hour	H_2O Mole %	CO Mole %	H_2 Mole %	CO_2 Mole %	CH_4 Mole %	C_2H_6 Mole %	Argon Mole %	N_2 Mole %	Total Mole %	Syngas MW lb/Mole	Syngas CO/CO_2 Ratio	Syngas LHV Btu/SCF
TC18-1	31	21.6	10.9	14.0	10.7	2.1	0.0	0.5	40.2	100.0	23.7	1.0	118
TC18-2	42	22.4	10.6	14.0	10.8	2.3	0.0	0.5	39.4	100.0	23.6	1.0	120
TC18-3	54	17.7	13.1	14.2	10.4	2.2	0.0	0.5	41.9	100.0	24.0	1.3	122
TC18-4	61	19.8	11.8	13.9	10.6	2.2	0.0	0.5	41.3	100.0	23.9	1.1	119
TC18-5	100	20.7	11.3	13.9	10.7	1.9	0.0	0.5	41.0	100.0	23.9	1.1	115
TC18-6	125	23.1	9.7	13.4	10.6	1.5	0.0	0.5	41.2	100.0	23.8	0.9	105
TC18-7	131	21.3	10.8	13.6	10.7	1.9	0.0	0.5	41.3	100.0	23.9	1.0	113
TC18-8	146	18.7	12.2	13.7	10.5	2.2	0.0	0.5	42.2	100.0	24.1	1.2	118
TC18-9	167	10.3	17.7	13.2	8.7	3.2	0.0	0.6	46.3	100.0	24.6	2.0	135
TC18-10	185	10.6	17.4	13.2	8.9	3.1	0.0	0.6	46.2	100.0	24.6	2.0	134
TC18-11	192	10.6	17.5	13.2	8.9	3.1	0.0	0.6	46.2	100.0	24.6	2.0	135
TC18-12	201	10.5	17.6	13.3	8.8	3.2	0.0	0.6	46.0	100.0	24.6	2.0	136
TC18-13	209	10.5	17.2	13.0	8.8	3.0	0.0	0.6	47.0	100.0	24.7	2.0	132
TC18-14	220	10.9	16.9	13.0	8.9	2.9	0.0	0.6	46.8	100.0	24.7	1.9	130
TC18-15	228	10.8	16.9	13.0	8.9	2.9	0.0	0.6	47.0	100.0	24.7	1.9	129
TC18-16	253	11.8	15.9	13.4	8.8	2.1	0.0	0.6	47.4	100.0	24.6	1.8	121
TC18-17	263	11.0	16.6	13.3	8.7	2.5	0.0	0.6	47.3	100.0	24.6	1.9	126
TC18-18	287	10.7	17.4	13.4	8.8	3.3	0.0	0.6	45.9	100.0	24.5	2.0	136
TC18-19	296	10.4	17.5	13.3	8.6	3.2	0.0	0.6	46.5	100.0	24.6	2.0	135
TC18-20	305	13.0	15.7	13.7	9.5	3.2	0.0	0.5	44.4	100.0	24.3	1.7	134
TC18-21	326	13.2	15.6	13.6	9.6	3.1	0.0	0.5	44.3	100.0	24.4	1.6	133
TC18-22	333	12.8	15.9	13.7	9.4	3.2	0.0	0.5	44.6	100.0	24.4	1.7	134
TC18-23	339	13.0	15.6	13.5	9.5	3.0	0.0	0.5	44.8	100.0	24.4	1.6	131
TC18-24	345	13.1	15.6	13.7	9.4	2.9	0.0	0.5	44.9	100.0	24.4	1.7	130
TC18-25	354	12.9	15.4	13.5	9.3	2.8	0.0	0.5	45.6	100.0	24.4	1.7	128
TC18-26	374	17.2	12.7	13.3	10.4	2.7	0.0	0.5	43.2	100.0	24.2	1.2	123
TC18-27	391	17.0	13.8	14.0	10.6	2.9	0.0	0.5	41.3	100.0	24.1	1.3	130
TC18-28	399	16.6	13.9	13.9	10.5	2.9	0.0	0.5	41.7	100.0	24.1	1.3	130
TC18-29	413	15.9	14.5	13.9	10.5	3.1	0.0	0.5	41.6	100.0	24.2	1.4	134
TC18-30	417	15.8	14.4	13.7	10.5	3.2	0.0	0.5	42.0	100.0	24.2	1.4	133
TC18-31	422	16.6	13.9	13.7	10.6	3.1	0.0	0.5	41.6	100.0	24.2	1.3	132
TC18-32	434	14.0	15.4	13.7	10.0	3.4	0.0	0.5	43.0	100.0	24.3	1.5	136
TC18-33	461	9.8	18.6	13.1	8.8	3.4	0.0	0.6	45.8	100.0	24.7	2.1	139
TC18-34	469	9.4	19.5	13.8	8.5	3.4	0.0	0.5	44.9	100.0	24.5	2.3	144
TC18-35	490	11.6	17.6	14.0	9.2	3.3	0.0	0.5	43.8	100.0	24.4	1.9	140

Table 3.1-5 Projected[1] Wet Syngas Composition, Molecular Weight, and Projected Heating Value
(Page 2 of 3)

Operating Period[2,3]	Average Relative Hour	H_2O Mole %	CO Mole %	H_2 Mole %	CO_2 Mole %	CH_4 Mole %	C_2H_6 Mole %	Argon Mole %	N_2 Mole %	Total Mole %	Syngas MW lb/Mole	Syngas CO/CO_2 Ratio	Syngas LHV Btu/SCF
TC18-36	509	12.5	16.8	13.9	9.5	3.1	0.0	0.5	43.7	100.0	24.4	1.8	136
TC18-37	530	12.2	17.2	14.1	9.4	3.1	0.0	0.5	43.5	100.0	24.3	1.8	139
TC18-38	541	12.2	17.1	14.4	9.2	2.4	0.0	0.5	44.2	100.0	24.3	1.9	132
TC18-39	556	12.4	16.9	14.2	9.3	3.0	0.0	0.5	43.7	100.0	24.3	1.8	137
TC18-40	588	11.7	16.4	13.4	9.0	2.9	0.0	0.6	46.0	100.0	24.5	1.8	131
TC18-41	611	10.3	18.6	14.4	8.5	3.1	0.0	0.5	44.6	100.0	24.3	2.2	141
TC18-42	620	9.0	19.7	14.2	7.9	3.3	0.0	0.5	45.4	100.0	24.4	2.5	144
TC18-43	632	9.0	20.0	13.8	8.2	3.8	0.0	0.5	44.6	100.0	24.4	2.4	150
TC18-44	638	9.7	19.1	13.7	8.5	3.6	0.0	0.5	44.8	100.0	24.5	2.2	146
TC18-45	645	11.2	17.9	14.2	8.9	3.1	0.0	0.5	44.2	100.0	24.3	2.0	140
TC18-46	653	9.4	19.5	14.0	8.3	3.4	0.0	0.5	44.8	100.0	24.4	2.4	145
TC18-47	659	9.4	19.5	14.2	8.2	3.4	0.0	0.5	44.7	100.0	24.4	2.4	146
TC18-48	681	10.5	18.4	14.0	8.7	3.1	0.0	0.5	44.9	100.0	24.4	2.1	139
TC18-49	689	10.6	18.2	13.8	8.8	3.1	0.0	0.5	45.0	100.0	24.5	2.1	139
TC18-50	699	10.3	18.2	13.7	8.6	3.1	0.0	0.5	45.5	100.0	24.5	2.1	138
TC18-51	708	10.6	17.3	13.4	8.7	3.0	0.0	0.6	46.5	100.0	24.6	2.0	133
TC18-52	717	10.4	17.4	13.1	8.7	3.0	0.0	0.6	46.7	100.0	24.7	2.0	133
TC18-53	729	10.9	17.3	13.6	8.7	3.3	0.0	0.5	45.6	100.0	24.4	2.0	137
TC18-54	735	11.3	16.5	13.1	8.9	2.7	0.0	0.6	47.0	100.0	24.7	1.8	127
TC18-55	754	12.6	15.3	12.7	9.6	3.1	0.0	0.6	46.1	100.0	24.7	1.6	128
TC18-56	765	11.5	16.5	12.9	9.3	3.2	0.0	0.6	46.0	100.0	24.7	1.8	132
TC18-57	783	11.7	16.1	12.8	9.3	2.9	0.0	0.6	46.7	100.0	24.7	1.7	127
TC18-58	801	11.5	17.7	14.3	9.0	2.7	0.0	0.6	44.2	100.0	24.3	2.0	136
TC18-59	825	10.6	18.5	14.0	8.8	3.4	0.0	0.6	44.1	100.0	24.4	2.1	143
TC18-60	834	11.1	17.9	14.1	8.9	3.0	0.0	0.6	44.4	100.0	24.4	2.0	138
TC18-61	850	10.5	18.8	14.2	8.8	3.6	0.0	0.6	43.6	100.0	24.3	2.1	147
TC18-62	857	10.5	18.4	13.9	8.7	3.2	0.0	0.6	44.7	100.0	24.4	2.1	141
TC18-63	863	10.2	19.0	14.2	8.6	3.6	0.0	0.6	43.9	100.0	24.3	2.2	147
TC18-64	918	9.8	19.8	15.1	8.1	3.6	0.0	0.6	42.9	100.0	24.0	2.4	153
TC18-65	924	10.7	18.9	15.2	8.4	3.4	0.0	0.6	42.8	100.0	24.0	2.3	149
TC18-66	932	10.2	19.1	15.0	8.2	3.5	0.0	0.6	43.4	100.0	24.1	2.3	149
TC18-67	939	10.1	19.4	15.3	8.1	3.2	0.0	0.6	43.4	100.0	24.0	2.4	147
TC18-68	954	9.8	19.3	15.0	8.0	3.3	0.0	0.6	43.9	100.0	24.1	2.4	147
TC18-69	960	9.8	19.7	15.4	7.9	3.1	0.0	0.6	43.5	100.0	24.0	2.5	148
TC18-70	965	9.7	19.8	15.6	7.8	3.0	0.0	0.6	43.6	100.0	24.0	2.5	147

Table 3.1-5 Projected[1] Wet Syngas Composition, Molecular Weight, and Projected Heating Value (Page 3 of 3)

Operating Period[2,3]	Average Relative Hour	H_2O Mole %	CO Mole %	H_2 Mole %	CO_2 Mole %	CH_4 Mole %	C_2H_6 Mole %	Argon Mole %	N_2 Mole %	Total Mole %	Syngas MW lb/Mole	Syngas CO/CO_2 Ratio	Syngas LHV Btu/SCF
TC18-71	979	9.7	19.8	15.3	7.9	3.1	0.0	0.6	43.6	100.0	24.0	2.5	148
TC18-72	988	9.3	20.2	15.4	7.7	3.1	0.0	0.6	43.7	100.0	24.0	2.6	149
TC18-73	993	9.5	19.9	15.7	7.7	2.8	0.0	0.6	43.7	100.0	23.9	2.6	146
TC18-74	1000	9.4	19.6	14.5	8.1	3.3	0.0	0.6	44.5	100.0	24.3	2.4	146
TC18-75	1006	9.1	20.0	14.6	7.9	3.4	0.0	0.6	44.3	100.0	24.2	2.5	149
TC18-76	1014	9.7	19.9	15.6	7.8	3.0	0.0	0.6	43.5	100.0	23.9	2.6	148
TC18-77	1032	9.3	20.0	15.1	7.8	3.3	0.0	0.6	43.9	100.0	24.1	2.6	149
TC18-78	1043	9.9	19.8	15.9	7.7	2.8	0.0	0.6	43.3	100.0	23.9	2.6	147
TC18-79	1079	20.6	12.3	15.1	10.6	1.7	0.0	0.5	39.0	100.0	23.6	1.2	122
TC18-80	1092	8.8	20.7	16.3	7.1	2.6	0.0	0.6	43.8	100.0	23.8	2.9	147
TC18-81	1101	8.0	21.1	15.0	7.1	3.2	0.0	0.6	45.0	100.0	24.1	3.0	149
TC18-82	1109	10.1	19.2	15.3	8.0	2.9	0.0	0.6	43.8	100.0	24.0	2.4	145
TC18-83	1116	10.6	19.2	16.3	7.9	2.5	0.0	0.6	43.0	100.0	23.7	2.4	144
TC18-84	1124	9.7	19.5	14.7	8.1	3.5	0.0	0.6	43.9	100.0	24.2	2.4	149
TC18-85	1131	10.2	19.5	15.5	8.1	3.4	0.0	0.6	42.7	100.0	23.9	2.4	151
TC18-86	1136	10.5	19.4	16.0	8.0	3.1	0.0	0.6	42.3	100.0	23.8	2.4	150
TC18-87	1169	9.4	19.7	14.6	8.0	3.2	0.0	0.6	44.5	100.0	24.2	2.4	145
TC18-88	1223	10.6	17.0	13.0	8.7	3.0	0.0	0.6	47.1	100.0	24.7	1.9	131
TC18-89	1234	10.2	17.2	12.9	8.6	2.7	0.0	0.6	47.9	100.0	24.8	2.0	128
TC18-90	1247	11.0	16.7	13.0	8.9	2.8	0.0	0.6	47.0	100.0	24.7	1.9	129
TC18-91	1254	10.0	17.6	13.0	8.6	2.9	0.0	0.6	47.4	100.0	24.7	2.1	131
TC18-92	1260	9.6	17.8	13.0	8.3	2.9	0.0	0.6	47.8	100.0	24.7	2.1	131
TC18-93	1287	10.8	17.9	13.7	8.9	2.6	0.0	0.5	45.6	100.0	24.6	2.0	133
TC18-94	1296	10.8	17.8	13.8	8.8	2.4	0.0	0.6	45.9	100.0	24.5	2.0	130
TC18-95	1317	10.6	18.1	14.0	8.7	2.5	0.0	0.5	45.6	100.0	24.5	2.1	133

Notes:
1. Adjustments are based on the following assumptions: only air nitrogen is in the syngas, the gasifier is adiabatic, and syngas is at the turbine inlet after the syngas cleanup processes.
2. Recycle gas was used as gasifier aeration during TC18-58 to TC18-87.
3. Transport air was used as coal conveying gas during TC18-94 and TC18-95.

Table 3.1-6 Coal Analysis

	Powder River Basin	
	Average Value	Standard Deviation
Moisture, wt%	18.04	1.08
Carbon, wt%	58.06	1.09
Hydrogen, wt%	3.64	0.13
Nitrogen, wt%	0.78	0.03
Oxygen, wt%	13.50	1.00
Sulfur, wt%	0.25	0.04
Ash, wt%	5.73	0.45
Volatiles, wt%	34.80	0.12
Fixed Carbon, wt%	41.43	1.08
Higher Heating Value, Btu/lb	9,619	144
Lower Heating Value, Btu/lb	9,281	136
CaO, wt %	1.22	0.08
SiO_2, wt %	2.15	0.26
Al_2O_3, wt %	0.97	0.09
MgO, wt %	0.19	0.02
Na_2O, wt %	0.93	0.14
Fe_2O_3, wt %	5.70	0.22
Na, wt % in ash	0.69	0.11
Ca/S, mole/mole	2.87	0.38

Notes:
1. All analyses are averages of coal feeder sample results.
2. Hydrogen in coal is reported separately from hydrogen in moisture.
3. Oxygen is calculated by difference.

Table 3.1-7 Standpipe Solids Analysis

Sample Number	Sample Date & Time	Sample Run Time Hours	SiO_2 Wt. %	Al_2O_3 Wt. %	FeO Wt. %	Other Inerts[1] Wt. %	$CaCO_3$ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB17696	6/28/2005 10:00	47	88.2	5.2	1.0	2.3	1.3	0.0	1.8	0.5	0.2	100.6
AB17738	6/29/2005 10:00	71	88.3	5.5	1.3	2.0	1.2	0.1	1.5	0.4	0.1	100.4
AB17763	6/30/2005 2:00	95	85.7	7.1	1.8	2.3	1.5	0.0	1.5	0.5	0.0	100.4

Notes:

1. Other inerts consist of P_2O_5, Na_2O, K_2O, BaO, & TiO_2.

Table 3.1-8 Loop Seal Solids Sample Analysis
(Page 1 of 2)

Sample Number	Sample[1,2] Date & Time	Sample Run Time Hours	SiO_2 Wt. %	Al_2O_3 Wt. %	FeO Wt. %	Other Inerts[3] Wt. %	$CaCO_3$ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB17808	7/1/2005 18:00	137	82.1	7.6	1.4	3.1	3.2	0.0	2.6	0.6	1.4	102.0
AB17889	7/3/2005 18:00	185	74.7	8.7	2.2	2.9	2.5	0.0	6.4	1.1	5.9	104.5
AB17892	7/4/2005 18:00	209	71.9	9.9	2.5	3.1	2.6	0.1	7.8	1.3	8.3	107.4
AB17894	7/5/2005 10:00	225	71.0	10.2	2.7	3.0	2.5	0.0	8.4	1.3	4.4	103.5
AB17959	7/6/2005 10:00	249	68.3	11.1	3.0	3.4	1.5	0.0	10.2	1.5	1.5	100.6
AB17995	7/7/2005 11:00	274	66.3	11.7	3.2	3.4	1.7	0.1	10.0	1.6	8.4	106.4
AB18046	7/8/2005 10:00	297	63.7	11.4	3.4	3.6	1.7	0.1	10.9	1.8	5.9	102.6
AB18083	7/9/2005 18:00	329	61.1	13.4	3.6	3.8	1.7	0.1	12.2	1.9	7.0	104.7
AB18085	7/10/2005 10:00	345	59.5	13.1	3.7	3.3	1.8	0.0	11.7	1.9	6.3	101.3
AB18125	7/12/2005 10:00	365	59.5	14.6	4.1	3.8	2.6	0.0	12.4	1.9	4.7	103.7
AB18146	7/13/2005 10:00	389	59.6	13.9	3.7	3.8	2.0	0.0	12.5	2.0	4.8	102.4
AB18175	7/14/2005 10:00	413	56.7	15.1	4.5	3.9	2.0	0.1	13.4	2.0	2.9	100.5
AB18202	7/15/2005 10:00	437	57.8	13.3	4.5	3.9	1.9	0.2	12.7	1.8	4.6	100.8
AB18308	7/16/2005 2:00	453	56.9	14.6	4.6	3.6	1.5	0.0	14.4	1.9	5.8	103.4
AB18313	7/17/2005 18:00	493	53.6	14.4	5.3	3.9	5.2	0.1	14.8	2.1	4.5	103.9
AB18361	7/18/2005 18:00	517	51.5	15.4	5.8	4.0	1.1	0.1	17.1	2.5	2.8	100.2
AB18378	7/19/2005 18:00	541	49.8	15.9	6.3	4.1	1.7	0.2	17.0	2.1	2.5	99.6
AB18410	7/21/2005 2:00	573	47.5	16.4	6.5	3.7	1.3	0.2	17.8	2.7	4.0	100.1
AB18469	7/22/2005 10:00	605	45.8	16.7	7.1	3.9	2.5	0.2	18.2	2.6	2.8	99.8
AB18480	7/23/2005 2:00	621	46.2	17.1	6.9	3.8	2.1	0.2	18.5	2.6	8.9	106.4
AB18534	7/24/2005 2:00	645	45.9	17.2	6.4	3.9	1.7	0.1	18.4	2.8	5.7	102.1
AB18546	7/25/2005 7:30	674	44.7	17.6	6.2	4.0	2.3	0.1	18.7	2.6	4.2	100.5
AB18621	7/25/2005 18:00	685	45.2	18.4	6.1	4.2	2.4	0.1	17.9	2.6	4.5	101.4
AB18623	7/26/2005 10:00	701	44.5	18.9	6.4	4.0	1.8	0.1	19.3	2.7	2.9	100.5
AB18644	7/27/2005 10:00	725	44.4	19.0	6.4	3.9	2.2	0.1	19.5	2.6	2.3	100.3

Table 3.1-8 Loop Seal Solids Sample Analysis
(Page 2 of 2)

Sample Number	Sample[1,2] Date & Time	Sample Run Time Hours	SiO_2 Wt. %	Al_2O_3 Wt. %	FeO Wt. %	Other Inerts[3] Wt. %	$CaCO_3$ Wt. %	CaS Wt. %	CaO Wt. %	MgO Wt. %	Organic Carbon Wt. %	Total Wt. %
AB18668	7/28/2005 10:00	749	44.1	19.1	6.1	4.2	1.9	0.1	19.3	2.4	2.7	100.0
AB18705	7/29/2005 10:00	773	44.1	19.1	6.0	3.9	1.8	0.1	19.4	2.6	2.6	99.7
AB18781	7/30/2005 2:00	789	43.8	18.8	6.3	4.1	1.9	0.1	20.3	2.9	1.1	99.3
AB18785	7/31/2005 10:00	821	51.8	16.0	5.1	4.0	2.0	0.1	16.0	2.2	1.9	99.1
AB18788	8/1/2005 8:05	843	51.6	16.4	5.2	3.8	2.2	0.1	16.5	2.3	2.7	100.7
AB18941	8/4/2005 18:00	925	49.8	16.3	5.0	3.6	3.5	0.2	15.2	2.2	7.6	103.4
AB18968	8/5/2005 18:00	949	48.6	18.5	5.5	3.9	2.3	0.2	16.5	2.2	2.5	100.2
AB18994	8/6/2005 18:00	973	47.9	19.1	5.4	4.0	2.6	0.2	15.7	2.1	5.0	102.2
AB19031	8/7/2005 18:00	997	47.9	19.6	5.7	4.2	2.5	0.2	16.1	2.1	1.9	100.1
AB19046	8/8/2005 18:00	1021	48.7	19.4	5.3	4.1	2.9	0.1	15.1	2.1	3.4	101.1
AB19105	8/10/2005 18:00	1069	46.3	21.4	6.2	4.4	2.3	0.0	16.5	2.2	0.2	99.4
AB19129	8/11/2005 2:00	1077	44.9	20.5	6.1	4.3	2.5	0.1	16.9	2.3	3.5	101.3
AB19134	8/11/2005 18:00	1093	44.5	19.4	5.9	4.3	2.7	0.1	17.0	2.3	6.0	102.3
AB19152	8/12/2005 14:00	1113	43.3	19.7	5.7	4.3	2.0	0.1	17.7	2.5	7.0	102.3
AB19168	8/13/2005 2:00	1125	44.5	19.2	5.8	4.1	2.0	0.1	18.1	2.3	7.2	103.3
AB19205	8/14/2005 10:00	1157	43.7	20.2	6.1	3.9	1.8	0.1	18.8	2.7	3.2	100.5
AB19281	8/18/2005 2:00	1171	66.9	12.5	3.2	2.9	1.5	0.1	9.5	1.3	3.6	101.6
AB19325	8/19/2005 10:00	1203	62.8	15.0	3.5	3.1	0.5	0.1	11.2	1.5	5.3	103.0
AB19389	8/20/2005 18:00	1235	60.2	15.8	3.9	3.5	0.5	0.1	12.3	1.7	1.7	99.7
AB19392	8/21/2005 18:00	1259	57.3	15.7	4.0	3.2	4.3	0.1	10.4	1.6	5.8	102.3

Notes:

1. August 1 to August 14 samples taken while recycle gas compressor was running.
2. Transport air was used as coal conveying gas from August 20 to August 21.
3. Other inerts consist of P_2O_5, Na_2O, K_2O, BaO, & TiO_2.

Table 3.1-9 PCD Solids Analysis
(Page 1 of 3)

Sample Number	Sample[2,3] Date & Time	Sample Run Time Hours	SiO_2 Wt. %	Al_2O_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
AB17667	6/27/2005 10:00	33	38.9	8.8	2.7	3.2	7.8	0.3	5.7	1.8	29.4	98.4	4,477	4,427
AB17675	6/27/2005 22:00	45	34.3	8.4	2.7	2.9	6.5	0.3	6.2	1.8	34.9	97.8	5,336	5,281
AB17689	6/28/2005 6:00	53	40.6	9.4	2.8	3.4	8.0	0.2	7.4	2.0	28.8	102.7	3,946	3,906
AB17695	6/28/2005 14:00	61	47.2	8.8	2.6	3.1	9.2	0.1	5.2	1.7	24.7	102.6	3,514	3,477
AB17766	6/30/2005 6:00	101	37.8	9.7	2.9	3.3	8.0	0.2	7.7	2.0	28.6	100.3	4,298	4,253
AB17787	7/1/2005 6:00	125	41.8	10.6	3.3	3.4	7.2	0.2	9.7	2.1	21.3	99.5	3,302	3,273
AB17803	7/1/2005 14:00	133	35.0	9.6	3.0	3.0	8.3	0.3	8.2	2.0	30.0	99.5	4,590	4,546
AB17825	7/2/2005 2:00	145	35.7	9.4	2.9	3.2	7.5	0.2	8.0	2.0	32.0	100.8	4,663	4,614
AB17871	7/3/2005 2:00	169	23.2	7.2	2.3	2.4	9.1	1.1	3.3	1.5	49.8	99.8	7,420	7,349
AB17875	7/3/2005 18:00	185	21.1	7.2	2.2	2.4	8.7	0.8	3.6	1.5	52.7	100.2	7,904	7,832
AB17878	7/4/2005 6:00	197	21.3	7.8	2.4	2.4	9.1	1.0	4.0	1.6	49.8	99.4	7,566	7,499
AB17879	7/4/2005 10:00	201	20.8	7.5	2.5	2.4	7.5	1.0	5.0	1.6	51.7	100.0	7,671	7,601
AB17881	7/4/2005 18:00	209	20.5	8.1	2.5	2.5	8.1	1.0	5.3	1.7	50.9	100.6	7,541	7,471
AB17885	7/5/2005 10:00	225	22.1	8.4	2.6	2.6	7.3	1.0	6.0	1.8	46.5	98.2	7,197	7,133
AB17886	7/5/2005 14:00	229	21.8	8.4	2.6	2.6	9.1	1.0	4.9	1.8	49.8	101.9	7,260	7,195
AB17964	7/6/2005 14:00	253	27.2	10.1	3.1	3.1	8.8	0.7	8.0	2.1	35.9	99.1	5,561	5,512
AB17986	7/7/2005 6:00	269	26.4	10.0	3.1	3.1	8.7	0.9	7.6	2.1	36.1	98.1	5,685	5,638
AB18000	7/7/2005 9:00	272	24.1	9.1	2.9	2.9	8.6	0.8	6.5	2.0	43.2	100.1	6,463	6,405
AB18026	7/8/2005 2:00	289	20.3	8.0	2.6	2.4	8.5	1.0	5.2	1.8	48.9	98.6	7,524	7,455
AB18027	7/8/2005 6:00	293	19.8	7.9	2.5	2.5	8.6	0.9	4.9	1.7	50.5	99.5	7,599	7,529
AB18032	7/8/2005 9:00	296	20.9	7.9	2.4	2.6	7.4	1.0	5.2	1.6	50.3	99.2	7,655	7,586
AB18087	7/8/2005 22:00	309	22.4	8.4	2.7	2.7	8.3	0.6	6.0	1.8	46.9	99.9	6,991	6,922
AB18093	7/9/2005 22:00	333	20.6	8.0	2.5	2.6	8.3	0.6	5.5	1.8	47.1	97.1	7,268	7,198
AB18095	7/10/2005 6:00	341	21.4	8.5	2.6	2.7	7.6	0.7	6.0	1.8	44.5	95.8	6,843	6,779
AB18096	7/10/2005 10:00	345	24.1	9.1	2.8	2.8	7.4	0.6	7.0	1.9	39.6	95.4	6,527	6,470
AB18127	7/12/2005 6:00	374	26.1	9.6	3.0	2.9	7.2	0.5	8.5	2.0	38.7	98.5	6,006	5,947
AB18128	7/12/2005 10:00	378	25.3	9.5	2.9	2.9	8.1	0.5	7.1	1.9	40.1	98.3	6,160	6,098
AB18142	7/13/2005 10:00	402	21.1	8.1	2.6	2.6	7.8	0.6	7.0	1.9	47.9	99.5	7,141	7,074
AB18163	7/13/2005 22:00	414	19.9	8.0	2.7	2.4	8.1	0.7	6.3	1.6	49.5	99.3	7,537	7,466
AB18164	7/14/2005 2:00	418	17.6	7.4	2.4	2.2	7.9	0.7	5.3	1.6	52.7	97.9	7,988	7,910
AB18165	7/14/2005 6:00	422	18.2	7.3	2.6	2.2	7.3	0.9	5.8	1.7	50.4	96.4	7,939	7,865
AB18177	7/14/2005 18:00	434	12.7	3.6	6.7	2.9	6.4	1.6	8.9	0.8	52.7	96.3	7,974	7,894
AB18316	7/15/2005 18:00	458	17.2	7.6	2.5	2.0	7.5	1.9	4.3	1.5	51.9	96.4	8,140	8,063
AB18318	7/16/2005 6:00	470	17.7	7.8	2.6	2.4	7.3	1.5	5.3	1.6	51.9	98.1	8,035	7,965
AB18323	7/17/2005 2:00	490	20.5	9.1	3.1	3.0	6.9	1.1	8.8	1.9	41.4	95.8	6,408	6,348
AB18328	7/17/2005 22:00	510	22.4	9.6	3.1	2.7	6.5	1.2	8.8	1.8	42.8	98.9	6,428	6,367

Table 3.1-9 PCD Solids Analysis
(Page 2 of 3)

Sample Number	Sample[2,3] Date & Time	Sample Run Time Hours	SiO_2 Wt. %	Al_2O_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
AB18332	7/18/2005 18:00	530	23.6	9.9	3.4	2.7	5.7	1.5	9.8	2.0	38.9	97.4	5,732	5,677
AB18366	7/19/2005 6:00	542	24.0	9.9	3.5	2.7	6.1	1.6	9.5	2.0	39.9	99.3	5,904	5,846
AB18382	7/19/2005 22:00	558	24.4	10.4	3.8	2.9	5.6	1.5	10.7	1.9	38.1	99.2	5,598	5,544
AB18409	7/21/2005 6:00	590	20.6	9.3	3.2	2.4	5.5	1.5	8.4	1.6	45.0	97.6	6,675	6,612
AB18457	7/22/2005 6:00	614	23.5	10.6	3.9	2.8	4.8	1.9	11.2	2.4	38.0	99.1	5,612	5,562
AB18470	7/22/2005 14:00	622	22.4	10.2	3.7	2.6	5.5	1.9	10.0	2.1	38.3	96.6	5,832	5,780
AB18484	7/23/2005 2:00	634	16.9	8.1	2.6	2.2	4.6	1.7	6.6	1.6	51.7	96.0	7,951	7,882
AB18485	7/23/2005 6:00	638	17.9	8.4	2.8	2.2	4.1	1.6	7.4	1.8	51.2	97.3	7,775	7,706
AB18487	7/23/2005 14:00	646	18.2	8.5	2.8	2.1	4.2	1.5	7.6	1.6	50.3	96.7	7,572	7,503
AB18523	7/23/2005 22:00	654	20.0	9.4	3.0	2.6	4.2	1.3	9.0	1.7	45.4	96.6	6,760	6,699
AB18525	7/24/2005 6:00	662	19.2	9.3	2.6	2.3	4.3	0.9	7.8	1.6	48.4	96.4	7,197	7,135
AB18530	7/25/2005 2:00	682	19.9	9.4	2.8	2.5	5.2	1.0	8.1	1.8	45.5	96.0	6,946	6,886
AB18557	7/25/2005 8:00	688	21.3	10.3	3.0	2.4	6.3	1.0	8.5	1.7	43.0	97.6	6,596	6,547
AB18559	7/25/2005 9:00	689	21.0	10.2	2.9	2.6	7.7	1.0	7.5	1.9	42.9	97.9	6,662	6,609
AB18561	7/25/2005 10:00	690	21.3	9.7	3.0	2.6	7.6	1.1	7.4	1.9	43.7	98.2	6,545	6,492
AB18563	7/25/2005 11:00	691	21.0	9.9	2.9	2.6	8.6	1.0	6.9	2.0	43.8	98.8	6,701	6,647
AB18602	7/25/2005 22:00	702	21.7	10.3	3.1	2.7	7.5	1.1	8.1	2.1	41.5	98.1	6,283	6,238
AB18603	7/26/2005 2:00	706	21.8	10.6	3.1	2.6	7.2	1.0	8.2	2.0	41.9	98.5	6,360	6,309
AB18604	7/26/2005 7:00	711	22.2	10.2	2.6	2.8	6.8	1.0	7.1	2.9	41.7	97.4	6,406	6,340
AB18605	7/26/2005 7:30	711	20.7	9.7	2.9	2.7	7.3	1.2	7.6	1.9	43.2	97.4	6,721	6,666
AB18607	7/26/2005 8:30	712	20.5	9.9	3.0	2.6	7.9	1.1	7.6	2.0	43.1	97.7	6,659	6,605
AB18609	7/26/2005 9:30	714	21.1	10.6	3.1	2.7	7.7	1.2	8.1	1.9	41.8	98.1	6,455	6,403
AB18611	7/26/2005 10:30	714	20.7	10.1	3.0	2.8	6.6	1.1	8.5	2.0	42.9	97.7	6,555	6,504
AB18612	7/26/2005 11:00	715	21.8	9.9	2.6	2.8	5.8	1.1	7.4	2.8	42.5	96.8	6,531	6,463
AB18613	7/26/2005 14:00	718	22.2	10.1	2.7	2.8	5.5	1.2	7.7	2.9	43.4	98.6	6,415	6,345
AB18625	7/26/2005 18:00	722	20.6	9.9	3.1	2.5	6.7	1.1	8.7	1.9	43.4	97.8	6,645	6,592
AB18639	7/27/2005 2:00	730	20.8	10.0	3.0	2.4	8.5	1.1	7.6	2.0	42.7	98.0	6,553	6,496
AB18641	7/27/2005 10:00	738	24.0	11.4	3.4	2.8	9.0	0.8	8.9	2.4	35.4	98.1	5,595	5,549
AB18663	7/28/2005 2:00	754	23.8	11.0	3.4	2.9	8.1	0.6	9.5	2.1	36.4	97.8	5,603	5,557
AB18688	7/28/2005 14:00	766	18.6	9.0	2.8	2.3	8.3	0.9	6.6	1.7	49.7	99.8	7,458	7,397
AB18701	7/29/2005 6:00	782	19.9	9.9	3.0	2.5	8.5	0.9	7.6	1.9	44.1	98.2	6,724	6,666
AB18749	7/30/2005 2:00	802	23.7	11.3	3.5	2.6	8.3	0.9	9.6	2.1	37.0	99.0	5,557	5,508
AB18756	7/31/2005 6:00	830	20.6	8.5	2.7	2.4	7.4	1.2	6.6	1.6	47.3	98.3	7,253	7,192
AB18758	7/31/2005 9:12	833	22.8	9.1	2.8	2.4	8.3	1.0	6.7	1.7	43.8	98.5	6,756	6,701
AB18760	7/31/2005 10:10	834	23.5	9.4	3.0	2.9	8.6	1.1	7.4	1.8	41.4	99.1	6,338	6,291
AB18762	7/31/2005 11:12	835	23.8	9.3	3.0	2.5	8.1	1.1	7.5	1.8	41.0	98.1	6,412	6,366

Table 3.1-9 PCD Solids Analysis
(Page 3 of 3)

Sample Number	Sample[2,3] Date & Time	Sample Run Time Hours	SiO_2 Wt. %	Al_2O_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
AB18764	7/31/2005 12:14	836	23.3	9.2	2.9	2.6	8.1	1.1	7.6	1.9	41.7	98.5	6,464	6,410
AB18768	8/1/2005 2:00	850	18.8	7.8	2.4	2.2	8.0	1.1	5.1	1.4	50.6	97.4	7,888	7,826
AB18771	8/1/2005 8:30	856	20.5	8.4	2.6	2.3	7.5	1.1	6.2	1.6	48.0	98.3	7,933	7,870
AB18773	8/1/2005 9:30	858	20.6	8.5	2.6	2.4	7.2	1.1	6.6	1.6	46.6	97.1	7,333	7,274
AB18776	8/1/2005 10:30	858	20.5	8.3	2.7	2.4	8.0	1.1	6.2	1.7	47.3	98.1	7,282	7,226
AB18778	8/1/2005 11:30	859	20.2	8.6	2.5	2.4	8.1	1.2	5.8	1.5	47.8	98.2	7,358	7,297
AB18857	8/1/2005 18:00	866	18.6	8.1	2.5	2.2	8.0	1.1	5.6	1.7	50.3	98.0	7,878	7,818
AB18939	8/4/2005 18:00	938	24.9	10.6	3.2	2.8	7.9	1.2	8.8	1.8	37.5	98.9	5,802	5,755
AB18959	8/5/2005 10:00	954	26.4	11.2	3.3	3.0	5.3	1.3	10.4	2.0	36.2	99.1	5,526	5,482
AB18964	8/5/2005 14:00	958	25.6	11.3	3.4	3.1	5.4	1.4	10.1	1.9	36.1	98.3	5,537	5,486
AB18970	8/5/2005 22:00	966	29.2	13.1	3.7	3.4	5.8	1.2	10.7	1.9	29.2	98.1	4,692	4,657
AB18989	8/6/2005 10:00	978	26.0	11.4	3.3	2.9	6.9	1.3	9.2	1.9	35.2	98.1	5,539	5,489
AB18997	8/6/2005 22:00	990	25.1	11.4	3.4	2.8	6.8	1.6	8.7	1.8	36.9	98.4	5,658	5,606
AB18998	8/7/2005 2:00	994	25.7	11.5	3.4	3.0	5.0	1.5	9.7	1.9	36.3	98.1	5,590	5,539
AB19011	8/7/2005 10:00	1002	23.7	10.5	3.1	2.8	5.1	1.4	8.5	1.8	41.5	98.3	6,306	6,252
AB19025	8/7/2005 18:00	1010	24.2	11.0	3.2	2.8	6.7	1.6	7.8	1.6	40.0	99.0	6,027	5,973
AB19026	8/7/2005 22:00	1014	26.2	11.9	3.3	2.8	6.5	1.4	8.4	1.9	37.3	99.7	5,370	5,317
AB19030	8/8/2005 14:00	1030	24.0	10.0	3.1	2.8	8.2	1.5	6.7	1.7	41.3	99.2	6,031	5,966
AB19048	8/8/2005 22:00	1038	24.5	11.3	3.2	2.9	8.3	1.5	7.0	1.6	38.9	99.2	5,812	5,756
AB19058	8/9/2005 2:00	1042	25.3	11.7	3.2	2.8	6.5	1.4	7.9	1.5	38.2	98.6	5,641	5,588
AB19106	8/10/2005 18:00	1082	34.5	15.4	4.6	4.0	6.5	0.2	14.5	2.5	18.4	100.6	2,594	2,571
AB19126	8/11/2005 6:00	1094	22.8	10.6	3.1	2.7	8.7	1.1	7.6	1.9	40.0	98.5	6,079	6,024
AB19128	8/11/2005 14:00	1102	19.2	9.6	3.1	2.6	8.7	1.5	7.4	1.9	44.7	98.6	6,882	6,817
AB19146	8/12/2005 6:00	1118	21.1	10.0	3.0	2.6	9.0	1.1	7.5	2.0	42.4	98.6	6,552	6,490
AB19149	8/12/2005 14:00	1126	17.1	8.6	2.8	2.2	8.3	1.4	6.5	1.9	49.0	97.8	7,590	7,516
AB19163	8/12/2005 18:00	1130	18.2	9.3	2.9	2.3	8.1	1.4	7.3	1.8	45.8	97.3	7,193	7,127
AB19170	8/13/2005 2:00	1138	18.1	9.2	3.0	2.5	8.6	1.3	7.2	1.9	46.9	98.5	7,175	7,105
AB19274	8/17/2005 22:00	1223	23.9	9.3	2.6	2.3	7.6	1.2	5.8	1.5	44.5	98.8	6,735	6,669
AB19296	8/18/2005 10:00	1235	25.4	10.0	3.0	2.6	6.9	1.2	7.8	1.8	39.8	98.5	6,111	6,053
AB19305	8/18/2005 22:00	1247	27.9	10.1	2.9	2.8	8.0	1.1	7.2	1.8	37.6	99.3	5,700	5,644
AB19307	8/19/2005 6:00	1255	26.2	10.0	2.8	2.6	8.8	1.0	5.9	1.6	41.1	99.8	6,250	6,191
AB19326	8/19/2005 10:00	1259	25.8	10.0	2.8	2.5	8.3	1.0	6.4	1.8	41.0	99.5	6,429	6,372
AB19394	8/20/2005 14:00	1287	27.6	10.6	3.2	2.8	8.2	1.0	8.5	1.9	35.5	99.3	5,505	5,456
AB19397	8/21/2005 2:00	1299	27.0	11.1	3.1	3.0	6.0	0.9	9.8	1.9	35.3	98.1	5,356	5,307
AB19402	8/21/2005 22:00	1319	25.7	10.8	2.9	2.7	6.3	0.7	8.6	1.8	39.3	98.8	5,933	5,880

Notes:

1. Other inerts consist of P_2O_5, Na_2O, K_2O, BaO, & TiO_2.
2. July 30 to August 13 samples taken while recycle gas compressor was running.
3. Transport air was used as coal conveying gas from August 20 to August 21.

Table 3.1-10 Historical Standpipe and PCD Solids Particle Sizes and Densities

Test Campaign	Fuel	Standpipe Maximum Particle Size SMD microns	Standpipe Steady State Part. Size SMD microns	Standpipe Minimum Bulk Density lb/ft^3	PCD Fines Average Particle Size SMD microns	PCD Fines St. Dev. Particle Size SMD microns	PCD Fines Average Bulk Density lb/ft^3	PCD Fines St. Dev. Bulk Density 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	289	none	71	15.2	2.3	14	3
TC18	Powder River Basin	170	165	82.41	11.4	1.3	15	2

Table 3.1-11 Carbon Conversion and Gasification Efficiencies
(Page 1 of 3)

Operating[1,2] Period	Average Relative Hours	Carbon Conversion %	Efficiency Raw Cold %	Efficiency Raw Hot %	Efficiency Projected[3] Cold %
TC18-1	31	92.8	50.8	82.6	66.3
TC18-2	42	93.0	50.5	82.5	66.8
TC18-3	54	93.1	52.2	83.4	68.2
TC18-4	61	92.9	50.5	82.7	67.2
TC18-5	100	95.1	53.0	86.0	68.0
TC18-6	125	94.3	45.8	82.5	64.7
TC18-7	131	95.1	50.0	84.3	67.6
TC18-8	146	93.5	52.4	84.6	67.5
TC18-9	167	91.6	56.2	83.1	70.9
TC18-10	185	89.6	55.0	81.6	69.1
TC18-11	192	89.6	55.4	81.7	69.2
TC18-12	201	89.5	55.4	81.4	69.3
TC18-13	209	89.0	53.1	80.1	68.0
TC18-14	220	89.3	53.6	80.9	68.1
TC18-15	228	90.3	54.2	81.5	68.8
TC18-16	253	92.0	51.4	80.8	68.6
TC18-17	263	92.3	52.3	81.1	69.5
TC18-18	287	90.8	54.5	80.8	70.2
TC18-19	296	89.1	53.6	79.0	68.7
TC18-20	305	88.8	52.8	79.0	67.6
TC18-21	326	90.2	54.3	80.7	68.7
TC18-22	333	90.2	54.3	80.6	69.0
TC18-23	339	90.3	53.8	80.6	68.4
TC18-24	345	89.4	51.6	78.7	67.3
TC18-25	354	90.0	51.2	78.8	67.4
TC18-26	374	92.0	50.1	81.5	67.4
TC18-27	391	91.1	52.5	82.3	68.1
TC18-28	399	89.8	51.7	81.1	67.1
TC18-29	413	89.7	52.5	81.2	67.6
TC18-30	417	90.0	53.2	81.7	67.9
TC18-31	422	90.5	53.5	82.0	67.9
TC18-32	434	88.7	52.0	79.9	67.6
TC18-33	461	88.4	51.7	79.2	67.9
TC18-34	469	89.1	52.8	80.0	69.2
TC18-35	490	91.8	53.0	81.8	70.7

Table 3.1-11 Carbon Conversion and Gasifier Efficiencies
(Page 2 of 3)

Operating[1,2] Period	Average Relative Hours	Carbon Conversion %	Efficiency Raw Cold %	Efficiency Raw Hot %	Efficiency Projected[3] Cold %
TC18-36	509	92.4	52.2	82.0	70.7
TC18-37	530	93.5	54.7	82.9	72.3
TC18-38	541	92.6	52.6	81.7	70.5
TC18-39	556	92.4	53.9	81.8	71.1
TC18-40	588	90.5	52.3	81.1	68.9
TC18-41	611	91.8	49.6	79.7	70.9
TC18-42	620	92.6	50.5	80.0	72.3
TC18-43	632	92.6	57.0	82.5	73.5
TC18-44	638	92.3	53.6	81.8	72.3
TC18-45	645	92.3	52.4	81.7	71.3
TC18-46	653	92.8	52.9	81.9	72.6
TC18-47	659	92.0	51.5	80.6	72.1
TC18-48	681	93.5	53.8	83.1	72.3
TC18-49	689	93.7	54.4	83.4	72.5
TC18-50	699	93.5	54.6	83.4	72.3
TC18-51	708	93.6	54.6	83.5	71.8
TC18-52	717	92.4	54.6	82.8	70.8
TC18-53	729	92.6	54.4	82.2	71.5
TC18-54	735	94.1	53.4	83.4	70.9
TC18-55	754	92.6	52.1	83.0	69.7
TC18-56	765	90.1	52.4	80.7	68.6
TC18-57	783	92.5	52.4	82.4	69.9
TC18-58	801	93.4	54.4	84.3	70.8
TC18-59	825	91.5	54.5	82.7	70.3
TC18-60	834	91.9	53.9	82.7	69.9
TC18-61	850	91.0	54.4	81.9	70.0
TC18-62	857	90.8	54.2	82.0	69.3
TC18-63	863	90.4	54.5	81.7	69.6
TC18-64	918	92.4	54.3	81.1	72.1
TC18-65	924	92.4	53.7	81.0	71.6
TC18-66	932	92.7	54.5	81.5	72.0
TC18-67	939	92.8	54.7	81.8	72.0
TC18-68	954	92.7	54.7	81.7	71.9
TC18-69	960	93.2	54.8	82.0	72.4
TC18-70	965	93.7	54.5	82.0	72.6

Table 3.1-11 Carbon Conversion and Gasification Efficiencies
(Page 3 of 3)

Operating[1,2] Period	Average Relative Hours	Carbon Conversion %	Efficiency Raw Cold %	Efficiency Raw Hot %	Efficiency Projected[3] Cold %
TC18-71	979	92.5	54.0	81.2	71.5
TC18-72	988	92.3	54.4	81.2	71.7
TC18-73	993	91.8	53.0	80.3	70.9
TC18-74	1,000	92.0	55.3	81.8	71.3
TC18-75	1,006	92.1	55.8	81.9	71.8
TC18-76	1,014	91.5	53.2	80.4	71.1
TC18-77	1,032	92.2	55.7	82.1	72.2
TC18-78	1,043	92.4	53.1	80.8	71.7
TC18-79	1,079	97.1	52.7	86.0	70.3
TC18-80	1,092	92.3	51.3	79.6	71.4
TC18-81	1,101	91.9	53.7	80.5	71.5
TC18-82	1,109	91.6	54.0	81.1	70.5
TC18-83	1,116	90.0	50.8	78.6	68.3
TC18-84	1,124	90.6	56.2	81.4	70.4
TC18-85	1,131	89.5	53.4	79.4	69.0
TC18-86	1,136	88.4	51.1	77.8	67.6
TC18-87	1,169	90.0	55.1	80.8	69.4
TC18-88	1,223	91.5	52.5	81.3	69.8
TC18-89	1,234	92.2	53.0	82.1	69.8
TC18-90	1,247	92.8	52.6	82.5	70.2
TC18-91	1,254	93.0	52.9	82.5	70.9
TC18-92	1,260	93.2	53.7	82.9	71.4
TC18-93	1,287	93.0	54.3	83.5	71.0
TC18-94	1,296	92.9	53.7	83.4	70.5
TC18-95	1,317	92.8	54.1	83.3	70.9

Notes:
1. Recycle gas was used as gasifier aeration during TC18-58 to TC18-87.
2. Transport air was used as coal conveying gas during TC18-94 and TC18-95.
3. Projection assumes that only air nitrogen in the syngas is from air and that the gasifier is adiabatic.

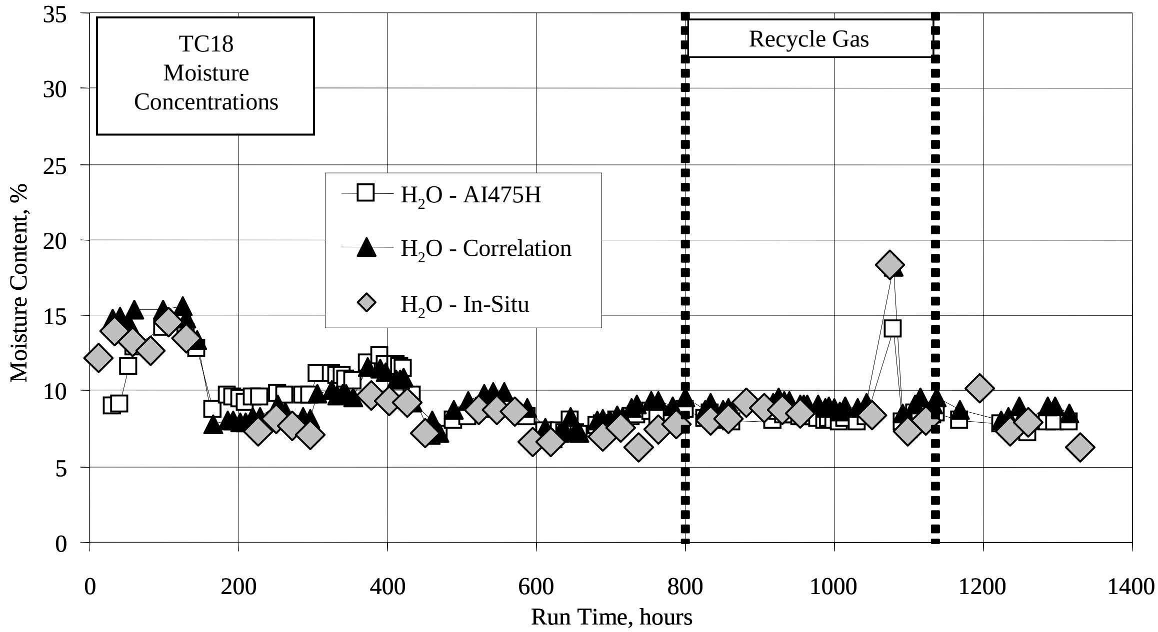

Figure 3.1-1 Syngas H_2O Concentrations

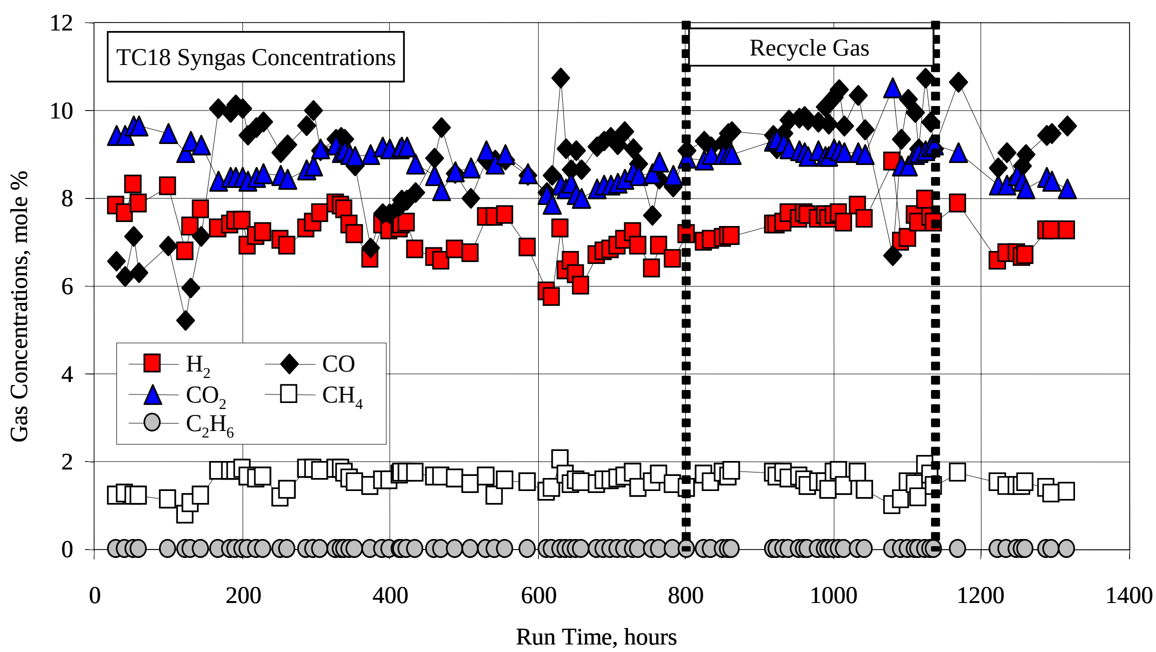

Figure 3.1-2 Wet Syngas Composition

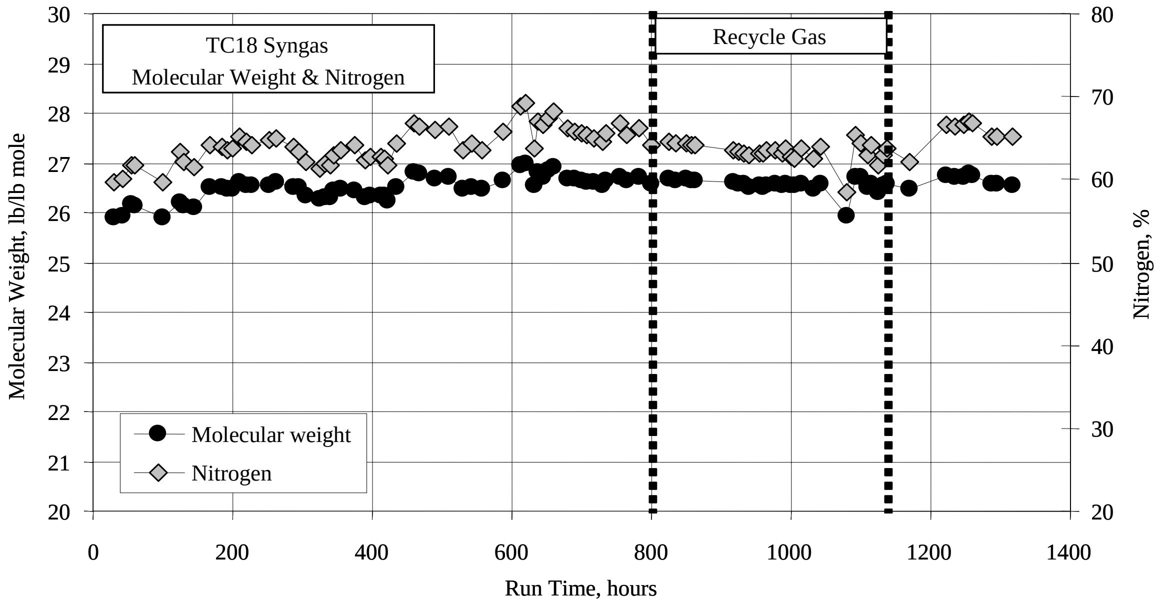

Figure 3.1-3 Wet Syngas Molecular Weight & Nitrogen Concentration

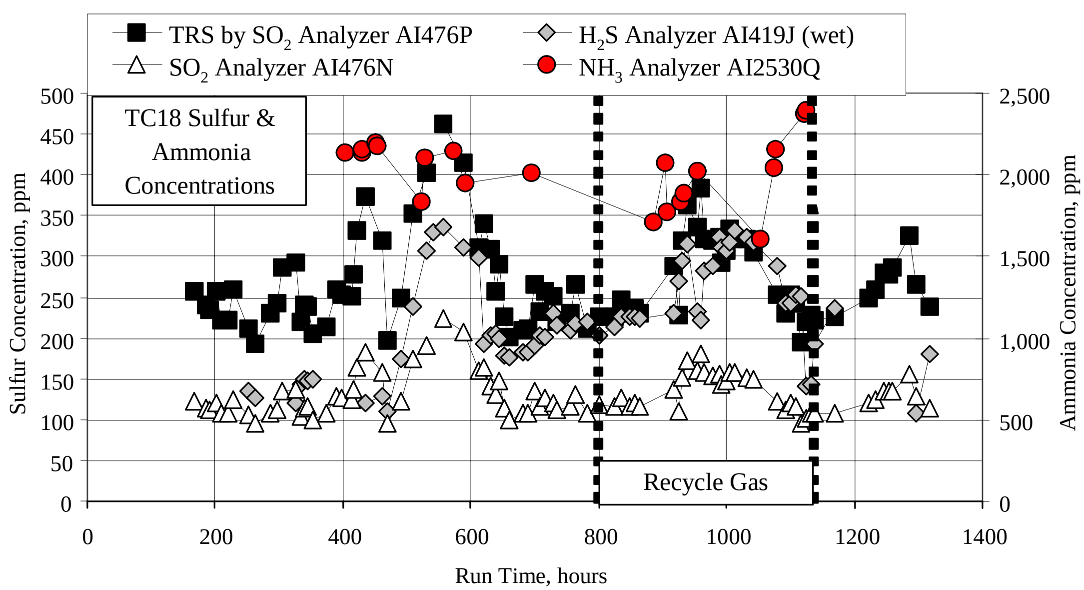

Figure 3.1-4 Syngas Sulfur Concentration

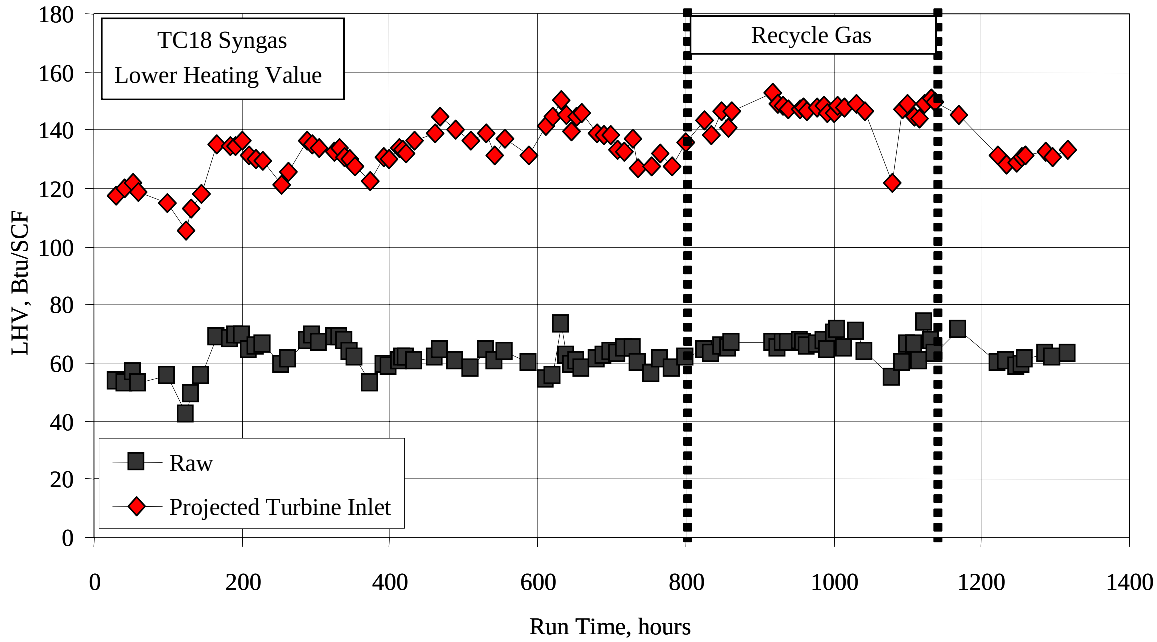

Figure 3.1-5 Syngas Lower Heating Values

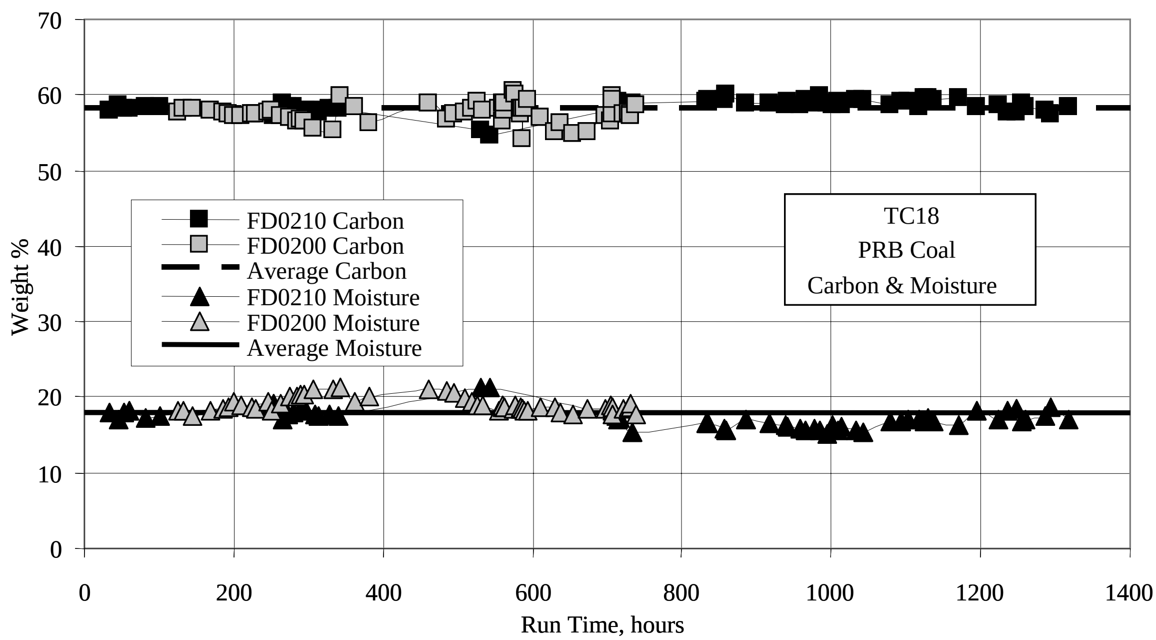

Figure 3.1-6 Coal Carbon & Moisture

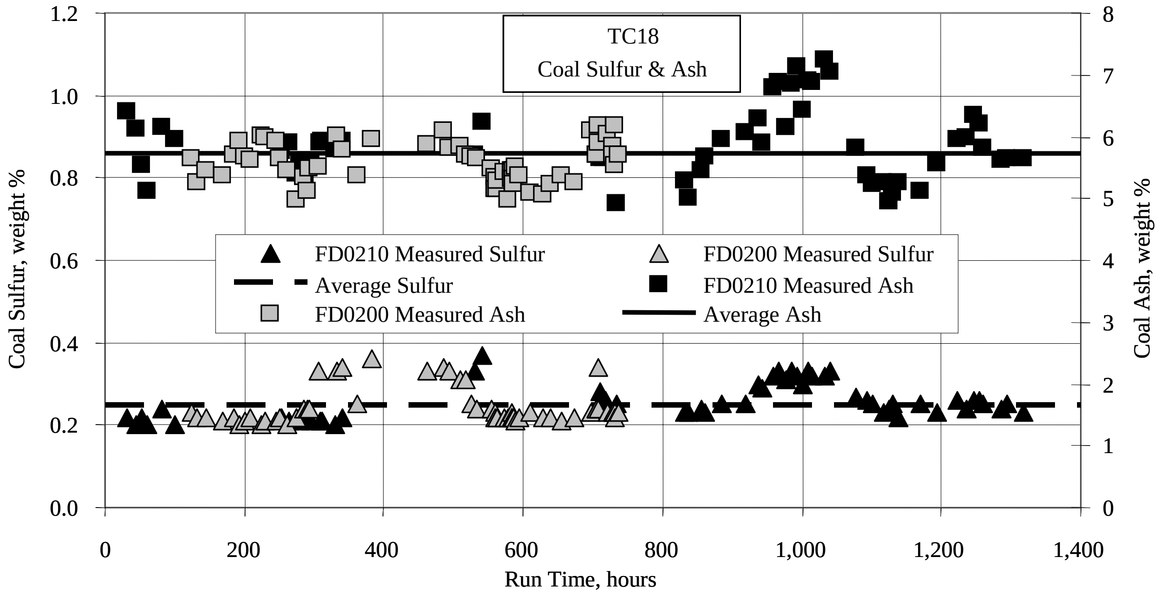

Figure 3.1-7 Coal Sulfur & Ash

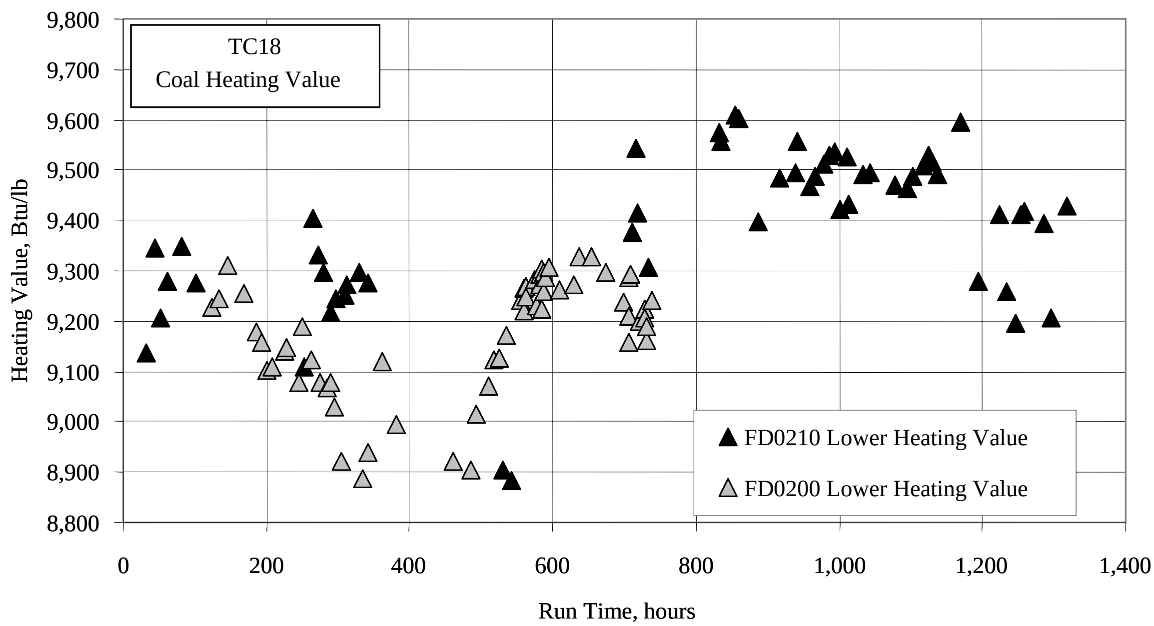

Figure 3.1-8 Coal Lower Heating Value

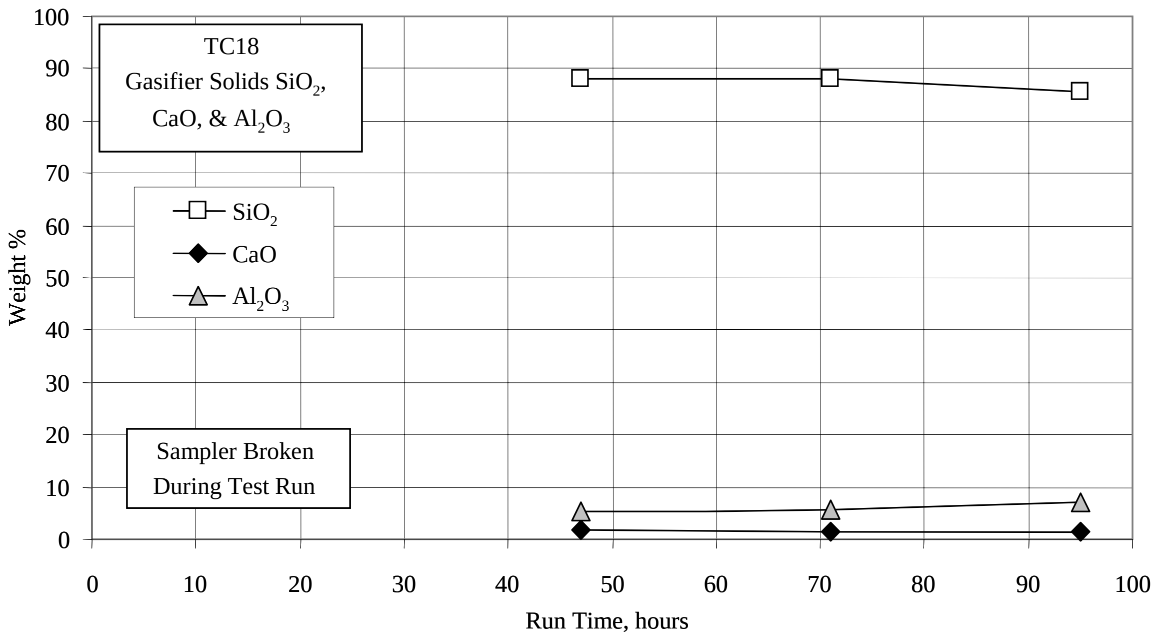

Figure 3.1-9 Standpipe Solids SiO_2, CaO, & Al_2O_3

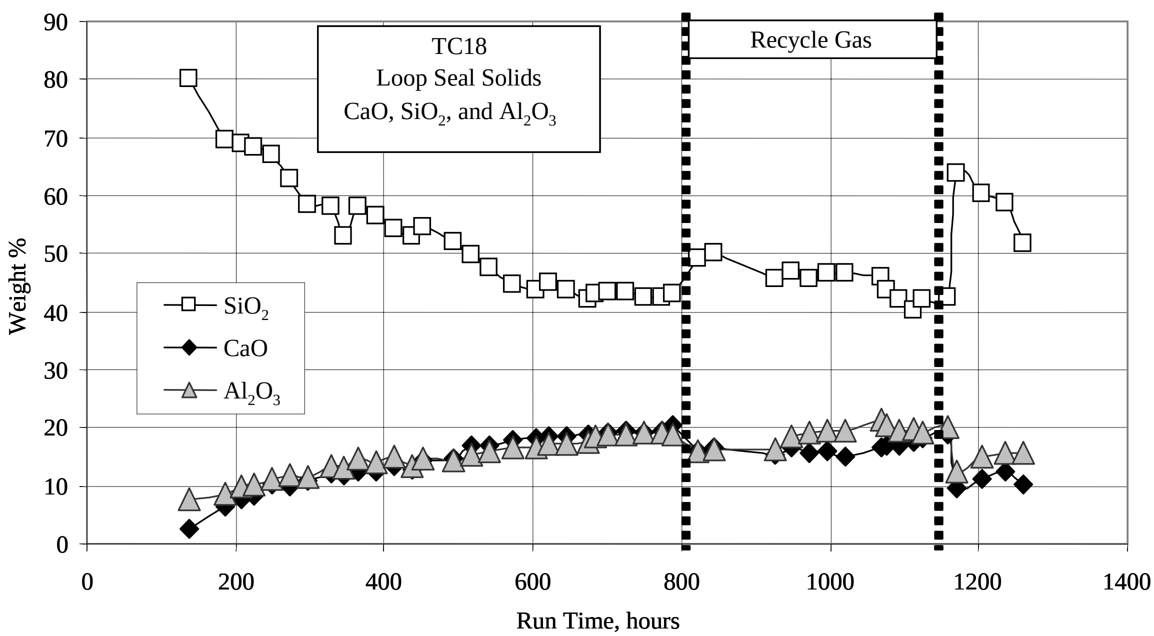

Figure 3.1-10 Loop Seal Solids SiO_2, CaO, & Al_2O_3

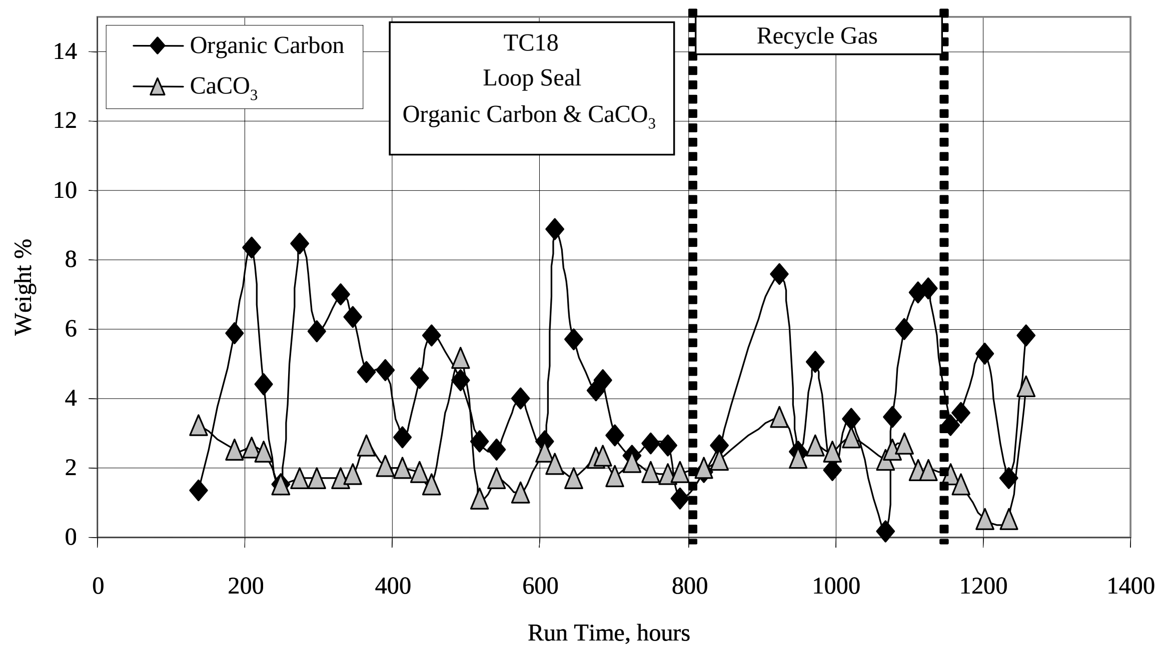

Figure 3.1-11 Loop Seal Solids Organic Carbon and $CaCO_3$

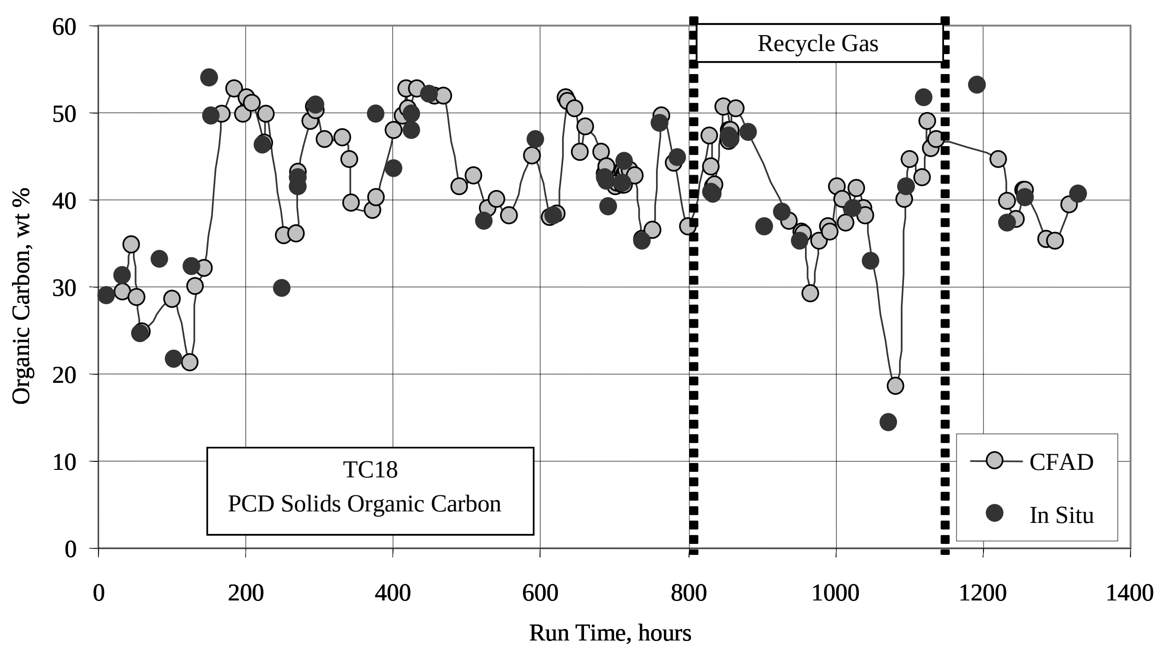

Figure 3.1-12 PCD Solids Organic Carbon

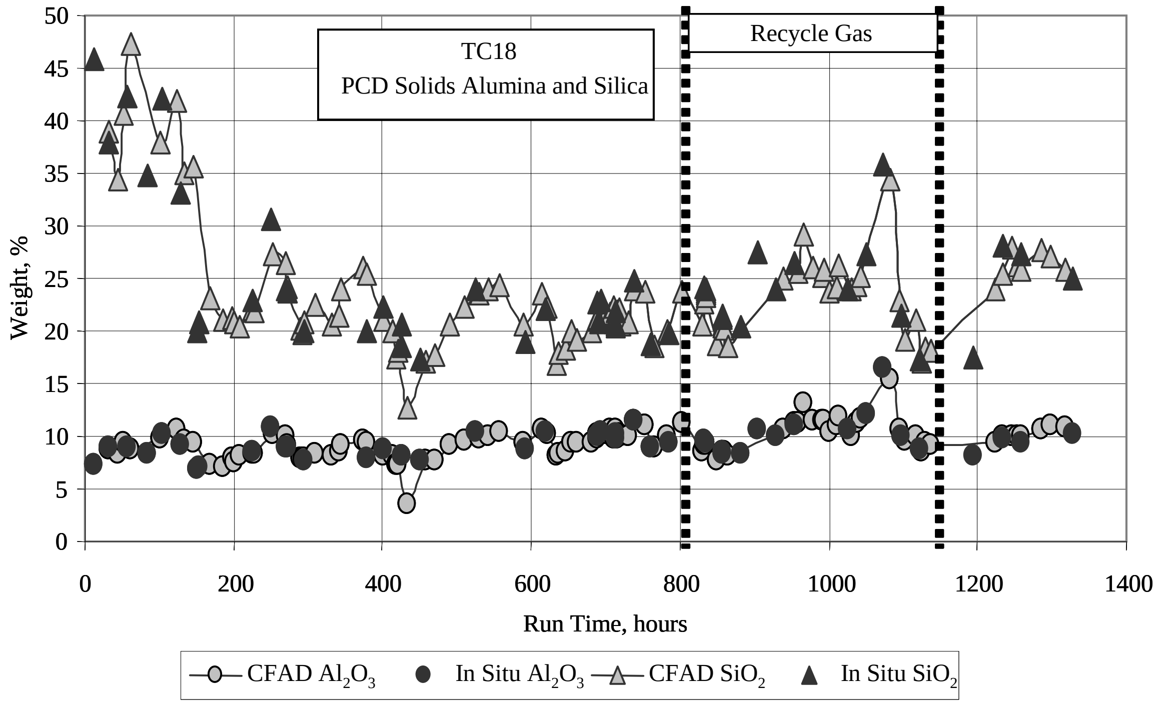

Figure 3.1-13 PCD Solids Alumina & Silica

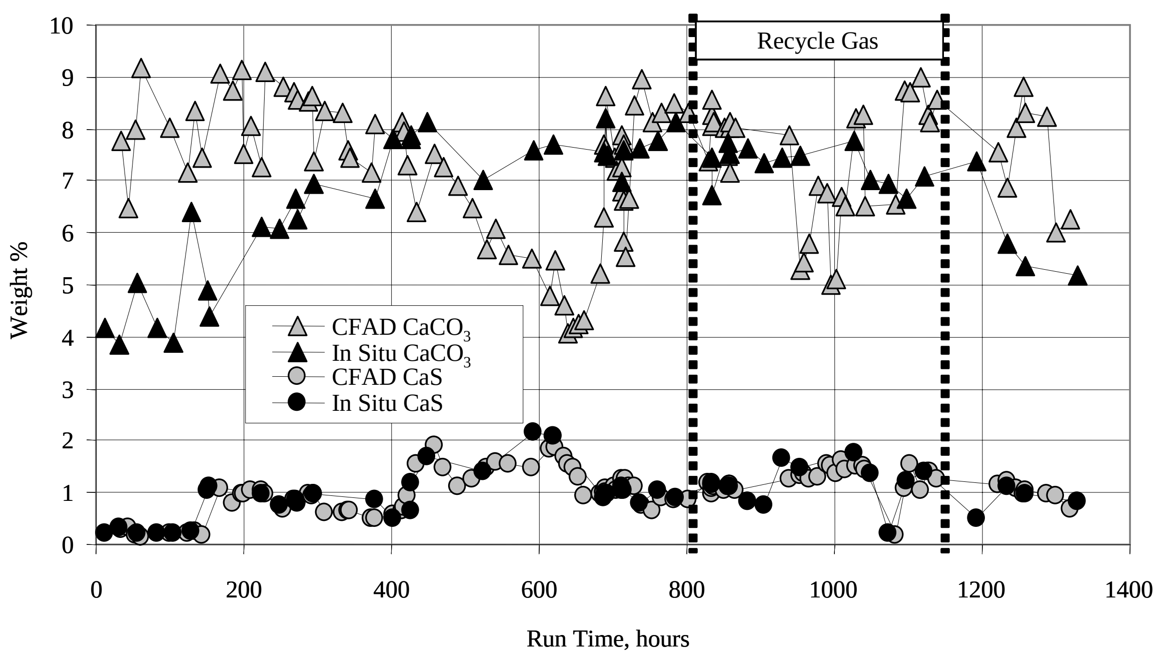

Figure 3.1-14 PCD Solids $CaCO_3$ & CaS

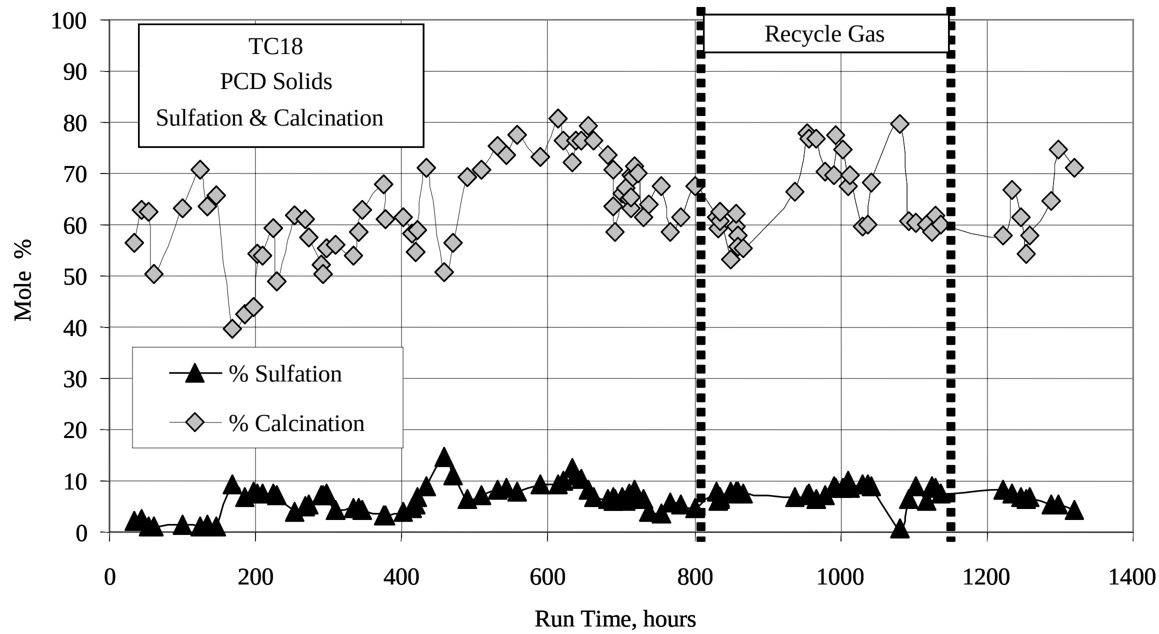

Figure 3.1-15 PCD Solids Sulfation & Calcination

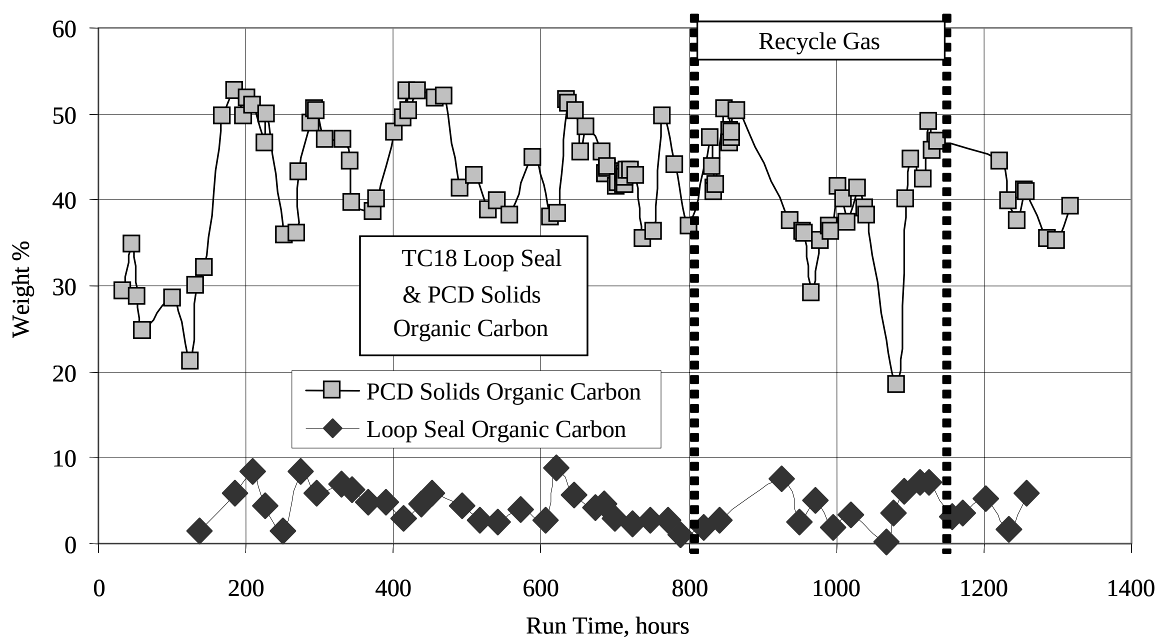

Figure 3.1-16 Loop Seal & PCD Solids Organic Carbon

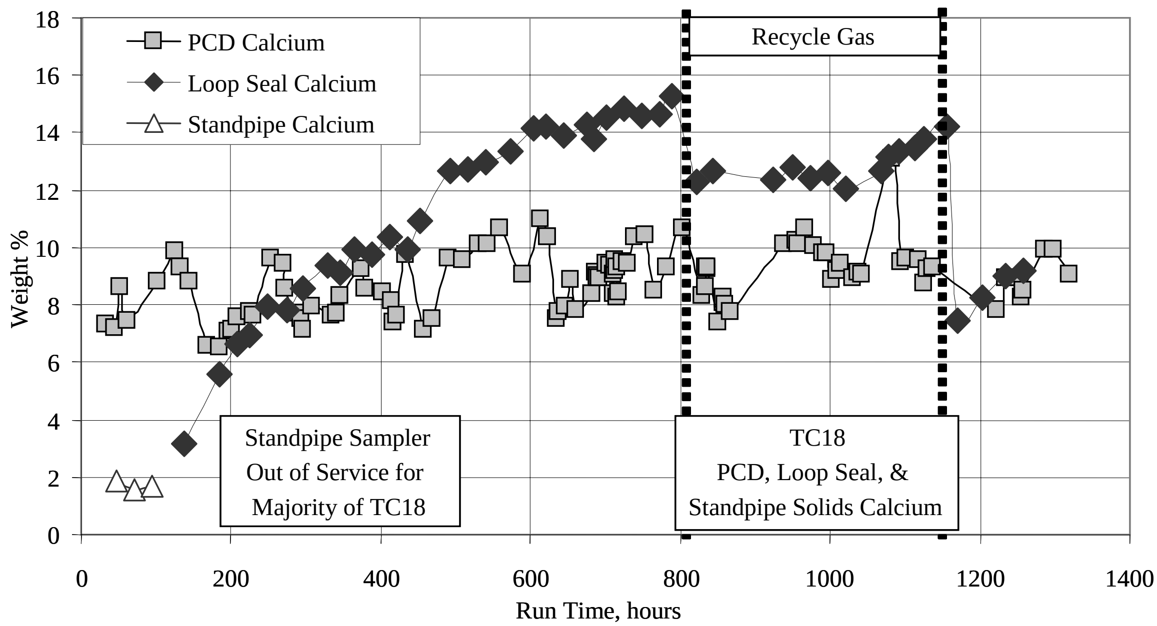

Figure 3.1-17 Standpipe, Loop Seal, & PCD Solids Calcium

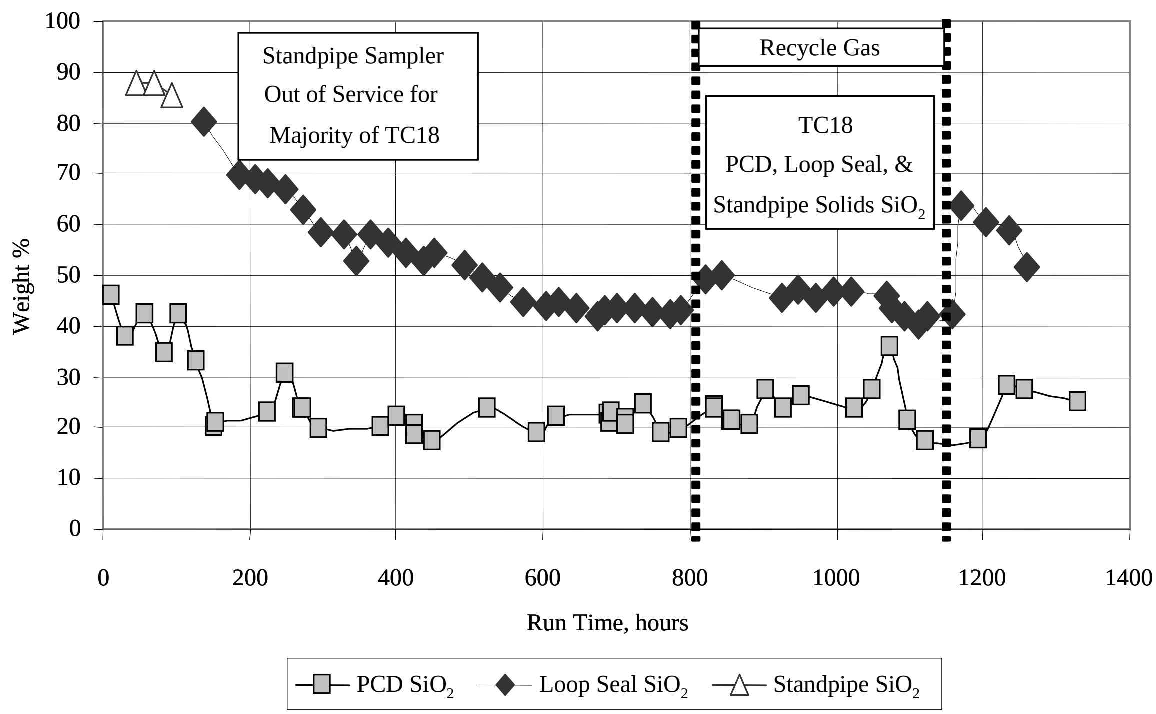

Figure 3.1-18 Standpipe, Loop Seal, & PCD Solids Silica

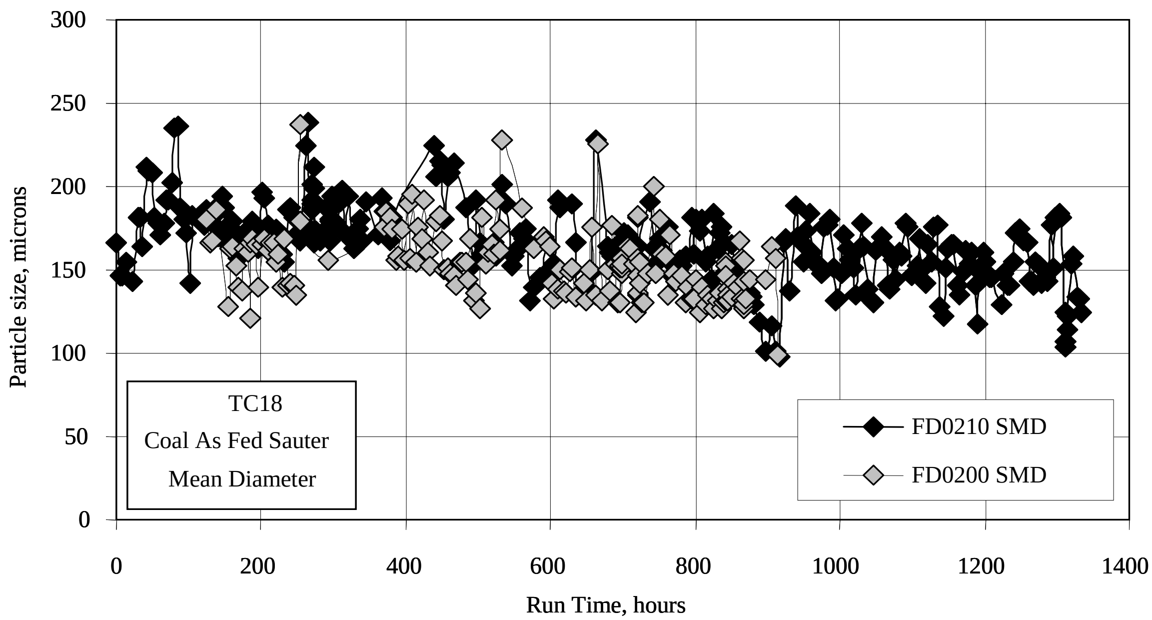

Figure 3.1-19 Coal Sauter Mean Diameter Particle Size

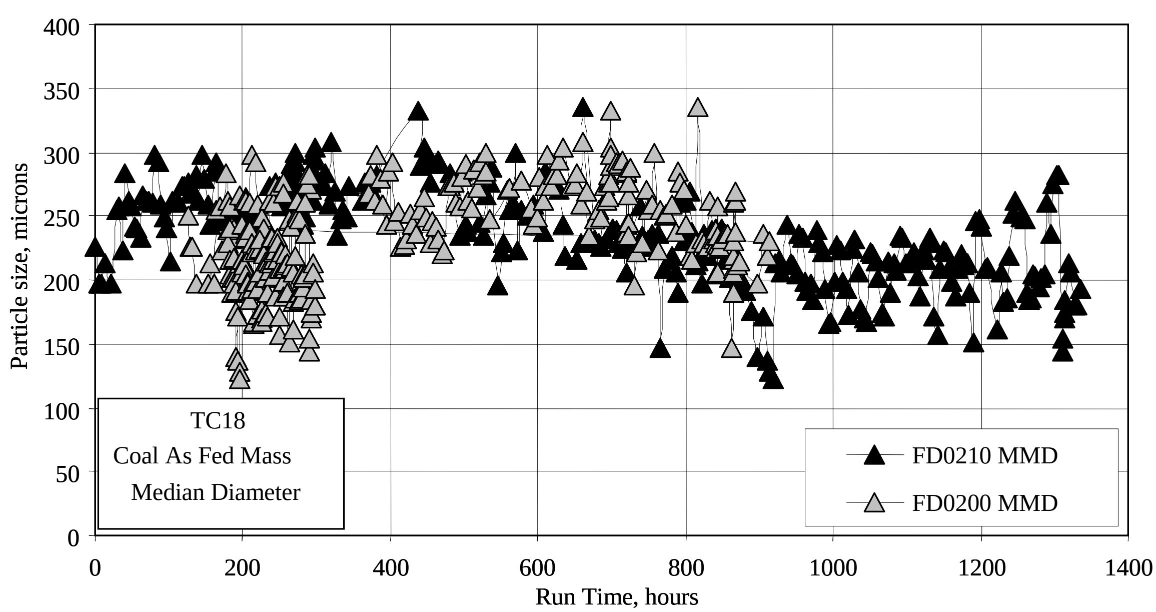

Figure 3.1-20 Coal Mass Median Diameter Particle Size

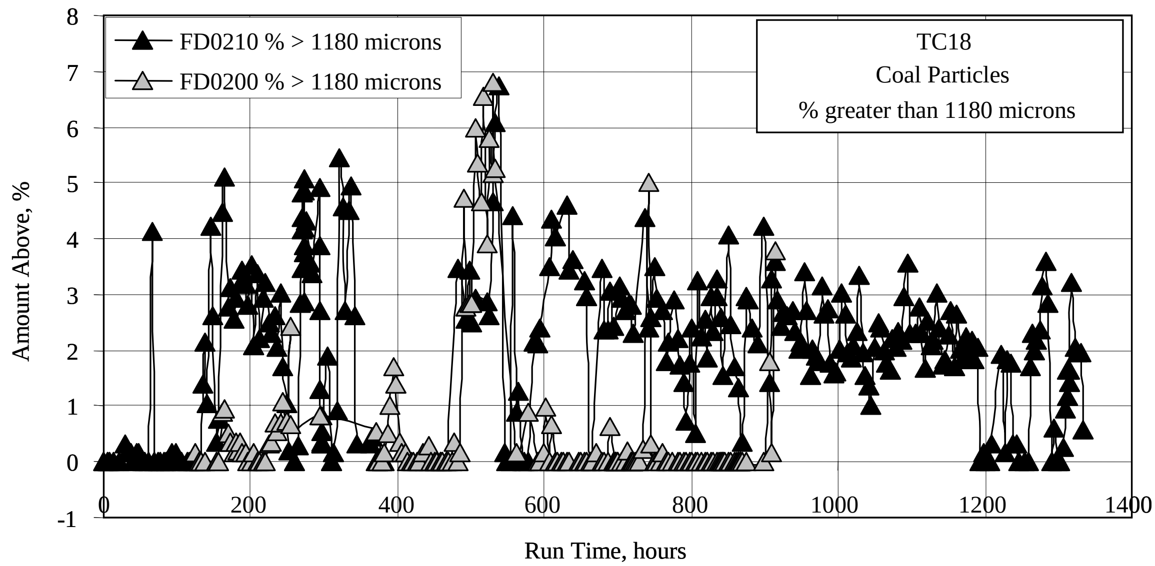

Figure 3.1-21 Percent Coal Oversize

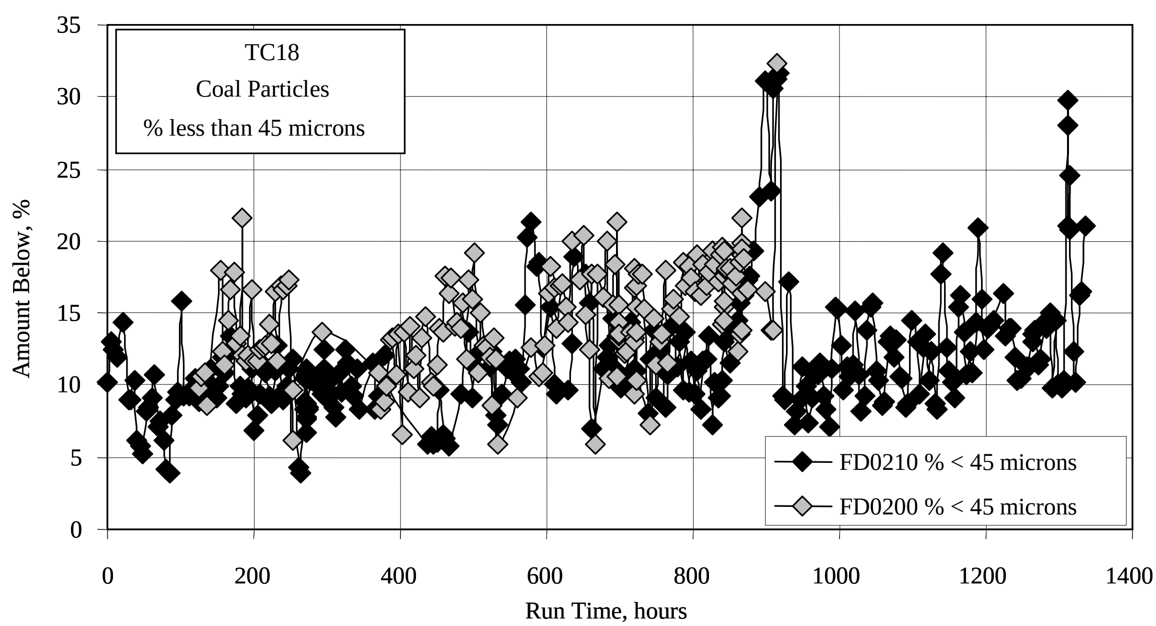

Figure 3.1-22 Percent Coal Fines

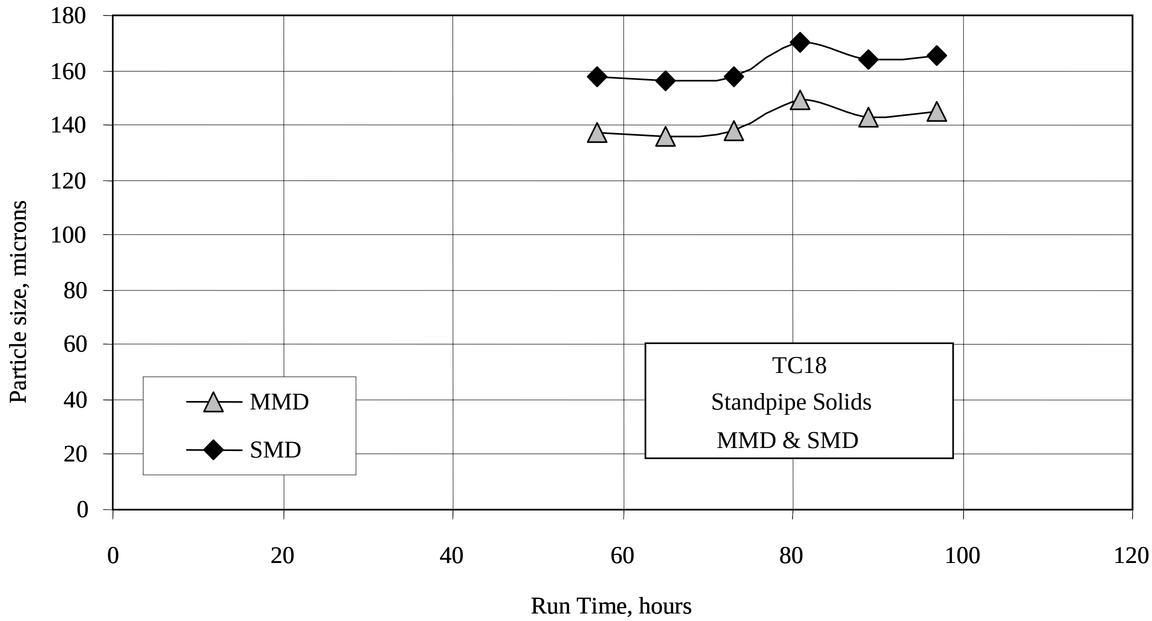

Figure 3.1-23 Standpipe Solids Particle Size

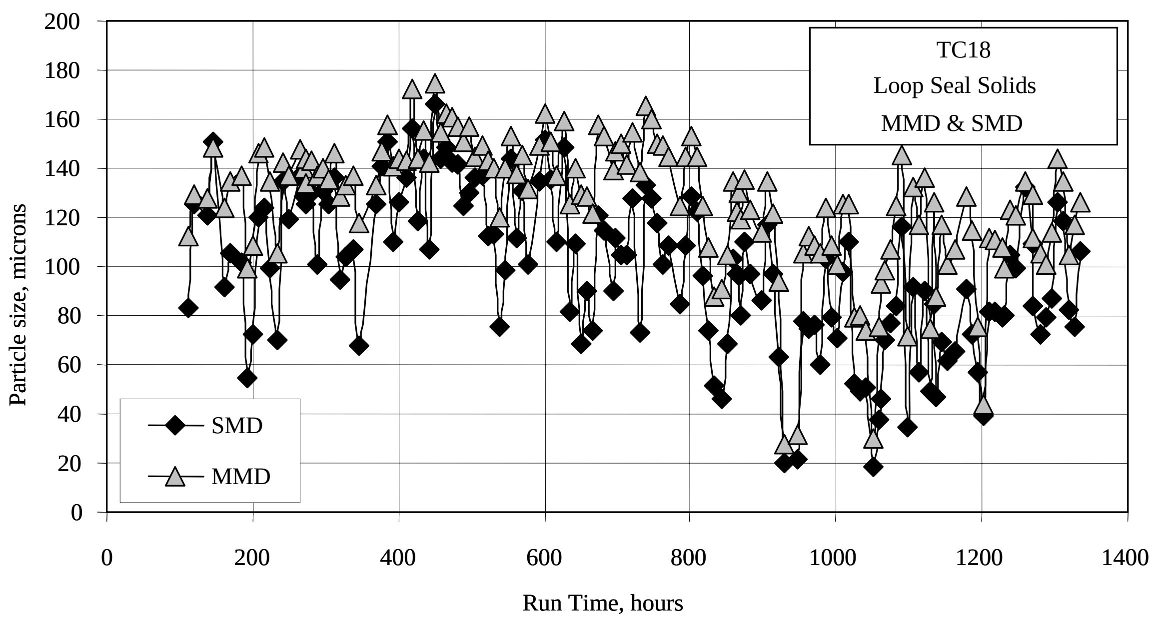

Figure 3.1-24 Loop Seal Solids Particle Sizes

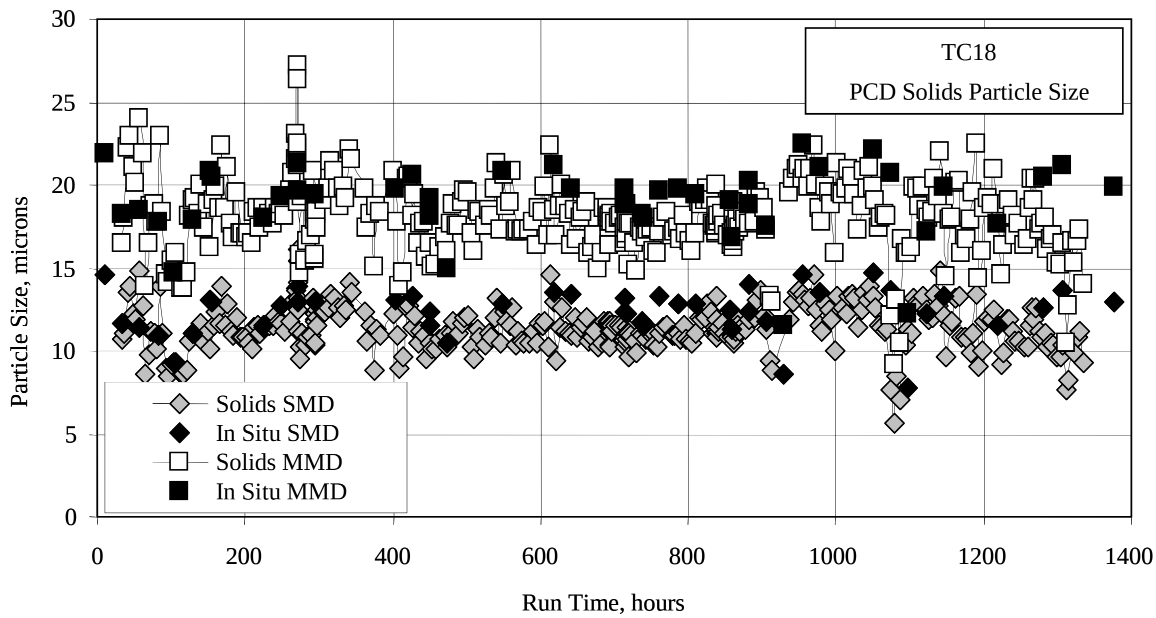

Figure 3.1-25 PCD Solids Particle Sizes

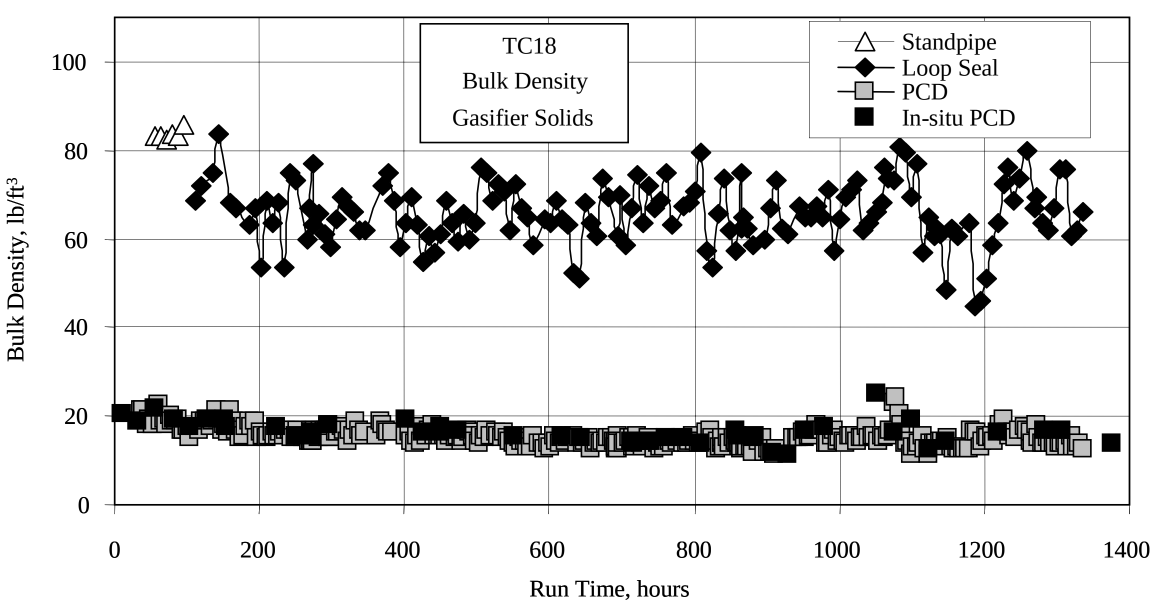

Figure 3.1-26 Standpipe, Loop Seal, & PCD Solids Bulk Density

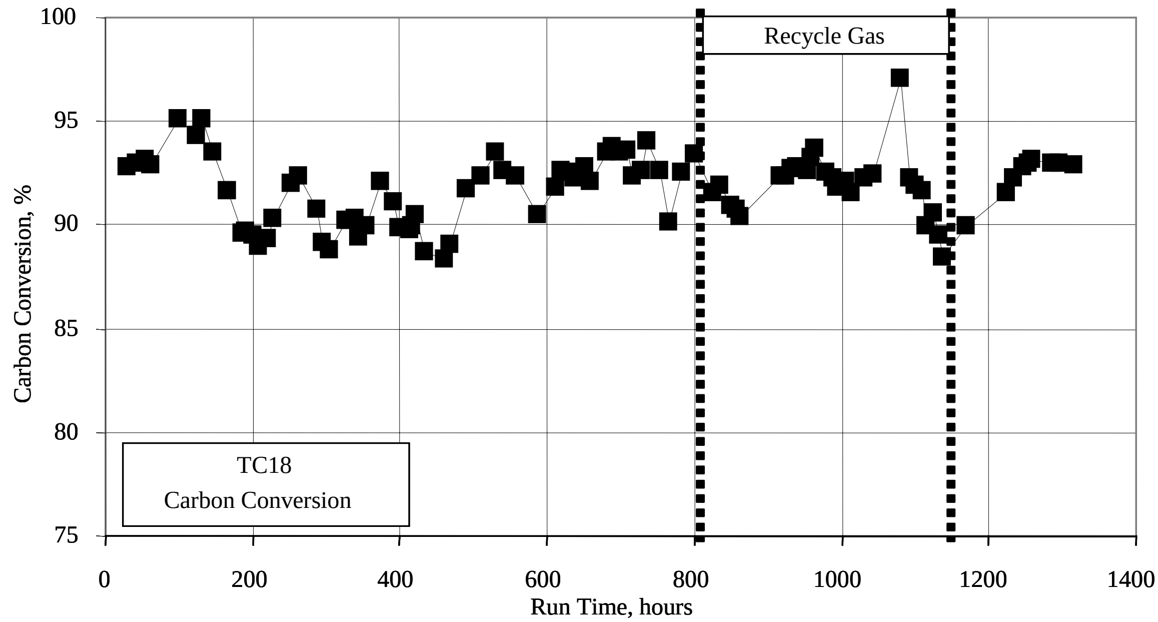

Figure 3.1-27 Carbon Conversion

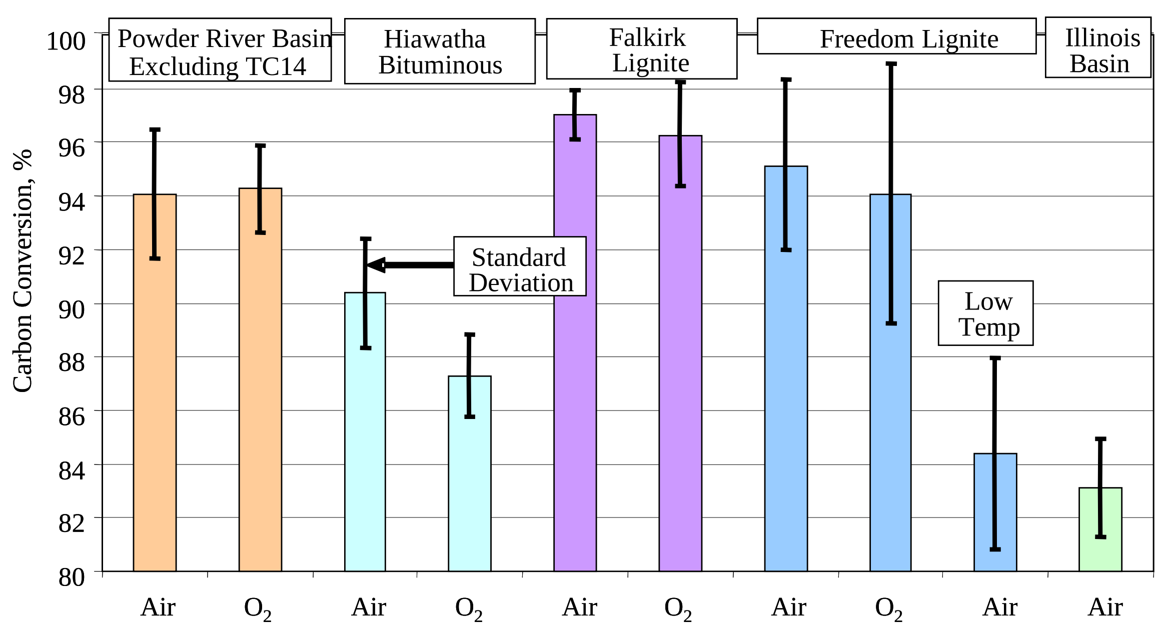

Figure 3.1-28 Carbon Conversion of Five Coals

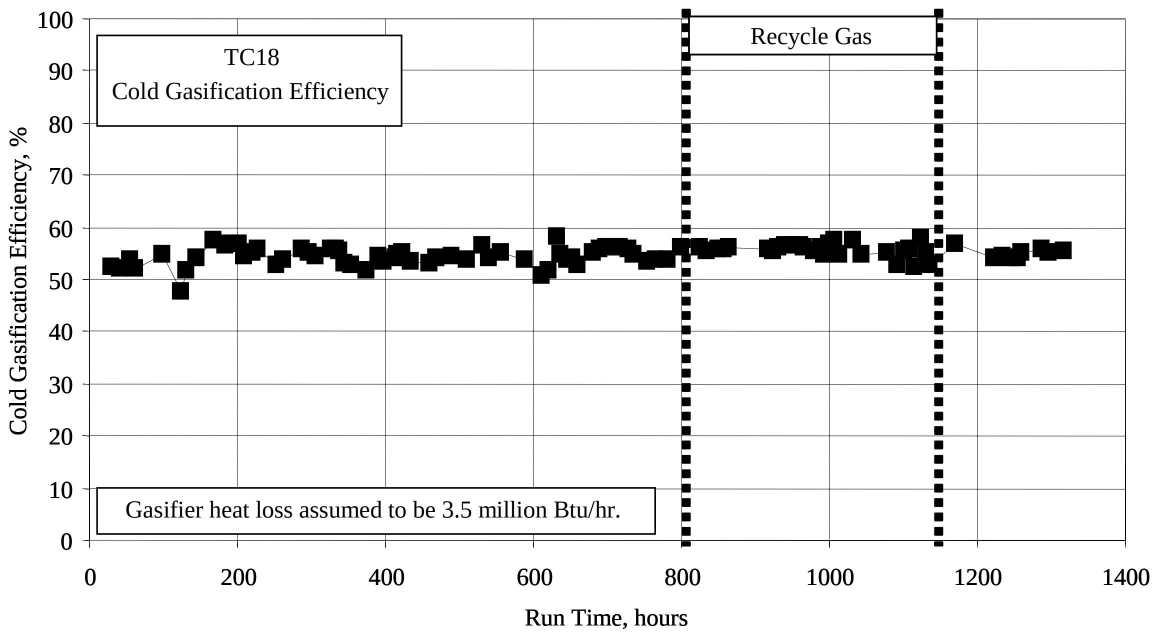

Figure 3.1-29 Cold Gasification Efficiency

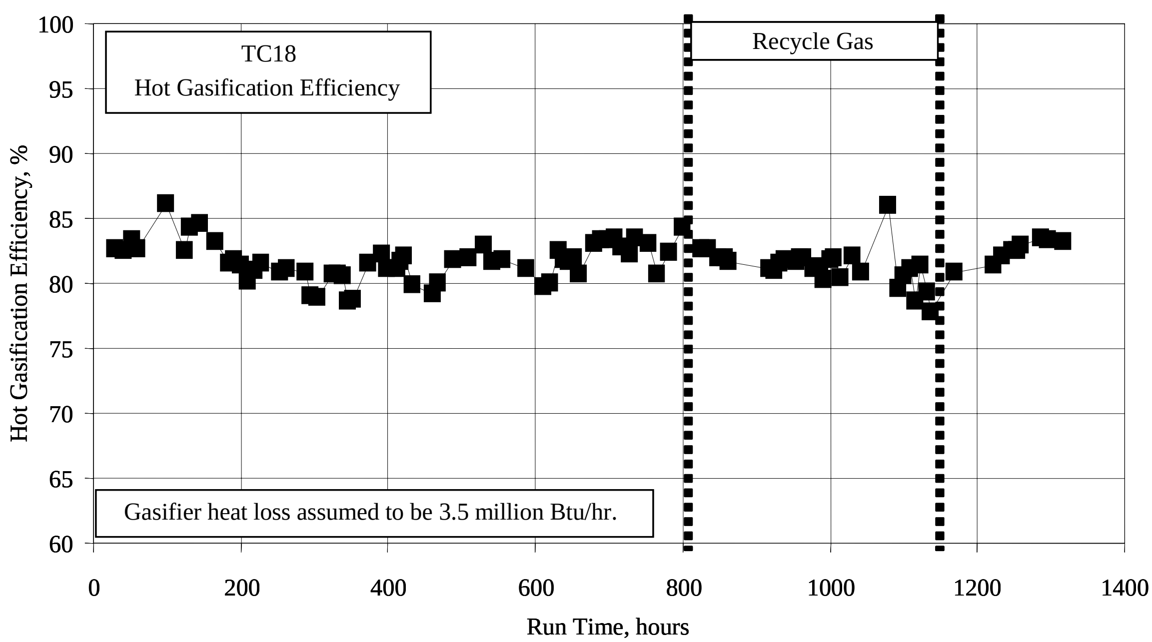

Figure 3.1-30 Hot Gasification Efficiency

3.2 PARTICULATE CONTROL DEVICE PERFORMANCE

3.2.1 Overview

The PCD operated reliably, achieving high collection efficiency while requiring no adjustments to operating parameters. However, anomalous operating conditions caused by a loss of coal feed led to the failure of titanium filter elements and an iron aluminide fuse failsafe, which caused the final system shutdown. This type of operating scenario would not be expected in a system with sufficient coal feeder redundancy.

3.2.2 Particle Mass Concentrations

In situ particulate sampling was performed at the PCD inlet and outlet using the SRI in situ batch sampling systems described in previous reports.

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 along with similar data from other gasification runs with PRB coal. The average inlet concentration was 17,800 ppmw (353 lb/hr) with a range of values from 12,300 to 23,600 ppmw. When plotted as a function of coal feed rate, the TC18 loadings fall slightly above many of the other PRB data when the gasifier recycle loop was working correctly. However, the slope of the data is more reasonable than the slope that was obtained when the recycle cyclone was working incorrectly in TC14.

The average inlet concentration without recycled syngas was 14,600 ppmw (295 lb/hr). When recycled syngas was in use, the inlet concentration averaged 17,400 ppmw (363 lb/hr). Although the recycled syngas test averaged 19 percent higher particle loading, this was not attributed to the use of recycled syngas but was due to differences in other operating factors.

PCD Outlet Mass Loadings. Particle concentrations measured at the PCD outlet are included in Table 3.2-1. Continuing the trend seen in previous test campaigns, a slightly elevated particulate loading was measured on the first few days of testing. After the PCD filter elements had been seasoned for a few days, the outlet concentration dropped below the lower limit of resolution of 0.1 ppmw and remained there for most of the test campaign.

At the end of TC18, there was a thermal event that occurred in the PCD that caused the failure of two filter elements including the Pall fuse failsafe located downstream of one of the elements. These failures occurred just after midnight on August 22 and effectively resulted in a hole through the tubesheet that was slightly larger than 1 inch. When a measurement of the PCD outlet was conducted at 8 a.m. the next morning a concentration of 1773 ppmw was measured. This is the highest leak rate ever measured at the PSDF, and even though the unit was shut down immediately after discovering the magnitude of the problem, an estimated 340 lbs of dust had leaked through the PCD during this period. Concerns over the disposition of this dust and about reentrainment during startup of TC19 resulted in a large effort to clean up the PCD outlet after TC18. As will be seen in the subsequent TC19 report, this effort was largely successful. This problem highlights the need for a reliable real-time particle monitor at the PCD outlet that can be used to make operational decisions.

Filter Element Sealing Issues. As discussed in the previous section, particle loadings measured at the PCD outlet during TC18 were initially elevated for several days before dropping below the limit of measurement resolution. This trend has been observed in the last five runs (TC14 through TC18). In all of those runs, the outlet particulate loading was initially ~ 0.2 to 0.4 ppmw and then gradually dropped to < 0.1 ppmw over the first few days of operation. As discussed in the TC17 report, this initial elevation and gradual decline in outlet particle loading is consistent with particle penetration through the fiber gaskets used to seal the filter elements and failsafes into the PCD tubesheet. Examination of the filter element gaskets used in TC16 and TC17 showed that the gasket filler material was severely degraded and that the interior of the gasket was filled with gasification ash particles. Since these observations show that gasification ash can penetrate into the fiber gaskets, it is possible that some of the particles could penetrate through the gaskets, at least until the gaskets become completely plugged with gasification ash.

As discussed in previous reports, lab measurements have shown that there is gas leakage through the fiber gaskets even when the gaskets are compressed with the maximum recommended bolt torque. These measurements have also shown that the gas leakage increases as torque is reduced. In addition to the concern about increased leakage, it is also possible that the loss of bolt torque could allow movement of the filter elements when they are backpulsed. If the ability of the gasket to cushion the element has been compromised due to the degradation of the gasket filler material, it is possible that this cyclical movement could result in long-term damage to the filter elements. Because of the potential for leakage and the concern about possible filter element damage, it is desirable to minimize any loss of torque in the bolts that hold the filter elements into the tubesheet. Torque measurements made after TC16 and TC17 showed that there was a substantial loss of torque. Torque values on the filter element bolts were typically reduced from an applied torque of 100 in-lbs before the run to an average final torque of about 30 in-lbs after the run. On the failsafe bolting, the average reduction was from 120 in-lbs to 70 in-lbs.

To address the loss of bolt torque, two techniques were tested in TC18. First, all of the filter element gaskets were pre-baked to minimize the effect of the gasket filler degradation. Secondly, Belleville conical washers were used on four sets of filter element bolts and on one set of failsafe holder bolts. The conical washers provide a spring action that was expected to minimize torque loss and to maintain more load on the gasket if some torque was lost. For the filter element installation at the PSDF, which uses four 5/16-inch bolts, the Belleville washers were specified to give a deflection of 0.008 inches at an applied load of 1478 lbs.

Torque measurements made after TC18 showed a major improvement in torque retention in the bolts equipped with the Belleville conical washers. On the filter element bolts that were originally torqued to 100 in-lbs, the average final torque was 90 in-lbs with the conical washers and only 50 to 60 in-lbs without conical washers. On the failsafe holder bolts that were originally torqued to 120 in-lbs, the average final torque was 110 in-lbs with conical washers and only 80 to 90 in-lbs without conical washers. Therefore, it is clear that the use of the conical washers resulted in a substantial improvement in torque retention. Since the average final torques measured after TC16 and TC17 were only 30 in-lbs for the filter element bolts and 70 in-lbs for the failsafe holder bolts, it appeared that the pre-baking of the gaskets also had some effect on the torque loss.

Based on the results achieved with the conical washers, it was decided that the washers would be used on all of the filter elements and failsafe holders in the next test campaign, TC19. Since the baking of the gaskets also appeared to have some beneficial effect, fiber gaskets in a rolled construction form (i.e., without any filler material) were tested prior to TC19.

3.2.3 Real-Time Particulate Monitoring

The PCME DustAlert-90 particulate monitor (referred to as the PCME) was operational throughout TC18. Although the instrument had some slightly elevated output indications during the course of TC18, these were not found to be related to actual PCD particle penetration. These elevated readings did not correlate with changes to any particular operating condition. Since this instrument uses a very sensitive electrometer to measure particle charge in the syngas, it may be affected by other types of electrical phenomena. However, at no time did its output exceed 10 percent when there was no particle leakage found.

Figure 3.2-2 shows the PCME output from midnight until 2 a.m. on August 22. A coal feeder problem caused oxygen breakthrough to the PCD that resulted in a thermal event. At 12:20 a.m., the filter elements and failsafe failed, producing an initial increase in PCME reading to about 50 percent. The coal feed was highly variable during the time from 12:20 until almost 1:00 a.m., and the inconsistency in coal feed accounts for the variations in PCME output during this period. Just before 1:00 a.m., the coal feed was restarted at a low rate, and the PCME reading went to nearly 100 percent and remained there. Because of the PCME indication, an in situ particle measurement was made. This measurement returned the value of 1,773 ppmw as discussed in the previous section, and the system was shut down.

The PCME provided a reliable indication of a catastrophic filter failure. In fact, the PCME has never failed to indicate a substantial PCD leak. However, it tends to have false positive indications. At the present time, the only sure way to segregate the true from false positives is to make manual measurements.

3.2.4 PCD Solids Analysis

PCD pressure drop, cleaning requirements, and bridging tendency can be influenced by changes in the characteristics of the solids being 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 effect of all 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 samples used for the laboratory drag measurements.

<u>In Situ Samples</u>. Figure 3.2-3 shows differential mass particle size distributions of the PCD inlet in situ samples. The average of all in situ measurements is compared to results from June 27, when very low dustcake drag was indicated. The drag differences will be investigated further in later sections of this report.

Hopper Samples. Figure 3.2-4 compares the differential mass percentage distributions for the average of all in situ samples with the hopper sample used for the TC18 lab drag measurements. Only minor differences are seen between the size distributions of all the samples. There was no indication hat the hopper samples chosen for the laboratory measurements were not representative in terms of particle size distribution.

3.2.4.2 Dustcake Observations and Thickness Measurements

The PCD dustcake was inspected, and cake thickness measurements were made during the PCD inspection following TC18. The dustcake on the iron aluminide filter elements was generally much thicker than usual and contained three distinct layers: a medium-gray outer layer, a black middle layer, and a white inner layer. The white inner layer appeared to be combusted ash from the thermal event discussed previously. It was presumably covered with gasification ash from subsequent operations after the thermal event. The individual layers were too thin to be sampled separately, so only bulk dustcake samples were collected from the top and bottom plenums.

On the titanium filter elements, the white inner layer was fused to the surface of the filter element and could not be removed. The bonded white inner layer was covered with a very thin layer of black char, which was much thinner than the dustcake on the iron aluminide elements. There was also an extremely thin black layer under the bonded white layer. The underlying black layer appeared to be some reaction product formed from the titanium. These observations suggest that some sort of bonding and/or reaction took place at the surface of the titanium elements, rendering them almost completely impermeable. This would explain why the outer char coating was very thin on the titanium elements.

Because of the surface transformation and damage of the titanium elements, dustcake thickness measurements were made only on the iron aluminide (FEAL) elements. The measurements are summarized in Table 3.2-2. After a normal shutdown of the PCD, the thickness of the residual cake is typically about 0.01 inches. The measurements in Table 3.2-2 confirm that the TC18 cake was much thicker than usual. The reason for the thicker cake is unclear, but it is possible that the thermal event caused a temporary increase in the tenacity of the gasification ash, resulting in the thicker residual cake.

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 samples used for the laboratory drag measurements, and the dustcake samples.

In Situ Samples. Tables 3.2-3 and 3.2-4 give the physical properties and chemical compositions of the in situ samples collected at the PCD inlet and the hopper samples selected for lab drag measurements. As shown in the tables, the gasification ash from the first few days of testing was characterized by lower LOI/NCC (loss on ignition/non-carbonate carbon), lower surface area, and (as discussed later) lower drag. As in previous tests with PRB coal, the TC18 data show a general increase in surface area with increasing NCC as shown in Figure 3.2-5. As noted in previous reports, this presentation of the data also shows that limestone addition produces an increase in the surface area at a given value of NCC. This effect presumably results from the extensive pore structure that is created in the limestone during calcination at high temperatures in the gasifier. The graph also shows that the ash from the Illinois Basin bituminous coal has a much lower surface area than the ash from the PRB coal, even though it has a relatively high NCC content.

Hopper Samples. For lab drag measurements, three hopper samples were selected: one from the initial test period when NCC and transient drag were low, one with steady state conditions without recycled syngas, and one from steady state conditions with recycled syngas. As shown in the tables, the first sample had much lower NCC and surface area. The samples from the two steady state tests with and without recycled syngas were essentially identical in terms of both physical properties and chemistry. This suggests that the use of recycled syngas in place of the nitrogen had very little or no effect on the particulate properties and chemical composition. All of the hopper samples appear to be very similar to the corresponding in situ samples collected under the same test conditions. This suggests that the hopper samples selected are representative samples for the laboratory drag measurements.

Dustcake Samples. Tables 3.2-5 and 3.2-6 give the physical properties and chemical composition of the dustcake samples taken following TC18. Some differences can be seen between the samples taken from the top and bottom plenums, but the differences are relatively minor except for the difference in MMD (6.3 microns on the top plenum and 11.6 microns on the bottom plenum). This difference could indicate some vertical segregation of particle sizes within the PCD, but there was no indication that this had any adverse impact on PCD operations.

As noted previously, the TC18 dustcake was unusually thick. As shown in the tables, the NCC/LOI of the TC18 dustcake was much lower than most of the TC18 in situ samples and hopper samples and much lower than the NCC/LOI of previous PRB dustcakes that were not subjected to a burnoff. Compared to the in situ/hopper samples and to the previous dustcakes, the TC18 dustcake was also higher in bulk density, lower in porosity, and lower in surface area. These differences in properties and chemistry suggest that the TC18 dustcake may have been affected by the thermal event. The thermal event could have produced a partially combusted cake that was more cohesive than normal. This would explain the observations that the TC18 cake was unusually thick and unusually low in NCC/LOI, surface area, and porosity.

3.2.4.4 Dustcake Flow Resistance

Lab Drag Measurements. Drag measurements were made on the three hopper samples described previously and on one of two drum samples that was collected for use in the PCD cold flow model. The results are illustrated in Figure 3.2-6 with drag plotted against mass median particle diameter (MMD). The non-carbonate carbon (NCC) contents of the samples are shown in the graph legend and, as expected, the dust cake drag values generally increase with increasing NCC. The results of a multiple linear regression relating drag to both MMD and NCC are shown on the graph as the two dashed lines that tend to form upper and lower bounds to the lab data. The lower dashed line is the regression calculated for 30 percent NCC while the upper line is for 60 percent NCC. The regression equation obtained was:

$$\text{Drag} = 10\text{\^{}}(2.522 - (0.725 * \text{Log(MMD)}) + 0.0076 * \text{NCC}), \text{ with an } r^2 = 0.94.$$

Although there is some scatter in the lab data, resulting in varying slopes for individual data sets, the regression appears to do a good job of averaging out the relationship. This equation will be used in a subsequent section to compare the lab measured drag to the actual PCD transient drag.

Transient PCD Drag. During each in situ sampling run at the PCD inlet, the PCD transient drag was calculated using the measured pressure drop, gas flow, and particle concentration (see Table 3.2-7). 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 Illinois Basin testing in TC17. There is considerable scatter in the data due to variations in particle size of individual tests, equipment configuration, process conditions, coal composition, and limestone addition. Nevertheless, the data show that there are definite distinctions between the data sets obtained with PRB coal, high sodium lignite, and Illinois Basin coal. These distinctions are to be expected based on the morphological differences between the ashes produced from lignite, PRB, and bituminous coal. Although the TC18 data show more scatter than in previous PRB tests, overall the data continue to show a definite trend toward increasing drag with increasing carbon content

Comparison of Lab Measurements with Transient Drag. Average lab and PCD drag values for all gasification test runs are summarized in Table 3.2-8 and plotted in Figure 3.2-8. The comparison shows excellent overall agreement (average difference of 8.4 percent), even though the difference is much higher for certain test campaigns. For TC18, the difference was 32.6 percent, which may seem quite large. Actually, a difference of 32.6 percent is reasonable considering all of the uncertainties in the various factors that go into these calculations. If, for example, 33 percent of the incoming gasification ash dropped out in the PCD before reaching the filter elements, the average calculated transient drag would be increased to a point that would give nearly perfect agreement with the average lab measured drag in TC18. Moreover, the plot of average lab versus PCD drag values (Figure 3.2-8) 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, indicating good agreement between the data sets.

3.2.5 Filter Element and Failsafe Testing

Filter element and failsafe testing continued during the test campaign. Long-term testing of iron aluminide elements continued, with several individual elements having accumulated over 7,000 hours of gasification exposure at the conclusion of TC18.

On-line failsafe testing was also conducted with the rupture disc failsafe tester used with the Pall iron aluminide fuse failsafe. It was discovered that the tester opened before it was activated (the rupture disc burst), so in situ sampling at the PCD outlet was performed, and this indicated good performance of the Pall fuse.

The failsafe tester, shown in Figure 3.5-9, consists of a pipe with a simplified Venturi device to measure flow and a double-burst disk arrangement to activate the test device. The tester is installed on the bottom plenum of the PCD. During operation, the failsafe device is exposed to particulate laden syngas by bursting the rupture discs with high pressure nitrogen. The burst discs, labeled 1 and 2 on the figure, maintain a seal, preventing the solids from penetrating the failsafe device. To initiate the test, the valve on the burst line is fully opened. The sudden increase in pressure ruptures Disc 1. This exposes Disc 2 to the tubesheet pressure drop, which is sufficient to rupture it. Once both burst disks are ruptured, the failsafe is exposed to the particulate laden syngas. This test setup allows the failsafe collection efficiency to be determined during a high solids concentration loading. With this test, the particulate loading to the failsafe should be close to the PCD inlet loading, typically 15,000 to 25,000 ppmw.

Unfortunately, the discovered that the tester opened before it was activated. Because this tester proved to be difficult to control in both TC17 and TC18, this test device design will no longer be used.

Despite the premature opening of the failsafe tester, the Pall fuse was successfully tested, and outlet loading was acceptably low (~0.3 ppmw) during the first test after the failsafe was exposed to particulate. Outlet loading samples later in the test campaign were below the lower limit of resolution, indicating that the failsafe plugged well.

Table 3.2-1 PCD In Situ Particulate Measurements
(Page 1 of 2)

	PCD Inlet					PCD Outlet				
Test Date	Run No.	Start Time	End Time	Particle Loading, ppmw	Particle Loading, lb/hr	Run No.	Start Time	End Time	H_2O Vapor, vol %	Particle Loading, ppmw
Sand Circulation with Startup Burner										
6/24/05	1	14:00	14:15	6100	109	1	13:00	15:00	4.3	0.34
PRB - Air Blown										
6/26/05	2	12:30	12:45	21400	483	2	10:45	14:45	12.6	0.39
6/27/05	3	9:30	9:45	19200	430	3	8:45	12:45	13.9	0.21
6/28/05	4	10:00	10:15	17900	394	4	8:30	11:05	13.3	0.19(1)
6/29/05	5	12:40	12:55	14600	379	5	9:15	13:15	12.6	< 0.10
6/30/05	6	9:20	9:35	15100	373	6	9:00	13:00	14.5	< 0.10
7/1/05	7	9:30	9:45	14800	347	7	8:45	12:45	13.5	< 0.10
7/2/05	8	8:30	8:45	23100	517	--	--	--	--	--
"	9	11:00	11:15	21300	471	--	--	--	--	--
7/5/05	10	9:30	9:45	19900	439	8	9:15	13:15	7.3	< 0.10
7/6/05	11	10:15	10:30	14400	282	9	10:00	14:00	8.2	< 0.10
7/7/05	12	8:15	8:30	19800	363	10	8:20	11:00	7.7	< 0.10
"	13	10:00	10:15	18000	334	--	--	--	--	--
7/8/05	14	8:30	8:45	23600	415	11	8:15	11:00	7.1	< 0.10
"	15	10:00	10:15	23400	412	--	--	--	--	--
7/12/05	16	9:15	9:30	17200	351	12	9:00	11:53	9.7	< 0.10
7/13/05	17	9:30	9:45	20700	453	13	9:00	13:00	9.3	< 0.10
7/14/05	18	9:15	9:30	17600	405	14	9:00	13:00	9.2	< 0.10
7/15/05	19	9:15	9:30	22700	481	15	9:00	13:00	7.2	< 0.10
7/18/05	20	12:45	13:00	16000	277	16(2)	8:45	13:09	8.8	< 0.10
7/19/05	--	--	--	--	--	17	9:45	13:45	8.8	< 0.10
7/20/05	--	--	--	--	--	18	9:30	13:30	8.6	< 0.10
7/21/05	21	8:45	9:00	20700	421	19	8:30	12:30	6.6	< 0.10
7/22/05	22	10:00	10:15	15400	267	20	9:00	13:00	6.6	< 0.10
7/25/05	23	8:30	8:45	(3)	(3)	21	8:15	11:45	7.0	< 0.10
"	24	10:15	10:30	12600	253	--	--	--	--	--
"	25	12:15	12:30	14200	282	--	--	--	--	--
7/26/05	26	8:15	8:30	14000	286	22	8:00	11:00	7.6	< 0.10
"	27	10:00	10:15	17500	359	--	--	--	--	--
7/27/05	28	9:01	9:16	14000	289	23	8:45	10:36	6.3(4)	< 0.10
7/28/05	29	9:00	9:15	18400	398	24	8:45	12:45	7.5	< 0.10
7/29/05	30	9:15	9:30	14900	308	25	9:00	13:00	7.8	< 0.10
7/31/05	31	8:45	9:00	17200	359	26	8:30	12:00	8.0	< 0.10
"	32	10:15	10:30	17000	354	--	--	--	--	--

Notes:
1. Mass includes both tar and ash.
2. Run paused for 24 minutes from 9:51 till 10:15 for coal mill upset.
3. Sample flow control malfunction. Result invalid.
4. Condenser stopped up with tar. All condensate may not have been recovered.

Table 3.2-1 PCD In Situ Particulate Measurements
(Page 2 of 2)

	PCD Inlet					PCD Outlet				
Test Date	Run No.	Start Time	End Time	Particle Loading, ppmw	Particle Loading, lb/hr	Run No.	Start Time	End Time	H_2O Vapor, vol %	Particle Loading, ppmw
8/1/05	33	8:30	8:45	16100	336	27	8:00	12:00	8.1	< 0.10
"	34	10:00	10:15	19400	404	--	--	--	--	--
8/2/05	35	9:30	9:45	14900	243	28	9:00	13:00	9.2	< 0.10
8/3/05	36	9:00	9:15	17300	273	29	8:45	11:58	8.9	< 0.10
8/4/05	37	9:00	9:15	16900	284	30	8:45	12:45	8.8	< 0.10
8/6/05	38	8:45	9:00	18400	310	31	8:30	12:30	8.5	< 0.10
8/8/05	39	9:15	9:30	19100	326	32	9:00	13:00	(5)	< 0.10
8/9/05	40	9:15	9:30	17200	273	33	9:00	13:00	8.4	< 0.10
8/10/05	41	9:00	9:15	12300	239	34	8:45	12:45	18.3	< 0.10
8/11/05	42	9:15	9:30	16100	249	35	9:00	13:00	7.3	< 0.10
8/12/05	43	9:15	9:30	20100	351	36	9:00	13:00	8.0	< 0.10
8/15/05	44	9:00	9:15	23100	392	37	8:30	12:30	10.2	< 0.10
8/18/05	45	9:15	9:30	18200	369	38	8:45	12:35	7.3	< 0.10
8/19/05	46	9:00	9:15	17500	350	39	8:30	12:30	7.9	< 0.10
8/22/05	47	8:15	8:30	16300	310	40	7:45	8:45	6.3	1773(6)

Notes: 5. Condensate spilled - no value.
6. PCD leak following thermal event.

Table 3.2-2 Thickness Measurements of Residual Dustcake

Element No.	Element Type	Failsafe Type	Thickness, in.		
			Top	Middle	Bottom
T-5	FEAL	Ext fuse up	0.0693	0.0895	0.0901
T-8	FEAL	Ext fuse SW down	0.0489	0.0850	0.0806
T-15	FEAL	Internal fuse	0.0610	0.0800	0.0906
B-5	FEAL	Ext fuse up	0.0492	0.0463	0.0213
B-15	FEAL	Internal fuse	0.0544	0.0654	0.0576
B-32	FEAL	PSDF	0.0696	0.0540	0.0513

Table 3.2-3 Physical Properties of In Situ and Hopper Samples
(Page 1 of 2)

Sample ID	Run No.	Sample Date	Bulk Density, g/cc	True Density, g/cc	Bulk Porosity, %	Surface Area, m^2/g	Mass Median Particle Size, µm
In-Situ Samples, Air-Blown							
AB19410	2	06/26/05	0.33	2.46	86.6	143.0	21.9
AB19411	3	06/27/05	0.30	2.44	87.7	147.0	18.2
AB19412	4	06/28/05	0.35	2.55	86.3	120.0	18.5
AB19413	5	06/29/05	0.31	2.37	86.9	N.M.	17.8
AB19414	6	06/30/05	0.28	2.60	89.2	N.M.	14.7
AB19415	7	07/01/05	0.31	2.48	87.5	N.M.	17.9
AB19416	8	07/02/05	0.31	2.18	85.8	N.M.	20.8
AB19417	9	07/02/05	0.28	2.24	87.5	N.M.	20.4
AB19418	10	07/05/05	0.28	2.33	88.0	N.M.	18.1
AB19419	11	07/06/05	0.25	2.61	90.4	N.M.	19.3
AB19420	12	07/07/05	0.26	2.33	88.8	N.M.	21.2
AB19421	13	07/07/05	0.25	2.41	89.6	N.M.	19.6
AB19422	14	07/08/05	0.29	2.19	86.8	177.0	19.4
AB19423	15	07/12/05	0.31	2.25	86.2	173.0	19.8
AB19424	16	07/13/05	0.26	2.37	89.0	N.M.	20.5
AB19425	17	07/14/05	0.26	2.34	88.9	201.0	19.2
AB19426	18	07/14/05	0.28	2.22	87.4	N.M.	18.2
AB19427	19	07/15/05	0.27	2.17	87.6	N.M.	15.0
AB19428	20	07/18/05	0.25	2.39	89.5	N.M.	20.9
AB19429	21	07/21/05	0.25	2.27	89.0	N.M.	21.2
AB19430	22	07/22/05	0.24	2.37	89.9	N.M.	19.8
AB19431	23	07/25/05	0.23	2.34	90.2	N.M.	19.7
AB19432	24	07/25/05	0.22	2.34	90.6	N.M.	18.9
AB19433	25	07/25/05	0.23	2.43	90.5	N.M.	18.8
AB19434	26	07/26/05	0.23	2.42	90.5	236.0	18.2

Table 3.2-3 Physical Properties of In Situ and Hopper Samples
(Page 2 of 2)

Sample ID	Run No.	Sample Date	Bulk Density, g/cc	True Density, g/cc	Bulk Porosity, %	Surface Area, m²/g	Mass Median Particle Size, µm
AB19435	27	07/26/05	0.23	2.35	90.2	267.0	18.0
AB19436	28	07/27/05	0.24	2.46	90.2	N.M.	19.6
AB19437	29	07/28/05	0.24	2.30	89.6	N.M.	19.8
AB19438	30	07/29/05	0.22	2.36	90.7	N.M.	19.4
AB19439	31	07/31/05	0.27	2.41	88.8	227.0	19.1
AB19440	32	07/31/05	0.25	2.50	90.0	223.0	16.9
AB19441	33	08/01/05	0.24	2.31	89.6	N.M.	18.9
AB19442	34	08/01/05	0.25	2.40	89.6	N.M.	20.3
AB19443	35	08/02/05	0.19	2.41	92.1	N.M.	17.5
AB19444	36	08/03/05	0.18	2.50	92.8	N.M.	11.5
AB19445	37	08/04/05	0.27	2.47	89.1	N.M.	22.5
AB19446	38	08/05/05	0.28	2.52	88.9	N.M.	21.0
AB19447	39	08/08/05	0.40	2.46	83.7	N.M.	22.1
AB19448	40	08/09/05	0.26	2.51	89.6	N.M.	20.7
AB19449	41	08/10/05	0.31	2.79	88.9	N.M.	12.2
AB19450	42	08/11/05	0.20	2.37	91.6	N.M.	17.1
AB19451	43	08/12/05	0.23	2.15	89.3	N.M.	19.9
AB19452	44	08/15/05	0.26	2.10	87.6	N.M.	17.6
AB19453	45	08/18/05	0.27	2.33	88.4	N.M.	20.4
AB19454	46	08/19/05	0.27	2.29	88.2	N.M.	21.1
AB19455	47	08/22/05	0.22	2.35	90.6	N.M.	19.9
Drum Samples							
AB19457	Drum 1	07/16/05	0.29	2.15	86.5	202.0	16.0
AB19458	Drum 2	07/16/05	0.29	2.15	86.5	200.0	14.8
Composite Hopper Samples Used for Lab Drag Measurements							
AB20774	Low-Drag	06/28/05	0.32	2.30	86.1	143.0	15.5
AB20775	No Recycle Gas	07/26/05	0.23	2.17	89.4	235.0	20.7
AB20776	Recycle Gas	07/31/05	0.23	2.15	89.3	236.0	23.3

Table 3.2-4 Chemical Composition of In Situ and Hopper Samples
(Page 1 of 2)

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 %
In-Situ Samples, Air-Blown								
AB19410	2	06/26/05	4.18	0.20	5.37	28.93	61.32	31.06
AB19411	3	06/27/05	3.84	0.20	8.03	31.26	56.67	33.32
AB19412	4	06/28/05	5.02	0.16	7.70	24.51	62.61	28.76
AB19413	5	06/29/05	4.16	0.20	8.45	33.03	54.16	37.07
AB19414	6	06/30/05	3.89	0.07	11.61	21.62	62.82	23.41
AB19415	7	07/01/05	6.39	0.22	8.55	32.20	52.63	35.96
AB19416	8	07/02/05	4.89	0.95	4.93	54.04	35.19	57.07
AB19417	9	07/02/05	4.39	1.06	5.72	49.50	39.33	54.34
AB19418	10	07/05/05	6.11	0.95	6.55	46.18	40.21	48.42
AB19419	11	07/06/05	6.07	0.71	11.37	29.70	52.15	31.32
AB19420	12	07/07/05	6.64	0.73	7.47	41.50	43.66	45.16
AB19421	13	07/07/05	6.25	0.97	7.54	42.46	42.77	45.40
AB19422	14	07/08/05	6.93	0.82	5.03	50.82	36.40	54.27
AB19423	15	07/12/05	6.66	0.82	5.26	49.85	37.41	54.09
AB19424	16	07/13/05	7.80	0.42	6.73	43.46	41.59	48.29
AB19425	17	07/14/05	7.80	0.53	6.59	47.96	37.12	51.08
AB19426	18	07/14/05	7.89	1.13	5.43	49.80	35.75	53.58
AB19427	19	07/15/05	8.14	1.63	3.28	52.09	34.86	56.92
AB19428	20	07/18/05	7.00	1.25	9.46	37.52	44.78	39.99
AB19429	21	07/21/05	7.59	2.04	6.07	46.96	37.34	49.69
AB19430	22	07/22/05	7.68	1.88	8.71	38.17	43.56	40.65
AB19431	23	07/25/05	7.55	0.80	6.90	42.43	42.32	45.89
AB19432	24	07/25/05	8.18	0.98	6.88	41.99	41.97	47.14
AB19433	25	07/25/05	7.48	0.84	8.34	39.17	44.17	43.21
AB19434	26	07/26/05	6.98	0.95	8.47	41.97	41.63	44.22

Table 3.2-4 Chemical Composition of In Situ and Hopper Samples
(Page 2 of 2)

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 %
AB19435	27	07/26/05	7.59	0.91	7.53	44.33	39.64	46.87
AB19436	28	07/27/05	7.64	0.64	9.95	35.19	46.58	38.09
AB19437	29	07/28/05	7.75	0.84	6.51	48.84	36.06	51.30
AB19438	30	07/29/05	8.14	0.79	7.47	44.71	38.89	48.24
AB19439	31	07/31/05	7.43	1.00	8.10	40.90	42.57	42.84
AB19440	32	07/31/05	6.70	1.00	8.26	40.67	43.38	43.84
AB19441	33	08/01/05	7.73	1.02	6.19	47.26	37.80	49.81
AB19442	34	08/01/05	7.52	1.02	6.39	46.84	38.23	49.49
AB19443	35	08/02/05	7.64	0.75	5.60	47.70	38.31	52.15
AB19444	36	08/03/05	7.34	0.69	7.08	36.96	47.93	40.22
AB19445	37	08/04/05	7.45	1.38	8.95	38.46	43.76	40.67
AB19446	38	08/05/05	7.48	1.38	8.68	35.11	47.35	37.51
AB19447	39	08/08/05	7.77	1.60	7.28	38.92	44.43	41.10
AB19448	40	08/09/05	7.02	1.27	9.18	32.96	49.57	34.96
AB19449	41	08/10/05	6.93	0.02	14.55	14.41	64.09	16.95
AB19450	42	08/11/05	6.64	0.93	8.85	41.47	42.11	44.81
AB19451	43	08/12/05	7.09	1.17	6.86	51.69	33.19	53.58
AB19452	44	08/15/05	7.36	0.42	5.74	53.17	33.31	56.62
AB19453	45	08/18/05	5.80	1.02	8.42	37.36	47.40	39.55
AB19454	46	08/19/05	5.34	0.91	7.85	40.13	45.77	42.70
AB19455	47	08/22/05	5.18	0.73	8.15	40.70	45.24	44.27
Average			*6.72*	*0.87*	*7.57*	*40.76*	*44.09*	*43.82*
Drum Samples								
AB19457	Drum 1	07/16/05	5.48	1.57	5.70	53.13	34.11	56.37
AB19458	Drum 2	07/16/05	5.41	1.60	5.67	54.26	33.06	56.39
Composite Hopper Samples Used for Lab Drag Measurements								
AB20774	Low-Drag	06/28/05	7.84	0.16	5.82	29.65	56.54	32.98
AB20775	No Recycle Gas	07/26/05	7.48	1.02	8.26	43.05	40.19	45.88
AB20776	Recycle Gas	07/31/05	8.48	0.90	7.24	42.98	40.40	45.63

Table 3.2-5 Physical Properties of Residual Dustcake

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 %
TC18 Bulk Dustcake from Top Plenum							
AB19464	08/26/05	0.34	2.53	86.6	145	6.3	29.6
TC18 Bulk Dustcake from Bottom Plenum							
AB19465	08/26/05	0.42	2.60	83.9	104	11.6	21.9

Table 3.2-6 Chemical Composition of Residual Dustcake

Sample ID	Sample Date	$CaCO_3$ Wt %	CaS Wt %	CaO Wt %	Non-Carbonate Carbon Wt %	Inerts (Ash/Sand) Wt %	Loss on Ignition Wt %
TC18 Bulk Dustcake from Top Plenum							
AB19464	8/25/05	5.25	0.74	13.96	25.32	54.73	29.61
TC18 Bulk Dustcake from Bottom Plenum							
AB19465	8/25/05	5.89	0.13	12.40	18.79	62.79	21.87

Table 3.2-7 Transient Drag Determined from PCD Performance and from Lab Measurements
(Page 1 of 2)

Run No.	ΔP/Δt, inwc/min	Δ(AL)/Δt, lb/ft²/min	FV, ft/min	MMD, µm	NCC, %	Drag, inwc/(lb/ft²)/(ft/min)		
						PCD	PCD@RT	Lab
Sand Circulation								
1	0.13	0.008	4.41	—	—	15	10	—
PRB Coal								
2	2.39	0.037	3.55	21.9	28.9	64	39	59
3	2.13	0.033	3.53	18.2	31.3	64	39	70
4	1.91	0.030	3.42	18.5	24.5	63	38	62
5	3.00	0.029	4.04	17.8	33.0	102	61	74
6	2.33	0.029	3.83	14.7	21.6	81	49	69
7	2.74	0.027	3.72	17.9	32.2	102	62	72
8	3.58	0.040	3.43	20.8	54.0	90	55	95
9	2.97	0.036	3.38	20.4	49.5	82	50	89
10	3.17	0.034	3.50	18.1	46.2	94	57	92
11	1.41	0.022	3.06	19.3	29.7	65	40	65
12	2.18	0.028	2.85	21.2	41.5	78	48	75
13	2.18	0.026	2.88	19.6	42.5	84	52	81
14	2.25	0.032	2.69	19.4	50.8	70	44	94
15	2.25	0.032	2.69	19.8	49.9	71	44	92
16	2.37	0.027	3.28	20.5	43.5	87	53	80
17	7.40	0.035	3.54	19.2	48.0	212	128	91
18	4.60	0.031	3.80	18.2	49.8	148	89	97
19	4.88	0.037	3.41	15.0	52.1	132	80	116
20	2.20	0.021	2.69	20.9	37.5	103	63	71
21	3.09	0.032	3.16	21.2	47.0	95	58	83
22	1.53	0.021	2.64	19.8	38.2	74	46	74
23	—	0.034	3.19	19.7	42.4	—	—	80
24	2.39	0.019	3.19	18.9	42.0	123	75	82
25	2.19	0.022	3.16	18.8	39.2	101	61	79
26	2.45	0.022	3.26	18.2	42.0	111	68	85
27	3.01	0.028	3.27	18.0	44.3	109	66	89

Note: Lab drag data calculated from linear regression to MMD and NCC

Table 3.2-7 Transient Drag Determined from PCD Performance and from Lab Measurements
(Page 2 of 2)

Run No.	ΔP/Δt, inwc/min	Δ(AL)/Δt, lb/ft²/min	FV, ft/min	MMD, µm	NCC, %	Drag, inwc/(lb/ft²)/(ft/min)		
						PCD	PCD@RT	Lab
28	2.89	0.022	3.56	19.6	35.2	130	79	71
29	3.00	0.031	3.58	19.8	48.8	98	59	90
30	2.83	0.024	3.33	19.4	44.7	119	72	85
31	2.87	0.028	3.47	19.1	40.9	104	62	80
32	2.85	0.027	3.46	16.9	40.7	104	63	87
33	3.05	0.026	3.40	18.9	47.3	118	71	90
34	2.50	0.031	3.39	20.3	46.8	80	48	85
35	2.21	0.019	2.59	17.5	47.7	118	73	96
36	2.18	0.021	2.50	11.5	37.0	103	64	108
37	2.13	0.022	2.89	22.5	38.5	97	60	68
38	1.94	0.024	2.88	21.0	35.1	81	50	68
39	1.92	0.025	2.91	22.1	38.9	76	47	70
40	1.63	0.021	2.68	20.7	33.0	78	48	66
41	1.75	0.018	3.45	12.2	14.4	95	58	70
42	1.53	0.019	2.59	17.1	41.5	79	49	88
43	2.07	0.027	2.98	19.9	51.7	76	47	94
44	2.69	0.030	2.75	17.6	53.2	89	54	105
45	2.85	0.028	3.36	20.4	37.4	100	60	72
46	2.89	0.027	3.30	21.1	40.1	107	64	74
47	2.26	0.024	3.19	19.9	40.7	94	57	77
AVG	2.64	0.027	3.20	19.0	40.8	97	59	82
Note: Lab drag data calculated from linear regression to MMD and NCC								

Nomenclature:

ΔP/Δt = rate of pressure drop rise during particulate sampling run, inwc/min.
Δ(AL)/Δt = rate of increase in areal loading during sampling run, lb/min/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 3.2-8 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	24.8	18.7	-27.8
TC18	PRB	59.0	82.0	32.6
Average		47.2	51.8	8.4

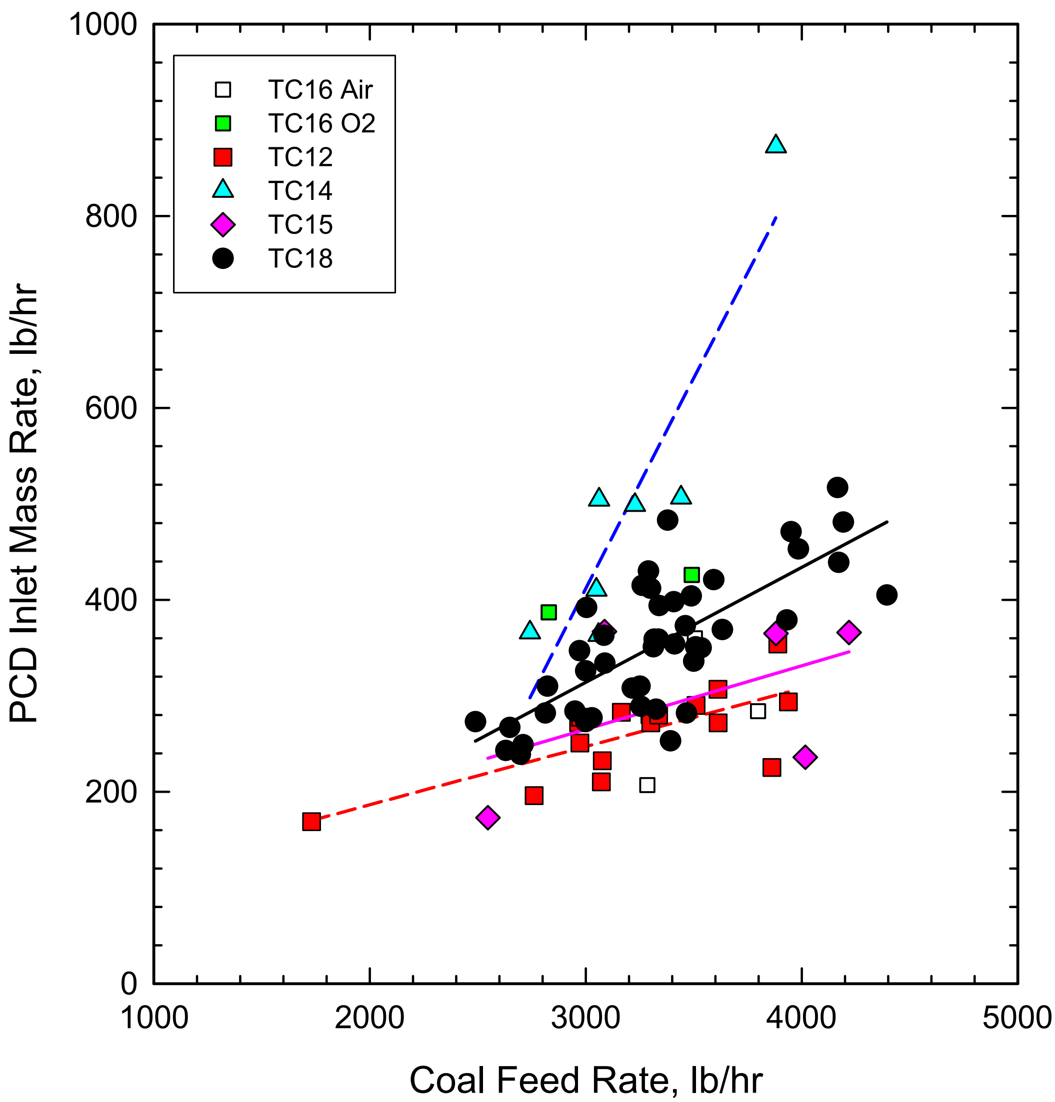

Figure 3.2-1 PCD Inlet Particle Concentration as a Function of Coal Feed Rate.

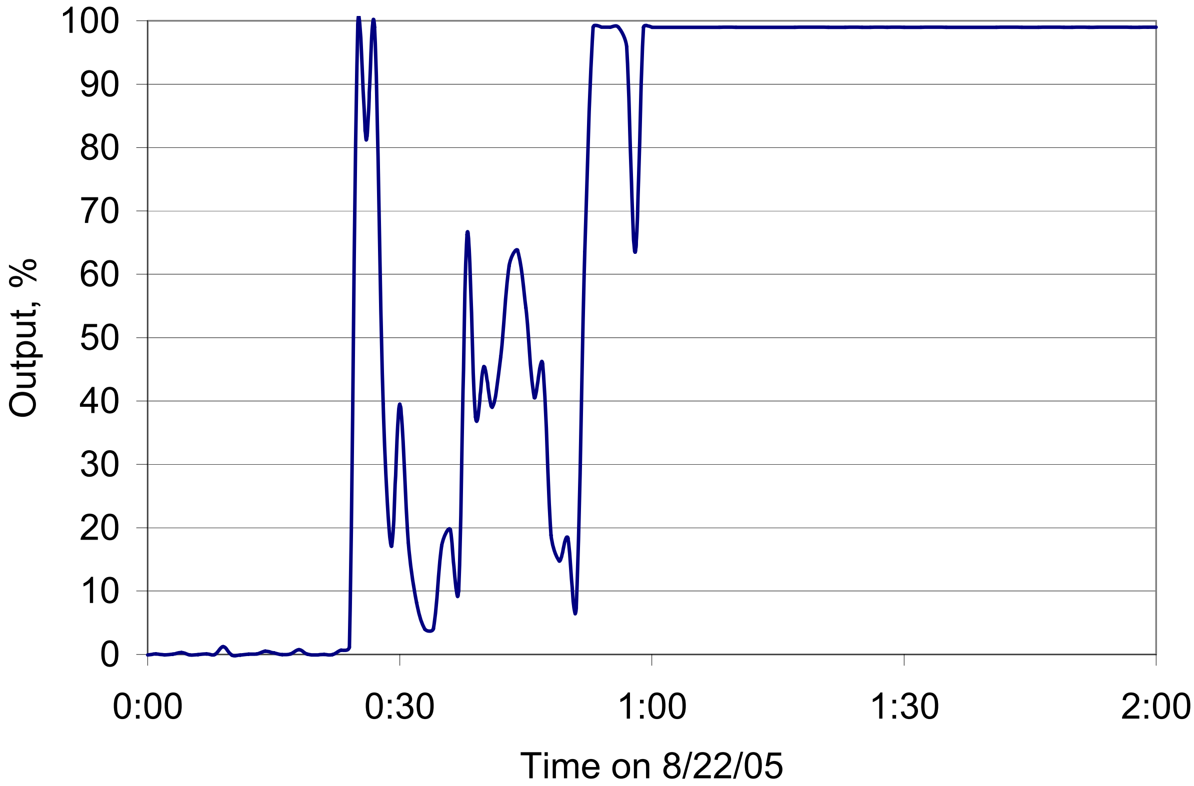

Figure 3.2-2 PCME Particulate Monitor Response to a Major Leak on August 22, 2005

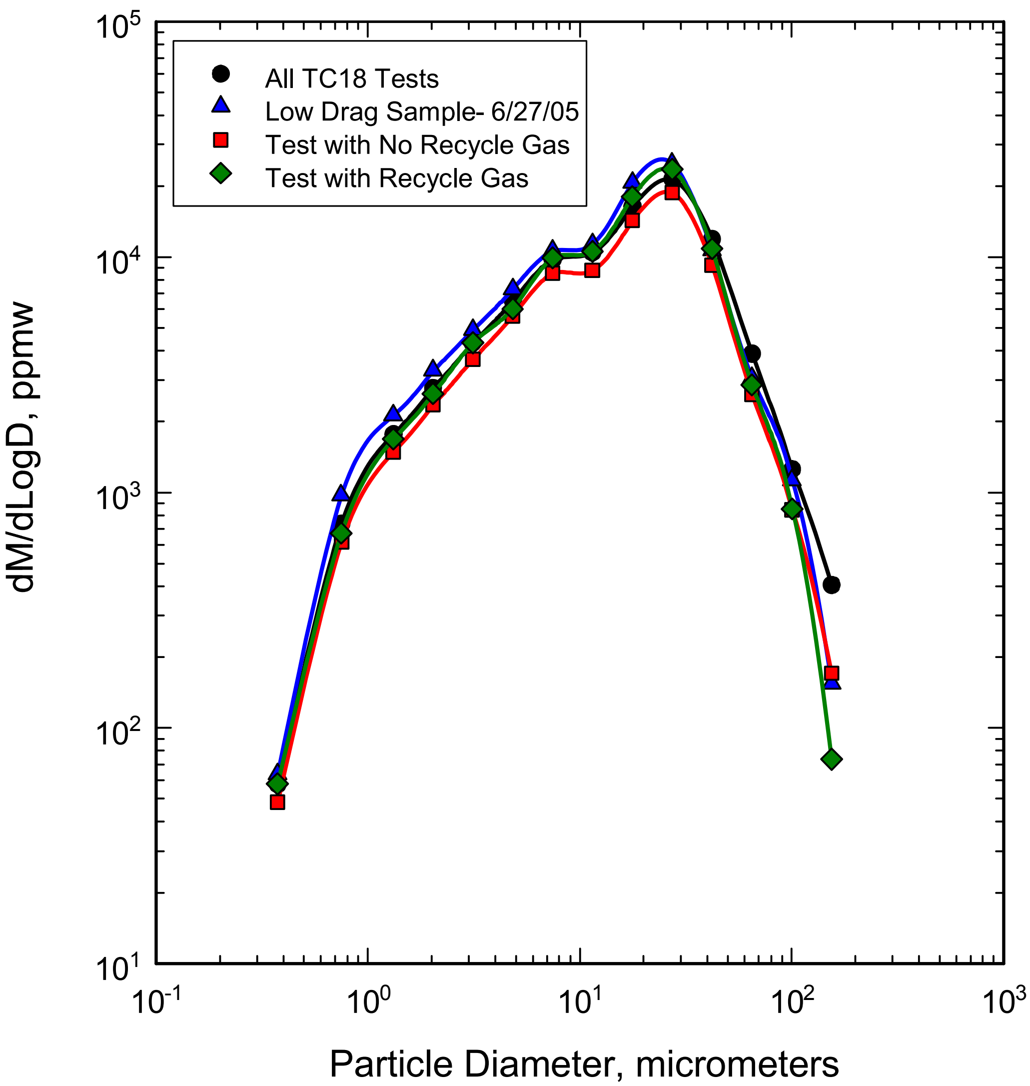

Figure 3.2-3 Comparison of Particle Size Distributions of In Situ Samples

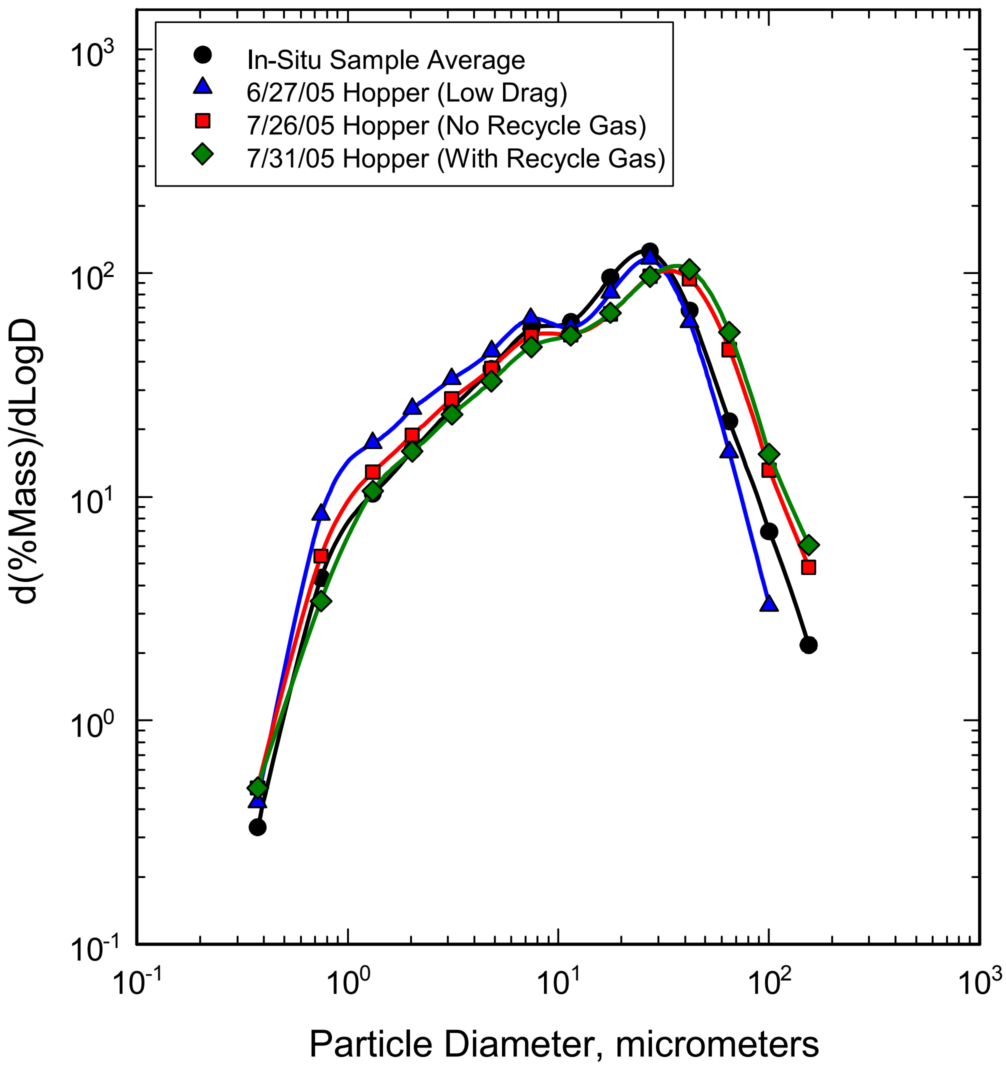

Figure 3.2-4 Comparison of Hopper and In Situ Particle Size Distributions

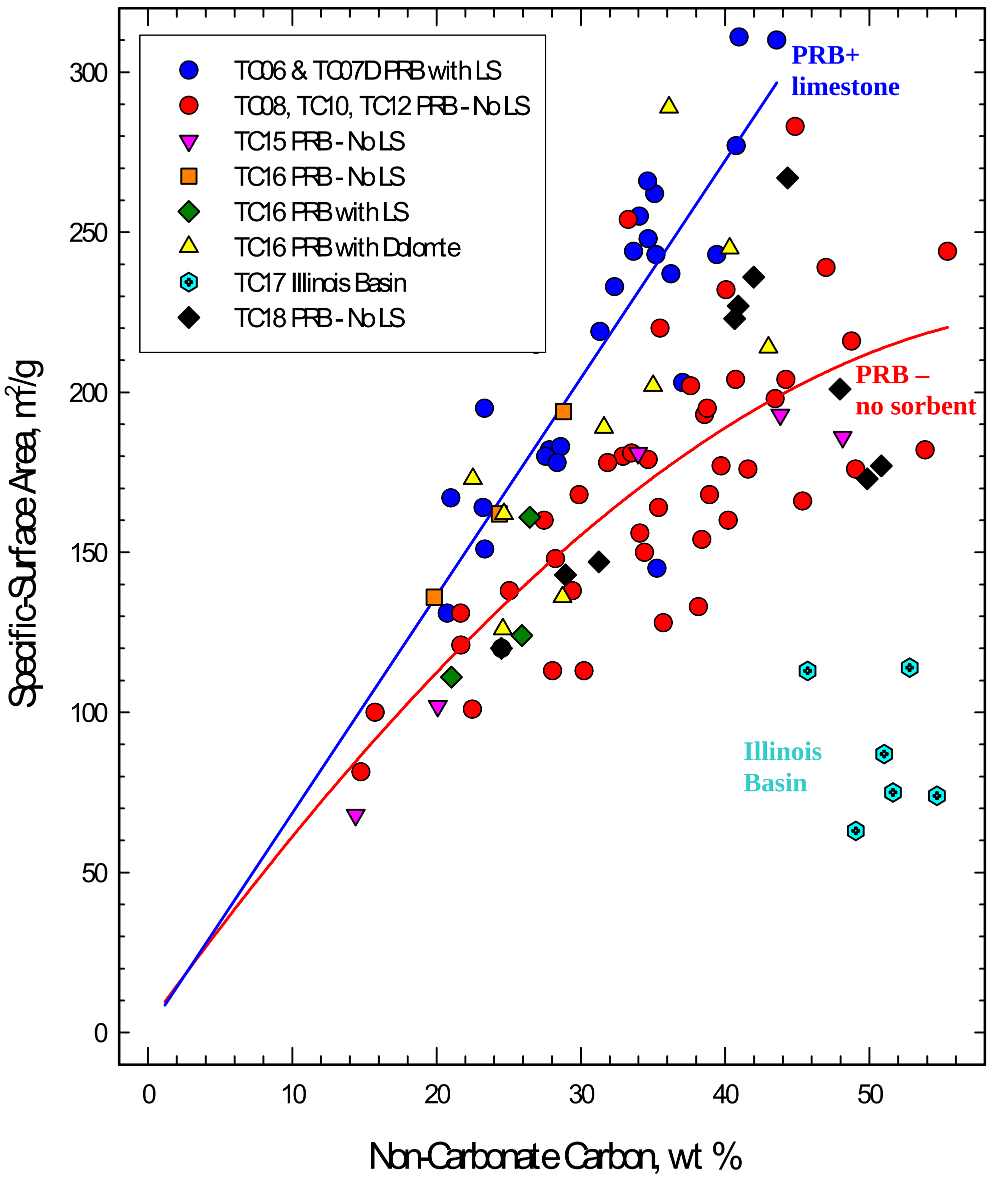

Figure 3.2-5 Effect of Non-Carbonate Carbon Content on Specific Surface Area

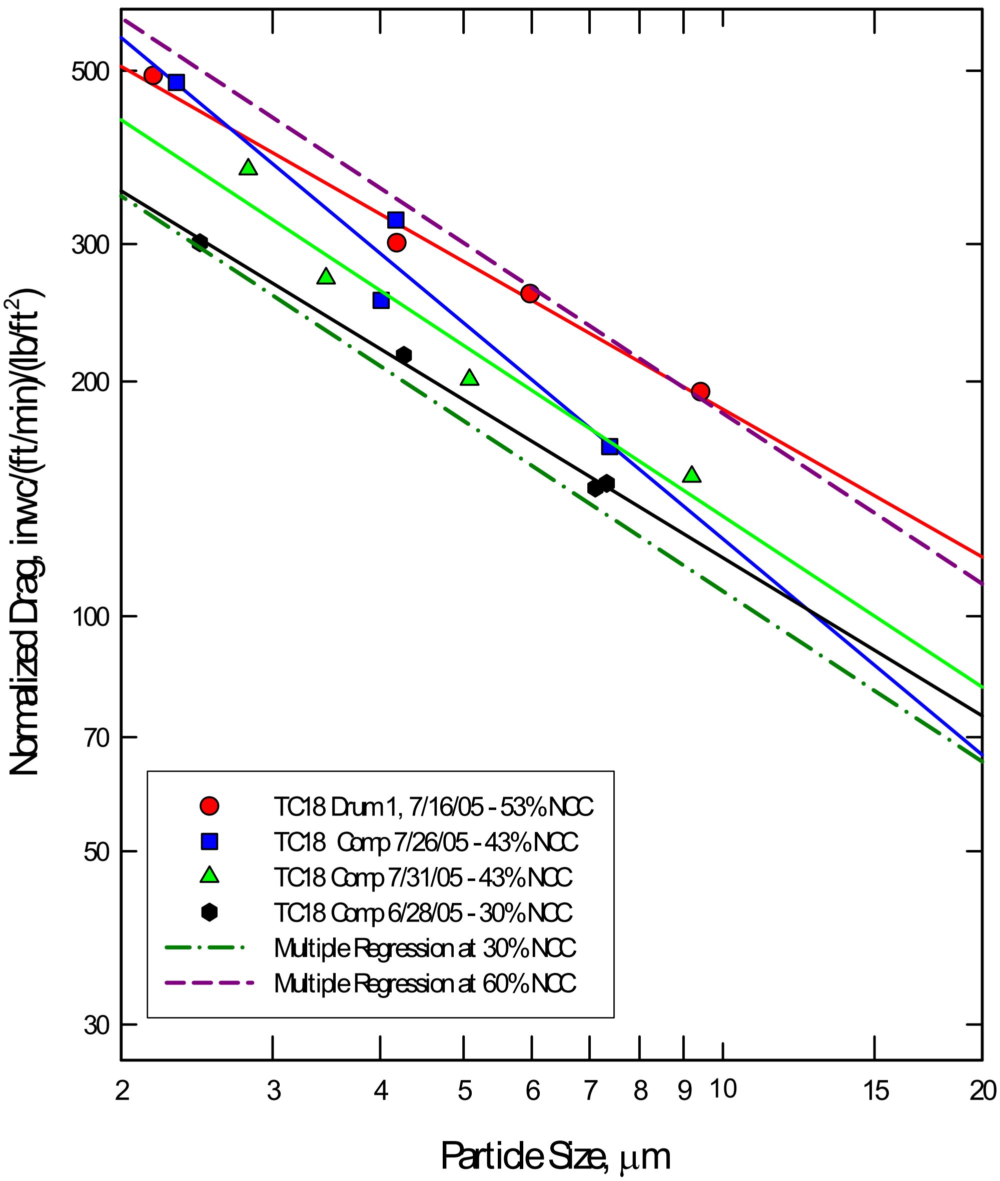

Figure 3.2-6 Lab-Measured Drag as a Function of Particle Size

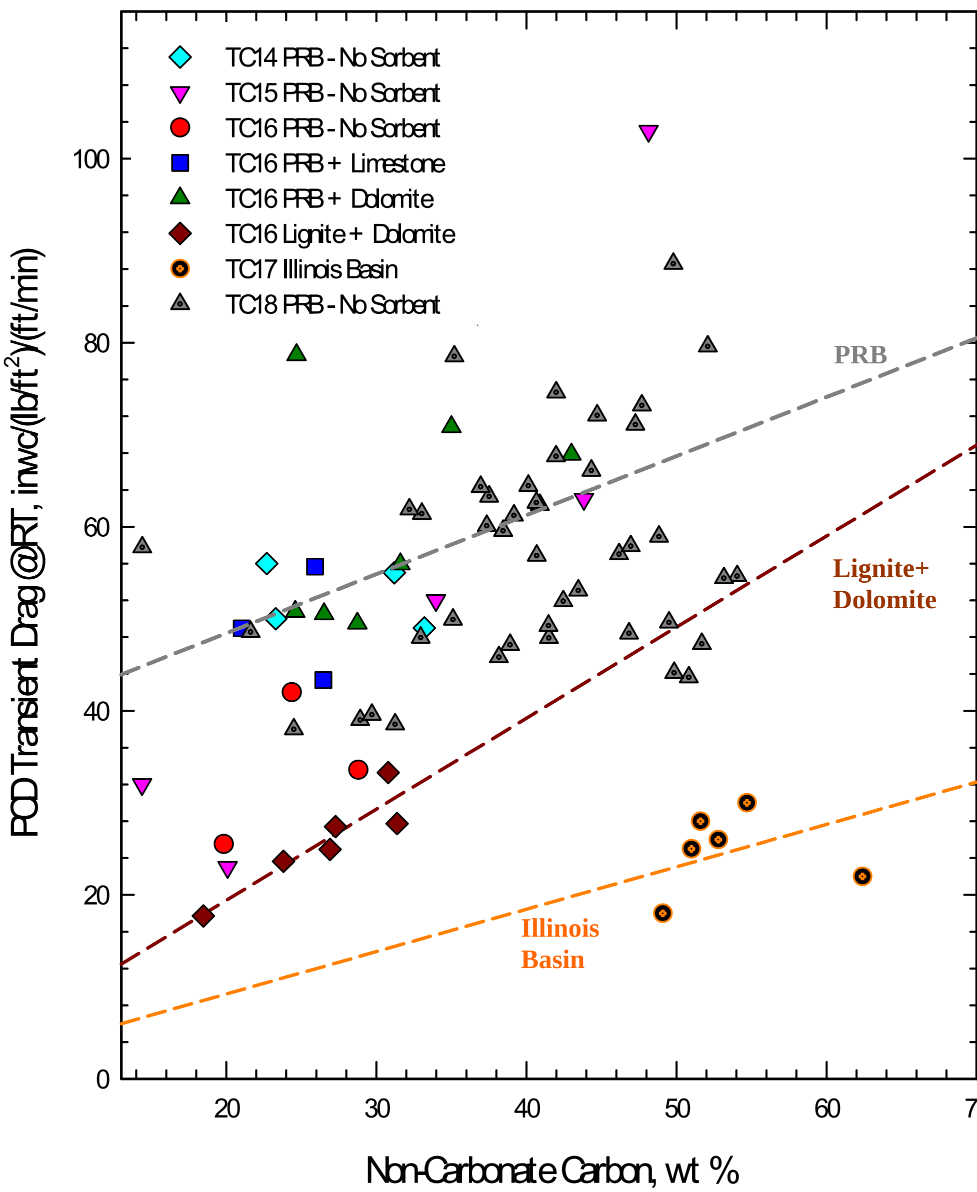

Figure 3.2-7 PCD Transient Drag versus Carbon Content of In Situ Samples

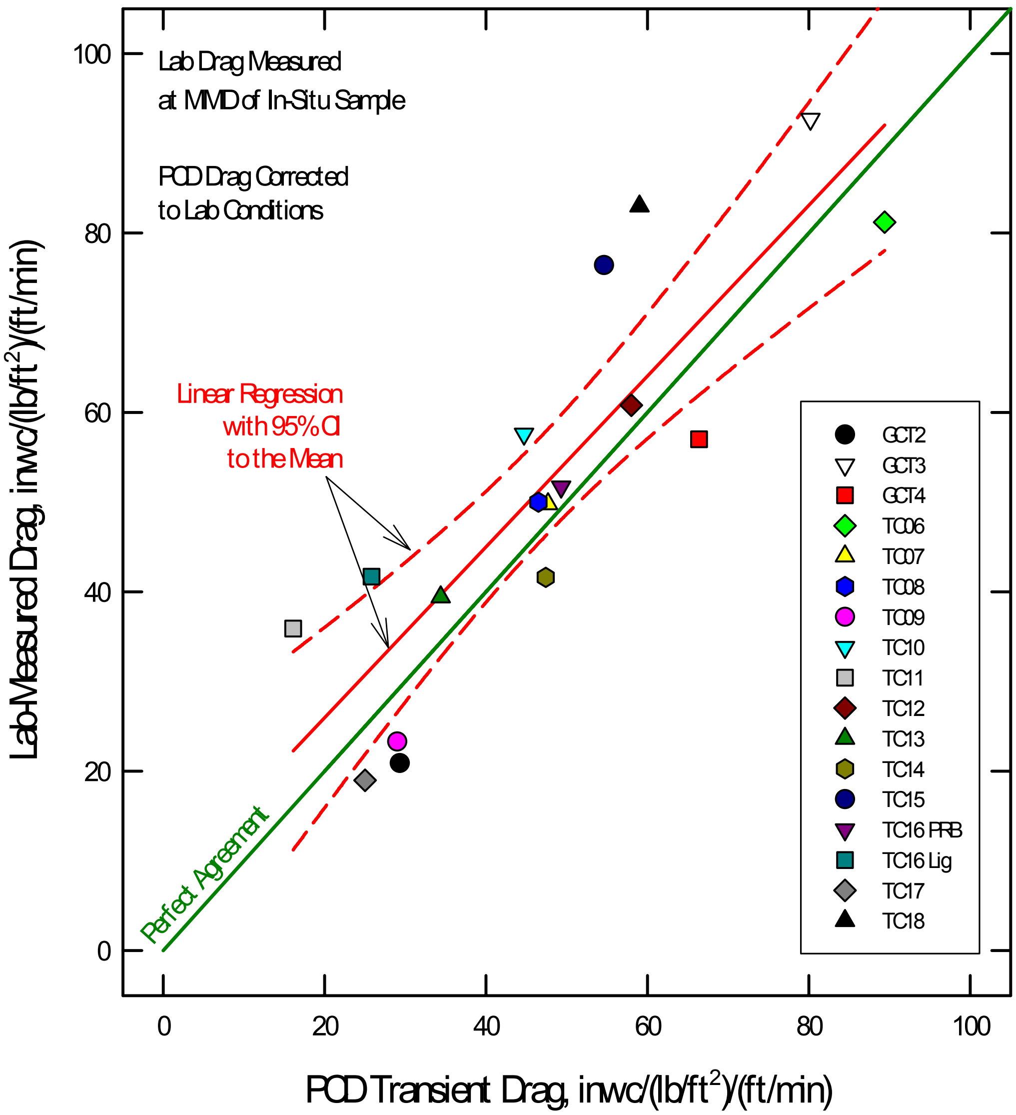

Figure 3.2-8 Comparison of PCD Transient Drag with Laboratory Measurements

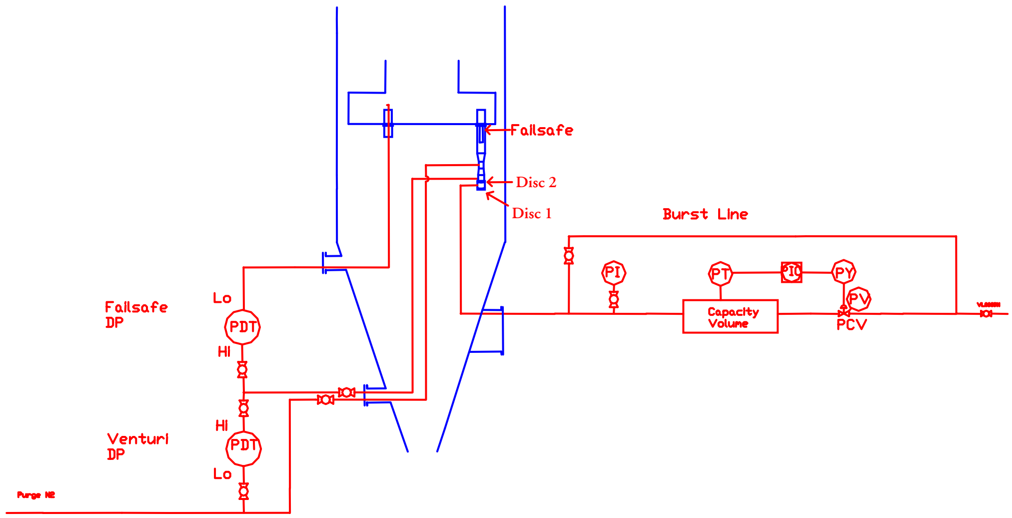

Figure 3.2-9   On Line Rupture Disc Failsafe Test Setup

3.3 PERFORMANCE OF OTHER SYSTEMS

3.3.1 Automation

Temperature control enhancements to address erroneous temperature measurements were successfully tested. A low temperature was manually forced in the control measurement block, and it was verified that the temperature controls rejected to manual. The scheme protected against erroneous readings well, until several thermocouples were reading incorrectly. The median select feature for controlling thermocouple selection caused some difficulties for the controller due to different tuning parameters set for each thermocouple. Fluctuations in the main air compressor discharge pressure caused some issues with the automatic control scheme because the air flow rates could not be controlled. The gasifier temperature controls performed well at typical gasifier operating conditions, maintaining both mixing zone and gasifier exit temperatures within 10°F of their respective set points. The load simulation controller was tested and some controller issues were identified. The PCD backpulse caused swings in syngas flow that proved to be problematic for the controller. Parametric testing in support of automation development was completed and will be used in future enhancements to the gasifier control schemes.

3.3.2 Coal Feed Systems

The original coal feeder ran for over 1,300 hours. The particle size distribution and moisture content were varied to continue evaluating the system operating envelope. The system fed coal particles ranging from 123 microns to 353 microns in mass median diameter (MMD) with an average particle size of 236 microns MMD. The average moisture content was 18 weight percent. Although the system ran well for most of the run, there were some problems, such as the lock vessel not venting due to overfilling. Since the vent valve is near the lock vessel, material may have backed up and wedged between the vent valve and a downstream orifice plate. The vent valve will be relocated to prevent this problem in the future.

The developmental coal feeder ran for over 600 hours (the most to date) until the system was shutdown to conserve nitrogen due to an external supply shortage. The system was automatically controlled by mass flow, and exhibited acceptable responses to set point changes, but had difficulty when a high percentage of fines or moisture was present in the coal. The feeder operated with different particle size distribution ranges to further evaluate the operating envelope. The system fed particles ranging from 133 to 353 microns MMD with an average particle size of 240 microns. The average moisture content was 19.6 weight percent. The minimum required coal conveying velocity was identified for each particle size distribution range tested in the developmental coal feeder.

3.3.3 Recycle Gas Compressor

The recycle gas compressor is a vertically mounted centrifugal compressor which operates at high temperature (nominally 500 to 600°F). The compressor was manufactured by Sundyne Corporation, and was designed to have a throughput of nominally 2,000 to 3,000 lb/hr. A schematic of the compressor is shown in Figure 3.3.3-1.

Overall, the recycle gas compressor system performed well throughout the test campaign. Initial commissioning, which was performed using nitrogen gas, was delayed because of excessive motor vibration. The motor was replaced, and the vibration issues were resolved. The compressor operated on recycled syngas for a total of 376 hours, with a syngas throughput of approximately 1,350 to 2,000 lb/hr. While the gasifier was operating with syngas for aeration, the raw heating value was about 8 to 10 percent higher than while operating with nitrogen. During an external nitrogen supply shortage, the Transport Gasifier continued to operate on recycled syngas. Without the recycle gas compressor, the test run would have been terminated as a consequence of the nitrogen shortage.

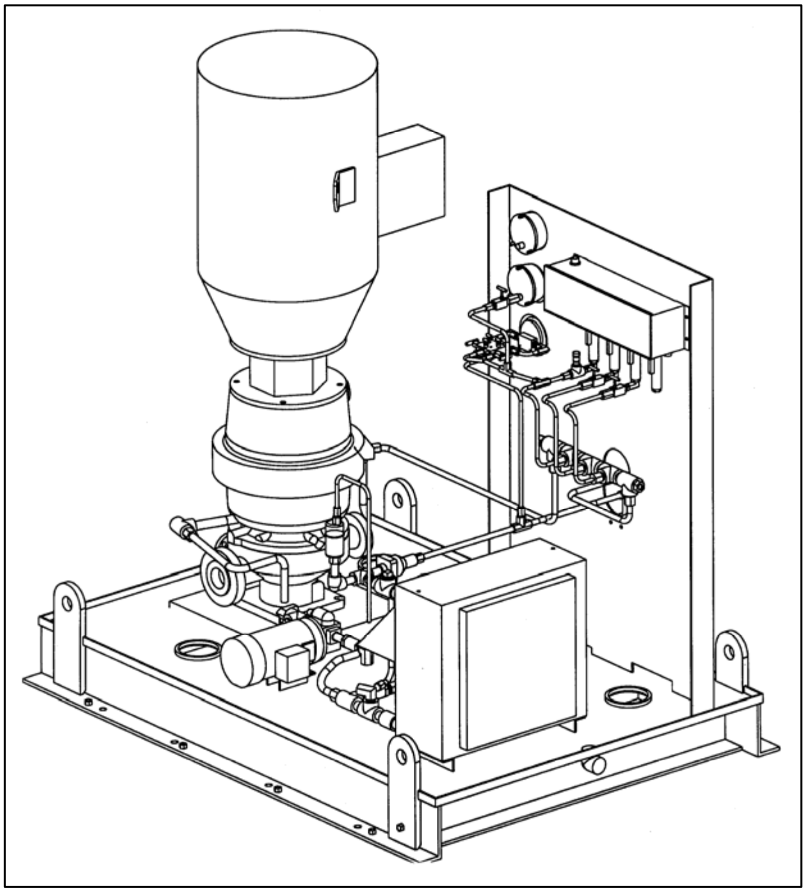

Figure 3.3.3-1 Recycle Gas Compressor

3.3.4 Continuous Fine Ash Depressurization (CFAD) System

Before the test run, the CFAD fines removal system was relocated and connected directly to the bottom of the PCD. CFAD operated well for over 1,300 hours and removed over 250 tons of gasification ash from the PCD. The system was fully available during operations and did not require maintenance. It was tested in many different configurations. CFAD successfully operated using an automated solids level controller. The fully automated operation was possible for all scenarios of operation.

3.3.5 Advanced Syngas Cleanup

During TC18, the syngas cleanup unit was operated to demonstrate fixed bed syngas desulfurization and ammonia cracking and hydrocarbon reforming using nickel-based catalysts.

Prior to TC18, Sud-Chemie nickel-based catalyst, G-117RR (previously tested in TC16 and TC17) was tested in the mini reactor (RX301) to check the catalyst reactivity and sulfur poisoning effect using ammonia, benzene, and sulfur dioxide. A bottled ammonia gas simulation test was conducted at 1650ºF and 2 to 10 psig pressure for 4 hours. The inlet ammonia concentration was maintained at 3000 ppm, while the outlet ammonia level measured below 20 ppm. A benzene bottle gas simulation test was conducted at 800 to 1600ºF and at pressures from 2 to 10 psig for 3 hours. The inlet benzene concentration was maintained at 1000 ppm. The outlet benzene level was obtained below 50 ppm at 1600ºF. A sulfur dioxide bottle gas simulation test was conducted at temperatures from 1000 to 1600ºF and at pressures of 2 to 10 psig for 8 hours. The inlet concentration of ammonia and benzene were maintained at 3000 and 550 ppm respectively. The outlet concentrations of both ammonia and benzene measured below 50 ppm.

The following modifications were completed at the gas cleanup unit during the outage preceding TC18:

- A new back pressure controller was installed in the downstream of the hot gas unit to maintain the operating pressure of the hot vessels.
- A new flow controller was installed in the upstream of the hot gas unit to maintain the syngas flow rate through the hot vessels.
- A new pressure transmitter was installed in the upstream of the hot unit to measure the syngas pressure inlet to the hot vessels.
- A new recirculating loop chiller was installed in the cold gas unit.
- Gas sampling lines for Fourier transform infrared (FTIR) spectroscopy analyzer and gas chromatography analyzer (GC) were modified.

The major accomplishments and observations in TC18 gas cleanup test run included the following:

Syngas desulfurization was demonstrated in the gas cleanup unit for 76 hours using a Sud-Chemie sulfur sorbent (RVS-1). The nominal properties of the sulfur sorbent are shown in Table 3.3.4-1, and operating parameters for the reaction are shown in Table 3.3.4-3. The H_2S concentration at the inlet of the vessel typically ranged from 300 to 400 ppm, and that at the outlet was below the detectable limit. The desulfurization test run ended before achieving sulfur breakthrough at the vessel outlet. The temperature, pressure, and flow profiles are shown in Figure 3.3.4-1, and the sulfur profile is shown in Figure 3.3.4-2.

The mini reactor was operated for 303 hours using Sud-Chemie nickel-based catalysts (G-117RR previously tested and G-31). The nominal properties of the nickle-based catalysts are shown in Table 3.3.4-2. The actual operating parameters for ammonia cracking and hydrocarbon reforming are shown in Table 3.3.4-4.

With the G-117RR catalyst, which was tested for 290 hours, the ammonia concentration was reduced by about 96 percent on average; benzene was reduced by about 75 percent; ethylene was reduced by 97 percent; acenaphthene was reduced by 97 percent; phenanthrene was reduced by 88 percent; and naphthalene was reduced by 81 percent. The hydrocarbon profiles for the G-117RR test are shown in Figures 3.3.4-3 to 3.3.4-8.

Ammonia cracking and hydrocarbon reforming testing was conducted for 13 hours using the G-31 catalyst, which had not been tested previously. About 99 percent reduction in ammonia, benzene, ethylene, acenaphthene, and phenanthrene was achieved. Naphthalene was reduced by about 98 percent. The hydrocarbon profiles while testing the G-31 catalyst are shown in Figures 3.3.4-9 to 3.3.4-14.

The gas cleanup unit syngas cooler was tested for 107 hours using a new recirculating loop chiller to collect the condensate from syngas. The syngas flow was maintained at 40 lb/hr. The syngas temperature at the cooler inlet was 500ºF, and that at the outlet was 120ºF. The gas velocity in the cooler tube was 65 ft/sec. The exchanger tube was not plugged with organics.

Table 3.3.4-1

Nominal Properties of Sulfur Sorbent

Sorbent	RVS-1
Chemical Composition	Wt%
Zinc Oxide	40 - 60
Calcium Sulfate	15 - 25
Calcium Oxide	5 - 10
Nickel Oxide	5 - 15
Bentonite	5 - 15
Silica, Quartz	<5
Physical Properties	
Shape	Spheres
Size, mm	3 - 4
Density, lb/ft^3	60 - 85

Table 3.3.4-2

Nominal Properties of Ammonia Cracking and Hydrocarbon Reforming Catalysts

Catalyst	G-117RR	G-31
Chemical Composition	Wt %	Wt %
Magnesium Oxide	75 – 90	---
Nickel Oxide	5 – 15	1 - 25
Calcium Oxide	1 - 5	---
Aluminum Oxide	1 - 5	75 - 99
Physical Properties		
Shape	Rings	Spheres
Size, mm	3 - 4	1
Density, lb/ft^3	55 - 75	65 -100

Table 3.3.4-3

Operating Parameters for Fixed Bed Syngas Desulfurization

Sorbent	RVS-1
Sorbent Manufacturer	Sud-Chemie
Gasifier Operation	Air Blown
Coal Type	PRB
Reactor	RX700A
Reactor Size	5.2” ID x 5’ Height
Reactor Material	310 SS
Sorbent bed mass, lb	16
Sorbent bed height, in	28
Syngas flow rate, lb/hr	30
Pressure, psig	200
Temperature, °F	610
Space Velocity, hr^{-1}	1990
Inlet H_2S, ppm	340

Table 3.3.4-4

Operating Parameters for Fixed Bed Syngas Ammonia Cracking and Hydrocarbon Reforming

Gasifier Operation	Air Blown	
Coal Type	PRB	
Mini Reactor	RX301	
Mini Reactor Size	1.5” ID x 4’ Height	
Reactor Material	310SS	
Catalyst	G-117RR	G-31
Catalyst Manufacturer	Sud-Chemie	Sud-Chemie
Sorbent bed mass, lb	0.3	0.3
Sorbent bed height, in	5	5
Syngas flow rate, SCFH	10 - 12	15 - 20
Pressure, psig	2 – 10	2 – 10
Temperature, °F	1650	1650
Space Velocity, hr^{-1}	2155	3430
NH_3 inlet, ppm	2040	2250
NH_3 outlet, ppm	86	6
Benzene inlet, ppm	860	824
Benzene outlet, ppm	210	8

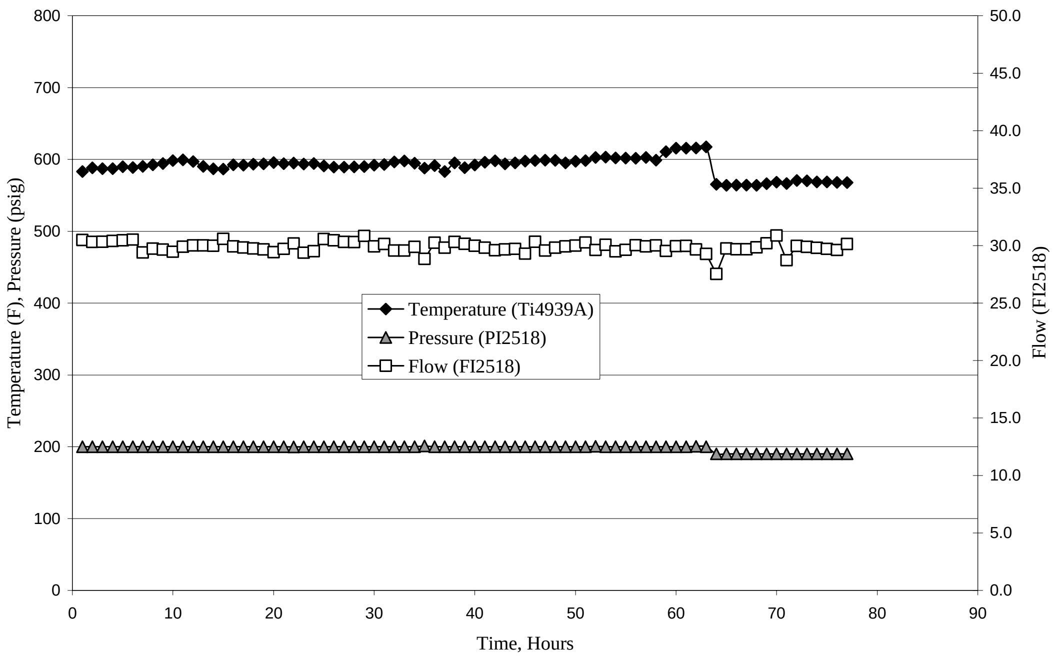

Figure 3.3.4-1 Temperature, Pressure and Flow Profiles, RVS-1 Sulfur Sorbent

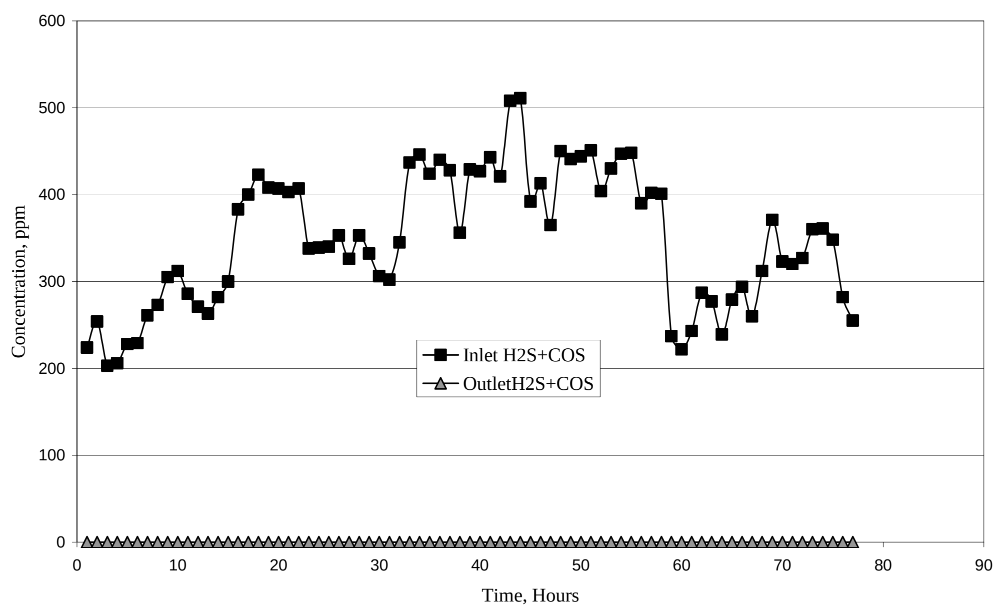

Figure 3.3.4-2 Sulfur Profile, RVS-1 Sulfur Sorbent

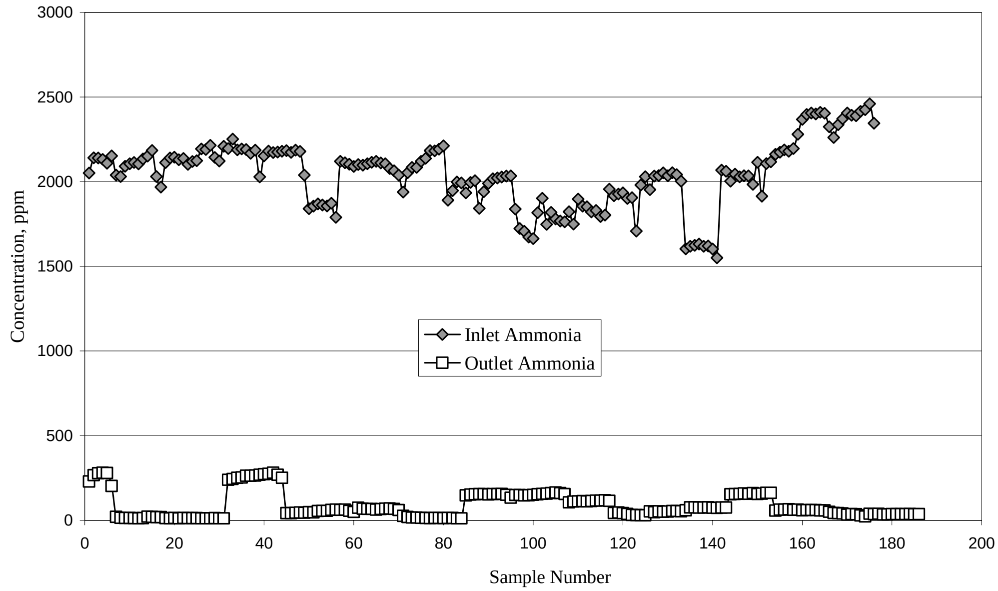

Figure 3.3.4-3 Ammonia Profile, G-117RR Catalyst

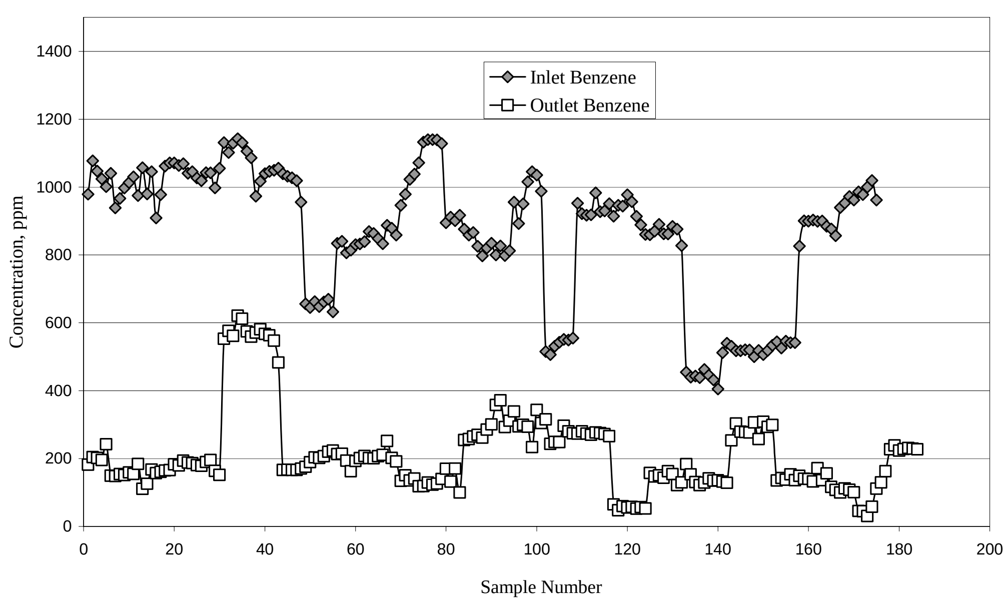

Figure 3.3.4-4 Benzene Profile, G-117RR Catalyst

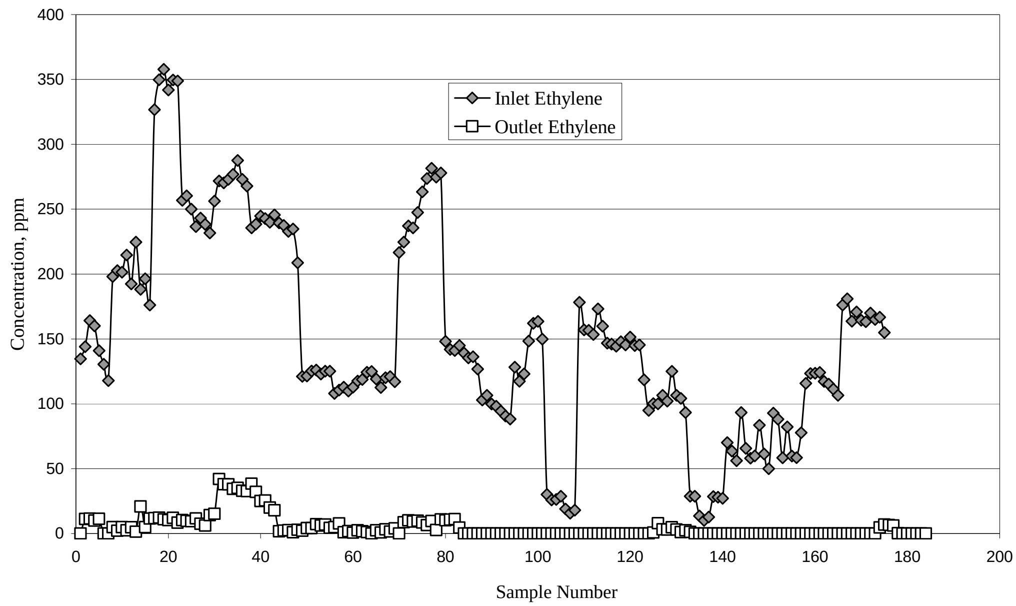

Figure 3.3.4-5   Ethylene Profile, G-117RR Catalyst

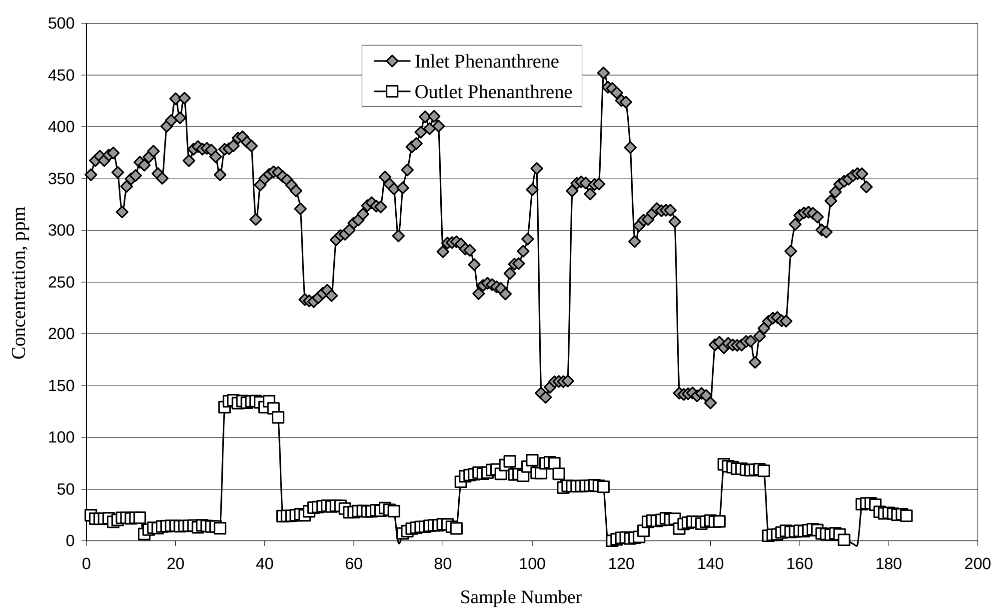

Figure 3.3.4-6   Phenanthrene Profile, G-117RR Catalyst

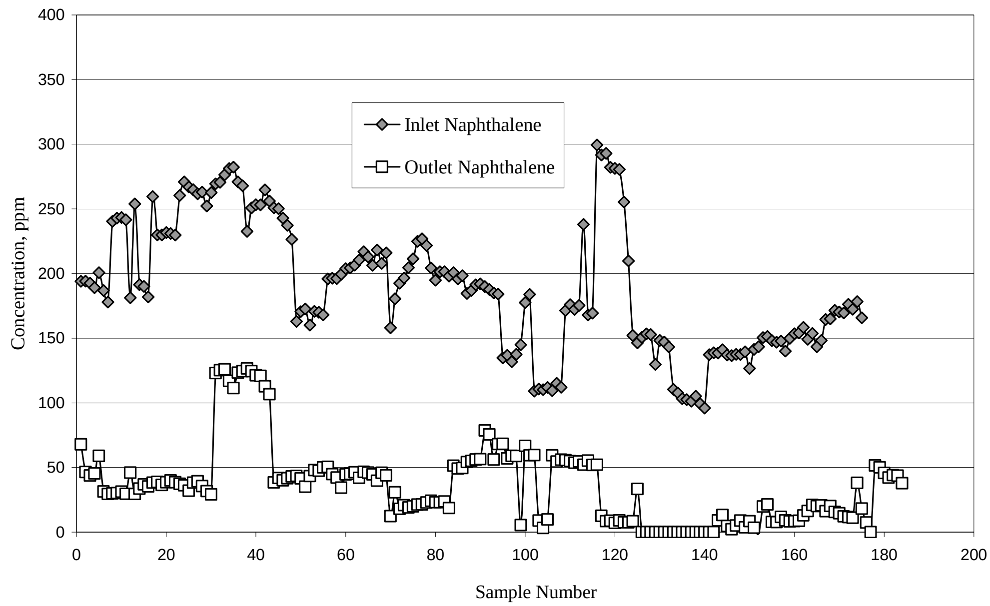

Figure 3.3.4-7 Naphthalene Profile, G-117RR Catalyst

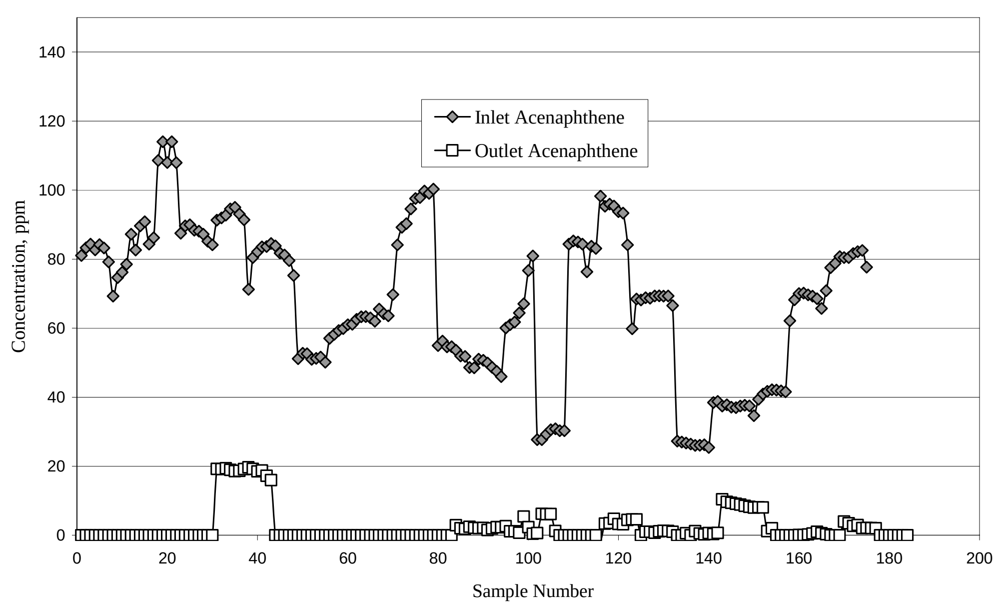

Figure 3.3.4-8 Acenaphthene Profile, G-117RR Catalyst

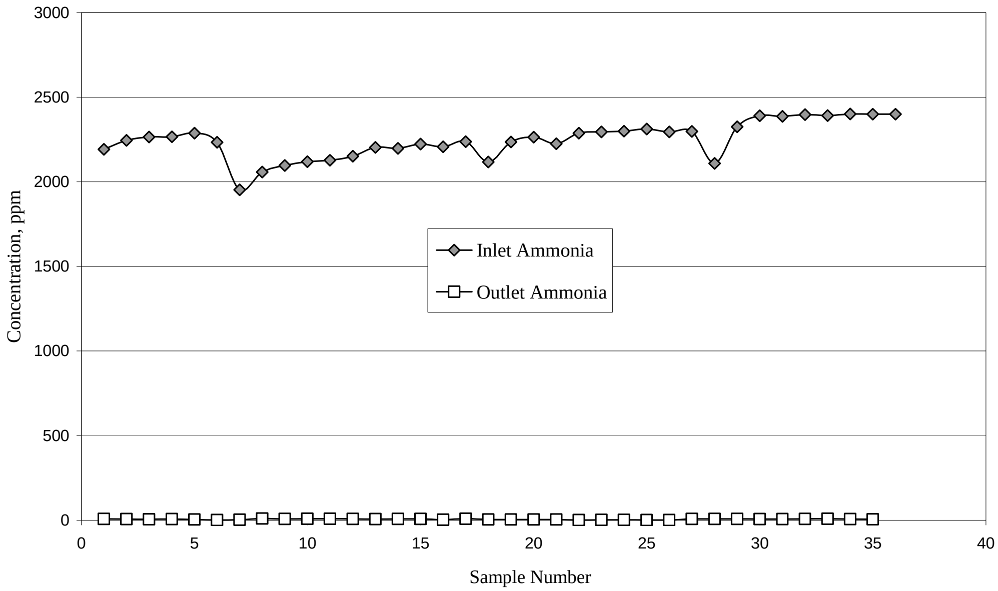

Figure 3.3.4-9 Ammonia Profile, G-31 Catalyst

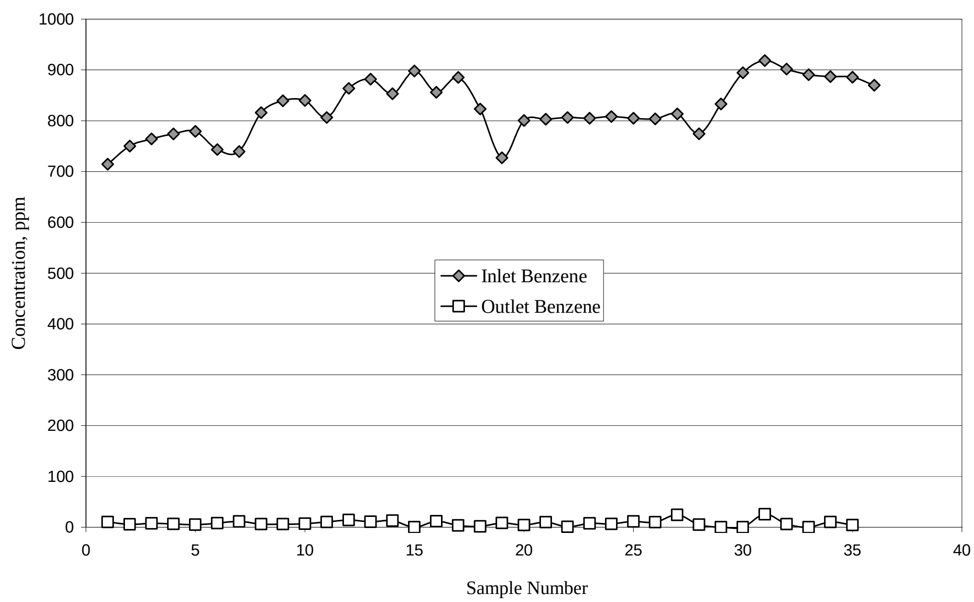

Figure 3.3.4-10 Benzene Profile, G-31 Catalyst

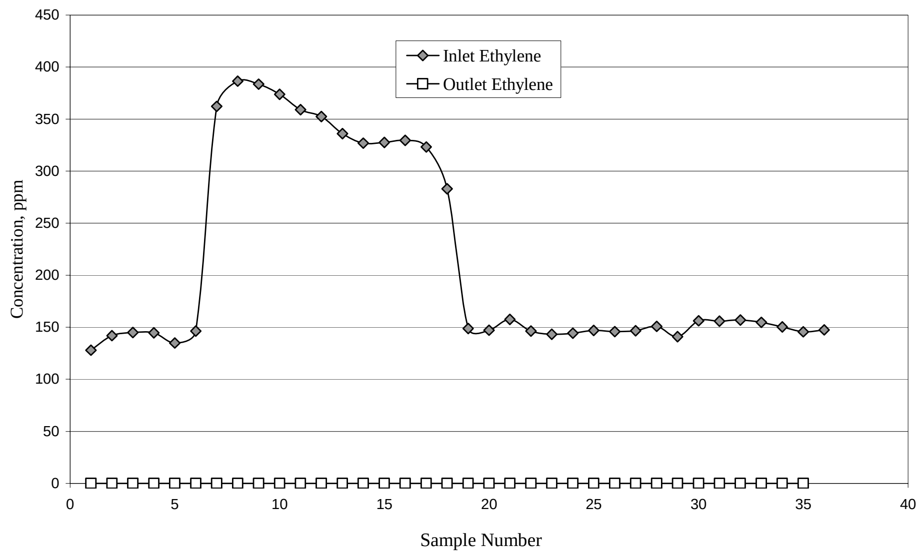

Figure 3.3.4-11 Ethylene Profile, G-31 Catalyst

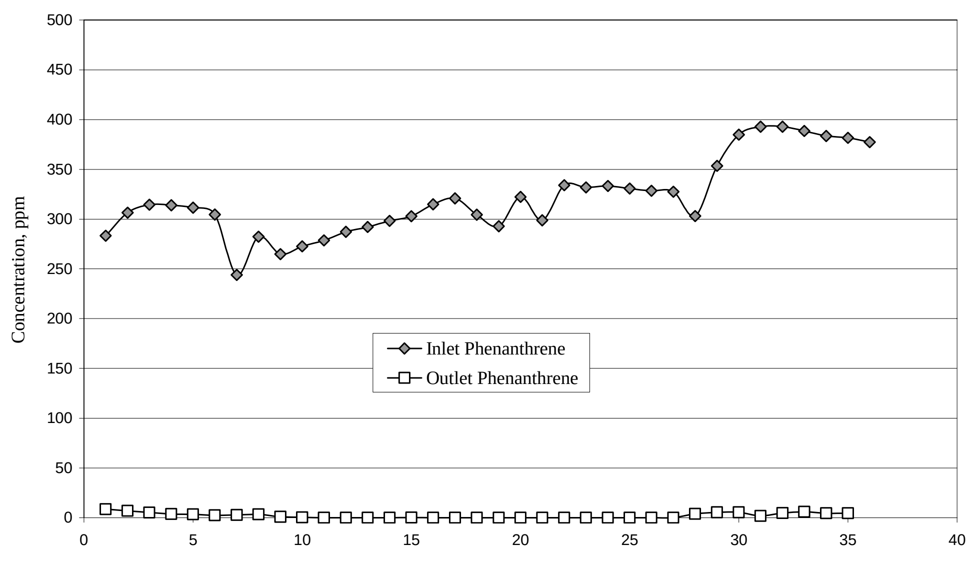

Figure 3.3.4-12 Phenanthrene Profile, G-31 Catalyst

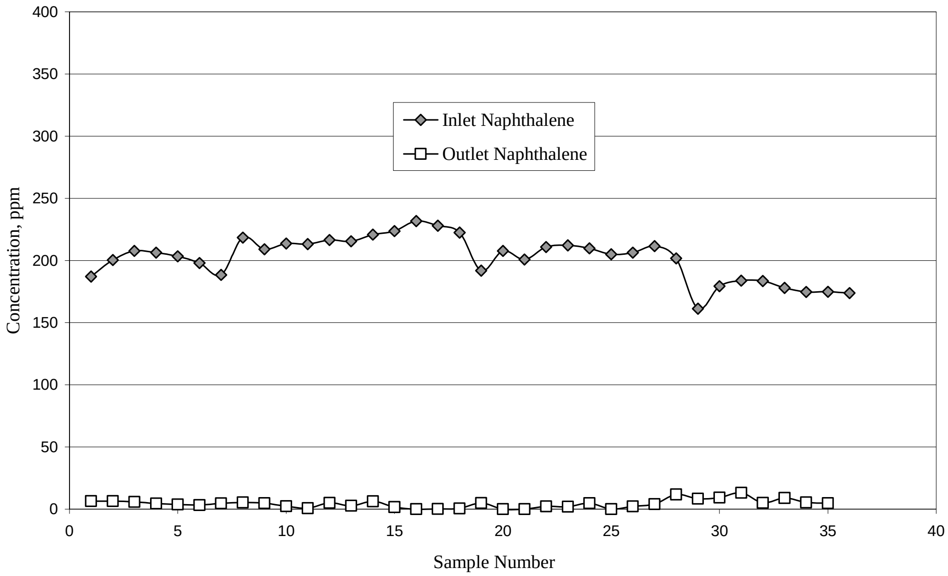

Figure 3.3.4-13   Naphthalene Profile, G-31 Catalyst

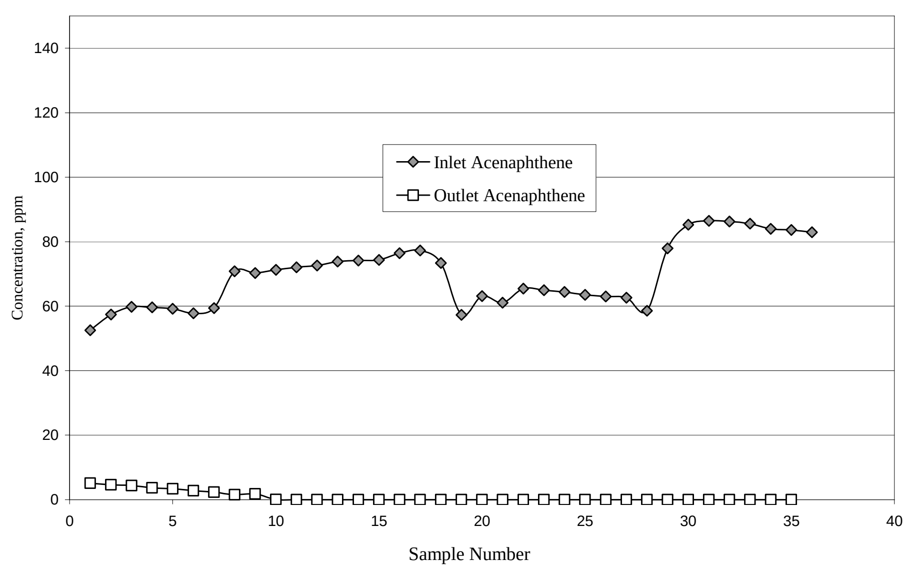

Figure 3.3.4-14   Acenaphthene Profile, G-31 Catalyst

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 to commission 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 Bucyrus limestone from Ohio. 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 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 Bucyrus limestone from Ohio 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 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 Bucyrus limestone from Ohio. Due to operational difficulties with the gasifier (stemming from instrumentation problems) the unit was taken offline several times. PCD operations were relatively stable considering the numerous gasifier upsets.

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. Both gasifier and PCD operations were stable during the test run.

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

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 even 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 of operations 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 percent 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. The gasifier operation was smooth, with the exception of the decrease in the primary cyclone efficiency which caused the gasifier to continuously lose bed material.

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 percent of the total energy to the PSB. The gasifier experienced stable operations in air blown mode and 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. The system was effective in reducing the sulfur content and achieving an absorption capacity between 19 and 30 percent.

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 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 syngas-to-PSB testing lasted for approximately 6 ½ hours at syngas flow rates up to 13,000 pph. The gasifier experienced stable operations with the bituminous coal.

TC18, the subject of this report, was a 1,342-hour test campaign to conduct short-term tests to evaluate gasifier, recycle gas, and PCD operations in air blown mode of operation using a blend of several PRB coals. TC18 began on June 23, 2005, and ended on August 22, 2005. A primary focus of TC18 was to commission the recycle gas compressor and test modification to the continuous fine ash depressurization (CFAD) unit. The recycle gas compressor operated for 376 hours, and increased the raw heating value approximately 8 to 10 percent. The CFAD system operated reliably for 1,300 hours, and removed over 250 tons of gasification ash from the PCD. Overall gasifier operation was smooth, with the exception of the inefficient primary cyclone, necessitating frequent addition of bed material.

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

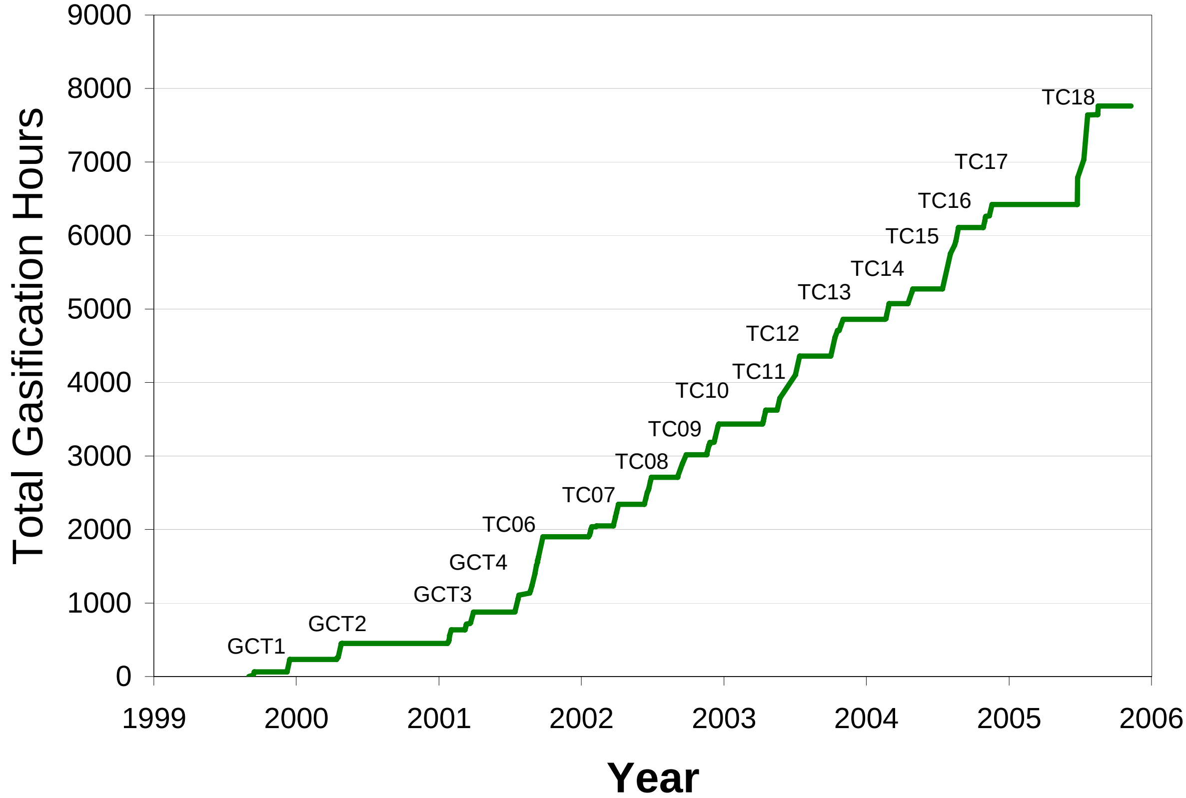

Figure A1-1 Operating Hours Summary for the PSDF Gasification Process

APPENDIX A2 EQUIPMENT LIST

Major Equipment in the Transport Gasifier Train

TAG NAME	DESCRIPTION
BR0201	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 Gasifier Loop
CY0207	Disengager in the Gasifier 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
FD0510	Spent Solids Transporter System
FD0530	Spent Solids Feeder System
FD0540	Continuous Fine Ash Depressurization (CFAD)
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
ME0540	Heat Transfer Fluid System
RX0201	Transport Gasifier
SI0602	Spent Solids Silo
SU0601	AFBC

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

TAG NAME	DESCRIPTION
BO2920	Auxiliary Boiler
BO2921	Auxiliary Boiler – Superheater
CL2100	Cooling Tower
CO2201A-D	Service Air Compressor A-D
CO2202	Air-Cooled Service Air Compressor
CO2203	High-Pressure Air Compressor
CO2601A-C	Reciprocating Nitrogen Compressor A-C
CR0104	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
HO0100	Reclaim Hopper
HO0105	Crushed Material Surge Hopper
HO0252	Coal Surge Hopper
HO0253	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	Uninterruptible Power Supply (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	Nitrogen 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 MATERIAL 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 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 ±10 percent. The relative difference (relative error) is defined as the Transport Gasifier feeds minus the products divided by the feeds ({In-Out}/In).

The main contributors to the material balance are the syngas flow rate (13,600 to 24,500 pounds per hour), the air flow rate (8,900 to 13,500 pounds per hour), the steam flow rate (20 to 1,600 pounds per hour), the nitrogen flow rate (3,000 to 6,600 pounds per hour), and the coal feed rate (2,400 to 4,600 pounds per hour). All of the data points with recycle gas fell within the ±10 percent range. Only one of the data points without recycle gas fell outside of the ±10 percent range; however, the data point was within ±13 percent.

The TC18 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 TC18 energy balances were mostly within ±10 percent error with the exception of three that fell within ±15 percent error. The energy entering the gasifier consisted of the coal, air, steam, and recycle gas fed to the Transport Gasifier. The nitrogen, oxygen and sand fed to the gasifier were considered to be at standard conditions (80°F) and, hence, had zero enthalpy. The nitrogen and oxygen feeds 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 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 TC18 carbon balance is shown in Figure A3-3. The carbon balance gives a measure of the accuracy of the carbon conversions. The most probably sources of error in the carbon balance are the coal feed rate measurement and the syngas flow rate measurement. All of the TC18 operating periods carbon balances were within ± 17 percent. Most of the carbon balances have a bias in that the data indicates that more carbon entered the system than left the system. This might indicate that the coal feed weigh cells were reading higher than actual or the syngas flow meter was reading lower than actual.

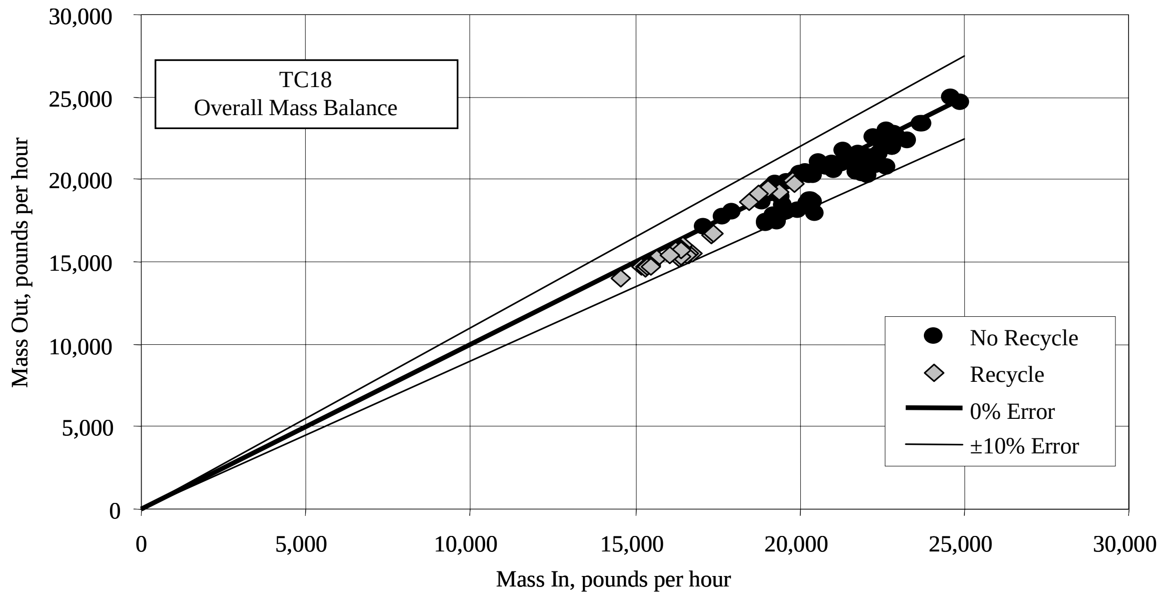

Figure A3-1 Mass Balance

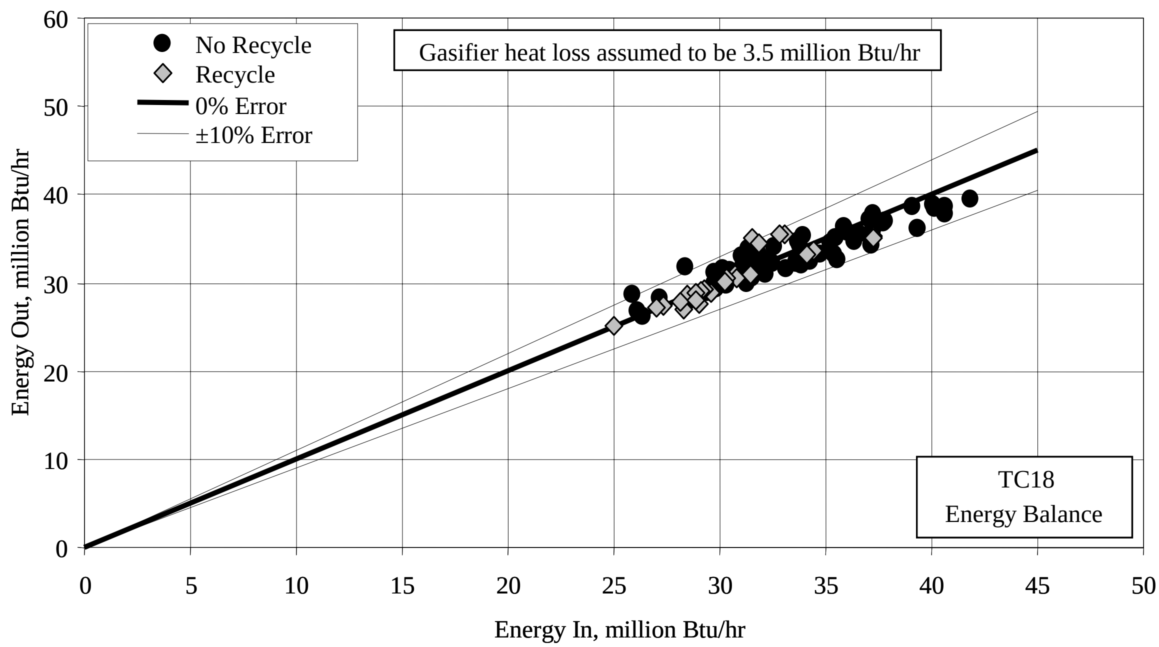

Figure A3-2 Energy Balance

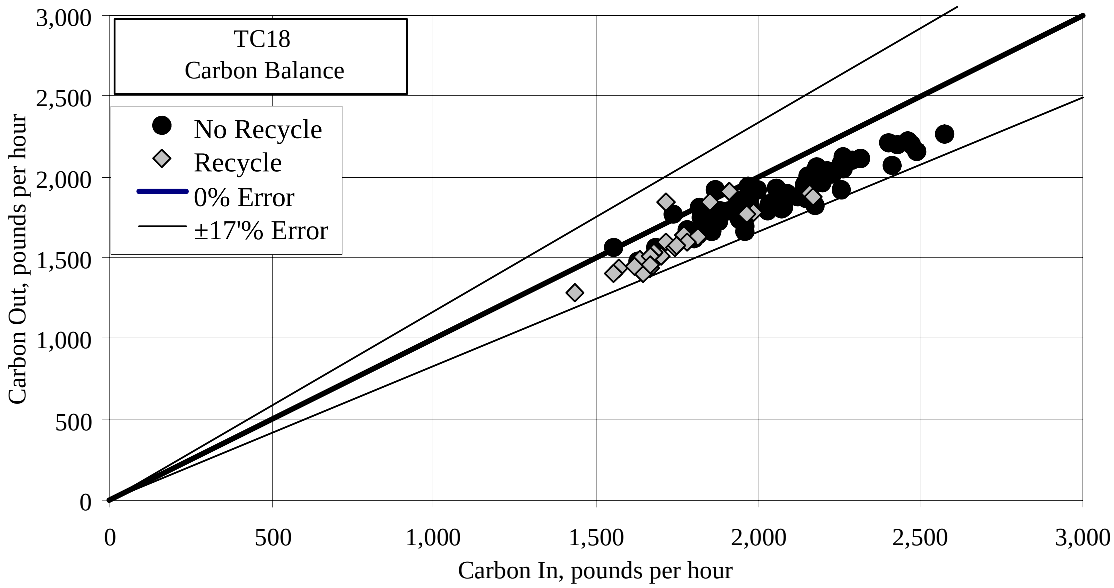

Figure A3-3 Carbon Balance

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APPENDIX A4 OPERATING TRENDS

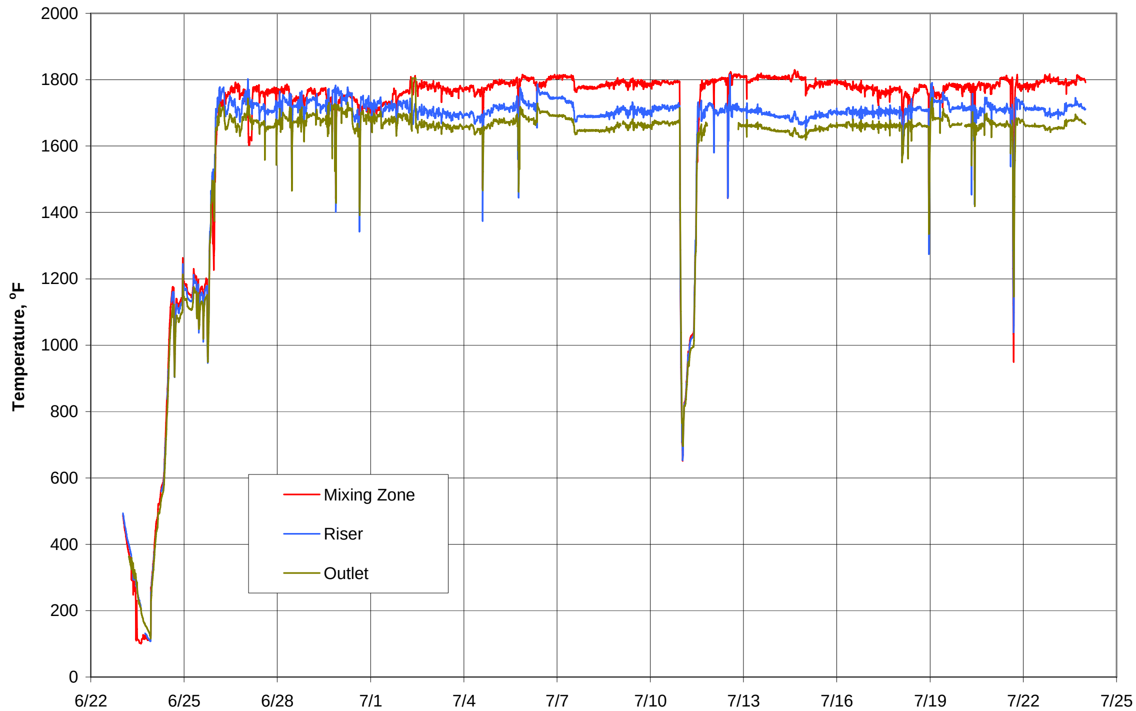

Figure A4-1 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 6/23/05 through 7/24/05

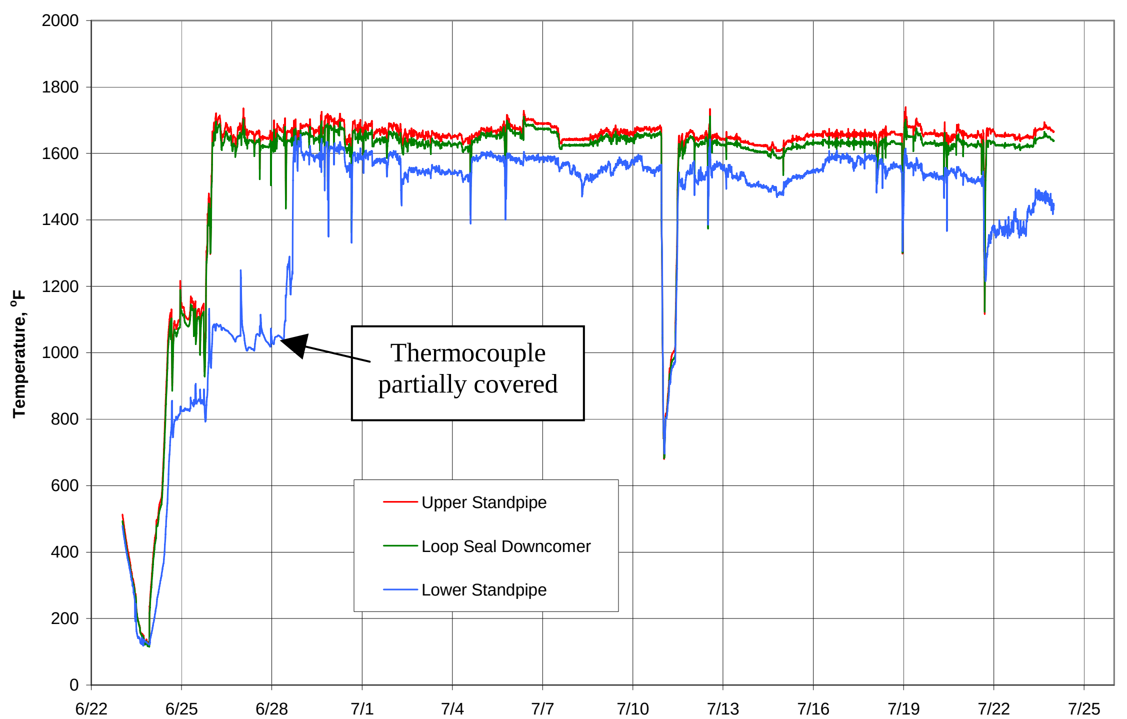

Figure A4-2 Standpipe and Loop Seal Temperatures, 6/23/05 through 7/24/05

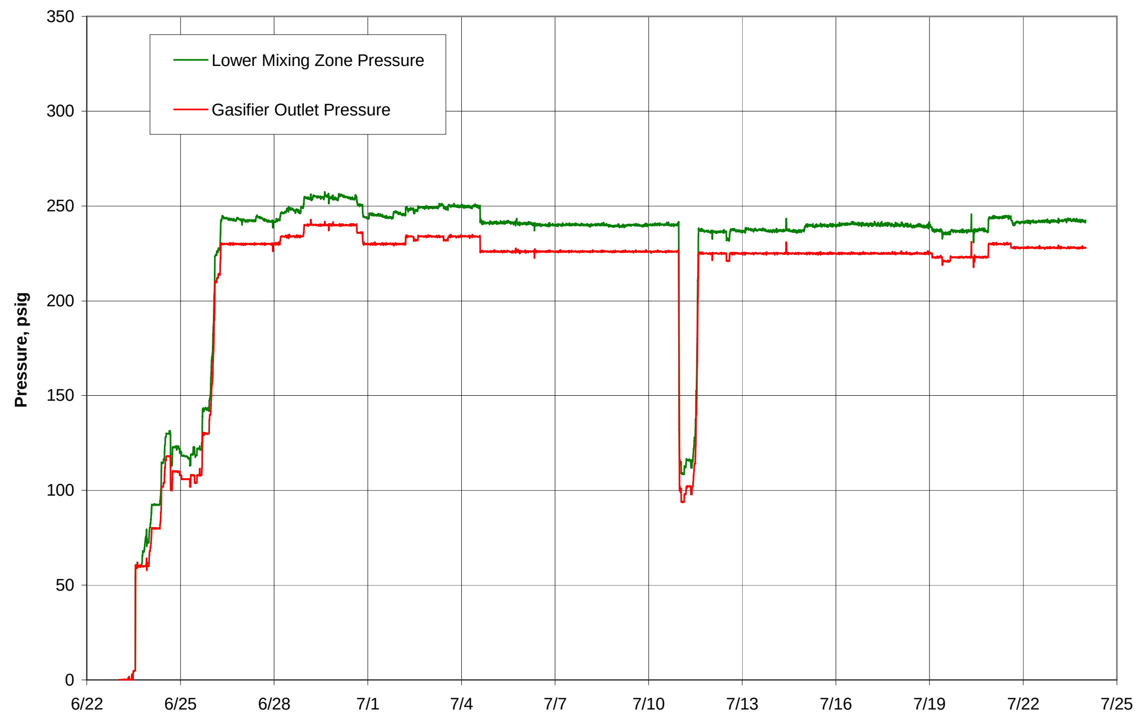

Figure A4-3 Gasifier Pressures, 6/23/05 through 7/24/05

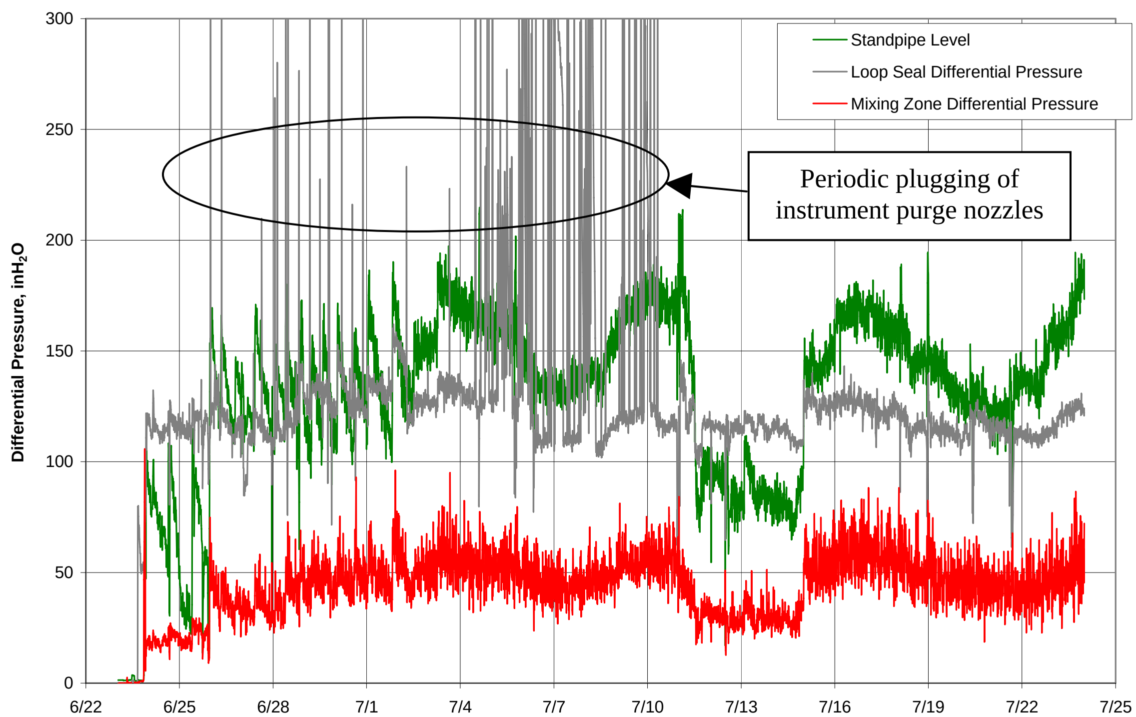

Figure A4-4 Gasifier Differential Pressures, 6/23/05 through 7/24/05

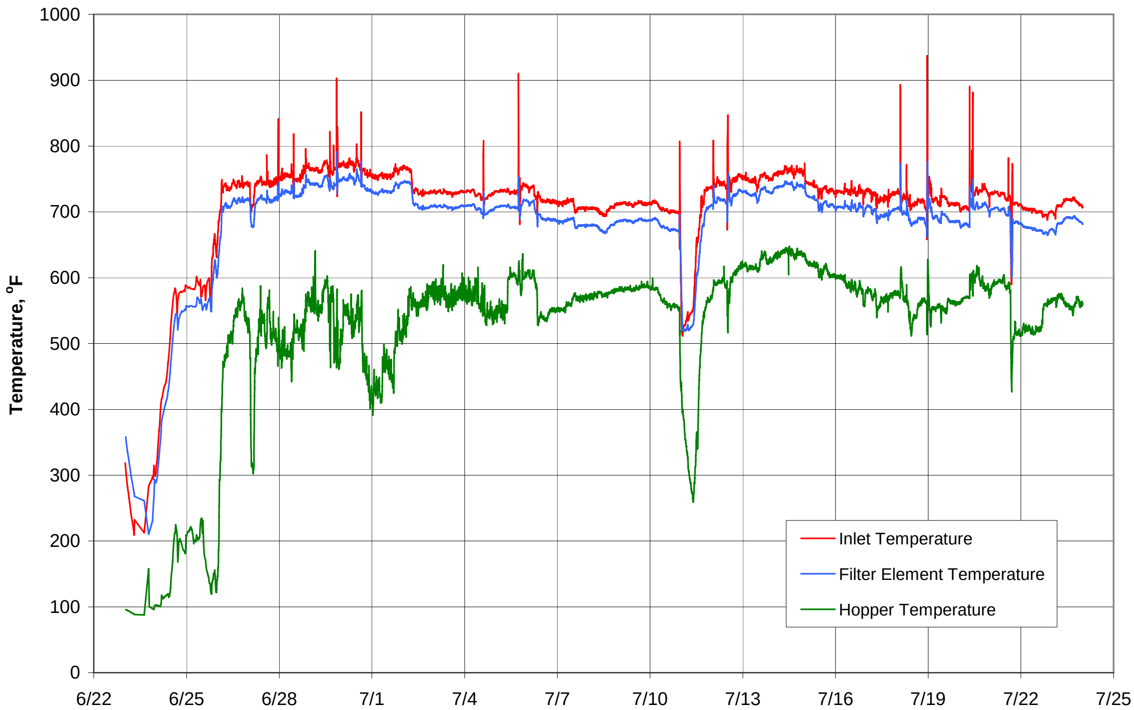

Figure A4-5 PCD Temperatures, 6/23/05 through 7/24/05

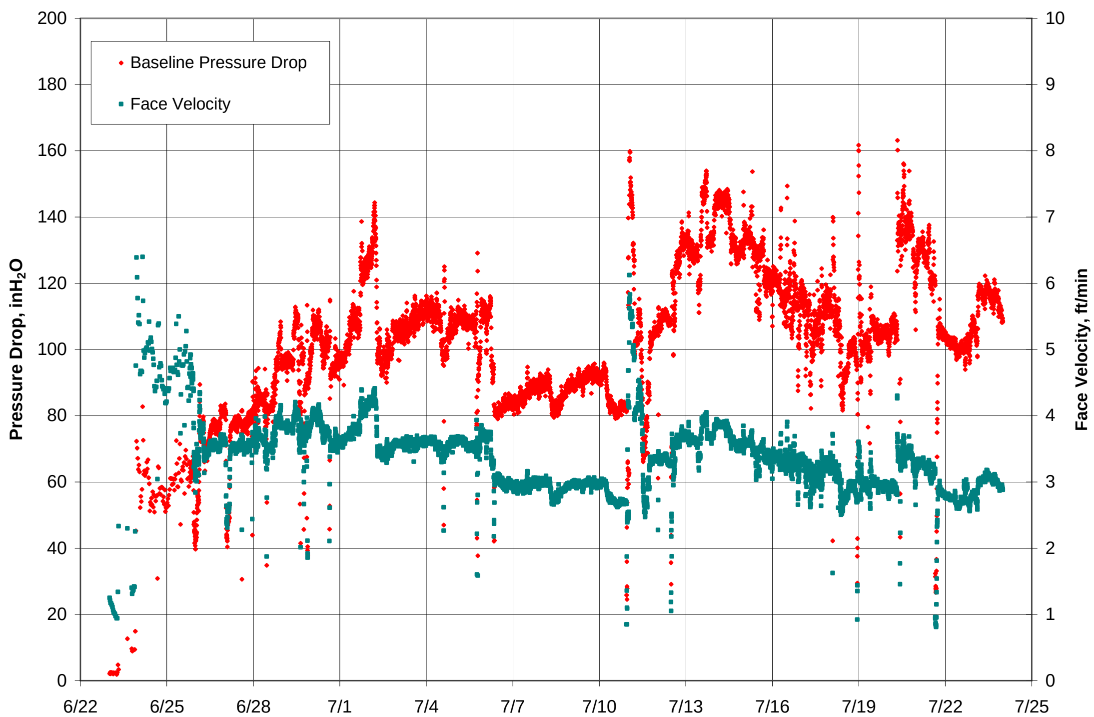

Figure A4-6 PCD Baseline Pressure Drop and Face Velocity, 6/23/05 through 7/24/05

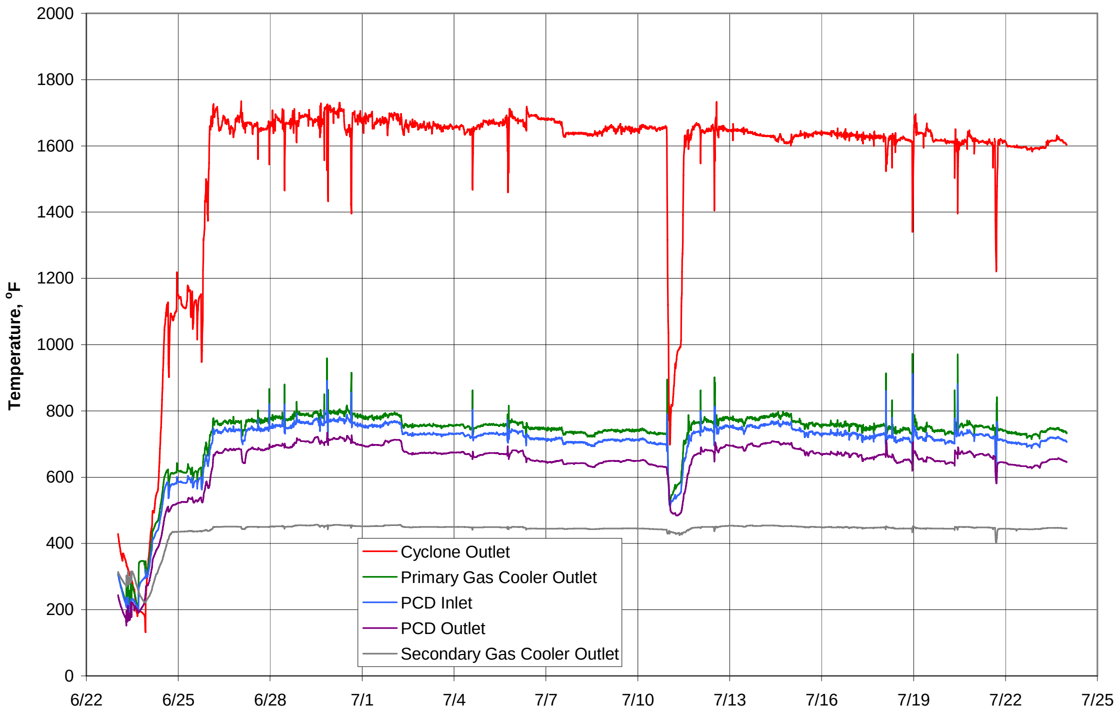

Figure A4-7 System Temperature Profile, 6/23/05 through 7/24/05

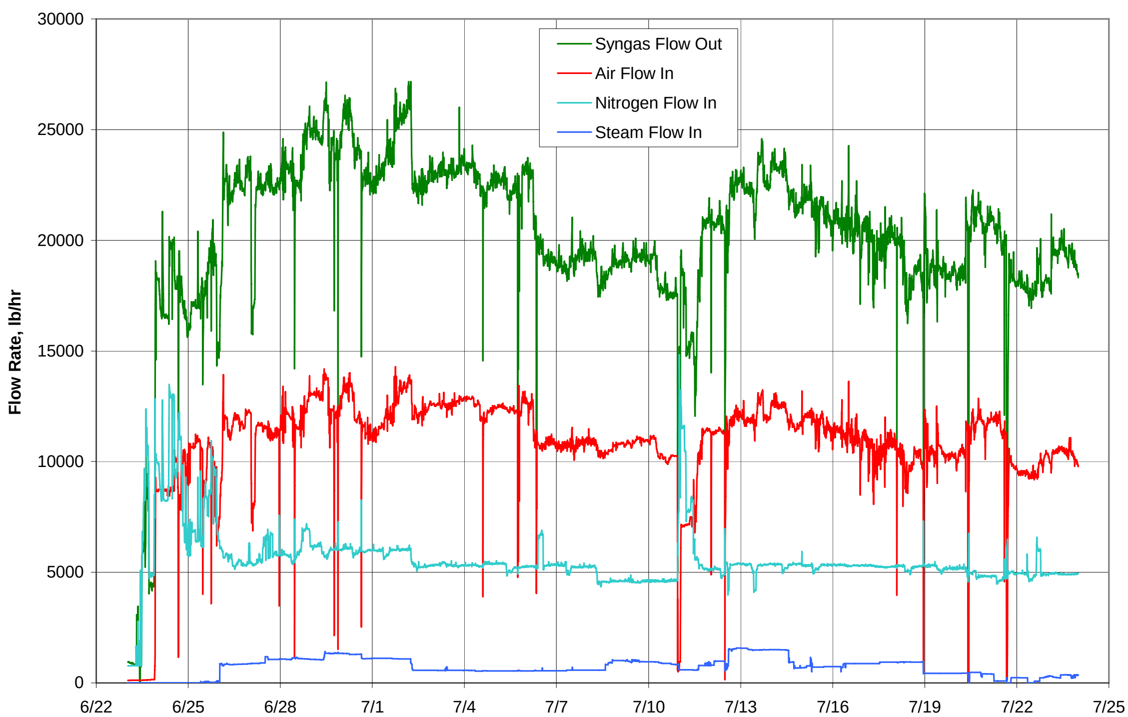

Figure A4-8 System Gas Flows, 6/23/05 through 7/24/05

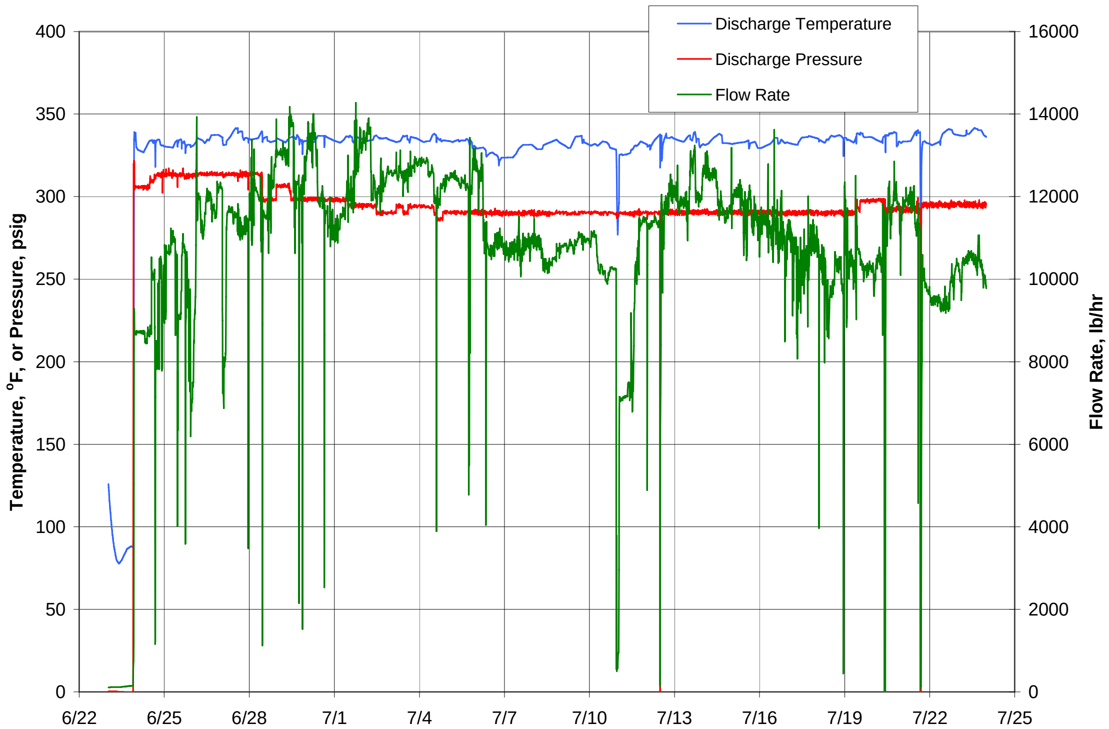

Figure A4-9 Main Air Compressor Operation, 6/23/05 through 7/24/05

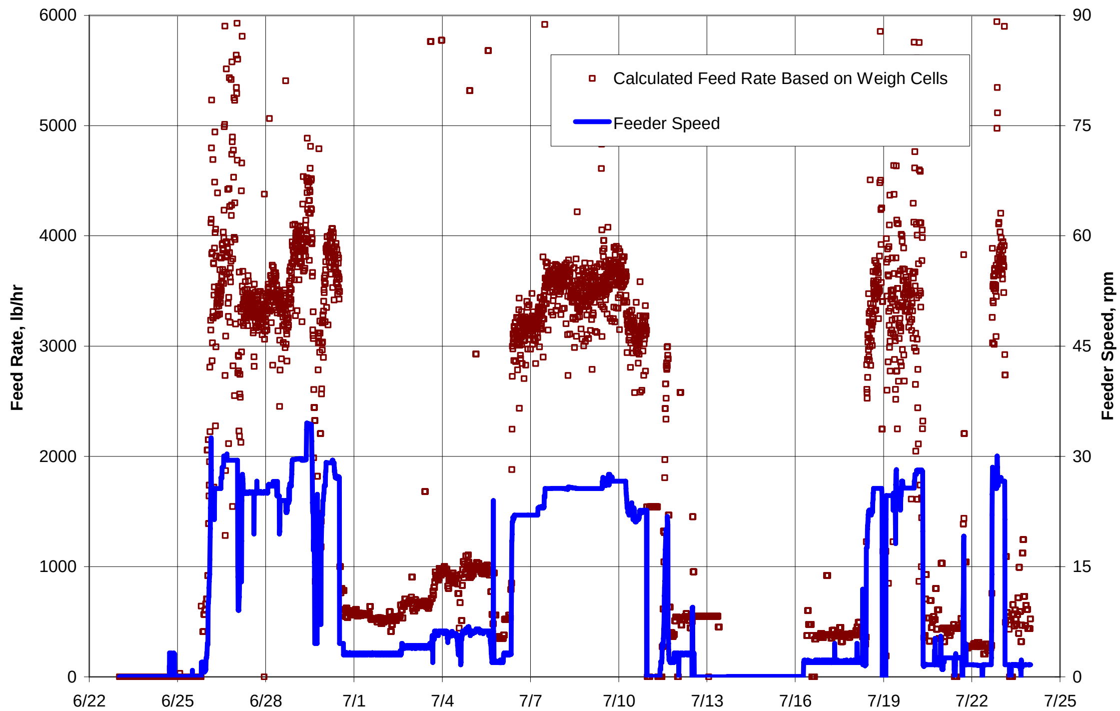

Figure A4-10 Original Coal Feeder Operation, 6/23/05 through 7/24/05

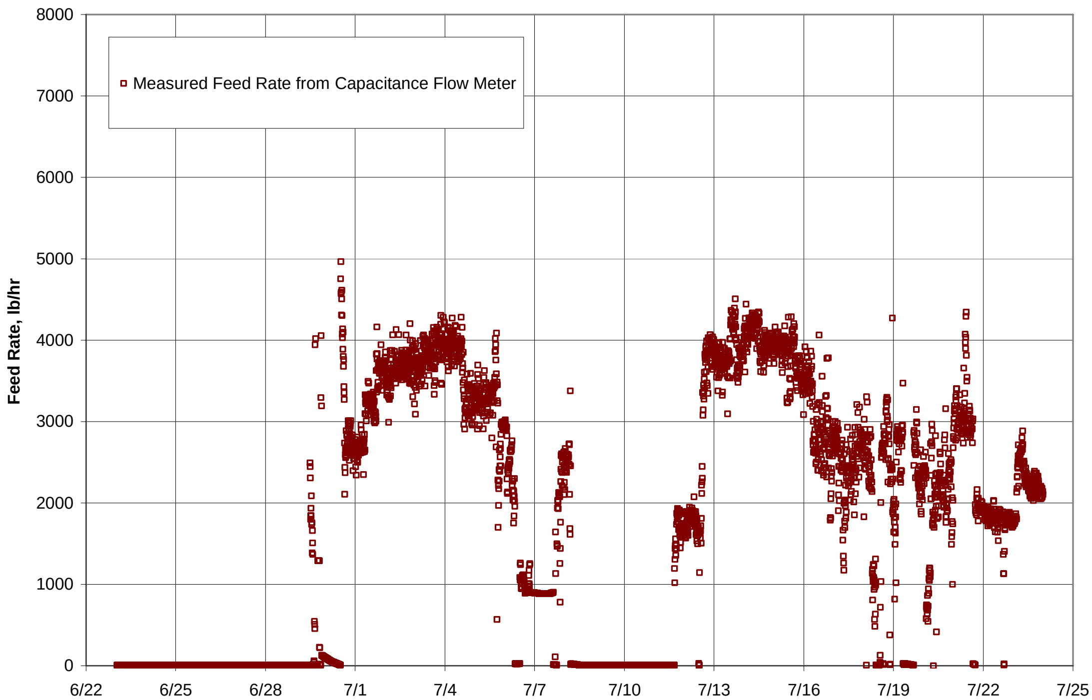

Figure A4-11 Developmental Coal Feeder Operation, 6/23/05 through 7/24/05

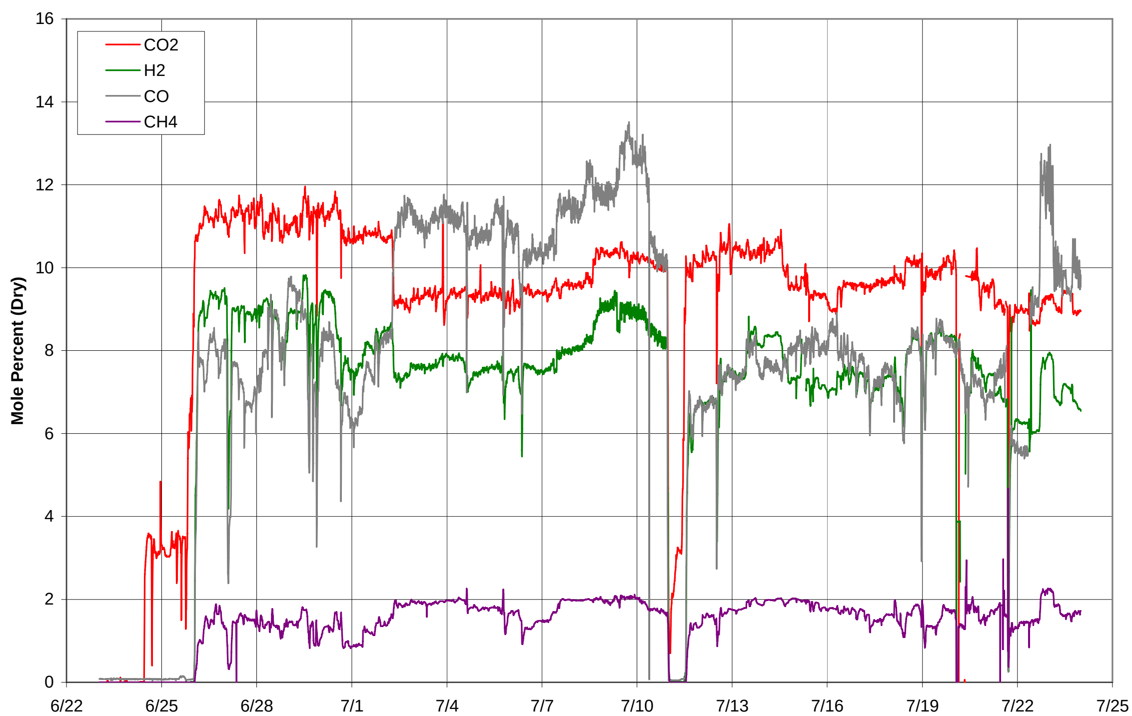

Figure A4-12 Syngas Analyzers, 6/23/05 through 7/24/05

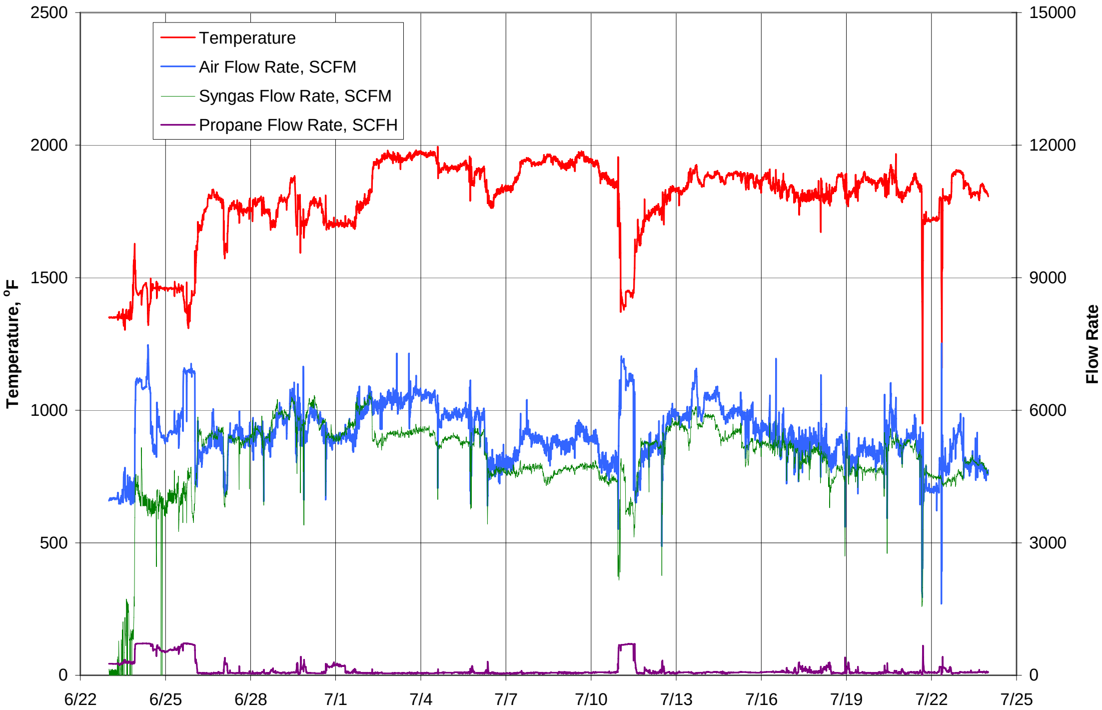

Figure A4-13 Atmospheric Syngas Combustor Operation, 6/23/05 through 7/24/05

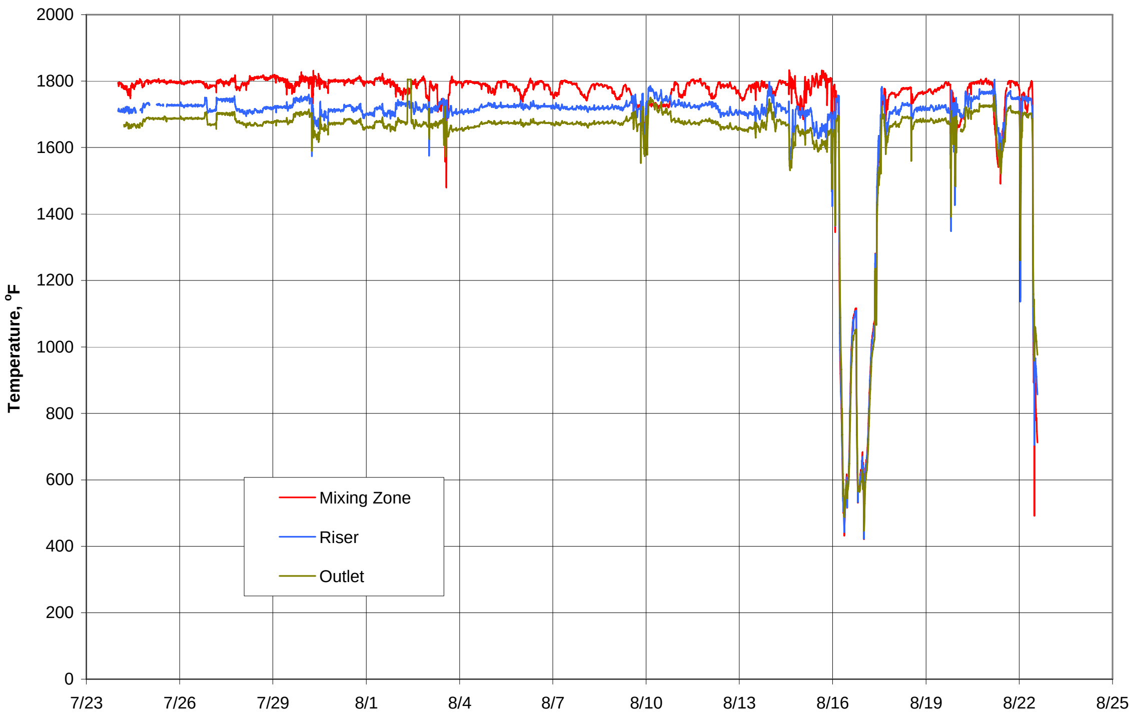

Figure A4-14 Gasifier Mixing Zone, Riser, and Outlet Temperatures, 7/24/05 through 8/23/05

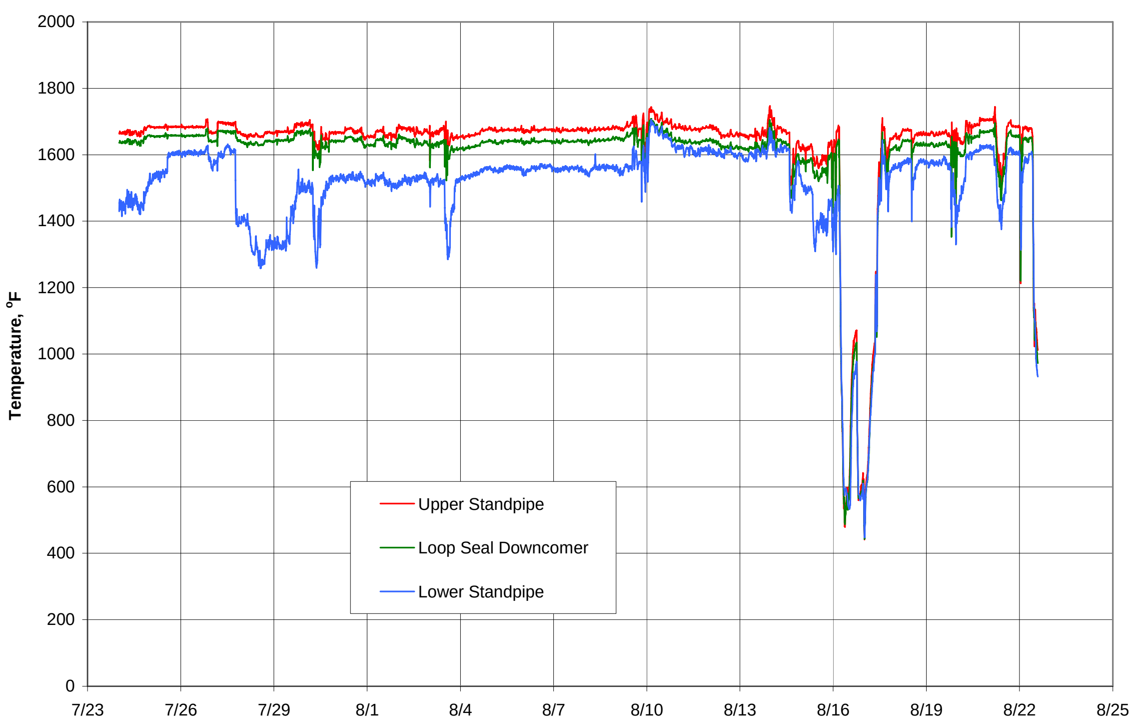

Figure A4-15 Standpipe and Loop Seal Temperatures, 7/24/05 through 8/23/05

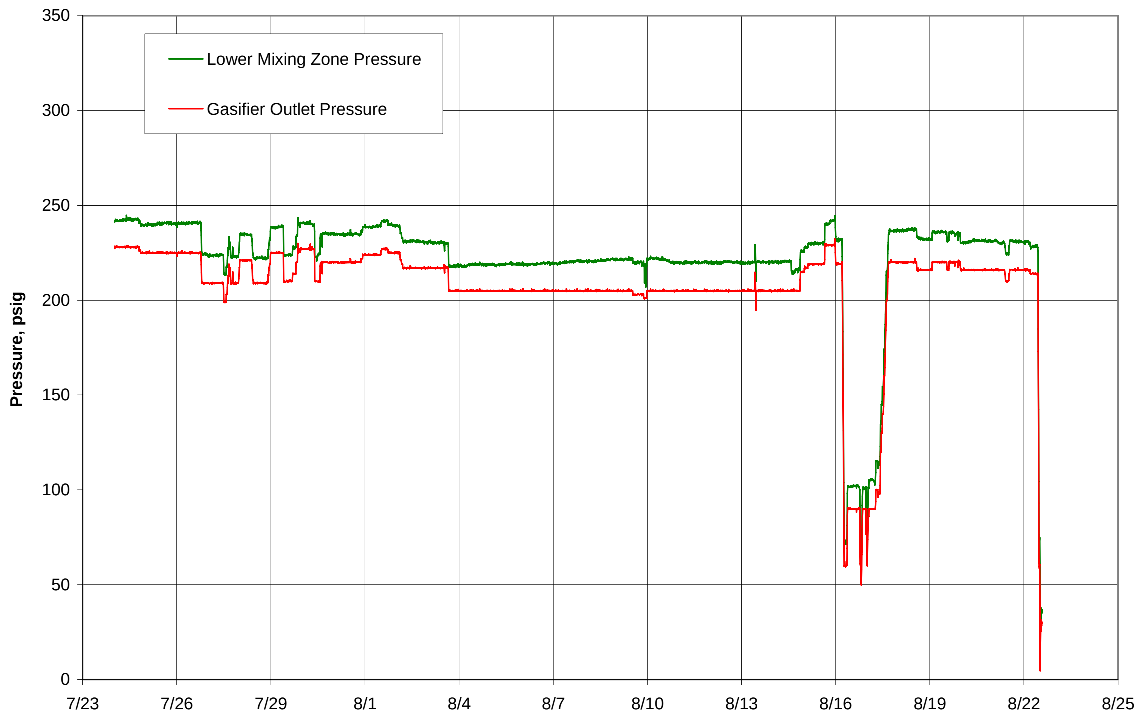

Figure A4-16 Gasifier Pressures, 7/24/05 through 8/23/05

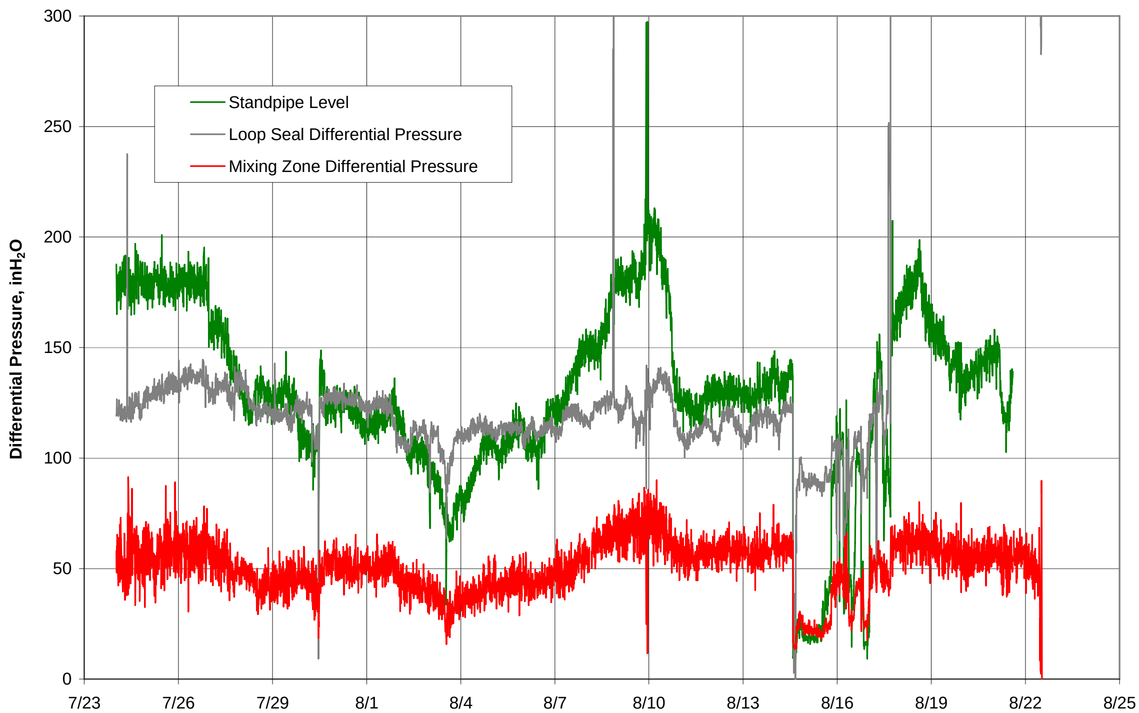

Figure A4-17 Gasifier Differential Pressures, 7/24/05 through 8/23/05

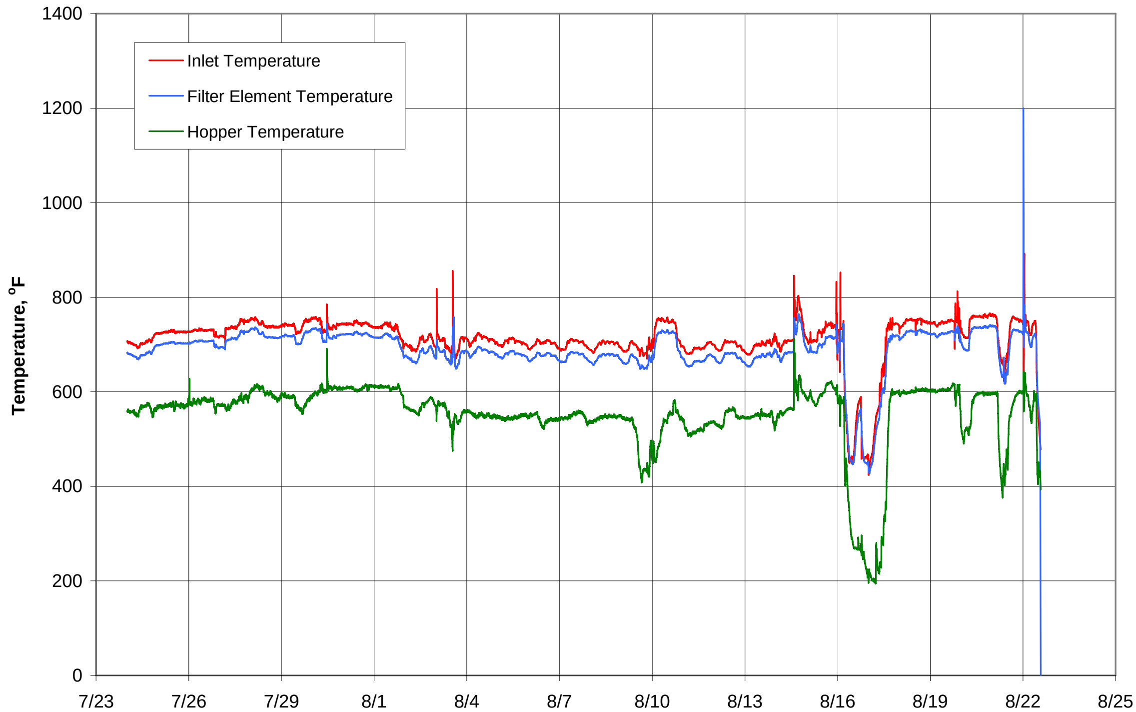

Figure A4-18 PCD Temperatures, 7/24/05 through 8/23/05

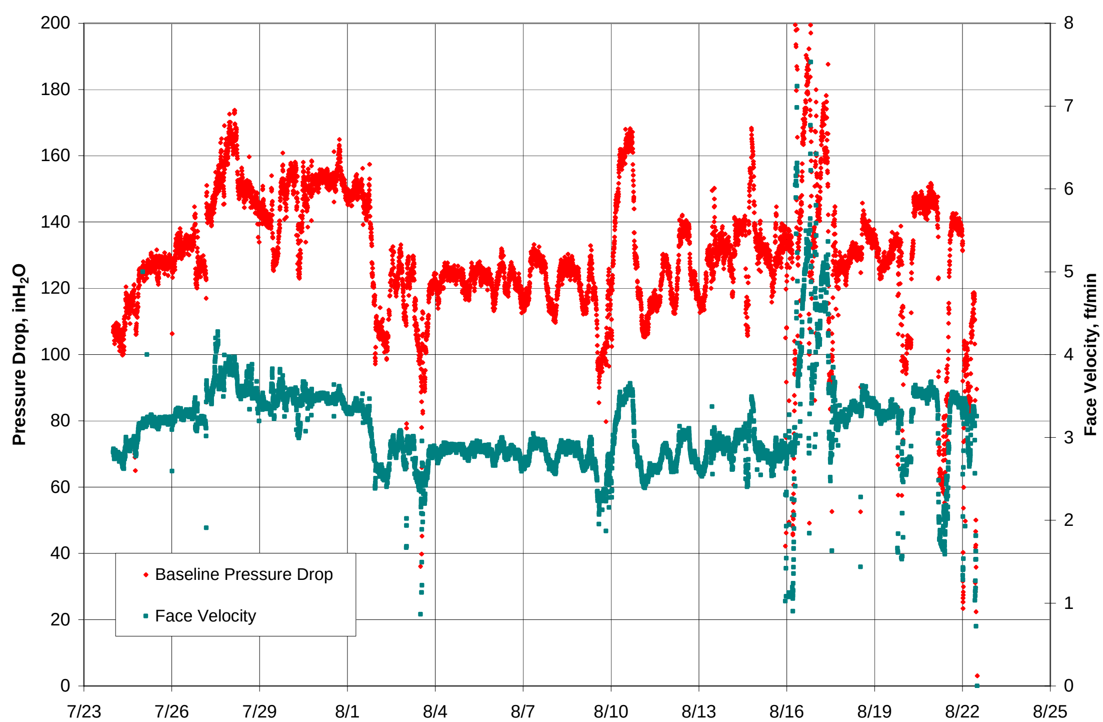

Figure A4-19 PCD Baseline Pressure Drop and Face Velocity, 7/24/05 through 8/23/05

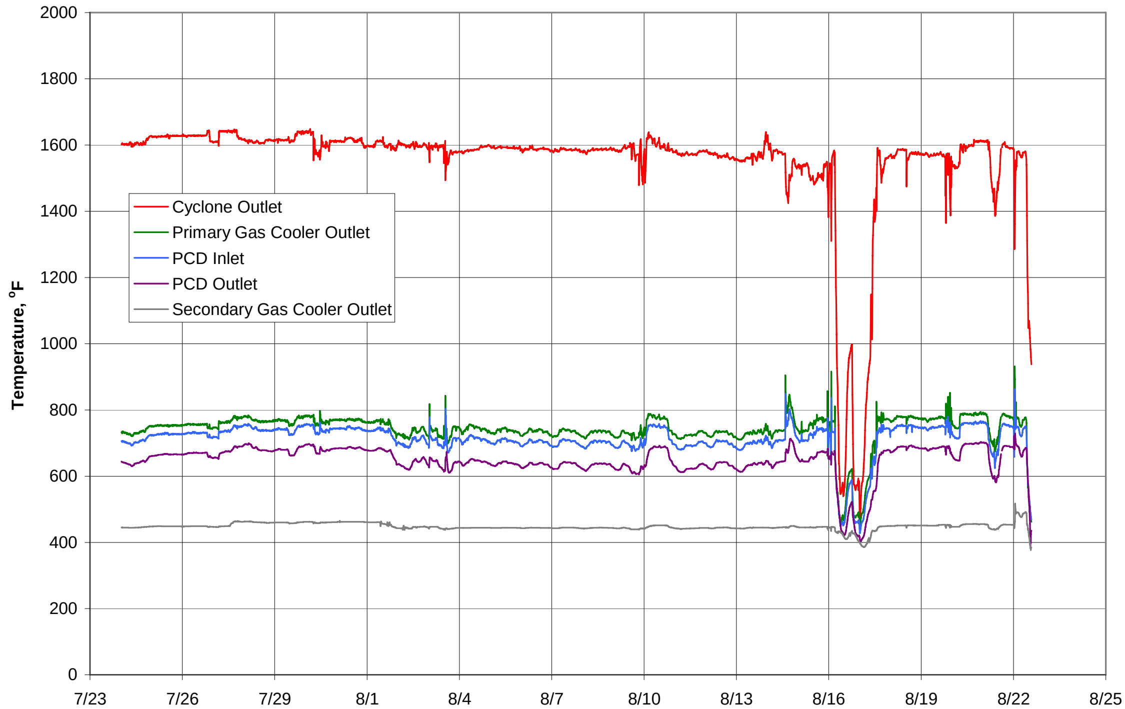

Figure A4-20 System Temperature Profile, 7/24/05 through 8/23/05

Figure A4-21 System Gas Flows, 7/24/05 through 8/23/05

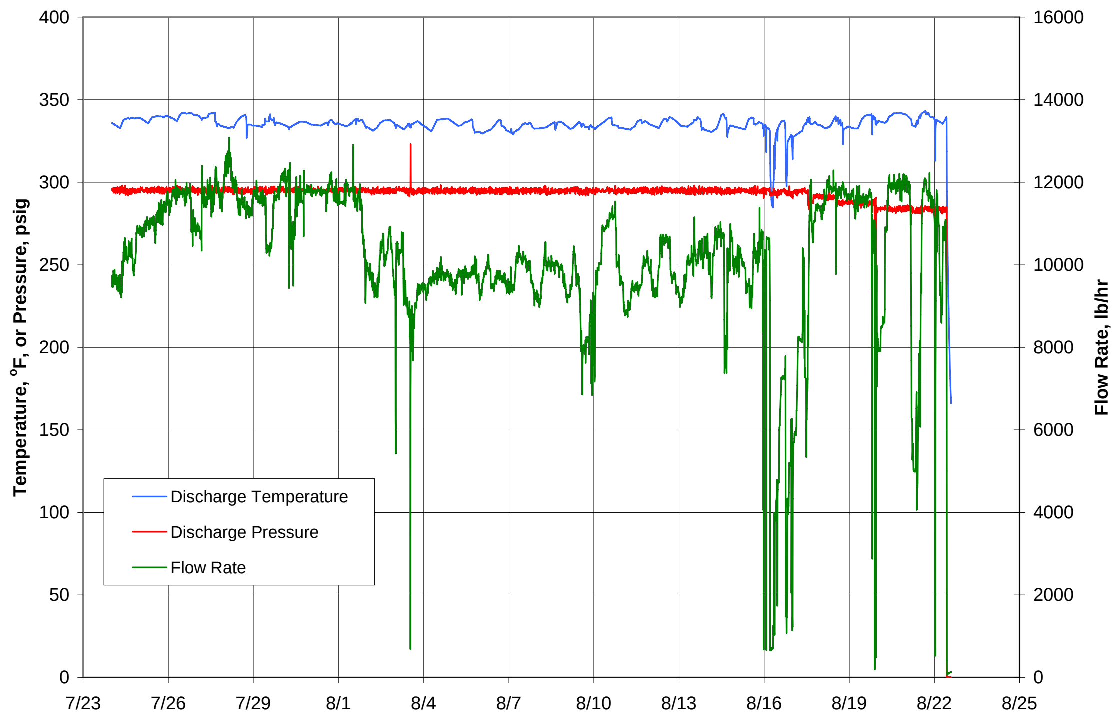

Figure A4-22 Main Air Compressor Operation, 7/24/05 through 8/23/05

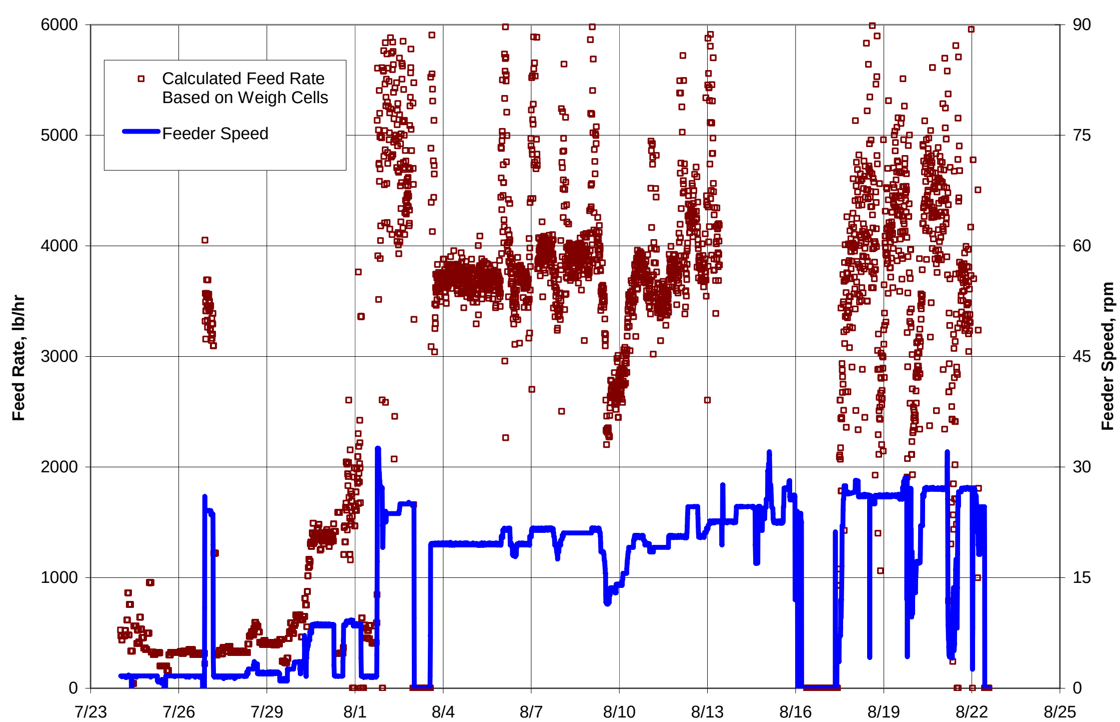

Figure A4-23 Original Coal Feeder Operation, 7/24/05 through 8/23/05

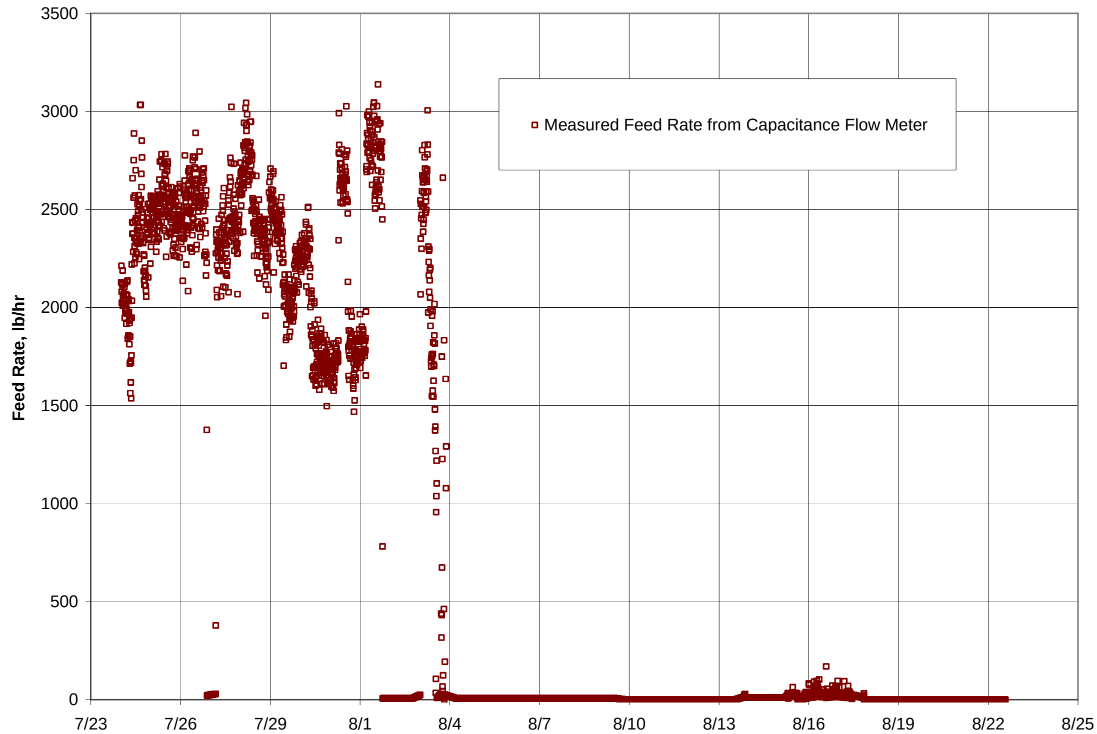

Figure A4-24 Developmental Coal Feeder Operation, 7/24/05 through 8/23/05

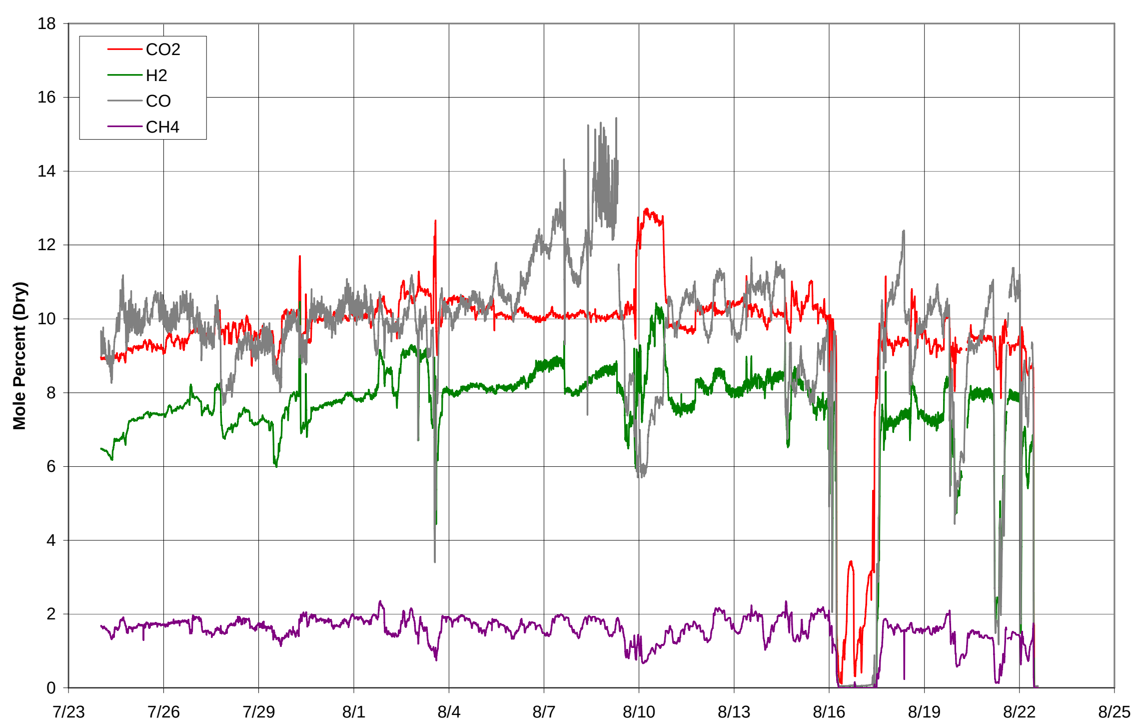

Figure A4-25 Syngas Analyzers, 7/24/05 through 8/23/05

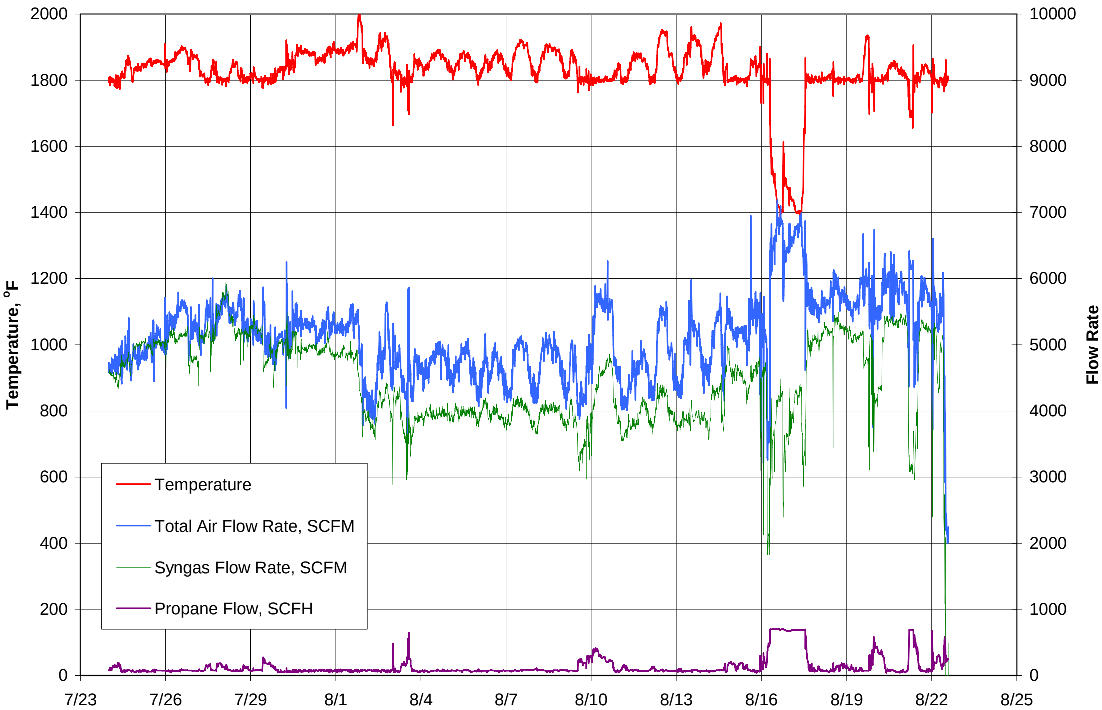

Figure A4-26 Atmospheric Syngas Combustor Operation, 7/24/05 through 8/23/05

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. 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 CO_2 and H_2O. Eliminating this additional coal reduces the syngas CO_2 and H_2O 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/CO_2 ratio will change due to the reduction in CO_2. 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 percent of the syngas flow rate from the gasifier and is available at 235ºF. The aeration steam flow rate is estimated to be 1.45 percent of the syngas flow rate from the gasifier and available at 660ºF.
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 CO_2, H_2O, and N_2 concentrations are reduced; the CO/CO_2 ratio change. Based on energy balance data, the heat loss for the PSDF Transport Gasifier is approximately 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 H_2O in the syngas by the amount the steam rate was reduced. The steam rate and the H_2O 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 #2, #3, and #4 all change the water gas shift equilibrium constant without changing the gasifier exit temperature. The commercial plant will operate at the PSDF gasifier exit temperature and hence have the same water gas shift equilibrium constant. The H_2O, CO_2, CO, and H_2 concentrations are then adjusted water gas shift equilibrium for the temperature of that particular operating period. The LHV could increase if H_2 and CO_2 are converted to H_2O and CO, since the LHV for CO is higher than H_2. The LHV will decrease if H_2O and CO are converted to H_2 and CO_2. 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 H_2 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 adjust 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 percent of the syngas nitrogen is removed in the oxygen blown projection, while only about 50 percent 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.

APPENDIX A6 FTIR HYDROCARBON RESULTS

Analyzer AI2530, a Fourier transform infrared (FTIR) analyzer, is used to analyze syngas for ammonia, H_2O, and several hydrocarbons. The FTIR concentrations are measured without any moisture removal and do not require any correction for H_2O. Due to syngas fouling of the instrument, the FTIR cannot be operated for more than a few minutes at time, and therefore the FTIR data is limited.

Figure A6-1 plots the FTIR ethane and FTIR ethylene concentrations and run time. The ethylene was higher than the ethane concentrations by between 100 to 200 ppm.

Figure A6-2 plots the FTIR propane and FTIR cyclopropane concentrations and run time. The propane concentrations were higher than the cyclopropane concentrations by about 2 to 25 ppm.

Figure A6-3 plots the FTIR benzene concentrations. Figure A6-4 plots the FTIR benzene concentrations against the cyclone exit temperature. As expected, higher gasifier temperatures crack more benzene and lead to lower benzene concentrations.

Figure A6-5 plots the FTIR naphthalene and FTIR phenanthrene concentrations. All of these concentrations are significantly less than the benzene concentrations. The phenanthrene concentrations are between 25 to 250 ppm higher than the naphthalene concentrations.

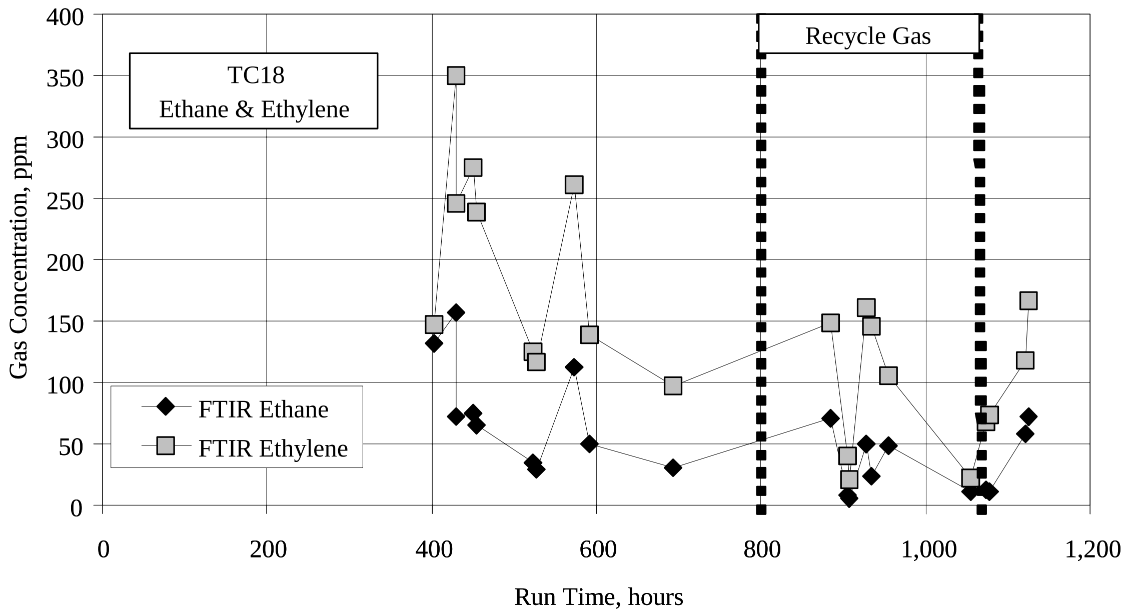

Figure A6-1 Ethane & Ethylene Concentrations

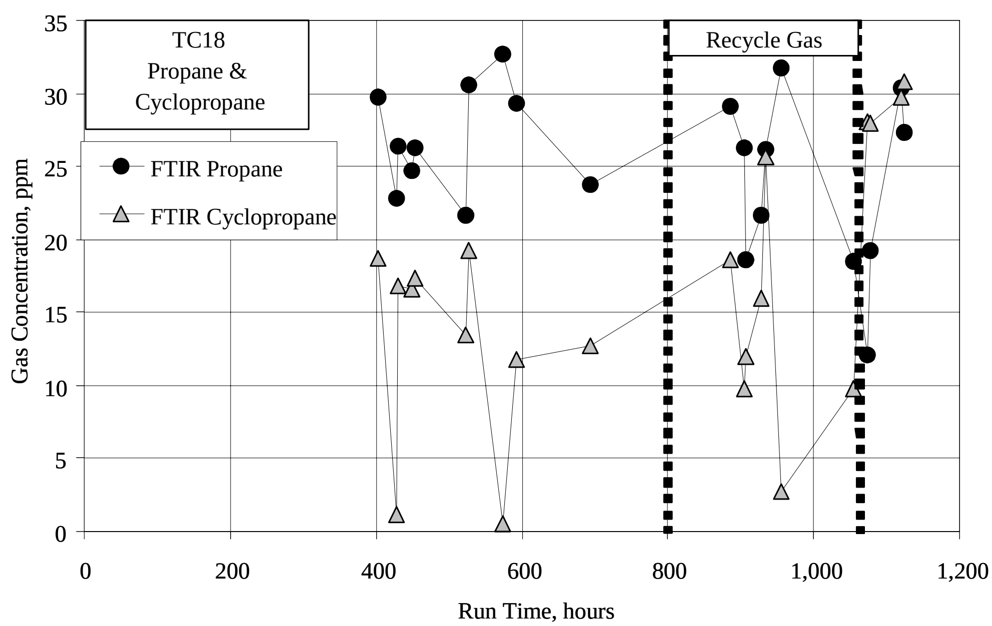

Figure A6-2 Propane & Cyclopropane Concentrations

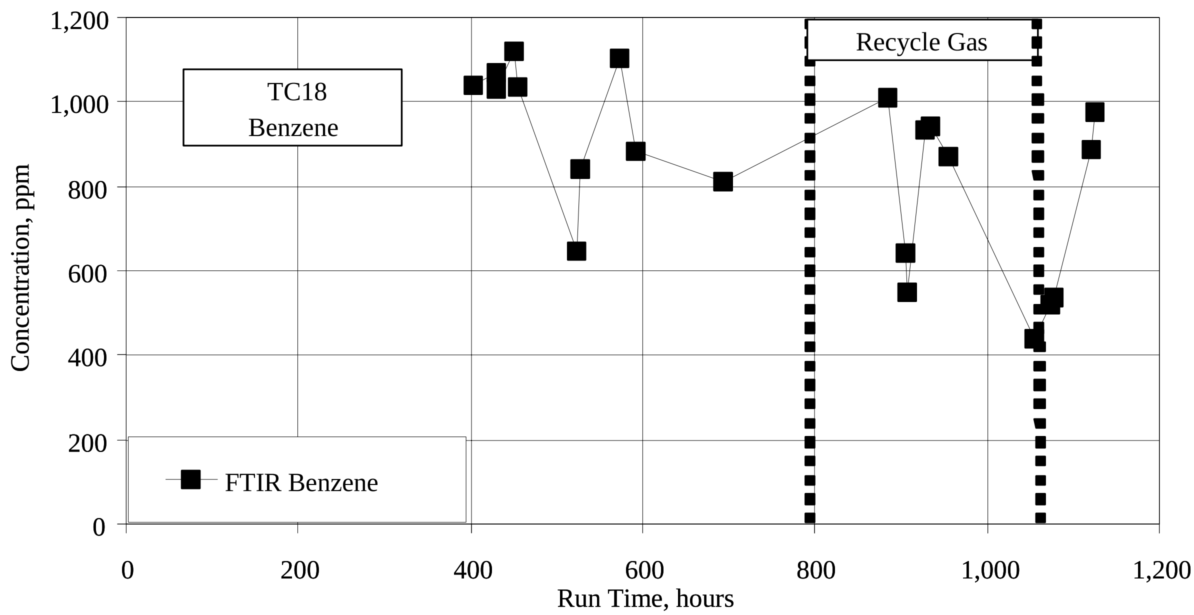

Figure A6-3 Benzene Concentrations

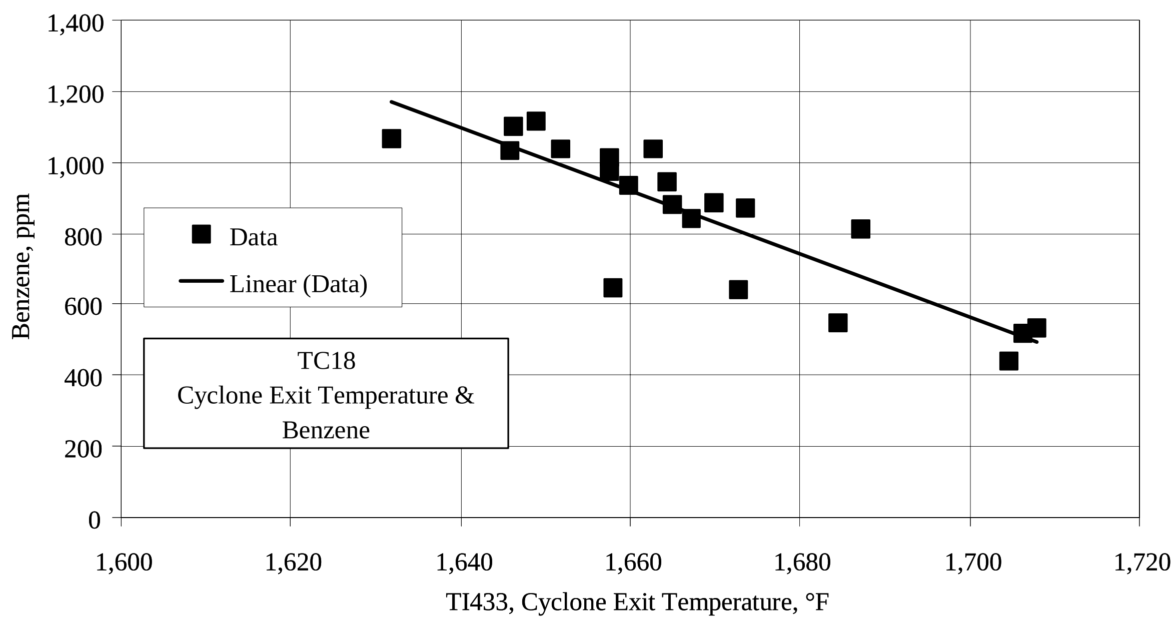

Figure A6-4 Cyclone Exit Temperature and Benzene Concentrations

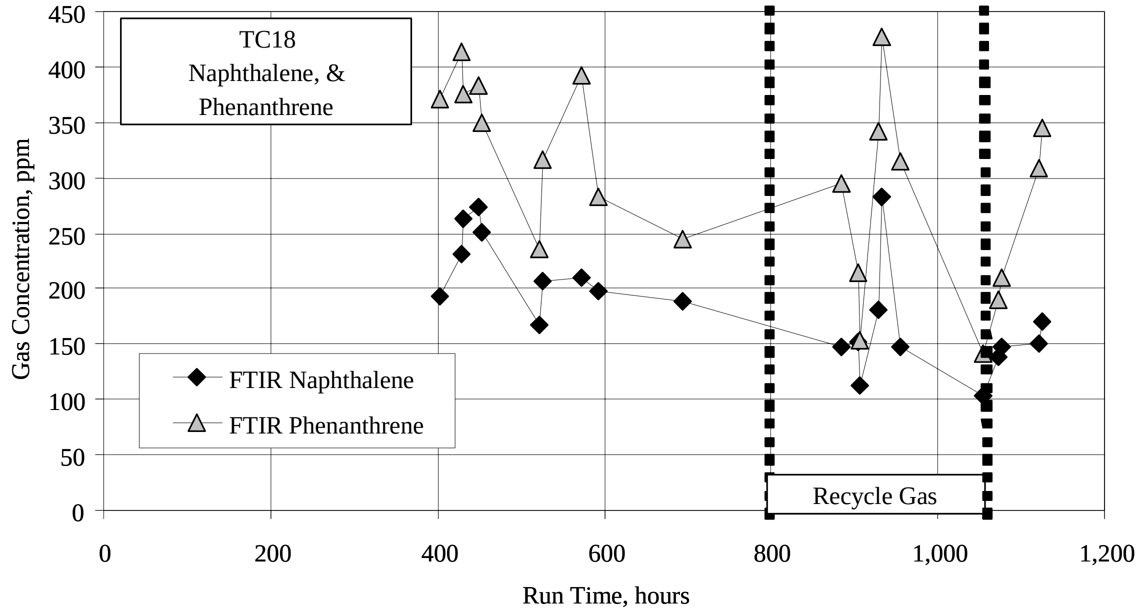

Figure A6-5 Naphthalene & Phenanthrene Concentrations