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ADVANCED DEVELOPMENT OF A PRESSURIZED ASH
AGGLOMERATING FLUIDIZED-BED COAL GASIFICATION SYSTEM

Topical Report

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Morgantown, West Virginia

By
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ADVANCED DEVELOPMENT OF A
PRESSURIZED ASH AGGLOMERATING FLUIDIZED-BED
COAL GASIFICATION SYSTEM

TOPICAL REPORT
PROCESS ANALYSIS

FY-1983

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ABSTRACT

KRW Energy Systems, Inc. is engaged in the continuing development of a pressurized, fluidized-bed gasification process at its Waltz Mill Site in Madison, Pennsylvania.

With the sponsorship of the U. S. Department of Energy, its precursor agencies, and the Gas Research Institute, Westinghouse Electric Corporation conducted this development from 1975 to March 1980 under Contract EF-77-C-01-1514, from March 1980 to January 1982 under Contract DE-AC21-80-ET14752, and from January 1982 to April 1984 under Contract DE-AC21-82MC19122. In April 1984, Contract DE-AC21-82MC19122 novated to KRW Energy Systems Inc., a new entity formed by Kellogg Rust Inc. and the Westinghouse Electric Corporation, which will continue the development at the Waltz Mill Site, formerly operated by Westinghouse Synthetic Fuels Division.

The overall objective of the program is to demonstrate the viability of the KRW process for the environmentally-acceptable production of low- and medium-Btu fuel gas from a variety of fossilized carbonaceous feedstocks and industrial fuels.

This report presents process analysis of the 24 ton-per-day Process Development Unit (PDU) operations and is a continuation of the process analysis work performed in 1980 and 1981 and reported in the 1982 final report, FE-14752-28. Included is work performed on FY-1983 PDU process data; gasification; char-ash separation; ash agglomeration; fines carryover, recycle, and consumption; deposit formation; materials; and environmental, health, and safety issues.

GLOSSARY

Agglomeration - growth of high ash particles by small partially melted ash particles attaching to each other until the resulting particle becomes too large to remain fluidized.

Analytical models - models using theoretical rather than empirical correlations to predict process behavior.

Bed carbon holdup - mass of solid carbon portion of fluidized bed.

Carbon utilization - fraction of total carbon fed to the gasifier which is gasified or combusted.

Carcinogens - cancer causing agents.

Char - material resulting from devolatilization of coal.

Char ash separation - process of classifying ash agglomerates from bed char by exploiting their difference of minimum fluidization velocity.

Coal - bituminous, subbituminous, or lignite carbonaceous fossilized material usually containing varying amounts of fixed carbon, volatile material, moisture, and ash.

Coke - material resulting from coal which has been through a coking process. Usually completely devolatilized.

Coke breeze - fine material carried off from coking process which has not been completely devolatilized but is characteristically low in reactivity.

Cold flow model - a fluidized bed which operates at ambient conditions but otherwise attempts to simulate hydrodynamic conditions of a fluidized coal gasifier.

Cold-wall - liquid cooled steel jacket, exposed to the product gas of the gasifier which cools entrained molten material impacting on its surface and thus inhibits deposit buildup.

Coupons - small samples of test material exposed to typical environment of coal gasifier off gas in order to test corrosion/erosion characteristics.

Dead-leg - portion of 90° turn of solids feed line which contains refractory.

Deposit control - means of avoiding buildup of molten material on the walls of vessels and spool pieces of a coal gasification process.

Devolatilization - process by which the volatile and moisture portions of coal are driven off leaving particles mostly made of solid carbon and ash.

Dipleg - section of cyclone where captured fines accumulate before they are transported back to the gasifier via lockhopper, non-mechanical valve, or solids injection device.

Dryer mill - device which crushes and dries coal feedstock to appropriate size and moisture content.

Elutriation - process by which fine material from a fluidized bed becomes entrained in the resulting off gas.

Feedhoppers - lockhoppers which feed fresh coal to the gasifier.

Feed line - pipe in which coal is pneumatically transported to the gasifier.

Feedstock - fresh feed material, i.e., bituminous, subbituminous, lignite, petroleum coke, coke breeze.

Fines - material which becomes entrained in the off gas of the gasifier and is typically smaller than 200 microns.

Fluid bed - section of fluidized bed in which the particles are being maintained between their minimum fluidization and terminal velocities.

Fluid bed density - mass per unit volume of a fluidized bed; includes particles and voids.

Freeboard - section of fluidized bed gasifier above the fluid bed in which entrained fines can become disengaged from the off gas.

Gasification - process where solid carbon is reacted with steam or carbon dioxide to form gaseous carbon monoxide and hydrogen.

Gasifier - device in which coal is partially combusted with oxygen and reacted with steam to form a gas mixture of CO, CO₂, H₂, CH₄, etc.

Grid plate - loosely defined as the section of a gasifier in which a conical transition from the ash annulus to a combustion zone is accompanied with a circular row of refractory-embedded nozzles which supply fluidizing gas.

Hardface refractory - refractory which when cast, results in a hard durable surface, but has low insulating ability.

Insulating refractory - refractory which when cast is relatively soft, but has enhanced insulating qualities. Used with hardface refractory in vessels requiring low heat loss but a durable internal surface.

Jet phenomena - occurrences evolving around the discharge of the oxidant tube into the fluidized bed.

L-valve - non-mechanical device by which fines accumulating in a standpipe are injected in a controlled fashion into the gasifier.

Lignite - a type of feedstock which is lower in grade than bituminous coal. Characteristically it has high moisture and volatile matter and reactivity compared to bituminous coal.

Lockhoppers - vessels in which coal is alternately charged then fed to the gasifier at system pressure.

Molten particle through-trap - a device which forces all elutriated material through a tortuous path of liquid-cooled surfaces in order to freeze any molten material before it passes further in the process.

Mutagens - a chemical agent which causes chromosome damage.

Non-mechanical valve - a device without any mechanically moving parts that is used to control the rate at which bulk solids are delivered from an elevated location (cyclone discharge), via gravity flow to a lower dilute phase receiver (fluidized bed gasifier), usually against an adverse pressure gradient.

Oxidant tube - generic term for the tri-coaxial central feed tube in which oxygen, steam, feedstock, and recycle gas are fed to the gasifier.

Plug flow - pattern of flow through a reactor in which no element of fluid overtakes or mixes with any other element ahead or behind. Lateral mixing may occur. The residence time in the reactor must be the same for all elements of fluid.

Priority pollutants - EPA list of particularly hazardous chemicals which must meet stringent standards before release to environment.

Process models - set of analytical results and experimental correlations which model solid, liquid, and gaseous phenomena occurring in the system.

Proximate analysis - analysis of coal or char consisting of volatile matter, fixed carbon, moisture, and ash in weight fractions.

Pyrolysis - method by which coke and charcoal are produced by heating. Volatile material is driven off leaving carbon and ash.

Quench scrubber - unit operation equipment in which hot and dirty off-gas from the gasifier is quickly cooled and particulate matter is separated from the gas by water spraying.

Radial feed - point at which solids are fed perpendicularly to the vertical surface of the gasifier.

Reactivity - rate at which a feedstock char reacts in a coal gasifier relative to coke breeze char.

Reactor - coal gasifier.

Recycle gas - portion of cleaned product gas which is compressed for use as process streams.

Scaleup - process of applying information gained from a small scale such as PDU, CFSF, and bench-scale to a larger scale gasifier and peripheral equipment.

Set point - reasonably steady conditions during a PDU test which are similar to conditions called for in a test plan.

Shock heating - process by which relatively cool feedstock particles are very suddenly brought up to high temperatures ($\sim 200^{\circ}\text{F}$) usually resulting in shattering of the particle.

Single-stage gasifier - gasification process in which devolatilization, gasification, combustion, and ash agglomeration occur in a single vessel.

Slagging - condition where heat is not sufficiently dissipated to avoid bulk ash melting or sintering.

Slip stream - a portion of a total product stream which is taken and used for a different purpose.

Solids injection device - device which can take accumulated solids from a lower pressure and inject those solids into a higher pressure.

Sparger - a ring from which a gas is equally distributed around the circumference.

Species - a particular type of chemical compound.

Specific gasification - rate of solid carbon gasification per unit mass of solid carbon holdup in the fluidized bed gasifier.

Spool - section of steel piping which is flanged on both ends.

Starwheel feeders - star-shaped wheel which has pockets in which material is placed. Solid feed rate is controlled by controlling the RPM of the starwheel.

Steady-state - portion of a set point in which various gasifier conditions remain steady according to predescribed criteria.

Subbituminous coal - a class of coal feedstock material which characteristically has less heating value per unit mass and more moisture than bituminous coal.

Thermal oxidizer - device in which all product gas and vent gases are thoroughly combusted.

Through-trap - cold wall device in which molten particles are selectively frozen in order to avoid wall deposition downstream.

Turndown - process by which the throughput of a gasification process is smoothly reduced in order to match the demand for its product gas.

Ultimate analysis - analysis of feedstock material which includes elemental carbon, oxygen, hydrogen, nitrogen, and sulfur.

Water-gas shift reaction - equilibrium reaction in which amounts of carbon monoxide and steam are in equilibrium with carbon dioxide and hydrogen. The associated equilibrium constant is a function of temperature.

Water injection - a means of avoiding wall deposition downstream of the gasifier by spraying liquid water into the gasifier freeboard, thus cooling the entire bulk off-gas and entrained solids.

Water quench - process of quickly cooling entire product off-gas with direct contact of water.

SECTION 1.0

INTRODUCTION

1.1 GENERAL

The ability to scale a fluidized-bed gasifier to large capacities is one attractive reason for the use of fluidized-bed technology. While empirical information is available and component models have been proposed (e.g. devolatilization, gasification, char combustion, jet phenomena, particle segregation, particle agglomeration), the understanding of phenomena occurring in coal gasification is limited. An improved understanding of key phenomena can result in process models that will provide a basis for effectively selecting design and operating criteria to meet specific application needs.

This report, part of a larger DOE/GRI program that includes experimental studies on a 24-ton per day process development unit (PDU), Figure 1-1, and laboratory experimental and analytical support studies, presents the status of work on process evaluation and the development of design criteria. The objectives for this task are:

- Assimilate PDU and laboratory test data to gain a better understanding of key process phenomena and equipment performance.
- Develop correlations or models to relate process phenomena to design and operating parameters that will permit commercial design to achieve a given performance.
- Interpret results and relate them to published coal gasification process information.
- Identify data needs required to develop integrated process models.
- Recommend PDU and laboratory tests to obtain data needs.

This work compiles PDU performance operability data. Performance data includes capacity (as a function of temperature, pressure, and coal characteristics); carbon loss in ash agglomerates (as a function of temperature, coal characteristics, and bed ash content); particle carryover (as a function of coal and velocity); cyclone performance; heat transfer; gas characteristics (water and sulfur concentrations); and solids characteristics. PDU operability data and hardware development includes oxidant tube design to avoid sintering, system control (e.g., ash annulus, bed level, bed temperature), and deposit formation and control.

Figure 1-1. KRW Process Development Unit (PDU)

The work carried out was directed to:

- Identify PDU data requirements based on analysis of available data base and model development needs.
- Implement procedures to obtain additional information from PDU data (e.g., ash concentrations as a function of particle size for char-ash separation and agglomeration analyses, and uncertainties in data).
- Develop models based on controlling phenomena that will be used in the continuing program to analyze PDU data--to permit a sound basis for scale-up--and will extend the prior work based on empirical correlations.
- Apply available performance models to PDU data to develop assessment of model prediction capability.

Specific areas investigated, program objectives, and current status are summarized in Table 1-1. These objectives are directly related to and are supported by the contract tasks of large-scale cold-flow modeling and the laboratory support work that includes experimental and analytical results on particle history, char-ash separation, jet phenomena, attrition, gasification kinetics, agglomeration and deposit formation/control. This document represents a status report on the areas where specific subtasks have been completed.

1.2 PROCESS DEVELOPMENT UNIT (PDU)

The advanced development of a pressurized, ash agglomerating fluidized-bed coal gasification system by KRW Energy Systems Inc. in Madison, Pennsylvania uses an oxygen-blown, nominal 24-ton per day process development unit (PDU) to generate medium-Btu gas from a variety of coal and lignite feedstocks. Originally conceived as a two-stage, air-blown, fluid-bed system for combined-cycle electric power generation, the PDU demonstrated successful operation using the stage 1 devolatilizer and the stage 2 gasifier on a variety of feedstock materials. When the technology was broadened to include oxygen instead of air to produce medium-Btu gas, several coal feedstocks were successfully processed directly in the gasifier without pretreatment in the devolatilizer.

During the subsequent 2 years of development, the PDU was operated as a single-stage unit using both oxygen-blown and air-blown modes to evaluate a variety of coal feedstocks. Goals for this second-generation gasifier technology were expanded to include production of medium-Btu gas for industrial use. Based on experimental breakthroughs in the process design, it was demonstrated that caking coals, highly reactive coals, and coals with low or high ash content could be processed successfully in a

TABLE 1-1

PROCESS ANALYSIS OVERVIEW

<u>Area</u>	<u>Objectives</u>	<u>Status</u>
PDU Process Data	● Identify data needs ● Develop format for data documentation ● Implement procedures to obtain data needs	● Data requirements identified based on performance and model needs ● Data format developed ● Procedures implemented to obtain data needs
Gasification	● Compare model for estimating reactor volume requirements and gas composition based on kinetic data to the PDU data base ● Develop basis for scaleup	● PDU bituminous coal test data analyzed and compared with model predictions ● PDU subbituminous and lignite test data analysis carried out during 1983
Char-Ash Separation	● Develop criteria for selecting operating conditions to achieve desired separation	● Additional PDU data reduction requirements identified and analyses implemented ● Analysis of initial results completed ● Analysis and laboratory separation tests completed using PDU materials from 1983 test programs
Ash Agglomeration	● Develop basis for estimating ash agglomeration rate	● Laboratory data on ash sintering behavior completed as part of lab support program ● Analysis of 1983 PDU data and lab data completed.
Fines Carryover, Recycle,	● Develop model for PDU entrainment	● PDU entrainment model developed and compared with PDU data

TABLE 1-1 (continued)

PROCESS ANALYSIS OVERVIEW

<u>Area</u>	<u>Objectives</u>	<u>Status</u>
Consumption	● Reconcile observed PDU cyclone efficiencies and cyclone performance prediction techniques ● Present design criteria for non-mechanical valves	● Cyclone performance evaluation and fines recycle performance assessment completed using 1983 test program data
Deposit Formation	● Develop understanding of deposit formation mechanisms ● Develop design criteria for deposit control	● Data collated on PDU deposit experience, deposit characteristics and control techniques
Environmental	● Project fate of major, minor, and trace elements throughout PDU system ● Compare model projections with PDU sampling results ● Correlate gas and solid composition data with PDU operating conditions	● Projections of chemical profiles were completed based on equilibrium calculations ● PDU data (1980-82) were reviewed and correlated with operating conditions
Materials	● Evaluation of materials of construction for fluid-bed gasifier systems ● Evaluation of coatings and specialty materials for corrosion/erosion prevention ● Assessment of PDU failure analyses and materials coupons test data	● Corrosion data from PDU vessel core samples analyzed and assessed ● Performance data on coal feed system, gasifier vessel, refractories, thermowells and quench system analyzed

single-stage gasification process. The single-stage configuration then became the prime design for the process in lieu of the two-stage fluidized-bed system. Concurrently, Gas Research Institute (GRI) became interested in the potential of this advanced process as a coal-to-synthetic natural gas (SNG) producing plant and, in a joint program with the Department of Energy, greater emphasis was placed on the development of the medium-Btu, oxygen-blown process. The program was also expanded to plan the inclusion of issues, such as heat recovery and related hardware systems outside the basic gasification module, which have yet to be implemented.

1.2.1 PDU Operation

Equipment and processing in the PDU can be broken into three functional areas for the production of a cleaned and cooled medium-Btu product gas, Figure 1-2, from a variety of feedstocks:

- Coal Preparation and Transport
- Gas Generation
- Gas Cleaning and Cooling

With its present state-of-the-art development, the PDU has demonstrated a broad flexibility which encompasses the use of:

- run-of-mine coals and lignites;
- fine-sized feedstocks; and,
- coals with different ash, sulfur, caking and reactivity characteristics.

In addition, although the PDU nominally operates at a 24 ton-per-day capacity, the process is projected to be efficient and economical for large-scale commercial plants because of its:

- Low steam and moderate oxygen use.
- High carbon utilization - low carbon in the ash and fines are consumed.
- High thermal efficiency - moderate operating temperatures with waste heat recovery.
- Component design and construction using readily available materials - no exotic or high technology materials.
- Environmental acceptability.

1.2.1.1 Coal Preparation and Transport

The coal preparation facility of the PDU, Figure 1-3, crushes the feed coal to a maximum size of 1/4 inch and dries it, if necessary, to reduce the moisture content. This is accomplished at the PDU by bulk, run-of-mine coal being dumped into a feed hopper (D-112) from which it is transported by a screw conveyor (T-100) and fed through a rotary feeder (T-102) to a hammermill (T-106) which has a velocity separator (C-136) to help control

Figure 1-2. Process Development Unit (PDU) Block Diagram

Figure 1-3. Process Development Unit (PDU) Simplified Schematic

coal size distribution. Drying the crushed coal is accomplished at the hammermill by hot (approximately 500°F) high-velocity combustion gases from an air heater (F-135) which sweep the coal to a cyclone (C-138) where the stream is separated. A screen sizer deposits the crushed coal into a screw conveyor (T-120) while the oversized material is passed through a bag filter (T-135) and recycled to the dryer mill. The other portion of the combustion gases with entrained fines is passed through a separate bag filter (T-113) where the fines residue is carried by a screw conveyor (T-134) to be deposited with the crushed coal from the other screw conveyor (T-120) into an unloading hopper (D-108). The separated combustion gas, consisting of carbon dioxide and water, is vented to the atmosphere from the bag filter.

Dried, crushed coal from the unloading hopper is transferred at a controlled rate through a relatively straightforward series of screw conveyors (T-161, T-207, T-104 and T-107), bucket elevators (T-103, T-105, and T-108), storage bins (D-101, D-102, and D-104), and belt weighers (T-121 and T-122) to the coal receiving lockhoppers (C-103A and C-109A) which cycle from atmospheric to system pressure. During normal operation, these receiving lockhoppers are alternately loaded with crushed coal and pressurized from ambient to approximately 230 psig with CO₂ from an outside source. Coal is transferred from the receiving lockhoppers to the feed lockhoppers at system pressure.

During startup, char is used from its storage bin (D-105) and associated bag filters (T-114 and T-115). Char is conveyed to the char receiving lockhopper (C-102A) using CO₂ through the conveying gas blower (K-106) as the pneumatic transport medium. Following the transfer, the char receiving lockhopper is pressurized to system pressure with CO₂; however, unlike the other two which are vented to the atmosphere, the char receiving lockhopper is vented to the thermal oxidizer. (The coal lockhoppers (C103 and C109) were originally vented to the thermal oxidizer also, but were changed to decrease depressurization time of the receiving lockhoppers.) All three lockhopper systems vent through sintered-metal leaf filters.

From this point through to the gasifier, the three feedstock transport systems operate identically. Feedstocks are transferred from the receiving lockhoppers to the feed lockhoppers (C-109B and C-103B for coal; C-102B for char) through manual or automatically controlled valves. Char feed through the C-102 lockhopper system is used as a startup material until operating temperatures are reached. Once system operating temperatures and pressures are stabilized, the coal feedstock is gradually introduced as the char feed is decreased until a steady-state condition using only coal is achieved.

Feedstock from the feed lockhoppers is controlled through starwheel feeders (T-124 and T-125 for coal, and T-132 for char) which, in turn, are controlled by a signal from a level indicating controller on the gasifier. Heated recycle gas is used as the coal transport medium.

1.2.1.2 Gasifier Operation

When the feedstock is injected into the gasifier (C-115), through the feed tube where it is mixed with oxygen, Figure 1-4, a portion of this coal combusts providing the heat necessary to devolatilize the remainder of the incoming coal, forming a char which reacts with steam to form hydrogen and carbon monoxide. As the bed of char circulates through the jet formed at the end of the central feed tube, the carbon in the char reacts with oxygen and steam and is consumed by combustion and gasification leaving particles that are rich in ash. The unique fluid dynamics of the gasifier allow the ash-rich particles to agglomerate with each other in larger spherical or raspberry type particles. These larger particles become dense enough to defluidize (precipitate) from the bed and collect in the ash annulus. These dry ash agglomerates are continuously removed from the gasifier through a rotary feeder and are transported to the ash handling system.

The performance of the agglomerator section in the process is tied to two distinct phenomena: ash agglomeration and char-ash separation. The degree of ash agglomeration for any coal is influenced by: (1) ash content of the fresh coal -- the more ash present in the coal translates to greater surface area of minerals exposed to the high temperatures in the combustion zone of the gasifier; and (2) melting properties of the coal ash-- the lower the threshold eutectic formation temperature or, more correctly, the closer the particle temperature is to the eutectic softening temperature, the easier it is to produce ash agglomerates.

Two factors which are to be considered in evaluating char-ash separation effectiveness are: (1) the tendency of the feedstock ash to form larger and/or denser agglomerates, that are separable from the smaller and/or less dense char particles that are rich in carbon; and (2) the velocity used in the char-ash separator which impacts the separation efficiency.

During operation, recycle gas flow to the char-ash separation zone is maintained at a sufficient rate to cool the ash particles prior to extraction from the gasifier and to minimally fluidize the ash particles in this region. The char-ash interface level is maintained on temperature control. A thermocouple in the annulus region is tied into the logic of the ash withdrawal rotary feeder. As the char-ash interface level rises, the temperature declines, and the rotary feeder responds by increasing its speed to remove more of the ash agglomerates. As the char-ash interface level declines, the temperature rises and the rotary feeder decreases its speed to maintain the inventory.

Raw product gas from the top of the gasifier passes through the refractory-lined primary and secondary cyclones (C-110 and C-120, respectively) where the entrained fines particles are recovered and pneumatically recycled to the gasifier. Carbon dioxide is normally used as the transport medium, although a provision is also made to use cleaned recycle gas.

Figure 1-4. Single-Stage Coal Gasifier

Ash withdrawal rate from the ash annulus is controlled by a large starwheel feeder. The ash is directed to either of two ash disposal lockhoppers (C-117A or C-117B) at system pressure. As one lockhopper is filling with ash for disposal, the other is isolated, depressurized, and emptied into a tote bin by gravity and CO₂ purging. System pressure is relieved through the ash vent filters (V-123 and V-124) to either the atmosphere or the thermal oxidizer.

From the tote bin, the ash is placed in the ash storage bin (D-106) by a bucket elevator (T-226) for disposal by truck. Carbon loss in the ash agglomerates is relatively small. Ash discharged from the gasifier typically has an ash content of 85 percent or higher. Other carbon losses are minimal, since the entrained fines from the cyclones are recycled to the gasifier and are consumed.

Leachability characteristics of the ash discharged from PDU tests were measured, assessed, and found to be in compliance with the Environmental Protection Agency Resource Conservation and Recovery Act (RCRA) regulations. General conclusions of this investigation were that: (1) residues from the KRW gasification process are non-hazardous; and (2) disposal of the ash agglomerates from the gasifier will cause less negative environmental impact than the disposal of coal ash residue from conventional power plants.

1.2.1.3. Gas Cleaning and Cooling

The product gas stream from the secondary cyclone is water quenched and scrubbed in a quench scrubber (C-111). An optional secondary stream is available for slipstream heat recovery tests from which it rejoins the primary stream downstream from the gas scrubbers. From the quench scrubber, where carryover fines are removed, the gas stream is cooled by a contact cooler (C-113) containing Pall rings, and is split: part goes to the thermal oxidizer (F-103) to be burned prior to exhausting to the atmosphere and the other part (being further cooled in gas cooling tower C-122, also containing Pall rings) goes for use as recycle gas to the gasifier for pneumatic transport of solids and fluidization of the ash annulus.

Fines from the quench scrubber collect in the fines settler (C-112) and are flushed to the tar water flash tank (Z-107), while fines from the contact cooler and the gas cooling tower are flushed directly to the tar water flash tank which empties into the separator pit where the collected solid residue is removed for contract disposal and the water is pumped to the makeup water system. Circulating water for the quench scrubber and the two cooling towers is also vented to the tar water flash tank and hence to the separator pit. The gas atmosphere in the separator pit is exhausted to the top of the stack of the thermal oxidizer by the separator pit exhaust fan (K-109).

SECTION 2.0

GASIFIER PERFORMANCE

This section encompasses the performance characteristics of the gasifier reactor; that is, correlation of reactor design parameters obtained in PDU tests for unit process models, which, in conjunction with computer and analytical models, are used in the scaleup design of the gasifier.

The KRW (PDU) ash agglomerating gasifier design has evolved to the point where operational problems due to clinker formation, ash withdrawal blockage, and oxidant tube sinter formations have been virtually eliminated. It has also been successfully converted from a 130 psig to a 230 psig oxygen-blown mode. After a series of 230 psig oxygen-blown tests, a comparison of gasifier behavior at both pressures was made at similar steam to coal ratios and temperatures. Areas found to be impacted by pressure were capacity, gas make, and methane production. Another area studied was turndown performance at fixed bed volume and variable temperature and at fixed temperature and variable bed volume.

2.1 GASIFICATION PERFORMANCE ANALYSIS

This section analyzes PDU performance in terms of gasification rate and product gas composition for selected tests and predicts performance using a gasification model and kinetic data developed on a thermogravimetric analyzer (TGA). It compares the predicted values with those measured in the PDU and analyzes the impact of measurement uncertainties on both the actual and the predicted PDU performance.

PDU gasification performance is measured in terms of the bed carbon holdup maintained to achieve the desired coal processing rate at the operating conditions for a given feedstock and product gas composition. The feedstock is characterized for reactivity on the TGA. Data on coal reactivity and the detailed kinetic data on carbon gasification reactions developed in a fluidized bed are used in a gasification model to predict the bed volume (carbon holdup) and product gas composition. Details of the gasification model are given in Section 2.1.1.1 and coal reactivity characterization on the TGA in Section 2.1.1.2.

2.1.1 Analysis and Prediction of PDU Performance with Pittsburgh Seam Coal

Using Pittsburgh seam coal as the feedstock, data from tests TP-018-2 through TP-028-1 (for which steady-state heat and mass balance data are available) were analyzed to evaluate the PDU performance. Freeboard temperature, the measured bed volume and specific gasification rate, and the estimated steam-to-carbon and recycle-gas-to-carbon ratios for these tests are given in Table 2-1.

TABLE 2-1

PDU PERFORMANCE WITH PITTSBURGH SEAM COAL AS THE FEEDSTOCK

PDU Test TP-0	Freeboard(1) Temp., °F	Actual Bed Vol.(2) ft ³	Steam/Carbon(3) mole/mole	R.G./Carbon(4) mole/mole	Specific Gasification(5) rate, hr ⁻¹
18-2(5)	1743	31.0	3.89	3.68	1.11
18-2(6)	1900	45.3	2.16	2.11	2.35
22-1(1)	1836	34.4	2.13	1.47	2.00
23-4(1A)	1821	30.2	2.01	0.93	1.55
23-4(1B)	1821	30.5	2.98	0.68	2.51
23-4(2A)	1786	30.9	2.63	1.27	1.45
23-4(2B)	1779	32.6	2.97	1.56	1.44
27-2(2B1)	1835	38.0	1.45	1.50	1.36
27-2(2B2)	1837	30.8	1.63	1.69	1.56
27-2(2B3)	1820	31.3	1.58	1.63	1.29
27-3(2A)	1858	31.6	1.68	2.55	1.62
27-3(2B)	1854	32.2	1.43	2.19	1.49
27-3(2C)	1850	33.3	1.46	2.15	1.52
27-3(2D)	1850	33.0	1.64	2.17	1.73
27-3(2E)	1855	33.6	1.89	2.42	1.88
27-3(2A1)	1838	28.7	1.62	1.93	1.46
28-1(1A)	1886	35.5	1.47	1.59	2.11
28-1(1B)	1890	35.4	1.37	2.02	1.82

1) Section of gasifier above the top surface of fluidized bed.

2) Volume of fluidized bed between the oxidant tube tip and the top of the bed.

3) Input steam fed to gasifier to solid carbon gasified.

4) Input recycle gas fed to gasifier to solid carbon gasified.

5) Solid carbon gasification rate to unit mass carbon hold up in fluidized bed.

The following assumptions were made in the analysis and prediction of PDU performance:

- Moisture-free hydrogen in coal reacts first to form methane (CH_4) hydrogen sulfide (H_2S) and ammonia (NH_3);
- Moisture-free hydrogen remaining will react with the moisture-free oxygen from coal or oxygen supplied to the gasifier to form water. This water formed is accounted for in the total steam supplied to the gasification section;
- If there is free oxygen from coal remaining, it is added to the amount of oxygen supplied to the gasifier; and,
- Volatile matter released from coal and the combustible components of the recycle gas react with the available oxygen. If any oxygen is still available, it will react with solid carbon from char. In most cases, no oxygen will be left for the combustion of solid carbon.

The bed carbon holdup (bed volume) required to process a given coal feedrate depends upon the rate at which carbon is gasified, the operating conditions, and the reactivity of the feedstock. Gasifier performance may be measured in terms of the percentage of fixed carbon gasified in addition to the other performance variables (carbon conversion, thermal efficiency, steam utilization, etc.) since the conversion of volatile carbon to gaseous form is instantaneous relative to carbon gasification. Combustion of solid carbon can be expected to be negligible since the combustion of carbon monoxide and hydrogen present in the recycle gas is two orders of magnitude faster than solid carbon combustion. Specific gasification rate was plotted as a function of freeboard temperature for low and high steam flow rates in Figure 2-1. The substantial effect of temperature on gasification rate for both low and high steam flow rates is visible in the figure as is the effect of steam/carbon ratio on the specific gasification rate.

Specific gasification rate is defined as the gasification rate of carbon per unit amount of carbon present in bed. Gasification rate was calculated from solid flow rates (coal feedrate, carbon losses through discharge of ash agglomerates, and fines escaping the cyclone) and proximate analysis of coal.

Gasification rate can also be computed on the basis of flow rates and composition of all inlet gas streams to the gasifier, product gas flow rate and composition, and the volatile matter in coal processed in the gasifier. This calculation is more involved and the effect of measurement errors on the calculated gasification rate is higher in this case since both solid and gas flow rates are required. Calculation of gasification rate based on solid flow rates only is expected to be more accurate than the other method.

Figure 2-1. PDU Performance With Pittsburgh Seam Coal

Steam and recycle gas flow to the gasification zone were computed by taking into account the moisture present in coal, steam formed, and recycle gas combusted in the combustion zone. The steam/carbon ratio is computed from the total steam flow rate to the gasification zone and the amount of carbon gasified rather than carbon feedrate to the gasifier.

Specific gasification rate was plotted as a function of freeboard temperature for low and high steam flow rates in Figure 2-1. The substantial effect of temperature on gasification rate for both low and high steam flow rates is visible in the figure as is the effect of steam/carbon ratio on the specific gasification rate.

Effect of steam/carbon ratio on specific gasification rate is plotted in Figure 2-2 for four narrow temperature ranges. The increase in specific gasification rate with an increase in steam/carbon ratio is substantial at all temperature levels.

Slopes of the lines drawn through the data at the three lower temperature levels, from a regression analysis, are nearly the same. The intercepts on these plots, i.e., the specific gasification rates when steam flow rate is zero, represent carbon gasification due to C-CO₂ reaction. An increase in the value of the intercept with an increase in temperature can be noted. Data presented here are too limited to extract such information.

The extent to which the water-gas shift reaction approaches equilibrium in the product gas is examined for the steady-state periods studied here. Equilibrium constant of the water-gas shift reaction, K_E , is defined as

$$K_E = Y_{CO_2} \cdot Y_{H_2} / Y_{CO} \cdot Y_{H_2O}$$

where Y is yield (volume fraction).

Measured values of this ratio are plotted as a function of temperature in Figure 2-3. Experimental values deviate substantially from the theoretical relation, indicating that the water-gas shift reaction did not reach equilibrium in the product gas. Note that small errors in the product gas composition result in an appreciable error in the K_E value. Further analysis is planned to correlate approach to shift equilibrium with gasifier operating conditions and feedstock reactivity.

2.1.1.1 Gasification Model

The objective of the modeling study is to estimate the bed volume (carbon holdup) required for a given coal processing rate based on the gasification rate of solid carbon, and to calculate the product gas composition and the

Figure 2-2. Effect of Steam/Carbon Ratio on PDU Performance

Figure 2-3. Experimental Shift Equilibrium Values

oxygen requirement of the gasifier. Important variables affecting the gasification rate are the bed temperature, steam flow rate, gas composition, and bed carbon holdup. Important reactions in the gasifier are:

Reaction rate can be expected to vary with bed height since the gas composition varies with bed height. The gasification rate has been estimated by dividing the fluidized bed into various zones and by using the rate equations developed for C-H₂O and C-CO₂ reactions from laboratory studies. Assumptions made in the study are:

- The methanation reaction ($C + 2H_2 \rightleftharpoons CH_4$) is at pseudoequilibrium, and the equilibrium constant is based on the PDU data as a function of temperature for different coals;
- The solids are at constant temperature throughout the bed which implies perfect mixing of solids;

- The gases are in plug flow--the composition of the gas remains unchanged in any zone and is the same as that at the outlet of the zone;
- The reaction rate is independent of the particle size-- this assumption has been confirmed for highly reactive coals;
- There are no radial gradients in the bed temperature or gas composition;
- Nitrogen, sulfur, and the required hydrogen from coal are released as ammonia and hydrogen sulfide during devolatilization;
- Part of the hydrogen present in coal is released as methane during devolatilization to satisfy the pseudomethanation equilibrium, and any excess hydrogen from coal is combusted to form water; and,
- The water-gas shift reaction is at equilibrium throughout the reactor.

2.1.1.1.1 Gasification Model Description

Considering a differential section of the bed, of height dx , and cross-sectional area A , and integrating over the bed, the carbon gasification rate in a fluidized bed can be represented by

$$F_1 = A\rho_B \int_0^X r_1 dx \quad (2-6)$$

$$F_2 = A\rho_B \int_0^X r_2 dx \quad (2-7)$$

where ρ_B is the fluid bed density, and F_1 and F_2 are the gasification rates due to reactions (2-1 and 2-2), respectively. Dividing the entire bed into n number of zones of height Δx , Equations (2-6 and 2-7) can be written as:

$$F_1 = A\rho_B \sum_{1}^n r_1 \Delta x \quad (2-8)$$

$$F_2 = A\rho_B \sum_1^n r_2 \Delta x \quad (2-9)$$

By replacing r_1 and r_2 with the equations developed from experimental studies, F_1 and F_2 can be expressed by the following equations:

$$F_1 = CM_n K_1 \sum [1/(1 + P_{CO}/K_1 P_{CO_2})] \quad (2-10)$$

$$F_2 = CM_n K_2 \sum [1/(1 + P_{H_2}/K_2 P_{H_2O})] \quad (2-11)$$

where M_n equals $A\rho_B \Delta x$ and represents the number of moles of carbon in each section of the bed, and C represents the reactivity of coal with reference to coke breeze.

The model presented here is admittedly an idealistic one based on plug flow of gases and uniform solid temperature. Gas flow characteristics, in reality, are between plug flow and backmix flow. From experimental data on the radial and axial profiles of the gas composition in the bed, we should be able to modify the model to take into account the actual gas flow patterns in the gasifier. Model representation is shown in Figure 2-4.

The water-gas shift reaction was assumed to be at local equilibrium (i.e., at equilibrium throughout the bed). Contribution of this reaction to the change in gas composition is calculated by forcing the gas composition in each zone to be at equilibrium. Equilibrium constant for the water-gas shift reaction, K_E , is defined and estimated as follows:

$$K = M_{CO_2} \cdot M_{H_2} / M_{CO} \cdot M_{H_2O} \quad (2-12)$$

$$\ln K = (11,321 - 31.08T + 3T \ln T - 2.08 \times 10^{-4} T^2 - 91,500/T) / RT \quad (2-13)$$

T = absolute temperature (K)

R = gas constant, cal/(gram mole) (K)

INPUT

BED TEMPERATURE
PRESSURE
ELUTRIATION RATE/COAL
CARBON LOSSES
COAL FEED RATE

OUTPUT

PRODUCT GAS RATE
AND COMPOSITION
BED VOLUME

Figure 2-4. Gasification Model Representation

If x moles of CO react with H₂O in a given zone for the water-gas shift reaction to reach equilibrium, then x is given by

$$X = (-B + \sqrt{B^2 - 4AC})/2A \quad (2-14)$$

where $A = 1 - K$

$$B = M_{CO_2} + M_{H_2} + K (M_{CO} + M_{H_2O})$$

$$C = M_{CO_2} \cdot M_{H_2} - K \cdot M_{CO} \cdot M_{H_2O}$$

Devolatilization and combustion processes are so rapid compared to the gasification process that they are essentially complete prior to gasification. Combustible components of recycle gas and volatile matter could be expected to react with oxygen preferentially to solid carbon since the gaseous combustion reactions are faster than the combustion of solid carbon by two orders of magnitude. The amount of volatile matter evolved is assumed to be the same as that corresponding to the proximate analysis reported as per ASTM procedure. It may not be the same, however, as pressure was shown to significantly affect the release of volatile matter. Combustion of solid carbon is assumed to yield CO₂ only. Any change in this assumption or in the assumption of the combustion of recycle gas and volatile matter is not expected to cause any significant effect on the prediction of bed volume.

2.1.1.1.2 Method of Computation

Bed carbon holdup is estimated from the given bed diameter, bed density, and an assumed bed height. A flow diagram for the calculations is shown in Figure 2-5. The bed is divided into a number of small zones (usually about 100 for the PDU). Gasification rate and the gas composition in each zone are estimated from the rate equations. Zones containing the same amount of carbon were selected to keep the numerical integration simple. If necessary, zones of variable carbon content can easily be incorporated in the program. Composition of the gas entering the gasification zone is estimated on the basis of the given steam/coal ratio, recycle gas/coal ratio, recycle gas composition, and assumptions on devolatilization and combustion. This gas composition is used to start the calculations on the first zone. Char gasification rate in the first zone determines the composition of the gas entering the second zone. After carrying out the calculations successively on each zone of the bed, the char gasification rate in the gasifier can be estimated.

Figure 2-5. Computation Scheme Flow Diagram

Estimation of the carbon gasification rate for the measured coal processing rate is given by the following equation:

$$F_g = \text{fixed carbon in coal-carbon losses from cyclone} \\ \text{fines not recycled and ash discharge} \quad (2-15)$$

From the difference between the estimated and the measured carbon gasification rates, a new bed carbon holdup can be assumed. Kinetic and material and energy balance calculations are iteratively performed to match the measured and estimated gasification rates in order to predict the required bed volume and the product gas composition and rate.

Input Parameters

Freeboard temperature
Coal reactivity
Coal composition (proximate and ultimate analyses, and heating value)
Coal and fines feed rate
Steam/coal ratio
Recycle gas/coal ratio
Elutriation/coal ratio
Carbon losses/coal ratio
Temperatures of all gas and solid streams

Output Parameters

Bed volume (carbon holdup) (Bed carbon holdup gives bed volume based on bed density and bed ash content.)
Product gas composition
Product gas flow rate
Oxygen flow rate

In view of the problems and uncertainties involved in measuring the bed temperature in the PDU it was decided to use freeboard temperature as a reasonable approximation of the bed temperature. Typically, freeboard temperature is measured at a distance ranging from a few inches to about 3 feet from the bed surface. It is generally higher than bed temperatures measured at some of the locations and lower than those measured at other locations, all locations being well above the combustion region. It is believed to represent an average bed temperature reasonably well. Steep temperature gradients exist in the jet region, but the volume occupied by the jet is relatively small compared to the total bed volume and is certainly small compared to the uncertainty introduced by measurement errors. If the temperature profiles of the bed are completely mapped, the gasification model can be expanded to take into account the temperature gradients present in bed.

2.1.1.2 Kinetic Studies

Reactivity of coal samples were studied in a steam-hydrogen-nitrogen ($H_2O-H_2-N_2$) atmosphere. The reference temperature and the gas composition for characterizing coal reactivity were based on expected operating conditions in the PDU and in commercial-size gasifiers. Literature and bench-scale data indicate that the $C-H_2O$ and $C-CO_2$ reactions are important in carbon gasification and that the $C-H_2$ reaction forming methane reaches pseudoequilibrium. Relative importance of the two reactions in a gasifier depends upon the type of coal and the operating conditions (e.g., steam/coal ratio). The $C-H_2O$ reaction was chosen to characterize eastern and midwestern coals. The $C-H_2O$ reaction accounts for a major part of the carbon conversion in the gasifier for the eastern coals, and these were processed initially in the PDU. For highly reactive coals, the $C-CO_2$ reaction may also be important.

Reactivity of coals changes with the extent of carbon conversion because of differences in the reaction behaviors of several macerals present and because of the changes taking place in the pore structure. A coal particle fed to the gasifier is converted to char and then varies in reaction behavior with carbon conversion until the desired carbon conversion (90 percent) is reached. (For all practical purposes, this can be taken as 100 percent). This is true whether the bed is operated at 25 percent or at 95 percent ash content, since in both cases very high carbon conversion is desired and can be achieved. Reaction rate, therefore, is integrated from 0 to 100 percent to calculate the average reaction rate at the reference temperature and gas composition.

Average relative reactivity of coals/chars is defined as the ratio of the average reaction rate of coal/char to the average reaction rate of coke breeze. Both fluidized bed studies and TGA data determined the ratio of the average reaction rate to the initial rate for coke breeze to be about 0.5. The rate equations developed from fluid-bed studies and the average relative reactivities using the relation that follows determined the reaction rates for coals/chars as a function of temperature and gas composition:

Determination of Average Rate Constant of Coal at T

Carbon - Steam Reaction

$$\begin{aligned} \text{Average relative reactivity of coal at T, } REAC_T &= \frac{\text{Average rate of coal at T}}{\text{Average rate of coke breeze at } 1800^\circ\text{F}} \quad (2-16) \\ &= \frac{\text{Average rate constant of coal at T}}{\text{Average rate constant of coke at } 1800^\circ\text{F}} \times \frac{1 + R'/K_2 \text{ at } 1800^\circ\text{F}}{1 + R'/K_2 \text{ at T}} \end{aligned}$$

where $R' = p_{H_2}/p_{H_2O}$

$$\text{Reactivity Factor for coal at T} = \frac{\text{Average rate constant of coal at T}}{\text{Average rate constant of coke at } 1800^{\circ}\text{F}}$$

$$= \text{REAC}_T \times \frac{(1 + R'/K_2) \text{ at T}}{(1 + R'/K_2) \text{ at } 1800^{\circ}\text{F}}$$

REAC_T is determined at $R' = 0.5$ on the TGA

$$\text{Reactivity Factor for coal at T} = \text{REAC}_T \times \frac{1 + 0.5/K_2}{1 + 0.5/0.084}$$

$$= \text{REAC}_T \times (1 + 0.5/K_2)/6.95$$

$$= \frac{r_{\text{av}} \text{ of coal at T}}{r_{\text{av}} \text{ of coke at } 1800} \times \frac{r_{\text{av}} \text{ of coal at } 1800}{r_{\text{av}} \text{ of coal at } 1800} \times (1 + 0.5/K_2)/6.95$$

$$= \text{REAC}_{1800} \times \frac{r_{\text{av}} \text{ of coal at T}}{r_{\text{av}} \text{ of coal at } 1800} \times (1 + 0.5/K_2)/6.95$$

Express the variation of the average rate of coal with temperature as

$$r_{\text{av}} = A \exp(-E/RT)$$

T_R = reference temperature, 1255K (1800°F)

E = activation energy, cal/gmole

$$\frac{r_{\text{av}} \text{ at T}}{r_{\text{av}} \text{ at } T_R} = \exp\left[\frac{E}{R} \left(\frac{1}{T_R} - \frac{1}{T}\right)\right]$$

$$\text{Reactivity Factor at T} = \text{REAC}_{1800} \times \exp\left[\frac{E}{R} \left(\frac{1}{1255} - \frac{1}{T}\right)\right] \times \frac{(1 + 0.5/K_2)}{6.95}$$

$$\text{Average rate constant for coal at T} = \text{Average rate constant of coke at } 1800^{\circ}\text{F} \times \text{Reactivity Factor at T}$$

$$= 0.3 \times 10^{-2} \times 6.95 \times 60 \times \text{Reactivity Factor at T}$$

$$= 1.25 \times \text{Reactivity Factor at T}$$

$$\text{Average rate constant for coal at T} = 1.25 \times \text{REAC}_{1800} \times \exp\left[\frac{E}{R} \left(\frac{1}{1255} - \frac{1}{T}\right)\right] \times \frac{(1 + 0.5/K_2)}{6.95}$$

$$= 0.18 \times \text{REAC}_{1800} \times \exp\left[\frac{E}{R} \left(\frac{1}{1255} - \frac{1}{T}\right)\right] \times (1 + 0.5/K_2) \text{ hr}^{-1} \quad (2-17)$$

A similar expression can be derived for the carbon - CO₂ reaction.

Carbon - CO₂ Reaction

$$\begin{aligned}
 \text{Average rate constant of coal at } T &= \text{Average rate constant of coke at } 1800^\circ\text{F} \times \text{Reactivity Factor at } T \\
 &= 0.1 \times 10^{-2} \times 5.292 \times 60 \times \text{REAC}_{1800} \times \exp\left[\frac{E}{R} \left(\frac{1}{1255} - \frac{1}{T}\right)\right] \\
 &\quad \times \frac{(1 + 0.5/K_1)}{(1 + 0.5/0.1165)} \\
 &= 0.06 \times \text{REAC}_{1800} \times \exp\left[\frac{E}{R} \left(\frac{1}{1255} - \frac{1}{T}\right)\right] \times (1 + 0.5/K_1) \text{ hr}^{-1} \quad (2-18)
 \end{aligned}$$

In the above analysis,

1. It is assumed that the same activation energy is applicable to both C-H₂O and C-CO₂ reactions; and,
2. That the same reactivity applies to both reactions in the absence of information.

The following rate equations, developed by Westinghouse, were used to calculate the average rates of C-CO₂ and C-H₂O reactions for coke breeze, respectively:

$$r_1 = k_1 / (1 + P_{\text{CO}}/K_1 P_{\text{CO}_2}) \quad (2-19)$$

$$r_2 = k_2 / (1 + P_{\text{H}_2}/K_2 P_{\text{H}_2\text{O}}) \quad (2-20)$$

where the rate constants k_1 and k_2 are given by Equations (2-17) and (2-18) and the equilibrium constants K_1 and K_2 are given by

$$K_1 = 5.86 \times 10^3 \times \exp(-27,000/RT) \quad (2-21)$$

$$K_2 = 2.25 \times 10^6 \times \exp(-42,600/RT) \quad (2-22)$$

2.1.1.2.1 Description of the Thermogravimetric Analyzer

The apparatus consists of a Du Pont 950 thermobalance and a 900 console, a data acquisition system controlled by a Mostek microprocessor, and a Wang 2200 computer. Samples are placed in a platinum mesh screen shaped into the form of a basket and suspended from the balance arm. A schematic of the TGA and the data acquisition system is given in Figure 2-6; a photograph of the TGA is shown in Figure 2-7.

2.1.1.2.2 Thermogravimetric Analysis (TGA) Methodology

A sample of coal/char of about 10 to 20 mg is placed in a platinum mesh basket, heated slowly (20°C/min) to the operating temperature in a mixture of nitrogen and hydrogen, and held at this temperature until the sample weight remains constant. Loss in weight during this period represents the total moisture and volatile matter present in the sample. A mixture of steam, nitrogen, and hydrogen is then introduced over the sample to conduct gasification, and the sample weight is recorded as a function of time. When the gasification rate becomes very slow, gasification is terminated, and air is introduced over the sample to combust the remaining carbon in order to determine the percentage of carbon present in the sample used. The data on weight loss as a function of time are converted to reaction rate as a function of carbon conversion.

Steam was generated by using a capillary tube and a preheated mixture of nitrogen and hydrogen, as shown in the schematic of the TGA. Reactant gas flow rates were measured by flow meters.

Data saved by the microprocessor were reduced by the Wang computer. Data points selected at 1 percent carbon conversion intervals were reduced to calculate the reaction rate as a function of carbon conversion. Reaction rate at any given time is defined as the weight of carbon reacted per minute per unit weight of carbon present at the time in the sample. The reaction rate is integrated over the carbon conversion range of 0 to 100 percent to calculate an average reaction rate. Reaction rate at 100 percent carbon conversion is taken as zero, and the rate is assumed to vary linearly with carbon conversion from the last data point to 100 percent carbon conversion.

Figure 2-6. Schematic Diagram of the TGA System for Coal Characterization

Figure 2-7. Thermogravimetric Analyzer for Coal Reactivity Characterization

Summary of Test Conditions

Sample size	-1.0 + 0.25 mm
Sample weight	10 to 20 mg
Temperature	982°C
Total gas flow rate	300 ml/min
N ₂ flow rate	150 ml/min
Steam flow rate	100 ml/min
H ₂ flow rate	50 ml/min
Rate of heating	20°C/min
Pyrolysis atmosphere	N ₂ -50%, H ₂ -50%
Pressure	atmospheric

2.1.1.3 Model Predictions

The prediction of bed volume was made as per the computation scheme shown in Figure 2-5 and as discussed in Section 2.1.1.1.2. An average relative reactivity of 5.9 was obtained for Pittsburgh seam coal on the TGA. This value was used in the model predictions.

Bed volume was predicted for various steady-state periods by means of the gasification model and kinetic data developed on the TGA. The ratio of predicted to measured bed volume is plotted as a function of freeboard temperature in Figure 2-8. Measured bed volume is plotted against predicted bed volume in Figure 2-9. The predicted bed volume is within +30 percent of the actual bed volume, except during three steady-state periods, showing that the agreement between the model and the PDU performance is very good.

Table 2-2 gives the coal feed rate, estimated carbon gasification rate from solid flow rates, and percent of fixed carbon gasified. Examining the last item, fixed carbon gasified is low (< 70%) for three of the four steady-state periods during which fines were not recycled as expected. But it is much lower for test TP-23-4(1B) compared to other tests where fines are recycled and hence the predicted volume is substantially lower than the measured bed volume in this case. These anomalies are probably due to measurement errors.

2.1.1.3.1 Sensitivity Analysis

The effect of uncertainties in measurement on the difference between predicted and observed performances of the PDU is shown in this section. A simple evaluation of the magnitude of the differences on the bed carbon holdup between the predicted value (by the gasification model) and the observed value in the PDU that can be expected on the basis of measurement errors alone is presented here.

Figure 2-8. Comparison of Model Predictions With PDU Data

Figure 2-9. Comparison of Measured and Predicted Bed Volume.

TABLE 2-2
GASIFICATION RATE WITH PITTSBURGH SEAM COAL AS THE FEEDSTOCK

<u>PDU Test TP-0</u>	<u>Coal Feed Rate lb/hr</u>	<u>Carbon Gasification Rate, lb/hr</u>	<u>% of Fixed Carbon (F.C.) Gasified and Combusted</u>	<u>Carbon Feed Rate in the Form of F.C., lb/hr</u>
18-2(5)*	505	184	63.7	289
18-2(6)*	995	389	68.3	569
22-1(1)	695	310	83.3	372
23-4(1A)	890	402	81.2	495
23-4(1B)	856	290	61.1	474
23-4(2A)	718	315	79.5	396
23-4(2B)	681	276	75.0	368
27-2(2B1)	1007	501	86.1	582
27-2(2B2)	884	428	84.9	504
27-2(2B3)	962	458	83.9	546
27-3(2A)*	1086	411	67.3	611
27-3(2B)*	1022	470	82.9	567
27-3(2C)	970	465	87.8	530
27-3(2D)	964	436	83.3	523
27-3(2D)	1073	412	69.5	592
27-3(2A1)	1043	465	79.3	586
28-1(1A)	960	490	94.4	519
28-1(1B)	982	487	91.7	531

* Indicates that char fines were not recycled.

<u>Variable</u>	<u>Typical Value in PDU</u>	<u>Percent Error</u>	<u>Absolute Error</u>
Freeboard temperature, °F	1800	+1.4	+25
Bed density, lb/ft ³	20	+10	+2
Bed carbon content, fraction	0.5	+10	+0.05
Bed height, ft	10	+10	+1.0

Bed carbon holdup (BCH) = Area x Bed height x Bed density x Bed carbon content

$$BCH = A \times (H \pm 1) (\rho_B \pm 2) (C_C \pm 0.05) \quad (2-23)$$

Taking the above values,

$$BCH = A \times (10 \pm 1) (20 \pm 2) (0.5 \pm 0.05)$$

$$BCH = 100A \left(\begin{smallmatrix} +33A \\ -27A \end{smallmatrix} \right) = 73A \text{ to } 133A$$

Thus the error in BCH = $\pm 30\%$

Express the gasification rate in the gasifier as Rate = Constant $\cdot \exp(-E/RT)$ in order to calculate the effect of error in temperature measurement on predicted bed carbon holdup. Take a modest value for the activation energy as 20,000 cal/gmole (36,000 Btu/lb mole)

Error estimate for the lower value of T

$$T_1 = 1775^\circ\text{F}; \quad T_2 = 1800^\circ\text{F}$$

$$\frac{R_1}{R_2} = \exp\left[-18,117.8 \left(\frac{1}{2235} - \frac{1}{2260}\right)\right] \quad (2-24)$$

Gasification rate $\propto$ BCH

With an error of $\pm 25^\circ\text{F}$ at 1800°F in temperature measurement, error in BCH = $\pm 9\%$.

Denote Δ as the difference between the measured and predicted values of bed carbon holdup per unit cross-sectional area.

$$\Delta = (\text{Measured} \pm 30\%) - (\text{Predicted value} \pm 9\%) = \pm 39\%/\text{lb of carbon}/\text{ft}^2$$

Thus, if the model can predict perfectly, the differences between the predicted and the observed values are in the range of ± 39 percent. If the predicted value differs from the real value, the value of Δ will be larger than ± 39 percent. In the above analysis, the errors in steam flow rate and coal feed rate have not been considered. The difference between the true observed and the true predicted values from the difference between the observed and predicted values can now be estimated.

Denote:

- x - observed value
- x^* - predicted value
- x^T - true observed value
- x^{*T} - true predicted value
- ϵ_1 - error in the observed value due to measurement errors
- ϵ_2 - uncertainty in the predicted value due to measurement errors in the operating parameters used for the prediction

As an example, take the predicted value as 100 and the observed value as 140.

$$\begin{aligned}
 x_1 &= x_1^T \pm \epsilon_1; \quad x_1^T = (x_1 \pm \epsilon_1) \\
 x_1^* &= x_1^{*T} \pm \epsilon_2, \quad x_1^{*T} = (x_1^* \pm \epsilon_2) \\
 x_1^T - x_1^{*T} &= (x_1 - x_1^*) \pm (\epsilon_2 - \epsilon_1) \\
 (x_1^T - x_1^{*T})^2 &= (x_1 - x_1^*)^2 + (\epsilon_2 - \epsilon_1)^2 \pm 2(x_1 - x_1^*)(\epsilon_2 - \epsilon_1) \\
 &= 40^2 + 21^2 \pm 2 \times 40 \times 21 \\
 &= 2041 \pm 1680 \\
 (x_1^T - x_1^{*T}) &= \sqrt{3721} \text{ to } \sqrt{361} \\
 &= 61 \text{ to } 19 \\
 \text{Define \% error as } &(x_1^T - x_1^{*T}) \times 100 / x_1^{*T} \\
 &= (x_1^T - x_1^{*T}) \times 100 / x_1^*
 \end{aligned}$$

Observed error = $(140-100)/100 = 40\%$
 True error = 19 to 61%

The above example shows the effect of measurement errors on the uncertainty of the difference between true observed and true predicted values. This uncertainty should be taken into account in evaluating the validity and adequacy of the model and in selecting desired improvements in the model.

2.1.1.4 Conclusions

For the PDU results reported here, the agreement between the model and the PDU performance is very good, despite the measurement uncertainties (resulting from errors at any given location and time, variation of measured value with time during steady-state periods, and spatial variation of some of the operating parameters throughout the gasifier bed). The difference between the measured and predicted bed volumes is greater than 30 percent at temperature close to 1900°F (1040°C). This will be examined in the future after additional kinetic data become available. There appears to be no reason to incorporate any refinements to the model for the feedstock examined here. Each steady-state period will be examined to see if the important variables for this study are steady within an acceptable range. If not, smaller steady-state periods will be selected for analysis.

The sensitivity analysis shown here points out the impact of measurement uncertainties on the comparison between the model and PDU performance. The differences between the model and the measured data are within the range of the measurement uncertainties for most of the steady-state periods.

No anomalies could be observed in the PDU performance in terms of the effect of either steam/carbon ratio or temperature. The specific gasification rate increased with both temperature and steam/carbon ratio, as predicted. No optimum value for steam/carbon ratio, so far as gasification performance is concerned, could be observed for the tests considered here.

2.1.2 Analysis and Prediction of PDU Performance with Western Coals

PDU performance with Wyoming Sub-C coal was analyzed in terms of gasification rate and predicted using a gasification model and kinetic data developed on carbon gasification reactions. PDU gasification performance was measured in terms of carbon gasification rate for a given bed carbon holdup (bed volume) at the operating conditions. The effect of steam/carbon ratio and freeboard temperature on specific gasification rate and the ability to predict PDU performance using the gasification model are examined in this section. A comparison of PDU performance with Wyoming Sub-C coal is made with that of Pittsburgh seam coal.

2.1.2.1 Analysis

The objective of this task was to analyze PDU performance with western coals in terms of gasification rate and product gas composition and to predict performance using a gasification model and available kinetic data on carbon gasification reactions. This section compares the predicted values with those measured in the PDU. Tests performed with Wyoming Sub-C coal and Texas and North Dakota lignites are analyzed herein.

Western coals represent the most reactive feedstocks processed in the PDU to date and it was of interest to determine if the effect of temperature and steam/carbon ratio on gasification rate would be similar to that of a low-reactive coal such as Pittsburgh seam coal. A comparison is made between Wyoming Sub-C and Pittsburgh seam coals in this section.

Steady-state data from tests TP-027-4 through TP-035-1 with Wyoming Sub-C coal feedstock, tests TP-033-1 and TP-034-3 with North Dakota lignite feedstock, and tests TP-023-4 through TP-034-3 with Texas lignite feedstock were analyzed to evaluate the PDU performance. Freeboard temperature, measured bed volume, specific gasification rate, and the estimated steam/carbon ratios for these tests are given in Tables 2-3, 2-4 and 2-5.

Specific gasification rate is plotted as a function of freeboard temperature for steam/carbon ratio range of 1.5 - 3.7 mole/mole for Wyoming Sub-C coal in Figure 2-10. The specific gasification rate increased with temperature just as expected. A comparison of PDU performance with Wyoming Sub-C and Pittsburgh seam coals is shown in Figure 2-11, assuming that the results of Pittsburgh seam coal can be extrapolated. At a freeboard temperature of 1640°F, the specific gasification rate of Wyoming Sub-C coal is about five times that of Pittsburgh seam coal.

Specific gasification rate is plotted as a function of freeboard temperature for North Dakota and Texas lignites in Figures 2-12 and 2-13. For North Dakota lignite, the gasification rate remained relatively constant in the temperature range of 1480 to 1540°F (804 to 838°C), while for Texas lignite from Monticello Mine, there is a large scatter in the data which indicates that there is an appreciable increase in gasification rate with an increase in temperature in the range of 1460 to 1640°F (793 to 893°C). The scatter in the data is reflected in the plots where predicted bed volumes are compared with observed values as shown later.

The uncertainty in the calculation of fines recycle rate has a significant impact on the estimation of gasification rate. Scatter in the data on gasification rate versus temperature might be reduced if the steady state of fines recycle rate is adopted as one of the steady-state criteria applied to PDU operation in this analysis.

Table 2-6 gives the coal feedrate, estimated carbon gasification rate from solid flowrates, and percent of fixed carbon gasified for Wyoming Sub-C coal. The last item varied from 62 to 88 percent. In the case of Pittsburgh seam coal, it varied from 61 to 94 percent. The percentage of fixed carbon gasified is plotted versus temperature for Wyoming Sub-C and Pittsburgh seam coals in Figures 2-14 and 2-15. It is fairly independent of operating temperature in the case of Wyoming Sub-C coal while it increased substantially with temperature in the case of Pittsburgh seam coal, except for a few data points. Tables 2-7 and 2-8 show the data for North Dakota and Texas lignites.

TABLE 2-3

PDU PERFORMANCE WITH WYOMING SUB-C COAL, STEAM/CARBON RATIOS

PDU Test TP-0	Freeboard Temp., °F	Steam/Carbon Ratio, Mole/Mole	Gasification Rate, lb/hr	Sp. Gasification Rate, hr ⁻¹	Predicted Bed Volume, ft ³	Measured Bed Volume, ft ³
27-4	1557	2.69	552	0.73	14	33.6
(M.1A)						
27-4	1591	2.62	594	0.98	14	36.4
(M.3A)+						
27-5	1567	3.65	375	1.19	11	31.4
(3A)+						
28-1	1587	2.00	788	1.31	25	36.1
(2A-1)						
28-1	1556	1.88	737	1.14	28	37.0
(2A-2)						
28-1	1557	1.80	758	1.15	27	36.1
(2B)						
29-1/030-1	1603	2.17	560	0.93	15	30.7
(4A)						
29-1/030-1	1612	2.01	685	1.16	16	30.1
(4B)						
33-2(1A-1)	1613	1.71	873	1.07	28	32.1
33-2(1B-1)	1604	1.57	684	1.30	33	31.6
33-2(1B-2)	1604	1.47	757	1.50	40	32.0
33-2(1B-3)	1636	1.62	691	1.56	30	32.3
34-1(1A)	1502	2.56	506	0.78	39	34.5
34-1(1A-1)	1517	2.46	510	0.82	41	34.5
34-1(2A)	1507	2.69	509	0.79	35	35.1
34-1(2B)	1506	2.06	546	0.88	52	34.2
34-1(2C)	1526	1.97	583	0.92	54	35.1
34-1(2D)	1509	2.24	534	0.85	51	35.1
35-1(1A)	1545	2.07	732	1.25	33	37.9
35-1(1B)	1539	2.15	720	1.33	35	37.8
35-1(1A1)	1557	2.08	755	1.67	42	33.2
35-1(1A2)	1559	2.13	742	1.24	32	37.8
35-1(2A)	1577	2.52	632	1.15	23	38.1
35-1(3A)	1517	2.42	683	1.29	31	35.9
35-1(4A)	1544	2.54	714	1.45	39	33.8

TABLE 2-4

PDU PERFORMANCE WITH NORTH DAKOTA LIGNITE, STEAM/CARBON RATIOS

PDU Test TP-0	Freeboard Temp., °F	Steam/Carbon Mole/Mole	Gasification Rate, lb/hr	Sp. Gasification Rate, hr ⁻¹	Predicted Bed Volume, ft ³	Measured Bed Volume, ft ³
33-1(3A)	1506	1.62	777	1.97	43	33.3
33-1(3A-1)	1531	1.99	652	1.58	30	34.5
33-1(3B-1)	1488	1.29	632	1.56	36	34.1
34-3(1A)	1509	2.16	619	1.78	33	31.6
34-3(1B)	1522	2.03	604	1.97	33	31.3
34-3(1C)	1534	2.10	636	1.74	31	31.3
34-3(1D)	1534	2.26	585	1.57	26	31.4
34-3(1F)	1516	2.26	606	1.75	29	31.6

TABLE 2-5

PDU PERFORMANCE WITH TEXAS LIGNITE, STEAM/CARBON RATIOS

PDU Test TP-0	Freeboard Temp., °F	Steam/Carbon Mole/Mole	Gasification Rate, lb/hr	Sp. Gasification Rate, hr ⁻¹	Predicted Bed Volume, ft ³	Measured Bed Volume, ft ³
<u>Monticello Mine</u>						
23-4(3)	1524	3.29	324	1.67	48	29.7
27-3(1A)	1543	3.93	226	0.49	10	32.5
27-5(1B-1)	1635	3.23	274	1.27	20	31.9
27-5(1C-1)	1519	3.37	513	1.32	25	20.6
27-5(1D-1)	1467	4.77	241	0.50	28	35.0
<u>Big Brown Mine</u>						
28-3(1A)	1537	1.92	500	1.48	42	33.5
28-3(2A)	1502	2.71	364	0.79	22	32.6
28-3(2B)	1506	2.49	382	0.98	30	33.0

Figure 2-10. Effect of Temperature on Gasification Rate for Wyoming Sub C Coal

Figure 2-11. PDU Performance with Wyoming Sub C Coal

Figure 2-12. Effect of Temperature on Gasification Rate for N. Dakota Lignite

Figure 2-13. Effect of Temperature on Gasification Rate for Texas Lignite

TABLE 2-6
PDU PERFORMANCE WITH WYOMING SUB-C COAL, FIXED CARBON GASIFIED

PDU Test TP-0	Coal Feedrate, lb/hr	F.C. Feedrate, lb/hr	Gasification Rate, lb/hr	% F.C. Gasified	Bed Ash %
27-4(M. 1A)	2357	908	565	62.3	27.4
27-4(M. 3A)*	2552	983	594	60.4	27.5
27-5(3A)*	2077	833	632	75.9	21.9
28-1(2A-1)	2368	902	788	87.5	20.8
28-1(2A-2)	2341	941	737	78.3	18.0
28-1(2B)	2293	922	758	82.3	18.4
29-1/030-1(4A)	2052	778	560	72.0	23.0
29-1/030-1(4B)	2190	819	685	83.6	24.2
33-2(1A-1)	2323	885	673	76.0	22.8
33-2(1B-1)	2271	909	681	74.9	19.8
33-2(1B-2)	2377	952	757	79.5	21.8
33-2(1B-3)	2256	888	691	77.8	31.1
34-1(1A)	1990	744	506	68.0	24.6
34-1(1A-1)	1837	684	510	74.6	24.3
34-1(2A)	1738	687	509	74.1	23.4
34-1(2B)	1716	693	546	78.0	22.1
34-1(2C)	1677	674	503	86.5	25.3
34-1(2D)	1686	661	534	80.8	27.6
35-1(1A)	2153	829	732	88.3	30.0
35-1(1B)	2134	800	720	90.0	35.0
35-1(1A1)	2313	881	755	85.7	33.8
35-1(1A2)	2249	838	742	88.5	29.9
35-1(2A)	2180	749	632	84.4	33.3
35-1(3A)	2493	846	683	80.7	35.1
35-1(4A)	2658	978	714	73.0	31.0

F.C. - Fixed Carbon

* No recycle of Fines

Figure 2-14. PDU Performance with Wyoming Sub C Coal

Figure 2-15. PDU Performance with Pittsburgh Seam Coal

TABLE 2-7

PDU PERFORMANCE WITH N. DAKOTA LIGNITE, FIXED CARBON GASIFIED

<u>PDU Test</u> <u>TP-0</u>	<u>Coal</u> <u>Feedrate,</u> <u>lb/hr</u>	<u>F.C.</u> <u>Feedrate,</u> <u>lb/hr</u>	<u>Gasification</u> <u>Rate,</u> <u>lb/hr</u>	<u>% F.C.</u> <u>Gasified</u>	<u>Bed</u> <u>Ash %</u>
33-1(3A)	2471	930	777	83.5	37.6
33-1(3A-1)	2358	883	652	73.8	30.3
33-1(3B-1)	2272	872	632	72.5	33.4
34-3(1A)	2185	690	619	89.7	22.8
34-3(1B)	2177	751	684	91.1	24.4
34-3(1C)	2187	730	636	87.1	20.2
34-3(1D)	2247	666	585	87.8	19.8
34-3(1F)	2110	732	606	82.8	26.7

F.C. - Fixed Carbon

TABLE 2-8

PDU PERFORMANCE WITH TEXAS LIGNITE, FIXED CARBON GASIFIED

<u>PDU Test</u> <u>TP-0</u>	<u>Coal</u> <u>Feedrate,</u> <u>lb/hr</u>	<u>F.C.</u> <u>Feedrate,</u> <u>lb/hr</u>	<u>Gasification</u> <u>Rate,</u> <u>lb/hr</u>	<u>% F.C.</u> <u>Gasified</u>	<u>Bed</u> <u>Ash %</u>
<u>Monticello</u> <u>Mine</u>					
23-4(3)	1546	422	324	76.8	59.7
27-3(1A)	1413	443	226	51.0	45.3
27-5(1B-1)	1088	334	274	82.0	76.2
27-5(1C-1)	3081	922	513	55.6	58.1
27-5(1D-1)	2049	568	241	42.4	53.6
<u>Big Brown</u> <u>Mine</u>					
28-3(1A)	1670	537	500	93.1	56.6
28-3(2A)	1616	562	364	64.8	46.9
28-3(2B)	1493	498	382	76.7	40.4

The percentage of fixed carbon gasified as a function of temperature for Texas lignite is shown in Figure 2-16. It varied from 42 to 93 percent showing a substantial increase with an increase in temperature. PDU performance with North Dakota lignite is much better with percentage of fixed carbon varying from 73 to 91 over the narrow temperature range of 1488 to 1534°F (809 to 834°C).

The reasons for poor gasifier performance in the case of some of the steady-state periods are due to appreciable loss of fines and appreciable loss of carbon in the discharge.

A comparison of the PDU performance with Pittsburgh seam coal, as reported earlier, with that of Wyoming Sub-C coal shows that the performance in the latter case is slightly lower, possibly due to lower operating temperatures. The lower operating temperatures coupled with low bed ash content are expected to result in lower ash agglomeration rate.

The effect of steam/carbon ratio on the specific gasification rate at temperatures of about 1500° and 1600°F (815 and 871°C) is shown in Figures 2-17 and 2-18. At both temperature levels, the gasification rate decreased with an increase in steam/carbon ratio. The effect of steam/carbon ratio at 1600°F (871°C) is greater than that at 1500°F (815°C). This behavior is in sharp contrast with that of Pittsburgh seam coal where the gasification rate increased appreciably with an increase in steam/carbon ratio at temperatures of 1821°F to 1900°F (994 to 1038°C). Based on these results, it is recommended that steam flowrates be reduced further, if possible, to increase the gasification rate.

The extent to which the water-gas shift reaction approaches equilibrium in the product gas was examined for the steady-state periods studied here for Wyoming Sub-C coal.

Measured values of the equilibrium constant are plotted as a function of temperature in Figure 2-19. The experimental values deviate substantially from the theoretical relation, indicating that the water-gas shift reaction did not reach equilibrium in the product gas. Note that small measurement errors in the product gas composition result in an appreciable error in the equilibrium constant of the water-gas shift reaction, K_E , value.

2.1.2.2 Model Predictions

Bed volume was predicted for various steady-state periods by means of the gasification model and kinetic data developed on carbon gasification reactions. Average relative reactivities used for these predictions are shown in Table 2-9. An activation energy of 25,000 cal/gmole was used based on the kinetic studies performed on the thermogravimetric analyzer with Wyoming Sub-C coal. Results showing the effect of temperature on the reaction rate of Wyoming Sub-C coal are given in Table 2-10 and are plotted in Figure 2-20 to determine the activation energy.

Figure 2-16. PDU Performance with Texas Lignite

Figure 2-17. Effect of Steam on Gasification Rate of Wyoming Sub C Coal

Figure 2-18. Effect of Steam on Gasification Rate of Wyoming Sub C Coal

Figure 2-19. Experimental Shift Equilibrium Values for Wyoming Sub C Coal

TABLE 2-9

REACTIVITIES USED FOR WESTERN COALS IN THE MODEL

<u>FEEDSTOCK</u>	<u>REACTIVITY</u>
Wyoming Sub-C Coal	18
North Dakota Lignite	30
Texas Lignite (Monticello Mine)	10
Texas Lignite (Big Brown Mine)	15

TABLE 2-10

EFFECT OF TEMPERATURE ON THE REACTIVITY OF WYOMING SUB-C COAL

<u>Temperature, °F</u>	<u>1800</u>	<u>1700</u>	<u>1600</u>	<u>1500</u>
<u>Reactivity Data*</u>				
Average reaction rate x 10 ² , min ⁻¹	26.60	21.29	13.80	6.47
Average relative reactivity	88.70	71.00	46.00	21.60
<u>Proximate Analysis, Wt.%</u>				
Moisture	17.57	15.24	15.83	12.27
Volatile Matter	35.05	39.43	40.48	36.11
Ash	5.72	4.31	6.26	10.42
Fixed Carbon	41.66	41.03	37.43	41.20
<u>Analysis of Char, Wt.%</u>				
Carbon	87.98	90.46	85.63	79.88
Ash	12.02	9.54	14.37	20.12

* Average of three or more tests.

Figure 2-20.. Char-Steam Reaction with Wyoming Sub C Coal

A comparison of the measured and predicted bed volumes is shown in Figure 2-21. Agreement between the two is generally within ± 50 percent. The ratio of predicted/measured bed volumes as a function of temperature is shown in Figure 2-22.

Based on the sensitivity analysis performed earlier, it was shown that with a very conservative estimate of measurement errors in freeboard temperature and density, carbon content and height of bed alone, the differences between the predicted and the observed values for a perfect model were in the range of ± 39 percent. Taking this into account, the agreement between the model and the PDU performance is reasonably good.

Comparisons of the measured and predicted bed volumes for North Dakota and Texas lignites is shown in Figures 2-23 and 2-24. Agreement between the measured and predicted values is very good for North Dakota lignite while the plot for Texas lignite shows larger differences between measured and predicted values resulting from a large scatter in PDU data obtained with this feedstock.

The true kinetic reactivities of Wyoming Sub-C coal and the two lignites were determined on a thermogravimetric analyzer and found to be in the range of 90 to 120 for Wyoming Sub-C coal and about 120 to 130 for Texas lignite from Big Brown Mine and North Dakota lignites. The overall reactivity of any feedstock depends upon the mass transfer rate from the bubble phase to the particulate phase and on the chemical reaction rate in the particulate phase. In the case of eastern coals, which are low-reactive, the reactivity determined on the TGA represents the overall reactivity since the mass transfer rate from the bubble phase to the particulate phase is very high compared to the chemical reaction rate. In the case of western coals, which are highly reactive, the magnitude of the mass transfer rate from the bubble phase to the particulate phase appears to be comparable to that of the chemical reaction rate and thus would have a strong influence on the overall rate. The computer model used in this study does not contain equations to estimate the mass transfer rate between the two phases. Hence it is recommended that the gasification model be expanded to include this estimation to be able to predict the PDU performance solely on the basis of kinetic data.

2.1.2.3 Conclusions

The following conclusions and recommendations are made based on the results of this study:

- The specific gasification rate increased with an increase in temperature for Wyoming Sub-C coal and Texas lignite. The temperature range covered for North Dakota Lignite is very narrow and no significant effect of temperature is noticed.

Figure 2-21. Comparison of Measured and Predicted PDU Bed Volumes for Wyoming Sub C Coal

Figure 2-22. Comparison of Model Prediction with PDU Data - Wyoming Sub C Coal

Figure 2-23. Comparison of Measured and Predicted Bed Volumes for N. Dakota Lignite

Figure 2-24. Comparison of Measured and Predicted Bed Volumes for Texas Lignite

- There is a significant decrease in the specific gasification rate of Wyoming Sub-C coal with an increase in the steam/carbon ratio for the narrow temperature ranges of 1515 and 1600 $\pm$ 15°F (824 and 871 $\pm$ 8°C). This is in sharp contrast with the behavior of Pittsburgh seam coal. The data on the two lignites are too limited to extract any information on the effect of steam.
- The gasification rate of Wyoming Sub-C coal is about five times higher than that of Pittsburgh seam coal based on the comparison shown in Figure 2-11.
- Inclusion of steady-state conditions in fines recycle rate in the definition of a steady-state period is expected to reduce the differences between predicted and observed results.
- As the rate determined on the TGA represents true kinetic rate, it should be combined with mass transfer rate calculations (from bubble phase to particulate phase) to obtain an overall reactivity for highly reactive coals such as Wyoming Sub-C coal and the two lignites studied here.
- The gasification model should be expanded to include calculations on mass transfer rate to predict behavior of western coals in the PDU.

2.1.3 Analysis and Prediction of PDU Performance with Midwestern Coals

Performance of the KRW Process Development Unit gasifier was analyzed in terms of the gasification rate and product gas composition for tests performed with Indiana No. 7, Western Kentucky, and Illinois No. 6 coals. Effect of temperature and steam flowrate on the specific gasification rate is shown. Bed volume (carbon holdup) was predicted for various tests by means of a gasification model and kinetic data developed on carbon gasification reactions, and the value was found to be in reasonable agreement with measured values for a large number of tests.

2.1.3.1 Analysis

The objective of this task was to analyze PDU performance in terms of gasification rate for various steady-state periods with midwestern coal feedstocks and to predict performance using a gasification model and kinetic data developed on carbon gasification reactions.

PDU gasification performance is measured in terms of the bed carbon holdup maintained to achieve the desired coal processing rate at the operating conditions for a given feedstock and product gas composition. The feedstocks are characterized for reactivity on a thermogravimetric analyzer (TGA). The detailed kinetic data on carbon gasification reactions developed on a fluidized bed test facility are used in the gasification model to predict the bed volume and product gas composition.

PDU steady-state data for tests TP-019-2 through TP-033-3 with Western Kentucky coal, tests TP-019-1 through TP-030-2 with Indiana No. 7 coal, and test TP-029-1 with Illinois No. 6 coal feedstocks were analyzed to evaluate the PDU performance. Freeboard temperature, measured bed volume and specific gasification rate and the calculated steam/carbon ratios for these tests are given in Tables 2-11 and 2-12.

The specific gasification rate for Indiana No. 7 coal is plotted as a function of freeboard temperature for the steam/carbon ratio range of 1.52 - 2.12 in Figure 2-25. There is appreciable scatter in the data which were obtained at the same pressure. It is surprising to see that the specific gasification rate decreased with an increase in freeboard temperature. A similar plot was shown for Western Kentucky coal in Figure 2-26. There is large scatter in these data as well. The scattered data show a trend in decrease of specific gasification rate with an increase in temperature. Data on Western Kentucky coal are too limited to extract any information on the effect of steam on the gasification rate. Data obtained with Illinois No. 6 coal are limited and are presented along with the data on Western Kentucky coal in Table 2-11.

Effect of steam/carbon ratio on gasification rate of Indiana No. 7 coal in the temperature ranges of 1803 to 1818°F (984 to 992°C) and 1875 to 1900°F (1024 to 1038°C) is shown in Figures 2-27 and 2-28. Regression lines are shown in both plots. The specific gasification rate decreased with an increase in steam/carbon ratio at both temperature ranges except in the steam/carbon ratio range of 1.5 - 1.75 indicating that the steam flows used were excessive from the point of view of gasification. These data indicate that the optimum for steam/carbon ratio is in the range of 1.5 to 1.8. The range of steam/carbon ratio used for Western Kentucky is too narrow to determine the effect of steam on gasification rate.

Examination of the gasification data of Western Kentucky coal shows that in the case of tests TP-019-2 and TP-019-3, the gasification rate was substantially lower than that of the rest of the tests despite the high steam/carbon ratios used for these tests. The reason for this low gasification rate could be either due to the unsteady-state nature of the tests and/or due to poor gas-solid contacting. Hence, these results were not used in model predictions. No effect on gasification rate can be seen from water injected in the freeboard. The low gasification rate obtained in tests TP-019-2 and TP-019-3 cannot be explained either in terms of the effect of pressure or in not injecting water in the freeboard.

Percentage of fixed carbon gasified as a function of freeboard temperature is shown in Figures 2-29 and 2-30 for Indiana No. 7 and Western Kentucky coals. Although there is substantial scatter in both plots, there is a trend showing a reduction in fixed carbon gasified with an increase in temperature. Lines are drawn based on regression analysis. Examining the data on specific gasification rate as well as percentage of fixed carbon

TABLE 2-11

PDU PERFORMANCE WITH WESTERN KENTUCKY AND ILLINOIS NO. 6 COALS

PDU Test TP-0	Freeboard Temp., °F	Steam/Carbon Mole/Mole	Gasification Rate, lb/hr	Sp. Gasification Rate, hr ⁻¹	Predicted Bed Volume, ft ³	Measured Bed Volume, ft ³
19-2(M.4A)	1759	6.32	121	0.21	2	46.6
19-2(M.4B)	1767	6.68	115	0.23	3	42.4
19-2(4C)	1766	5.17	152	0.30	4	44.0
19-3(1A)	1790	2.69	348	0.36	5	43.3
19-3(1B)	1779	3.90	247	0.28	4	48.2
19-3(1C)	1808	2.55	355	0.53	6	45.2
19-3(1D)	1799	3.03	285	0.41	5	46.8
27-4(5A-1)	1814	1.67	576	1.30	21	33.2
27-4(5A-2)	1805	1.49	626	1.66	31	34.7
27-4(5B-1)	1851	1.45	849	2.37	34	31.1
27-4(5B-2)	1829	1.37	826	2.46	42	33.6
27-4(6A)	1913	1.29	915	2.62	52	44.0
32-2(3A)	1807	1.21	507	1.43	35	34.8
32-2(3A-1)	1860	1.11	629	1.78	47	31.7
32-2(3B-1)	1880	1.50	473	1.36	29	33.4
32-2(3B-2)	1872	1.39	510	1.45	34	35.1
32-2(3B-3)	1813	1.27	489	1.46	39	33.8
32-2(3B-4)	1851	1.11	626	1.04	48	35.0
32-2(3C-1)	1824	1.06	734	2.10	57	34.5
33-3(2A)	1836	1.37	510	1.25	27	35.8
Illinois No. 6						
29-1(1A)	1834	1.80	390	1.13	15	33.1
29-1(3A)	1786	1.74	314	0.92	15	32.1

TABLE 2-12

PDU PERFORMANCE WITH INDIANA NO. 7 COAL

PDU Test TP-0	Freeboard Temp., °F	Steam/Carbon Mole/Mole	Gasification Rate, lb/hr	Sp. Gasification Rate, hr ⁻¹	Predicted Bed Volume, ft ³	Measured Bed Volume, ft ³
19-1(4)	1709	3.22	155	0.25	6	46.2
19-2(1)	1721	2.80	271	0.49	9	46.4
19-2(M.2A)	1737	2.09	308	0.47	9	46.8
19-2(M.2B)	1717	2.19	293	0.46	10	46.3
25-1(2A)	1845	2.80	235	0.69	5	29.2
25-1(2B)	1818	4.46	145	0.42	2	28.8
25-1(2C)	1817	3.43	189	0.66	4	27.9
25-1(3A)	1828	2.43	254	0.80	7	29.5
25-1(3B)	1811	2.35	263	0.96	9	27.6
25-1(4)	1806	2.53	208	1.06	8	25.1
25-2(3A-1)	1792	2.07	338	1.09	20	42.4
25-2(3A-2)	1809	2.50	253	1.07	13	36.5
25-2(3A-3)	1803	2.33	283	1.14	14	35.6
28-2(1A)	1877	2.01	441	1.45	16	35.3
28-2(2A)	1899	2.05	496	1.62	19	33.4
28-2(2B)	1891	2.12	477	1.50	17	33.4
28-2(2C)	1874	1.57	640	2.39	39	34.3
28-2(2D)	1890	1.52	688	2.32	39	33.7
28-2(2E)	1891	1.89	556	1.84	25	33.4
28-2(2F)	1822	1.57	721	2.56	47	33.8
28-2(2G)	1826	1.63	697	2.84	53	34.1
28-2(2H)	1819	1.74	678	2.59	44	33.8
28-2(2I)	1889	1.77	720	2.46	37	35.0
28-2(2J)	1875	1.76	708	2.47	37	33.8
30-2(1B-1)	1831	2.04	471	1.60	18	34.0
30-2(1B-2)	1845	1.68	514	1.65	25	32.1

Figure 2-25. PDU Performance with Indiana No. 7 Coal

Figure 2-26. PDU Performance with W. Kentucky Coal

Figure 2-27. Effect of Steam on Gasification Rate with Indiana No. 7 Coal, Temperature Range of 1803°F to 1818°F

Figure 2-28. Effect of Steam on Gasification Rate with Indiana No.7 Coal, Temperature Range of 1874°F to 1900°F

Figure 2-29. PDU Performance with Indiana No. 7 Coal

Figure 2-30. PDU Performance with W. Kentucky Coal

gasified as a function of temperature shows that there is no advantage in operating the gasifier at temperatures higher than 1800°F (982°C) for these two feedstocks.

Tables 2-13 and 2-14 give the coal feedrate, estimated carbon gasification rate from solid flowrates, and percent of fixed carbon gasified for Western Kentucky and Indiana coals. The last item varied from 30 to 91 percent for Western Kentucky coal. PDU performance with Western Kentucky coal in tests TP-019-2 and TP-019-3 was relatively poor and is reflected in the low bed volumes predicted for these tests. In the case of Indiana coal, tests TP-025-1 and TP-025-2 showed poor performance with fixed carbon gasified varying from 31 to 56 percent.

The extent to which the water-gas shift reaction approaches equilibrium in the product gas was examined for Indiana and Western Kentucky coals. Measured values of this ratio are plotted as a function of temperature in Figures 2-31 and 2-32 for Indiana and Western Kentucky coals. The experimental values deviate substantially from the theoretical relation, indicating that the water-gas shift reaction did not reach equilibrium in the product gas. Large scatter in earlier test data (19 and 25 series) for Indiana No. 7 coal indicates that the measurement of water content in the product gas probably has substantial error since the composition of other components could be measured fairly accurately.

2.1.3.2 Model Predictions

Bed volume (bed carbon holdup) was predicted for various steady-state periods by the gasification model and the available kinetic data. The ratio of the predicted to actual bed volumes is plotted as a function of temperature for Western Kentucky coal in Figure 2-33. Predicted bed volumes are compared with the measured values in Figure 2-34. Agreement between the two is generally within ± 50 percent.

For Indiana No. 7 coal, the ratio of the predicted to actual bed volume as a function of temperature is shown in Figure 2-35. A comparison of the predicted and measured bed volume is shown in Figure 2-36. The predictions are generally within ± 50 percent of the measured bed volume.

Reactivity data developed on the TGA as well as the reactivities used for the model predictions for the three feedstocks whose behavior in the PDU is analyzed in this study are given in Table 2-15.

TABLE 2-13

PDU PERFORMANCE WITH WESTERN KENTUCKY AND ILLINOIS NO. 6 COALS

PDU Test TP-0	Coal Feed- Rate lb/hr	Fixed Carbon Feedrate lb/hr	Gasification Rate, lb/hr	% Fixed Carbon Gasified	Bed Ash, %
19-2(M.4A)	810	398	121	30.4	28.3
19-2(M.4B)	723	352	115	32.7	32.9
19-2(4C)	806	395	152	38.5	33.1
19-3(1A)	890	460	348	75.7	21.6
19-3(1B)	945	468	247	52.8	26.6
19-3(1C)	1113	545	355	65.1	31.0
19-3(1D)	1030	493	285	57.8	31.5
27-4(5A-1)	1781	918	576	62.7	29.8
27-4(5A-2)	1582	833	626	75.2	38.0
27-4(5B-1)	2222	1133	849	74.9	36.2
27-4(5B-2)	2037	1075	826	76.8	38.9
27-4(6A)	2300	1206	915	75.9	33.6
32-2(3A)	966	557	507	91.0	27.8
32-2(3A-1)	1445	711	629	88.2	26.6
32-2(3B-1)	1468	732	473	64.6	25.5
32-2(3B-2)	1454	725	510	70.3	26.1
32-2(3B-3)	1347	676	489	72.3	26.5
32-2(3B-4)	1340	761	626	82.3	24.9
32-2(3C-1)	1497	850	734	86.4	23.5
33-3(7A)	1458	729	510	70.0	20.9
Illinois No. 6					
29-1(1A)	1152	574	390	67.9	29.2
29-1(3A)	855	422	314	74.4	26.6

TABLE 2-14

PDU PERFORMANCE WITH INDIANA NO. 7 COAL

PDU Test TP-0	Coal Feed Rate lb/hr	Fixed Carbon Feedrate lb/hr	Gasification Rate, lb/hr	% Fixed Carbon Gasified	Bed Ash, %
19-1(4)	742	321	155	43.3	22.9
19-2(1)	866	423	271	64.1	30.8
19-2(M.2A)	835	410	308	75.1	24.5
19-2(M.2B)	832	422	293	69.4	29.2
25-1(2A)	1075	513	235	45.8	33.4
25-1(2B)	996	476	145	30.5	28.9
25-1(2C)	1010	482	189	39.2	45.2
25-1(3A)	1070	526	254	49.2	31.6
25-1(3B)	1082	532	263	49.4	39.0
25-1(4)	987	471	208	44.2	39.0
25-2(3A-1)	1239	608	338	55.6	24.6
25-2(3A-2)	1094	529	253	47.8	37.1
25-2(3A-3)	1135	562	283	50.4	36.5
28-2(1A)	1406	692	441	63.7	42.7
28-2(2A)	1522	758	496	65.4	31.7
28-2(2B)	1509	752	477	63.4	27.0
28-2(2C)	1595	796	640	80.4	31.5
28-2(2D)	1712	857	688	80.3	26.8
28-2(2E)	1729	868	556	64.1	28.9
28-2(2F)	1760	875	721	82.4	29.4
28-2(2G)	1706	848	697	82.2	34.9
28-2(2H)	1686	832	678	81.5	33.1
28-2(2I)	2015	994	720	72.4	27.9
28-2(2J)	1934	954	708	74.2	28.1
30-2(1B-1)	1363	682	471	69.1	36.2
30-2(1B-2)	1378	701	514	73.3	30.5

Figure 2-31. Experimental Shift Equilibrium Values for Indiana No. 7 Coal

Figure 2-32. Experimental Shift Equilibrium Values for W. Kentucky Coal

Figure 2-33. Comparison of Model Predictions with PDU Data - W. Kentucky Coal

Figure 2-34. Prediction of PDU Performance with W. Kentucky Coal

Figure 2-35. Comparison of Model Predictions with PDU Data - Indiana No. 7 Coal

Figure 2-36. Prediction of PDU Performance with Indiana No. 7 Coal

TABLE 2-15
REACTIVITY DATA ON MIDWESTERN COALS

<u>Feedstock</u>	<u>Reactivity Used For Model Predictions</u>	<u>Reactivity Determined On TGA For Coal</u>
Indiana No. 7	8.0	6.5
Western Kentucky	10.0	17 to 18
Illinois No. 6	10.0	16.0

2.1.3.3 Conclusions

- There is a decrease in specific gasification rate with an increase in temperature and steam/carbon ratio for Western Kentucky and Indiana coals. These results indicate that there is no advantage in operating the gasifier with these feedstocks at temperatures higher than 1800°F.
- The predicted bed volumes are in reasonable agreement with measured bed volumes for the feedstocks considered here.
- The reactivity of Indiana No. 7 coal determined on the TGA is close to that used for the predictions. For Western Kentucky coal, the reactivity used for the predictions is less than the reactivity determined on the TGA indicating possible existence of mass transfer effect (from bubble phase to particulate phase).

2.1.4 Nomenclature

A	cross-sectional area of gasifier
C	reactivity of coal with reference to coke breeze
E	activation energy
F ₁	gasification rate due to C-CO ₂ reaction
F ₂	gasification rate due to C-H ₂ O reaction
F _g	measured gasification rate in the PDU
k ₁	reaction rate constant for C-CO ₂ reaction
k ₂	reaction rate constant for C-H ₂ O reaction
K ₁	equilibrium constant for the oxygen exchange reaction in the case of C-CO ₂ reaction
K ₂	equilibrium constant for the oxygen exchange reaction in the case of C-H ₂ O reaction
K _E	equilibrium constant of the water-gas shift reaction

P_{CO}	partial pressure of CO component in the gas, other partial pressures are denoted similarly
r_1	reaction rate due to C-CO ₂ reaction
r_2	reaction rate due to C-H ₂ O reaction
T	absolute temperature, K
Y_{CO}	mole fraction of CO in the product gas, composition of other components are denoted similarly
ρ_B	fluid-bed density

2.2 ASH-AGGLOMERATOR-SEPARATOR PERFORMANCE

The performance of the agglomerator section in the process has been tied to two distinct phenomena.

- Ash Agglomeration - As the bed of char circulates through the jet formed at the end of the central feed tube, the carbon in the char is consumed by combustion and gasification, leaving particles that are rich in ash.
- Char-Ash Separation - The underlying principle of ash separation is to exploit the difference in the fluidization velocities of bed char and agglomerated ash.

The char-ash separator is an adjunct module, where a recirculating portion of the bed material is separated into light and heavy fractions. The light fraction is returned to the bed and the heavy fraction is removed as ash. Two factors that were considered are:

- The tendency of the feedstock ash to form larger and/or denser agglomerates that are separable from the smaller and/or less dense char particles that are rich in carbon.
- The amount of recycle gas used in the char-ash separator which impacts the separation efficiency.

From an operational standpoint, the ash withdrawal process was demonstrated to be reliable and independently controllable for all coal tests. From a process performance standpoint the agglomerator/separator has showed varying degrees of effectiveness in producing high ash content material (> 80%) in the gasifier discharge. The degree of ash agglomeration for any coal is influenced by the ash content of the fresh coal and the melting properties of the ash. Feedstocks tested in the PDU have been ranked in terms of their ease of agglomerability. It was also possible to create an operating map relating steady state ash content in the bed with that in the withdrawal.

2.2.1 Theoretical Analysis

The KRW PDU agglomerating gasifier performs combustion, gasification, ash agglomeration, and char-ash separation functions with their associated

intricate inter-relationships. Preliminary results of an in-depth analysis of the possible effects of combustion, gasification, and ash agglomeration rates, and performance of the char-ash separator on ash concentration in the materials withdrawn from the char-ash separator are presented herein. Preliminary PDU data are also cited to corroborate the theory developed to date. Factors identified in this analysis will help to interpret the data obtained from the PDU and the phenomena observed in the PDU, to guide the PDU operation, and to provide PDU operational limitations.

Under current PDU operating procedures, samples are collected hourly from the bed and bottom withdrawal, and every four hours from the carryover and recycled fines to determine their respective ash content. The bottom solid withdrawal rate is then adjusted to maintain the overall gasifier ash balance determined from the following ash conservation equation:

$$F_{1C}X_{1C} + F_{1R}X_{1R} = F_2X_2 + F_3X_3 \quad (2-25)$$

where:

- F_{1C} = coal feed rate to the gasifier
- X_{1C} = ash concentration in the coal
- F_{1R} = fines recycle rate to the gasifier
- X_{1R} = ash concentration in the recycled fines
- F_2 = entrainment rate from the gasifier
- X_2 = ash concentration in the entrained fines
- F_3 = bottom withdrawal rate from the char-ash separator
- X_3 = ash concentration in bottom withdrawal

To satisfy Equation (2-25) alone, the gasifier operation is at a quasi-steady-state where the ash alone is balanced. Although nothing is said about the total inventory of the bed material and its ash concentration, Equation (2-25) can be satisfied while the amount of bed material and the ash concentration of the bed material are changing with time.

At true steady-state condition, both ash and carbon should be balanced. The carbon balance can be written as:

$$F_{1C}(1 - X_{1C}) + F_{1R}(1 - X_{1R}) = F_2(1 - X_2) + F_3(1 - X_3) + G \quad (2-26)$$

where G = the rate of carbon consumed by combustion and gasification.

The steady-state bottom withdrawal rate, F_S , and its steady-state ash concentration, X_S , can be obtained from Equations (2-25) and (2-26) to be:

$$F_S = F_{1C} + F_{1R} - F_2 - G \quad (2-27)$$

$$X_S = \frac{F_{1C}X_{1C} + F_{1R}X_{1R} - F_2X_2}{F_{1C} + F_{1R} - F_2 - G} \quad (2-28)$$

Both the steady-state bottom withdrawal rate, F_S , and the steady-state ash concentration in the material withdrawn from the bottom, X_S , are uniquely determined by Equations (2-27) and (2-28) based on the material balance consideration. Thus, unless the desirable bottom withdrawal happens to require an ash concentration corresponding to the steady-state ash concentration, X_S expressed in Equation (2-28), the operation will be inherently unsteady.

If the ash concentration in the bottom withdrawal, X_3 , is different from the steady-state ash concentration, X_S , the rate of bottom withdrawal, F_3 , will have to be at the following rate to maintain the overall ash balance:

$$F_3 = F_{1C} \frac{X_{1C}}{X_3} + F_{1R} \frac{X_{1R}}{X_3} - F_2 \frac{X_2}{X_3} \quad (2-29)$$

Since X_3 is not a steady-state value, there will be a net change in bed weight due primarily to carbon which can be expressed as:

$$\begin{aligned} \frac{dW_1}{dt} &= \frac{d[W_1(1 - X_1)]}{dt} \\ &= F_{1C}(1 - \frac{X_{1C}}{X_3}) + F_{1R}(1 - \frac{X_{1R}}{X_3}) - F_2(1 - \frac{X_2}{X_3}) - G \end{aligned} \quad (2-30)$$

Integrating Equation (2-30) yields:

$$W_1 = W_{10} + [F_{1C}(1 - \frac{X_{1C}}{X_3}) + F_{1R}(1 - \frac{X_{1R}}{X_3}) - F_2(1 - \frac{X_2}{X_3}) - G] t \quad (2-31)$$

Where W_{10} is the bed weight at $t = 0$, i.e., the bed weight at the time the ash balance operating procedure expressed in Equation (2-29) is imposed. The change of ash concentration in the bed, X_1 , can then be calculated from Equation (2-31) to be:

$$X_1 = \frac{X_{10}}{1 + [F_{1C}(1 - \frac{X_{1C}}{X_3}) + F_{1R}(1 - \frac{X_{1R}}{X_3}) - F_2(1 - \frac{X_2}{X_3}) - G] t} \quad (2-32)$$

Thus, depending on the time the ash balance procedure is imposed, the changes in the bed weight and the bed ash concentration will be different and can be calculated from Equations (2-31) and (2-32).

Bed weight changes can also be expressed in terms of the steady-state ash concentration in the bottom withdrawal, X_S , as:

$$\frac{dW_1}{dt} = F_{1C} \frac{X_{1C}(X_3 - X_S)}{X_3 X_S} + F_{1R} \frac{X_{1R}(X_3 - X_S)}{X_3 X_S} - F_2 \frac{X_2(X_3 - X_S)}{X_3 X_S} - (G - G_S) \quad (2-33)$$

For simplicity, assume that $G = G_S$. This is probably a good approximation because, except for the changes in carbon content in the bed, other operating conditions are not changed. The changes of carbon content in the bed is likely to be gradual and represents a small fraction of the total carbon content in the bed. Equation (2-33) can then be simplified to:

$$\frac{dW_1}{dt} = \frac{(X_3 - X_S)}{X_3 X_S} (F_{1C} X_{1C} + F_{1R} X_{1R} - F_2 X_2) \quad (2-34)$$

Since $F_{1C} X_{1C} + F_{1R} X_{1R} > F_2 X_2$

$$\frac{dW_1}{dt} > 0 \text{ and } \frac{dX_1}{dt} < 0 \text{ if } X_3 > X \quad (2-35a)$$

and

$$\frac{dW_1}{dt} < 0 \text{ and } \frac{dX_1}{dt} > 0 \text{ if } X_3 < X \quad (2-35b)$$

Hence, the bed weight and the ash concentration in the bed can either be increasing or decreasing with time after the ash balance operating procedure is imposed, depending on whether the ash concentration in the bottom withdrawal is larger or smaller than the steady state ash concentration as shown in Equations (2-35a and 2-35b). The steady-state ash concentration in the bottom withdrawal, X_S , depends on the operating conditions and the reactivity of the feedstock, while the ash concentration in the actual bottom withdrawal, X_3 , depends on the ash agglomeration characteristics in the bed and the char-ash separation efficiency in the char-ash separator.

From this analysis, it is theorized that the gasifier can be operated at the true steady-state only when the ash concentration in the bottom withdrawal is equal to the steady-state ash concentration; i.e., $X_3 = X_S$. There are three possible steady-state operational modes:

Mode 1: $X_3 = X_1 = X_S$ and no ash agglomeration in the bed.

This mode of operation will occur when there is no agglomeration of ash in the bed and thus no particle separation is possible in the char-ash separator. The operation reaches the steady state when $X_3 = X_1 = X_S$. The biggest single disadvantage of this mode of operation is that the bed ash concentration is similar to that in the bottom withdrawal. To obtain a high ash concentration in the bottom withdrawal, the bed ash concentration has to be similarly high. More operational problems such as deposit formation and slagging are expected with a high ash bed. High ash concentration from the bottom withdrawal for this mode of operation is feasible only for coals with high reactivities such as lignites because only coals with high reactivities will have a high steady-state ash concentration, X_S (see Equation (2-28)). In fact, certain lignite tests in the PDU confirm this analysis.

Mode 2: $X_3 = X_S$ and the ash agglomeration rate is less than the ash withdrawal rate.

A necessary condition for the gasifier to operate at a steady-state under this operational mode is to satisfy the following equation:

$$F_{1C}(1 - \frac{X_{1C}}{X_C}) + F_{1R}(1 - \frac{X_{1R}}{X_C}) - F_2(1 - \frac{X_2}{X_C}) - G + Af(\frac{X_a}{X_C} - 1) = 0 \quad (2-36)$$

Depending on the ash agglomeration rate, A , and the char-ash separation efficiency represented by f , Equation (2-36) may not be satisfied and the gasifier may not be able to be operated at a true steady-state under this mode. At steady-state, the ash concentration in the bottom withdrawal can be expressed as:

$$X_3 = X_S = \frac{A}{F_3} \times a + (1 - \frac{A}{F_3})X_C \quad \text{for } f = 1 \quad (2-37)$$

Under this mode of operation, the ash concentration in the bottom withdrawal is always less than that in the ash agglomerates based on equation (2-37). At the limiting case where the ash agglomeration rate is zero; i.e., $A = 0$, Equation (2-37) reduces to $X_3 = X_S = X_1$, similar to the operation in Mode 1.

Mode 3: $X_3 = X_S = X_a$; $A = F_3$; and $f = 1$

At the other limit where the ash agglomeration rate, A , is equal to the bottom withdrawal rate, F_3 , and the char-ash separation is complete, Equation (2-37) becomes $X_3 = X_S = X_a$, where X_a is the ash concentration of the agglomerates. This is the most favorable steady-state operational mode because the ash concentration in the bottom withdrawal has the highest ash content among the three steady-state operational modes, while the ash concentration in the bed can be maintained at a relatively low level because there is no net accumulation of ash agglomerates in the bed.

Based on the current PDU operation philosophy, the theoretical analysis indicated that most of the PDU tests conducted so far were probably at quasi-steady-state where ash alone was balanced. True steady-state operations require that the ash concentration in the bottom withdrawal, X_3 , be equal to the steady-state ash concentration, X_S , dictated by the kinetics of coal combustion, gasification and other process variables as expressed in Equation (2-37). There are only three steady-state operational modes possible depending on coupling effect of ash agglomeration rate and char-ash separation efficiency. To obtain high ash concentration from the bottom withdrawal, the steady-state ash concentration has to be high. This is generally true for coals of high reactivities such as lignites. However, because of high reactivity, lignites tend to be gasified at a lower temperature which may not be conducive for formation of ash agglomerates. Thus, the steady-state operational mode for coals of high reactivities such as lignites tend to be Mode 1 where $X_3 = X_S = X_1$ at steady state.

For coals of low reactivities, such as coke breeze or Pittsburgh seam coals, the bed has to be operated at temperatures substantially higher so that the steady-state ash concentration will be high based on Equation (2-28). The higher temperatures will promote ash agglomeration and the steady-state operational mode tends to be Mode 3 where $X_3 = X_S = X_a$ at steady-state. Though this mode of operation is the most favorable one, the ash agglomeration rate should be very carefully controlled. If the ash agglomeration rate, A , is higher than the steady state ash withdrawal rate, F_S , expressed in Equation (2-27), ash agglomerates can accumulate in the bed quickly and render additional operational problems such as deposit formation and slagging.

For coals of medium reactivities such as Pittsburgh seam coal, the steady-state operational modes can be either Mode 2 or Mode 3. For operational Mode 2, Equation (2-36) has to be satisfied to attain the true steady-state and for operational Mode 3, the ash agglomeration rate has to exactly equal the ash withdrawal rate and char-ash separation should be complete. Those exact conditions are relatively difficult to attain during actual operation. Without precise knowledge of ash agglomeration rate and combustion and gasification kinetics, the true steady-state can be attained only through an empirical approach.

The current analysis provides a preliminary theoretical basis for interpretation of phenomena observed in the PDU. Further analysis can be carried out in conjunction with PDU data analysis. A sound operational procedure may thus evolve. Critical experiments in the PDU can also be recommended to test the theory.

2.2.2 Operational Analysis

The ash concentration in the gasifier bottom withdrawal stream affects the carbon conversion efficiency of the process; therefore, a high ash concentration is desirable. However, this has not always been achievable in the PDU. For example, steady state bottom withdrawal ash concentrations averaged 58 percent for 1982 PDU tests (excluding North Dakota lignite testing). However, ash withdrawal concentrations of 90 percent or more were consistently achieved during the first four PDU tests of 1983. The purpose of this discussion is to identify those factors which influence the ash withdrawal concentration and those which contributed to the higher bottom withdrawal-ash concentration during 1983 PDU testing. This analysis will be based on observations and data generated during operation of the PDU.

2.2.2.1 Process Parameters

Basically the gasifier bottom withdrawal-ash concentration is controlled by three factors: 1) ash agglomeration rate, 2) ash balance, and 3) char-ash separation. All three must be in proper adjustment to obtain a high ash concentration in the gasifier bottom withdrawal stream. If ash is not agglomerating at a rate equal to or greater than the annulus withdrawal rate, then a high ash concentration in the bottom withdrawal stream will not be achieved. The same is true if the annulus withdrawal rate is greater than the rate required to achieve a high ash withdrawal concentration. Once produced, ash agglomerates must be separated from char particles to achieve a high bottom withdrawal-ash concentration. This requires proper adjustment of annulus velocities.

2.2.2.1.1 Ash Agglomeration Rate

Feedstock Size Distribution

As previously noted, high bottom withdrawal-ash concentrations were achieved during 1983 PDU testing relative to 1982 PDU tests. The feedstock particle size was also reduced during 1983 PDU testing to eliminate feedline plugs. PDU data were reviewed to determine if the difference in particle size distribution was responsible for the higher bottom withdrawal-ash concentration. Feedstock size distribution for two different Wyoming coal PDU tests is plotted in Figure 2-37. As can be seen, the size distributions are very similar. However, the bottom withdrawal-ash concentrations are very different. The relatively high ash concentration achieved during TP-034-1 indicates that feedstock particle

Figure 2-37. Feedstock Size Distribution for Wyoming Coal

size was not the cause of the poor bottom withdrawal-ash concentration during TP-028-1. This is not to say that the feedstock particle size does not influence ash agglomeration, but that the higher ash withdrawal concentrations achieved during 1983 PDU tests should not be attributed to a change in the feedstock particle size distribution.

Temperature

Until recently gasifier temperatures were measured only at the wall of the gasifier. It was recognized that these measurements did not reflect actual internal gasifier temperatures, especially in the jet region. A Non-Slagging Traversing Thermocouple Probe (NSTTP) was developed to measure the actual radial temperature profile of the gasifier. One NSTTP probe was located ~30 inches above the oxidant tube. This provided a gasifier radial temperature profile in the jet region.

Ash agglomeration requires partial melting or softening of the feedstock ash or of some component of the ash. Ash fusion temperatures have been used in the coal-fired boiler industry as indicators of the fouling or slagging tendency of a particular coal. These temperatures indicate the onset or degree of melting of the ash. The lowest of the ash fusion temperatures is the initial deformation temperature. The initial deformation temperature is the temperature at which a prepared standard cone of ash just begins to fuse or show evidence of deformation at the top of the cone as it is being heated in a standard test furnace. Measured temperature in the gasifier jet region, using the NSTTP, were compared to the ash initial deformation temperature of the feedstock to determine if a relationship exists between the two. These data are shown on Table 2-16. The initial deformation temperatures (IDT), under oxidizing and reducing conditions, are listed. Though the gasifier atmosphere is predominantly reducing, the IDT under oxidizing conditions is believed more pertinent to ash agglomeration because ash agglomeration is believed to occur in or around the jet. As shown on Table 2-16, the ash IDT under oxidizing conditions is significantly above the maximum temperature measured in the gasifier jet region. However, ash agglomeration did occur in each of these tests. In fact, greater than 90 percent ash was achieved in the bottom withdrawal during TP-034-3 and TP-035-1 even though the ash initial deformation temperature is above the jet temperature. Thus, the ash IDT cannot be used as a direct indication of the ability to form ash agglomerates.

Currently there is no reliable method of estimating the agglomerability of a feedstock prior to PDU testing. An attempt has been made to qualitatively rank the agglomerability of feedstocks previously gasified in the PDU. The intent is to correlate this ranking with some property of the feedstock. Some arbitrary assumptions were made in developing this ranking. Therefore, the ranking may not be totally accurate, but is believed to be generally correct.

TABLE 2-16

GASIFIER JET TEMPERATURE VS. ASH INITIAL DEFORMATION TEMPERATURE

PDU Test	Feedstock	*Temperature at Wall (°F)	*Maximum Radial Temperature (°F)	Ash Initial Deformation Temperature (°F)	
				Oxidizing	Reducing
TP-033-2	Wyoming	1620	2050	2200	2120
TP-033-3	W. Kentucky	1835	2180	2370	2120
TP-034-3	N. Dakota Lignite	1575	1915	2380	2370
TP-035-1	Wyoming	1550	1725	2200	2120

*Measured approximately 30 inches above tip of oxidant tube.

TABLE 2-17

ASH AGGLOMERABILITY INDEX

Coal Type	PDU Test	Steady State	Withdrawal Ash %	Bed Ash %	*Agglomeration Index	Ash Initial Deformation Temp (°F)	
						Oxidizing	Reducing
N. Dakota	TP 031 3	1C	99.0	23.0	426	2130	1960
Wyoming	TP-034-1	2C	85.2	25.3	290	2200	2120
Pittsburgh (Federal Mine)	TP-034-2	2B	89.9	31.1	261	2350	2080
W. Kentucky	TP-03303	2A	77.9	28.9	210	2370	2120
Δ Pittsburgh (Westland Mine)	TP-027-3	2C	86.5	45.5	164	2560	2400
Indiana No. 7	TP-028-2	2C	68.8	31.5	150	2380	2330
RSA	TP-0302	6A-1	89.2	55.9	143	2340	2420
Brookville	TP-033-1	2A	77.0	43.4	139	2350	2030
Ohio No. 9	TP-027-2	1B	67.0	40.4	114	2590	2380
U. Freeport	TP-032-1	1A-1	58.4	32.1	105	2550	2230
W. Virginia	TP-030-3	2A	51.7	27.4	98	2670	+2800

Δ Initial Deformation Temperatures based on TP-028-1 Pittsburgh Coal Sample

* Agglomeration Index = $\frac{(\text{Withdrawal Ash \%})}{(\text{Bed Ash \%})}$

Note: Data are confined to those steady states in which the sparger flow was >500 lb/hr, the booster flow >300 lb/hr and the annulus temperature interface was at TE-504-9 or above.

The process by which the agglomerability ranking was developed consisted of reviewing data reports for all PDU tests since the inception of the single-stage gasifier. Data for the steady state in which the highest bottom withdrawal ash concentration was achieved for a particular coal were extracted. The assumption was made that this steady state represented an optimization of the bottom withdrawal-ash concentration based on a variety of gasifier and annulus operating conditions. Although this assumption may not be valid in every case it is probably true in most. In addition, specific operational conditions had to be met to qualify the data for consideration; that is, the annulus booster flow and the sparger flow had to exceed a minimum mass flow rate, and the annulus temperature interface elevation had to be higher than a certain elevation.

These criteria are believed to represent a condition which provides the most favorable operating conditions for producing a high ash concentration in the bottom withdrawal stream. Also this provides a comparison between feedstocks at similar operating conditions. The bed ash concentration was also factored into this analysis. A 60 percent ash withdrawal concentration is much easier to achieve if the bed ash concentration is 50 percent than if it is only 20 percent. Therefore, an agglomeration index was arbitrarily defined as follows:

$$\text{Agglomeration Index} = \left(\frac{\text{Bottom Ash Concentration}}{\text{Bed Ash Concentration}} \right) \left(\frac{\text{Bottom Ash Concentration}}{\text{Bed Ash Concentration}} \right)$$

The results of this analysis are shown on Table 2-17.

There appears to be some correlation between a high agglomeration index and a low initial deformation temperature (oxidizing) although some inconsistency is present. The two coals at the extremes of this ranking have both been observed to correspond with this ranking very well. With North Dakota lignite (highest agglomeration index) relatively high bottom withdrawal-ash concentrations were achieved even during the initial operation of the PDU with this feedstock. Bottom ash-withdrawal concentrations exceeded 90 percent during five of the six steady states of TP-034-3 with North Dakota lignite. In contrast, the highest steady state ash withdrawal concentration achieved during TP-030-3 with West Virginia coal was 52 percent even though operational parameters were varied in an attempt to increase this concentration.

The purpose of Table 2-17 is to suggest that there may be a relationship between the initial ash deformation temperature and the ash agglomeration temperature. This is not to suggest, however, that the initial ash

deformation temperature is the temperature at which agglomeration occurs. It has been shown in Table 2-16 that gasifier temperatures even in the jet region are significantly below the initial ash deformation temperature.

The initial ash deformation temperature in the feed coal and bottom withdrawal was compared to determine if a chemical transformation might be resulting in an ash with an initial deformation temperature below gasifier operating temperatures. The results for Pittsburgh seam and Wyoming coal are shown on Table 2-18. There is no significant change in the initial ash deformation or softening temperatures during gasification and combustion with these feedstocks.

Studies were conducted to determine the ash sintering temperature for feedstocks processed in the PDU. This is the temperature at which an ash particle begins to shrink. Coal feedstock samples were ashed at low temperature to prevent distortion of the ash. This ash was then heated under a hot stage microscope. The transformation of the ash was videotaped as the sample was heated. The temperature at which shrinkage was observed was then noted. Results of these tests are shown on Table 2-19. Initial sintering temperatures were compared to the agglomerability index presented in Table 2-17. The outer extremes of both tables are in agreement (i.e., North Dakota lignite and West Virginia coal). There are similarities and differences between these limits. A comparison of the sintering temperatures with the jet region temperatures presented on Table 2-16 shows that the jet temperature is higher than the sintering temperature. Therefore, if the sintering temperature is indeed an indication of the agglomeration temperature then agglomeration should occur at these conditions. In these cases, char-ash separation efficiency would control the bottom withdrawal-ash concentration.

Ash Composition Versus Ash Agglomeration Rate

Chemical composition of the ash of each feedstock processed in the PDU was reviewed to determine if a correlation existed between a particular ash component and the agglomerability of that feedstock. A tabulation of ash compositions for each feedstock processed in the PDU is shown on Table 2-20. No correlation was found between the agglomerability of that feedstock (as shown on Table 2-17) and its ash components. North Dakota lignite however has a relatively high sodium content. Since it is the only coal tested in the PDU with a high sodium content no conclusions can be made on the effect of sodium on the agglomerability of a feedstock based solely on the PDU data base. However, North Dakota lignite has the highest ash agglomeration index value shown on Table 2-17. Agglomeration and slaying have been studied extensively for coal fired boilers and combustors. Research indicates that "coals which contain more than 3 to 5 percent Na_2O in the ash are prime candidates for agglomeration" and "sodium-rich inorganic constituents may promote partial melting of ash and bed materials by acting as a flux to produce low melting temperature

TABLE 2-18
PRE AND POST GASIFICATION ASH FUSION TEMPERATURES

Sample	PDU Test	Initial Deformation Temp (°F)		Ash Softening Temp (°F)	
		Oxidizing	Reducing	Oxidizing	Reducing
Wyoming Coal	TP-034-1	2200	2120	2260	2170
Wyoming Withdrawal Ash	TP-034-1	2310	2220	2370	2280
Pittsburgh Coal	TP-034-2	2350	2080	2460	2160
Pittsburgh Withdrawal Ash	TP-034-2	2350	2060	2460	2140

TABLE 2-19
ASH SINTERING TEMPERATURES

<u>Feedstock Ash</u>	<u>Sintering Temperature (°F)</u>
North Dakota Lignite	1575
Pittsburgh No. 8	1650
Montana Rosebud	1655
Western Kentucky	1660
Illinois No. 6	1745
Upper Freeport	1750
Texas Lignite	1770
Ohio No. 9	1810
RSA	1815
Indiana No. 7	1855
Wyoming	1940
French	2020
West Virginia	2290

TABLE 2-20
PDU FEEDSTOCK ASH COMPOSITIONS

Coal Type	% Ash In Coal	% of Ash (Oxide Basis)						
		SiO ₂	Al ₂ O ₃	Fe ₂ O ₃	CaO	MgO	Na ₂ O	K ₂ O
Pittsburgh (Federal Mine)	11.2	42.0	20.0	17.3	7.9	1.1	1.4	1.3
Pittsburgh (Westland Mine)	7.4	44.4	24.2	17.3	0.13	0.44	0.46	1.69
W. Kentucky	8.3	50.3	20.6	18.9	2.0	0.8	0.7	3.2
Indiana #7	9.7	54.4	24.6	6.1	0.10	0.77	0.13	2.81
Brookville	13.2	38.9	22.6	23.5	0.33	0.21	1.01	1.52
Ohio #9	13.6	54.6	24.0	17.1	1.8	0.6	0.3	1.9
Upper Freeport	14.2	48.4	20.8	23.3	0.05	0.54	0.28	2.5
Illinois #6	11.8	45.0	18.8	12.8	3.8	1.4	1.4	2.2
West Virginia	9.0	54.5	30.3	4.6	0.08	0.32	0.43	1.7
RSA	19.1	47.7	24.8	3.2	1.5	2.0	0.4	0.5
Wyoming	8.7	37.7	19.9	5.0	12.9	4.0	0.9	0.6
Texas Lignite (Monticello Mine)	25.4	32.5	19.3	5.3	22.2	3.2	-	-
Texas Lignite (Big Brown Mine)	14.7	46.5	11.5	7.5	8.0	2.3	0.02	0.5
N. Dakota Lignite	10.7	34.5	14.0	9.1	5.6	0.9	5.4	1.5

phases."(1) These quotes were made based on observations of ash agglomeration under conditions different from those in the PDU. However, the ease with which agglomeration is achieved with North Dakota lignite indicates that sodium may promote agglomeration in the PDU also.

Ash agglomeration rate does not appear to limit the bottom withdrawal-ash concentration achieved in the PDU at the typical PDU operating conditions for a particular feedstock type. High bottom withdrawal-ash concentrations have been achieved with all types of coals (bituminous, sub-bituminous, and lignite) during 1983 PDU testing. Annulus operating conditions appear to be more important than any inherent tendency of the ash to agglomerate. However, this may not be true if the gasifier is operated at conditions significantly different from those used in the past (particularly temperature).

2.2.2.1.2 Ash Balance

To obtain a high ash concentration in the bottom withdrawal, the ash withdrawal rate must be adjusted to maintain ash balance at a high ash withdrawal concentration. For example, if 100 lb/hr of ash is fed to the gasifier (1000 lb/hr coal at 10 percent ash) and 48 lb/hr of fines is lost to the water system (120 lb/hr at 40 percent ash) then the ash withdrawal rate must be 58 lb/hr to achieve 90 percent ash in the bottom withdrawal.

A comparison was made of ash withdrawal rates for 1983 Wyoming coal PDU tests versus pre-1983 Wyoming coal tests to determine whether high fines loss rates caused by cyclone inefficiencies prior to 1983 resulted in such low bottom withdrawal rates as to exceed lower operability limit of the bottom ash-withdrawal starwheel feeder. For example, if the ash input to the gasifier was 100 lb/hr and the ash lost to the water system was 90 lb/hr (225 lb/hr fines at 40 percent ash) then the required ash withdrawal rate to achieve ash balance at 90 percent ash would be 11 lb/hr. This is below the lower operability limit of the PDU bottom ash-withdrawal system. This limitation may have affected the ability to obtain a high bottom withdrawal-ash concentration in a few isolated cases. For example, the bottom withdrawal rate to maintain ash balance at 90 percent ash for steady state 1B-3 of TP-033-2 was 16 lb/hr. The PDU ash withdrawal system cannot accurately withdraw ash at such a low rate. However, in most cases, the lower operability limit of the bottom withdrawal system was not exceeded. Seven Wyoming coal PDU tests were reviewed comprising 23 steady states. Only during one of the 23 steady states was the required bottoms withdrawal rate to obtain 90 percent ash below 35 lb/hr. Of those steady states with a bottom withdrawal-ash concentration of less than 70 percent the calculated withdrawal rate to achieve 90 percent ash averaged 69 lb/hr.

An ash balance requires an accurate estimation of the ash lost to the water system. For a 100 lb/hr ash input and a 48 lb/hr ash loss to the water

(1) "AFBC Bed Material Performance With Low Rank Coals" presented at 12th Biennial Lignite Symposium, May 18-19, 1983.

system, the bottom withdrawal rate must be 58 lb/hr to obtain a 90 percent ash concentration. If the actual loss to the water system was 58 lb/hr, an error of 20 percent, a 58 lb/hr bottoms withdrawal rate would result in a bottom withdrawal-ash concentration of 72 percent. Recent improvements to the isokinetic measurement system have improved the accuracy and reliability of ash loss measurements. However, prior to 1983 there was a significant discrepancy in many cases between isokinetic measurements of ash lost to the water system and loss rates determined by ash balance. Therefore, some cases of low bottom withdrawal-ash concentrations may have been due to an improper ash balance resulting from inaccurate estimations of the ash lost to the water system during test operation.

Annulus Temperature Interface

During normal PDU operation, a sharp temperature gradient exists in the gasifier annulus. This is referred to as the temperature interface. This interface is a result of the heat balance produced by solids and gas flow into and out of the annulus and zones of fluidized and non-fluidized char-ash solids. The temperature interface is normally maintained at a relatively fixed elevation by adjustment of the ash withdrawal rate and the annulus recycle gas flows (booster and sparger). The effect the location of the temperature interface has on the bottom withdrawal-ash concentration was reviewed for six Wyoming coal PDU tests. In general, with all other operating parameters being equal, the higher the temperature interface elevation the higher the bottoms withdrawal ash concentration.

2.2.2.1.3 Char-Ash Separation

Annulus Velocities

Char is separated from ash in the gasifier annulus based on the concept of minimum fluidization velocity. The annulus velocity is adjusted such that small dense particles defluidize while large less-dense char particles remain fluidized. The velocity is maintained below the minimum fluidization velocity of the ash but above the minimum fluidization velocity of the char.

As noted previously, higher bottom withdrawal-ash concentrations were achieved during 1983 PDU testing in comparison to prior years. There has also been a trend toward higher annulus flow (which translates to higher annulus velocities). For the five PDU tests conducted with Wyoming coal prior to 1983 the coal annulus flow (booster plus sparger) averaged 672 lb/hr. The corresponding average ash withdrawal concentration was 49 percent. The coal annulus flow for 1983 PDU testing with Wyoming coal (TP-034-1 and TP-035-1) was 987 lb/hr with a corresponding average ash withdrawal concentration of 75 percent. There were two steady states during TP-035-1 which included full fines recycle and the same temperature interface elevation. When the coal annulus flow was 1215 lb/hr, the bottom withdrawal-ash concentration averaged 84 percent. When the total annulus

flow was 1383 lb/hr, the bottom withdrawal-ash concentration averaged 93 percent. Thus, annulus velocity rather than annulus flow is influencing the bottoms withdrawal ash concentration.

Fines Recycle

The use of a non-mechanical valve for hot fines recycle was successfully demonstrated during the first PDU test of 1983 (TP-034-1). This also coincided with the first PDU test in which a high bottom withdrawal-ash concentration was achieved with Wyoming coal. Subsequently high bottom withdrawal-ash concentrations and successful non-mechanical valve operation were attained throughout 1983 PDU testing. Set points were designed for TP-035-1 to determine if the hot fines recycle generated by the non-mechanical valve was responsible for the higher bottom withdrawal-ash concentrations. During set point 1 fines were recycled via the non-mechanical valve and secondary injection device (SID) as normally done. However, during set point 4 there was no fines recycle. During set point 1 the ash concentration in the discharge was in the range of 84 to 93 percent, whereas in set point 4 the ash content ranged between 70 and 75 percent. These results showed that the non-mechanical valve fluidized dipleg permitted gasifier operation at higher cyclone efficiencies, thereby improving the agglomerator performance.

2.2.2.2 Conclusions

- The annulus temperature interface elevation has a significant impact on the bottom withdrawal-ash concentration.
- Higher annulus velocities during 1983 PDU testing in comparison to historical levels have contributed to higher bottom withdrawal-ash concentrations during 1983 PDU testing.
- Higher bottom withdrawal-ash concentrations were achieved with the hot fines recycle apparatus which provided higher recycle efficiencies.
- At typical PDU operating conditions for a particular feedstock the bottom withdrawal-ash concentration is not constrained by the ash agglomeration rate.
- Gasifier temperatures in the jet region are below initial ash deformation temperatures. Therefore initial ash deformation temperature cannot be used directly as an indication of the agglomerability of a particular feedstock. However, there may be a relationship between the initial sintering temperature (measured on a hot stage microscope) and the actual ash agglomeration temperature.

2.3 FEEDSTOCK CHARACTERIZATION

The purpose of this task was to characterize the chemical and physical properties of the feedstocks processed in the PDU. This includes ultimate and proximate analyses, ash fusion properties, ash elemental analysis, sulfur form, chlorine/fluorine content, hardgrove grindability index, free swelling index, and reactivity. The particle size distribution produced from the coal crusher-dryer system was also included with this task. These data are then used to project feedstock performance in the PDU, or as an aid in other PDU studies (e.g., ash agglomeration and cyclone deposit control).

2.3.1 Background

The following issues are important when characterizing the feedstocks processed in the PDU.

- A. Coal Source It is important to identify the source of coal used in the PDU. Chemical and physical properties of a particular classification of coal vary significantly from mine to mine. Also, information on a particular coal can be obtained from literature if its origin is known.
- B. Coal Sizing Requirements for Gasifier Two different coal top sizes have been used in the PDU, 1/4-inch and 6 mesh. These sizes can be controlled through adjustments made to the coal crusher-dryer system. Also, pneumatic feed line performance and reliability are believed to be related to particle size distribution.
- C. Coal Composition Chemical and physical properties of feedstocks are needed to project the performance of that feedstock in the PDU. This information is also used to aid other gasifier studies.
- D. Fines Generation During Crushing Operation Fines generated during the coal crushing operation must be kept to a minimum. Cyclone inefficiencies will result in some of these fines being carried to the product gas cooling system and lost. The lost carbon in these fines reduces the overall carbon conversion efficiency.

Table 2-21 lists the feedstocks processed in the PDU, the mode and hot hours of operation, and the dates the tests were conducted.

A new coal crushing and drying system was added to the PDU and was first used for coal preparation in August 1980 (TP-027-4) to size classify the coal feedstock and dry the coal to a surface moisture content of less than 5 percent. Drying is necessary so reliable transfer of the feedstock can be obtained in the material handling train. Fines generation data from the

TABLE 2-21
PDU COAL FEEDSTOCK

Coal Feedstock	Test No.	Month/ Year	Hot Hours of Coal	Mode
Pittsburgh No. 8 Adamsburg Mine (R-O-M Washed)	TP-016	4-5/78	113	Air
Pittsburgh No. 8 Champion Mine(R-O-M Washed)	TP-013-3	12/77	16	Air
	TP-017	5/78	161	Air
	TP-018-2	7/78	45-1/2	O ₂
	TP-018-3	8/78	69-1/2	O ₂
	TP-018-4	9/78	47-1/2	O ₂
	TP-018-5	9-10/78	85	O ₂
	TP-020-2	5/79	164-1/2	Air
	TP-022-1	6/79	53	O ₂
	TP-022-2	6-7/79	57	O ₂
	TP-023-3	12/79	50-1/2	O ₂
	TP-023-4	3/80	134-3/4	O ₂
	TP-027-2	5/80	64-1/2	O ₂
	TP-027-3	6-7/80	147-1/2	O ₂
	TP-028-1	6/81	33-1/2	O ₂
	TP-034-2	6-7/83	435	O ₂
Pittsburgh No. 8 Federal Mine	TP-035-2	12/83	226-1/2	O ₂
Ohio No. 9 Dannock Mine	TP-023-1	7-8/79	30	O ₂
	TP-023-2	8/79	28	O ₂
	TP-025-2	11/79	14-1/2	Air
	TP-027-1*	4/80	69	O ₂
	TP-027-2	5/80	62-1/2	O ₂
W. Kentucky No. 9 Hamilton Mine (R-O-M Washed)	TP-019-2	11/78	45	O ₂
	TP-019-3	12/78	56	O ₂
	TP-027-4	8/80	66-1/2	O ₂
	TP-032-2	5/82	95-1/2	O ₂
	TP-033-3	12/82	94 1/3	Air & O ₂
Indiana No. 7 Minnehaha Mine (R-O-M Washed)	TP-013-3	12/77	27	Air
	TP-019-1	10/78	37-1/4	O ₂
	TP-019-2	11/78	95-1/2	O ₂
	TP-030-2	5/81	61-1/2	O ₂
	TP-028-2	7/81	144	O ₂ & Mixed
Indiana No. 7 Sycamore Mine (R-O-M Washed)	TP-025-1	9/79	146	Air
	TP-025-2	11/79	61-1/4	Air
	TP-027-1*	4/80	69	O ₂
Rosebud Coal Montana (R-O-M Unwashed) Subbituminous "B"	TP-019-1	10/78	63-1/2	O ₂
Texas Lignite (R-O-M Unwashed)	TP-023-4	3/80	11-1/2	O ₂
Texas Lignite Monticello Mine (R-O-M Unwashed)	TP-027-3	6-7/80	47	O ₂
	TP-027-5	10/80	85	O ₂
Wyoming (Wyodak Mine) (R-O-M Unwashed)	TP-013-3	12/77	22	Air
	TP-015	4/78	36	Air
	TP-027-4	8/80	50-1/2	O ₂
	TP-027-5	10/80	24	O ₂
	TP-029-1/ -030-1	4/81	60	O ₂
	TP-028-1	6/81	39	O ₂
	TP-033-2	10-11/82	75-1/2	O ₂
	TP-034-1	3-4/83	148-1/4	O ₂
	TP-035-1	9-10/83	291-1/4	O ₂

* Indiana No. 7 Ohio No. 9 blended during drying operation.

TABLE 2-21 (Continued)

PDU COAL FEEDSTOCK

<u>Coal Feedstock</u>	<u>Test No.</u>	<u>Month/ Year</u>	<u>Hot Hours of Coal</u>	<u>Mode</u>
Illinois No. 6	TP-031-1	12/80	12	0 ₂
Delta Mine	TP-029-1/	4/81	112-1/2	0 ₂
(R-U-M Unwashed)	-030-1			
Upper Freeport Seam	TP-023-3	12/79	22	0 ₂
(R-O-M Unwashed)				
Upper Freeport	TP-032-1	3-5/82	97-1/3	0 ₂
E-Seam, Elton Mine	TP-032-2	5/82	53-1/2	0 ₂
(Unwashed)				
RSA Coal	TP-031-1	12/80	97	0 ₂
Bosjesspruit Colliery	TP-030-2	5/81	53	0 ₂
(R-O-M Unwashed)	TP-031-2	10/81	46	0 ₂
Texas Lignite	TP-028-3	9/81	112-1/2	0 ₂
Big Brown Mine				
(R-O-M Unwashed)				
West Virginia	TP-030-3	11/81	172-1/2	0 ₂
Big Mountain Mine				
(R-O-M Washed)				
Arco Coke	TP-031-3	12/81	88-1/4	0 ₂
Watson Refinery				
Chevron Coke	TP-031-3	12/81	61-1/4	0 ₂
El Secundo, Calif.				
N. Dakota Lignite	TP-033-1	7-8/82	67-1/2	0 ₂
Indianhead Mine	TP-034-3	8/83	216-2/3	0 ₂
(R-O-M Unwashed)				
Brookville	TP-033-1	7-8/82	103-1/4	0 ₂
Ace Drilling Co.				

crusher-dryer mill were obtained for W. Kentucky, Illinois, West Virginia, Wyoming, Texas Lignite, RSA, Indiana, and Pittsburgh coals, as well as for Chevron and ARCO petroleum cokes.

2.3.2 Experimental Results

The chemical and physical properties of all feedstocks tested in the PDU are shown in Table 2-22. Most of these results were obtained from sample analyses performed by the PDU Chemistry Lab or an outside testing laboratory. A few data points were obtained from the Penn State Coal Data Bank. Particle size distribution of each feedstock as it was fed to the gasifier is shown in Table 2-23. Some of the variation is due to using a 1/4-inch or a 6 mesh top screen on the coal crusher-dryer.

2.3.2.1 Analysis of Results

- A. Coal Composition Nine different coals and two petroleum cokes were tested in the PDU during the 1980-1981 test period. The wide range of properties of these feedstocks demonstrated the versatility of the gasifier. Ash content of the feedstock varied from 0.2 percent with ARCO petroleum coke to almost 23 percent with Texas Lignite (Monticello Mine). Non-caking coals, such as Wyoming Sub-C, and highly caking coals, such as Pittsburgh No. 8, were successfully processed. High sulfur content feedstocks, such as ARCO petroleum coke (2.5 percent), were run along with low sulfur content coals, such as RSA (0.50 percent). Both highly reactive coals (e.g., Lignite) and coals of relatively low reactivity (e.g., Pittsburgh No. 8) were also run.
- B. Particle Size Distribution Particle size distribution for four coals crushed and dried with the same top screen on the coal crusher-dryer is shown in Figure 2-38. These data have been averaged from belt weigher samples taken as the coal was being conveyed to the coal feed lockhoppers. The data indicate that a fairly consistent particle size distribution is being obtained for material fed to the gasifier.

2.3.3 Conclusions

- The PDU is capable of processing coals with a broad range of chemical and physical properties.
- The particle size distribution of the feedstock to the gasifier has been fairly consistent.
- Although PDU tests were conducted with feedstocks processed through two top screen sizes (1/4 inch and 6 mesh), the average particle size was not significantly varied. Further testing with

TABLE 2-22
FEEDSTOCK PROPERTIES

Proximate, wt.% as received	Pittsburgh No. 8 Adamsburg Mine	Pittsburgh No. 8 Westland Mine	Upper Freeport Renton Mine	Indiana No. 7 Minnehaha Mine	Indiana No. 7 Sycamore Mine	Upper Freeport Elton Mine
Volatile Matter	34.82	36.03	36.30	30.52	28.85	16.29
Fixed Carbon	53.68	54.07	52.78	49.92	49.55	63.94
Moisture	2.05	2.51	5.56	9.42	11.14	5.90
Ash	9.45	7.39	10.92	10.14	10.46	13.87
Ultimate, wt.%						
Carbon	74.81	74.44	74.38	64.83	65.54	71.51
Hydrogen	5.42	5.25	5.03	5.58	5.37	4.40
Oxygen (by difference)	7.29	9.59	6.33	17.31	17.36	6.21
Nitrogen	1.54	1.51	1.42	1.52	1.52	1.36
Sulfur	1.49	1.82	1.92	0.62	0.75	2.65
Proximate, wt.% as received	Wyoming Sub-C Wyodak Mine	W. Kentucky No. 9 Hamilton Mine	Montana Sub-B Rosebud	Texas Lignite Monticello Mine	Texas Lignite Big Brown Mine	N. Dakota Lignite Indian Head Mine
Volatile Matter	33.37	34.94	43.55	31.93	35.85	33.67
Fixed Carbon	40.18	51.53	26.16	30.67	29.26	37.82
Moisture	18.28	4.98	22.17	14.85	18.31	17.76
Ash	8.17	8.55	8.09	22.56	16.58	10.74
Ultimate, wt.%						
Carbon	54.47	69.17	53.98	44.96	47.93	52.66
Hydrogen	5.83	5.36	6.01	4.84	5.60	3.77
Oxygen (by difference)	30.24	12.74	30.44	25.80	28.07	31.13
Nitrogen	0.82	1.56	0.85	0.88	0.88	0.99
Sulfur	0.47	2.63	0.63	0.67	0.94	0.70
Proximate, wt.% as received	Illinois No. 6 Delta Mine	RSA Bosjesspruit	West Virginia Big Mountain Mine	Ohio No. 9 Bannock Mine	Brookville Ace Drilling	Pittsburgh No. 8 Federal No. 2 Mine
Volatile Matter	32.89	22.61	31.54	30.87	22.77	35.35
Fixed Carbon	47.80	54.63	57.96	51.87	62.10	48.27
Moisture	7.04	4.35	1.38	3.54	1.91	2.97
Ash	12.29	18.41	9.12	13.72	13.21	13.42
Ultimate, wt.%						
Carbon	64.66	61.56	75.40	66.65	72.61	69.83
Hydrogen	5.12	3.64	4.76	4.87	4.10	5.03
Oxygen (by difference)	14.40	14.28	8.50	11.63	5.26	7.25
Nitrogen	1.34	1.61	1.41	1.17	1.49	1.30
Sulfur	2.20	0.50	0.81	1.96	3.32	3.17
Proximate, wt.% as received	L-60 Coke Breeze	L-4-40 Coke Breeze	ARCO Petroleum Coke	Chevron Coke	Rockwell Char	
Volatile Matter	4.76	2.19	9.67	9.00	2.35	
Fixed Carbon	82.31	83.27	89.23	80.37	70.65	
Moisture	1.38	1.19	0.87	10.24	7.21	
Ash	11.55	13.56	0.21	0.39	19.79	
Ultimate, wt.%						
Carbon	84.21	83.27	89.73	80.82	70.40	
Hydrogen	0.78	0.76	3.75	4.55	2.12	
Oxygen (by difference)	1.89	0.73	1.58	0.64	5.20	
Nitrogen	0.85	0.75	1.39	1.45	1.28	
Sulfur	0.72	0.93	2.47	1.91	1.21	

TABLE 2-22 (Continued)

FEEDSTOCK PROPERTIES

	Pittsburgh No. 8 Adamsburg Mine	Pittsburgh No. 8 Westland Mine	Upper Freeport Renton Mine	Indiana No. 7 Minnehaha Mine	Indiana No. 7 Sycamore Mine	Upper Freeport Elton Mine
Chlorine, %	0.14	0.11	0.44	0.05	0.05	0.04
Fluorine, ppm	60	53.3	60	70	70	90
Sulfur Forms, %						
Sulfur	1.41	1.60	1.11	0.70	0.86	3.13
Pyritic Sulfur	0.49	1.21	0.19	0.17	0.22	2.14
Sulfate Sulfur	0.18	0.01	0.45	0.01	0.20	0.04
Organic Sulfur	0.74	0.39	0.47	0.53	0.44	0.95
	Wyoming Sub-C Wyodak Mine	W. Kentucky No. 9 Hanilton Mine	Montana Sub-B Rosebud	Texas Lignite Monticello Mine	Texas Lignite Big Brown Mine	N. Dakota Lignite Indian Head Mine
Chlorine, %	0.02	0.27	0.03	0.04	0.02	0.01
Fluorine, ppm	69.2	20	40	50	61.07	60
Sulfur Forms, %						
Sulfur	0.41	2.63	1.17	0.72	0.85	0.68
Pyritic Sulfur	0.07	1.49	0.52	0.04	0.48	0.30
Sulfate Sulfur	0.05	0.12	0.23	0.02	0.01	0.08
Organic Sulfur	0.28	1.02	0.42	0.66	0.37	0.31
	Illinois No. 6 Delta Mine	RSA Eosjesspruit	W. Virginia Big Mountain Mine	Ohio No. 9 Bannock Mine	Brookville Ace Drilling	Pittsburgh No. 8 Federal No. 2 Mine
Chlorine, %	0.10	0.02	0.14	0.11	0.12	0.09
Fluorine, ppm	94.7	107.2	40	30	80	100
Sulfur Forms, %						
Sulfur	2.39	0.69	0.85	1.85	2.39	2.66
Pyritic Sulfur	1.23	0.21	0.13	0.57	2.66	1.01
Sulfate Sulfur	0.02	0.04	0.01	0.45	0.11	0.02
Organic Sulfur	1.15	0.45	0.72	0.83	0.63	1.63
	L-60 Coke Breeze	L-4-40 Ccke Breeze	ARCO Petroleum Coke	Chevron Coke	Rockwell Char	
Chlorine, %	0.10	0.03	0.02	0.02	0.35	
Fluorine, ppm	20.00	< 10.00	< 10.00	< 10.00	70.00	
Sulfur Forms, %						
Sulfur	0.82	0.90	2.26	2.09	1.37	
Pyritic Sulfur	0.05	0.15	0.00	0.00	0.13	
Sulfate Sulfur	0.03	0.02	0.01	0.01	0.14	
Organic Sulfur	0.74	0.73	2.25	2.08	1.10	

TABLE 2-22 (Continued)

FEEDSTOCK PROPERTIES

Analyses, wt.% of Ash	Pittsburgh No. 8 Adamsburg Mine	Pittsburgh No. 8 Westland Mine	Upper Freeport Renton Mine	Indiana No. 7 Minnehaha Mine	Indiana No. 7 Sycamore Mine	Upper Freeport Elton Mine
SiO ₂	54.00	44.38	47.40	54.86	60.90	44.46
Al ₂ O ₃	29.30	24.17	26.40	22.18	25.40	23.03
CaO	2.70	0.13	1.80	0.30	1.80	0.09
MgO	0.90	0.44	0.70	0.75	1.50	0.22
Na ₂ O	0.40	0.46	0.20	0.15	0.30	0.15
K ₂ O	1.90	1.69	2.60	2.49	3.10	2.52
Fe ₂ O ₃	7.80	17.30	19.90	5.95	7.00	19.78
SO ₃	1.40	1.10	1.00	0.20	0.02	- -
Analyses, wt.% of Ash	Wyoming Sub-C Wyodak Mine	W. Kentucky No. 9 Familton Mine	Montana Sub-B Rosebud	Texas Lignite Monticello Mine	Texas Lignite Big Brown Mine	N. Dakota Lignite Indian Head Mine
SiO ₂	34.51	50.30	35.40	32.50	48.69	34.54
Al ₂ O ₃	16.13	20.60	19.30	19.30	14.33	14.02
CaO	3.18	2.00	17.80	22.20	5.06	5.59
MgO	2.29	0.80	4.40	3.20	2.04	0.86
Na ₂ O	0.48	0.70	0.30	- -	0.08	5.43
K ₂ O	0.30	3.20	0.10	- -	0.74	1.46
Fe ₂ O ₃	2.26	18.90	5.60	5.30	7.06	9.07
SO ₃	16.40	2.00	16.30	14.30	14.30	- -
Analyses, wt.% of Ash	Illinois No. 6 Delta Mine	RSA Bosjesspruit	W. Virginia Big Brown Mine	Ohio No. 9 Bannock Mine	Brookville Ace Drilling	Pittsburgh No. 8 Federal No. 2 Mine
SiO ₂	45.00	48.59	54.40	54.60	38.93	43.59
Al ₂ O ₃	18.76	26.74	31.28	24.00	22.64	19.76
CaO	3.77	1.06	0.94	1.80	0.33	4.31
MgO	1.41	2.67	0.32	0.60	0.21	1.02
Na ₂ O	1.41	1.34	0.00	0.30	1.02	0.89
K ₂ O	2.24	0.38	1.40	1.90	1.52	1.69
Fe ₂ O ₃	12.83	3.19	7.24	17.10	23.50	15.03
SO ₃	7.40	- -	0.61	0.20	- -	- -
Analyses, wt.% of Ash	L-60 Coke Breeze	L-4-40 Coke Breeze	ARCO Petroleum Coke	Chevron Coke	Rockwell Char	
SiO ₂	51.92	50.15	26.27	31.53	50.33	
Al ₂ O ₃	28.70	28.82	9.61	6.25	16.87	
CaO	1.96	2.24	1.28	3.35	0.46	
MgO	0.61	1.11	3.67	1.10	0.66	
Na ₂ O	0.68	0.62	4.42	0.44	0.91	
K ₂ O	0.69	0.62	1.69	0.82	1.76	
Fe ₂ O ₃	14.66	15.63	7.11	15.05	20.66	
SO ₃	0.34	0.37	- -	- -	- -	

TABLE 2-22 (Continued)
FEEDSTOCK PROPERTIES

Ash Properties	Pittsburgh No. 8 Adamsburg Mine		Pittsburgh No. 8 Westland Mine		Upper Freeport Renton Mine		Indiana No. 7 Minnehaha Mine		Indiana No. 7 Sycamore Mine		Upper Freeport Elton Mine	
	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing
Initial Deformation, °F	2650	2550	2560	2400	2610	2420	2380	2330	2610	2460	2550	2232
Ash Softening, °F	2740	2650	2630	2560	2710	2580	2480	2420	2720	2660	2624	2332
Ash Fusion, °F	2780	2750	2660	2620	2750	2660	2560	2700	2760	2750	2654	2418
Calorific Value, Btu/lb	13,290		12,570		13,270		12,100		13,000		13,059	
Free Swelling Index	8.0		5.5		6.5		1.5		2.0		8.4	
Hardgrove Grindability	57.0		70.5		58.0		55.0		53.2		95.2	
Reactivity	5.0		3.7		--		8.0		6.1		--	
Ash Properties	Wyoming Sub-C Wyodak Mine		W. Kentucky No. 9 Hamilton Mine		Montana Sub-B Rosebud		Texas Lignite Monticello Mine		Texas Lignite Big Brown Mine		N. Dakota Lignite Indian Head Mine	
	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing
Initial Deformation, °F	2200	2220	2370	2120	2290	2240	2390	2350	2220	2180	2130	1950
Ash Softening, °F	2270	2290	2440	2200	2350	2320	2460	2410	2310	2270	2220	2070
Ash Fusion, °F	2340	2360	2490	2320	2420	2400	2540	2500	2500	2460	2240	2140
Calorific Value, Btu/lb	9,396		12,580		10,177		7,500		8,350		8,500	
Free Swelling Index	0.0		5.0		0.0		0.0		0.0		0.0	
Hardgrove Grindability	46.0		62.9		50.4		68.4		45.0		38.0	
Reactivity	59.8		--		71.0		28.2		118.0		122.0	
Ash Properties	Illinois No. 6 Delta Mine		RSA Bosjesspruit		W. Virginia Big Mountain Mine		Ohio No. 9 Bannock Mine		Brookville Ace Drilling		Pittsburgh No. 8 Federal No. 2 Mine	
	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing
Initial Deformation, °F	2370	2200	2340	2420	2670	2700	2590	2380	2350	2030	2338	2118
Ash Softening, °F	2440	2290	2500	2470	2800+	2800+	2680	2440	2410	2130	2408	2188
Ash Fusion, °F	2510	2370	2600	2590	2800+	2800+	2720	2520	2440	2230	2453	2340
Calorific Value, Btu/lb	11,540		10,170		13,200		12,228		12,160		13,276	
Free Swelling Index	2.5		0.5		7.5		3.0		6.17		--	
Hardgrove Grindability	57.4		65.0		52.0		53.2		74.0		--	
Reactivity	5.0		49.4		11.9		4.7		2.2		--	
Ash Properties	L-60 Coke Breeze		L-4-40 Coke Breeze		ARCO Petroleum Coke		Chevron Coke		Rockwell Char			
	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing	Oxidizing	Reducing		
Initial Deformation, °F	2530	2410	2400	2320	2170	2330	--	--	2300	2060		
Ash Softening, °F	2600	2500	2550	2420	2520	2390	--	--	2450	2210		
Ash Fusion, °F	2720	2600	2680	2580	2600	2540	--	--	2540	2580		
Calorific Value, Btu/lb	12,400		12,355		15,470		15,580		11,270			
Free Swelling Index	0.0		0.0		0.5		0.5		0.5			
Hardgrove Grindability	--		--		76.0		--		--			
Reactivity	1.0		--		1.9		1.2		49.5			

TABLE 2-23
PARTICLE SIZE DISTRIBUTION

Solids Size Distribution	Pittsburgh No. 8 Adamsburg Mine		Pittsburgh No. 8 Westland Mine		Upper Freeport Renton Mine		Indiana No. 7 Minnehaha Mine		Indiana No. 7 Sycamore Mine		Upper Freeport Elton Mine	
Mesh Size	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³
+6	0.78	- - -	0.06	- - -	0.60	- - -	3.78	37.9	1.01	- - -	9.96	46.4
-6 +10	17.36	42.3	2.98	37.5	2.57	- - -	8.79	40.4	7.59	41.73	13.91	45.9
-10 +20	40.72	40.9	40.15	42.1	28.68	42.3	22.87	40.4	37.53	40.75	26.31	45.75
-20 +40	18.98	38.3	31.05	65.9	23.59	40.0	21.83	39.4	23.32	39.02	17.01	41.5
-40 +70	12.00	36.2	14.65	41.0	18.20	39.2	18.57	37.2	16.64	37.46	13.59	34.1
-70 +100	3.97	33.7	5.95	40.4	10.96	38.0	7.67	36.8	6.42	35.58	5.62	32.25
-100 +200	6.19	32.8	5.17	31.9	15.39	32.0	16.43	32.1	7.49	33.99	13.99	29.2
Solids Size Distribution	Wyoming Sub-C Wycdak Mine		W. Kentucky No. 9 Hamilton Mine		Montana Sub-B Rosebud		Texas Lignite Monticello Mine		Texas Lignite Big Brown Mine		N. Dakota Lignite Indian Head Mine	
Mesh Size	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³
+6	8.05	37.5	0.00	- - -	2.63	- - -	0.07	- - -	4.35	42.7	6.36	36.17
-6 +10	17.34	36.7	8.73	39.87	15.82	41.2	9.40	37.57	15.92	39.9	9.98	38.41
-10 +20	31.24	37.6	32.68	39.63	39.21	39.7	35.08	40.01	39.45	40.1	26.31	45.75
-20 +40	18.36	37.3	23.89	39.51	21.39	38.8	21.95	40.06	21.30	38.7	17.01	41.5
-40 +70	10.40	37.1	17.29	39.60	11.01	38.3	13.50	39.28	10.73	38.7	13.59	34.1
-70 +100	3.92	38.6	6.63	37.37	4.07	36.8	4.73	37.38	3.38	39.2	5.62	32.25
-100 +200	10.20	32.1	10.79	37.56	5.88	32.3	15.28	35.89	4.94	38.6	13.99	29.2
Solids Size Distribution	Illinois No. 6 Delta Mine		RSA Bosjesspruit		W. Virginia Big Mountain Mine		Ohio No. 9 Bannock		Brookville Ace Drilling		Pittsburgh No. 8 Federal No. 2 Mine	
Mesh Size	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³
+6	0.78	- - -	7.00	47.1	6.41	38.64	14.87	45.09	4.15	47.12	15.43	44.1
-6 +10	14.71	42.16	14.05	47.1	13.68	38.73	29.84	45.72	8.96	46.74	20.10	44.4
-10 +20	37.47	41.62	30.72	47.0	34.40	40.69	32.97	45.78	28.93	44.08	30.66	45.6
-20 +40	18.72	40.54	17.64	46.1	20.29	39.23	10.92	41.96	20.42	43.49	14.04	43.6
-40 +70	11.16	37.21	12.12	46.1	10.29	38.37	5.51	40.58	12.99	42.56	7.62	43.5
-70 +100	4.93	33.29	5.16	42.9	3.11	38.08	1.79	46.82	4.79	45.22	2.67	40.5
-100 +200	12.23	34.02	13.30	39.1	11.82	32.03	4.10	33.51	19.77	36.30	9.49	35.3
Solids Size Distribution	L-60 Coke Breeze		L-4-40 Coke Breeze		ARCO Petroleum Coke		Chevron Coke		Rockwell Char			
Mesh Size	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³	wt. %	Bulk Density lb/ft ³		
+6	0.00	- - -	20.58	27.30	3.55	30.90	0.71	- - -	- - -	- - -		
-6 +10	2.77	27.70	28.59	28.60	10.49	34.70	5.52	37.74	- - -	- - -		
-10 +20	20.84	32.30	33.71	29.70	30.48	38.00	34.23	42.07	28.08	7.16		
-20 +40	23.34	35.50	14.51	33.10	19.75	38.10	24.65	42.19	15.95	8.14		
-40 +70	23.61	39.00	1.74	32.00	14.75	38.20	16.17	41.18	19.18	18.38		
-70 +100	12.19	40.80	0.28	- - -	5.80	36.30	5.52	38.81	15.95	18.76		
-100 +200	17.25	43.50	0.59	- - -	15.18	34.00	13.18	33.54	13.16	18.16		
-200 +270	- - -	- - -	- - -	- - -	- - -	- - -	- - -	- - -	7.50	18.40		

Figure 2-38. Particle Size Distribution vs. Coal Type - Composite

Figure 2-39. Automatic Ash Withdrawal

varying feedstock particle sizes over a series of narrow ranges is required to fully assess the effect of coal size on gasifier performance.

2.4 SYSTEM INTEGRATION AND CONTROL

Achievement of prolonged steady state operation and smooth, orderly, and timely load follow operation of the gasifier can be enhanced by automatic control and integration of key input and output flows. During the 1981 testing year, several automatic control loops were installed and tested with very good results. Load follow characteristics of the gasifier were being studied to demonstrate the practicability of the Westinghouse gasifier to perform in such a manner.

2.4.1 Automatic Ash Annulus Control

This control loop existed in various forms, using a temperature difference (ΔT) between two thermocouples located in the ash annulus of the gasifier. Testing was limited, due to an inability to determine the correct ΔT to maintain, in order to operate the ash annulus with consistency. The loop was modified to its present form, using a single temperature signal to the controller. This control scheme thereafter gave excellent results. Figure 2-39 is a simplified schematic of the control loop.

Figures 2-40 and 2-41 are time plots showing manual and automatic control of the ash withdrawal rate, respectively, using RSA coal as the feedstock. The top portion of the plots indicates the ash withdrawn from the annulus during each hour. The middle and bottom portions of the plots indicate the bulk density and ash concentrations by weight percent of each hour sample. As can be seen in Figure 2-40, during manual operation of the withdrawal there is the possibility of withdrawing too much ash. The result of this is to draw out some bed material, diluting the high ash material. Also, the average ash concentration over the time period shown was 76 weight percent, and the actual ash concentration of each hourly sample was very unsteady. Figure 2-41 shows that even though there were large swings in the amount of ash being withdrawn from one hour to the next, the ash concentration of the samples did not vary greatly. This indicates that only as much as was produced was removed. The automatic ash annulus control loop, therefore, smooths out the operation of the ash withdrawal and helps to increase carbon utilization by not removing lower ash bed material.

2.4.2 Automatic Bed Level Control

Bed level is controlled at a given height and freeboard temperature by adjusting the coal feed rate. This is accomplished by adjusting the speed of the starwheel feeder below the coal lockhopper. Prior to June 1981, the control room operator controlled gasifier bed height by monitoring either the pressure differential recorders of the bed or the nuclear densitometer

Figure 2-40. Manual Ash Annulus Control RSA Coal - TP-031-1

Figure 2-41. Automatic Ash Annulus Control RSA Coal

bed level recorder and making a manual adjustment to the speed controller of the appropriate starwheel. Figure 2-42 shows the variation of bed height and freeboard temperature that results from manual bed level control of bed height.

During 1981 an automatic bed level control was installed on the PDU to address the problems mentioned above. Figure 2-43 is a schematic of the bed level control loop. The control room operator can select either the nuclear densitometer bed level signal or a filtered pressure differential signal to the level controller. The level controller then adjusts the coal feed starwheel speed, as required, to maintain the desired bed level. During the testing year, it was found that either the nuclear densitometer or the pressure differential transmitter signal can be used with equal success in controlling bed height. Figure 2-45 shows automatic bed level control of the gasifier with Pittsburgh seam coal.

2.4.3 Automatic Temperature Control

Automatic temperature control is not usually necessary in a steady-state operation of the gasifier. The desired freeboard temperature is achieved by first building the char bed to a desired level and adjusting oxygen flow, as needed. If the bed level controller is functioning properly, the freeboard temperature, as shown previously, should stabilize. However, in a load follow situation, the gasifier will have to be turned down or up and stabilized as quickly as possible. This would be difficult to achieve without some automatic control of the temperature. Since load follow and turndown of the gasifier will be studied in future testing, a control scheme needed to be developed. The Cascade Ratio Temperature Control loop was initially selected as the control scheme on the PDU. This control scheme should provide a smooth and orderly transition from a base load set point to a "turned down" set point via steam demand modulation.

The temperature control loop was installed late in 1981 and tested only in steady-state operation during TP-030-3. Figure 2-44 is a plot of freeboard temperature and bed height, with both bed height and freeboard temperature automatically controlled during a portion of TP-030-3. The increase in the freeboard temperature from 1800°F to 1850°F (982 to 1010°C) in the middle of the plot is the result of a transition between set points.

2.4.4 Transient Response

To demonstrate and control the load follow characteristics of the gasifier, it is necessary to study the response of the gasifier to various transients imposed on it. This is necessary to determine: (1) how fast the gasifier reacts to a transient; (2) how fast any automatic control loop must respond; and (3) if the gasifier can be made to respond more quickly through intervention by an integrated control scheme involving several automatic control loops acting in conjunction, using feed forward signals.

Figure 2-42. Manual Bed Height Control

Figure 2-43. Bed Level Control

Figure 2-44. Automatic Bed Height Control and Automatic Temperature Control

Figure 2-45. Automatic Bed Height Control

Transient response testing was initiated during TP-030-3. Transients imposed on the gasifier were 10 percent CO₂ step increase, 10 percent steam step increase, 10 percent oxygen step decrease, and total solids/transport gas interruption. The step increases in CO₂ and steam did not noticeably affect the gasifier with the exception of the product gas composition. Larger increases of these flow rates will be necessary to determine their effect on the gasifier. The effect of the oxygen and solid feed transients, however, were clearly observed.

The oxygen transient was a 10 percent cut in the flow of oxygen to the air tube. All other inputs and conditions were held constant. Figure 2-46 shows the cut in oxygen flow and the resulting drop in freeboard and bed temperatures, as well as the effect on the superficial bed density in the combustion zone and the lower annulus. With the cut in oxygen, a cut in jet and fluidization velocities resulted also.

The oxygen transient was not a smooth, single step decrease, but a number of small decreases over several minutes. This transient should be repeated in future testing, so that the response from a single step decrease can be studied. Figure 2-47 shows the effect of the oxygen step decrease on the product gas composition.

The solids feed interruption was unplanned but did provide another transient response. Figure 2-48 shows the effect that the coal feed interruption had on the product gas composition. Once the operators detected the coal and fines feed interruption, the following corrective actions were taken: (a) reduced the oxygen flow; (b) isolated the feed line; and (c) dislodged the plug through a bleedline by reverse flow of carbon dioxide. Figure 2-49 shows the response of the gasifier in the time interval between the onset of the problem and the initiation of correction actions.

When fines and coal feed are interrupted without the oxygen flow being turned down, the char bed is the only fuel source for the oxygen to combust with. As the bed becomes consumed, the bed height falls and the heat sink capacity that the bed provides is lessened. Consequently, the bed and freeboard temperatures begin to rise. The bed-ash concentration would influence how fast the bed would become consumed and the temperature would increase. The higher the bed ash, the faster the response.

Figure 2-47. Wet Product Gas Analysis O₂ Step Decrease

Figure 2-48. Wet Product Gas Analysis Following Coal Feed Interruption

Figure 2-49. Transient Response - Coal and Fines Feed Shut Off

SECTION 3.0

FINES CARRYOVER AND RECYCLE

3.1 ISSUES

An important objective of the gasification development program is to demonstrate carbon conversion in the process to be greater than 90 percent. The percent carbon conversion in the KRW gasifier,

$$\frac{\text{carbon in coal} - \text{carbon loss}}{\text{carbon in coal}} \times 100$$

is principally influenced by the loss of char fines through the fines collection and recycle system. Issues related to fines recycle performance are:

- A. Fines Elutriation: Estimation of fines elutriation rate as a function of coal feedstock and gasifier freeboard velocity.
- B. Fines Consumption: Estimation of fines consumption rate as a function of gasifier temperature, recycled fines temperature, and fines injection point, and evaluations of the effect of recycled fines on gasifier operation.

3.2 BACKGROUND

Carryover of fine char particles from the KRW gasifier, Figure 3-1, is influenced by one purely physical process, fines elutriation from the fluidized bed, and two physio-chemical processes, fines generation and fines consumption. To study fines consumption and the effects of fines recycle on gasifier operation, similar set points were run with a coal during which: (1) no fines were recycled, (2) primary fines only were recycled, and (3) both primary and secondary fines were recycled. Fines recycle data were also analyzed to estimate the generation of fines from the fresh coal feed caused by the shock heating and attrition occurring in the fluidized bed. The generation rate was determined by assuming that all particles (coal and recycle fines) fed to the gasifier with a terminal velocity less than the freeboard velocity would be elutriated unless consumed.

3.3 FINE PARTICLE CARRYOVER ESTIMATION BASED ON PDU DATA

3.3.1 General

It is important to be able to predict fines carryover rate in gasification system design. An estimate is needed for selection of the recycle technique, specification of component sizes, and evaluation of the overall

Figure 3-1. Mechanistic Diagram for Fines Carryover from the Gasifier

process heat and material balances. Recycle rate magnitude can also affect the gasifier design itself.

The fines carryover rate will depend on numerous complex rate phenomena occurring simultaneously - elutriation, attrition, agglomeration, gasification, combustion, and cyclone performance. Any model that is to be used for entrainment prediction must consider all these events in some manner, yet a theoretical prediction is clearly not possible. An empirical model, with simple assumptions about these phenomena, is appropriate.

In the following text, such a simple model is developed in which the various mechanisms of fines formation and removal are treated parametrically, in the context of a fines balance around the gasifier. A substantial body of PDU data from 1980-1983 was used to evaluate the parameters. Additionally, use of the model for entrainment rate prediction is discussed.

3.3.2 Model Development

Consider the particles composing the fluidized bed to be of two types: fines that are entrainable, and coarse particles that are not entrainable. The dividing line between them can be considered to be the size of that particle the terminal velocity of which equals the freeboard gas velocity at the conditions of interest.

3.3.2.1 Phenomena

The model consists of a steady-state fines material balance around the gasifier. Entrainment rate is considered to be the sum of four net rates of addition of fines to the gasifier. Each of these fines addition rates represents a distinct source of fines - i.e., a distinct measurable weight or flow of solids. Each of the four addition rates is a net rate in that some of the fines so introduced may be consumed by agglomeration, devolatilization, or chemical reaction prior to entrainment. Fines are assumed to leave the gasifier only as entrained material in the product gas stream. Each of the fines addition rates is modeled by a single fines addition parameter. The fines sources are:

- Fines transported in with the coal feed.
- Attrition of large coal particles caused by the thermal shock of entering the hot gasifier bed, together with associated events such as rapid devolatilization.
- Relatively slow break-up of large bed particles caused by shrinkage or disintegration into smaller particles by (1) chemical consumption including combustion, pyrolysis, and gasification, and

(2) by attrition of large bed particles caused by mechanical action, such as collisions between particles, collisions between particles and walls, and abrasion.

- Recycled fines that are neither consumed by agglomeration or chemistry.

No attempt is made to actually model the complex mechanisms, only the sources.

3.3.2.2 Fine Coal Feed

Fine particles are usually present in the coal feed and they contribute to entrainment at the rate

$$dF \quad (3-1)$$

where d = fines addition constant, (lb fines surviving)/(lb fines fed)

F = fine non-volatile coal feed rate, lb/hr

Note that in the definition of " F ", "fine" is defined relative to the particle size that can be entrained at the given conditions, and that only non-volatiles are considered. Parameter " d " will be less than one if a fraction of this fine coal feed is consumed by gasification prior to entrainment.

3.3.2.3 Fines Generation from Coal Feed

Coarse coal feed also contributes to carryover because of disintegration or shrinkage caused by thermal shock upon its introduction into the gasifier and the associated fast reactions and devolatilization that occur. The rate of fines addition by this source is

$$bzC \quad (3-2)$$

where b = fines addition constant, (lb 149 μ m or smaller fines formed)/(lb coarse coal fed)

z = correction factor (lb fines formed)/(lb 149 μ m or smaller fines formed)

C = coarse non-volatile coal feed rate, lb/hr

Parameter " b " may be considered a normalized coal feed attrition constant. Correction factor " z " allows this parameter to be used at a variety of operating conditions, as explained in paragraph 3.3.2.6.

3.3.2.4 Fines Generation from Slow Bed Processes

Fines are also created by the slow breakup and shrinkage of coarse bed particles at the rate

$$azW \quad (3-3)$$

Where a = fines addition constant, (lb 149 μ m or smaller fines formed per hour)/(lb bed material),

W = bed weight, lb,

and z is as before.

Again " a " may be considered a normalized bed attrition constant, although in fact it represents a very complex set of physical and chemical effects. Technically, the weight of coarse material in the bed should be used rather than the total bed weight but this would greatly complicate the model, since we would need a way to predict fines concentration in the bed. However, PDU data have consistently shown that only a small fraction of the bed consists of fines so that the difference between the coarse and total bed weights is insignificant relative to our ability either to measure it or predict it. This consideration holds true for fluidized-bed gasification in general because of the wide size distribution of solids present. It is, therefore, justifiable to use the total bed weight.

3.3.2.5 Recycle Fines Survival

The rate at which recycled fines are added to the bed is

$$cR \quad (3-4)$$

where c = fines addition constant, (lb fines surviving)/(lb fines fed)

R = fines recycle rate, lb/hr

The recycle fines addition constant " c " is one minus the fraction of recycled fines consumed per pass.

3.3.2.6 Effects of Operating Conditions

In several of the fines generation expressions developed previously, the correction factor " z " is used so as to allow using the same fines addition constants at different operating conditions. For instance, if two PDU runs were made at different freeboard velocities, such that 200 μ m and 250 μ m particles were the largest entrainable, then the attrition constants derived would give rates of formation of < 200 μ m and < 250 μ m particles, respectively. Similarly, if it were desired to design a gasifier in which < 300 μ m particles were entrained, it would

be difficult to use an attrition constant relative to a smaller size. The factor "z" allows such a computation by making the fines addition constants independent of operating conditions.

We can determine "z" for any PDU data set from the particle size distribution analysis of the carryover material, and a calculation based on freeboard velocity, pressure, temperature, and gas composition of the largest particle size (d_m) that can be entrained. Similarly, we can predict "z" for other conditions by using the operating conditions to predict d_m , and making an assumption about particle size distribution. These two similar procedures, one for PDU data, and one for design, are explained in paragraphs 3.3.4 and 3.3.5, respectively.

The data do not support making generalizations about other effects of operating conditions.

3.3.3 Integrated Model

Expressions (3-1), (3-2), (3-3), and (3-4) are added to give the net rate at which fines are added to the bed. At steady state, this is also the entrainment rate, E (lb/hr):

$$E = azW + bzC + cR + dF \quad (3-5)$$

3.3.4 Data

Data from 71 steady-states from 15 PDU tests during the period 1980-1983 were used to evaluate the four parameters in Equation (3-5), by first calculating the values of E, W, C, R, F, and z for each point, and then correlating.

The data are summarized in Table 3-1, the entries of which were derived from the data in the appropriate test reports as follows:

1. The coal feed rate was taken directly from the appropriate report.
2. The fines recycle rate (R) was assumed to be the average of the primary fines collection rate and the fines feed rate as listed in the report, since these should be duplicate measurements.
3. The entrainment rate (E) was determined by forcing an ash balance on the gasifier.

$$E = (Gf_c - Af_a + Rf_r)/f_r \quad (3-6)$$

where

- G = coal feed rate, lb/hr
- f_c = fraction ash in coal feed
- A = ash withdrawal rate, lb/hr
- f_a = fraction ash in ash withdrawal stream
- f_r = fraction ash in recycled fines

TABLE 3-1
SUMMARY OF PDU DATA

TEST	023-4					027-1			027-2					
COAL	Pittsburgh				Tx Lign	Ohio - Indiana			Ohio #9			Pittsburgh		
POINT	1A	1B	2A	2B	3	1B-1	1B-2	1C	1A	1B	M1B	2B-1	2B-2	2B-3
Coal feed rate, lb/hr (G)	890	856	718	681	1546	1451	1430	1435	1139	1227	950	1007	884	962
Fines recycle rate, lb/hr (R)	284	282	248	258	422	240	274	224	140	181	0	151	124	181
Entrainment rate, lb/hr (E)	405	533	359	404	540	597	575	556	349	407	153	362	254	244
Bed weight, lb (W)	318	336	271	316	273	491	445	436	425	432	383	350	271	288
Freeboard gas velocity, ft/s	1.93	1.88	1.83	1.82	1.72	2.02	2.15	1.92	1.63	1.74	1.44	1.44	1.41	1.45
Max. particle size, μm (d_m)	279	280	272	271	286	321	334	308	274	284	245	249	246	250
Fraction coal $< d_m$ (f_d)	.08	.29	.046	.045	.155	.14	.105	.135	.073	.115	.057	.06	.062	.042
Coal fines feed rate, lb/hr (F)	45	162	21	19	177	131	98	127	55	93	36	39	35	26
Coal coarses feed rate, lb/hr (C)	513	397	429	399	638	805	836	812	693	712	592	615	532	584
Correction factor (Z)	1.34	1.46	1.48	1.37	1.55	1.69	1.56	1.57	1.38	1.38	1.34	1.35	1.28	1.24

TABLE 3-1 (Continued)

SUMMARY OF PDU DATA

TEST	027-3							027-4						
COAL	Tx Lign	Pittsburgh						Wyoming Sub-C		Western Kentucky				
POINT	1A	2A	2B	2C	2D	2E	2A-1	M1A	M3A	5A-1	5A-2	5B-1	5B-2	6A
Coal feed rate, lb/hr (G)	1413	1086	1022	970	964	1073	1043	2337	2552	1781	1582	2222	2037	2300
Fines recycle rate, lb/hr (R)	484	0	70	62	47	0	329	492	0	350	406	398	376	524
Entrainment rate, lb/hr (E)	925	275	271	238	171	131	436	1054	500	778	671	772	729	917
Bed weight, lb (W)	614	336	355	409	353	379	300	714	628	446	439	369	395	471
Freeboard gas velocity, ft/s	1.44	1.45	1.39	1.48	1.59	1.58	1.62	2.04	1.96	1.68	1.68	1.96	1.84	1.76
Max. particle size, μm (d_m)	229	247	241	253	271	268	270	290	281	278	278	319	301	290
Fraction coal $< d_m$ (f_d)	.24	.045	.035	.05	.055	.043	.033	.19	.18	.23	.16	.23	.18	.31
Coal fines feed rate, lb/hr (F)	192	32	23	32	35	30	22	203	210	246	153	304	224	433
Coal coarses feed rate, lb/hr (C)	609	685	641	605	603	673	656	867	959	824	805	1019	1020	963
Correction factor (Z)	1.27	1.29	1.26	1.54	1.31	1.24	1.16	1.50	1.38	1.42	1.27	1.80	1.36	1.35

TABLE 3-1 (Continued)

SUMMARY OF PDU DATA

TEST	027-5				029-1/030-1					030-2					
COAL	Tx. Lignite			Wy. Sub C	Illinois #6		Wyoming Sub C			Indiana #7		RSA			
POINT	1B-1	1C-1	1D-1	3A	1A	3A	4A	4B	6C	1B-1	1B-2	6A-1	62-A	6B-1	6C-1
Coal feed rate, lb/hr (G)	1088	3081	2049	2077	1152	855	2052	2190	171C	1363	1378	1396	1626	1541	1387
Fines recycle rate, lb/hr (R)	396	0	0	0	254	196	359	245	31C	361	324	347	343	373	369
Entrainment rate, lb/hr (E)	575	763	522	527	438	331	827	810	637	567	563	592	587	626	626
Bed weight, lb (W)	708	347	808	469	440	407	567	536	515	421	401	689	744	825	880
Freeboard gas velocity, ft/s	0.87	1.40	1.11	0.99	1.07	.90	1.24	1.35	1.55	1.22	1.15	1.17	1.25	1.29	1.37
Max. particle size, μm (d_m)	174	231	202	190	206	186	222	233	275	222	215	215	223	228	236
Fraction coal $< d_m$ (f_d)	.18	.19	.18	.24	.18	.17	.27	.185	.14	.18	.18	.27	.17	.17	.25
Coal fines feed rate, lb/hr (F)	104	288	181	240	129	88	259	185	106	149	149	278	202	190	253
Coal coarses feed rate, lb/hr (C)	475	1229	825	761	589	428	701	814	651	663	678	752	985	926	758
Correction factor (Z)	1.30	1.19	1.13	1.17	1.19	1.19	1.27	1.31	1.45	1.23	1.38	1.27	1.36	1.46	1.33

TABLE 3-1 (Continued)

SUMMARY OF PDU DATA

TEST	031-1						032-1			032-2				033-1	
COAL	RSA						Upper Freeport			Upper Freeport		W. Kentucky		Brookville	
POINT	2A	2A-1	2B-1	3A	3A-1	4A	1A-1	1B-1	1B-2	1A	1B	3B-1	3B-2	1B-1	1B-2
Coal feed rate, lb/hr (C)	1228	1265	1005	1494	1385	1102	1174	1152	1333	1014	989	1468	1454	1414	1424
Fines recycle rate, lb/hr (R)	387	366	259	369	282	0	0	0	0	136	242	0	306	516	456
Entrainment rate, lb/hr (E)	494	663	616	574	501	326	245	351	520	462	722	327	605	659	898
Bed weight, lb (W)	729	804	824	770	764	722	228	243	258	191	303	425	464	440	451
Freeboard gas velocity, ft/s	1.14	1.20	1.00	1.29	1.17	1.24	1.01	1.22	1.34	1.04	1.02	1.27	1.48	1.30	1.32
Max. particle size, μm (d_m)	212	220	202	229	218	224	199	227	259	202	199	236	253	234	239
Fraction coal $< d_m$ (f_d)	.24	.25	.135	.18	.155	.23	.30	.42	.38	.15	.15	.16	.175	.28	.20
Coal fines feed rate, lb/hr (F)	216	236	100	196	157	186	291	406	420	128	125	137	148	306	209
Coal coarses feed rate, lb/hr (C)	684	707	638	894	856	622	680	561	686	726	708	718	699	786	835
Correction factor (Z)	1.25	1.28	1.22	1.25	1.36	1.30	1.10	1.13	1.21	1.11	1.14	1.22	1.24	1.22	1.23

TABLE 3-1 (Continued)

SUMMARY OF PDU DATA

TEST	034-2				034-3				
COAL	Pittsburgh				Wyoming				
POINT	2A	2B	2C	2D	1A	1B	1C	1D	1F
Coal feed rate, lb/hr (C)	1015	1118	1221	1218	2185	2177	2187	2247	2110
Fines recycle rate, lb/hr (R)	2506	2368	2538	2580	3121	2537	2375	2484	3305
Entrainment rate, lb/hr * (E)	2560	2425	2583	2611	3227	2636	2535	2616	3502
Bed weight, lb (W)	444	466	475	450	410	412	412	420	481
Freeboard gas velocity, ft/s	1.15	1.29	1.37	1.40	1.52	1.54	1.51	1.58	1.47
Max. particle size, μm (d_m)	240	270	280	285	315	315	315	320	305
Fraction coal < d_m (f_d)	0.19	0.15	0.24	0.10	0.12	0.21	0.15	0.19	0.22
Coal fines feed rate, lb/hr (F)	119	103	180	75	96	194	129	156	202
Coal coarses feed rate, lb/hr (C)	509	584	570	670	735	729	762	666	718
Correction factor (Z)	1.44	1.55	1.59	1.60	1.70	1.70	1.70	1.71	1.61

and all variables were taken from the report, except R which was calculated as above.

4. The bed weight was calculated as

$$W = [(H - 13) 3.14 + 6.5] \rho_b \quad (3-7)$$

where H = reported bed depth, ft
 ρ_b = reported bed density (14' to 20'), lb/ft³
3.14 = cross-sectional area, ft²
6.5 = annulus volume, ft³

The first term in brackets represents the approximate gasification section volume, and the second term the approximate combustion zone volume.

5. For the non-mechanical valve tests, TP-034-2 and TP-034-3 the entrainment and recycle rates were determined by isokinetic rate data.
6. The maximum particle size (d_m) was calculated to be that particle size the terminal velocity of which equaled the freeboard velocity. System pressure, freeboard temperature, and gas composition, as reported, were used to calculate gas density and viscosity for this calculation. A particle density of 60 lb/ft³ was assumed. Standard drag relationships were used to relate particle size and velocity.
7. The fraction (f_d) of the coal feed less than " d_m " was determined from the coal particle size distribution.
8. The coal fines feed rate was calculated as

$$F = G(1 - f_v)(f_d) \quad (3-8)$$

where f_v = fraction volatiles (including moisture) in the coal feed.

9. The coarse coal feed rate was then

$$C = G(1 - f_v)(1 - f_d) \quad (3-9)$$

10. The correction factor "z" for each data point was determined from d_m as calculated above and from the particle size distribution of the entrained material as follows:

If log (particle size) is plotted against "percent less than" for the material collected by the primary cyclone on linear-scaled paper, the result is a nearly straight line except that an upward curve occurs near 100 percent because of the limited entrainability of larger particles. It would be possible to estimate "z" by taking the ratio of the percent less than " d_m " to the percent less than 149 μ m using this graph, but this would be faulty because the curve represents entrained fines rather than

generated fines, and it is the latter which is of interest. In this study, it is postulated that the relatively straight-line segment of the graph represents the relative creation rates of fines, and that the line may be extrapolated to the larger particle sizes for purposes of determining "z". This is justifiable if the particle sizes on the straight line section all have their terminal velocities exceeded by the freeboard velocity, which is true for the PDU data.

With this rationale, the straight line portion of the particle size distribution for collected primary fines in each of the PDU data points was fit by relationship of the form

$$\log (d) = MP + B \quad (3-10)$$

where: d = particle size (μm)
 P = "percent less than"
 M, B = parameters

The values of M and B were calculated using the two points at $149 \mu\text{m}$ (100 mesh) and $74 \mu\text{m}$ (200 mesh) directly from the appropriate data report.

Then

$$z = P(d_m)/P(d_{149}) \quad (3-11)$$

3.3.5 Correlation

3.3.5.1 Procedure

Equation (3-5) was rearranged to give

$$\frac{E}{zW} = a + b\frac{C}{W} + c\frac{R}{zW} + d\frac{F}{zW} \quad (3-12)$$

This is a linear equation in which each of the bracketed expressions is a variable that was calculated for each data point using the values in Table 3-1. A multiple linear regression analysis (least squares) was then applied to determine the best values of the parameters a , b , c , and d .

3.3.5.2 Results

The data regression results were grouped together according to such parameters as coal type and type of primary recycle system. Carbon and ash were treated individually in order to try to better understand the sensitivity of the model to the operational variables. Table 3-2 lists the values generated by the data regression for the various grouping of the data. No attempt was made to group all 71 available data points (58 with lockhopper mode recycle and 13 with non-mechanical valve mode recycle) due to the marked increase in elutriation rate measured during the non-mechanical valve tests. This suggests that the physio-chemical

TABLE 3-2

LINEAR REGRESSION PARAMETERS FOR ELUTRIATION MODEL

<u>Grouping</u>	<u>All Points (Lockhopper Recycle)</u>	<u>All Points (Non-Mechanical Valve Recycle)</u>	<u>Carbon Only (Lockhopper Recycle)</u>	<u>Ash Only (Lockhopper Recycle)</u>	<u>Wyoming and Lignite Coals (Lockhopper Recycle)</u>
Number of Data Points	58	13	58	58	11
a	0.00	-0.015	0.06	-0.09	0.13
b	0.27	0.15	0.19	0.55	0.07
c	0.78	0.96	0.75	0.94	0.63
d	0.57	0.33	0.57	0.53	1.97

<u>Grouping</u>	<u>Wyoming and Lignite Coals (All Points)</u>	<u>W. Kty., Indiana, Illinois Ohio-Ind. Blend</u>	<u>RSA Coal</u>	<u>Pgh. Coals (Lockhopper Recycle)</u>	<u>Pgh. Coals (All Points)</u>
Number of Data Points	20	14	10	23	27
a	0.019	0.24	0.20	-0.25	-0.20
b	0.17	0.12	0.05	0.34	0.33
c	0.88	0.75	0.55	1.02	0.93
d	1.41	0.55	0.59	0.88	0.52

processes (i.e. fines generation and consumption) may be greatly dependent upon the recycle mode and, therefore, it would be improper to lump these data points together. Additionally, from a practical standpoint, the limited data available from the non-mechanical valve tests would have little significance when regressed with the lockhopper data and would not greatly change the parameters from the "lockhopper only" case. Data from the two recycle modes were grouped together for individual coal types (due to limited data available). Additionally, the "lockhopper only" data was regressed for each coal type.

A review of Table 3-2 suggests some interesting conclusions; however, some of the parameters make no physical sense (i.e. values of c and d which are greater than one). Therefore, it is concluded that the attempted breakdown of the model by coal type, ash and carbon was unsuccessful at this time. This is probably due to the limited availability of data covering a wide enough range of operating conditions to provide a significant correlation.

- Of particular interest is that the values for 'a' are very low or negative, which suggests some net consumption of the fines in the bed may be occurring.
- A relatively large fraction of the coal fines feed is entrained. This appears less significant for non-mechanical valve tests, however, the reason for this reduction is unclear.
- A substantial fraction of the coarse coal becomes fines (by attrition and other mechanisms) upon first entering the gasifier, contributing directly to the entrainment. Again, this appears less significant in the non-mechanical valve tests.
- A significant amount of the recycled fines survive to be re-entrained. Methods must be examined to improve recycle fines consumption (i.e. feed point location, temperature, mixture with oxygen) or an alternative method for fines utilization should be developed. It should be noted that the non-mechanical valve test results are based upon limited testing (i.e., 13 points) and isokinetic rate data which, at present, cannot be verified by independent measurement. Further testing is required.

Given that

$$R = E\eta \quad (3-13)$$

where: η = cyclone efficiency,

we may write:

$$E = (0.27 Cz + 0.57 F)/(1 - 0.78\eta), \quad (3-14)$$

which is the predictive equation for fines entrainment utilizing lockhopper fines recycle. Additionally we may write:

$$E = (-0.015 zW + 0.15 Cz + 0.33 F)/(1 - 0.96\eta), \quad (3-15)$$

which is the predictive equation for fines recycle utilizing the non-mechanical valve fines recycle.

3.3.6 Use of Correlation for Prediction

3.3.6.1 Procedure

Equations (3-14) and (3-15) may be used to predict entrainment rates for various designs and sets of operating conditions. The following procedure is used:

1. Estimate cyclone efficiency (η).
2. Knowing the freeboard velocity, gas density, and viscosity, calculate the particle size (d_m) that has its terminal velocity equal to the freeboard velocity, assuming a particle density of 60 lb/ft³.
3. Estimate the coal feed rate (G), fraction volatiles (f_v) and particle size distribution. Determine the fraction (f_d) of the coal feed particles less than " d_m ". The coal fines (F) and coal coarse (C) feed rates are then determined with Equations (3-8) and (3-9), respectively.
4. Determine the "z" factor, based on the calculated value of " d_m " above.

The following expression was derived by regression analysis of all the PDU data in Table 3-1:

$$P = (\log(d) - 1.707) / 0.00641. \quad (3-16)$$

It represents a typical distribution of particle sizes and can be used to estimate "z". Combining Equations (3-11) and (3-16) yields:

$$z = \frac{\log(d_m) - 1.707}{\log(149) - 1.707} \quad (3-17)$$

5. Calculate the weight of the bed material (W).
6. Determine the entrainment rate from Equations (3-14) or (3-15).

The above analysis does not give an estimation of the ash content of the entrained material. An analysis of the data in Table 3-1 gave the following regression:

$$y = 0.088 + 0.612x \quad (3-18)$$

where: y = ash fraction of recycled material
 x = ash fraction of bed material

Equation (3-18) can be used to estimate carryover composition given a bed composition.

3.3.6.2 Example

Consider the following case:

Coal feed rate (G)	= 20,000 lb/hr
fraction volatiles (f_v)	= 0.2
fraction < 200 μm	= 0.10
fraction < 300 μm	= 0.15
Bed weight (W)	= 4000 lb
Cyclone efficiency	= 0.85
Freeboard velocity	= 1.5 ft/s
Pressure	= 300 psia
Temperature	= 1850°F
Molecular weight of gas	= 25

From this information the gas density is 0.30 lb/ft^3 and the viscosity is $3 \times 10^{-5} \text{ lbm/ft-sec}$. A particle of density 60 lb/ft^3 and of $270 \mu\text{m}$ (d_m) size would have a terminal velocity of 1.5 ft/s. Therefore, about 13 percent of the coal feed is less than " d_m " so, from Equations (3-8) and (3-9), the fine coal feed rate (F) is 2080 lb/hr, and the coarse coal feed rate (C) is 13,920 lb/hr. We estimate " z " from Equation (3-17) to be 1.55. Then the entrainment rate (E) from Equation (3-14) is 20,800 lb/hr.

The correlation can also be used to estimate the sensitivity of the entrainment rate to various parameters. As examples, Tables 3-3 through 3-6 show the effects of several parameters on entrainment rates, where in each table all other parameters are assumed to be those determined above. From these tables note that coal feed rate, cyclone efficiency, and freeboard velocity are important in determining entrainment rates, but that coal fines content is not.

TABLE 3-3
EFFECT OF COAL FEED RATE

<u>Coal Feed Rate, lb/hr</u>	<u>Entrainment Rate, lb/hr</u>
5000	5200
10000	10400
20000	20800
30000	31200
40000	41600

TABLE 3-4
EFFECT OF CYCLONE EFFICIENCY

<u>Efficiency</u>	<u>Entrainment Rate, lb/hr</u>
0	7000*
0.5	11500
0.6	13200
0.7	15400
0.8	18600
0.85	20800
0.9	23500
0.95	27100
0.99	30800
1.00	31900**

* No recycle.

** Ideal recycle.

TABLE 3-5
EFFECT OF COAL FINES CONTENT

<u>% < 270 μm</u>	<u>Entrainment Rate, lb/hr</u>
0	19900
5	20300
10	20600
13	20800
20	21300
25	21700

TABLE 3-6
EFFECT OF FREEBOARD VELOCITY

<u>Velocity ft/sec</u>	<u>Particle Size (d_m), μm</u>	<u>% < d_m</u>	<u>Z</u>	<u>Entrainment Rate, lb/hr</u>
0.5	125	6	0.84	11700
1.0	199	10	1.27	17400
1.5	270	13	1.55	20800
2.0	336	17	1.76	23300
3.0	486	28	2.10	27000

3.3.7 Conclusions

A simple material balance model for fine particle entrainment from a gasifier has been developed. The model treats fine particles as being added to the gasifier bed from four net sources - fine coal feed, coarse coal disintegration, bed particle disintegration, and fines recycle. The mechanisms of addition are not addressed. Extensive PDU data have been used to evaluate lump rate parameters for the various sources. We show how to use the model in prediction of entrainment rates. We believe the model will produce reasonable rough estimates of carryover.

3.3.8 Recommendations for Future Work

- Continue to develop separate model equations for different recycle modes.
- Continue attempts to correlate by coal type.
- Incorporate new test data into correlation.
- Attempt to verify the rate data generated by the isokinetic sampling since the model parameters are primarily dependent upon the reliability of this data.
- Attempt non-mechanical valve recycle at other feed locations to improve consumption.

3.4 NON-MECHANICAL VALVE EXPERIENCE AT THE PDU

3.4.1 General

At the Process Development Unit, fines collection and recycle is accomplished via a two-stage cyclone system utilizing a non-mechanical (i.e. stand pipe) configuration for the primary recycle and an eductor-type system for the secondary recycle. The non-mechanical configuration was first installed and operated during October 1982, prior to that the primary recycle was via a lockhopper arrangement.

From October 1982 through December 1983, eight gasifier tests were conducted with the non-mechanical valve recycle system in operation, (TP-033-2, TP-033-3, TP-I-83-1, TP-034-1, TP-0-34-2, TP-034-3, TP-035-1, and TP-035-2). During the initial tests, the emphasis was on startup and obtaining steady recycle operation, while emphasis in later tests was on performance evaluation via data analysis. (Data from the most recent tests, TP-035-1 and TP-035-2, were not intended to be included in this evaluation, however some TP-035-1 data were included due to the limited non-mechanical valve operating data previously available.)

3.4.2 Operational Modes

The PDU primary cyclone was designed to operate as a "classical" L-valve. This means the level of fines in the standpipe would be maintained at a sufficiently high level to insure that the collected fines would be recycled to the gasifier in an unfluidized (packed-bed) mode. Operational experience during 1982-83, however, has demonstrated difficulty in maintaining reliable (steady) recycle of fines with the packed bed due to difficulty with the fines level detection, control response, and bridging of fines in the standpipe. Two additional modes of standpipe operation have been observed during PDU operation: fluidized and flushing. Flushing is described as a combination of packed and fluidized operation.

3.4.2.1 Packed-Bed Mode

The packed-bed mode of dipleg operation was attempted during all the 1982-1983 gasifier tests. Difficulty in maintaining continuous operation resulted in the longest period of stable packed-bed operation being approximately nine hours. The packed-bed mode is characterized by maintaining the top of the fines level in the range of a nuclear densitometer (45 feet). The densitometer detects the level and varies the aeration gas to control the fines recycle rate. Several difficulties were experienced with this operation including slow control system response resulting in varying fines level and recycle rates, difficulty with the calibration of the densitometer which caused inaccuracies in level detection, plugging of pressure transmitter taps, bridging in the standpipe due to the low solids velocity and narrow pipe diameter, and condensation on the fines resulting from high heat loss from the standpipe. The bridging problem was the greatest hindrance in obtaining steady-state operating data; however, this is seen as a problem of scale (i.e., the small pipe diameter and low solids velocity result from low PDU elutriation rates). Large-scale standpipes could conceivably operate in a packed bed mode.

3.4.2.2 Fluidized Mode

During recent PDU tests, the fines recycle system has been operated in a fluidized mode, with continuous recycle of fines occurring for periods of up to 65 hours. The fluidized mode is characterized by increasing the aeration gas to fluidize the fines without attempting to control the fines level. The fines leg then operates in a fluidized mode with the level below the detection range of the densitometer. Fines recycle to the gasifier has been confirmed via gasifier and standpipe temperature profiles. The fluidized mode has several advantages, such as the temperature profile in the standpipe is more uniform (and higher) than the packed-bed mode, and the fines level is self regulating. The fluidized mode of operation can, however, lead to slugging in a narrow standpipe (such as the PDU). Reliable pressure measurements of the standpipe are

infrequent due to pluggage of the pressure taps, therefore, it is not known if there is slugging. Another disadvantage to the fluidized mode is the limited operational data available for analysis.

3.4.2.3 Flushing Mode

A third mode of operation has been identified from the PDU tests and has been labeled the flushing mode. The flushing mode appears to be a combination of the packed-bed and fluidized modes resulting from stagnation of solids due to low aeration or recycle rates. The flushing mode is characterized by the level of packed solids increasing above the packed-bed set point resulting in increased aeration. The level suddenly drops below the detectable level (probably to the fluidized level), the aeration then decreases and the packed bed is re-established. This cycle occurs in approximately 100 seconds. This mode is not recommended for recycle since the recycle rate is cyclic.

3.4.3 Aeration Gas Requirements

Figure 3-2 presents a comparison of the aeration gas rate to the solids recycle rate for the non-mechanical valve recycle operation during TP-034-2, TP-034-3, and TP-035-1. A line is also shown to illustrate the "predicted" aeration requirement (assuming direct volume replacement-- $1 \text{ ft}^3 \text{ gas}/1 \text{ ft}^3 \text{ solids}$). In actual "L" valve operation, the aeration requirement (per pound of solids) decreases as the solids rate increases so that the line would begin to curve upward possibly through the data points.

Additionally, the aeration rates are average values which may have varied by up to 50 percent during the steady state. Figure 3-2 does show, however, that the required aeration is not widely different from predictions, and can therefore be extrapolated with some confidence.

3.4.4 Pressure Recovery

A comparison was made between the measured system pressure drops (i.e. gasifier, transfer line, and cyclone) and the measured standpipe pressure drop (for TP-034-2, and TP-034-3) with the results presented in Table 3-7. If it is assumed that half the cyclone loss must be recovered by the standpipe, then the standpipe pressure drop and the total system pressure loss are approximately equal, indicating that the major system pressure losses have been accounted for in the standpipe design. In addition, the height of fines in the standpipe has been calculated, based upon the measured standpipe pressure drop and the measured fines density, and are listed in Table 3-7. The average calculated fines height (fluidized) is approximately 10 feet, which is much less than the packed-bed height of approximately 45 feet. It is likely, however, that the actual fluidized height is higher, due to bed expansion. Also, the packed-bed height was based on a "worst case" design, i.e., maximum pressure drop and minimum fines density.

Figure 3-2. Non-Mechanical Valve Aeration Requirements

TABLE 3-7
COMPARISON OF BED PRESSURE DROP TO CHAR LEG PRESSURE DROP

TP-034-1				
<u>Set Point</u>	<u>2A</u>	<u>2B</u>	<u>2C</u>	<u>2D</u>
Bed Pressure Drop, psid	0.80	0.80	0.90	0.80
Transfer Line Drop, psid	0.19	0.27	0.21	0.21
Cyclone Pressure Drop, psid	1.49	1.07	1.68	1.70
Total (Bed + Transfer + 1/2 Cyclone)	1.74	1.61	1.95	1.86
Measured Char Leg ΔP , psid	1.44	1.60	1.66	1.83
Calculated Fines Height, ft	9.4	10.2	10.7	11.5
Fines Bulk Density, lb/Ft ³	22.0	22.6	22.4	23.0

TP-34-3					
<u>Set Point</u>	<u>1A</u>	<u>1B</u>	<u>1C</u>	<u>1D</u>	<u>1F</u>
Bed Pressure Drop, psid	0.81	0.84	0.83	0.84	0.85
Transfer Line Drop, psid	0.06	0.04	0.02	0.01	0.06
Cyclone Pressure Drop, psid	2.00	1.95	2.01	2.10	2.04
Total (Bed + Transfer + 1/2 Cyclone)	1.87	1.86	1.86	1.90	1.93
Measured Char Leg ΔP , psid	1.74	1.99	1.96	1.99	2.23
Calculated Fines Height, ft	11.4	13.6	11.8	11.5	12.8
Fines Bulk Density, lb/ft ³	22.0	21.0	24.0	25.0	25.0

3.4.5 Conclusions

Based upon the limited operational experience with the non-mechanical valve system, the following conclusions can be made:

- The non-mechanical valve system can successfully recycle fines to the gasifier bed.
- The non-mechanical, valve system has demonstrated recycle of fines to the gasifier in three distinct modes of operation.
- The non-mechanical valve system has demonstrated the ability to recover from upsets (i.e., loss of level).

3.4.6 Recommendations

The following recommendations are made based on the non-mechanical valve operation.

- Operate the non-mechanical valve recycle system in a fluidized mode due to ease and smoothness of operation.
- Additional testing of the non-mechanical valve fines recycle system is necessary to gather data during long duration steady state operation.

SECTION 4.0

DEPOSIT CONTROL

Buildup of mineral matter in the gasifier-to-cyclone spool piece, primary cyclone barrel and downcomer, and in some cases the secondary cyclone, has been observed in many single-stage PDU gasifier tests. This buildup has been so severe in some cases as to almost totally block the gasifier-to-cyclone spool piece passage, necessitating shutdown of the test. The source of these deposits is the ash in the coal feedstock. Rate of deposit formation varies significantly from coal to coal. Certain PDU operating parameters are also believed to be a factor in the rate of buildup.

Four areas which need to be expanded to predict and control deposit formation are:

- A. To gather data and identify mechanisms for deposit formation. One of the key elements in developing methods to prevent cyclone deposit formation is to determine how these deposits are formed by collecting and analyzing the composition of deposit samples. From these analyses and a knowledge of the major constituents of the ash in the feedstock, a mechanism for deposit formation can be postulated and an approach to circumventing the mechanisms can be formulated.
- B. To investigate the effects of coal ash composition and gasifier operating conditions on the deposition potential of various feedstocks. The rate of formation of cyclone deposits is believed to be related to gasifier operating conditions and the ash composition of various feedstocks. Some coals form deposits much faster than others. Correlations between gasifier operating conditions, ash composition of the feedstock, and rate of deposit formation are still being investigated.
- C. To provide input into the PDU test program so that the data base on cyclone deposition can be expanded. Additional PDU testing is necessary to confirm deposit formation mechanisms and to test control devices. The PDU test program must be tailored to provide this data within the constraints of other test objectives.
- D. To develop and test component hardware capable of preventing cyclone deposition without compromising the thermal efficiency of the process. Devices have been tested and proposed for testing that are intended to prevent deposit formation. Modifications of previously tested devices and testing of new devices are necessary to confirm their effectiveness.

4.1 BACKGROUND

Deposition of fines in the outlet pipe of the gasifier and cyclone barrel has been experienced since the inception of the oxygen-blown gasifier development program in mid-1978. During the 1978 and 1979 test programs, deposition leading to run termination was evidenced during all tests using Pittsburgh No. 8 seam coal at gasifier freeboard temperatures of 1800 to 1850°F. Samples of deposit material, Tables 4-1 and 4-2, consistently indicated a high concentration of ash (greater than 90 percent) thereby suggesting that deposition occurs by one or both of the following mechanisms:

- A. Certain constituents of the coal ash form low-melting eutectics on the surface of the char particles. Char fines that are carried over from the gasifier impinge on the impact surfaces of the cyclone barrel, causing adhesion of the eutectic material of the char fines to the refractory wall. The residual carbon in the deposit is then gasified, leaving a matrix of fused mineral matter as a deposit.
- B. Low melting minerals in the coal ash are released from the char during reactions in the gasifier. These liquid droplets are carried over and adhere to the refractory wall of the cyclone.

4.2 PDU EXPERIENCE WITH DEPOSIT CONTROL METHODS

To date, three concepts for deposit control have been tested:

- Water Quench
- Cyclone Cold Wall
- Particle Through-trap

Table 4-3 lists the PDU tests that included deposit control methods and summarizes observations on the effectiveness of those methods.

4.2.1 Water Injection

In this deposit control option, water was injected into the gasifier freeboard to cool the product gas stream to a temperature below the melting point of the binding material, thereby freeze-cooling the sticky particles before they entered the cyclone. Water injection was first tested in TP-023-1 with Ohio No. 9 coal. This coal had a high pyritic sulfur content and was expected to cause massive deposits, rich in iron sulfides. Only a thin layer of ash was deposited in the primary cyclone after TP-023-1 (Figure 4-1); however, an x-ray analysis of deposit samples showed no iron-sulfide (Table 4-2). This initial success in reducing deposits with

TABLE 4-1
DEPOSIT FORMATION IN THE CYCLONE

<u>Conclusions</u>	<u>Supporting Analyses</u>
Mechanism of particle arrival at the cyclone wall: inertial flight and impaction	- Particle size analysis ($r_p \cong 30 \mu$) - Deposit pattern (build-up appears to have started from opposite wall to the inlet pipe and grown back toward the inlet)
Mechanism of particle adhesion: molten, sticky FeS and possibly eutectics of K-Al-Si oxide systems acting as a binding matrix for ash build-up	- X-ray, microprobe and chemical analyses of deposit samples - Phase diagrams of FeO-FeS and K_2O-SiO_2 systems - Thermodynamic calculations indicating the presence of K_2SiO_3 and $KAlSi_3O_8$ in the gasifier

TABLE 4-2

SUMMARY OF X-RAY ANALYSES OF SAMPLES OF CYCLONE DEPOSITS

PITTSBURGH SEAM COAL -- (TP-018-2; TP-022-2; TP-023-2)

Major - SiO_2 ; Fe_{1-x}S ; FeS ; Fe_2O_3 ; $(\text{Mg Fe})_2\text{SiO}_4$ - ring
woodite; $(\text{Fe})_2\text{SiO}_4$ - fayalite; $(\text{Al}_{1.96}\text{Fe}_{0.04})_2\text{O}_4$ -
spinel

Minor - FeS_2 ; $(\text{Fe})_2\text{SiO}_4$; $(\text{Mg Fe})_2\text{SiO}_4$; $\text{CaAl}_2\text{Si}_2\text{O}_8$ -
anorthite

Trace - CaSO_4 and MgO

WESTERN KENTUCKY COAL -- (TP-027-4)

Major - FeS

Minor - Fe_2O_3

OHIO NO. 9 COAL AND WATER INJECTION -- (TP-023-1)

Major - Fe_2O_3

Minor - SiO_2

Trace - $(\text{Fe})_2\text{SiO}_4$; $(\text{Mg Fe})_2\text{SiO}_4$

INDIANA NO. 7 COAL -- (TP-028-2) - Primary and secondary cyclone
deposits.

Major - $(\text{Fe})_2\text{SiO}_4$; SiO_2 ; FeS (only in through-trap deposit)

Minor - $(\text{Fe, Mg})\text{SiO}_3$; $(\text{Mg, Fe})\text{Si}_2\text{O}_4$ - spinel

RSA COAL -- (TP-031-2) - Deposits from primary and secondary cyclone
downcomers.

Major - $(\text{Na, Ca})(\text{Al, Si})_4\text{O}_8$ - bytownite

Minor - not identified

WEST VIRGINIA COAL -- (TP-030-3) - Primary and secondary cyclone
deposits.

Major - $\text{Fe}_2(\text{SiO}_4)$ or FeAl_2O_4 - spinel; SiO_2

Minor - $\text{Fe}_2(\text{SiO}_4)$; Al_2SiO_5

Trace - $\text{Mg}_{.17}\text{Fe}_{.83}(\text{SiO}_3)$ - ortho-ferrosilite

TABLE 4-3

DEPOSIT CONTROL METHODS TESTED AT THE KRW PROCESS DEVELOPMENT UNIT

<u>PDU Test</u>	<u>Date Month/Year</u>	<u>Control Method</u>	<u>Coal Feed</u>	<u>Observations on Deposit</u>
TP-023-1	7/79	Water injection	Ohio No. 9	Very small deposit in primary cyclone.
TP-025-2	11/79	Water injection	Indiana No. 7	No deposit observed.
TP-023-3	12/79	Water injection	Pittsburgh	Cyclone plug; deposits in primary cyclone and inlet pipe.
TP-023-4	3/80	Water injection and cold-wall insert	Pittsburgh	No deposit observed.
TP-027-1	4/80	Water injection and cold-wall insert	Indiana-Ohio Blend	No deposit observed.
TP-027-2	5/80	Water injection and cold-wall insert	Ohio No. 9 and Pittsburgh	No deposit observed.
TP-027-3	6/80	Water injection and cold-wall insert	Texas lignite and Pittsburgh	No deposit observed.
TP-027-4	8/80	Reduced water injection and cold- wall insert	Wyoming W. Kentucky	Cyclone plug; deposits in primary cyclone and inlet pipe.
TP-027-5	10/80	Water injection and cold-wall insert	Texas Lignite Wyoming	No deposit observed.
TP-031-1	12/80	Intermittent water injection and cold- wall insert	RSA Coal	No deposit observed.
TP-029-1/ TP-030-1	4/81	Through trap; intermittent water injection and cold- wall insert	Illinois No. 6 and Wyoming	No deposit observed.

TABLE 4-3 (Continued)

DEPOSIT CONTROL METHODS TESTED AT THE KRW PROCESS DEVELOPMENT UNIT

<u>PDU Test</u>	<u>Date Month/Year</u>	<u>Control Method</u>	<u>Coal Feed</u>	<u>Observations on Deposit</u>
TP-030-2	5/81	Through-trap; intermittent water injection and cold- wall insert	Indiana No. 7 RSA Coal	No deposit observed in cyclones; some deposit in through- trap.
TP-028-1	6/81	Through-trap; water injection for Pittsburgh coal and cold-wall insert	Pittsburgh and Wyoming	No deposit observed in cyclones; through-trap partially plugged.
TP-028-2	7/81	Through-trap removed after first set point; reduced water injection and cold-wall insert	Indiana No. 7	Deposits in spool piece, primary and secondary cyclones; no deposit on surface of cold-wall insert.
TP-032-1	4/82	Full cold-wall cyclone	Upper Freeport	No deposit observed.
TP-032-2	5/82	Full cold-wall cyclone	Upper Freeport and Western Kentucky	No deposit observed.
TP-033-1	7/82	Full cold-wall cyclone	Brookville and N. Dakota Lignite	No deposit observed.
TP-033-2	10/82	Second design through-trap	Wyoming	Preliminary results; no deposit observed.
TP-033-3	12/82	Second design through-trap	W. Kentucky	Preliminary results; no deposit observed.

Figure 4-1. Deposit Pattern in Primary Cyclone and Spool Piece

water injection tended to support the conclusion that iron sulfides were entrained by the product gas as sticky particles which struck the cyclone walls forming a binding matrix. Subsequent tests with high-fouling coals (Pittsburgh and Western Kentucky), as well as with moderate or low-fouling coals (Indiana, RSA, Texas lignite, Wyoming Sub-C), proved water injection to be a successful, reliable method of controlling cyclone deposit formation.

The parameter of great importance in the application of this control method was the product gas temperature after water injection. A review of PDU operating data indicated the presence of a "threshold" product gas temperature below which deposits would not form. In Figure 4-2, the product gas temperatures are plotted against the pyritic sulfur contents of the coal feeds. The open circles and triangles, representing data points taken during tests without deposits, lie to the left of the dashed line, which can be taken as the locus of the threshold temperatures of the various coals tested in the PDU. The solid line represents the locus of the temperatures above which deposits would form with the coals of corresponding pyritic sulfur contents. Between the two lines is the uncertainty band caused by the uncertainty in the data collected in the PDU. Threshold temperature was shown to be low for high pyritic sulfur coals and became higher as the pyritic sulfur content in the coal decreased. When water injection was used as a deposit control method, the amount of water injected into the gasifier freeboard would have to be appropriate for cooling the product gas to below the threshold temperature of the corresponding coal feed. As shown in Figure 4-3, during a few PDU tests with water injection (TP-023-2, TP-027-4, TP-028-2, and TP-031-1), the product gas temperature was allowed to exceed the threshold temperature through reduction in the amount of water injected, resulting in cyclone deposits.

4.2.2 Cold-Wall Cyclone

The concept of the cold-wall cyclone was developed on the basis of observations by Moza and Austin⁽¹⁾ that "a liquid slag droplet falling onto a metal surface sticks to the surface if the metal temperature is above a critical temperature and does not stick if it is below this temperature," which is called the "sticking temperature," T_s .

The physical process occurring in the adhesion or non-adhesion of the droplet can be tentatively described as follows. The droplet is still molten when it strikes the metal, but the contact with the metal causes an immediate cooling and solidification. The surface tension of slag-metal-air creates an adhesion and there may be chemical reaction, but

- (1) MOZA, and Austin, "A New Test for Characterizing the Slag Deposition Properties of a Coal Ash: The Sticking Temperature," Journal of the Institute of Energy, March 1979.

Figure 4-2. Effect of Product Gas Temperature on Cyclone Deposits

Figure 4-3. Product Gas Temperatures Recorded During PDU Tests With Water Injection

the solidification causes stress due to contraction or expansion of the material bonded to the surface. When the metal temperature is above T_s , the adhesion force is strong enough to allow a continuing bond to the surface after solidification is complete.

4.2.2.1 PDU Experience with Cold Wall

A water-cooled cyclone wall insert was designed and installed in the PDU for test TP-023-4 in March 1980. The insert was made of 310 stainless steel and covered a 7 x 11-inch area on the cyclone barrel, directly opposite the cyclone inlet pipe (Figure 4-4). It was embedded into the refractory of the cyclone wall, and its water-cooled metal wall was kept at a temperature estimated to be 600°F (316°C) or lower. Since water injection also had been used in test TP-023-4, there was no deposit in the cyclone and the cold-wall insert was clean. The insert remained in the cyclone for subsequent PDU tests and its surface remained free of deposit after every test. After 711 hours of PDU operation the insert was removed and the loss of wall thickness caused by erosion was measured as about 0.015 inch. A replacement cold-wall insert of similar design was installed in the cyclone in test TP-027-4 and was operating until it was damaged in test TP-028-2. (As a result of a site power failure, the water-circulation pump failed, and stagnant water turned to steam. The resulting steam pressure ruptured the insert wall.) A third cold-wall insert replaced the damaged one and remained in the cyclone until the installation of the full cold-wall cyclone in TP-032-1, April 1982.

The cold-wall inserts were used primarily to test the concept of freeze-cooling of sticky particles and were often operated concurrently with primary deposit control, such as water injection or the through-trap. In tests where water injection was not used, however, or where the amount of water injected was reduced so that ash deposits were formed in various areas of the cyclones and spool pieces (TP-027-4, TP-028-2, TP-031-1, and TP-031-3), the metal surface of the cold-wall inserts remained clean, thus validating the freeze-cooling concept.

Because of the success of the cold-wall inserts in avoiding ash buildup, a full cold-wall cyclone was designed and operated as a primary deposit control device during three PDU tests: TP-032-1 with Indiana and Upper Freeport coals, TP-032-2 with Upper Freeport and Western Kentucky coals, and TP-033-1 with Brookville and North Dakota coals. With the exception of Indiana coal and lignite, the coals used in these tests had high pyritic sulfur contents and were expected to be highly fouling. The full cold-wall cyclone included four sections, covering the gasifier outlet, the spool piece, the cyclone inlet, and the cyclone barrel (Figure 4-5).

A summary of the number of hot hours of cold wall testing in the PDU is as follows:

Figure 4-4. Diagram of Primary Cyclone (C-119) With Water-Cooled Wall Insert

Figure 4-5. Diagram of Full Cold Wall Cyclone

<u>PDU Test</u>	<u>Total Hot Hours of Operation</u>	<u>Total Coal Feed Hot Hours</u>
TP-032-1	162.8	13.5 (Indiana) 97.3 (Upper Freeport)
TP-032-2	155	53.5 (Upper Freeport) 95.5 (W. Kentucky)
TP-033-1	204	103.3 (Brookville) 67.5 (N. Dakota Lignite)
Total	521.8	430.6

4.2.2.2 Surface Metal Temperature Calculation

It is known that deposit build-up on the metal surface can be prevented by maintaining the surface temperature in contact with product gas below a specific temperature. This temperature was considered the most important variable in cold wall operation. Thermocouples embedded in the wall of the cold wall were originally intended to provide an indirect measurement of the metal surface temperature in contact with the gas. However, at least one of the four embedded thermocouples was known to be providing erroneous measurements. This was indicated by the fact that in some cases the temperature measured midway through the metal wall was actually below the average water temperature in that particular cold wall section. Since at least one of the thermocouples was providing misleading data, it was decided not to use any of these thermocouples to determine the surface metal temperature in contact with product gas.

Surface metal temperature of the cold wall was calculated on-line using a computer program based on gas side and water side heat rates. For the water side the following equation was used:

$$Q_w = M_w C_{p_w} (T_{wo} - T_{wi}) \quad (4-1)$$

where:

Q_w = Water side heat rate
 M_w = Mass flow rate of water
 C_{p_w} = Heat capacity of water
 T_{wo} = Temperature of water exiting cold wall
 T_{wi} = Temperature of water entering cold wall

M_w , T_{wo} , and T_{wi} were measured. C_{p_w} is known and assumed to be constant.

The water side and gas side heat rates should be equivalent for the cold wall. Therefore, the value of Q_w calculated above was substituted into the following gas side heat rate equation to determine the overall heat transfer coefficient.

$$Q_g = Q_w = UA\Delta T$$

where:

- Qg = Gas side heat rate
- U = Overall heat transfer coefficient
- A = Heat transfer area of cold wall (gas side)
- ΔT = Gas side temperature drop

Rearranging:

$$U = \frac{Q_g}{A \Delta T} \quad (4-2)$$

The temperature drop (ΔT) was determined as follows:

$$Q_g = (M_g C_{pg} + M_f C_{pf})(T_{gi} - T_{go})$$

where:

- Mg = Mass flow rate of product gas
- Mf = Mass flow rate of elutriated fines
- Cpg = Heat capacity of product gas
- Cpf = Heat capacity of fines
- Tgi = Temperature of gas entering cold wall
- Tgo = Temperature of gas exiting cold wall

Mg, Mf, and Tgo were measured values. Cpg and Cpf are known (for a particular composition). Qg was assumed to be equal to Qw.

Equation (4-2) was then rearranged to determine the gas inlet temperature.

$$T_{gi} = T_{go} + \frac{Q_w}{(M_g C_{pg} + M_f C_{pf})} \quad (4-3)$$

The temperature drop is then:

$$\Delta T = T_{gi} - T_{go}$$

which is substituted into Equation (4-2) to determine the overall heat transfer coefficient.

The water side heat transfer coefficient was then calculated as follows:

$$h_w = 0.023 \left(\frac{K_w}{D_e} \right) (Re)^{0.8} (Pr)^{0.33} \quad (4-4)$$

where:

- hw = Water side heat transfer coefficient
- Kw = Thermal conductivity of water
- De = Hydraulic diameter
- Re = Reynold's number
- Pr = Prandtl number

Kw is constant at a given temperature.
 De was calculated from cold wall dimensions.
 Re was calculated based on water flow rate.
 Pr is constant at a given water temperature.

The gas side heat transfer coefficient was then calculated as follows:

$$\frac{1}{U} = \frac{1}{h_g} + \frac{1}{h_w} + \frac{t}{K} \quad (4-5)$$

where:

h_g = Gas side heat transfer coefficient
 K = Thermal conductivity of the metal
 t = Metal wall thickness

Rearranging:

$$h_g = \frac{1}{\frac{1}{U} - \frac{1}{h_w} - \frac{t}{K}} \quad (4-6)$$

U and h_w were calculated above.
 t was known
 K is constant at a given temperature.

The gas side metal surface temperature was then calculated from the following equation:

$$Q_g = Q_w = h_g A (T_g - T_m) \quad (4-7)$$

where:

T_g = Average gas temperature

$$T_g = \frac{T_{gi} + T_{go}}{2}$$

$$T_m = \frac{T_g - Q_w}{h_g A}$$

The difference between the gas side and water side heat rates was used as an indication of the accuracy of this calculational method.

$$Q_w = M_w C_{pw} (T_{wo} - T_{wi}) \quad (4-8)$$

$$Q_g = (M_g C_{pg} + M_f C_{pf})(T_{gi} - T_{go}) \quad (4-9)$$

$$\% \text{ Error} = \frac{Q_w - Q_g}{Q_w} \times 100$$

The online computer calculation used during the PDU tests assumed that the Prandtl number was constant and the water viscosity and thermal conductivity did not vary with water temperature. However, following completion of all cold wall testing, the data were reevaluated using more precise calculations. Specifically, the Prandtl number, water viscosity, and water thermal conductivity were compensated for temperature variation. This refinement in the calculation of the gas side metal wall temperature resulted in temperatures approximately 15 percent higher in comparison to the online computer calculations. Only the results from the more precise calculation are presented.

4.2.2.3 Cold Wall Operating and Performance Parameters

A summary of cold wall operating parameters for the three PDU tests is shown on Table 4-4. The first PDU test (TP-032-1) was a proof of concept test. Therefore, the surface metal temperature was maintained conservatively low because of the unknown deposit threshold temperature. This temperature was increased the following two tests in an effort to identify the temperature at which deposits begin to stick to a metal surface. Maximum surface metal temperature achieved during any one-hour period of the three PDU tests was 742°F. This occurred from 1830 to 1930 hours on 8-11-82 (TP-033-1). Average surface metal temperatures over the duration of the tests were significantly below the maximum. This variation between the maximum and average surface metal temperature was due predominantly to changes in PDU operating conditions which resulted in higher product gas or fines elutriation rates. The cold wall surface metal temperature was not adjusted until steady PDU conditions were achieved. This resulted in short periods where the surface metal temperature exceeded pre-established limits. Temperature profiles of the product gas/fines, surface metal, and cooling water are shown in Figures 4-6, 4-7, and 4-8 for each PDU test.

Deposition did not occur on any of the cold wall sections during any of the three PDU tests. Neither did deposits occur downstream of the cold wall. As shown on Table 4-4, the product gas/fines temperature drop across the four cold wall sections exceeded 400°F in each test. Thus the product gas/fines temperature was below 1350°F at the exit of the cold wall. This is below the deposit "sticking" temperature for each of the coals tested.

Average gas film heat transfer coefficients for each cold wall section are also shown in Table 4-4. The heat transfer coefficient for cold wall section A is significantly higher than those for sections B, C, and D. This is due to the greater gas impact on the surface created by the cyclone swirl in this section.

TABLE 4-4

SUMMARY OF COLD WALL OPERATION PARAMETERS

	<u>TP-032-1</u>	<u>TP-032-2</u>	<u>TP-033-1</u>
Maximum Gas Side Metal Surface(a) Temperature (°F)	479	665	742
Average Gas Side Metal Surface(b) Temperature (°F)			
Section A	398	525	538
Section B	330	406	420
Section C	354	416	412
Section D	401	521	556
Average Temperature Drop Across Cold Wall (°F)	479	491	454
Average Gas Inlet Temperature (°F)	1786	1840	1745
Average Gas Film Heat Transfer Coefficient (Btu/hr x ft ² °F)			
Section A	107	137	160
Section B	64	65	77
Section C	69	74	79
Section D	76	95	119
Average Gas Side/Water Side Heat Rate Difference (%)	12	5	3
Average Fines Load (lb/hr)	364	409	803
Average Cyclone Inlet Velocity (ft/sec)	45	49	58
Coal Feed Hours	111	149	171

(a) Maximum temperature during any one-hour period.

(b) An average of hourly averages over the entire test.

Figure 4-6. Cold Wall Temperature Profiles - TP-032-1

Figure 4-7. Cold Wall Temperature Profiles - TP-032-2

Figure 4-8. Cold Wall Temperature Profiles - TP-033-1

4.2.2.4 Correlative Cold Wall Experience

All test results to date prove that the cold-wall or water-jacketed external cyclones can be operated downstream of the gasifier without ash buildup on the cyclone walls. Rate of erosion of the cyclone walls, however, remains to be determined, as does the problem of ash deposition downstream of the cooled cyclone when the temperature of the outlet gas from the cyclone was greater than the deposit threshold temperatures shown in Figure 4-2.

4.2.3 Through-Trap

The through-trap was designed to provide a number of cooled surfaces of impaction in the path of the sticky particles of binding material. Rendered nonsticky through freeze-cooling, the particles would no longer be able to form deposits on the cyclone wall. Successful performance of the through-trap as a deposit control device was based on three major assumptions:

- Through-trap design would allow for impaction at its cooled surfaces of all sticky particles (large and small) carried over in the gas stream.
- Cooled surfaces would be kept at temperatures low enough to freeze-cool all sticky particles.
- Should the gas temperature at the outlet of the through-trap be higher than the deposit threshold temperature shown in Figure 4-2, frozen, nonsticky particles would not be reheated and melt in the gas stream as they traveled to the cyclone.

4.2.3.1 PDU Experience with Through-Trap

The through-trap was first tested in the PDU during TP-029-1 and TP-030-1, with Illinois and Wyoming coals. Subsequent PDU tests with the through-trap included TP-030-2, with Indiana No. 7 and RSA coals; TP-028-1, with Pittsburgh seam and Wyoming coals; and TP-028-2, with Indiana No. 7 coal. Of the coal feedstocks used, Illinois and Pittsburgh seam were expected to be moderately to highly fouling, while Indiana No. 7, RSA, and Wyoming were ranked as moderately to low fouling.

The first test of 1981 (TP-029-1/030-1) was carried out using Illinois No. 6 and Wyoming coals as the feedstock. During the initial portion of the test (Illinois No. 6) the ΔP (pressure drop) increased across the through-trap in several instances. The increase was reduced to normal levels by turning on the water injection into the gasifier. The second portion of the test with Wyoming coal had no pressure increases and water injection was not used. Post-test inspection of the through-trap, gasifier outlet piping and the cyclones revealed no evidence of deposition in the system.

The third and fourth tests of 1981 (TP-028-1 and 028-2, respectively) showed results similar to TP-030-2. Water injection was required shortly after startup and used throughout the Pittsburgh seam portion of the test. Wyoming coal was used in the latter segment of the run and no water injection was required. TP-028-1, with Indiana No. 7 as the feedstock, showed the same results as the previous tests. During the first set point, water injection was used to control pressure increases across the through-trap. The test was shut down after the first set point due to a gasifier flange leak. During the shutdown, inspection of the through-trap revealed a water leak in the tube bundle that could not be repaired. The through-trap was removed and the test was restarted using water injection to prevent deposition downstream of the gasifier. During the last day of the test, water injection was turned off. After approximately 8 hours, the pressure increase across the gasifier and cyclone required that the water injection be turned on to prevent further pressure rise.

This second through-trap was tested in the PDU during TP-033-2 with Wyoming coal and TP-033-3 with Western Kentucky coal. Results of these tests are currently being analyzed, but post-test inspections revealed no ash deposit in the through-trap nor downstream from it.

To date, results using the through-trap have been inconclusive. The assumptions that formed the basis for its successful performance as a deposit control device have not been fully verified, since with the first design the through-trap was used concurrently with water injection, when the coal feeds were moderately to highly fouling; and with the second design, the time of operation with the highly fouling Western Kentucky coal was too short for deposit accumulation. A long-duration PDU test with a highly fouling coal and a high temperature is needed to prove that the through-trap could be an effective deposit control device, especially when the gas temperature at the trap outlet is higher than the deposit threshold temperature of the coal feed.

SECTION 5.0

ENVIRONMENTAL, HEALTH AND SAFETY CONSIDERATIONS

While environmental, health and safety considerations have been a concern during the development of the Westinghouse coal gasification process, the paramount effort, in common with most development projects, has been on process and operating controls and refinements. Thus, qualification of the permissible released concentrations of possible priority gaseous, liquid, and solid waste pollutants, in view of all environmental, health and safety aspects, remains somewhat incomplete at this time. Issues which must be addressed are:

- A. Thorough assessment of trace elements in the product gas and particulate cleaning systems.
- B. Identification and quantification of regulated and unregulated pollutants in gas, waste water, and solid streams.
- C. Acquisition of a more complete understanding of the PDU chemical reactions which will lead to the development of a sound thermodynamic model that can be used to accurately predict stream compositions prior to gas quench.

5.1 BACKGROUND

The product fuel gases generated during gasification, consisting primarily of H_2 , CO , CH_4 , CO_2 , H_2O , N_2 , H_2S and COS , are generally considered to also contain contaminants such as alkali and trace metals and particulates. Emission of the major species, metals or particulates is assessed as a potential environmental, health or safety concern when the quantity released approaches primary pollutant levels. A separate Environmental, Health and Safety Topical Report, FE-19122-23, has been issued in preliminary form which details analytic and process data obtained from the PDU over approximately two and a half years, starting in mid-1980, in an effort to establish relationships between the data and the initial feed composition, and system operating parameters.

Ash agglomerates, produced by the PDU process, and the PDU recycle cooling water with its accumulation of coal fines were analyzed to assess these materials against the toxicity standards established under the Resource Conservation and Recovery Act (RCRA). Through leachability tests the evaluation established that neither were considered hazardous by EPA definition.

Identification of the chemical species and compounds that are of environmental, health, and safety (EH&S) concern during the operation of the PDU is essential not only to monitor the work performed at the PDU, but also to verify and support modeling studies of the gasifier. Developing a comprehensive data base is a long term objective of PDU operations.

Knowledge of those species likely to be of EH&S importance in large-scale facilities provides a basis for selecting the parameters of concern at the PDU. In large-scale facilities, the chemical parameters of concern are expected to be present in controlled discharges and emissions, intermediate streams, and fugitive emissions. Future monitoring at the PDU will address these identified parameters.

The approach taken is to identify both regulated and unregulated pollutants of concern for each media separately (i.e., gas, waste water, and solid streams). Regulated parameters are those which are commonly covered by environmental permits and/or which have been identified in the Clean Air Act, Clean Water Act, Resource Conservation and Recovery Act (RCRA) or Occupational Safety and Health Act (OSHA) regulations. Unregulated parameters can also be of concern if they are toxic or hazardous and have a potential to occur in coal gasification effluents or intermediate streams.

Large-scale EH&S concerns can be assessed at the PDU only if the streams of interest exist at the test facility and if scaleable information can be obtained. Comparison of the PDU process flow scheme with that for a typical large-scale project using the KRW gasifier indicates that there are major differences between these facilities. It is important to recognize the differences so as to maximize the value of all work done at the PDU. Key differences include:

- Feedstock rate

PDU	24 tons/day
Large-scale	2400 tons/day

- Gas handling

PDU - Raw gas leaving the secondary cyclone at the PDU is sent to the quench scrubber for residual fines removal, gas cooling, and removal of ammonia, sulfur species, and organics from the gas. The quench scrubber will be replaced with a venturi scrubber and a slurry filter for solids concentration and removal. Gas leaving the quench scrubber is further cooled and scrubbed in the product gas cooler. Cooled gas leaving the product gas cooler is split into two streams: the major portion of the gas is incinerated in a natural-gas-fired thermal oxidizer; the smaller gas stream is used for recycle gas after additional cooling and recompression. Flashed gases containing organic compounds, sulfur species, and ammonia are sent to the thermal oxidizer for incineration. A portion of the water in the separator pit is also pumped to the thermal oxidizer for incineration.

Large Scale - In most designs proposed, raw gas from the cyclones is first cooled in a waste heat boiler. Particulate scrubbing is accomplished with a venturi scrubber and the fines are concentrated in a pressurized filter.

- Gas processing

PDU - None

Large Scale - The following unit operations are proposed for gasifier downstream processing: sour water stripping and ammonia recovery, CO shift and COS hydrolysis, acid gas removal, and sulfur recovery.

- Waste Water treatment

A major concern of the waste water generated at the PDU is that it may not be truly representative of the waste water that will be produced at a large scale facility, because:

- The equipment at the PDU used to quench, scrub and cool the hot gas is radically different and there are no unit operations for downstream processing.
- The relatively large amount of makeup water entering the quenching/cooling system at the PDU creates a more dilute waste water than would be expected at a large-scale plant.

Gas stream parameters of EH&S concern can be broadly classified into three categories: (1) regulated criteria pollutants, (2) unregulated hazardous or toxic species prevalent in intermediate streams, and (3) unregulated species potentially present in fugitive or controlled emissions. The first category involves particulates, SO₂, CO, NO₂, and O₂, hydrocarbons, and Pb, only a few of which are important at the PDU (i.e., particulates, CO, and hydrocarbons). The second category is concerned mainly with sulfur- and nitrogen-containing species such as COS, CS₂, H₂S, NH₃, and HCN. The final category includes volatile organic compounds, hydrocarbons, and numerous trace elements. Knowledge of other gas stream parameters is also required because of their influence on the overall chemistry of the system.

Gas stream treatment systems at large-scale facilities will be designed to meet existing regulations and provide a safe working environment. Design of appropriate systems will be dependent upon characterizing the raw gas from the gasifier for the design base coal. Certain sulfur species will require particular attention, not only from EH&S aspects, but also from process design considerations. For example, certain acid gas removal processes have difficulty in removing COS, and a hydrolysis unit is

required for conversion of COS to easily remove gases such as H₂S and CO₂. Sulfur species analyses will therefore require closer scrutiny at the PDU in the future since the design basis for treatment systems depends on gasifier product gas analysis. Limited data are currently available on COS, mercaptans, thiophenes, CS₂, and other trace sulfur species from the PDU.

Major gas streams of EH&S interest follow and the relationship between the PDU and a large-scale plant is identified. The parameters of concern are derived from all categories previously discussed.

- Vent gases from coal and ash lockhoppers should be similar in composition between the PDU and a large-scale plant.

Parameters of concern: CO, CO₂, H₂, O₂, N₂, CH₄, H₂O
sulfur species
NH₃, HCN
C₁-C₆ hydrocarbons
volatile organics
particulates

- Raw product gas from the gasifier at the PDU, after passing through cyclones, is quenched in a raw gas quench system. Large-scale designs would likely call for a waste heat boiler and a venturi scrubber after the cyclones for more efficient particulate scrubbing.

Parameters of concern: CO, CO₂, H₂, O₂, N₂, CH₄, H₂O
sulfur species
NH₃, HCN, SCN
volatile organics
condensable organics
trace elements (vapor)
particulates

Gaseous sulfur species in the parameters of concern include: H₂S, COS, CS₂, CH₃SH, C₂H₅SH, (CH₃)₂S, and SO₂. Volatile organics in gas samples include species with six to twelve carbon atoms. Condensable organics include carbon numbers C₈ and greater. Trace elements in the parameters of concern include: As, Sb, Pb, Se, Hg, Al, B, Be, Ca, Co, Cu, F, Fe, K, Mn, Mo, Na, Ni, P, Si, Ti, Tl, V, and Zn.

The major waste water streams of EH&S concern at large-scale gasification facilities will be gas cooling condensates and quench blowdown. These streams should be given primary consideration since they are expected to:

- contain the highest concentrations of environmentally significant parameters;

- contribute significantly to the waste water treatment facility inflow; and,
- pose the greatest source of technical uncertainty to waste water treatment system designers.

In comparison, other possible plant process streams are either relatively clean (i.e., methanation condensate), or only contribute small volumes of waste water (i.e., acid gas removal condensates, sulfur recovery waste water). Successful treatment of non-process waste waters, such as coal pile runoff and storm water drainage, have already been accomplished in a number of related industries.

Parameters of concern for waste water streams include both regulated and unregulated parameters. Table 5-1 presents a listing of these parameters combined into classes or groups where appropriate. Table 5-2 identifies example compounds in other parameter groups which present EH&S concerns. Where possible, potential surrogates for each parameter group are presented based on EPA methodology. A surrogate parameter can be used to represent a parameter group if:

- the presence or absence of a limited number of compounds within a class of organic compounds can indicate the potential significance of that class of compounds in a given stream; and,
- the potential for concern of different compounds in a group is broadly the same.

The surrogate may either be the summation of a group of compounds or may be a single compound.

Regulated waste water parameters include water quality parameters identified by EPA or the individual state, priority (toxic) pollutants specified by EPA, and other parameters found on application forms for waste water discharge permits. Groundwater quality parameters have also been included in this table since they may be of concern if a facility is located near a sole-source aquifer. Unregulated waste water stream parameters include thiosulfate, thiocyanate, organic sulfur and nitrogen species, and other trace organic compounds. These species are of concern since they are hazardous and have been identified during analysis of waste water samples collected at the PDU or at other fluidized-bed pilot plants. Unregulated organic compounds are represented by groups or by surrogate compounds that can be used to monitor a class of compounds. The basis of the surrogate approach and an identification of individual group or surrogate class is presented in Table 5-3.

Two types of solid wastes will be generated at large-scale gasification facilities. The first includes solid waste streams whose composition is unique to the gasification process (i.e., cyclone fines, bottom ash, or

TABLE 5-1
PARAMETERS OF CONCERN: WASTE WATER STREAMS

	<u>REGULATED</u>			<u>UNREGULATED</u>
	<u>Water Quality</u>	<u>Priority (Toxic) Pollutant</u>	<u>Effluent Limitation</u>	
<u>General</u>				
pH	X		X	
TDS	X			
TSS	X		X	
Alkalinity	X		X	
Oil and Grease	X		X	
<u>Nitrogen and Sulfur Species</u>				
NH ₃ -N	X			
NO ₃ -N	X			
NO ₂ -N	X			
SO ₄	X			
SO ₃			X	
S			X	
S ₂ O ₃				X
CN	X	X		
SCN				X
HCN				X

TABLE 5-1 (Continued)

PARAMETERS OF CONCERN: WASTE WATER STREAMS

	<u>REGULATED</u>		<u>UNREGULATED</u>
	<u>Water Quality</u>	<u>Priority (Toxic) Pollutant</u>	<u>Effluent Limitation</u>
<u>Metals and Trace Elements</u>			
Alkalies (e.g., Ca, Mg, Na, K)			X
Toxic Trace Elements (Sb,As,Be,Cd,Cr,Cu, Pb,Hg,Ni,Se,Ag,Th,Zn)	X	X	
Others (B,Ba,Co,Fe,Mn,Mo,V)			X
<u>Organics</u>			
BOD	X		X
COS	X		X
TOC	X		X
Volatile Compounds*		X	
Acid Compounds*		X	
Base-neutral Compounds*		X	
Aldehydes			X
Carboxylic Acids			X
C ₂ -C ₄ Cyanides			X
Amines			X
Phenols			X**
Benzenes			X**

* Individual compounds listed in Table 5-2.

** Some compounds in this group are regulated.

TABLE 5-2

EPA PRIORITY (TOXIC) POLLUTANT CONSTITUENTS BY PARAMETER GROUP

Inorganics:

Sb, As, Be, Cd, Cr, Cu, Pb, Hg, Ni, Se, Ag, Th, Zn, Total cyanides, Total phenols

Organics:

<u>Volatiles</u>	<u>Acid Extractables</u>	<u>Base Neutrals</u>
Acrolein	2-Chlorophenol	Acenaphthene
Acrylonitrile	2, 4-Dichlorophenol	Acenaphthylene
Benzene	2, 4-Dimethylphenol	Anthracene
bis (Chloromethyl) ether	4, 6-Dinitro-o-cresol	Benzidine
Bromoform	2, 4-Dinitrophenol	Benzo(a)anthracene
Carbon tetrachloride	2-Nitrophenol	Benzo(a)pyrene
Chlorobenzene	4-Nitrophenol	Benzo(b)fluoranthene
Chlorodibromomethane	p-Chloro-m-cresol	Benzo(ghi)perylene
Chloroethane	Pentachlorophenol	Benzo(k)fluoranthene
2-Chloroethylvinyl ether	Phenol	bis(2-Chloroethoxy)methane
Chloroform	2, 4, 6-Trichlorophenol	bis(2-Chloroethyl)ether
Dichlorobromomethane		bis(2-Chloroisopropyl)ether
Dichlorodifluoromethane		bis(2-Ethylhexyl)phthalate
1, 1-Dichloroethane		4-Bromophenyl phenyl ether
1, 2-Dichloroethane		Butyl benzyl phthalate
1, 1-Dichloroethylene		2-Chloronaphthalene
1, 2-Dichloropropane		4-Chlorophenyl phenyl ether
1, 3-Dichloropropylene		Chrysene
Ethylbenzene		Dibenzo(a,h)anthracene
Methyl bromide		1, 2-Dichlorobenzene
Methyl chloride		1, 3-Dichlorobenzene
Methylene chloride		1, 4-Dichlorobenzene
1,1,2,2-Tetrachloroethane		3, 3-Dichlorobenzidine

TABLE 5-2 (Continued)

EPA PRIORITY (TOXIC) POLLUTANT CONSTITUENTS BY PARAMETER GROUP

<u>Volatiles</u>	<u>Acid Extractables</u>	<u>Base Neutrals</u>
Tetrachloroethylene		Diethyl phthalate
Toluene		Dimethyl phthalate
1,2-Trans-dichloroethylene		Bi-n-butyl phthalate
1,1,1-Trichloroethane		2,4-Dinitrotoluene
1,1,2-Trichloroethane		2,6-Dinitrotoluene
Trichloroethylene		Di-n-octyl phthalate
Trichlorofluoromethane		1,2-Diphenylhydrazine
Vinyl chloride		Fluoranthene
		Fluorene
		Hexachlorobenzene
		Hexachlorobutadiene
		Hexachlorocyclopentadiene
		Hexachloroethane
		Indeno(1,2,3-c,d)pyrene
		Isophorone
		Naphthalene
		Nitrobenzene
		N-Nitrosodimethylamine
		N-Nitrosodi-n-propylamine
		N-Nitrosodiphenylamine
		Phenanthrene
		Pyrene
		1,2,4-Trichlorobenzene
		2,3,7,8-TCDD

TABLE 5-3
CONSTITUENTS OF PARAMETER GROUPS

<u>Parameter Group</u>	<u>Example Constituents</u>	<u>Potential Surrogates</u>	<u>Reason for Concern**</u>
Aldehydes	Formaldehyde, acetaldehydes, benzaldehydes	Formaldehyde, acetaldehydes	T, C, N I, N
Carboxylic acids	Formic acid	C ₁ -C ₃ carboxylic acids	I
C ₂ -C ₄ cyanides	Methyl cyanide, ethyl cyanide, isocyanides	C ₂ -C ₄ cyanides	T
Amines	Methyl amine, ethyl amine	Methyl and ethyl amines	T, I
Phenols	Catechol, alkyl phenols, cresols, naphthols	Total phenols, C ₁ -C ₃ alkyl phenols	T, C, N
Benzenes	Ethylbenzene, trimethylbenzene	C ₁ -C ₃ alkyl benzenes	T

**I = demonstrated toxicity; C - carcinogen or suspect carcinogen;
I = irritant to humans; N = subject of a NIOSH criteria document.

waste water treatment sludges); the second includes solid waste whose composition is the same as found in existing industries (i.e., boiler fly ash, FGD sludge, catalyst or solvents). Only the first type poses issues of EH&S concern at the PDU, since the major solid streams at this facility are bottom ash, cyclone fines, and separator pit sludge.

RCRA is the principal regulatory act covering the storage, treatment and disposal of solid hazardous waste. PDU wastes will be considered hazardous if they display the characteristics of toxicity, ignitability, reactivity or corrosivity, and if they are not excluded under RCRA exemption. The major EH&S concern presented by solid wastes is the transport of hazardous materials by leachate movement from waste storage to nearby surface or ground waters. Table 5-4, which presents the parameters of concern for solid wastes, contains only leachable or extractable constituents. This table includes leachable inorganics, some of which are identified in the RCRA regulations, as well as unregulated leachable or extractable organic compounds. Many of the parameters listed are presented in groups or are represented by surrogates.

Many of the chemical species mentioned above create certain health or safety-related issues in coal gasification plants. These issues are in addition to the environmental concerns raised by these chemicals because of their possible release into gaseous, aqueous or solid waste media. The issues principally involve the potential for worker or public exposure to the compounds produced in the plant. As an illustration of the various sources of concern, Table 5-5 summarizes the types of possible occupational exposures that have been identified in a high-Btu coal gasification plant. Exposures listed indicate the kinds of chemical parameters which will be of prime EH&S importance in any gasification plant.

Studies on the health effects of most of the compounds mentioned in the previous discussion have been performed and occupational exposure standards or permissible exposure limits have been established. For example, the following species likely to be present in coal gasification plants have been the subject of such studies or standard setting:

NH ₃	Coal tar products	Methanol
As	Coke-oven emissions	Ni
Benzene	Cresol	NO _x
Be	HCN	Phenol
Cd	Fluorides	Se
CO ₂	HCl	SiO ₂
CO	H ₂ S	SO ₂
CS ₂	Isopropylether	Toluene
Cr	Pb	V
Coal dust	Mn	Xylene
	Hg	

TABLE 5-4

PARAMETERS OF CONCERN: SOLID WASTES

Parameter Groups	Parameters	Regulatory Status		
		RCRA	CWA	Unregulated
<u>Leachable inorganics:</u>				
● RCRA Trace Elements	As, Ba, Cd, Cr, F ⁻ , Pb, Hg, NO ₃ -N, Se, Ag	X		
● Other Trace Elements	Sb, Be, Cu, Ni, Th, Zn		X	
● RCRA Groundwater Quality Parameters	Cl ⁻ , Fe, Mn, Na, SO ₄	X		
<u>Leachable organics:</u>				
● Priority (toxic) pollutants	See Table 5-2		X	
	- volatile			
	- acid extractables			
	- base neutrals			
● Others	See Table 5-3			X
	- aldehydes			
	- carboxylic acids			
	- amines			
	- phenols			
	- benzenes			
<u>Extractable organics:</u>				
● Priority (toxic) pollutants	See Table 5-2		X	
● Others	See Table 5-3			X
<u>Miscellaneous:</u>				
RCRA Contamination Parameters	pH, specific conductance X Total organic carbon Total organic halogens			

TABLE 5-5

POTENTIAL OCCUPATIONAL EXPOSURES IN COAL GASIFICATION

Unit Process	Potential Exposures
Coal handling and preparation	Coal dust, noise, fire
Coal feeding	Coal dust, noise, gaseous toxicants, and asphyxia
Gasifier operation	Coal dust, high-pressure hot raw gas, high-pressure oxygen, high-pressure steam, fire, noise
Ash removal	Heat stress, high-pressure steam, high-pressure oxygen under impact conditions, hot ash, and dust
Quenching	High-pressure hot raw gas, hot tar, hot tar oil, hot gas-liquor, fire, noise
Shift conversion	High-pressure hot raw gas, high-pressure hot shifted gas, high-pressure steam, tar, tar oil (naphtha), hydrogen cyanide, fire, catalyst dust, heat stress
Gas cooling	High-pressure hot raw gas, hot tar, hot tar oil, hot gas-liquor, fire, heat stress, noise
Gas purification	Sulfur-containing gases, methanol, naphtha, cryogenic temperatures, high-pressure steam, noise
Methanation	High-pressure rectisol product gas, high-pressure methanated gas, steam, nickel carbonyl, nickel catalyst dust, fire, noise
Sulfur removal	Hydrogen sulfide, other sulfides, and sulfur oxides
Gas-liquor separation	Tar oil, tar, gas-liquor with high concentrations of phenols, ammonia, hydrogen cyanide, hydrogen sulfide, carbon dioxide, trace elements, and noise
Phenol and ammonia recovery	Phenols, ammonia, acid gases, gas-liquor, ammonia recovery solvent, and fire
By-product storage	Tar, tar oil, phenols, ammonia, methanol, phenol recovery solvent, and fire

5.2 SUMMARY OF RESULTS

The principal emphasis has been to review the PDU data which were generated throughout the 1980-1982 campaigns, and to correlate the gas and solid composition data with feed composition, operating parameters, and possible pollutant standards levels. In an attempt to determine the fate of the elements and species in the feed materials during gasification, an equilibrium model was employed to project the stability of the gas, solid, and condensed phases which are most likely formed throughout the PDU at given set point operating conditions.

5.2.1 Mineral Species

A generalized reaction scheme was developed to illustrate the partitioning of the ash constituents in solids during gasification. The preliminary attempt does not reflect such factors as feed rank, initial feed composition or reaction temperature. Data generally indicate that Al_2O_3 , K_2O and Na_2O have a greater tendency to be elutriated in the ash material released from the gasifier freeboard, while SiO_2 and Fe_2O_3 tend to be retained within the withdrawal ash. Calcium oxide tends to be evenly distributed in the elutriated fines and withdrawal ash, while MgO slightly favors partitioning within the withdrawal ash fraction.

At equilibrium, the estimated solids composition formed at operating conditions consists principally of KAlSi_3O_8 , $\text{NaAlSi}_3\text{O}_8$, V_2O_5 , $\text{Fe}_2\text{O}_3 \cdot \text{Al}_2\text{O}_3$, NaAlSiO_4 (North Dakota lignite), CaSiO_3 (North Dakota lignite), SiO_2 (RSA, Western Kentucky, and Brookville coal), $\text{CaO} \cdot \text{MgO} \cdot 2\text{SiO}_2$ (Upper Freeport coal and North Dakota lignite), MgAl_2O_4 (RSA, Western Kentucky, and Brookville coal), MgSiO_3 (RSA and Upper Freeport coal), FeS (Upper Freeport, Western Kentucky and Brookville coal), and FeO (North Dakota lignite). The nature of the incoming feed composition, operating parameters and, indirectly, recycled gas and oxidant gas composition influences the resulting gasifier ash composition.

5.2.2 Sulfur Compounds

In reviewing the reported sulfur analyses, between 33 and 94 percent of the total background sulfur in the PDU feed materials is calculated to be released during gasification. The principal reported released sulfur species is H_2S , with secondary contributions of COS . Sulfur analyses are used to rank four coal feed materials as a function of decreasing total sulfur release in the following manner:

Pittsburgh seam > Indiana No. 7 > West Virginia > Wyoming subbituminous.

Sulfur partitioning data indicate that the higher pyritic sulfur containing coals release a substantially greater percentage of their background sulfur concentration, and that conversion into the sulfate constituents formed in the resulting char agglomerates is primarily dependent on the quantity of

iron available for the reaction with steam, and possibly oxygen, and/or recycle gas. Organic sulfur is observed to be completely released and not recaptured by either the inorganic or organic constituents in the presence of steam, and the recycle gas environment. Total sulfur release from these four coals increases as the gasifier operating temperature increases.

A completely different sulfur release mechanism is suggested during gasification of RSA coal. Loss of sulfur is indicated from the pyritic fraction, with subsequent conversion and retention of sulfur within the sulfatic and organic fractions. Further effort is required to establish the criteria for variations in the sulfur release based on the initial feed composition and rank.

Sulfur is projected to be totally released from pyrite or iron disulfide during the gasification of RSA coal and North Dakota lignite, while partial retention of sulfur is projected for the Upper Freeport, Western Kentucky, and Brookville coal systems. The concentration of H_2S which is projected to be released during gasification at equilibrium conditions exceeds the reported H_2S concentrations. One possible explanation for this discrepancy is that the calculated H_2S equilibrium concentration is determined at the gasifier temperature which is higher than the reported raw product gas temperature. Alternately, the system may not have attained equilibrium for the release of H_2S from the gasifying coal particles, or sulfidation has resulted along the structural metal surfaces in the reducing environment.

5.2.3 Assessment of Raw Gas, Liquor and Sludge Analyses

The objective of the 1983 EH&S characterization program at the PDU was to provide screening information on major process streams for parameters of EH&S concern and to identify major issues which could be addressed in the 1983-86 program. A representative summary of the EH&S characterization results for one coal feedstock and one lignite feedstock is reported in Tables 5-6 through 5-8 for the raw gas, separator pit liquor and sludge. Results are presented as observed ranges for the parameters of concern, and they take into account all data from the two tests together with trace organic information. These tests were conducted during the gasification of different coals under a variety of process conditions.

The large ranges reported for many parameters are most likely related to variations in gasifier conditions (e.g., coal feed and temperature), as well as variations in quench water recycle operation and detailed sampling procedures. Data are summarized as ranges to allow a general characterization of the PDU gasifier and its variability rather than a specific focus on individual numbers. Because of limitations experienced in process control, sampling, and analytical procedures, it is believed premature to relate the observed results to individual test points or process conditions. The major causes of the observed variations in parameter concentrations have not yet been firmly established.

TABLE 5-6

TEST PLAN TP-033-1 CHARACTERIZATION RESULTS
SET POINTS 1, 1A and 2* (BROOKVILLE COAL)

Parameter	Miniscrubber (lb/10 ⁶ lb, dry gas)	Separator Pit (mg/l)
Number of samples:	3	2
pH	---	7.1 - 7.6
Alkalinity	< 25	---**
BOD	< 10 - 30	2 - 19
COD	< 20 - 300	---
TOC	1970 - 2420	170 - 174
CN ⁻	< 60	2.18 - 6
SCN ⁻	< 20	8 - 51
NH ₃ -N	100 - 120	---
TKN	3 - 25	---
Organic-Nitrogen	4	---
Sulfide (S ⁼)	300 - 700	< 0.03
Sulfate	20 - 400	---
TDS	< 25	---
Antimony	< 3.2	< 0.2 - 0.4
Beryllium	< 0.12	< 0.005 - 0.016
Boron	25	1.6 - 2.4
Cadmium	< 0.12	< 0.005 - 0.2
Calcium	2 - 4	---
Chromium	< 0.32	< 0.002 - 0.68
Copper	< 0.25	< 0.002 - 0.44
Iron, total	10 - 60	---
Lead	< 1.25	< 0.05 - 0.39
Magnesium	1 - 3	---
Mercury	< 0.02	< 0.0008
Nickel	< 0.48	< 0.002 - 0.49
Potassium	< 0.2 - 0.7	---
Sodium	2 - 60	---
Zinc	< 0.5 - 1.0	< 0.002 - 0.46

* Trace organics are reported in Table 5-8. Trace elements were not reported when below detection limit in all samples.

** Parameter not analyzed.

TABLE 5-7

TEST PLAN TP-033-1 CHARACTERIZATION RESULTS
 SET POINT 3*
 (NORTH DAKOTA LIGNITE)

<u>Parameter</u>	<u>Miniscrubber (lb/10⁶ lb, dry gas)</u>	<u>Separator Pit (mg/l)</u>
Number of samples:	1	1
pH	---	8.6
Alkalinity	< 20	---**
BOD	< 20	6
COD	100	6
TOC	1100	380
CN ⁻	40	< 1
SCN ⁻	< 10	60
NH ₃ -N	185	---
TKN	< 22.3	---
Organic-Nitrogen	18.5	---
Sulfide (S ⁼)	1100	< 0.03
Sulfite	< 40	---
Sulfate	260	---
TDS	< 20	---
Arsenic	< 0.16	< 0.05
Calcium	1	---
Iron, total	0.5	---
Magnesium	0.5	---
Selenium	0.5	---
Sodium	5	---

* Trace organics are reported in Table 5-8. Trace elements were not reported when below detection limit in all samples.

** Parameter not analyzed.

TABLE 5-8
TRACE ORGANICS RESULTS, TEST PLAN TP-033-1

Parameter ⁺	Brookville Coal*		North Dakota Lignite**		Post Test
	Separator		Separator		Sample
	Miniscrubber	Pit	Miniscrubber	Pit	Separator Pit
	(lb/10 ⁶ lb)	(mg/l)	(lb/10 ⁶ lb)	(mg/l)	(mg/l)
<u>Priority Pollutants</u>					
Benzene	<0.48 - <0.59	0.08 - 0.68	630	0.32	0.24
Acenaphthene	<0.03 - <0.043	0.02 - 0.04	<0.034	0.03	0.15
Acenaphthylene	<0.03 - <0.043	0.02 - 0.04	<0.034	0.02	0.09
Fluoranthene	<0.03 - <0.043	<0.01 - 0.01	<0.01	<0.01	<0.01
Fluorene	<0.03 - <0.043	<0.01	0.02	0.02	0.14
Naphthalene	<0.03 - <0.043	0.3 - 0.6	0.5	0.5	2.68
Phenanthrene	<0.03 - <0.043	0.02 - 0.03	0.04	0.04	0.11
Pyrene	<0.03 - <0.043	0.06 - 0.13	0.02	0.02	<0.01
Anthracene	<0.03 - <0.043	<0.01	<0.034	<0.01	0.01
<u>Tentatively Identified Compounds</u>					
Methanethiol		Present			
Thiophene		Present			Present
Dimethyl disulfide		Present			
Chlorinated propane		Present			
Benzenes (ethyl, methyl)		Present		Present	Present
Acetic acid esters		Present			
Tetrahydro thiopyranones		Present		Present	
Trithiolanes		Present			
Cyclopropanes		Present			
Hexathiopane				Present	Present
Pyridenes				Present	Present
Cyclohexane				Present	
Alkyl alcohol				Present	
Tetrahydronaphthalene				Present	
Aliphatic alcohol				Present	
Alkylated PAH				Present	
1,1-Biphenyl					Present
Biphenylene					Present
Dibenzofuran					Present
Phenylene					Present
Indene					Present
Methyl pentanone					Present
Trichloro, trifluoro-ethane					Present

* Includes samples from set points 1, 1A and 2.

** Includes samples from set point 3.

+ Not reported if below the detection limit.

The major purpose of the raw gas characterization reported here was to evaluate trace constituents of EH&S concern. Raw gas was observed to contain many parameters of EH&S concern at levels within the range expected for fluidized bed gasification processes. For instance, alkalinity, biological oxygen demand (BOD), chemical oxygen demand (COD), phenolics, and benzene concentrations were at levels similar to those found in raw gas from another fluidized-bed process. Sulfur and nitrogen species were at somewhat lower levels. Total organic carbon (TOC) levels were variable and occasionally very high. The specific components of the TOC load have not yet been quantified. Data indicate that raw product gas from the PDU is generally characterized by no observable tars and oils and low levels of toxic trace organic constituents, such as phenolics, benzene and polycyclic aromatic hydrocarbons (PAH). Trace element levels are also low and most likely a function of the particular coal being gasified.

Quenching of the raw gas produces a waste water which contains appreciable quantities of alkalinity, ammonia, sulfides, cyanides, thiocyanates, and TOC. Separator pit liquor approximates the content of a quenched waste water, although levels of volatile and highly-sorbable constituents are probably reduced because of reaction in the pit. If present at similar levels in large-scale commercial facilities, all these constituents will need to be substantially removed in waste water treatment facilities before process condensates can be discharged to receiving waters or before they can be recycled as makeup water for in-plant reuse.

Separator pit liquor was the principal liquid stream characterized in the 1982 program, with occasional supplemental sampling of quench recycle water and raw water. Parameters of concern for design of waste water treatment systems that were detected include TOC, ammonia, sulfide, thiocyanate, cyanide, trace elements and trace organic compounds. Small amounts of phenol, benzene and other polycyclic aromatic hydrocarbon (PAH) compounds were present in the separator pit liquor. The range of TOC values could not be satisfactorily accounted for by the organic parameters quantified by this study. It is likely that some of the TOC comprises other compounds which were tentatively identified in the separator pit liquor such as alkyl benzenes, carboxylic acids, and alcohols. Cyanides and thiocyanates were found at quite variable levels in the separator pit. If these levels are typical of large-scale plant quench liquor, they can be reduced to environmentally acceptable levels by either biological oxidation or physical/chemical treatment.

A number of trace elements were detected during the 1982 program, e.g., arsenic and iron. Levels of these constituents are expected to be controlled by levels in the coal feedstocks used. Based on the preliminary information obtained, it appears that conventional precipitation/clarification unit operations can effectively reduce these elements to levels acceptable for discharge to receiving waters.

5.3 RECOMMENDATIONS

Principal issues which remain to be resolved include:

- Identification of the inorganic and organic priority pollutants, and their respective concentrations as gas, liquid, condensed, or solid phase pollutants.
- Identification of the variance in the solids flow rates throughout the PDU, and in the chemical compositions of the various solid products.
- Identification of trace and minor species or elemental concentrations in the initial feed, product solids or gases, and in the miniscrubber and quench scrubber solutions.
- Definition of the solubility of the various gas and solid species in the scrubber solutions through bench-scale leaching studies, or computer developed models such as the ECES/FRACHEM programs by OLI Systems.
- Evaluation of and/or revision to the chemical analytical techniques; establishment of lower analytical detection limits for species defined as priority pollutants.
- Refinement of the sampling procedures (i.e., elimination of compositing; initiation of multiple sampling and chemical analysis).
- Refinement of the equilibrium model to define the influence of recycled fines on bed and effluent gas composition.
- Updating of the organic data base; updating of trace species data base for priority pollutants; expansion of the equilibrium calculation capacity to incorporate inorganic and organic species into a single system.
- Identification of the gasification partitioning mechanisms of the various species within the PDU solids.
- Evaluation of the trace and minor species volatilization reactions in the gasifier, surface chemical reactions and/or absorption/condensation reactions of volatiles in the effluent stream with entrained fines.

SECTION 6.0

MATERIALS EVALUATION

Early operation of the Process Development Unit (PDU) was mainly concerned with proving that the process was viable. Strict attention was not paid to materials behavior. Thus, characterization of materials behavior in the KRW PDU is somewhat difficult due to the lack of comprehensive records on exposure histories. In addition, the varied operating conditions experienced in the PDU makes evaluation of materials behavior difficult in random locations without the instrumentation required to adequately describe the specific operational environments.

Meaningful materials exposure can be performed in the PDU environment if experiments are adequately designed so that a detailed exposure history can be obtained for each test. Such a program would result in the following issues being addressed:

- A. Evaluation of materials which are economically attractive for commercial applications.
- B. Evaluation of coatings for corrosion/erosion prevention.
- C. Assessment of PDU failure analyses and materials coupons test data.

6.1 BACKGROUND

Materials testing at the PDU has been conducted in several areas. For example, the oxidant tube has had various types of plasma spray coatings applied to evaluate the ability of coatings to protect the base metal. To date, more than 2000 hours of operation has been accumulated in this ongoing evaluation. Additionally, sections of the PDU feed line, originally constructed of Inconel 800 and later of Incoloy 825, have been replaced with Type 816L stainless steel; sample plugs have been removed from the containment shells of the gasifier vessel and primary cyclone for evaluation; and, sample coupon racks have been installed at various points throughout the gasification process and periodically removed for evaluation.

6.2 COAL FEED SYSTEM EXPERIENCE

6.2.1 Coal Lockhoppers

The PDU coal lockhoppers are constructed of plain carbon steel approximately 1/2-inch thick. A core sample from a lockhopper cone shows that the material has eroded approximately 1/16 inch over approximately one year (8400 hours) of exposure to coal fines. The present design is considered adequate for commercial application with the addition of a hardened, smooth liner for the cone area of the vessel.

6.2.2 Coal Feed Lines

Coal feed lines on the PDU were originally built from Inconel 800 pipe. Problems with weld cracking were experienced during early PDU runs which were attributed to the use of improper material. The Inconel 800 pipe was supplied in the sensitized condition leading to weld cracking. Feed lines were reconstructed using Incoloy 825 and no further problems with weld failures were encountered.

Erosion problems with the PDU feedlines have been a continuing problem. Elbows making 45- or 90-degree directional changes were eroded in 4 to 24 hours of operation. A design change of the 90-degree elbows incorporating a "dead-leg" type of construction with a sacrificial metallic or ceramic insert proved to be successful for reliable operation. The original PDU design required feed lines to meet an 1100°F operating temperature. Subsequent PDU operation has shown that the feed line temperatures never exceed 400°F and most of the system operates at temperatures of approximately 200°F. Thus, there is no need for feed lines to be fabricated from expensive materials such as Incoloy 825. A section of the PDU feed line, fabricated from 316L, has been in operation in the PDU for more than 2000 hours without signs of degradation. In all probability, carbon steel would operate satisfactorily at temperatures below 400°F. The ultimate feed line design for the KRW process would probably be a carbon steel pipe with a refractory liner to achieve optimum service life.

6.2.3 Oxidant and Coal Feed Tube

PDU experience with the oxidant tube has shown two areas for concern. One is the outer surface of the tube which is exposed to the internal gasifier environment. An outer tube, fabricated from Incoloy 825, showed metal loss of 0.060 inch after 652 hours of operation. This extrapolates into a corrosion rate of 0.75 inch/year. A new outer tube was fabricated of Incoloy 825 and coated with a plasma spray coating of MgO.ZrO₂. This coating was exposed to approximately 1400 hours of operation. At the end of that time the tube showed pockets of corrosion where the coating had spalled off. The exact reason for the coating spall is not known. It is possible that the plasma sprayed base coat was attacked through coating cracks. This base coat was a high nickel alloy coat recommended to be used in conjunction with the ceramic coat. High nickel content materials are particularly susceptible to attack in coal gasification atmospheres containing sulfur.

To date, the approach of using a protective coating seems to be the most promising avenue to pursue for developing a reasonable lifetime for the oxidant tube. In a DOE-sponsored program at Argonne National Laboratory, some promising results on plasma sprayed coatings have been noted. Type 310 SS coated with a high cobalt base coat and overlayed with a CaO stabilized ZrO₂ has shown good thermal shock and corrosion resistance in a simulated coal gas environment. Also, aluminized materials show some

promise of success in this application. Testing of these options in the PDU is recommended to determine whether or not they would be acceptable in a commercial application.

The other area of concern in the oxidant tube assembly is the central feed tube. This tube is subject to erosion similar to that of the coal feed lines. In 1982 a PDU coal feed tube failed catastrophically after approximately 2000 hours of operation. Post test inspection revealed that the coal feed tube was eroded to the point where much of the original tube was missing and only a thin shell of approximately one half of the original tube remained. Therefore, it seems unlikely that a metallic feed tube would provide long service life. A ceramic insert of fused Al_2O_3 or a similar hard refractory material is a possible solution for commercially viable feed tube design. Testing of this concept in the PDU via a removable ceramic tube insert should be carried out.

6.3 GASIFIER EXPERIENCE

6.3.1 Shell Material

Experience with the PDU gasifier has revealed no major problems with the design in terms of the vessel shell material and the refractory lining. The type of insulating and hardface refractory used in the gasifier has withstood up to 8000 hours of operation with no indication of failure. HW-26 castable refractory was used for the insulating layer and Castalast-G from Harbison-Walker was used for the hardfacing. In addition, an intermediate alumina refractory (Quigley 30-50) has been exposed in the piping outlet of the cyclone for more than 1100 hours with no sign of spalling or cracking.

The shell of the gasifier is plain carbon steel, ASTM A-515 Gr 70, and has performed adequately. The gasifier operates in such a manner as to keep the vessel shell below the dew point of the product gas. It was speculated that this operating mode could cause acidic condensation at the metal/refractory interface leading to excessive shell corrosion. A core sample removed from the gasifier shell after more than 8000 hours of operation showed that no degradation of the vessel wall could be detected.

During the span of time that the PDU has been operating, modifications have been made to the gasifier which required refractory replacement in sections of the vessel. There was some concern that operating below the dew point would result in refractory degradation in the insulating layer at or near the i.d. of the vessel containment shell. In all cases of refractory removal, no areas of degradation were noted either in the refractory or on the i.d. of the vessel wall. Thus, the refractory system used in the KRW system appears to offer adequate protection for the vessel wall.

6.3.2 Thermowells

Various thermowell materials have been tested in the PDU since inception. Hastaloy-X and C-276 were found to corrode completely after 300 to 400 hours of operation. Type 310 SS has shown a varied lifetime ranging from 450 to 2500 hours, depending upon operating conditions. This will need further investigation.

6.4 PRODUCT GAS PIPING EXPERIENCE

The PDU experience with product gas flow through refractory lined pipe is the same as that noted for the gasifier. No problems have been noted using the two-layer refractory system.

6.5 CYCLONE EXPERIENCE

The cyclone on the PDU is constructed the same way as the gasifier and hot product gas piping. One such cyclone, after 6000 hours of operating time, showed no evidence of refractory degradation or erosion.

6.6 QUENCH SYSTEM EXPERIENCE

The gas cooling system on the PDU consists of a quench scrubber (C-111) and two sequential gas coolers (C-113 and C-122). All vessels are fabricated from ASTM A-515 Gr 70 carbon steel. Product gas enters the quench scrubber at 1100°F to 1700°F, depending upon the mode of operation, and is cooled to approximately 350°F. The bulk of the coal fines are removed from the gas stream and the product gas continues through the remaining gas coolers where more particulates are removed from the gas stream. The gas stream is split upstream from the last gas cooler (C-122), part of the stream going to the thermal oxidizer to be consumed and the remaining gas going into C-122 to be cooled and piped to the recycle gas compressor where it is brought back to system pressure for use in the process.

Material concerns in the PDU quench system are concentrated in the vessels, the piping between the vessels, and in the dump lines where water mixed with particulates are let down to atmospheric pressure. In the case of the vessels, ultrasonic wall thickness measurements indicate that up to 0.070 inch of wall thickness has been removed in selected areas after approximately 8000 hours of operation. A sample of vessel wall removed from the quench scrubber was examined metallographically and found to have lost approximately 0.060 inch of material from the original thickness.

Ultrasonic testing of the carbon steel piping between vessels show metal losses of up to 0.100 inch. The exact exposure time of this material is not known exactly due to occasional replacement of piping in the PDU. In the most optimistic case, if the piping under discussion is the original PDU line, then the corrosion rate is approximately 0.100 inch/year under PDU operating conditions.

Scrubber and cooler dump lines are frequently changed due to corrosion or corrosion/erosion degradation. The carbon steel lines are exposed to high-velocity, particle-laden water dumped into the separation system. In a commercial plant this water would probably be recirculated and cleaned at pressure, thus eliminating the excessive erosive wear experienced in the PDU.

6.7 PUMPS

Two types of pumps are used on the PDU. In the PDU makeup pit, vertical 6-stage centrifugal pumps are used to provide PDU makeup water. These pump particle-laden water and, as a result of their design and service requirements, frequency of repair is high. The most prominent problems are corrosion and erosion of the main shaft and impeller bearings.

Other pumps in the PDU are single-stage centrifugal types which provide make-up water for the quench vessels and pump water from the separator pit to the thermal oxidizer. Early operation of these pumps resulted in short lifetimes due to high corrosion rates. The pump casings and impellers were cast carbon steel. Various castings and treatments were evaluated to protect the pump internals, the most successful being chrome plating. The present single-stage pumps in the PDU are fabricated from CD4MCu, a high chromium cast material, and have chromium plating on the impeller and casing. This has proved to be satisfactory in cases of normal operation at the pump speeds used in the PDU. In rare cases where process upsets have resulted in high particle loadings through make-up pumps, some erosion of the CD4MCu impellers has resulted.

6.8 RECOMMENDATIONS

In general, materials testing in PDU type operating conditions tend to result in overpredictions of corrosive rates. This is true because most experiments are of the tag-along type and are subject to frequent process upsets, frequent shutdowns and long idle periods exposed to air. Even under these stringent conditions, these types of tests still provide useful trends when comparing one material against another as candidate materials for construction of large-scale applications.

Materials experiments can be carried out in a manner to minimize the effect of the off-normal operations by designing tests to operate using a slip stream where a partial flow of the desired environment is diverted into the testing area. This diversion will occur only when the conditions are normal and is shut off and back purged with a neutral environment at off-test and shutdown conditions. Data from this type of testing would approximate large-scale operations more closely.

The main need for materials testing is for future plants. Near-term large-scale plants will be fabricated from materials that have limited exposures in coal gas atmospheres and are of a reasonable cost. Data will

be generated in these initial large-scale plants on the lifetime of the materials in terms of corrosion and erosion. The value of PDU testing is not to accurately predict the lifetimes of materials for the initial large-scale plants, but to compare the behavior of the initial materials with alternate materials so that if the initial selections are not adequate, one may have a rationale for a selection of more suitable materials for second generation large-scale plants.

APPENDIX A PDU PROCESS DATA

Process data from the KRW PDU is generated in several categories. This data consists of both continuous and periodically recorded information during a test run. Solid, liquid, and gas samples are also periodically taken throughout the test run. At the conclusion of a test, the declared set points are analyzed to determine steady state portions. Samples taken during steady state periods are separated for analyses. These analyses are then averaged for input into a heat and material balance. An exception is that all liquid samples are individually analyzed and recorded. Most of these various types of data have been collected for each test run on the PDU and they form the data base which is continuously being expanded and refined for the Westinghouse pressurized fluidized-bed gasifier PDU.

These various types of data can be categorized into four general types:

- Type 1 data are directly measured parameters whose instantaneous readings are recorded by a computer every 10 seconds. This data can be played back immediately for analysis and includes gas flow rates, solid feed or withdrawal rates, temperatures, pressures, pressure differentials, product and recycle gas composition, and results of calculations from these data.
- Type 2 data are those which are measured and recorded on a hand generated time plot. Included in this type are: ash withdrawal rate; ash withdrawal ash concentration and bulk density; bed ash concentration and bulk density; primary and secondary fines ash concentration and bulk density; and, coal feed ash concentration and bulk density.
- Type 3 data are detailed chemical analyses of various solid, liquid and gas samples. Included are ultimate and proximate analyses of feedstock, bed char, ash withdrawal, primary fines, secondary fines, and lost fines. Additional inclusions are: chemical analyses of the ash component of both feedstock and ash withdrawal; product gas analyses from various miniscrubber samples; and, analyses of various liquid samples taken from the quench scrubber to detect the existence of any organic constituents. The analyses for samples obtained during non-steady state conditions are performed only upon special request.
- Type 4 data are continuously recorded on strip charts and they duplicate the data periodically recorded by the computer as Type 1 data.

The process used to derive heat and material balances from raw data is shown in Figure A-1. A chronological history of the various feedstock for which data has been accumulated is shown in Table A-1. A summary method for determining how well a period during a set point met steady-state criteria is shown in Figure A-2.

Of the four types of data listed in Table A-2, types 1, 2, and 3 from the steady-state periods are compiled and/or averaged into a Process Summary and a detailed heat and material balance which can be incorporated into a Process Flow Diagram, as shown by the examples in Figures A-3 and A-4 for a Pennsylvania coal and a North Dakota lignite.

Table A-3 lists those specific points in the system from which Type 1 data are periodically recorded by the computer. Table A-4 lists Type 2 data collected by units and frequency. Type 3 data collection along with the samples on which the various analyses are performed are shown in Table A-5 and the analyses forms are shown in Figures A-5 through A-10.

Figure A-11 shows the Process Summary Form, Figure A-12 shows the Carbon Utilization form, Figure A-13 shows the Gasifier Heat and Material Balance Flow Schematic and Figure A-14 shows the Heat and Material Balance format.

Figure A-1. Derivation of PDU Test Heat and Materials Balances

TABLE A-1
CHRONOLOGICAL HISTORY OF PDU COAL FEEDSTOCKS

<u>Coal Feedstock</u>	<u>Test No.</u>	<u>Month/Year</u>	<u>Hot Hours of Coal</u>	<u>Mode</u>
Pittsburgh No. 8 Adamsburg Mine (R-O-M Washed)	TP-016	4-5/78	113	Air
Pittsburgh No. 8 Champion Mine (R-O-M Washed)	TP-013-3	12/77	16	Air
	TP-017	5/78	161	Air
	TP-018-2	7/78	45-1/2	O ₂
	TP-018-3	8/78	69-1/2	O ₂
	TP-018-4	9/78	47-1/2	O ₂
	TP-018-5	9-10/78	85	O ₂
	TP-020-2	5/79	164-1/2	Air
	TP-022-1	6/79	53	O ₂
	TP-022-2	6-7/79	57	O ₂
	TP-023-3	12/79	50-1/2	O ₂
	TP-023-4	3/80	134-3/4	O ₂
	TP-027-2	5/80	64-1/2	O ₂
	TP-027-3	6-7/80	147-1/2	O ₂
	TP-028-1	6/81	33-1/2	O ₂
Pittsburgh No. 8 Federal Mine	TP-034-2	6-7/83	435	O ₂
	TP-035-2	12/83	226-1/2	O ₂
Ohio No. 9 Bannock Mine	TP-023-1	7-8/79	30	O ₂
	TP-023-2	8/79	26	O ₂
	TP-025-2	11/79	14-1/2	Air
	TP-027-1*	4/80	69	O ₂
	TP-027-2	5/80	62-1/2	O ₂
W. Kentucky No. 9 Hamilton Mine (R-O-M Washed)	TP-019-2	11/78	45	O ₂
	TP-019-3	12/78	56	O ₂
	TP-027-4	8/80	66-1/2	O ₂
	TP-032-2	5/82	95.5	O ₂
	TP-033-3	12/82	94-1/3	Air & O ₂
Indiana No. 7 Minnehaha Mine (R-O-M Washed)	TP-013-3	12/77	27	Air
	TP-019-1	10/78	37-1/4	O ₂
	TP-019-2	11/78	95-1/2	O ₂
	TP-030-2	5/81	61-1/2	O ₂
	TP-028-2	7/81	144	O ₂ & Mixed
Indiana No. 7 Sycamore Mine (R-O-M Washed)	TP-025-1	9/79	146	Air
	TP-025-2	11/79	61-1/4	Air
	TP-027-1*	4/80	69	O ₂

* Indiana No. 7 and Ohio No. 9 blended during drying operation.

TABLE A-1 (Continued)

CHRONOLOGICAL HISTORY OF PDU COAL FEEDSTOCKS

<u>Coal Feedstock</u>	<u>Test No.</u>	<u>Month/Year</u>	<u>Hot Hours of Coal</u>	<u>Mode</u>
Rosebud Coal Montana (R-O-M Unwashed) Subbituminous "B"	TP-019-1	10/78	63-1/2	O ₂
Texas Lignite (R-O-M Unwashed)	TP-023-4	3/80	11-1/2	O ₂
Texas Lignite Monticello Mine (R-O-M Unwashed)	TP-027-3 TP-027-5	6-7/80 10/80	47 85	O ₂ O ₂
Wyoming (Wyodak Mine) (R-O-M Unwashed)	TP-013-3	12/77	22	Air
	TP-015	4/78	36	Air
	TP-027-4	8/80	50-1/2	O ₂
	TP-027-5	10/80	24	O ₂
	TP-029-1/ -030-1	4/81	60	O ₂ & Mixed
	TP-028-1	6/81	39	O ₂
	TP-033-2	10-11/82	75-1/2	O ₂
	TP-034-1	3-4/83	148-1/4	O ₂
	TP-035-1	9-10/83	291-1/4	O ₂
Illinois No. 6 Delta Mine (R-O-M Unwashed)	TP-031-1	12/80	12	O ₂
	TP-029-1/ -030-1	4/81	112-1/2	O ₂
Upper Freeport Seam (R-O-M Unwashed)	TP-023-3	12/79	22	O ₂
Upper Freeport E-Seam, Elton Mine (Unwashed)	TP-032-1	3-5/82	97.3	O ₂
	TP-032-2	5/82	53.5	O ₂
RSA Coal Bosjesspruit Colliery (R-O-M Unwashed)	TP-031-1	12/80	97	O ₂
	TP-030-2	5/81	53	O ₂
	TP-031-2	10/81	46	O ₂
Texas Lignite Big Brown Mine (R-O-M Unwashed)	TP-028-3	9/81	112-1/2	O ₂

TABLE A-1 (Continued)
CHRONOLOGICAL HISTORY OF PDU COAL FEEDSTOCKS

<u>Coal Feedstock</u>	<u>Test No.</u>	<u>Month/Year</u>	<u>Hot Hours of Coal</u>	<u>Mode</u>
West Virginia Big Mountain Mine (R-O-M Washed)	TP-030-3	11/81	172-1/2	O ₂
Arco Coke Watson Refinery	TP-031-3	12/81	88-1/4	O ₂
Chevron Coke El Segundo, Calif.	TP-031-3	12/81	61-1/4	O ₂
N. Dakota Lignite Indianhead Mine	TP-033-1 TP-034-3	7-8/82 8/83	67-1/2 216-2/3	O ₂ O ₂
(R-O-M Unwashed)				
Brookville Ace Drilling Co.	TP-033-1	7-8/82	103-1/4	O ₂

STEADY STATE CRITERIA PARAMETERS
FOR TP- _____ OXYGEN BLOWN
_____ AIR BLOWN

PAGE NO. _____

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SET POINT													
STEADY STATE													
DATE AND TIME													
PARAMETERS FOR CRITERIA #1	BAND	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time
Summation of Coal Feeder RPM	+15%												
Steam Feed Rate	+ 5%												
Oxygen Feed Rate	+ 5%												
Total Recycle Gas	+ 5%												
Freeboard Temp.	+30°F												
Operating Pressure	+3PSI												
Bed Elevation	+12in												
Bed Ash (SC-22)	+ 5%												
PARAMETERS FOR CRITERIA #2	BAND	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time	Exceeded By	Time
Product Gas HHV	+10 Btu/scf												
Fines Collect/Recycle Rate	+10%												
Ash Conc. (C117)	+10%												
Product Gas	+10%												

Figure A-2 Steady State Criteria Parameters

TABLE A-2

DATA TYPES

<u>Type</u>	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>
Description	Directly measured; recorded every 10 seconds	Recorded on a hand- generated time plot	Detailed chemical analysis of solid, liquid and gas samples	Directly measured; continuously recorded on strip charts
Availability	Can be played back immediately for analysis		Only for steady- state periods	
Information Sampled	See Table A-3	See Table A-4	See Table A-5	See Table A-3

TABLE A-3
SUMMARY OF TYPE 1 DATA

<u>Type Measurement</u>	<u>No. Sampled and/or Recorded</u>
Flows:	
Gas flows	30
Liquid flows	7
Pressure differentials:	
Solids transport	8
Gasifier axial	9
Fines collection (cyclone)	8
Fines transport	7
Lockhoppers	7
Miscellaneous	2
Pressure:	
C-115 gasifier	3
Lockhoppers	12
Utilities	9
Recycle gas	2
Water system	5
Cyclones	4
Isokinetics	2
Weigh cells:	
Feedstock	6
Char	3
Starwheel feeders:	
Feedstock	2
Char	2
Ash withdrawal	1
Level transmitters:	
Water system	6
Gasifier	1
Cyclone	1
Gas chromatograph measurements:	
Product and recycle gas	24
Temperature:	
Thermocouples	146
Calculated parameters:	
Velocities, product and recycle gas molecular weights, solid flow rates, bed densities, constants, etc.	262

TABLE A-4
TYPE 2 DATA

	<u>Units</u>	<u>Frequency</u>
Ash withdrawal weight	lb	hourly
Withdrawal ash concentration	wt%	hourly
Withdrawal bulk density	lb/ft ³	hourly
Bed char ash concentration	wt%	hourly
Bed char bulk design	lb/ft ³	hourly
Coal feed ash concentration	wt%	1 every 4 hours
Coal feed bulk density	lb/ft ³	1 every 4 hours
Primary fines ash concentration	wt%	1 every 4 hours
Primary fines bulk density	lb/ft ³	1 every 4 hours
Secondary fines ash concentration	wt%	1 every 4 hours
Secondary fines bulk density	lb/ft ³	1 every 4 hours

TABLE A-5
TYPE 3 DATA

I. <u>Proximate Analysis - weight percent</u>	<u>Solid sample on which performed</u>
a) Fixed carbon	a) Feedstock
b) Volatile matter	b) Bed char
c) Moisture	c) Ash withdrawal
d) Ash	d) Primary fines
	e) Secondary fines
	f) Lost fines
II. <u>Ultimate Analysis - weight percent</u>	<u>Solid sample on which performed</u>
a) Carbon	a) Feedstock
b) Hydrogen	b) Bed char
c) Oxygen	c) Ash withdrawal
d) Nitrogen	d) Primary fines
e) Sulfur	e) Secondary fines
f) Ash	f) Lost fines
III. <u>Elemental Analysis of Ash - weight percent</u>	<u>Solid sample on which performed</u>
a) SiO ₂	a) Feedstock
b) Al ₂ O ₃	b) Ash withdrawal
c) CaO	
d) MgO	
e) Na ₂ O	
f) K ₂ O	
g) Fe ₂ O ₃	
h) SO ₃	
IV. <u>Sieve analysis and bulk density</u>	<u>Solid samples on which performed</u>
	a) Feedstock
	b) Bed char
	c) Ash withdrawal
	d) Primary fines
	e) Secondary fines
	f) Lost fines
V. <u>Mean particle diameter (μ)</u>	<u>Solid sample on which performed</u>
	a) Primary fines
	b) Secondary fines
	c) Lost fines

TABLE A-5 (Continued)

TYPE 3 DATA

VI. <u>Miniscrubber - Distilled Water</u> <u>(PPM - dry basis)</u>	<u>Gas stream on which performed</u>
a) BOD b) Alkalinity as CaCO_3 c) Arsenic d) Cadmium e) Chloride f) Chromium g) Copper h) Lead i) Mercury j) Nickel k) Selenium l) Silver m) Dissolved solids n) Zinc o) Thallium p) Boron q) Beryllium r) Antimony s) pH	a) Product gas before quench
VII. <u>Miniscrubber Acid Scrub</u> <u>(ppm - dry basis)</u>	<u>Gas stream on which performed</u>
a) COD b) Calcium c) Iron d) Magnesium e) Ammonia as N f) Potassium g) Sodium h) Organic Nitrogen i) T.O.C. j) Thiocyanate	a) Product gas before quench
VIII. <u>Miniscrubber Base Scrub</u> <u>(ppm - dry basis)</u>	<u>Gas stream on which performed</u>
a) Total Cyanide b) Sulfate c) Sulfite d) Sulfide e) Thiosulfate as $\text{Na}_2 \text{S}_2 \text{O}_3$	a) Product gas before quench

TABLE A-5 (Continued)

TYPE 3 DATA

IX. <u>Quench Scrubber Water Analysis</u>	<u>Liquid stream on which performed</u>
a) Solids wt%	a) C-111 blowdown
b) pH	b) C-113 blowdown
c) Alkalinity, Total to pH4 mg/l	c) C-122 blowdown
d) BOD 5 day mg/l	d) Pit
e) Total Organic Carbon mg/l	
f) Chloride mg/l	
g) Total Cyanide mg/l	
h) Fluoride mg/l	
i) Ammonia Nitrogen mg/l	
j) Total Kjeldahl Nitrogen mg/l	
k) Permanganate Value, 4 hours mg/l	
l) Phenols mg/l	
m) Sulfate mg/l	
n) Total Sulfide mg/l	
o) Total Sulfur mg/l	
p) Thiocyanate mg/l	
q) Thiosulfate mg/l	
r) Volatile Organic Acids mg/l	
s) COD mg/l	

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 1 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 2 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 3 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 4 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 6 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 7 of 10)

MATERIAL BALANCE													
STREAM NUMBER	C-122 EFFLUENT		RECYCLED PHOSPHATE GAS		PHOSPHATE GAS		GAS FROM WITNESS		WASTE WATER FROM C-122		WASTE WATER FROM C-122		TO THE
STREAM NAME	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	TO THE
SOLIDS													
CARBON													
HYDROGEN													
NITROGEN													
SULFUR													
OXYGEN													
ASH													
TOTAL	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	
LIQUIDS													
H ₂ O													
HCN													
NH ₃													
TOTAL	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	
GAS													
H ₂													
CO													
CO ₂													
H ₂ S													
CS ₂													
N ₂													
NH ₃													
AT													
H ₂ O													
QA													
TOTAL	122.84	22.22	94.87	22.22									
MOLE WEIGHT (GAS)	26.46	26.46	26.46	26.46									
TEMPERATURE (°F)	140	140	140	140									
PHOSPHATE (PPM)	245	245	245	245									
TOTAL LB/HR	5726												

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 8 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 9 of 10)

Figure A-3. Process Flow Diagram, Pennsylvania Coal (Sheet 10 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 1 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 2 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 3 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 4 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 5 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 6 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 7 of 10)

MATERIAL BALANCE													
STREAM NUMBER	4	5	6	7	8	9	10	11	12	13	14	15	16
STREAM NAME	CH22 EFFLUENT	RECYCLED PRODUCT GAS	PRODUCT GAS	SEPARATED BY SEPARATOR	WASTE WATER FROM SEPARATOR	WASTE WATER FROM GAS COMPRESSION	WASTE TO DISPOSAL						
SOLIDS	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR	WT %	LB/HR
CARBON													
HYDROGEN													
NITROGEN													
SULFUR													
OXYGEN													
ASH													
TOTAL													
LIQUIDS	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR
H2O	7840								7840				7840
H2S													
TOTAL	7840								7840				7840
GAS	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR	MOLE %	LB/HR
CH4			4.54	5.33	4.54	5.33							
CO			25.05	29.54	25.05	29.54							
CO2			37.48	44.16	37.48	44.16							
H2			16.05	19.07	16.05	19.07							
N2			0.22	0.26	0.22	0.26							
CO													
N2			0.33	0.39	0.33	0.39							
H2O													
AF			1.53	1.82	1.53	1.82							
TOTAL			111.65	133.01	111.65	133.01							
MOL. WEIGHT (GAS)			25.05	29.54	25.05	29.54							
TEMPERATURE °F			140	140	140	140							
PRESSURE PSIA			245	245	245	245							
TOTAL LB/HR	7840		2865		336						2865		

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 8 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 9 of 10)

Figure A-4. Process Flow Diagram, N.D. Lignite (Sheet 10 of 10)

TP-
STEADY STATE -

PROXIMATE WT. %	FEEDSTOCK SC-9 +SC-5		BED (SC-22)		FINES (SC-41)				ASH LOCKHOPPER (C-117)		
Volatile Matter Fixed Carbon Moisture Ash <u>ULTIMATE, WT. %</u> Carbon Hydrogen Oxygen (By Difference) Nitrogen Sulfur											
SOLIDS SIZE DISTRIBUTION	WT. %	BULK DENSITY LB/FT ³	WT. %	BULK DENSITY LB/FT ³	WT. %	BULK DENSITY LB/FT ³	WT. %	BULK DENSITY LB/FT ³		WT. %	BULK DENSITY LB/FT ³
<u>Mesh Size</u> +6 -6 +10 -10 +20 -20 +40 -40 +70 -70 +100 -100 +200 -200 +270 -270 +400 -400									<u>Mesh Size</u> +4 -4 +8 -8 +12 -12 +16 -16 +20 -20 +40 -40 +70 -70		

Figure A-5 Solids Analysis

TP-
STEADY STATE-

	Fines (SC-44)	Fines (SC-41)	
Solids Size Distribution	WT. %	WT. %	
<u>Micron Size</u>			
1.1			
1.3			
1.5			
1.8			
2.1			
2.5			
2.9			
3.5			
4.1			
4.8			
5.7			
6.7			
7.9			
9.4			
11.1			
13.0			
15.4			
18.2			
21.4			
25.3			
29.9			
35.2			
41.6			

FIGURE A-6 SOLIDS SIZE DISTRIBUTION

TP-
ANALYSIS OF SOLIDS
SAMPLE NUMBER - _____

STEADY STATE

PERCENT AS RECEIVED

Sulfur							
Pyritic Sulfur							
Sulfate Sulfur							
Organic Sulfur							
Chlorine							
Fluorine							

FIGURE A-7 SOLIDS ANALYSIS

TP-
ANALYSIS OF SOLIDS
SAMPLE NUMBER - _____

STEADY STATE

PERCENT AS RECEIVED

Chlorine

Fluorine

FIGURE A-8 HALOGEN ANALYSIS

ANALYSIS OF ASH

TP-

STEADY STATE

SAMPLE

Analyses, Wt. % of Ash	Feedstock (SC-5 and SC-9)	Collected Fines (SC-20)	Escaped Fines (Isokinetic)	Ash (C-117)
SiO ₂				
Al ₂ O ₃				
CaO				
MgO				
Na ₂ O				
K ₂ O				
Fe ₂ O ₃				
SO ₃				

FIGURE A-9 ASH ANALYSIS

TP-

MINISCRUBBER ANALYSIS OF RAW PRODUCT GAS
(Water Soluble Compounds Only)

STEADY STATE
SAMPLE NUMBERS

pH

Parts Per Million By Weight (Dry Basis)

Arsenic

Chloride

Total Cyanide

Fixed Ammonia Nitrogen

Free Ammonia Nitrogen

Total Kjeldahl Nitrogen

Lead

Phenol

Potassium

Sodium

Vanadium

FIGURE A-10 RAW PRODUCT GAS ANALYSIS

SUMMARY OF OPERATING DATA
GASIFIER OXYGEN TEST TP-____
GASIFIER AIR TEST TP-____

SET POINT		1	2	3	4	5
TEST RUN DATE AND TIME ( )	Unit	Hours	Hours	Hours	Hours	Hours
<u>GASIFIER TEMPERATURES*</u>						
TE-504-4 29'-3 1/2", 315°	°F					
TE-507-1 23'-4 1/16", 0°	°F					
TE-507-12 23'-4 1/16", 120°	°F					
TE-504-5 22'-1 3/8", 315°	°F					
TE-504-6 20'-3 3/8", 315°	°F					
TE-504-7 15'-1 7/8", 330°	°F					
TE-507-3 12'-7/16", 180°	°F					
TE-507-2 12'-7/16", 0°	°F					
TE-507-11 10'-11", 225°	°F					
TE-126 9'-5", 330°	°F					
TE-507-4 8'-5", 120°	°F					
TE-504-9 6'-11", 90°	°F					
TE-507-10 6'-2", 225°	°F					
TE-121 6'-2", 330°	°F					
TE-504-10 5'-2", 220°	°F					
TE-504-8 1'-9", 270°	°F					
<u>FLUIDIZED BED PARAMETERS</u>						
Average Bed Height	feet					
System Pressure	psig					
PDT-22 Bed Density, 22'-31"	lb/ft ³					
PDT-23 Bed Density, 20'-22"	lb/ft ³					
PDT-24 Bed Density, 14'-20"	lb/ft ³					
PDT-26 Bed Density, 12'-14"	lb/ft ³					
PDT-122 Bed Density, 7'-12"	lb/ft ³					
PDT-121 Bed Density, 5 1/2'-7"	lb/ft ³					
PDT-120 Bed Density, 4'-5 1/2"	lb/ft ³					
PDT-119 Bed Density, 4'-12"	lb/ft ³					

*0° PLANT NORTH, CLOCKWISE

FIGURE A-11 OPERATING DATA SUMMARY (Sheet 1 of 4)

SUMMARY OF OPERATING DATA
GASIFIER OXYGEN TEST TP-____
GASIFIER AIR TEST TP-____

SET POINT		1	2	3	4	5
TEST RUN DATE AND TIME ( )	Unit	Hours	Hours	Hours	Hours	Hours
<u>REACTOR FLUID DYNAMIC DATA</u>						
Upper Annular Velocity (TE-504-9)	fps					
Middle Annular Velocity (TE-504-10)	fps					
Above Sparger Ring Velocity (TE-504-10)	fps					
30" Freeboard Gas Velocity	fps					
24" Bed Gas Velocity	fps					
Oxidant Tube Velocity	fps					
Auxiliary Tube Steam Velocity						
<u>OVERALL PROCESS RATES</u>						
Recycle Gas/Coal Ratio, MAF**	lb/lb					
Steam/Coal Ratio, MAF	lb/lb					
Total Moisture/Coal Ratio, MAF	lb/lb					
O ₂ /Coal Ratio, MAF	lb/lb					
Gas Make/Coal Ratio, MAF	scf/lb					
<u>OXIDANT TUBE INPUTS</u>						
Air Flow, FT-37	lb/hr					
Air Heater Temperature, TE-503-9	°F					
Oxygen Flow, FT-15	lb/hr					
Steam Flow, FT-35	lb/hr					
Diluent CO ₂ Flow, FT-16	lb/hr					
O ₂ /CO ₂ /H ₂ O Heater Temp. TE-507-5	°F					
Oxygen Concentration in Feed Gas	%					

**Moisture and Ash Free

FIGURE A-11 OPERATING DATA SUMMARY (Sheet 2 of 4)

SUMMARY OF OPERATING DATA
GASIFIER OXYGEN TEST TP-____
GASIFIER AIR TEST TP-____

SET POINT		1	2	3	4	5
TEST RUN DATE AND TIME ( )	Unit	Hours	Hours	Hours	Hours	Hours
<u>GAS INLET CONDITIONS</u>						
Recycle Gas to Grid, FT-12	lb/hr					
Recycle Gas Temp., TE-67	°F					
Steam to Grid, FT-34	lb/hr					
Steam Temp., TE-506-10	°F					
Recycle Gas to Booster, FT-11	lb/hr					
Recycle Gas Temp., TE-67	°F					
Steam to Booster, FT-36	lb/hr					
Steam Temp., TE-506-12	°F					
Recycle Gas to Sparger, FT-48	lb/hr					
F-113 Heater Temp., TE-503-2	°F					
Auxiliary Tube Steam, FT-38	lb/hr					
Steam Temp., TE-506-9	°F					
Water Injection to Gasifier	GPH					
Temperature	°F					
Axial Feed Transport Gas, FT-60	lb/hr					
Axial Feed Temp., TE-503-6	°F					
Axial Feed Transport Gas, FT-21	lb/hr					
Axial Feed Temp., TE-505-2	°F					
Axial Feed Transport Gas, FT-61	lb/hr					
Axial Feed Temp., TE-505-6	°F					
Radial/Axial Feed Transport	lb/hr					
Gas, FT-62						
Radial/Axial Feed Temp.,	°F					
TE-512-8						
Axial Feed Temp. to Gasifier	°F					
TE-505-1						

FIGURE A-11 OPERATING DATA SUMMARY (Sheet 3 of 4)

SUMMARY OF OPERATING DATA
GASIFIER OXYGEN TEST TP-____
GASIFIER AIR TEST TP-____

SET POINT		1	2	3	4	5
TEST RUN DATE AND TIME ( )	Unit	Hours	Hours	Hours	Hours	Hours
<u>SOLID INPUTS AND OUTPUTS</u>						
Axial Fines Feed Material	-					
Axial Feed Rate, WR-14	lb/hr					
Axial Coal Feed Material	-					
Axial Feed Rate, WR-27-WR-29	lb/hr					
Cyclone Collection Rate, WR-19	lb/hr					
Carryover to Water System*	lb/hr					
Ash Withdrawal Rate	lb/hr					
<u>PRODUCT GAS ANALYSIS (DRY BASIS)</u>						
Carbon Monoxide (CO)	%					
Carbon Dioxide (CO ₂)	%					
Methane (CH ₄)	%					
Nitrogen (N ₂)	%					
Oxygen (O ₂)	%					
Hydrogen (H ₂)	%					
Hydrogen Sulfide (H ₂ S)	%					
Higher Heating Value, Dry Basis	Btu/scf					
Moisture Content in Raw Gas	%					
<u>SOLID ANALYSIS</u>						
Ash Content - Radial Feed	%					
Ash Content - Coaxial Feed	%					
Ash Content - Bed	%					
Ash Content - Agglomerate	%					
Ash Content - Cyclone Fines	%					

*Estimated from quench water samples and/or isokinetic probe.

FIGURE A-11 OPERATING DATA SUMMARY (Sheet 4 of 4)

Steady State	Wc	Xc	Wa	Xa	Wp	Xp	Net Carbon Consumed #/hr

Steady State	Carbon in Coal	Total Carbon Out	Carbon Conversion

FIGURE A-12 CARBON UTILIZATION

Figure A-13 Heat and Material Balance Flow Schematic

GASIFIER HEAT AND MATERIAL BALANCE

PAGE 1

TEST NO. : SAMPLE
POINT NO. :DATE :
TIME :ISSUED BY :
COMPLETED BY :DATE :
DATE :

STREAM NO.	1	2	3	4	5	6
STREAM DESCRIPTION	STEAM TO GRID	STEAM TO AIR TUBE	AIR	CO2 PURGE	GRID GAS	SPARGER RING
TOTAL LB/HR	0.00	0.00	0.00	0.00	0.00	0.00
TEMPERATURE F	0.0	0.0	0.0	0.0	0.0	0.0
GAS LB/HR	0.00	0.30	0.00	0.00	0.00	0.00
SOLID LB/HR	0.00	0.30	0.00	0.00	0.00	0.00
LIQUID LB/HR	0.00	0.30	0.00	0.00	0.00	0.00
MOLECULAR WEIGHT	0.000	0.000	0.000	0.000	0.000	0.000
CO VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
CO2 VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
H2 VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
CH4 VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
N2 VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
O2 VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
H2O VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
GAS CHARACTERIZATION						
H2S PPM	0.000	0.000	0.000	0.000	0.000	0.000
COS PPM	0.000	0.000	0.000	0.000	0.000	0.000
SOLIDS						
CARBON DRY %	0.000	0.000	0.000	0.000	0.000	0.000
HYDROGEN DRY %	0.000	0.000	0.000	0.000	0.000	0.000
OXYGEN DRY %	0.000	0.000	0.000	0.000	0.000	0.000
NITROGEN DRY %	0.000	0.000	0.000	0.000	0.000	0.000
SULFUR DRY %	0.000	0.000	0.000	0.000	0.000	0.000
ASH DRY %	0.000	0.000	0.000	0.000	0.000	0.000

THE INPUT STREAMS ARE AS FOLLOW
LINE NO. STREAM NO.

GAS

1	5	GRID FLOW	RECYCLE GAS
2	6	SPARGER FLOW	RECYCLE GAS
3	7	BOOSTER GAS	RECYCLE GAS
4	8	TRIAXIAL TRANSPORT GAS	RECYCLE GAS
5	10	C120 FINES FEED TRANSPORT GAS	RECYCLE GAS

C-113 TOP TEMPERATURE 0.0 F. SYSTEM PRESSURE 0.0 PSIG
C-122 TOP TEMPERATURE 0.0 F. K-101 SUCTION PRESSURE 0.0 PSIG

Figure A-14 Heat and Material Balance Format (Sheet 1 of 3)

GASIFIER HEAT AND MATERIAL BALANCE

PAGE 2

TEST NO. :SAMPLE		DATE :		ISSUED BY :		DATE :	
POINT NO. :		TIME :		COMPLETED BY :		DATE :	
STREAM NO.		7	8	9	10	11	12
STREAM DESCRIPTION		BOOSTER GAS	COAL FEED AND TRANS. G.	OXYGEN TO TUBE	C120 CYCLOHE TRANS. G	LIQUID WATER INJECTION	PRODUCT GAS
TOTAL	LB/HR	0.00	0.00	0.00	0.00	0.00	0.00
TEMPERATURE	F	0.0	0.0	0.0	0.0	0.0	0.0
GAS	LB/HR	0.00	0.00	0.00	0.00	0.00	0.00
SOLID	LB/HR	0.00	0.00	0.00	0.00	0.00	0.00
LIQUID	LB/HR	0.00	0.00	0.00	0.00	0.00	0.00
MOLECULAR WEIGHT		0.000	0.000	0.000	0.000	0.000	0.000
CO	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
CO2	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
H2	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
CH4	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
N2	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
O2	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
H2O	VOLUME %	0.00	0.00	0.00	0.00	0.00	0.00
GAS CHARACTERIZATION							
H2S PPM		0.000	0.000	0.000	0.000	0.000	0.000
COS PPM		0.000	0.000	0.000	0.000	0.000	0.000
SOLIDS							
CARBON	DRY %	0.000	0.000	0.000	0.000	0.000	0.000
HYDROGEN	DRY %	0.000	0.000	0.000	0.000	0.000	0.000
OXYGEN	DRY %	0.000	0.000	0.000	0.000	0.000	0.000
NITROGEN	DRY %	0.000	0.000	0.000	0.000	0.000	0.000
SULFUR	DRY %	0.000	0.000	0.000	0.000	0.000	0.000
ASH	DRY %	0.000	0.000	0.000	0.000	0.000	0.000

THE INPUT STREAMS ARE AS FOLLOW
 LINE NO. STREAM NO.

		GAS	
1	5	GRID FLOW	RECYCLE GAS
2	6	SPARGER FLOW	RECYCLE GAS
3	7	BOOSTER GAS	RECYCLE GAS
4	8	TRIAXIAL TRANSPORT GAS	RECYCLE GAS
5	10	C120 FINE FEED TRANSPORT GAS	RECYCLE GAS

C-120 TOP TEMPERATURE 0.0 F. SYSTEM PRESSURE 0.0 PSIG

C-120 TOP TEMPERATURE 0.0 F. K-101 SUCTION PRESSURE 0.0 PSIG

Figure A-14 Heat and Material Balance Format (Sheet 2 of 3)

GASIFIER HEAT AND MATERIAL BALANCE

PAGE 3

TEST NO. : SAMPLE
POINT NO. :DATE :
TIME :ISSUED BY :
COMPLETED BY :DATE :
DATE :

STREAM NO.		13	14		
STREAM DESCRIPTION		ASH WITH-DRAWAL	BED SAMPLE (SC-22)	HEAT LOSS	TOTAL CLOSURE
TOTAL	LB/HR	0.00	0.00	0.00	0.00
TEMPERATURE	F	0.0	0.0	0.0	0.0
GAS	LB/HR	0.00	0.00	0.00	0.00
SOLID	LB/HR	0.00	0.00	0.00	0.00
LIQUID	LB/HR	0.00	0.00	0.00	0.00
MOLECULAR WEIGHT		0.000	0.000	0.000	0.000
CO	VOLUME %	0.00	0.00	0.00	0.00
CO2	VOLUME %	0.00	0.00	0.00	0.00
H2	VOLUME %	0.00	0.00	0.00	0.00
CH4	VOLUME %	0.00	0.00	0.00	0.00
H2	VOLUME %	0.00	0.00	0.00	0.00
O2	VOLUME %	0.00	0.00	0.00	0.00
H2O	VOLUME %	0.00	0.00	0.00	0.00
GAS CHARACTERIZATION					
	H2S PPM	0.000	0.000	0.000	0.000
	COS PPM	0.000	0.000	0.000	0.000
SOLIDS					
CARBON	DRY %	0.000	0.000	0.000	0.000
HYDROGEN	DRY %	0.000	0.000	0.000	0.000
OXYGEN	DRY %	0.000	0.000	0.000	0.000
NITROGEN	DRY %	0.000	0.000	0.000	0.000
SULFUR	DRY %	0.000	0.000	0.000	0.000
ASH	DRY %	0.000	0.000	0.000	0.000

THE INPUT STREAMS ARE AS FOLLOW
LINE NO. STREAM NO.

GAS

1	5	GRID FLOW	RECYCLE GAS
2	6	SPARGER FLOW	RECYCLE GAS
3	7	BOOSTER GAS	RECYCLE GAS
4	8	TRIAXIAL TRANSPORT GAS	RECYCLE GAS
5	10	C100 FINES FEED TRANSPORT GAS	RECYCLE GAS

C-110 TOP TEMPERATURE 0.0 F, SYSTEM PRESSURE 0.0 PSIG
C-122 TOP TEMPERATURE 0.0 F, K-101 SUCTION PRESSURE 0.0 PSIG

Figure A-14 Heat and Material Balance Format (Sheet 3 of 3)