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**Second-Generation Pressurized Fluidized-Bed Combustion
Plant -- Conceptual Design and Optimization of
A Second-Generation PFB Combustion Plant
Phase 2**

**Annual Report
October 1991 - September 1992**

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**For
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Office of Fossil Energy
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October 1992

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Section 1

INTRODUCTION

Section 1

INTRODUCTION

1.1 GENERAL

After many years of experimental testing and development work, coal-fired pressurized fluidized bed (PFB) combustion combined-cycle power plants are moving toward reality. Under the U.S. Department of Energy's Clean Coal Technology Program, a 70-MWe PFB combustion retrofit, utilizing a 1525°F gas turbine inlet temperature, has been built and operated as a demonstration plant at the American Electric Power Company's Tidd Plant in Brilliant, Ohio. As PFB combustion technology moves closer and closer to commercialization, interest is turning toward the development of an even more efficient and more cost-effective PFB combustion plant. The targeted goals of this "second-generation" plant are a 45-percent efficiency and a cost of electricity (COE) that is at least 20 percent lower than the COE of a conventional pulverized-coal (PC)-fired plant with stack gas scrubbing. In addition, plant emissions should be within New Source Performance Standards (NSPS) and the plant should have high availability, be able to burn different ranks of coal, and incorporate modular construction technologies.

In response to this need, a team of companies led by Foster Wheeler Development Corporation (FWDC) and consisting of:

- Gilbert/Commonwealth (G/C)
- Institute of Gas Technology (IGT)
- Westinghouse Electric Corporation, Power Generation Business Unit (W-PGBU)
- Westinghouse Electric Corporation, Science & Technology Center (W-STC)

has embarked upon a 5-year program to develop this new type of plant.

1.2 DESCRIPTION OF PROPOSED SECOND-GENERATION PLANT

The proposed plant is relatively simple and is a logical extension of first-generation-type plant technology (gas turbine inlet temperatures of approximately 1600°F).

It utilizes a "steam-cooled" PFB combustor (PFBC)—fluidized bed combustion heat release is absorbed by water/steam-cooled tubes—operating at 14 atm/1600°F with a conventional 2400 psi/1000°F/1000°F/2.5 in. Hg steam cycle. To reach efficiency levels that are higher than first-generation-type PFB combustion plants, power output is shifted from the steam turbine to the gas turbine via high excess air (greater than 100 percent rather than 30 percent) and an increased gas turbine inlet temperature (2000°F and higher).

Most of the first-generation PFB combustion plant designs have been based on a 1650°F bed temperature, even though test units have operated as high as 1750°F, because higher temperatures increase the alkali vapor release rate, in-bed tubewall temperatures, and the risk of bed/ash sintering and agglomeration—all of which can lead to gas turbine and PFBC hot corrosion and fluidizing problems.

To achieve a significantly higher gas turbine inlet temperature without increasing bed temperature, topping combustion has been incorporated into the proposed cycle. In this arrangement a fuel supply subsystem generates a coal-derived low-Btu fuel gas that is burned to increase the gas turbine inlet temperature. The fuel supply subsystem is a coal pyrolysis unit/carbonizer operating at approximately 1600°F and in parallel with the PFBC. Its fuel gas is burned in a topping combustor immediately upstream of the gas turbine with high excess air flue gas from the PFBC.

The coal char residue from the carbonizer is burned in the PFBC along with the balance of the plant coal, if any. Lime-based sorbent is injected into the carbonizer and PFBC to minimize carbonizer tar yield and desulfurize the gases from both units. The carbonizer is configured as a jetting fluidized bed to minimize tendencies toward agglomeration, and the PFBC takes the form of a refractory-lined circulating PFBC (CPFBC) with an external heat exchanger.

The 1600°F exhaust gases leaving the carbonizer and the CPFBC contain char, sorbent, coal, and fly ash—all of which can erode and foul downstream equipment. To prevent erosion and fouling, a hot gas cleanup (HGCU) system, consisting of ceramic cross-flow filters and assisted by cyclone separators, cleans these gases of their particulates (solids loading ≤ 20 ppm) before the gases enter the fuel gas topping combustor and the gas turbine. Ceramic candle filters, screenless granular-bed filters, etc., are also candidate alternatives for the cross-flow filter should their performance and economics be found superior.

The carbonizer fuel gas yield, composition, and heating value are very important in determining the efficiency of this plant. Assuming 2.9 percent sulfur Pittsburgh No. 8 coal and Plum Run dolomite are used as feedstocks and a plant operating with a 1500°F carbonizer, a 2100°F topping combustor exit temperature is optimum; and a plant efficiency of 43.6 percent and a cost of electricity (COE) about 19 percent lower than that of a conventional PC-fired plant with a scrubber are projected. Operation with a 1600°F carbonizer temperature increases the optimum topping combustor temperature to 2350°F and the plant efficiency to 44.9 percent; the COE becomes 22 percent lower than that for a conventional PC-fired plant.

Figure 1.1 is a heat and material balance for the referenced 14 atm/1600°F carbonizer second-generation PFB combustion plant operating with a 2350°F gas turbine. Since the carbonizer, CPFBC, FBHE, gas turbine, and stack gas heat recovery steam generator operate as an integrated subsystem, they

form a modular building block for achieving plants with varying electrical output. Although only one module is shown in Figure 1.1, the flow rates listed are the totals for the plant, and they are divided equally between the two modules.

The plant efficiency and COE data given indicate an incentive for higher carbonizer operating temperatures. However, experimental tests must be conducted to identify the maximum temperature for carbonizer operation without creating agglomeration and NO_x and alkali release problems. Although data are lacking, a 1500 to 1600°F and possibly 1700°F operating range is anticipated for the plant.

The key components in the proposed second-generation plant are the carbonizer, CPFBC, ceramic cross-flow filter, and topping combustor. Unfortunately, none of these components has been operated at proposed plant operating conditions, and experimental tests must be conducted to explore/determine their performance throughout the proposed plant operating envelope.

1.3 PROPOSED PROGRAM

A three-phase program plan has been identified to address these risks and provide test data to validate plant performance and cost estimates. The program is structured into three phases, with go-no go decision points between each phase. During Phase 1, the system was defined and optimized, the sensitivity of COE parametrically investigated, R&D needs identified, a program plan for solving the needs developed, commercialization potential addressed, and preliminary planning undertaken for Phases 2 and 3. The major thrust of Phase 2 is to design, construct, test, and evaluate the performance of the key components of the proposed plant (i.e., carbonizer, CPFBC, ceramic cross-flow filter, and multiannular swirl burner topping combustor)—all as separate components. In Phase 3 these key components will be designed, constructed, and tested at a 1.1-MWe (equivalent) scale as a fully integrated subsystem. This effort will allow component performance data to be collected on a larger scale and, most important, integrated subsystem performance data obtained and proposed plant operating modes/philosophies investigated. In Phases 2 and 3, the commercial/baseline plant design and R&D program plan prepared in Phase 1 will be updated to reflect the results of Phases 2 and 3 testing.

1.4 PHASE 2 SCHEDULE AND OBJECTIVES

Figure 1.2 is the schedule and milestones originally planned for Phase 2. The objective of each of the work elements shown is discussed in this section. The balance of this report contains a discussion of the technical work performed by the team during the period shown on the report cover.

8. Submit test plan revisions for approval.
9. Begin shakedown.
10. Begin testing.
11. All testing completed.
12. Data reduction completed.
13. Submit topical report for approval.
14. DOE approves topical report.

and CPFBC with request to fabricate.

CPFBC.

Figure 1.2 Original Program Schedule—Phase 2

Task 1—Test Plan

Objective: To prepare an overall logical Test Plan which, when executed by the team members, addresses the technology R&D gaps for the various system components identified in Phase 1, Task 2.

Workscope:

- Prepare a detailed Test Plan for evaluating individual component performance and developing a data base from which predictive correlations can be generated. This document will detail all test activities and identify decision points, costs, and schedules.

Task 2—Design

Objective: To submit a detailed design package for all components and, where applicable, test facilities.

Workscope:

- Prepare detailed drawings, compile costs, and determine fabrication/erection schedules.

Task 3—Fabrication/Modification and Shakedown

Objective: To fabricate, erect, check, and debug test facilities for the carbonizer/CPFBC, ceramic cross-flow filter, and topping combustor.

Workscope:

- Modify facilities, as required, and procure new components to enable testing of the carbonizer, CPFBC, ceramic cross-flow filter, and topping combustor.

Task 4—Test Plan Implementation

Objective: To conduct tests in the test facilities to characterize the performance of the various components under investigation.

Workscope:

- Establish test conditions according to Test Plan specifications.
- Conduct performance evaluation tests of the carbonizer, CPFBC, ceramic cross-flow filter, and topping combustor.
- Record and analyze all data.

Task 5—Phase 2 Topical Report

Objective: To deliver to the DOE COTR a topical report that details the results of the Phase 2 testing.

Workscope:

- Reduce and analyze data from Task 4 testing.
- Characterize the effects of operating parameters.
- Evaluate the test results in relationship to the designs and economics of both the baseline plant design and the Phase 3 Integrated subsystems test facility.
- Prepare a topical report that ties together the Test Plan, modifications, execution, and results, including:
 - Details of the impact that Phase 2 results have on the conceptual design (e.g., changes in specific design features, performance parameters, or operating economics)
 - Details of the impact that Phase 2 results have on the integrated program plan (e.g., schedules, costs, and scale of facilities)
 - An update of the plan for Phase 3 activities in light of Phase 2 results.
- Submit the edited report in draft form to the DOE COTR for approval at least 30 days before the completion of Phase 2.

Task 6 -Program Management/Reporting

Objective: To provide overall guidance and control of the various test programs so that the greatest amount of information is obtained for the least expenditure.

Workscope:

- Manage all aspects of the program, including monitoring, cost tracking, reporting, and coordinating (by each team member).
- Prepare and transmit regular reports to the DOE COTR, as required by the contract.

Section 2.

TASK 2--DESIGN

Section 2

TASK 2—DESIGN

2.1 OBJECTIVE

The Task 2 objective was to submit a detailed design package for all components and, where applicable, test facilities.

2.2 OVERVIEW

2.2.1 Carbonizer/CPFBC Pilot Plant

In April 1990 FWDC began constructing a second-generation PFB combustion pilot plant at its John Blizard Research Center in Livingston, New Jersey. The facility is designed to test a second-generation carbonizer and then a CPFBC/FBHE. A ceramic cross-flow filter will be installed on both units as a final stage of particulate control. In the event problems should develop with the ceramic cross-flow elements, a "backup" candle element assembly is being provided that can be inserted within the same vessel to replace the cross-flow filter. Initially, the pilot plant will be operated in a coal devolatilization/carbonization mode; later it will be modified to facilitate coal/char combustion. Lime-based sorbents will be pneumatically injected into the fluidized bed to control gaseous sulfur emissions. The coal and sorbent will be pressurized in separate lock hoppers, with screw feeders controlling their feed rate into a common, pneumatic transport line that injects these materials into the carbonizer. Operating in the carbonizing mode, the maximum gross heat input will be $8.5(10^6)$ Btu/h, and the low-Btu fuel gas generated by the facility will be burned in a natural-gas-fired incinerator before venting it to the atmosphere. A carbonizer designed by FWDC and a ceramic cross-flow filter supplied by W-STC will be the first components tested.

Since a new facility and a new process (carbonization) are involved, the pilot plant will initially be operated with a two-cyclone particulate-removing gas cleanup (HGCU) system. To minimize interactions, the stainless steel cyclones have been placed in separate, refractory-lined pressure vessels and provided with separate ash cooling and depressuring systems. After sufficient operating experience is gained and the composition of the carbonizer-generated fuel gas is ascertained, the second-stage cyclone will be removed. The ceramic cross-flow filter will then be connected to the system.

The gas exhausting from the second-stage cyclone or from the ceramic cross-flow filter is lowered to atmospheric pressure as it passes through a choked-flow orifice. The high-velocity orifice jet discharges into an Incoloy-shrouded refractory-lined chamber, where a nitrogen-atomized nozzle injects spray water to cool the gas to approximately 350°F. Although the cooling is accomplished by a dry quench, a wire mesh demister is provided at the base of the spray tower to remove any water droplets or oil condensate.

The cooled gas passes through a baghouse filter, is incinerated, and is then discharged to the atmosphere through an elevated stack.

Figure 2.1 is a simplified schematic of the carbonizer pilot plant. The carbonizer is a 30-in.-O.D. x 34 ft-6 in.-tall refractory-lined pressure vessel. It has a 19-ft-deep jetting fluidized bed and a 9 ft 5-3/4 in.-tall freeboard (Figures 2.2 and 2.3). The carbonizer I.D. increases from the bottom to the top in three steps to enhance solids circulation and limit slugging and elutriation. The I.D. is 10 in. for the first 3 ft-7 in. of bed height, 12 in. to the top of the bed, and 14 in. in the freeboard. Air, coal, and lime-based sorbent are injected into the unit at a 40- to 60-ft/s jet velocity through a central, vertical, 1-in. Sch 80, stainless steel feed pipe at the bottom of the unit. Low-Btu fuel gas leaves through a 3-in.-I.D. radial nozzle at the top. Spent bed material and char drain through the annulus that surrounds the feed pipe, enter the packed-bed cooling section, and are cooled to 300°F by counterflowing nitrogen before leaving through a 4-in. drain. A pressurized screw feeder in the drain line controls the material drain rate/bed height and discharges to a lock hopper for depressuring and ultimate drum disposal.

The carbonizer will be operated with a 3 ft/s superficial gas velocity in the lower 10-in.-dia. section. Operating pressures will range from 10 to 14 atm and temperatures from 1500 to 1800°F. Although Pittsburgh No. 8 coal and Plum Run dolomite crushed to 1/8 in. x 0 will be the primary feedstock combination, Illinois No. 6 and Eagle Butte coals and Longview limestone will also be tested.

After carbonizer testing, the plant will be modified to incorporate a CPFBC and an external FBHE. Coal (later char generated during the carbonizer program) will be burned in the CPFBC with lime-based sorbents to control sulfur releases.

The coal and sorbent will be pressurized in separate lock hoppers, with screw feeders controlling their feed rate into a common pneumatic transport line that injects these materials into the CPFBC.

The CPFBC is a 30-in.-O.D. x 34 ft-6 in.-tall pressure vessel. The primary zone is approximately 4 ft 6-3/8 in. tall, and the secondary zone is 24 ft 4-5/8 in. tall. Each is refractory lined to yield diameters of 5 and 8 in. respectively. Coal, sorbent, and pneumatic transport air are injected at the bottom of the unit at a 40- to 60-ft/s jetting velocity through a central, vertical, 1-in. Sch 80 stainless steel pipe. An outer, concentric, 2-in. Sch 40 pipe injects fluidizing air around and at the base of the feed pipe. A nitrogen-aerated packed-bed cooler at the bottom of the CPFBC cools spent bed material to 300°F before depressuring in a lock hopper and disposal. Secondary air injection ports are provided at two elevations to study staging effects.

Figure 2.2 10-in. Carbonizer Test Unit

Figure 2.3 Carbonizer—Primary Cyclone Arrangement

The heat released during the combustion process is absorbed by a sorbent/fly ash mixture continuously circulated between the CPFBC and the FBHE. A cyclone separator atop the FBHE and a nonmechanical L-valve at the bottom control the circulation of solids entering the CPFBC 14-3/8 in. above the fluidizing air.

The FBHE is a 42-in.-O.D. x 34 ft-6 in.-tall pressure vessel, refractory lined to yield an 18-in. square bed and freeboard section. A 39-in.-tall (bottom-to-top tube centerline height) water-cooled tube bundle in the bed, consisting of eight 1-in.-O.D. Incoloy 800H tubes, cools the solids before their return to the CPFBC. An air-sparger-pipe arrangement injects fluidizing air at the bottom of the bed and allows solids to flow downward into the L-valve or through the bed-drain cooling section. A screw feeder immediately below the FBHE controls the bed drain rate and bed height. Raising and lowering the bed height controls the amount of tube surface immersed in the bed and thus the bed and CPFBC solids return temperature. The fluidizing air leaves the top of the FBHE and enters the CPFBC as secondary air. A dip-leg J-valve arrangement at the bottom of the cyclone provides the necessary CPFBC/ FBHE gas seal. Figure 2.4 depicts the interconnected CPFBC/FBHE.

2.2.2 Specialty Gas Measurements

Opacity Measurements. In an electricity-generating second-generation PFB combustion plant, a hot gas cleanup system consisting of a cyclone and a barrier-type ceramic filter will be provided to protect the gas turbine from erosion. Since ceramic filter element failures leading to gas turbine erosion are possible, it is desirable to continuously monitor the particulate level exiting the filter. Because of the high pressure and temperatures involved, no systems are commercially available for this service. As a step toward developing such a system, a conventional opacity meter (isolated from the process pressure and temperature by a nitrogen-purged dual quartz lens assembly) is provided downstream of the ceramic barrier (cross-flow) filter. A U-shaped pipe section (shown in Figure 2.5) facilitates these measurements. The downflow section has a 5-in. I.D. and sufficient straight/calming length to facilitate cascade impactor or isokinetic particle sampling measurements; this dual measurement will permit an accurate assessment of filter performance. The opacity beam will be shot down the 3-in. I.D. upflow leg, and attempts will be made to calibrate the opacity meter vis-a-vis the impactor or isokinetic measurements. At a minimum, we hope this opacity meter system will alert operating personnel to a sudden increase in particle breakthrough, thereby providing a qualitative (if not a quantitative) measure of filter performance.

Figure 2.6 is the general arrangement drawing of the opacity metering system. Figure 2.7 depicts the quartz lens assemblies that isolate the opacity meter transmitter and receiver from the process pressure and temperature. A nitrogen purge on the front face of the upstream lens keeps it clear of dust and protects it from the high process gas temperature. A similar (but two-pass) opacity meter system was tested in the Grimethorpe PFB facility. To

Figure 2.4 Integrated CPFBC/FBHE Unit

Figure 2.5 Cascade Impactor, FOAM, and Opacity Measurement Locations

Figure 2.6 Opacity Measurement System—General Arrangement

Figure 2.7 Opacity Meter Quartz Lens Assembly

minimize beam reflections, the Grimethorpe quartz windows had to be inclined to form a 70-deg angle with the light beam. The opacity meter purchased for the carbonizer/CPFBC pilot plant (a single-pass unit) should be less sensitive to the window-beam angle. This sensitivity was verified by placing a quartz window in the beam and varying the window-to-beam angle from 90 deg to approximately 45 deg. The relatively small change in light transmittance was easily compensated for by increasing the gain in the receiver amplifier. For obvious cost savings and simplicity, the windows have been oriented perpendicular to the beam. The chamber between the two quartz lenses will be pressurized to about half the process pressure. Any significant change in this pressure will indicate a failure of either of the two lenses and signal the upstream ball valve to close immediately. For safety purposes a double block-and-bleed ball valve arrangement has been provided between the process and each lens assembly.

Alkali Monitoring. The amount of alkali vapors contained in the gas entering and exiting the ceramic cross-flow filter is to be measured at the locations shown in Figure 2.5. These measurements are to be made with the DOE/METC Fiber Optic Alkali Meter (FOAM) and cross checked with an in-situ probe developed by W-STC.

The carbonizer fuel gas is combustible, and all equipment in the pilot plant tower must conform to the National Electric Code (Hazardous Gas Environment/Class I-Division II, Group C&D). However, since the FOAM system is not Code-approved for this environment, it will be used during CPFBC testing only.

The Westinghouse probe is shown in Figure 2.8. The probe draws an isokinetic sample through a high-temperature filter in the tip for removal of particulates, but not alkali vapors. The gas then passes into the cooling section to condense and collect alkalies. Cooling is achieved as a continuous supply of high-pressure gas flows into the concentric tube assembly in the center. The sample gas passes through the outer annulus and cools; vapor condenses on the walls.

After a sufficient sampling time, the entire probe is withdrawn through a double block-and-bleed-valve arrangement. Alkali compounds are collected from the cooling section using a rinse-and-recovery approach. Solid-phase alkalies are extracted from the particulate filter. The alkali compounds will be characterized to identify water-soluble, acid-soluble, and insoluble components. On-site analysis at FWDC's laboratories will be completed by selective-ion electrodes, atomic absorption, or inductively coupled plasma spectroscopy.

Isokinetic Particle Sampling. These measurements are being made with the alkali gas sampling probes fitted with a tip specifically designed for isokinetic sampling.

A. PITTSBURGH VALVE & FFG CO
41 MEAD AVE P.O. BOX 4155
PITTSBURGH PA 15202

B. HEID TOOL SUPPLY
2265 BLACK CREEK RD
MUSKEGON, MICHIGAN 49444

C. FRICK & LINDSAY CO
NEVILLE ISLAND-PGM., PA. 15225

D. SAGINAW STEERING GEAR DIV.
SAGINAW, MICHIGAN 48605

E. LEE SPRING CO.
30 MAIN STREET
BROOKLYN, N.Y. 11201

TITLE ALKAL PROOF			
ASSEMBLY			
DWG 6486F15			
REV 2			
ITEM	PART NAME	QTY	SIZE REFERENCE INFORMATION
01	PISTON ASST	ONE	
02	BASE	ONE	
03	SIDE RAIL	ONE	
04	SIDE RAIL	ONE	
05	MSG	ONE	
06	BRAKE	ONE	
07	PLATE	ONE	
08	DISC	ONE	
09	HANDLE	ONE	
10	BRKT	ONE	
11	BRKT	ONE	
12	BRAKE BASE	ONE	
13	SCREW	ONE	
14	REAR STOP	ONE	
15	FRONT STOP	ONE	
16	BUSHING	ONE

2.3 WORK PERFORMED

2.3.1 Carbonizer and Specialty Gas Measurements

The design of the carbonizer pilot plant, together with specialty gas measurement systems (i.e., freeboard gas sampling, alkali gas sampling, isokinetic particle sampling, and an opacity meter system), was completed in FY91. The pilot plant and the carbonizer were shown schematically in Figures 2.1 and 2.3. The plant was formally dedicated on October 23, 1991. The first test run was conducted the week of November 18, 1991, and the originally planned test program was completed the week of June 15, 1992.

On June 1, FWDC received approval from METC to interrupt the planned conversion of the carbonizer to the 8-in. CPFBC/FBHE configuration shown in Figure 2.4 and to conduct one test run with the carbonizer in the 8-in. circulating fluidized bed configuration depicted in Figure 2.9. To accomplish this, a refractory lined L-valve instead of the FBHE was installed under the primary cyclone, yielding the arrangement depicted in Figures 2.10 and 2.11. General arrangement, instrumentation, and L-valve detailed drawings were completed. The thermocouples, differential pressure-measuring transmitters, and control valves required for the circulating carbonizer test were ordered. Bids were solicited and orders placed for the porous-pipe-type aerators and pressure taps required for the L-valve.

Cascade Impactor Measurements. Southern Research Institute (SRI) has been contracted to provide two sampling systems to measure loadings and size distribution of the particles entering and leaving the pilot plant particle-collecting ceramic cross-flow filter.

The general layout of the sampling system planned for the cross-flow filter inlet is shown in Figure 2.12. Samples are obtained by allowing particulate-laden process gas to flow out through a 1/2-in. pipe mounted to a 6-in. block-and-bleed ball valve and into a vessel containing five cyclones (piped in series) for inertial separation of particle-size fractions. The cyclones are followed by an absolute filter to collect the smallest size fraction. The filtered gas then passes out of the vessel, where it is cooled in three stages. In the first cooling stage, which consists of a 40-in. length of 1/2-in. Incoloy 800H pipe, the gas temperature is lowered to below 1500°F. In the next cooling stage, a coiled 1/2-in. Type 316 stainless steel tube, it is lowered to below 250°F. Further cooling occurs as the gas passes through a knockout pot and flow-control valves. The sample gas (at ambient pressure after passing through the control valves) then vents through a final filter, a silica gel column, and flow turbine for measurement of sample flow rate and volume.

Initially, a cascade impactor (rather than the series cyclone train) was planned for sampling at the inlet of the cross-flow filter. However, the anticipated mass loading and size distribution of the particulate matter at this location revealed that the capacity of existing cascade impactors was two orders of magnitude too low for meaningful samples and data. Requirements

Figure 2.9 8-in. Circulating Bed Carbonizer Test Unit

Figure 2.10 Circulating Bed Carbonizer—General Arrangement

Figure 2.11 Circulating Bed Carbonizer Pilot Plant

Figure 2.12 Cross-Flow Filter Inlet Particle Sampling System

associated with the temperature and pressure make fabrication of an impactor with sufficient capacity impractical. Several grams per size fraction are anticipated for sampling runs of a few hours, obtainable with the existing series cyclone sampler.

The sampling system planned for the cross-flow filter outlet, shown in Figure 2.13, is similar to the one for the inlet, except that an existing cascade impactor (Andersen Mark V) has sufficient capacity, requiring only changes in the housing and interstage seals to accommodate pressure and temperature. The particulate mass loadings at this location are expected to be substantially less than the 8 to 40 ppm(v) expected for alkali vapors. Thus it is important to avoid condensation of alkali before the substrates are collected (the condensed vapors would add substantially to the sample weight). The sampling system has been designed to minimize cooling of the sample gas. We expect to limit the gas sample temperature loss to 50°F until the particulate samples are collected in the cascade impactor. This reduction in temperature will occur where sample gas exits the process piping. The ball valve and impactor housing will be heated externally, and the cyclone vessel will be heated internally.

The designs of both systems were completed; but because of the long delivery of the 270 psig/1650°F process isolation valves, neither system was available for use during the carbonizer test program. A cost estimate submitted to FWDC in June indicated it would cost an additional \$223,000 to procure materials, fabricate, and check out these systems. Because neither of these systems was available for the carbonizer test program, an isokinetic particle sampling tip was designed for the Westinghouse alkali sampling probes. Attempts will be made to use these probes to take the same measurements planned for the SRI systems. In the meantime, the SRI effort has been placed on hold until the effectiveness of the Westinghouse particle sampling probe is evaluated.

Alkali Gas Measurements. In 1992 an alkali sampling probe was designed, fabricated, and supplied by the W-STC to measure the alkali vapor content of both the carbonizer fuel gas and the CPFBC combustion gas. The alkali probe design has also been conceived of as functioning for the dual purpose of isokinetic particle sampling. A probe of this type was designed and operated by W-STC on the KRW gasifier PDU; it successfully provided this dual function. A variety of on-line analysis and batch sampling probe methods have been proposed and tested to measure the particulate and vapor alkali content of coal-derived gas streams. Because of the inherent difficulties in making on-line alkali measurements, and the questionable accuracy and interpretation of these methods, W-STC has chosen to use simple batch measuring methods that do not offer continuous, on-line measurements, but are reliable and capable of providing periodic data to characterize the alkali gas stream content.

Figure 2.13 Cross-Flow Filter Outlet Particle Sampling System

The types of batch systems considered by developers include a measured gas sample through:

- An absorbing fluid, followed by chemical analysis of the fluid
- A packed bed of adsorbent, or sorbent particles, followed by chemical analysis of the solids
- A cooled probe to condense the alkali vapors on the interior of the probe, followed by washing of the probe and analysis of the wash.

Since condensation of alkali vapor is difficult to avoid at various points in any measuring system, and because simple washing and chemical analysis of the wash are simpler and more accurate than chemical analysis of an adsorbent/or sorbent bed, W-STC has selected the condensation route as the best batch approach. Any of the other probe systems should require washing and analysis of the wash to account for condensation losses. W-STC has previously used this type of a probe successfully at three different locations:

- A bench-scale test facility at the W-STC site operated at high pressure and high temperature to measure the capture of alkali by packed beds of sorbents such as emathlite pellets. This configuration was a simple fixed-probe system requiring no on-line withdrawal.
- A Textron/AVCO direct coal-fired gas turbine combustor system operated at high pressure and temperatures of about 1850°F under oxidizing gas conditions—again, a simple, fixed probe.
- The KRW gasifier PDU operated at high pressures and temperatures of 1600°F and lower, under reducing gas conditions. This arrangement included a motorized probe withdrawal system.

A general arrangement drawing of the alkali probe designed for the FWDC facility was shown in Figure 2.8. The sampling tip extends into the center of the gas pipe. Separate sampling tips have been designed for alkali sampling and for particle sampling so that optimum performance can be achieved for each function. A hot filter is placed within the hot end of the probe to collect the particulate phase carried by the gas for separate alkali analysis; separate water and acid washes can be used to characterize the alkali vapor constituents that are condensed within the probe. The cold finger of the probe is cooled by an annular flow of nitrogen. The probe is enclosed in a pipe pressure boundary. Double block-and-bleed valves permit the removal of the probe, cleaning and replacement of the internals, and on-line reinsertion. A hand crank is operated to extract and insert the probe. During the probe design phase, FWDC provided significant input on the probe design relating to the design safety criteria. SRI provided input on the sampling tip design criteria for isokinetic particle sampling. Initial operation of the probe also led to probe modifications suggested by FWDC to improve operation of the probe and the quality of the data produced.

2.3.2 Ceramic Cross-Flow Filter

The cross-flow filter vessel installed at the FWDC facility is shown in Figure 2.14. The refractory-lined, pressure vessel has a 50-in. O.D. and is 206 in. tall. Process gas enters the vessel tangentially and flows primarily over the top of a cylindrical shroud that protects the cross-flow filter elements from direct gas impingement and promotes general downflow of the gas through the array of filter elements. We expected that downflow of gas through the vessel would be desirable to minimize reentrainment of fly ash. The vessel contains ten cross-flow filter elements arranged vertically on two parallel plenum pipes. It was designed to operate at a nominal carbonizer gas flow rate of about 120 aft^3/min and a nominal CPFBC gas flow rate of about 400 aft^3/min .

At the beginning of the 1991 reporting period, the filter vessel was installed at the FWDC test site; internals and the balance of the filter equipment were also at the site, but they were not scheduled to be installed/activated until after the carbonizer parametric test program was completed (in lieu of the barrier filter, the pilot plant was utilizing a second-stage cyclone). At that time, the only remaining design issue for the cross-flow filter vessel was the size of the taper gap that should be used at the base of the shroud. A reduced gap at the skirted bottom of the shroud promotes gas flow upward, over the shroud, but with a greater possibility of fly ash accumulation at the base of the shroud in the gap region.

Cold flow model testing had previously been conducted at W-STC, using an exact duplicate cross-flow filter Plexiglas cold model, with resolution of gap size as one of the objectives of the testing. This cold model testing, conducted earlier in the program, concluded that the major factor in selecting a gap size is the cohesive nature of the carbonizer fly ash. The cold model showed that a recirculation pattern always exists within the vessel, with gas swirling primarily downward near the inside wall and flowing moderately upward between the two filter plenums. The recirculation intensity increases as the gas inlet velocity increases and is not very sensitive to the size of the gap at the base of the shroud. The nature of the fly ash appears to control the filter cleaning performance more than the nature of the gas circulation pattern in the vessel. That is, if the fly ash, by its nature, does not agglomerate well as a filter cake, then reentrainment will occur—regardless of the gas circulation pattern in the bed. If the fly ash contains components that stick to the filter elements, then the cleaning performance can not be modified by control of the gas flow pattern. If the fly ash agglomerates well as a cake, then little reentrainment is expected—even if considerable upflow of gas exists in the filter vessel. SRI completed characterization of two samples of fly ash obtained from the first carbonizer test series completed. One sample was taken from the secondary cyclone catch and the other from the bag-house catch—the cross-flow filter receives a combination of these two. A list of the properties determined by SRI is given in Table 2.1.

Table 2.1 Carbonizer Fly Ash Physical Properties (SRI)

Properties Determined	Cyclone Ash	Baghouse Ash
Aerodynamic mean diameter, μm	20	---
Volumetric mean diameter, μm	9.5	7.3
True particle density, g/cm^3	2.03	1.96
Compacted bulk porosity, %	77	66
Estimated dust cake porosity, %	89	80
Specific surface area, m^2/g	6.8	4.1
Morphology factor	22	9.5
Drag-equivalent diameter, μm	1.30	0.97
Tensile strength, N/m^2	0.3	1.1
Relative gas flow resistance, $\text{H}_2\text{O}\cdot\text{min}\cdot\text{ft}/\text{lb}$	2.3	12

Although SRI concludes that these fly ashes are not as cohesive as many they have characterized, their conclusion is not a clear basis for judging the material.

W-STC previously explored the use of a simple tap density test as a means to judge the cohesive nature of powders, and this test was applied to several filter fly ashes to provide a data base for comparison. In this test, the fly ash is initially loosely packed in a column, followed by mechanical tapping, with the bulk density determined periodically. The ratio of the tap density to the initial loose-packed density is plotted as a function of the number of taps. The general guideline is that the powder can be classified as cohesive if this ratio exceeds 1.4 within 100 taps. The technique is highly reproducible and simple, and gives meaningful results. The results are shown in Figure 2.15 for Texaco fly ash (a very noncohesive fly ash suspected of being reentrained during filter testing), for Grimethorpe "red" fly ash (a very cohesive fly ash difficult to pulse clean in filter testing), and for the two carbonizer fly ashes. The carbonizer fly ashes fall between the two extremes represented by the Texaco and the Grimethorpe fly ashes. The cyclone catch meets the criteria for a cohesive powder, while the baghouse fly ash becomes cohesive after some consolidation. This cohesive behavior may promote good filter performance, indicating that the filter cake may be relatively easily pulse cleaned with relatively little reentrainment.

Additional particle samples from later carbonizer tests (secondary cyclone catch and baghouse catch) with three coals and two sorbents were tested to estimate their possible performance in the filter tests before the

Figure 2.15 Dependence of Tap/Bulk Density Ratio on Number of Taps

TR8 test campaign (the first test run to utilize a ceramic cross-flow filter). The simple tap-density test was conducted with 50% cyclone/50% baghouse mixtures of each sample. The tap-density ratio (ultimate tap density over the initial, loose powder density) has been measured for the highly reentrained Texaco fly ash as about 1.2, and for the difficult-to-clean Grimethorpe "red" fly ash at about 2.1. Other fly ashes with tap-density ratios between these two have been relatively easily cleaned with little reentrainment. Of course, other factors, such as the fly ash acid/base characteristics, may also need to be considered in this simplified correlation. The resultant tap density ratios were measured as:

- Pittsburgh No. 8 coal and Plum Run dolomite: 1.6
- Illinois No. 6 coal and Longview limestone: 1.45
- Eagle Butte coal and Longview limestone: 1.97

Based on the correlation, the Eagle Butte coal/Longview limestone carbonizer fly ash would be expected to be the most difficult of the three to pulse clean. The other two feedstock combinations would be expected to perform well, and very limited reentrainment would be expected with all three.

On the Texaco gasifier, W-STC has operated a filter vessel with nearly the same size and gas flow rate as that of the filter supplied to FWDC. The performance of the Texaco filter was reviewed to extract any information that might relate to the FWDC carbonizer cross-flow filter. The major differences in the Texaco and FWDC filter vessels are that the Texaco filter has a radial gas inlet and the FWDC vessel has a tangential gas inlet. The Texaco filter was operated with two significantly differing configurations, with:

- Four cross-flow filter elements and a simple deflection baffle on the radial inlet
- Eight cross-flow filter elements and a shroud with a 5/8-in. taper gap.

A comparison of the data has shown that the Texaco filter performance did not change too much with these significant changes in the design of the internals, within the uncertainty of the data; this comparison seems to confirm conclusions drawn from the cold flow model testing.

Based on the cold flow model test results and the Texaco filter testing, the size of the taper gap in the shroud for the carbonizer cross-flow filter vessel was set at 5/8 in.—the same as was used in the Texaco filter vessel. Based on inspections conducted in the Texaco filter vessel, little dust accumulation is expected within the shroud.

During 1992, new pulse gas eductors with superior performance have been designed to replace the existing venturi-type eductors in the cross-flow filter vessel. This new design was based on improved design procedures developed since the original eductors were fabricated.

Some further programming improvements were also completed for the cross-flow filter control and data acquisition system to ensure that the filter operator could make rapid changes in key control parameters. Other provisions for manual pulse cleaning and independent data recording have been added.

2.3.3 Ceramic Candle Filter

Early in 1992 the team decided that it would be prudent to have a backup candle filter design ready for installation in the same pressure vessel in the event that operating problems, such as difficult filter cake cleaning, arose with the cross-flow filter. A backup candle filter design was completed. The backup candle filter contains 22 candle elements, and the assembly fits into the existing filter tubesheet. A new shroud was also designed for the candle filter, extending the shroud below the tips of the candle elements. The new shroud was designed with no skirt taper at its base. The general layout of the candle filter is shown in Figure 2.16. The internals consist of two semicircular plenums, each holding 11 candle elements and each independently pulse cleaned. Two alternative candle layouts were considered (Figure 2.17), and the top layout was selected for the design. The candle filter elements are Coors/Westinghouse mullite candles, 1.5 m long. These are a new generation of candles made from the superior, oxide-based mullite ceramic—the same material used in the cross-flow filter elements.

2.3.4 Topping Combustor

Overview. In FY90 and FY91, four MASB tests were conducted at UTSI. The first three tests utilized a syngas that simulated the expected carbonizer fuel gas—two with a 12-in. MASB operating at approximately 2100°F and the third with a 14-in. MASB operating at approximately 2300°F. In the fourth test, distillate oil and natural gas were fired separately in the modified 12-in. MASB. Although the results of these tests have been presented in the previous annual reports, analyses continued in FY92 to determine why the conversion of fuel-bound nitrogen (FBN) to NO_x was higher in these units than expected (20 to 30 percent rather than 10 to 12 percent). Computational fluid dynamic (CFD) and chemical kinetic models prepared of the MASB have identified:

- The 800°F syngas temperature used at UTSI was too low to allow the desired primary zone ammonia reactions to occur (commercial plant value will be 1600°F). As a result, we have recommended that the preheat be raised to 1200°F, the highest that apparently can be achieved with commercially available equipment.

Figure 2.16 Candle Filter—General Arrangement

Figure 2.17 Alternative Candle Arrangements Considered for Carbonizer Barrier Filter

- Modify the primary zone design to permit the syngas to be mixed with a portion of the combustion air before it enters the fuel-rich ignition zone. This premixing should speed up rich-zone combustion and enable the completion of the pyrolysis reactions that destroy ammonia and hydrogen cyanide compounds before the oxygen-rich zones are reached.

The modeling work involved in these efforts is described in detail in the following subsections. Additional 12- and 14-in. MASB tests aimed specifically at NO_x suppression were prepared and were released to proceed in September 1992.

A multi-purpose Power Systems Development facility is to be constructed at a Southern Company Services, Inc., site in Wilsonville, Alabama. This facility will incorporate and test a second-generation/advanced PFB module with ≈ 7 MWe equivalent rating. In support of this effort, Westinghouse will design and supply two 18-in. MASBs (one will be a spare) to burn the fuel gas generated in the plant carbonizer. A full-scale replica of the 18-in. MASB flow path boundaries will be built by Westinghouse and tested at UTSI. Although the detailed design of the 18-in. UTSI MASB test unit is not scheduled to begin until October 1992, much of the referenced modeling efforts have centered on establishing a new primary zone arrangement for the unit.

Computational Fluid Dynamic (CFD) Modeling of the MASB.

General. As discussed in earlier reports, some of the experiments conducted at UTSI with the 2100 and 2350°F combustors showed FBN conversions greater than desired (i.e., in the range of 20 to 30 percent, whereas the desired conversion is 10 to 12 percent). In an attempt to determine why FBN conversion was this high, the data were analyzed in more detail, including the preparation of a comprehensive CFD model of the MASB. With regard to requirements, this model should correctly model fluid mechanics, combustion reaction, and NO_x formation. It should be able to determine the effect of varying such parameters as swirl angle, airflow through each swirler, preheat temperature, and change in gas composition. It should also be a tool for determining desired hardware and operating changes to improve performance.

An additional reason for development of a model is that actual testing of the MASB with low Btu gas is very expensive. Thus a CFD model could easily pay for itself by avoiding only a few tests.

Approach. In selecting and developing a model of the Westinghouse MASB, a number of factors were considered. The model had to be capable of:

- Handling complex, swirling, recirculating flow
- Handling combustion reactions
- Handling both equilibrium and kinetic reactions
- Predicting NO_x levels
- Handling complex, swirling boundary conditions
- Handling turbulent flow
- Running two-dimensional and three-dimensional problems
- Running with moderate computing resources and reasonable run times
- Restarting calculations for small changes in input parameters

It also had to incorporate the following attributes:

- Automatic mesh generation
- Automatic time-step control to maintain required accuracy constraints
- Built-in or easily added graphic output
- Low cost of code acquisition
- Availability of source code.

In general, these features would be difficult, if not impossible, to obtain in any single code. Essentially, development of such a code "from scratch" would also be impossible within the available time and budget.

With these criteria in mind, KIVA-11 (Amsden, 1989) was selected as the base program code for modeling the MASB. KIVA is considered an advanced experimental code—not a production program. Thus an expert user must often make changes to the actual code itself to model a particular problem. This procedure is contrary to various commercial codes that allow the user to solve wide ranges of problems by simply changing the input data. KIVA meets most of the desired criteria, and it has a modular construction that allows other criteria to be met.

KIVA-11 is a computer program that calculates time-dependent, two- or three-dimensional chemically reactive fluid flows with sprays. KIVA is applicable to laminar or turbulent flows; subsonic, transonic, or supersonic flows; and single-phase or dispersed two-phase flows.

KIVA solves the unsteady equations of motion of a turbulent, chemically reactive mixture of ideal gases, coupled to the equations for a single-component, vaporizing fuel spray.

As received at UTSI, KIVA was set up to run on a Cray supercomputer. It was also configured to model an internal combustion (piston) engine. KIVA can handle liquid fuel droplet evaporation and combustion, although this feature has not been used to date for MASB application. It can also handle moving boundaries (i.e., piston motion), which is also not required. KIVA can be run for either a two- or three-dimensional geometry. Mesh generation is partially automated for simple geometries.

A basic feature of KIVA is vectoring, which allows it to run efficiently on Cray supercomputers. Access to a Cray was not readily available at UTSI, but UTSI does have an Alliant mini-supercomputer that is architecturally similar to the Cray computers. Both machines have vectoring FORTRAN compilers, thus the programmer can optimize at the FORTRAN code level.

The first effort performed was to get KIVA to run on the Alliant at UTSI. Basically, this was done in two stages. In the first stage, KIVA was installed on the Alliant, with a minimum number of changes made to get the program to run the test case distributed with the program.

After making the appropriate changes, the test case was run, and results were compared with reported values. Debugging continued until reasonable results were obtained.

The next step in getting KIVA to run on the Alliant was to attempt to optimize the program. Changes made to optimize the program included:

- Use of the Alliant compiler vector optimization switches

- Conversion of the patched vector merge function calls to "IF-THEN-ELSE" statements
- Conversion of the Cray compiler directive "CVD\$IVDEP" to the more-or-less equivalent Alliant directive "CDIR\$NODEPCHK". Programmers found that these switches were functionally identical for vectoring, but were not identical when Alliant concurrence was attempted.

With the optimization completed, the test problem ran in about 10 minutes on the Alliant 4 CE complex or 20 minutes on a single CE with full optimization. Thus the program was set up to run on a single CE with full optimization. For comparison, the test program executes in about 1 minute on a Cray Y-MP. Later in the project, a Sun SPARCstation II was used; it runs at about twice the speed of a single CE on the Alliant.

Once KIVA was running efficiently for the test case supplied, the effort began to convert the program to simulate the Westinghouse MASB. As received, the program was set up to simulate an internal combustion engine. This simulation included spray of a liquid hydrocarbon fuel into an enclosed chamber with a moving piston. Initially, the chamber was filled with air. The piston further compressed the fuel. At a specified crank angle, an ignitor was simulated, the liquid fuel droplets were evaporated/moved/collided, and the fuel was vapor burned. By contrast, the MASB required:

- Inflow and outflow boundaries, with inflow boundaries located in multiple axial planes.
- Inflow of air with both axial and swirl components.
- Low-Btu fuel feed through a large central opening (rather than a small nozzle).
- Fuel with multiple fuel components (i.e., CH_4 , H_2 , and CO) vs. a KIVA assumption that only the first of the species is fuel. This assumption was changed in the program code.
- Various gas temperature, composition, and velocity components at different radial positions.
- Complex combustion chamber outline.

KIVA has a great deal of flexibility. However, much of this flexibility is directly related to options needed for internal combustion engine simulations. The mesh generator can produce grids by inputting only a piston, a head outline, or a combination. For the MASB, the "piston" shape was used to generate the mesh for the MASB shape, with some modification.

KIVA has a limited inflow and outflow capability. Basically, a gas of a fixed composition could enter at a single inflow plane at axial distance $z = 0.0$. Similarly, the gas leaves at a single plane at $z = \text{constant}$.

For the MASB simulation, a single outflow plane is sufficient. However, multiple inflow planes are required. The MASB outline is read in as a piston outline in the KIVA input file. Several inflow parameters were re-dimensioned from single variables to arrays of dimension NXP, where NXP is the number of x vertex locations. The inlet z (axial) and y (swirl) velocities, and species densities can be read from the input data file as a function of the radial position.

KIVA uses a flag (BOTIN = 1.0) to identify an inflow boundary simulation. Similarly, TOPOUT = 1.0 flags that an outflow boundary is being used. By searching for BOTIN most of the areas of the program that required modification were identified.

The general method used to modify the inflow boundary conditions was to replace the statements that stepped across the $z = 0.0$ plane with statements that stepped around the piston outline. Changes also were made in the definition of certain vertex flags to correctly identify which vertices were active (i.e., inside the combustor) and which were inactive.

Basically, routines that were modified were those where boundary conditions were initialized or set. Subroutine RINPUT was modified to read in the flow velocity, temperature, species density, and turbulence length scale arrays. Subroutine SETUP was modified to change settings of some cell and vertex flags. Subroutines BC, BCFC, and CCFLUX were modified to allow for the new inflow boundaries.

Results of Syngas Simulation Modeling. As a time-dependent program, KIVA requires a complete set of initial conditions (i.e., at time = 0.0) as well as well-defined boundary conditions.

At time = 0.0, the combustor is filled with a gas (usually air or nitrogen) at 1600°F with no velocity in any direction and at approximately 150 psig.

At time $t > 0.0$, an inflow of 1600°F air is specified to flow in through the swirlers. This air has an axial and a swirl velocity, both of which are specified. The syngas enters through the center area at a defined axial velocity (see Figure 2.24 on p. 42). The syngas swirl velocity is usually taken as zero, although a swirl velocity may be specified. The syngas is set to have an amount of mixing air premixed with it. The syngas inlet temperature is one of the variables in these runs.

In KIVA, the fuel is defined to be material component one. The combustion of the fuel is by a specified kinetic reaction. For our tests with syngas, however, there are two fuel components, methane and hydrogen. (For a real syngas, carbon monoxide would also be a significant fuel component.)

In a single kinetic reaction for methane combustion to carbon dioxide and water, no allowance is made for the hydrogen combustion. Hydrogen could be assumed to be at equilibrium, handled by the equilibrium reaction set. Calculations made with only a kinetic reaction for methane gave results that did not seem reasonable, however. In specific, the peak flame temperature would climb too high.

After considerable experimentation and discussions with others, a three-kinetic reaction combustion model was chosen. This reaction set includes:

- Methane partial combustion to carbon monoxide and hydrogen
- Hydrogen combustion
- Carbon monoxide combustion

Values for these reaction rate equations were taken from Westbrook and Dryer (1984), Dryer and Glassman (1972), and Westbrook (1991). Small reverse reaction rates were also added to avoid negative or zero concentrations.

Three kinetic reactions were also included in the kinetic reaction set for a simple thermal NO_x model. (This is probably not very meaningful with the FRN in the real syngas.)

In addition to the six kinetic reactions (three for combustion and three for thermal NO_x), five equilibrium reactions were included. These reactions are for disassociation and radical formation reactions that absorb considerable energy and moderate the flame temperature.

The initial calculations made were for the Westinghouse 2100°F MASB, with the flow-deflector installed. The main goal for these calculations was to examine the primary substoichiometric upstream zone of the combustor. For this purpose, only the upstream area of the combustor (including the syngas injection, first three swirlers, and flow deflector) was modeled. (Note: The flow accelerates through the flow deflector, as seen in following figures shown later in this section, and no recirculation or back-flow occurs. Thus the downstream and upstream of the combustor are effectively uncoupled, except for pressure. Because the overall pressure drop in this combustor is very low, the effect of downstream pressure drop on the upstream pressure can be neglected.)

In most of the following results, 4840 grid points were used in the calculations, representing a grid dimension of approximately 0.5 cm by 0.5 cm which is still fairly coarse. However, KIVA is robust and usually gives reasonable results for relatively coarse grids. (Note: Earlier calculations had modeled the entire combustor with an even coarser grid, but investigators thought that the coarse grid might have been part of the difficulty in obtaining and maintaining ignition. Physically, for a cell to support

combustion, the cell temperature must be above the fuel ignition temperature, with the fuel and oxygen concentrations within the flammability limits. As seen in later plots, the center core of fuel and outer annulus of hot air results in very high concentration and temperature gradients in the radial direction. Thus a grid that is too coarse may not be able to get temperatures and concentrations suitable for combustion. Finer grids can resolve these high-gradient zones much better, but only at the price of much longer computation times.)

This grid represents a compromise between grid resolution and computation time. With additional grid points, the computation time increases for two reasons. The first reason is simply the larger number of points that must be computed, with computation times roughly proportional to the number of grid points. A more significant problem is the constraint that, during any time step, the fluid cannot flow through more than some fraction of a cell anywhere in the computational grid. (Note: This is a numerical constraint that can be understood in a physical sense. If the fluid were allowed to traverse more than one grid step in a single time step, adjacent cells would be effectively uncoupled. That is, one cell would be dependent upon what happens in cells some distance away, but not upon adjacent cells. Numerically, multiple independent solutions can develop. For example, if odd-numbered cells are mutually dependent and even-numbered cells are mutually dependent, but odd- and even-numbered cells are mutually independent, two mutually independent solutions can be calculated.)

With a grid of 4840 points, a computation requires approximately 24 hours on a Sun SPARCstation 11.

A consideration is whether a converged solution has been reached. KIVA does a time-dependent calculation, and steady-state solutions are obtained by calculation until no significant change occurs between successive time steps. When KIVA is run for these calculations, plots are generated after certain redefined time increments. These plots are compared to determine when to stop computations. (Note: In practice this is not exactly true. In previous calculations with generally larger grid spacing, investigators often found that autoignition did not occur. In these cases an ignitor was simulated. Sometimes, after achieving ignition and turning the ignitor off, the flame would be swept out of the combustor. Of course, we know from experimental data that this does not occur, and combustion is stable in the MASB. In these time-dependent calculations with KIVA, the syngas enters a combustor filled with 1600°F air. Ignition thus starts on the leading edge of the slug of syngas. As the computation continues, the leading edge of the syngas "hot spot" travels by convection downstream through the combustor. Ignition seems to propagate outward from this center zone and radially outward, igniting the swirler air/fuel mixture at the air/fuel interface. Thus calculations were continued until the hot leading edge of the syngas exited by convection, but calculations stopped before the flame could be swept out. This method provides the closest simulation of actual test results at the minimum computational time. The correctness of this assumption will be tested in future simulations.)

Syngas Simulation Results. The first simulation reported was of the 2100°F MASB with 800°F syngas, as tested for Westinghouse at UTSI. Figure 2.18 shows the computational grid used for this run. This grid is 22 x 2 x 110, for a total of 4840 grid vertices.

Figures 2.19 through 2.22 show plots of temperature (in Kelvin) at four times during the computation. On all contour plots:

- "min" and "max" = minimum and maximum values found in the mesh
- "l" and "h" = values of the low and high contours (also shown as "L" and "H" on the plots)
- dq = difference between adjacent contour lines.

The plots are automatically scaled, thus leading to uneven contour line values.

Figures 2.19 through 2.22 show the ignition (high contour) at the front edge of the fuel "slug," which is moving toward the downstream end of the combustor as time progresses. Also note that the low-temperature contour penetrates very far into the combustor in Figure 2.22, and that this contour is at 789 K (960°F)—a temperature too low for any significant chemical reaction to take place.

Figure 2.22 indicates what we believe is the cause of the high conversion rates of ammonia to NO_x seen in the MASB tests conducted at UTSI. Discussions with Dr. Janos Beér (inventor of the MASB and a consultant to Westinghouse) confirm this. What is desired in the MASB is to have a sub-stoichiometric, but reactive, zone in the upstream, fuel-rich area of the combustor (i.e., up to the third-stage swirler for this combustor). However, for significant chemical reaction to be occurring, this temperature probably needs to be at least 1030 K (1400°F). The syngas is not mixing and reacting with the air introduced through the first two swirlers fast enough for a high bulk gas temperature in the primary region.

Figure 2.23 shows a velocity plot for this simulation. Note the low velocity of the entering syngas and how the flow accelerates through the flow deflector, very much as through a nozzle. (Note: With the original 2100°F combustor, the cross-sectional area of the flow deflector was approximately half the area just upstream and downstream of it, causing an excessive pressure drop and resulting in some of the syngas "bypassing" the flow deflector by reverse flow through the third-stage swirlers. Thus combustion took place in the third-stage swirler, causing thermal damage to the metal. Note also that no recirculation occurs in the fuel-rich zone, which would stabilize the flame and increase the temperature in this zone.)

Figure 2.24 illustrates the normal velocity (i.e., swirl velocity) as a contour plot. (Note: The swirl does not reach the center core of the combustor.)

kiva-ii version date: 050189 run date: 06/20/91 14:04:31
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 1600 F nrk=6 win=1524
t= 2.00051E-02 cycle 1315 crank=-180.00 deg atdc
cross-section through j= 1 (on the right)

Note: The KIVA plots in this section show low as "L" and high as "H"; arrows on the figures show where these levels occur. Values are given in the legend above the figure.

Figure 2.18 Run 1: Computational Grid

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
temp across j= 1 plane at t= 1.00000E-02 cycle 437 crank=-180.00 deg atdc
min= 6.34030E+02 max= 2.28752E+03 l= 7.99379E+02 h= 2.12217E+03 dq= 1.65349E+02

Figure 2.19 Run 1: Temperature (K) at $t = 1^{-2}s$

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
temp across j= 1 plane at t= 1.50006E-02 cycle 858 crank=-180.00 deg atdc
min= 6.33717E+02 max= 2.29075E+03 l= 7.99420E+02 h= 2.12505E+03 dq= 1.65703E+02

Figure 2.20 Run 1: Temperature (K) at $t = 1.5^{-2}s$

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
temp across j= 1 plane at t= 2.00028E-02 cycle 1271 crank=-180.00 deg atdc
min= 6.29482E+02 max= 2.22179E+03 l= 7.88712E+02 h= 2.06256E+03 dq= 1.59231E+02

Figure 2.21 Run 1: Temperature (K) at $t = 2^{-2}s$

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 800 F nrk=6 min=1524
temp across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
min= 6.30119E+02 max= 2.22126E+03 l= 7.89233E+02 h= 2.06214E+03 dq= 1.59114E+02

Figure 2.22 Run 1: Temperature (K) at $t = 2.5^{-2}s$

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 min=1524
velocity across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
um= 6.28692E+03 um= 5.67800E+03 um= 1.52567E+04
umtan= 0.62869E+04 umtan= 0.90949E-12 umtan= 0.15257E+05

Figure 2.23 Run 1: Velocity Plot (cm/s)

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 min=1524
norm vel across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
min=-8.76977E+01 max= 5.26909E+03 l= 4.47901E+02 h= 4.73341E+03 dq= 5.35670E+02

Figure 2.24 Run 1: Normal (Swirl) Velocity (cm/s)

Figures 2.25 and 2.26 depict concentrations (as mass fractions) of two fuel components (CH_4 and H_2) for Run 1. (Note: The plots show how far they penetrate into the combustor.)

Figure 2.27 shows concentration of H_2O , which is primarily a product of combustion. As could be predicted, the high concentration of H_2O corresponds closely with the peak temperature (Figure 2.22). This is also shown in Figure 2.28, which is a plot of thermal NO_x . (Note: Most of the MASB NO_x comes from the FBN; therefore, this plot can only be used to confirm the zone of highest temperature, which will correspond with the zone of highest thermal NO_x production.)

As a comparison to the 800°F case, which simulates test conditions encountered at UTSI, a simulation was run with 1600°F syngas (Run 2). This run more closely simulates conditions that would be encountered in an integrated pilot plant or commercial facility. The entering syngas velocity remained the same in these runs; therefore, the stoichiometry is not the same as in Run 1.

Figure 2.29 illustrates the axial velocity for this case, and Figure 2.30 the normal velocity. Comparison of Figures 2.24 and 2.30 shows that the swirl is transmitted into the core at about the same rate.

Figure 2.31 depicts the temperature plot for Run 2. Comparison with Figure 2.22 shows that the hot combustion zone extends much farther upstream, almost to the second swirler. The center core is also much hotter. In this case, the center core is hot enough for pyrolysis reactions and reaction of ammonia to occur (refer also to chemical kinetic modeling studies discussed later in this section).

Figures 2.32 and 2.33 show plots of the methane and hydrogen concentrations for Run 2. The hydrogen reacts much farther upstream in the combustor. In fact, most of the hydrogen reacts immediately in the premix area (i.e., with mixing air). Figure 2.34 shows H_2O concentrations for this case.

A conclusion that can be reached from Runs 1 and 2 is that an 800°F syngas does not generate a reactive reducing zone, while a 1600°F syngas should. However, in the test rig presently operating at UTSI, the maximum syngas temperature obtainable is only 800°F. This subject will be discussed later in this report.

After discussions among Dr. Beér, Westinghouse, and UTSI, several possible changes were suggested. The obvious solution—installing a syngas heater at UTSI that is capable of temperatures greater than 800°F—was not considered at this point. Basically, all of the suggestions made revolved around getting more hot, vitiated air into the fuel-rich zone of the combustor. These suggestions included:

- Adding vitiated mixing air to the nozzle

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
ch4 across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
min=-5.01713E-08 max= 6.53228E-02 l= 6.53223E-03 h= 5.87905E-02 dq= 6.53228E-03

Figure 2.25 Run 1: CH₄ Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 800 F nrk=6 win=1524
h2 across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
min= 0.00000E+00 max= 6.26303E-03 l= 6.26303E-04 h= 5.63672E-03 dq= 6.26303E-04

Figure 2.26 Run 1: H₂ Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 800 F nrk=6 win=1524
h2o across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
min= 7.43932E-04 max= 1.16125E-01 l= 1.22820E-02 h= 1.04587E-01 dq= 1.15381E-02

Figure 2.27 Run 1: H₂O Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/21/91 16:56:26
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
no across j= 1 plane at t= 2.50017E-02 cycle 1649 crank=-180.00 deg atdc
min=-6.98994E-14 max= 8.81399E-05 l= 8.81399E-06 h= 7.93259E-05 dq= 8.81399E-06

Figure 2.28 Run 1: NO_x Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/20/91 14:04:31
 ted k050189 21x1x109 fine mesh 2-d cy (x) low-btu 1600 F ntk=6 min=1524
 velocity across j= 1 plane at t= 2.00051E-02 cycle 1315 crank=-180.00 deg atdc
 um= 5.99827E+03 um= 5.67800E+03 um= 1.52187E+04
 umtan= 0.59983E+04 umtan= 0.90949E-12 umtan= 0.15219E+05

Figure 2.29 Run 2: Axial Velocity (cm/s)

kiva-ii version date: 050189 run date: 06/20/91 14:04:31
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 1600 F nrk=6 win=1524
norm vel across j= 1 plane at t= 2.00051E-02 cycle 1315 crank=-180.00 deg atdc
min=-5.79723E+01 max= 5.27921E+03 l= 4.75746E+02 h= 4.74549E+03 dq= 5.33718E+02

Figure 2.30 Run 2: Normal Velocity (cm/s)

kiva-ii version date: 050189 run date: 06/20/91 14:04:31
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 1600 F nrk=6 min=1524
temp across j= 1 plane at t= 2.00051E-02 cycle 1315 crank=-180.00 deg atdc
min= 1.04385E+03 max= 2.35791E+03 l= 1.17526E+03 h= 2.22650E+03 dq= 1.31406E+02

Figure 2.31 Run 2: Temperature (K)

kiva-ii version date: 050189 run date: 06/20/91 14:04:31
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 1600 F nrk=6 min=1524
ch4 across j= 1 plane at t= 2.00051E-02 cycle 1315 crank=-180.00 deg atdc
min=-2.47493E-08 max= 6.56348E-02 l= 6.56346E-03 h= 5.90713E-02 dq= 6.56349E-03

Figure 2.32 Run 2: CH₄ Concentration (Mass Fraction)

Figure 2.33 Run 2: H₂ Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/20/91 14:04:31
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 1600 F nrk=6 win=1524
h2o across j= 1 plane at t= 2.00051E-02 cycle 1315 crank=-180.00 deg atdc
min= 4.87254E-03 max= 1.17978E-01 l= 1.61831E-02 h= 1.06668E-01 dq= 1.13106E-02

Figure 2.34 Run 2: H₂O Concentration (Mass Fraction)

- Combining the two upstream swirlers
- Adding a flow deflector following the first or second swirler, or both
- Lengthening the fuel-rich zone before the third-stage swirler
- Changing the fuel-rich zone stoichiometry.

A number of simulation runs were made to look at the fluid dynamics of some of these alternatives—either singly or in combination.

The most change possible seemed to be to combine the airflow that was premixed with the fuel (about 10 percent of the total) with the other primary zone air entering through the first two swirlers. These would then be combined into a single swirler at the entrance to the combustor, allowing maximum residence time for the fuel and air to react. The computational mesh for this geometry, shown in Figure 2.35, was used for Runs 3 through 5.

Run 3 was made with 1600°F syngas for comparison with Run 2. Figure 2.36 shows the normal velocity for Run 3. The swirl is rapidly transmitted into the center with this arrangement. Figure 2.37 shows the temperature profile for this run. The high-temperature contour extends well back in the combustor. The central core is also quite hot.

Figure 2.38 depicts H_2 concentrations and the rapid reaction of the hydrogen.

Run 4 examined the effect of the single upstream swirler at $z = 0.0$ for the 800°F syngas. Figure 2.39 shows that the swirl is still rapidly transmitted into the central core, as with Run 3.

Figure 2.40 confirms that there still exists a cold core where the temperature is too low for any significant reaction to take place. The hot combustion zone still occurs too far downstream. Figure 2.41 shows that mixing of H_2 is occurring farther upstream with this configuration than occurred during the run illustrated in Figure 2.26, so this move is in the right direction.

Run 5 was generated to determine the effect of doubling the swirl velocity for the single swirler at $z = 0.0$ case. Figure 2.42 shows that doubling the inlet swirl effectively doubles the swirl throughout the combustor (compare to Figure 2.39).

Figure 2.43 illustrates the calculated temperature of Run 5. Apparently, combustion is taking place somewhat upstream of the position seen in Figure 2.40. Figure 2.44 shows that the H_2 does not penetrate as far into the combustor for this Run as for Run 4.

kiva-ii version date: 050189 run date: 06/23/91 10:10:45
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 min=1524
t= 2.50123E-02 cycle 1702 crank=-180.00 deg atdc
cross-section through j= 1 (on the right)

Figure 2.35 Computational Mesh

Figure 2.36 Run 3: Normal Velocity (cm/s)

kiva-ii version date: 050189 run date: 06/21/91 16:56:31
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 1600 F nrk=6 min=1524
temp across j= 1 plane at t= 2.50032E-02 cycle 1772 crank=-180.00 deg atdc
min= 1.03712E+03 max= 2.37215E+03 l= 1.17062E+03 h= 2.23865E+03 dq= 1.33503E+02

Figure 2.37 Run 3: Temperature (K)

kiva-ii version date: 050189 run date: 06/21/91 16:56:31
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 1600 F nrk=6 min=1524
h2 across j= 1 plane at t= 2.50032E-02 cycle 1772 crank=-180.00 deg atdc
min= 0.00000E+00 max= 1.50232E-03 l= 1.50232E-04 h= 1.35209E-03 dq= 1.50232E-04

Figure 2.38 Run 3: H₂ Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/23/91 10:10:45
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 800 F nrk=6 win=1524
norm vel across j= 1 plane at t= 2.50123E-02 cycle 1702 crank=-180.00 deg atdc
min=-5.65450E+00 max= 5.37239E+03 l= 5.32150E+02 h= 4.83458E+03 dq= 5.37804E+02

Figure 2.39 Run 4: Normal Velocity (cm/s)

kiva-ii version date: 050189 run date: 06/23/91 10:10:45
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
temp across j= 1 plane at t= 2.50123E-02 cycle 1702 crank=-180.00 deg atdc
min= 6.35654E+02 max= 2.23284E+03 l= 7.95373E+02 h= 2.07312E+03 dq= 1.59719E+02

Figure 2.40 Run 4: Temperature (K)

kiva-ii version date: 050189 run date: 06/23/91 10:10:45
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 min=1524
h2 across j= 1 plane at t= 2.50123E-02 cycle 1702 crank=-180.00 deg atdc
min= 0.00000E+00 max= 6.09436E-03 l= 6.09436E-04 h= 5.48492E-03 dq= 6.09436E-04

Figure 2.41 Run 4: H₂ Concentration (Mass Fraction)

kiva-ii version date: 050189 run date: 06/24/91 16:46:17
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
norm vel across j= 1 plane at t= 2.5002E-02 cycle 1757 crank=-180.00 deg atdc
min=-7.8901E+00 max= 1.0832E+04 l= 1.0761E+03 h= 9.7482E+03 dq= 1.0840E+03

Figure 2.42 Run 5: Normal Velocity (cm/s)

kiva-ii version date: 050189 run date: 06/24/91 16:46:17
ted k050189 21x1x109 fine mesh 2-d cy (W) low-btu 800 F nrk=6 win=1524
temp across j= 1 plane at t= 2.5002E-02 cycle 1757 crank=-180.00 deg atdc
min= 6.3530E+02 max= 2.2598E+03 l= 7.9775E+02 h= 2.0973E+03 dq= 1.6245E+02

Figure 2.43 Run 5: Temperature (K)

kiva-ii version date: 050189 run date: 06/24/91 16:46:17
ted k050189 21x1x109 fine mesh 2-d cy (H) low-btu 800 F nrk=6 win=1524
h2 across j= 1 plane at t= 2.5002E-02 cycle 1757 crank=-180.00 deg atdc
min= 0.0000E+00 max= 6.0963E-03 l= 6.0963E-04 h= 5.4866E-03 dq= 6.0963E-04

Figure 2.44 Run 5: H₂ Concentration (Mass Fraction)

Modeling and simulation of various configurations of the MASB is an ongoing effort at UTSI and Westinghouse. These reported results have been very useful in providing insight into changes that would be most effective in meeting program goals—changes that will be discussed later in this report.

Natural Gas Simulation Results. Westinghouse is developing the MASB as a low- NO_x burner, with emphasis on fuels that contain FBN. However, in most of the applications that have been investigated, there exists the necessity to fire the combustor with a clean fuel (typically natural gas) for start-up conditions and when the coal-derived fuel-gas generator is not operating. Therefore, UTSI and Westinghouse have attempted to obtain information on the firing of clean fuels. A test was conducted at UTSI (reported in the FY91 annual report) in which both natural gas and No. 2 fuel oil were burned in the 12-in. (2100°F) MASB. This section summarizes some results of modeling the natural gas (methane) combustion in this combustor.

Figure 2.45 shows the computational grid used for the natural gas modeling. It is just the 12-in. combustor geometry with the flow deflector removed. Run 6 was conducted at the "design" operating point of 0.1-lb/s nozzle cooling air and 0.1 lb/s of natural gas (CH_4) flowing through this center 4-in.-dia. cylinder. An assumed temperature of 200°F was used for the CH_4 and cooling air. No swirl was assumed for the fuel and nozzle cooling air.

Vitiated air at 1600°F was used as the oxidizer, corresponding with actual test conditions for high oxygen levels. A full-height first-stage swirler was used for the simulation, even though this swirler was partially blocked for the actual test. No third-stage flow deflector was used, corresponding to actual test conditions.

Figure 2.46 illustrates the initial velocity vectors for this run. The velocity of the fuel plus nozzle cooling air is very low (i.e., the arrows look like dots compared with the axial velocity arrows of the swirlers entering the combustor. For earlier syngas tests the velocity in this region was 1500 to 2000 cm/s; but for this run, the axial velocity was 170 cm/s, a factor of about 10 lower. This may be the root cause of the metal damage noted in the first-stage swirler and nozzle sections of the combustor, as reported in the FY91 report.

Figure 2.47 plots velocity vectors for the converged solution. While difficult to see on this plot, there is an annular recirculation zone just upstream of the first swirler. This recirculation zone effected the high metal temperatures observed.

Figure 2.48 shows that the swirl velocity is penetrating well into the center of the combustor. The swirl penetration upstream of the first swirler corresponds to the recirculation zone.

Figure 2.49 shows the temperature profiles for this run. The high-temperature zone extends well upstream of the first-stage swirler. Where

kiva-ii version date: 050189 run date: 08/10/91 16:56:04
ted k050189 21x1x102 fine mesh 2-d cy (W) nat gas nrk=6
t= 0.0000E+00 cycle 0 crank=-180.00 deg atdc
cross-section through j= 1 (on the right)

Figure 2.45 Run 6: Computational Grid

kiva-ii version date: 050189 run date: 08/10/91 16:56:04
ted k050189 21x1x102 fine mesh 2-d cy (W) nat gas nrk=6
velocity across j= 1 plane at t= 1.0000E-05 cycle 1 crank=-180.00 deg atdc
um= 1.6464E+03 vm= 5.6780E+03 wm= 4.4268E+03
umtan= 0.1646E+04 vmtan= 0.9095E-12 wmtan= 0.4427E+04

Figure 2.46 Run 6: Initial Velocity

kiva-ii version date: 050189 run date: 08/10/91 16:56:04
ted k050189 21x1x102 fine mesh 2-d cy (W) nat gas nrk=6
velocity across j= 1 plane at t= 4.0008E-02 cycle 4127 crank=-180.00 deg atdc
um= 8.7994E+02 vm= 5.7034E+03 wm= 3.8035E+03
umtan= 0.8799E+03 vmtan= 0.9095E-12 wmtan= 0.3803E+04

Figure 2.47 Run 6: Converged Velocity

kiva-ii version date: 050189 run date: 08/10/91 16:56:04
ted k050189 21x102 fine mesh 2-d cy (W) nat gas nrk=6
norm vel across j= 1 plane at t= 4.0000E-02 cycle 4127 crank=-180.00 deg atdc
min= 1.7294E-02 max= 5.3643E+03 l= 5.3645E+02 h= 4.8279E+03 d ζ = 5.3643E+02

Figure 2.48 Run 6: Swirl Velocity

kiva-ii version date: 050189 run date: 08/10/91 16:56:04
ted k050189 21x1x102 fine mesh 2-d cy (W) nat gas nrk=6
temp across j= 1 plane at t= 4.0000E-02 cycle 4127 crank=-180.00 deg atdc
min= 4.4331E+02 max= 2.4290E+03 l= 6.4188E+02 h= 2.2385E+03 dq= 1.9857E+02

Figure 2.49 Run 6: Temperature (K)

this high temperature contour of over 3500°F touches the metal, it is certainly hot enough to cause damage to an uncooled metal cylinder.

Figure 2.50 shows the CH₄ concentration profile. The methane is burned out in a relatively short distance. A second run (Run 7) was made with CH₄ and nozzle cooling airflow as actually run during the test (i.e., 0.14 lb/s CH₄ and 0.24 lb/s nozzle cooling air. The inlet velocity is now 280 cm/s, still very low. The results were very similar to Run 6.

One conclusion that can be reached from the model results, as well as from the post-test examination of the MASB, is that care must be used in designing an MASB to operate with both a low-Btu syngas and natural gas. The vast difference in volumetric flows causes considerable change in the fluid mechanics, leading to hot spots such as those seen in both the model and the actual experiment.

Chemical Kinetic Modeling of the MASB. The computational fluid dynamic (CFD) model of the MASB computes the flow field of the MASB when firing the low-Btu syngas required for this application. This model includes a very simplified chemical kinetics model utilizing three overall kinetic reactions for combustion of the main fuel components (CH₄, CO, and H₂) along with a simple three-equation thermal NO_x production model. Other components are considered to be at equilibrium in the CFD model.

In the design application for this MASB, the fuel will consist of a low-Btu syngas produced by a fluidized bed carbonizer (a partial gasifier). This gas will be provided to the topping combustor at 1400 to 1600°F. Vitiated air at 1400 to 1600°F is also provided to the topping combustor. A small amount of air is introduced in the burner nozzle area. Additional air is introduced at two swirlers also in the fuel-rich zone. This total "primary" air is selected so that the equilibrium flame temperature in this zone is in the range of 2700 to 3000°F, corresponding to a fuel-to-air ratio of approximately 1.6.

The CFD model, when set up for this case, shows a rapid temperature rise resulting from the nozzle injection air. Additional combustion occurs primarily at the outside of the burner downstream of the two primary air swirlers.

In the tests conducted at UTSI, the major difference is that a "synthetic" syngas is used. This fuel is produced by blending several components (currently N₂, CH₄, and H₂) and heating this gas in an indirectly fired heater to 800°F. This temperature was selected as a compromise, since heaters to produce this temperature were readily available and moderately priced. A vitiation heater provides oxidant at 1600°F.

When the CFD model is run for the UTSI conditions, there is no immediate temperature rise because of the nozzle area air addition (about 10 percent of the total air). All primary zone reactions occur at the outside of the

kiva-ii version date: 050189 run date: 08/10/91 16:56:04
ted k050189 21x1x102 fine mesh 2-d cy (W) nat gas nrk=6
ch4 across j= 1 plane at t= 4.0000E-02 cycle 4127 crank=-180.00 deg atdc
min=-3.7175E-09 max= 4.9463E-01 l= 4.9463E-02 h= 4.4517E-01 dq= 4.9463E-02

Figure 2.50 CH₄ Concentration

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2 of 2

combustor and downstream of the two swirlers (i.e., in an area where the fuel-to-air is near 1.0), resulting in a cold, cigar-shaped inner core of syngas. The temperature in this zone is too low for significant chemical reaction to take place.

These model results can be somewhat confirmed by the observation that when syngas flow is started in MASB tests, metal temperatures in the nozzle area drop, indicating that reaction is not occurring (see Figures 2.51 and 2.52 extracted from the FY91 Annual Report).

The reason that the bulk gas temperature in the primary zone is important is that the "real" syngas will contain a significant amount of NH_3 . It is important that most of this NH_3 be converted to N_2 and not to NO_x .

Although the CFD model does not include any NH_3 reactions, it can be inferred that no ammonia reaction occurs in the cold core of the syngas. Reactions would occur at the air/syngas interface downstream of the first two swirlers and in the quench and burnout zones downstream of the third swirler. This might not occur in the real burner, since the primary zone core temperature is much hotter.

These conclusions are supported by the observation that the test results to date show that the NH_3 conversion to NO_x is significantly higher than expected, as seen in the FY91 Annual Report. (At UTSI, NH_4OH solution is injected into the 800°F syngas in amounts corresponding to predicted NH_3 levels in the real system.)

In an attempt to determine the action to take, the team decided to try modeling the kinetics of the reactions taking place.

Two chemical kinetics programs have been used to model the combustor kinetics. One program is the Premixed One-Dimensional Flame (PROF) code (Kendall and Kelley, 1978) and the second is the General Chemical Kinetics Code (GCKP84) (Bittker and Scullin, 1984). An updated version of the latter code (GCKP87) was actually used for these calculations. Both of these programs can handle one-dimensional or batch processes. Calculations can be made for adiabatic conditions, or with a specified temperature profile or specified heat flux. UTSI performed calculations with the PROF code to check the calculations made at Westinghouse using GCKP87.

The kinetic reaction set consisted of 145 reactions, as shown in Table 2.2. Most of these reactions came from Crawford, et al, 1983; Crawford, 1984, and the original sources are documented in these papers. Additional reactions were provided by Dr. Crawford. All forward reactions in this table

TEST DATE: APR. 3, 1991
TIME = 13.15 - 13.20

Figure 2.51 MASB Wall Temperatures With Only Preheater Firing

TEST DATE: APR. 3, 1991
TIME = 13.70 - 13.80

Figure 2.52 MASB Wall Temperatures at 2312°F Firing Temperature

Table 2.2 Chemical Reactions Considered

REACTION						PRE EXP FACT (MOLE-CM-SEC)	TEMP EXP	ACTIVATE (CAL/MOL)
						A	N	E
M	+	CO2	=	CO	+	O	.3800E+31	-4.0000 134021.
CO	+	OH	=	CO2	+	H	.1500E+08	1.3000 -765.
CO	+	O2	=	CO2	+	O	.3200E+13	.0000 50000.
M	+	H2	=	H	+	H	.2100E+16	.0700 103830.
OH	+	O	=	HO2	+	M	.1600E+21	-1.5000 0.
H	+	O2	=	HO2	+	M	.3000E+16	.0000 -1000.
M	+	H2O	=	OH	+	H	.2300E+25	-2.0000 122600.
H	+	OH	=	H2	+	O	.2700E+18	-.9400 14690.
H	+	HO2	=	OH	+	OH	.2500E+15	.0000 1900.
OH	+	H2	=	H	+	H2O	.2500E+14	.0000 5200.
OH	+	OH	=	H2O	+	O	.6000E+13	.0000 1000.
OH	+	O	=	H	+	O2	.6300E+12	.5000 0.
HO2	+	H	=	O2	+	H2	.2500E+14	.0000 700.
HO2	+	O	=	OH	+	O2	.5000E+14	.0000 1000.
HCN	+	OH	=	CN	+	H2O	.2000E+12	.6000 5000.
CN	+	CO2	=	NCO	+	CO	.3700E+13	.0000 0.
CN	+	OH	=	NCO	+	H	.6200E+14	.0000 0.
CN	+	H2	=	HCN	+	H	.6200E+13	.0000 5300.
CN	+	O	=	CO	+	N	.6300E+12	.5000 0.
NCO	+	H	=	NH	+	CO	.5000E+12	.5000 6870.
CO	+	HNO	=	CO2	+	NH	.1000E+12	.5000 21000.
H	+	NO	=	HNO	+	M	.2000E+17	.0000 0.
NO	+	O	=	NO2	+	M	.1400E+22	-1.8200 0.
M	+	N2O	=	N2	+	O	.4000E+15	.0000 51400.
H	+	HNO	=	H2	+	NO	.1000E+14	.0000 2500.
H	+	N2O	=	OH	+	N2	.8000E+14	.0000 15000.
NH	+	NO	=	H	+	N2O	.9000E+10	.7500 0.
H	+	NO2	=	OH	+	NO	.3200E+15	.0000 1500.
HNO	+	OH	=	H2O	+	NO	.3600E+14	.0000 0.
HNO	+	O	=	OH	+	NO	.5000E+12	.5000 0.
HNO	+	O	=	NH	+	O2	.1000E+12	.5000 7000.
OH	+	N2O	=	HO2	+	N2	.3200E+14	.0000 15000.
N2	+	O	=	NO	+	N	.1840E+15	.0000 76250.
O	+	NO	=	N	+	O2	.2360E+10	1.0000 38650.
NO	+	H	=	N	+	OH	.2220E+15	.0000 50500.
NH	+	H	=	N	+	H2	.6300E+12	.5000 8000.
NH	+	OH	=	H2O	+	N	.5000E+12	.5000 2000.
NH	+	OH	=	NO	+	H2	.5000E+12	.5000 2000.
NH	+	O	=	NO	+	H	.6300E+12	.5000 0.
NH2	+	H	=	NH	+	H2	.1400E+12	.6700 4300.
NH2	+	OH	=	NH	+	H2O	.3000E+11	.6800 1300.
NH2	+	O	=	NH	+	OH	.9200E+12	.5000 0.
NH2	+	NH2	=	NH3	+	NH	.1700E+12	.6300 3600.
NH3	+	OH	=	NH2	+	H2O	.1760E+13	.0000 1700.
NH3	+	H	=	NH2	+	H2	.2460E+14	.0000 17070.
NH2	+	NO	=	N2	+	H2O	.5000E+13	.0000 0.
NH3	+	O	=	NH2	+	OH	.1500E+13	.0000 6000.
NH3	+	O2	=	NH2	+	HO2	.5000E+12	.5000 56000.
NO	+	HO2	=	NO2	+	OH	.5000E+12	.5000 3520.
NH	+	OH	=	HNO	+	H	.5000E+12	.5000 5620.
CO	+	N2O	=	CO2	+	N2	.3000E+12	.0000 17200.

Table 2.2 (Cont) Chemical Reactions Considered

REACTION						PRE EXP FACT (MOLE-CM-SEC)	TEMP EXP	ACTIVATE (CAL/MOL)
						A	N	E
M	+	HCO	=	H	+	CO	.1550E+15	.0000 14700.
HCO	+	H	=	CO	+	H2	.2000E+15	.0000 0.
HCO	+	O	=	CO	+	OH	.5500E+13	.5000 0.
HCO	+	O	=	CO2	+	H	.5500E+13	.5000 0.
HCO	+	OH	=	CO	+	H2O	.3000E+13	1.0000 0.
HCO	+	O2	=	CO	+	HO2	.5000E+12	.5000 830.
NH2	+	O	=	H	+	HNO	.2100E+13	.0000 0.
CO2	+	N	=	NO	+	CO	.2000E+12	-.5000 8000.
NO	+	NO	=	N2O	+	O	.1290E+13	.0000 63800.
HNO	+	HNO	=	N2O	+	H2O	.3000E+12	.0000 3500.
NO	+	HNO	=	N2O	+	OH	.2000E+13	.0000 26000.
NO	+	NO	=	N2	+	O2	.3100E+14	.0000 63100.
N	+	N	=	N2	+	M	.3000E+15	.0000 -900.
THIRDBODY								
H		9.0	OH		9.0	N	9.0	O 9.0
END								
N	+	O	=	NO	+	M	.6400E+17	-.5000 0.
THIRDBODY								
CO2		2.0	H		9.0	OH	9.0	H2O 6.0
N		9.0	O		9.0	END		
O	+	O	=	O2	+	M	.4000E+19	-1.0000 0.
SH	+	S	=	H	+	S2	.6310E+12	.5000 0.
SH	+	H	=	S	+	H2	.6310E+12	.5000 0.
H2S	+	H	=	SH	+	H2	.9030E+13	.0000 1700.
SH	+	SH	=	H2S	+	S	.1000E+15	.0000 1420.
OH	+	H2S	=	H2O	+	SH	.1400E+14	.0000 894.
SO	+	OH	=	H	+	SO2	.6460E+15	.0000 2100.
S	+	OH	=	SO	+	H	.6310E+12	.5000 0.
SH	+	O	=	SO	+	H	.6310E+12	.5000 0.
O	+	H2S	=	SH	+	OH	.4400E+13	.0000 3298.
SH	+	O	=	S	+	OH	.6310E+12	.5000 8000.
SO	+	O2	=	SO2	+	O	.1810E+12	.0000 5600.
S	+	O2	=	SO	+	O	.6310E+12	.5000 0.
O	+	S2	=	SO	+	S	.6310E+12	.5000 0.
SO	+	SO	=	SO2	+	S	.4100E+12	.0000 3230.
S	+	H2O	=	SH	+	OH	.4400E+13	.0000 28000.
S	+	S	=	S2	+	M	.1000E+19	-1.0000 0.
S	+	O	=	SO	+	M	.5500E+14	.0000 1788.
S	+	H	=	SH	+	M	.7200E+16	.0000 0.
M	+	H2S	=	SH	+	H	.2000E+15	.0000 74000.
SO	+	O	=	SO2	+	M	.1220E+23	-1.8400 0.
SO2	+	O	=	SO3	+	M	.4000E+21	-4.0000 5300.
SO3	+	O	=	SO2	+	O2	.2800E+15	.0000 12000.
N	+	SO	=	NO	+	S	.6310E+12	.5000 2000.
NO	+	S	=	SN	+	O	.1000E+13	.5000 34800.
SN	+	O	=	SO	+	N	.6310E+12	.5000 8000.
N	+	SN	=	N2	+	S	.6310E+12	.5000 0.
N	+	SH	=	SN	+	H	.6310E+12	.5000 8000.
SO	+	HNO	=	SO2	+	NH	.6310E+12	.5000 8000.
NO	+	SN	=	N2O	+	S	.6310E+12	.5000 8000.
N	+	SO2	=	NO	+	SO	.2000E+12	.5000 8000.

Table 2.2 (Cont) Chemical Reactions Considered

REACTION					PRE EXP FACT (MOLE-CM-SEC)	TEMP EXP	ACTIVATE (CAL/MOL)
					A	N	E
NH2	+	SH	=	H2S	+	NH	.6310E+12 .5000 0.
CS2	+	O	=	CS	+	SO	.5000E+14 .0000 1900.
CS2	+	O	=	COS	+	S	.1700E+13 .0000 1200.
CS2	+	O	=	CO	+	S2	.3200E+13 .0000 1000.
COS	+	O	=	CO	+	SO	.6000E+14 .0000 5500.
COS	+	O	=	CO2	+	S	.1200E+15 .0000 11000.
CS	+	O	=	CO	+	S	.1100E+14 .0000 41000.
CS2	+	S	=	CS	+	S2	.1500E+14 .0000 0.
COS	+	S	=	CO	+	S2	.9000E+12 .0000 2500.
COS	+	SO	=	CO2	+	S2	.1000E+09 .0000 0.
CS	+	O2	=	CO	+	SO	.5500E+08 .0000 2000.
CS	+	O2	=	COS	+	O	.3100E+10 .0000 9500.
CS	+	SO	=	CO	+	S2	.1000E+14 .0000 0.
CS	+	S	=	CS2	+	M	.5100E+25 -3.0000 0.
CO	+	S	=	COS	+	M	.2900E+26 -3.6000 0.
HCN	+	O	=	NCO	+	H	.5200E+13 .0000 8100.
NCO	+	O	=	NO	+	CO	.5620E+14 .0000 0.
NCO	+	NO	=	N2O	+	CO	.1900E+14 .0000 0.
HCN	+	O	=	CN	+	OH	.2700E+10 1.5800 26600.
M	+	CH4	=	CH3	+	H	.1000E+18 .0000 88400.
CH4	+	CH2	=	CH3	+	CH3	.1300E+13 .7000 20000.
CH4	+	H	=	CH3	+	H2	.5000E+11 1.0000 10000.
CH4	+	OH	=	CH3	+	H2O	.3000E+14 .0000 5000.
CH4	+	O	=	CH3	+	OH	.1900E+15 .0000 11700.
CH3	+	OH	=	CH2	+	H2O	.2000E+12 .7000 2000.
CH2	+	H	=	CH	+	H2	.3200E+12 .7000 5000.
CH2	+	OH	=	CH	+	H2O	.5000E+12 .5000 6000.
CH2	+	OH	=	CH3	+	O	.5000E+12 .5000 6000.
CH2	+	H2	=	CH3	+	H	.3000E+13 .0000 7000.
CH2	+	O	=	HCO	+	H	.5000E+12 .5000 4000.
CH	+	CO2	=	HCO	+	CO	.1000E+11 .5000 6000.
CH	+	OH	=	HCO	+	H	.1000E+11 .5000 10000.
CH	+	O	=	CO	+	H	.5000E+12 .5000 0.
CH	+	O2	=	HCO	+	O	.5000E+12 .5000 6000.
CH4	+	CN	=	HCN	+	CH3	.3200E+12 .7000 5000.
CH3	+	HNO	=	CH4	+	NO	.5000E+12 .5000 0.
CH2	+	N2	=	HCN	+	NH	.1000E+15 .0000 60000.
HCO	+	N	=	HCN	+	O	.1000E+15 .0000 0.
HCN	+	N	=	CH	+	N2	.2500E+12 .0000 16000.
HCN	+	O	=	CH	+	NO	.1000E+15 .0000 72000.
HCO	+	N	=	CH	+	NO	.1000E+15 .0000 48600.
CH3	+	O	=	CH2O	+	H	.2600E+15 .0000 2000.
CH2	+	OH	=	CH2O	+	H	.1000E+14 .0000 5000.
CH2	+	O2	=	CH2O	+	O	.5000E+12 .5000 7000.
CH2O	+	H	=	HCO	+	H2	.1300E+11 1.0000 3200.
CH2O	+	OH	=	HCO	+	H2O	.3200E+11 1.0000 0.
CH2O	+	O	=	HCO	+	OH	.2000E+12 1.0000 4400.
HCO	+	HO2	=	CH2O	+	O2	.1000E+15 .0000 3000.
CH2	+	NO	=	CH2O	+	N	.1600E+13 .0000 7000.

are of the Arrhenius form:

$$\text{Rate} = A T^N \exp (-E/RT)$$

where

A = Preexponential Factor (mole·cm·s)
T = Temperature (K)
N = Temperature Exponent (dimensionless)
E = Activation Energy (cal/mole)
R = Gas Constant

In the table, M represents any gaseous third-body species. All third-body efficiencies are 1.0 except for those identified with a THIRDBODY statement following the reaction, where the efficiencies for certain species are as given. This reaction set includes reaction rates for all fuel components of the proposed Wilsonville syngas (Table 2.3) except for C₂'s. For these calculations the weight fraction of C₂'s was added to the CH₄.

Two things should be noted regarding Table 2.2. The first is that reactions of NH₃ and related nitrogen compounds are included, as are reactions of H₂S and related sulfur compounds. The second is that this reaction set has been effectively used to model NO_x production from NH₃ and other compounds in the coal-fired MHD radiant furnace and heat-recovery system, as reported in Crawford.

In Table 2.3 the syngas compositions are as provided by Foster Wheeler to Westinghouse and UTSI. The 1247 Btu/lb (LHV) fuel represents the product of a 70:30 coal/sorbent water slurry feed to the carbonizer. The 1908 Btu/lb fuel is the present "design" fuel. In the kinetic model calculations, this fuel was used without modification, other than with C₂'s as described.

In Table 2.3 the 1600°F vitiated air composition requires some explanation. It has been assumed that air with a 65°F dew point was heated to 1600°F by the combustion of methane. The composition given is for this vitiated air at equilibrium conditions, the reason for amounts of OH, HO₂, NO, NO₂, and N₂O in the vitiated air. The OH and HO₂, in particular, were thought to help "trigger" the combustion reactions. The level of NO, however, is higher than really expected. Some of these species would be present in the vitiated air both at UTSI and at Wilsonville.

The kinetic calculations that have been performed are basically for a premixed, adiabatic batch reaction. The vitiated air and syngas are assumed to be completely premixed at time = 0.0. The initial mixture temperature is based on the weighted average of 1600°F vitiated air and syngas at the temperature that is the main variable in these runs. The fuel-to-air ratio has been adjusted (by trial and error) to yield equilibrium temperatures between about 2700 and 3000°F (1760 and 1920 K).

Table 2.3 Wilsonville Syngas Composition

Element	1247 Btu/lb Fuel		1908 Btu/lb Fuel		1600°F Vitiated Air	
	Wt%	Mole Fraction	Wt%	Mole Fraction	Wt%	Mole Fraction
CO ₂	26.087	0.15399	20.424	0.12535	4.9736	0.03190
CO	4.210	0.03905	14.245	0.13737		
OH					3.84E-5	6.381E-7
O ₂					15.256	0.13458
H ₂	0.412	0.05083	0.412	0.05521		
HO ₂					2.87E-6	2.456E-8
H ₂ O	16.637	0.23991	3.792	0.5685	5.3157	0.08329
NO					1.148E-2	1.080E-4
NO ₂					9.24E-4	5.979E-6
N ₂ O					3.24E-6	2.075E-8
N ₂	48.473	0.44954	55.968	0.53965	74.442	0.75011
NH ₃	0.153	0.00233	0.227	0.00360		
H ₂ S	0.101	0.00077	0.119	0.00094		
CH ₄ *	3.927	0.06359	4.813	0.08104		
*Includes weight % C ₂ 's.						

Figures 2.53 through 2.55 show plots of some key components of the combustion products as a function of time. (Detailed compositions are given in Tables 2.4 through 2.6.) Figure 2.53 shows results for reaction of syngas and vitiated air (both at 1600°F). The temperature rises to near equilibrium conditions (1930 K or 3020°F) within the first 10 ms. Note that this equilibrium temperature is higher than desired in the fuel-rich zone of the MASB; however, the conclusions to be drawn will remain the same.

Figure 2.54 shows results for 800°F syngas mixed with 1600°F vitiated air. This mixture does not react significantly during the 90-ms duration of the run.

Figure 2.55 illustrates results for 1200°F syngas mixed with 1600°F air. The sharp temperature rise, indicating the zone of combustion, occurs at about 50 ms, thus indicating that an ignition delay may occur.

Figure 2.53 1600°F Syngas/1600°F Air

Figure 2.54 800°F Syngas/1600°F Air

Figure 2.55 1200°F Syngas/1600°F Air

Table 2.4 1908 Btu/lb Syngas—1600°F Oxidizer—1600°F Syngas

INITIAL CONCENTRATION TIME(S) TEMP.(K)	0.0 1144.	1.E-2 1931.	2.E-2 1931.	5.E-2 1932.	9.E-2 1933.	Equilibrium 1934.
SPECIES	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION
CO ₂	7.25E-02	1.10-01	1.10E-01	1.10E-01	1.10E-01	1.10E-01
CO	5.97E-02	5.79E-02	5.79E-02	5.78E-02	5.78E-02	5.78E-02
O	0.00E+00	5.24E-08	5.30E-08	5.39E-08	5.46E-08	5.54E-08
OH	3.60E-07	3.12E-05	3.14E-05	3.17E-05	3.19E-05	3.22E-05
H	0.00E+00	4.27E-05	4.29E-05	4.31E-05	4.33E-05	4.35E-05
O ₂	7.60E-02	1.87E-07	1.89E-07	1.92E-07	1.95E-07	1.98E-07
H ₂	2.40E-02	1.85E-02	1.84E-02	1.84E-02	1.84E-02	1.83E-02
HO ₂	1.39E-08	2.67E-10	2.67E-10	2.70E-10	2.72E-10	2.78E-10
H ₂ O	7.18E-02	1.51E-01	1.51E-01	1.51E-01	1.51E-01	1.51E-01
HCN	0.00E+00	7.53E-07	1.39E-07	1.29E-08	1.40E-08	1.34E-08
CN	0.00E+00	4.74E-11	8.91E-12	7.92E-13	9.14E-13	2.78E-10
NCO	0.00E+00	3.24E-10	6.25E-11	5.06E-12	6.93E-12	2.78E-10
N	0.00E+00	8.47E-10	8.15E-10	7.23E-10	5.90E-10	2.78E-10
NH	0.00E+00	2.36E-09	1.86E-09	1.22E-09	7.95E-10	2.78E-10
HNO	0.00E+00	6.00E-08	5.01E-08	3.04E-08	1.69E-08	2.78E-10
NO	6.10E-05	3.50E-04	2.91E-04	1.77E-04	9.90E-05	6.00E-06
NO ₂	3.38E-06	1.90E-09	1.58E-09	9.75E-10	5.47E-10	2.78E-10
N ₂ O	1.17E-08	6.58E-09	3.55E-09	2.03E-09	1.37E-09	2.78E-10
N ₂	6.58E-01	6.61E-01	6.61E-01	6.61E-01	6.61E-01	6.61E-01
NH ₂	0.00E+00	3.49E-08	2.06E-08	1.81E-08	1.68E-08	2.78E-10
NH ₃	1.57E-03	5.94E-06	3.15E-06	2.73E-06	2.53E-06	4.72E-07
HCO	0.00E+00	1.01E-08	1.01E-08	1.01E-08	1.01E-08	1.01E-08
SH	0.00E+00	5.27E-06	5.28E-06	5.25E-06	5.23E-06	5.20E-06
S	0.00E+00	3.96E-07	3.99E-07	3.98E-07	3.98E-07	3.98E-07
S ₂	0.00E+00	4.03E-07	4.04E-07	3.98E-07	3.94E-07	3.90E-07
H ₂ S	4.10E-04	2.68E-05	2.68E-05	2.65E-05	2.62E-05	2.60E-05
SO	0.00E+00	2.00E-05	2.00E-05	2.00E-05	2.00E-05	2.00E-05
SO ₂	0.00E+00	3.54E-04	3.54E-05	3.55E-04	3.55E-04	3.56E-04
SO ₃	0.00E+00	1.11E-07	7.48E-08	2.40E-08	7.01E-09	2.78E-10
SN	0.00E+00	1.90E-09	1.46E-09	1.39E-09	1.60E-09	1.01E-08
CS ₂	0.00E+00	1.61E-11	1.94E-11	1.90E-11	1.87E-11	2.78E-10
CS	0.00E+00	1.53E-10	1.54E-10	1.53E-10	1.52E-10	2.78E-10
COS	0.00E+00	2.99E-06	3.02E-06	2.98E-06	2.96E-06	2.93E-06
CH ₄	3.52E-02	1.01E-06	7.61E-07	3.17E-07	9.56E-08	2.78E-10
CH ₃	0.00E+00	5.38E-08	4.05E-08	1.70E-08	5.16E-09	2.78E-10
CH ₂	0.00E+00	1.90E-10	1.44E-10	6.10E-11	1.86E-11	2.78E-10
CH	0.00E+00	1.28E-12	9.80E-13	4.19E-13	1.29E-13	2.78E-10
CH ₂ O	0.00E+00	7.34E-09	7.28E-09	7.17E-09	7.10E-09	2.78E-10

Table 2.5 1908 Btu/lb Syngas—1600°F Oxidizer—800°F Syngas

INITIAL CONCENTRATION TIME(S) TEMP.(K)	0.0 951.	1.E-2 952.	3.E-2 952.	6.E-2 952.	9.E-2 952.	Equilibrium 1761.
SPECIES	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION	MOLE FRACTION
CO ₂	7.25E-02	7.25E-02	7.25E-02	7.25E-02	7.25E-02	1.12E-01
CO	5.97E-02	5.97E-02	5.97E-02	5.97E-02	5.97E-02	5.56E-02
O	0.00E+00	2.83E-12	4.13E-12	4.38E-12	4.49E-12	2.78E-10
OH	3.60E-07	6.52E-14	2.51E-12	2.67E-12	2.79E-12	5.18E-06
H	0.00E+00	1.44E-12	8.90E-12	9.38E-12	9.80E-12	1.14E-05
O ₂	7.60E-02	7.59E-02	7.59E-02	7.59E-02	7.59E-02	2.78E-10
H ₂	2.40E-02	2.39E-02	2.39E-02	2.39E-02	2.39E-02	2.02E-02
HO ₂	1.39E-08	1.19E-05	1.26E-05	1.37E-05	1.49E-05	2.78E-10
H ₂ O	7.18E-02	7.18E-02	7.18E-02	7.18E-02	7.18E-02	1.49E-01
HCN	0.00E+00	1.89E-16	2.87E-16	4.38E-16	6.18E-16	1.54E-08
CN	0.00E+00	8.05E-23	6.53E-23	6.88E-23	7.43E-23	2.78E-10
NCO	0.00E+00	7.04E-22	5.71E-22	6.01E-22	6.49E-22	2.78E-10
N	0.00E+00	4.64E-20	2.62E-20	1.89E-20	1.83E-20	2.78E-10
NH	0.00E+00	4.08E-15	-2.04E-16	-1.35E-17	3.93E-19	2.78E-10
HNO	0.00E+00	-2.61E-10	-3.53E-11	-2.20E-12	-1.20E-13	2.78E-10
NO	6.10E-05	-2.33E-09	2.67E-09	2.65E-09	2.56E-09	6.63E-07
NO ₂	3.38E-06	6.39E-05	6.39E-05	6.39E-05	6.39E-05	2.78E-10
N ₂ O	1.17E-08	3.01E-09	2.59E-09	2.72E-09	2.93E-09	2.78E-10
N ₂	6.58E-01	6.58E-01	6.58E-01	6.58E-01	6.58E-01	6.61E-01
NH ₂	0.00E+00	-3.44E-11	2.09E-11	2.25E-11	2.33E-11	2.78E-10
NH ₃	1.57E-03	1.56E-03	1.56E-03	1.56E-03	1.56E-03	7.67E-07
HCO	0.00E+00	6.55E-14	8.33E-14	8.88E-14	9.46E-14	2.78E-10
SH	0.00E+00	-1.41E-09	1.13E-09	1.07E-09	1.10E-09	7.70E-06
S	0.00E+00	2.78E-16	-1.16E-15	-8.72E-17	5.75E-17	2.32E-07
S ₂	0.00E+00	5.08E-12	5.08E-12	5.08E-12	5.07E-12	1.91E-06
H ₂ S	4.10E-04	4.08E-04	4.08E-04	4.08E-04	4.08E-04	1.10E-04
SO	0.00E+00	1.27E-12	-2.42E-12	-1.81E-13	1.19E-13	1.23E-05
SO ₂	0.00E+00	1.38E-06	1.38E-06	1.38E-06	1.38E-06	2.65E-04
SO ₃	0.00E+00	5.94E-12	2.05E-11	4.24E-11	6.44E-11	2.78E-10
SN	0.00E+00	4.02E-13	3.97E-13	3.93E-13	3.92E-13	2.78E-10
CS ₂	0.00E+00	5.86E-26	2.36E-25	5.04E-25	7.70E-25	2.78E-10
CS	0.00E+00	4.51E-21	4.58E-21	4.59E-21	4.59E-21	2.78E-10
COS	0.00E+00	3.92E-09	3.91E-09	3.91E-09	3.91E-09	1.05E-05
CH ₄	3.52E-02	3.52E-02	3.52E-02	3.52E-02	3.52E-02	2.78E-10
CH ₃	0.00E+00	-1.89E-10	1.04E-09	1.10E-09	1.15E-09	2.78E-10
CH ₂	0.00E+00	-9.28E-15	-3.56E-17	7.14E-19	8.61E-19	2.78E-10
CH	0.00E+00	-9.85E-21	2.11E-22	4.01E-26	-6.20E-27	2.78E-10
CH ₂ O	0.00E+00	3.95E-08	3.21E-08	2.35E-08	1.84E-08	2.78E-10

Table 2.6 1908 Btu/lb Syngas—1600°F Oxidizer—1200°F Syngas

INITIAL CONCENTRATION TIME(S) TEMP.(K)	0.0 1048.	1.E-2 1050.	2.E-2 1051.	3.E-2 1052.	4.E-2 1119.	5.E-2 1287.	6.E-2 1853.	9.E-2 1852.	Equilibrium 1847.
CO ₂	7.25E-02	7.25E-02	7.25E-02	7.26E-02	7.37E-02	7.92E-02	1.13E-01	1.12E-01	1.11E-01
CO	5.97E-02	5.97E-02	5.97E-02	5.97E-02	5.92E-02	5.74E-02	5.43E-02	5.48E-02	5.68E-02
O	0.00E+00	3.35E-10	5.81E-10	2.08E-09	7.53E-08	6.20E-08	1.32E-08	1.31E-08	1.19E-08
OH	3.60E-07	4.32E-10	8.45E-10	3.56E-09	1.65E-07	7.02E-07	1.43E-05	1.43E-05	1.34E-05
H	0.00E+00	1.18E-09	2.09E-09	7.45E-09	2.47E-07	1.37E-07	2.26E-05	2.29E-05	2.29E-05
O ₂	7.60E-02	7.58E-02	7.57E-02	7.55E-02	7.08E-02	5.74E-02	5.08E-08	4.81E-08	4.09E-08
H ₂	2.40E-02	2.37E-02	2.36E-02	2.33E-02	1.47E-02	2.89E-04	1.79E-02	1.82E-02	1.92E-02
HO ₂	1.39E-08	7.28E-05	1.01E-04	1.56E-04	2.11E-04	2.13E-04	7.17E-11	6.81E-11	2.78E-10
H ₂ O	7.18E-02	7.20E-02	7.21E-02	7.24E-02	8.17E-02	1.02E-01	1.49E-01	1.49E-01	1.50E-01
HCN	0.00E+00	1.77E-14	4.64E-14	8.95E-14	5.20E-13	1.10E-10	2.16E-06	2.27E-06	1.43E-08
CN	0.00E+00	3.02E-20	4.34E-20	7.12E-20	2.23E-18	3.84E-16	6.28E-11	6.56E-11	2.78E-10
NCO	0.00E+00	3.28E-19	4.73E-19	7.77E-19	2.87E-17	5.51E-15	5.15E-10	5.29E-10	2.78E-10
N	0.00E+00	1.70E-18	1.67E-18	4.52E-18	1.24E-14	2.05E-13	1.89E-09	6.40E-10	2.78E-10
NH	0.00E+00	6.61E-15	2.02E-14	1.96E-13	7.50E-11	3.45E-10	1.61E-08	3.81E-09	2.78E-10
HNO	0.00E+00	4.21E-11	1.38E-10	7.69E-10	1.24E-08	5.08E-09	5.41E-09	3.93E-09	2.78E-10
NO	6.10E-05	5.46E-08	7.26E-08	1.58E-07	3.44E-06	2.15E-06	2.73E-05	2.23E-05	2.09E-06
NO ₂	3.38E-06	6.39E-05	6.39E-05	6.38E-05	6.15E-05	7.52E-05	9.08E-11	7.27E-11	2.78E-10
N ₂ O	1.17E-08	5.21E-08	7.33E-08	1.11E-07	3.2E-07	1.72E-06	3.04E-09	1.24E-09	2.78E-10
N ₂	6.58E-01	6.58E-01	6.58E-01	6.58E-01	6.62E-01	6.67E-01	6.63E-01	6.62E-01	6.61E-01
NH ₂	0.00E+00	1.65E-09	2.94E-09	1.06E-08	3.69E-07	1.08E-06	1.65E-06	3.64E-07	2.78E-10
NH ₃	1.57E-03	1.56E-03	1.57E-03	1.57E-03	1.57E-03	1.55E-03	4.97E-04	1.09E-04	5.96E-07
HCO	0.00E+00	3.96E-12	6.12E-12	1.59E-11	7.08E-10	4.77E-09	6.12E-09	6.28E-09	2.78E-10
SH	0.00E+00	6.31E-08	1.08E-07	3.09E-07	2.31E-06	1.30E-06	3.81E-06	5.94E-06	6.61E-06
S	0.00E+00	1.31E-13	3.82E-13	3.12E-12	1.97E-10	7.94E-11	1.98E-07	3.03E-07	3.24E-07
S ₂	0.00E+00	2.96E-11	2.97E-11	3.44E-11	2.67E-09	3.56E-09	3.10E-07	7.70E-07	9.24E-07
H ₂ S	4.10E-04	4.06E-04	4.06E-04	4.04E-04	2.66E-04	3.42E-05	3.02E-05	4.72E-05	5.47E-05
SO	0.00E+00	2.17E-10	6.33E-10	5.13E-09	2.78E-07	9.01E-08	1.73E-05	1.64E-05	1.67E-05
SO ₂	0.00E+00	3.12E-06	3.54E-06	5.51E-06	1.43E-04	3.79E-04	3.56E-04	3.35E-04	3.25E-04
SO ₃	0.00E+00	1.51E-10	3.30E-10	5.64E-10	5.89E-09	3.82E-07	6.49E-07	4.90E-07	2.78E-10
SN	0.00E+00	3.00E-12	5.23E-12	2.15E-11	2.00E-09	9.47E-09	4.69E-09	3.66E-09	2.78E-10
CS ₂	0.00E+00	3.06E-23	7.28E-23	1.17E-22	5.89E-20	1.48E-17	3.65E-12	5.20E-11	2.78E-10
CS	0.00E+00	4.61E-19	4.73E-19	5.33E-19	8.63E-17	7.21E-15	6.20E-11	1.65E-10	2.78E-10
COS	0.00E+00	5.36E-09	6.06E-09	9.40E-09	1.96E-07	2.86E-07	2.61E-06	4.88E-06	5.70E-06
CH ₄	3.52E-02	3.52E-02	3.52E-02	3.52E-02	3.52E-02	3.27E-02	8.20E-04	7.27E-04	2.78E-10
CH ₃	0.00E+00	1.27E-07	2.40E-07	9.55E-07	4.82E-05	3.72E-04	2.39E-05	2.11E-05	2.78E-10
CH ₂	0.00E+00	1.18E-14	4.23E-14	7.04E-13	1.53E-09	4.46E-08	4.42E-08	3.89E-08	2.78E-10
CH	0.00E+00	-1.10E-19	5.88E-21	2.59E-19	2.16E-14	9.34E-13	1.69E-10	1.48E-10	2.78E-10
CH ₂ O	0.00E+00	1.76E-07	8.54E-08	1.09E-07	3.24E-05	1.02E-04	5.27E-08	4.72E-08	2.78E-10

More recent calculations with kinetic programs at UTSI, W-PGBU (Orlando), and W-STC (Pittsburgh) show that if the equilibrium temperature is carefully controlled to yield an equilibrium temperature of 2850°F (1840 K), NO_x levels are reduced, although some NH₃ remains in the fuel-rich zone after 20 to 30 ms.

These chemical kinetic calculations indicate that it is necessary to maintain a hot, fuel-rich zone in the combustor, to react the NH₃ to N₂. Along with the CFD modeling and data analysis, they lead to the redesign of the fuel-rich zone of the MASB described in the next section.

Conclusions and Recommendations. A significant amount of time in FY92 was spent analyzing the MASB. The data from all tests were reviewed in light of both the CFD modeling results and the chemical kinetic studies. The results support the conclusions and recommendations that follow.

Figures 2.56 and 2.57 show preburner and MASB temperature profiles measured during the test of the 2350°F combustor on April 3, 1991. These figures show that it is possible to get a uniform temperature pattern from the preburner, and when this occurs, the pattern from the MASB is also very good. However, Figures 2.58 and 2.59 taken from the test on July 19, 1991, show that the preburner temperature profile had deteriorated significantly and may have contributed to the relatively nonuniform temperatures exiting the MASB. Therefore, one recommendation is some modification and experimentation with the preburner to correct this nonuniform distribution.

When the conditions required for the Wilsonville plant are considered, we find that there are actually two different conditions required from the preburner. The first case is for 1600°F preheated air, as has been used in all tests to date at UTSI. At Wilsonville, this simulates the case when the PFBC is operating and generating 1600°F vitiated air. The second case is to simulate the condition when the PFBC is off-line, such as during start-up. In this case the compressor air will be at about 700°F and contain the full 23.2 wt% oxygen present in air.

For testing at UTSI, these preburners represent two distinct cases and designs. The 1600°F preburner represents further modification of the present UTSI design. With regards to the 700°F burner entering temperature (BET) condition, the approach that will be taken is to fire the preburner with sufficient air to achieve an acceptable flame temperature. Then this vitiated air will be diluted with bypass air to achieve 700°F air at the burner exit and entrance to the MASB. This air will contain the full 23.2 wt% oxygen, since the oxygen level can be independently varied during tests at UTSI by varying the amount of cryogenic oxygen added. Figures 2.60 and 2.61 are sketches of the two different preburners required; they are being fabricated for use during the 18-in. Wilsonville MASB tests.

Because the preburner can have a significant effect on the performance of the MASB, the two preburners will be tested to ascertain their performance

Figure 2.56 Preburner Exhaust Temperature Pattern, 4/3/91 Test

Figure 2.57 MASB Exhaust Temperature Pattern, 4/3/91 Test

Figure 2.58 Preburner Exhaust Temperature Pattern During 7/19/91 Test

Figure 2.59 MASB Exhaust Temperature Pattern During 7/19/91 Test

Figure 2.60 1600°F Preburner for Multiannular Swirl Burner Test at UTSI, Tullahoma, Tennessee 5/5/92

Figure 2.61 700°F Preburner for Multiannular Swirl Burner Test at UTSI, Tullahoma, Tennessee 5/4/92

characteristics before coupling them to the MASB. At the end of FY92, pre-burner construction and installation were nearing completion. Preburner tests will be conducted during the first quarter of FY93. A partially open plate will be installed into the test rig to simulate the MASB pressure drop.

A second area of concern that has been discussed during the modeling and kinetic studies is that the 800°F syngas available at UTSI is not hot enough to obtain reaction of the NH_3 . Figures 2.51 and 2.52 (shown earlier) confirm that this effect is real and measurable. Figure 2.51 depicts MASB wall temperatures with the preburner firing only. Note that the upstream metal temperatures are in the range of 1500°F, as is expected with 1600°F preheater firing only. However, when syngas is added to achieve a combustor exit temperature of 2312°F, the metal temperature of the upstream metal actually drops by about 200°F, indicating that combustion is not taking place in the upstream end of the MASB. (If it were, a temperature rise would be seen.)

Concern about the low core temperature achieved at UTSI has led the Westinghouse project team to do several things to increase the temperature in the fuel-rich zone. The first is to recommend that a 1200°F syngas heater be purchased and installed at UTSI. The 1200°F temperature is a compromise that can be commercially obtained at a reasonable price and guaranteed to operate as desired. If a little bit higher temperature is required to kick off reactions, it can be provided by an ignitor or pilot light. However, the recirculation that is being added (discussed in the following paragraphs) will probably alleviate the requirement for a constant ignitor.

The modeling and kinetic results, along with the requirements of the MASB's inventor Dr. Beér—require that the fuel-rich zone be hot—but not too hot. Ideal conditions are about 2850°F, with a fuel-to-air ratio of about 1.6. The Westinghouse project team is currently redesigning the upstream portion of the MASB. Two combustors are being redesigned. One is the 14-in.-dia. combustor that was tested in 1991, and the second is the 18-in.-dia. combustor designed to simulate Wilsonville operation.

The combustor designs are still not complete; however, their features include:

- Elimination of the first two stages of swirlers. Their airflow will be combined with the 10-percent mixing air, and all will be added at the upstream end of the combustor with swirl.
- Addition of the syngas (or natural gas) to the swirling primary air at the combustor inlet, allowing maximum residence time and simulating a premixed burner, although with swirl.
- Use of a "blunt body" at the center of the burner and an expanded section immediately following the inlet of the fuel and air. These modifications should ensure a recirculation zone that stabilizes the flame and temperature and ensures ignition.

- Addition of a device to ratio the amount of primary to secondary air. The primary air will all enter at the upstream end of the combustor, and the secondary air will be added in three stages of swirlers.

Westinghouse has been authorized to design, fabricate and test a 14-in. MASB, incorporating these features; it is to be tested at UTSI during the second quarter of FY93. The 1200°F syngas heater may not be available for this test. Therefore, we plan to include either a constant ignitor or a pilot light to kick off combustion. However, the recirculation pattern should maintain ignition once the combustor is operated near design conditions.

Westinghouse has also been authorized to design, build, and test an 18-in. combustor at UTSI. This combustor will be "internally" similar to the burner to be designed for Wilsonville. There may be minor external differences, simply because of the difference in mounting hardware required. The Wilsonville combustor will be fabricated and delivered to Wilsonville based on the 14-in. MASB test data. Although differences in performance between the 14- and 18-in. MASBs are expected to be minimal, should they prove significant, knowledge gained from the 18-in. MASB tests will be available in time to be incorporated in the spare 18-in. MASB being constructed for Wilsonville.

2.3.5 CPFBC Pilot Plant

With the exception of the FBHE-to-CPFBC secondary air and the FBHE cooling water piping, both of which were completed in August and September 1992, the design of the CPFBC pilot plant shown in Figure 2.62 was essentially completed in FY91. With regard to specialty gas measurements, the Westinghouse alkali gas sampling probe will be supplemented by measurements taken with the METC-developed fiber optic alkali meter (FOAM) system. Successful operation of the FOAM requires that the gas sample be kept at 1600°F during its passage to the nearby FOAM system to prevent alkali condensation. Valves are required in the sampling system for shutoff, control, and the bypassing of a particle-removing filter (alkali measurements are to be made with and without the presence of particulate).

Because of the high pressure and high temperature involved (270 psig at 1650°F) an Incoloy 800H, 2300-lb valve rating is required. These valves are massive and expensive; they are long-lead delivery items that will be difficult to heat/maintain at 1600°F. Consequently, FWDC is investigating the feasibility of fabricating the sampling system from Type 316SS and placing all pressure parts inside a suitably designed pressure vessel. As shown in Figure 2.63, electric clamshell heaters will heat the sampling system to 1600°F, and internal insulation will isolate the pressure vessel from high temperature. At 1600°F, galling of valve parts is a potential concern. A valve manufacturer offering Stellite-coated, nonrotating stems suitable for on-off and metering service ($C_v = 0.08$) was located. Two valves (one of each type), costing approximately \$200 each, were purchased for modification and baking in a 1600°F oven to determine whether these relatively inexpensive valves can be made to work.

Figure 2.62 CPFBC Pilot Plant

Figure 2.63 FOAM Sampling System Arrangement Proposed for CPFBC Testing

A photograph of the as-received valve proposed for the FOAM system is shown in Figure 2.64. Figure 2.65 shows how the valve was modified to place the stem thread and seals outside the 1600°F system environment. All of the stem packing seal parts were removed and replaced with new stem and stem extenders and a seal housing. The stem alignment feature of the original packing design was duplicated in the end of the seal extender that is welded into the valve body. The Stellite valve stem is welded into a stem extender which mates with the seal housing. Viton O-rings seal the housing to the extenders. The threaded connections between the stem extender and the housing are for flow control. Double O-rings are used between both the stem extender and seal extender and the housing. Nitrogen at a pressure slightly higher than system pressure is supplied between the seals so that any leak, should it

Figure 2.64 As-Received FOAM System Valve

Figure 2.65 Modified Valve Assembly (RD-923-68)

occur, will be nitrogen overboard or into the process—a hot gas leak cannot occur. The valve housing was modified by welding shut the six vent holes in the body. The inlet and outlet ports were welded shut, and new ones were drilled for the 1/4-in. tubing, as shown in Figure 2.65. Figure 2.66 is a photograph of the modified valve assembly.

The valve was tested at 1600°F to determine whether tight shutoff and adequate flow control could be achieved. Because the housing seals would not be under pressure during the test, they were tested before and after the test with 210-psi nitrogen applied to the seal pressure port. None of the seals leaked. The hot test setup is shown in Figure 2.67. After the valve reached the test temperature (1600°F), leak and flow tests were performed. Six tests were performed over a 3-hour period (valve temperature between 1600 and 1638°F). Tight shutoff was achieved in all tests with a very small handle torque. Flow control over the full 2.5 ft³/h range was easily accomplished. The calculated test rate of 0.37 sft³/h could be maintained. At the end of the test, the seal housing temperature had stabilized at 170°F—well below the 400°F allowable for Viton O-rings. The valve handle was warm to the touch, but could be held firmly without discomfort. Disassembly of the valve showed the seal and stem to be in excellent condition, with no signs of galling.

Having demonstrated that the modified valves will not leak or gall, the final design and procurement of materials required for the FOAM sampling systems was completed.

Figure 2.66 Modified Valve

Figure 2.67 Hot-Test Setup

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Section 3

TASK 3—FABRICATION/MODIFICATION AND SHAKEDOWN

Section 3

TASK 3—FABRICATION/MODIFICATION AND SHAKEDOWN

3.1 OBJECTIVE

The Task 3 objective is to fabricate, erect, and commission test facilities for the carbonizer/CPFBC, ceramic cross-flow filter, and topping combustor.

3.2 CARBONIZER PILOT PLANT

Excavation for the pilot plant foundations began in April 1990; by September 30, the tower steel had been erected and all pressure vessels were on the site. By September 30, 1991, pilot plant construction and checkout had been completed, and two short-duration shakedown runs had been successfully completed. The pilot plant was publicly dedicated on October 23, and the first test run was conducted the week of November 18.

With the exception of the ceramic cross-flow filter, the construction and checkout of the carbonizer pilot plant was essentially completed in FY91. Upon completion of the parametric portion of the carbonizer test program, the assembly of the ceramic cross-flow was completed, and the unit was connected to the pilot plant (replacing the second stage cyclone) for testing in May. To accomplish this, the fines dip-leg cooler and lock hopper, which were previously positioned under the cyclone, were moved under and connected to the cross-flow filter vessel. The pilot plant was operated from November 1991 through June 1992 (as originally planned) in the jetting fluidized bed configuration (shown in Figure 3.1). On June 1, METC released FWDC to interrupt the conversion of the plant to the CPFBC configuration to permit one test run of an 8-in.-I.D. circulating bed carbonizer. The primary cyclone dip-leg cooler and drain lock hopper were removed. The 12-in. nozzle plug was removed from the side of the carbonizer, and the double corrugated expansion joint shown in Figure 3.2 was temporarily installed. The vertical pipe run and cross forming the cyclone fines recirculation L-valve were moved into position, cut to size, and their slip-on flanges were tack welded in place. While the L-valve piping was being welded and pressure tested, the carbonizer bed drain screw and lock hopper were removed, and the carbonizer was lowered to the floor section by section. The ceramic inserts that convert the 10-, 12-, and 14-in.-I.D. sections of the unit to a constant 8-in. dia. were installed and the unit, together with its bed drain screw and lock hopper, was reassembled.

The carbonizer bubbling bed primary cyclone was removed, the larger-capacity CPFBC primary cyclone was installed, and the internal piping was appropriately modified.

Figure 3.1 Carbonizer Pilot Plant Schematic (Rev. A)

Figure 3.2 Circulating Bed Carbonizer—General Arrangement

The natural-gas-fired stack incinerator, which has been a constant NO_x-CO emissions concern to us, was removed and reassembled with the new 10-ft-long spool piece shown in Figure 3.3. The incinerator has been lengthened to allow us to operate it with the staged combustion needed to reduce its NO_x emissions. Since the discharge elevation of the stack cannot be changed, the incinerator was lengthened downward; and all supports, piping, and instrumentation were modified appropriately. Automatic valving and metering were installed; they will permit regulation of incinerator primary and secondary air split from the control room.

Testing with the ceramic cross-flow filter identified the need to modify the element mounting arrangement. Rather than delay the circulating bed carbonizer test to accommodate this change, the cross-flow elements were replaced with the backup candle assembly.

The installation of the 22 candle filter elements was completed, the unit was reassembled, and all instrumentation was installed. A 1-1/2 in. bypass line and valve were provided around the 8-in. pressure-test valve at the demister outlet. With the 8-in. valve closed, the 1-1/2 in. bypass valve will be throttled to control the velocity in the carbonizer during bed heat-up. The plant, modified as shown in Figure 3.4, was pressure tested; and on August 19, hot air from the electric air heater was used to dry out/cure the standpipe, L-valve, and new refractory linings in the carbonizer. After cool-down these components were inspected and found satisfactory in appearance; the plant was then made ready for circulating bed operation.

3.3 FILTERS/PROBES

3.3.1 Ceramic Cross-Flow Filter

Early in 1991, 12 of the 30 cross-flow filter elements needed for the Phase 2 program were purchased from Coors and qualified (by visual inspection and permeability tests) for the test program. These filter elements were fabricated by Coors using previously established techniques and specifications. The balance was fabricated by Coors, using improved specifications and fabrication techniques. The new cross-flow filter manufacturing technique applied by Coors incorporated improved quality control. Recent NDE of paired filter plates has demonstrated the importance of improved quality control with respect to plate uniformity, control of loads used during firing, and improved firing temperature control. The geometry and dimensions of the new elements are identical to the past generation of elements. The 20 new cross-flow filter elements have been received; and we have visually inspected them, checked critical dimensions, and completed permeability measurements. Although there is ample reason to expect that these new elements will be superior in their durability, some characterizations and process testing will be performed before W-STC ships them to FWDC. They will not be qualified by W-STC during in-house filter element testing until 1993.

Figure 3.3 Incinerator With 10-ft-long Spool Piece Extension

Figure 3.4 Circulating Bed Carbonizer Pilot Plant

The new pulse gas eductors were fabricated and installed. Installation of the cross-flow filter elements in the plenums and of all the filter internals was completed in May 1992. The carbonizer heat-up curves measured in the initial carbonizer testing were evaluated for their impact on the cross-flow filter. Although the carbonizer is heated at the rate of 200°F/h, upon ignition at 900°F, temperatures rise rapidly and heating rates of 400 to 600°F/h can occur at the filter location. These increases are no greater than have been experienced in W-STC simulator testing or than are expected during the upcoming testing at the AEP Tidd PFBC plant. Thus no concern for the heat-up of the carbonizer filter has been identified. The system leak test was completed, and the pulse cleaning system functions were confirmed operable during cold shakedown testing.

The objectives of the carbonizer cross-flow filter testing were reviewed:

- Demonstrate short-term durability and operability of the cross-flow filter components over a broad range of carbonizer conditions.
- Demonstrate short-term ability to effectively clean the cross-flow filter elements over a broad range of carbonizer conditions.
- Characterize, for each carbonizer condition, the representative filter vessel pressure drop and pulse frequency behavior.

The cross-flow filter control and operating strategy for carbonizer test campaign TR8 was established before testing. The carbonizer test matrix and the required response to various filter performance conditions were factored into the strategy outlined in Table 3.1.

The process data recorded by W-STC instrumentation during the filter test operation are:

- Filter tubesheet pressure drop
- Eductor pressure drop in each plenum
- Gas temperature in filter vessel (dirty and clean sides)
- Tubesheet temperatures
- Filter element temperatures (four points)
- Pulse tank pressure change during pulse
- Pressure in each plenum.

Table 3.1 Cross-Flow Filter Operating and Control Strategy Pulse Control Strategy

- Time delay between pulsing of the two plenums:
 - Do not pulse simultaneously, avoid dead-heading the carbonizer system
 - Select minimum time between pulsing the two plenums (set at 20 seconds), making the data more easily interpreted.
- Initiate operation on "DP Mode", with the trigger DP set at 7.5 in. H₂O above the baseline to build sufficient cake thickness, but to avoid filling the channels.
- Operate with maximum pulse tank pressure (700 psig) and a valve-open time of 0.25 s (0.05 s to open and 0.2 s to close) to achieve acceptable pulse intensity and to avoid patchy cleaning.
- Switch to Timer Mode, and let the filter DP increase—if the pulse frequency during DP Mode becomes greater than about once every 5 minutes and does not appear to be leveling to a steady condition.
- Continue test operation on sudden loss of filter DP (probably a broken filter) unless the condition is catastrophic.

Test Termination Criteria

- An alarm sounds if filter peak DP exceeds 9 psi on both the W-STC and the FWDC instruments.
- Request FWDC to modify carbonizer conditions if filter DP alarm is sounded.
- Request FWDC to shut down the test if the filter baseline DP exceeds 200 in. H₂O for more than 1 hour.

FWDC instrumentation also provides applicable data:

- Total filter vessel pressure drop
- Total gas volumetric and mass flow rate to the filter
- Filter vessel dust hopper accumulation (periodic)
- Status of carbonizer conditions and operation
- Inlet dust loading samples

- Outlet dust load samples
- On-line opacity meter
- Baghouse pressure drop and pulse frequency.

Using these data, several on-line filter condition and performance estimates can be made:

- Face velocity
- Inlet dust loading
- Outlet dust loading
- Filter element permeability
- Cake permeability and cake thickness
- Flow uniformity through the two plenums, and possible leak indication
- Pulse gas delivery
- Pulse intensity and valve-open time.

3.3.2 Ceramic Candle Filter

The fabrication, assembly, and inspection of the backup candle filter internals was completed in 1992. Gasket sets for the candles were received and inspected. The backup candle filter internals were stored at W-STC until required at the FWDC test site for TR9. The 30 Coors candle filter elements for the backup candle filter were received and inspected. Color variations were found in the candles that may correlate with their locations in the Coors ovens during fabrication. Permeability tests on the candles have been completed. Cold-strength testing on one candle has been completed with very favorable results. NDE testing (time-of-flight) was performed on all of the candles.

3.3.3 Alkali and Particle Sampling Probes

Two alkali/particle probes were fabricated, assembled, inspected, and shipped by W-STC to FWDC. Alundum extraction thimbles were inserted into the tip of the probes to collect particulate material. The high-temperature adhesive used to hold the thimble in place was selected from two candidates after oven testing of their effectiveness. W-STC provided both installation support and operating instruction for the probes. Some spare probe internal components were also fabricated so that alkali and particle sampling could continue efficiently during sample analysis periods.

Analytical and operating procedures were prepared for the probes. The analytical procedures specified by W-STC produce the total sodium and potassium and the water-soluble sodium and potassium content of the collected fly ash particles on the probe filter thimble, and of the condensed alkali on the cold finger. Probe handling procedures were also established to minimize losses and contamination.

The alkali probe phenomena and expected performance of the probe were reviewed based on W-STC past test experience with alkali probes, alkali getter reaction testing, and the filter material alkali exposure testing. Based on this review, some recommendations were proposed for taking alkali measurements during the carbonizer test campaign. The possible alkali transformations in the probe are:

- Alkali vapors (Na and K primarily in the forms of chlorides and hydroxides) may react with the fly ash particles collected on the filter thimble, reducing the amount of vapor reaching the condensation finger and generating more "insoluble" alkali in the collected fly ash. W-STC has measured this rate of reaction in extensive alkali getter testing. The rate of this reaction is relatively slow; and with the relatively short residence time of the sample gas in contact with the fly ash, little change in the vapor-phase alkali content will occur by this route. This phenomenon is of little concern for the probe located after the cross-flow filter, where dust loading is very low.
- Alkali vapors may react with the filter thimble material, reducing the vapor-phase alkali and increasing the thimble insoluble alkali content. Because the thimble contains significant insoluble alkali in its original state, and this alkali content is highly variable, it would be difficult to account for this conversion. W-STC has measured this rate in tests exposing ceramic materials to alkali vapors. Again, this reaction is slow at the temperatures of the sampled gas, and the contact time between the sampled gas and the thimble is so short that little change in the gas-phase vapor content will occur.
- Alkali vapors may be released from the collected fly ash and from the thimble because of a reaction with the HCl contained in the fuel gas. This mechanism could release water-soluble alkali species, but would not have much impact on insoluble alkali in the fly ash and thimble (based on past testing).
- Alkali may be released from the cement holding the thimble in place, and if the cement contains significant alkali, interpretation of the final chemical analyses could be difficult. Some of the water-soluble alkali contained in the relatively small amount of cement could be released, but the poor contact of the cement with the sampled gas would make this release very slow. The use of the low-alkali cement minimizes this as an issue.
- Condensation of alkali vapors on the collected fly ash and thimble may occur if the thimble region is cooler than the sampled fuel gas. This is

the most likely transformation that we expect, but it leads to having water-soluble alkali added to the fly ash and the thimble, which will be evident from the chemical analyses.

The general conclusion is that only the water-soluble alkali measurements are of significance to the vapor-phase alkali estimates. Although significant evidence exists for these expectations, they can only be confirmed by conducting controlled experiments using simulated fuel gas and injected alkali.

The following recommendations were proposed to achieve the highest alkali probe measurement accuracy:

- Minimize the fly ash collected on the probe by directing the sampling tip downstream.
- Ensure the highest gas sampling rate and thimble zone temperature by using the alkali probe configuration that provides the maximum exposure of the filter thimble to the hot gases.
- Eliminate the thimble (and its cement) in the alkali probe located after the cross-flow filter.
- Use the identified, low-sodium cement in the alkali probe located before the cross-flow filter.
- Sample for as long a period of time as possible for each test point.
- Wash the thimbles with deionized water before use in the tests to minimize their water-soluble alkali content.

The maximum and minimum vapor-phase alkali content of the fuel gas is estimated from the probe measurement results by following the relationships:

Maximum Fuel Gas Vapor Alkali Content (Na,K) =

$$\frac{[(\text{Water soluble alkali collected in the condensation finger}) + (\text{Soluble alkali collected on fly ash and collected on the thimble})]}{[(\text{Gas sampling rate})(\text{Gas sampling time})]}$$

Minimum Fuel Gas Vapor Alkali Content (Na,K) =

$$\frac{[(\text{Soluble alkali collected in the condensation finger}) - (\text{Change in soluble alkali content of the thimble and its cement})]}{[(\text{Gas sampling rate})(\text{Gas sampling time})]}$$

The fuel gas sampling rate is about 980 g/h of fuel gas. If there is 1 ppm(w) sodium vapor contained in the fuel gas, for example, then the sodium vapor sampling rate will be about 1 mg/h. If sampling can continue for 4 hours, about 4-mg soluble sodium could be collected in the probe, resulting in a wash concentration of about 20-ppm sodium following the analytical procedure.

The thimble, weighing about 9 g, contains about 1.8-mg water-soluble sodium in its initial, unwashed state. Little change is expected to the insoluble alkali content of the thimble. If all of the soluble alkali were driven off the thimble and collected in the condensation finger of the probe, it would be accounted for as fuel gas vapor-phase alkali. Washing the thimbles will minimize this error, and the error can at least be identified and included as part of the uncertainty of the measurement.

If about 0.1 g of cement is used, with a water-soluble sodium content of 0.05 wt%, the possible addition of soluble sodium to the condensation finger would be about 0.05 mg.

3.4 TOPPING COMBUSTOR

WPGBU began fabrication of the 700 and 1600°F preburners (shown in Figures 2.51 and 2.52 in Section 2), which will be required in the 18-in. MASB tests being conducted in the next fiscal year.

3.5 CPFBC PILOT PLANT

The conversion of the pilot plant to the CPFBC configuration was interrupted to perform one circulating bed carbonizer test run. As the reporting period ended, laboratory test results from this run were being compiled. Assuming the preliminary results confirm the validity of the test, conversion to the final CPFBC configuration is expected to start in October 1992.

Section 4

TASK 4—TEST PLAN IMPLEMENTATION

Section 4

TASK 4—TEST PLAN IMPLEMENTATION

4.1 OBJECTIVE

The Task 4 objective is to conduct tests that will characterize the performance of the key second-generation PFB combustion components (i.e., carbonizer, CPFBC, ceramic cross-flow filter, and topping combustor).

4.2 CARBONIZER PILOT PLANT

In November 1991 FWDC began operating a PFB combustion pilot plant at its John Blizard Research Center in Livingston, New Jersey. The facility has been designed to test a second-generation plant carbonizer [1] and then, after modification, a CPFBC/FBHE. Ceramic barrier filters provided by Westinghouse-STC are being used with these units to demonstrate particulate control capabilities.

Because a new facility and a new process (carbonization) were involved, the pilot plant began operation with a two-cyclone HGCU system. To minimize interactions, the stainless steel cyclones were placed in separate, refractory-lined pressure vessels with separate ash cooling and depressuring systems. After completing the planned carbonizer test matrix and ascertaining that the carbonizer fuel gas composition was compatible with a ceramic barrier filter (no tars or oil vapors in the gas), the second-stage cyclone was replaced first by a ceramic cross-flow filter and then by a ceramic candle filter.

The carbonizer has been operated in two modes—first as a jetting fluidized bed (Figures 2.1 and 2.2) and second as a circulating fluidized bed (Figures 2.9 and 2.11).

4.2.1 Jetting Fluidized Bed Mode

Operation. Carbonizer operation was smooth and controlled throughout the entire test program, including commissioning of the unit. Figure 4.1 is a typical start-up curve. The carbonizer and downstream components were heated to 900°F at approximately 150 to 200°F/h by hot air blown through the unit.

Upon reaching 900°F, sorbent was pneumatically injected into the carbonizer until a 19-ft expanded bed height was reached. During this filling operation, the cold sorbent and transport air cooled the unit to ≈550°F,

despite continued air heater operation. Once filling stopped, the unit was reheated. Upon reaching 925°F, airflow was reduced to ≈ 650 lb/h, and low-volatile petroleum coke was pneumatically injected into the carbonizer at about 300 lb/h. This air-to-coke ratio was estimated to yield a 1600°F carbonizer temperature. As Figure 4.1 shows, the coke partially burned/carbonized; bed temperature immediately rose to $\approx 1600^\circ\text{F}$. With a fixed coke flow rate, the airflow rate was trimmed to control the rate of temperature rise. The carbonizer was held at $\approx 1600^\circ\text{F}$ for 4 hours, with the coke-generated fuel gas heating—the farthest downstream component (the choked-flow orifice) to at least 1200°F. Coke was used during this heat-up period because its low volatile content ensured that no hydrocarbon vapors would be present that could condense on cold downstream components. After about 4 hours of operation, the coke was replaced with the coal of interest. The change to coal was smooth and virtually unnoticeable.

Gas Sampling. The fuel gas generated by the carbonizer is sampled at two different locations in the plant and analyzed within 24 hours by a gas chromatograph to determine the gas composition. The first sample, taken at the demister outlet, simply allows the slightly pressurized gas to flow into

Figure 4.1 Typical Carbonizer Start-Up Curve

and fill each of two 3-L Teflon bags. Hydrochloric acid (HCl) is injected into one bag to convert the ammonia in the fuel gas to ammonium chloride (NH_4Cl) to prevent its decomposition while awaiting analysis by the gas chromatograph. Similarly, sodium hydroxide is injected into the other bag to convert hydrogen sulfide to sodium sulfide or sodium hydrosulfide ($\text{Na}_2\text{S}/\text{NaHS}$). The second sampling involves extracting gas from the top of the carbonizer via the system shown in a simplified schematic (Figure 4.2). Immediately upon leaving the carbonizer, the gas is cooled to 400 to 600°F via a steam-jacketed pipe and particles are removed by a steam-traced cyclone and porous metal filter. Vapors are condensed by cooling the gas to approximately 60°F in a water-cooled heat exchanger; the gas is then collected in a knockout pot for analysis/to check for the presence of any tars or oils. A valve depressures the gas to ambient, sets the sampling rate at 250 sft³/L, and delivers the gas to a totalizing dry gas meter.

The second gas sample is taken immediately downstream of this meter in a 3-L Teflon bag. The system facilitates determination of the fuel gas water content and provides a cross check on the gas composition measured at the demister outlet (except for ammonia and hydrogen sulfide, which are water soluble and possibly partially absorbed by the condensate).

From the standpoint of the solids fed into and withdrawn from the plant during a test/setpoint, material samples were collected for analysis every time a feed system lock hopper was filled and a drain hopper was discharged. A setpoint was held for 4 hours, and gas samples were collected at about 1-hour intervals during this period. Allowing time for transition from setpoint to setpoint, for establishing steady-state conditions, and for tending to problems, we typically completed a setpoint every 12 hours. A test run typically involved starting preheat of the facility on Monday and shutting down the plant on Friday, with operation occurring around the clock during this period.

Based on the analyses of the gas and solids samples and the flow rate and temperature measurements of all streams entering and leaving the plant, mass, heat, and element balances were prepared for each setpoint. The balances closures (total out divided by total in) were checked to validate the accuracy of the setpoint.

Test Runs. Our jetting fluidized bed carbonizer test program consisted of eight test runs, comprising 37 setpoints and 533 hours of operation. Two test runs (TR3 and TR7) were aborted during start-up before establishing their first setpoints because of equipment problems unrelated to the process. Although Pittsburgh No. 8 bituminous coal and Ohio Plum Run dolomite were most frequently used, Illinois No. 6 and Wyoming Eagle-Butte coals and Alabama Longview limestone were also tested. As shown in Table 4.1, the Pittsburgh coal typically had a 3.5-percent sulfur content and a Free Swelling Index of 6.5; it is a highly caking coal that has caused agglomeration problems in other fluidized bed gasification programs. Thus, it was purposely selected for use in our test program to demonstrate the capability of the carbonizer

Figure 4.2 Freeboard Sampling System

Table 4.1 Typical Carbonizer Feedstock Analysis

Description	Petroleum Coke	Coal		
		Pittsburgh No. 8	Illinois No. 6	Eagle Butte
Feed Size	-1/8"	-1/8"/-3/16"	-1/8"	-1/8"
Free Swelling Index		6.5	4	0
Proximate, wt%				
Fixed Carbon	83.96	51.06	45.05	36.16
Volatile Matter	14.97	34.23	33.49	31.39
Ash	0.94	12.61	10.59	5.58
Moisture	0.13	2.10	10.87	26.87
Ultimate, wt%				
C	89.89	69.45	62.76	50.77
H	4.42	4.62	4.40	3.60
O	0.73	5.70	6.78	11.49
N	2.30	2.03	1.57	0.98
S	1.59	3.50	3.05	0.72
Ash	0.94	12.61	10.59	5.58
Moisture	0.13	2.10	10.87	26.87
HHV	15,743	12,724	11,172	8,729

process. The Eagle Butte, a subbituminous coal, contained 0.7-percent sulfur and 27-percent moisture. To preclude the risk of forming and condensing hydrocarbon vapors at start-up, when the plant "back end" was cold, the carbonizer was started up with a low-volatile petroleum coke. The transition to coal was made after about 4 hours, when the choked-flow orifice temperature reached approximately 1200°F. Typical analyses for the sorbents are presented in Table 4.2.

As shown in Table 4.3, 83.4 tons of four different fuels have been successfully carbonized—45.3 tons of Pittsburgh No. 8, 15.4 tons of Eagle Butte, 15.3 tons of Illinois No. 6, and 7.4 tons of coke. Table 4.4 identifies the operating conditions tested and presents test results.

Throughout this program, the carbonizer has operated at approximately 3 ft/s superficial gas velocity (measured in the 10-in.-I.D. section). Temperatures, pressures, and steam injection rates have ranged from 1500 to 1800°F, 10 to 14 atm, and 0 to 0.4 lb/lb coal.

The first four setpoints were conducted at approximately 10 atm(g), utilized Pittsburgh No. 8 coal and Plum Run dolomite, and investigated the

Table 4.2 Typical Carbonizer Sorbent Analysis

Description	Plum Run Dolomite	Longview Limestone
Supplier	Highland Stone Div. of Davon, Inc., Hillsboro, Ohio	Longview Lime Saginaw, Alabama
Feed Size	-1/8 in.	-1/8 in.
Composition		
CaCO ₃	55.96	96.90
MgCO ₃	42.82	1.91
S	0.19	0.15
Ash	1.03	1.05

effects of varying the carbonizer temperature between 1500 and 1700°F. The balance closures for the four setpoints (Table 4.5) were deemed good. Because of the high surface-to-volume ratio of the pilot-scale carbonizer and the cooling effect of the nitrogen from pressure tap purges and the internal drain cooler, the pilot-scale unit needs a higher air-to-coal ratio to reach the temperature predicted for a commercial plant. Air-to-coal ratios ranged from 1.42 to 2.50 and yielded carbon conversions of 28 to 43 percent. Sulfur-capture efficiencies were very good, with 88 to 95 percent of the released sulfur being captured by the dolomite. The coal nitrogen that was released to the fuel gas and appeared as ammonia ranged from 45 to 83 percent. In all cases the amount of sulfur and ammonia in the fuel gas was significantly less than that estimated for a commercial plant [2]; hence commercial plant emissions performance should be better than previously estimated.

Carbonizer fuel gas samples extracted from the top of the carbonizer and from a point downstream of the dry-quench spray tower were analyzed by a gas chromatograph and are in excellent agreement. The predominant gas components (in decreasing order) are nitrogen, carbon monoxide, carbon dioxide, hydrogen, and methane. Figure 4.3 compares gas sampled from the two locations during the first four Pittsburgh No. 8 coal test points on a nitrogen-free (vol%) basis; agreement was excellent. An increase in carbonization temperature and hence air-to-coal ratios raised the carbon monoxide level and lowered the methane level; other predominant components remained relatively unchanged. Analyses of the remaining 33 setpoints is under way.

4.2.2 Circulating Fluidized Bed Carbonizer

The circulating fluidized bed carbonizer test program entailed two test runs comprising four setpoints and 94 hours of operation, as shown in Table 4.6. The first test run was aborted during start-up before establishing the first setpoint because of an equipment problem unrelated to the process

Table 4.3 Carbonizer Operating Summary*

Description	Petroleum Coke	Pittsburgh No. 8	Illinois No. 6	Wyoming Eagle Butte	Plum Run Dolomite	Longview Limestone
Operating Time						
Shakedown Runs 1-2	8	5	---	---	13	---
Test Runs 1-8	59	289	76	96	317	203
Subtotal	67	294	76	96		
Total	533				330	203
Material Consumed (tons†)						
Shakedown Runs 1-2	1	0.8	---	---	0.7	---
Test Runs 1-8	6.4	44.5	15.3	15.4	12.3	5.8
Subtotal	7.4	45.3	15.3	15.4		
Total	83.4				13.0	5.8
*Jetting fluidized bed. †2000 lb.						

Table 4.4 Summary of Carbonizer Test Conditions

Setpoint	Date (Time)		Coal Type*/ Flow Rate (lb/h)	Sorbent Type†/ Flow Rate (lb/h)	Air (lb/h)	Steam (lb/h)	Pressure (Psig)	Temperature (°F)
Jetting Fluidized Bed Mode								
TR1.1	11/20/91	(03:45-07:45)	P/329	D/89	625	0.0	153	1610
TR1.2	11/21/91	(01:55-05:55)	P/235	D/67	585	0.0	133	1695
TR1.3	11/21/91	(11:15-15:05)	P/354	D/98	670	0.0	147	1645
TR1.4	11/21/91	(19:35-23:35)	P/400	D/91	560	0.0	125	1510
TR2.1	12/19/91	(01:50-05:50)	P/234	D/118	609	0.0	141	1700
TR2.2	12/19/91	(14:00-18:00)	P/315	D/94	546	0.0	132	1605
TR2.3	12/20/91	(01:30-05:30)	P/182	D/83	615	0.0	138	1800
TR2.4	12/20/91	(10:00-14:00)	P/376	D/99	884	0.0	195	1790
TR2.5	12/20/91	(16:00-18:00)	P/508	D/160	881	0.0	199	1690
TR4.1	03/10/91	(20:00-00:00)	P/259	D/98	680	62.2	151	1670
TR4.2	03/11/92	(04:00-08:00)	P/297	D/90	658	115.8	156	1600
TR4.3	03-12/92	(02:00-06:00)	P/207	D/86	630	95.2	126	1710
TR4.4	03/12/92	(17:22-21:00)	P/253	D/95	565	78.4	127	1608
TR4.5	03/12/92	(04:20-09:50)	P/370	D/106	900	151.7	193	1730
TR4.6	03/13/92	(13:00-17:00)	P/396	D/106	818	87.1	175	1725
TR4.7	03/13/92	(22:00-02:00)	P/450	D/182	775	54.0	178	1600
TR4.8	03/14/92	(02:00-06:00)	P/338	D/150	763	55/154.0	181	1590
TR5.1	03/24/92	(11:00-25:00)	P/229	D/60	608	89.3	135	1695
TR5.2	03/24/92	(99:00-23:00)	P/236	D/119	688	89.7	154	1708
TR5.3	03/25/92	(05:00-09:00)	P/237	L/70	720	99.5	147	1715
TR5.4	03/25/92	(14:30-down)	P/198	L/24	563	99.0	130	1706
*P=Pittsburgh No. 8; W=Wyoming Subbituminous; I=Illinois No. 6 †D=Dolomite; L=Limestone								

Table 4.4 (Cont) Summary of Carbonizer Test Conditions

Setpoint	Date (Time)	Coal Type*/ Flow Rate (lb/h)	Sorbent Type†/ Flow Rate (lb/h)	Air (lb/h)	Steam (lb/h)	Pressure (Psig)	Temperature (°F)
TR6.1	04/21/92 (21:00-01:00)	W/344	L/60	750	0.0	147	1583
TR6.2	04/22/92 (04:00-08:00)	W/329	L/67	650	0.0	132	1505
TR6.3	04/22/92 (14:10-18:10)	W/424	L/45	815	0.0	157	1513
TR6.4	04/22/92 (21:00-01:00)	W/478	L/55	1000	0.0	192	1610
TR6.5	04/23/92 (04:00-08:00)	W/462	L/17	975	0.0	190	1638
TR6.6	04/23/92 (18:00-22:00)	I/304	L/36	630	112.5	144	1603
TR6.7	04/23/92 (23:30-03:30)	I/232	L/39	570	46.4	121	1605
TR8.1	06/10/92 (20:00-01:00)	W/300	P/41	613	45.0	130	1470
TR8.2	06/11/92 (03:00-08:00)	W/310	P/55	700	46.5	137	1585
TR8.3	06/11/92 (13:30/09:30)	W/393	P/27	785	0.0	141	1587
TR8.4	06/12/92 (21:00-02:00)	I/320	P/70	860	51.2	113	1642
TR8.5	06/13/92 (07:00-13:30)	I/427	P/104	1010	128.1	170	1713
TR8.6	06/14/92 (21:00-04:00)	P/380	P/38/109	916	95.0	158	1700
TR8.7	06/15/92 (04:00-21:00)	P/344	P/111	890	103.2	155	1705
TR8.8	06/16/92 (02:00-06:00)	P/360	P/80	838	90.0	144	1700
TR8.9	06/16/92 (08:00-12:30)	P/350	P/84	800	56.0	136	1710
TR8.10	06/16/92 (14:30-19:30)	P/350	P/40	725	56.0	127	1720
Circulating Fluidized Bed Mode							
TR10.1	09/01/92 (00:00-60:00)	P/525	P/143	1170	150.0	116	1700
TR10.2	09/01/92 (16:30-20:30)	P/232	P/32	630	0.0	60	1700
TR10.3	09/02/92 (00:00-04:00)	P/300	P/25	840	0.0	75	1725
TR10.4	09/02/92 (06:00-10:00)	P/286	P/25	638	0.0	69	1600
*P=Pittsburgh No. 8; W=Wyoming Subbituminous; I=Illinois No. 6 †D=Dolomite; L=Limestone							

Table 4.5 Test Run 1 Balances and Summaries

Closure*	1.1	1.2	1.3	1.4
Mass Balance	0.982	0.953	0.937	0.996
Heat Balance	0.967	0.990	0.865	0.954
Elemental				
C	1.015	1.042	0.918	0.957
H	1.085	1.004	0.992	1.177
O	1.049	0.988	1.040	1.169
N	0.955	0.938	0.935	1.013
S	1.378	1.118	0.962	0.828

*Out divided by in.

Figure 4.3 Pittsburgh No. 8 Carbonizer Fuel Gas Compositions

Table 4.6 Circulating Bed Carbonizer Summaries

Run Data	Petroleum Coke	Pittsburgh No. 8	Plum Run Dolomite	Genstar Limestone
Operating Time (hrs)				
Test Run 9	14	21	35	---
Test Run 10	4	55	6	53
Subtotal	18	76		
Total	94		41	53
Feedstock Consumption (tons)*				
Test Run 9	0.7	1.3	0.9	---
Test Run 10	0.4	7.5	0.2	1.5
Subtotal	1.1	8.8		
Total	9.9		1.1	1.5
*2000 lb.				

(leaking bag in baghouse filter). Pittsburgh No. 8 Coal and Genstar Limestone (see Table 4.7 for a typical analysis of the latter) were the feedstocks used during the four setpoints completed in Test Run 10.

The carbonizer operated at a nominal 10 ft/s velocity at temperatures of 1600 and 1700°F and pressures of 5 and 9 atm. When this section of the report was written, data analyses had not yet been completed. However, the macroscopic performance of the jetting and circulating beds appears similar. Because of its simplicity, the jetting fluidized bed appears best for second-generation PFB combustion plants.

Table 4.7 Typical Circulating Bed Carbonizer Sorbent Analysis

Description	Plum Run Dolomite	Genstar Limestone
Supplier	Highland Stone Div. of Davon, Inc. Hillsboro, Ohio	Genstar Stone Products Company Hunt Valley, Maryland
Feed Size	-1/8 in.	16 x 200 mesh
Composition		
CaCO ₃	55.96	96.0
MgCO ₃	42.82	2.0
S	0.19	0.0
Ash	1.03	2.0

4.3 CERAMIC CROSS-FLOW FILTER

On May 26 the first hot testing of the cross-flow filter was initiated in TR7, with the testing started using the Eagle Butte coal and Longview limestone. Preheated air began flowing through the plant at 0700, with building of the dolomite bed begun in the afternoon and continuing most of the day. Pressure drop did not build up across the filter during this whole dolomite feeding period. Coke feeding began early the next day and continued until about 0600, when transition to coal feed was started. When coke feeding began, pulse cleaning of the filter was initiated because of the pressure drop increase measured. By about 0630 the carbonizer bed had become completely coal fed. Behavior of the filter appeared normal at this point, with low pressure drops and very stable pulse cleaning—very close to the performance predicted for the filter (13 to 20 in. H₂O tubesheet pressure drop and 10- to 20-min pulse cleaning cycles). At about 1000 to 1100, the plant pressure was observed to be building, and the coal feed rate to the carbonizer was decreased. The cross-flow filter performance still appeared normal, with acceptable pressure drop and cleaning frequencies. Some of the instrumentation had been incorrectly configured, so some of the process data were not available. In particular, the pulse intensity in one of the two plenums in the filter could not be measured during the test. At about 1530 on May 27, after about 40 hours of operation, the test crew decided that the plant should be shut down because all indications were that the critical choked-flow orifice was plugged. Also, a significant amount of fly ash had been removed from the plant baghouse during the test period, giving rise to the expectation that a leak had occurred in the filter, although none of the filter instrumentation indicated such a leak.

After cooldown of the equipment, the pipe section containing the critical choked-flow orifice was opened, and the orifice was found plugged with refractory with refractory fibers of the type used to insulate sections of the filter tubesheet. Inspection of the cross-flow filter tubesheet revealed one slot, about 1 in. long by 1/8 in. wide that had not been seal welded; W-STC estimated that this slot could have accounted for a maximum dust leak of about 1 lb/h. In addition, one of the pulse cleaning pipes had become misaligned so that its nitrogen pulses were bouncing off the tubesheet and had blown the unprotected soft insulation out of the unit.

The tubesheet fiber insulation was repaired, and a metal liner was installed to protect the fiber blanket from direct gas impingement during future testing. The open slot in the tubesheet was welded closed, and new pulse tube positioners were installed to prevent misalignment of the pulse tubes. All instrumentation errors were corrected.

Following the repair of the cross-flow filter, Test Run 8 was restarted on June 8. During the testing, several setpoints were completed with Wyoming Eagle-Butte coal, Illinois No. 6 coal, and Pittsburgh No. 8 coal. The filter pressure drops were low again (8 to 20 in. H₂O), pulse-cleaning frequency was at 6- to 50-min intervals, and the filter cake was easily cleaned during the testing, even at very high dust loadings, estimated to be up to

about 35,000 ppm(w). This is a very significant test result, leading to expectations of very effective commercial performance of the carbonizer-filter system. Plant baghouse dust removal monitoring, isokinetic particle sampling of the fuel gas particulate content, and the clean condition of fuel gas samples withdrawn downstream of the filter indicated that the filter particle-removal performance was very good. After about 100 hours of testing, filter performance data began to indicate a possible leak, as illustrated by the data in Figure 4.4, which shows the volumetric gas flow through each of the two cross-flow filter plenums. The leak indication is the divergence of these two flows, with the Plenum 1 flow dropping to near zero. Carbonizer testing was continued even after it was clear that a significant leak existed somewhere in the filter. The carbonizer test matrix was completed on June 17. The cross-flow filter conditions and performance are summarized in Table 4.8 for the six setpoints recorded during TR8.

Preliminary inspection of the filter internals after system cooling and removal of the head and internals revealed the following:

- Some of the filter holddown clamps had come loose.
- On five of the mounts, the ceramic gaskets were displaced, bulging on one end and causing a gas path leak.
- On one filter, at the bottom of the plenum, the flange compression bars had dropped off the flange, the ceramic gasket and pads were missing from the mount, and the filter flange was broken.
- One other filter element showed evidence of a possible delamination crack.
- No other filters appeared damaged.
- The lowest filter element on each of the two plenums appeared considerably cleaner than the filter elements at higher positions.

Further detailed inspection of the internals was conducted, and the recorded run data were evaluated. The following evaluations are continuing:

- Identification of the probable mechanisms for the mount failures and development of recommended mount and gasket improvements
- Examination of the filter performance data trends and development of performance conclusions for each of the feedstocks and carbonizer conditions tested
- Characterization of the ash collected in the filter during the several setpoints to continue compilation of the data base being developed on fly ash filter cake behavior
- Examination of the filter elements used in the testing.

Figure 4.4 Gas Flow Through Filter Plenums

Table 4.8 Summary of TR8 Cross-Flow Filter Conditions and Performance

Conditions	Wyoming Eagle Butte			Illinois No. 6		Pittsburgh No. 8
Setpoint	1	2	3	4	5	6
Setpoint Duration (h)	5	5	20	5	30	7
Carbonizer Temp. (°F)	1500	1600	1600	1600	1700	1700
Filter Temp. (°F)	1100	1200	1250	1300	1450	1460
Inlet Dust Loading [ppm(w)]	8,420*	3,460*	28,010*	5,692*	26,982†	5,681†
Outlet Loading [ppm(w)]	2†	0†	---	0†	982§	831†
Tubesheet ΔP (in. H ₂ O)†	8	8	9	12	13/18	12
Trigger ΔP (in. H ₂ O)	17	17	18	18	20	20
Pulse Cycle Time (min)	30	50	20	8	8	240

*Based on filter hopper material collection.

†Based on single alkali probe collection (nonisokinetic).

§Based on isokinetic particle probe sample (after leak initiation).

¶13 to 4 in. H₂O pressure loss caused by eductor.

The permeability of some the undamaged cross-flow filter elements was measured after cleaning. A comparison of the new and used cross-flow filter element pressure drops (Figure 4.5) shows that the ratio of the conditioned filter permeability to the new permeability is about 0.2—typical of testing in other field and in-house units. The range of cross-flow filter element pressure drops measured at the nominal face velocity of 1.8 ft/min during the FWDC carbonizer tests is also superimposed on the figure, after adjustment to low-temperature conditions. This range falls directly on the pressure drop measured for the cleaned, used filter elements, indicating that the pulse cleaning in the FWDC carbonizer filter tests had been very effective. Examination of the cross-flow filter elements also showed that carbon was deposited on the clean and dirty sides of the filter channels as well as throughout the pores, but this carbon layer had no influence on filter permeability. We do not know when this carbon deposition occurred during testing.

The carbonizer fines collected in the cross-flow filter showed low flow resistance and appear to be easily cleaned without reentrainment. The baseline pressure drop of 8 to 12 in. H₂O with a nominal pulse cycle of 6 to 50 min is attractive compared with the Texaco entrained gasifier filter performance of 100 to 200 in. H₂O baseline pressure drop with a nominal pulse cycle time of 1 to 3 min.

Figure 4.5 Comparison of Cross-Flow Permeability

The conclusion reached after evaluating the filter element clamping incident is that the use of Haynes 556 bolts with a Type 316SS post increased bolt stressing, but was not a contributing factor to the gasket loss because of the relatively low filter temperature (1100 to 1450°F). The problem arose from the fact that the gaskets were not properly fabricated; the thickness of the two short sides was not uniform, preventing uniform loading when the filter elements were installed. Also, the gaskets were not fully captured by the plenum structure.

Such an occurrence should be eliminated by using new, properly fabricated gaskets and by modifying the recess in the existing mounts so that the gasket will be captured/contained. Other modifications to the filter element installation procedure have been identified that will ensure proper gasket integrity.

4.4 CERAMIC CANDLE FILTER

In parallel with the FWDC conversion of the jetting fluidized bed carbonizer to circulating bed operation, the backup candle filter internals were installed in the filter vessel. TR9 was the first attempt by FWDC to operate the carbonizer in a circulating bed mode, the main objectives of the testing being to explore the feasibility of circulating carbonizer operation and determine whether it offered any obvious/significant advantages over jetting fluidized bed operation. The conditions to be faced by the candle filter were uncertain with respect to dust loading and the nature of the fly ash. The nominal conditions of the test were:

System pressure:	95 psia
Filter temperature:	1500°F
Fuel gas flow:	250 aft^3/min
Filter face velocity:	3.8 ft/min

The 22 candles and their locations in the two plenums were given the designations shown in the layout in Figure 4.6.

TR9 began with coke ignition at about 1700 hours on August 25, after establishing a 600°F filter temperature. Coal feeding began at about 0300 on August 26, establishing a filter outlet temperature of about 1440°F by 0600. The pressure differential existing across the candle filter tubesheet during the first 12 hours of TR9 is plotted in Figure 4.7. Based on earlier cross-flow filter tests, initial operation of the candle filter appeared normal with a baseline pressure drop of about 20 in. H_2O . After coal feeding began and flow rates were increased, both the pulse cleaning frequency and the baseline pressure drop began to increase, as is expected with the rise in dust loading that occurred. Dust loadings were high—at least 50,000 ppm(w). The pulse period approached about 5 min between 0630 and 0830 and then became short enough that the trigger pressure drop was increased, as well as the pulse reservoir pressure (from 300 to 500 psig). This increase in trigger pressure drop did not result in a longer pulse period, contrary to expectations, even though pulse delivery increased from about 0.5 to 1.0 lb gas/pulse. At about 1000, after about 7 hours of operation with coal, the vent line of the candle filter lock hopper plugged, and the vessel could not be drained for about 5 hours. After the hopper drain problem was recognized, the feed rates of coal, sorbent, and air to the carbonizer were slowed and eventually stopped, while efforts were made to clear the line. Nitrogen purges remained on in the carbonizer and cooled the candle filter at about 200°F/h. After clearing the lock hopper vent line, 1423 lb of solids were drained from the hopper, using the filter vessel spargers to assist in solids removal.

FWDC measured the bulk density of the 1423 lb of solids drained (about 44.5 lb/ft^3) from the filter vessel and estimates that the accumulation of solids in the filter vessel may have submerged the lower 31 in. of the candles. Of course, the solids level profile within the filter vessel is unknown, but subsequent inspection of the filter internals indicated a fill depth based on the patterns on the candles that confirms this estimate.

Figure 4.6 Candle Filter Arrangement—Plan View

Figure 4.7 Candle Filter Tubesheet Pressure Differential During TR9

The system underwent a "hot" restart at about 1930. While waiting for the establishment of Setpoint 1 test conditions, a baghouse leak developed that forced the shutdown of the carbonizer pilot plant at about 2300 on August 26.

Inspection of the filter vessel internals was made following system cooldown, when soft insulation and coal char were found on the screen that protects the choked-flow orifice, indicating a filter leak. One broken candle was found in Plenum 1, broken 15 in. below its flanged end. This was candle C-26, at Location 1-11. The broken piece was found in the hopper. The filter was replaced, and the other candles were visually inspected, checked for tightness in their holders, and found undamaged. Preliminary inspection of the broken candle showed that the break occurred near a "fill line" in the candle, a line formed during manufacture. It is speculation to assert that this resulted in the break. Review of the filter plenum flow data collected during the last half of August 26 is shown in Figure 4.8. The two plenum flow rates, reported as face velocities on each plenum, are shown as a function of time. The two plenum flows are identical until about 1930, during the restart of the test. At this time the plenum flows diverged, with the flow in Plenum 1 rising, and the flow through Plenum 2 falling significantly. This is believed to be when the candle broke. There are several scenarios and mechanisms proposed as to why this candle broke, and the possible feasibility of these is being checked.

With regard to the increase in pressure differential shown in Figure 4.7, this may be attributed to the filling of the filter vessel with ash (hopper drain bridged/blocked) rather than to the ramping of flow rates. Since problems began after transitioning to coal feed, FWDC speculates that a sticky ash condition may have developed that caused solids to accumulate in the filter vessel and eventually block the vent line of the drain lock hopper below it. Since a sticky ash condition had not been experienced in previous start-ups, the carbonizer solids circulation rate is suspected as being too low when coal feed began (the mass of hot solids circulating through the carbonizer was not large enough to heat the highly caking Pittsburgh No. 8 coal through its sticky regime before particulate escaped to the barrier filter).

After locating and plugging the leaking bag in the baghouse filter and completing the inspections and repair of the candle filter, TR10 was begun.

Preheat of the carbonizer for TR10 was initiated on August 30. Coke feed was started at about 2145. Prior to coke feeding, the filter established a 20 in. H₂O baseline pressure drop and was pulse cleaned three times over about a 4-hour period at a 50 in. H₂O trigger pressure drop. During coke feeding the rate of rise of the filter pressure drop indicated the possibility of broken candle elements. Based on the plenum flow measurements, the investigators guessed that at least one filter had broken on Plenum 1, and this could have occurred as early as 1330 on August 31, during the systems preheat and at the very first pulse cleaning event. Operation of the carbonizer continued through September 3, for a total of about 50 hours, with the filter

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Figure 4.8 Candle Filte Face Velocities During TR9

maintaining a pressure drop, but with very infrequent pulse cleaning. Following termination of TR10 and system cooling, the filter vessel was opened for inspection. The following observations were made:

- The pulse tubes were aligned and not bent.
- The tubesheet was loaded with dust up to the top of the expansion web (about 11 in. high).
- The "high-side" pressure tap used to measure Plenum 1 flow was broken (eroded). The Plenum 2 tap was intact.
- Ash was tightly packed (bridged) between the candle elements and over their full lengths (except the lower few inches), forming a cylindrical cast. This ash was crusty and had to be physically removed.
- The filter vessel cone was filled with ash.
- Of the 11 candles on Plenum 1, 7 were broken at the tips (3 to 5 in. from the bottom). The broken candles were at positions 1-7, 1-6, 1-5, 1-3, 1-2, 1-8, and 1-9 on Figure 4.6.
- After cleaning the bridged dust from the candles, the unit was lifted. At this point, one filter on Plenum 2 (at location 2-2 on Figure 4.6) broke off.
- Significant dust was found within all of the candles, broken and unbroken, on both plenums.
- Filling of the filter vessel with ash, as experienced in TR9, should be prevented. Long-term operation of the filter with no pulse cleaning, as experienced in TR10, should not be allowed.
- The fact that all of the candle failures occurred on one plenum indicates that the failures were related to various system conditions and were not caused by filter candle properties.
- The conditions that the candle filter was subjected to after the TR10 broken-candle event were far from design conditions [that is, the dust loading ranged from 50,000 to 200,000 ppm(w)].

4.5 SAMPLING AND DATA EVALUATION

4.5.1 Freeboard Gas Sampling

This system (shown earlier in Figure 4.2) extracts a 350 sft³/h gas sample from the top of the carbonizer. After particulate removal and cooling to approximately 90°F, the condensate and a gas bag sample are collected for analyses. The freeboard sampling system performed very well throughout the

jetting fluidized bed carbonizer operating mode; data were in excellent agreement with the gas sample collected at the demister outlet. Because of the high solids loading in the carbonizer fuel gas in the circulating bed mode, the sample line continually plugged. Thus gas analyses were based on the demister samples.

4.5.2 Carbonizer Alkali Probe Data Evaluation

The alkali probe data collected during TR8 and TR10 have been compiled, and the following evaluations are under way:

- Review of data trends
- Comparison of the data with alkali probe data collected earlier by W-STC at other facilities
- Error analysis estimate for the data
- Comparison of test data trends with equilibrium estimates
- Analysis of the alkali content of bulk solids drained from the carbonizer, the primary cyclone, and the filter vessel for comparison with the probe solids
- Interpretation of the significance of the data and conclusions
- Recommendations for improved probe features and improved sampling procedures.

The data indicate some clear trends:

- Under all test conditions, very little alkali was found condensed on the probe cold finger, and the amount condensed was much less than equilibrium estimates, was less than turbine alkali standards, and was not very sensitive to the gas temperature at the probe.
- The solids collected on the probe filter cartridge contained a large amount of water-soluble alkali, and this content of water-soluble alkali increased significantly as the gas temperature entering the probe was increased.
- As the gas temperature of the probe sample rose, the alkali content of the fuel gas (based on the total water-soluble content of the collected solids and cold finger condensate) increases significantly. These total fuel gas alkali levels are close to those expected based on equilibrium estimates (Figure 4.9).
- Since the internal filter in the probe was near to the temperature of the sampled gas, the water-soluble alkali collected could not have been present because of pure condensation; it must have been present as a result of physical absorption. If the alkali vapor had reacted with the collected

The isokinetic particle sampling measurements were made at the barrier filter inlet and outlet, as shown in Figure 4.10, with the alkali probe fitted with an isokinetic tip. Since these measurements were made via the same probe, alkali measurements were given a higher priority, and only six particulate samples were taken (three in each of Test Runs 8 and 10) to investigate/

4.5.3 Isokinetic Particle Sampling

- Losses of alkali vapor (in the form of alkali aerosol) passing through the probe trap, could result in significant errors in the probe results. Such losses have not yet been identified.
- Either the alkali vapor was effectively removed by the collected solids within the probe or it was never released from the solids. The fact that the content of water-soluble alkali increased with a higher gas temperature can be interpreted to indicate that there was more alkali vapor available to accumulate on the solids as the temperature increased.
- solids, the alkali would not have been identified as water soluble, but would have been in the insoluble form.

Figure 4.9 Alkali Probe Measurement Results

Figure 4.10 Cascade Impactor, FOAM, and Opacity Measurement Locations

develop the feasibility of these measurements. A stainless steel internal filter for temperatures below 1500°F and a ceramic filter for temperatures at or above 1500°F were utilized/evaluated. The internal ceramic filter cracked both times it was used, and the sample was lost. As a result, Hastalloy X metal filters have been ordered for high-temperature service. The four remaining samples were being analyzed/evaluated as the reporting period ended.

4.5.4 Opacity Measurement System

The Opacity Measurement System shown in Figure 4.11 has worked during cold tests of the ceramic-barrier filter. The first time it was activated during carbonizer operation (Test Run 8), an evaporated condensate film believed to have formed during plant heatup/start-up prevented its operation. The film was cleaned off, but when the system was reexposed to the fuel gas, a cross-flow filter dust leak resulted in a 100-percent opacity reading. The system was next activated in Test Run 10. Although a change in start-up/operating procedures prevented the formation of the condensate film experienced in Test Run 8, a similar candle filter dust leak also resulted in a 100-percent opacity reading. Removal and examination of the pressure-isolation lens assemblies in Test Runs 8 and 10 indicates that the nitrogen purges are keeping the lenses free of dust films. However, because of the barrier filter dust leaks experienced to date, the system has not yet been calibrated vis-a-vis isokinetic particle sampling measurements.

4.5.5 Carbonizer Sulfur-Removal Data Evaluation

Test data from the carbonizer tests (TR1, TR2, TR4, TR5, and TR6) have been reviewed, and compilation of the data has been completed. The sulfur-removal data will be correlated using the approach applied previously to KRW in-bed desulfurization data, with the carbonizer sulfur-removal model as the basis for correlation. The sulfur-removal performance of the carbonizer will also be compared with that of the KRW gasifier, since the operating conditions are similar.

FWDC sent samples of Plum Run dolomite and Longview limestone to W-STC, and chemical analyses of the two sorbents have been completed. The chemical analyses are listed in Table 4.9. TGA of the Plum Run dolomite and Longview limestone was partially completed during the reporting period. The sorbent reactivity at 1600 and 1700°F, in both H₂S and SO₂ gases, is being tested. Tests will also be conducted to provide information on the ability to oxidize the sulfided sorbents to the sulfated condition. The testing provides empirical parameters that are used by W-STC in the carbonizer and CPFBC sulfur-removal models.

Table 4.9 Sorbent Chemical Analyses, wt% (Antech Ltd.)*

Element	Longview Limestone	Plum Run Dolomite
Aluminum	0.053	0.048
Calcium	32.0	21.0
Iron	0.096	0.25
Potassium	0.031	0.026
Magnesium	0.32	12.0
Silicon	0.80	0.20
Sodium	<0.01	0.023

*Method Reference: U.S. Environmental Protection Agency, 1987, Test Methods for Evaluating Solid Waste, SW-846, 3rd ed., Office of Solid Waste and Emergency Response, Washington, DC.

Silicon Analysis: Perkin-Elmer, Analysis of Silicates and Carbonates—Lithium Metaborate Fusion Procedure, March 1973.

4.6 GENERAL CONCLUSIONS

With regard to the overall carbonizer test program, at no point have any tar or oil vapors been found/condensed from the fuel gas. This is a very important finding because their presence might jeopardize barrier filter operation by causing filter blinding and blocking. Freeboard and demister gas samples have been in close agreement, indicating that no shifting in gas composition is occurring in the downstream piping; the high carbonizer operating temperature, relatively long in-bed gas residence time (approximately 8 seconds for the jetting fluidized bed configuration), high pressure, exposure to lime-based sorbents, and method of coal/air feed (coal and air injected at the same point) destroy all tar and oil vapors in the bed.

The solids residue from the process is a char/sorbent mixture containing fine carbon particles. The residue is essentially free flowing and not sticky, except possibly at very low sorbent-to-coal ratios. If highly caking bituminous coals are operated at a low temperature (1500°F) and a low sorbent-to-coal mass ratio (0.15), a sticky ash condition can develop. After several hours of operation, this condition can lead to particle growth/agglomeration and a loss of fluidization. However, these operating conditions are far outside the range of commercial plant interests and pose no risk to successful commercial plant operation.

With regard to ceramic barrier filter operation, one of the major technical uncertainties of the proposed commercial plant was whether the

carbonizer-generated fuel gas would contain tar or oil vapors or sticky ash that would blind barrier filters. Fortunately, the fuel gas and residues from the four different fuels posed no problem to the barrier filter; the filter cakes were easily cleaned, there was little change in filter permeability, and there were no indications of significant particle reentrainment during pulse cleaning.

With regard to topping combustor development, no testing occurred during this reporting period. Analysis of previously collected test data has revealed the need to redesign the combustors primary premix rich-burn zone to minimize the conversion of carbonizer fuel gas ammonia to NO_x . In September 1992, Westinghouse PGBU was released to proceed with this redesigning, and testing of the modified combustor should be completed by the end of 1992.

In summary, the carbonizer test program has been very successful. It has demonstrated that carbonizer operation is smooth and controlled, emissions are lower than were previously estimated for a commercial plant, and the char/sorbent residue and fuel gas from the process appear compatible with particle-capturing ceramic barrier filters.

Having successfully completed the carbonizer test program, the pilot plant is being modified for circulating pressurized fluidized bed operation. This modification primarily involves replacement of the L-valve shown in Figure 2.5 (Section 2) with the external fluidized bed heat exchanger shown in Figure 2.4 (also Section 2). In the next effort, the effectiveness and emissions of circulating pressurized fluidized bed combustion will be investigated in a series of tests in an 8-in. I.D. CPFBC fed with coal and then with the carbonizer char/sorbent residue. CPFBC testing, which will also focus on converting the calcium sulfide generated in the carbonizer to calcium sulfate, will commence in December and be completed by May 1993.

4.7 REFERENCES

1. A. Robertson, et al. "Second-Generation Pressurized Fluidized Bed Combustion Plant Carbonizer Development Work," IECEC-90, 1990.
2. A. Robertson, et al., "Second-Generation PFB Combustion Plant Performance and Economics," 1988 Seminar on Fluidized-Bed Combustion Technology for Utility Applications, Palo Alto, California.

Section 5

TASK 5—PHASE 2 TOPICAL REPORT

Section 5

TASK 5—PHASE 2 TOPICAL REPORT

5.1 OBJECTIVE

The objectives in Task 5 are to conduct a conceptual design study aimed at evaluating the cost competitiveness of small, industrial-sized, second-generation PFB combustion plants and to deliver to the DOE COTR a topical report that details the results of the Phase 2 testing and updates the Phase 1 electric utility commercial plant design.

5.2 STATUS

5.2.1 Small Industrial Plant Study

In a previous reporting period, a conceptual design study was conducted to evaluate the technical and economic feasibility of using the second-generation system for industrial cogeneration. Conceptual plant designs were developed, and cost estimates were prepared for these plants as well as for alternative conventional cogeneration plants. Conventional cogeneration plants utilize atmospheric fluidized bed combustion (AFBC) with back-pressure steam turbines or gas-fired cogeneration systems. Six cases were selected for study. Two of the cases, referred to as the large plants, were based on a 21-MWe Westinghouse 151 gas turbine. These plants would be built for a large industrial user, and they were designed for maximum steam generation. The four remaining plants were based on two, parallel, 3.5-MWe Allison 501 gas turbines. These plants, referred to as the small plants, were designed for minimum physical size, minimum capital cost, and simple operation. One of the small plants is a 1-1/2 generation plant that uses natural gas in the topping combustor instead of generating fuel gas from the coal. When the six cases are referred to collectively, the term "second-generation systems" is used, even though one case is actually a 1-1/2 generation system.

An economic analysis compared the plants developed in this study with conventional cogeneration systems that have comparable steam outputs. The following assumptions were made:

- All the electricity generated by the various systems could be used within the facility.
- Differences between the electrical outputs of the different systems would have to be satisfied with purchased power.

Assuming a 20-year plant life and costs typical of an Ohio Valley location, the resultant cost of steam was calculated for the study cases and the comparison plants.

Under the base-case cost assumptions, the lowest cost of steam in the larger plant category was the AFBC cogeneration comparison plant. The lowest cost of steam in the smaller plants would be achieved with a conventional natural-gas-fired gas turbine cogeneration system. Second-generation systems produce a greater amount of electrical power per pound of process steam generated than the comparison plants; thus the latter must purchase varying amounts of electrical power. As a result, higher rates for purchased power lower the cost of steam for the second-generation system relative to the other options. How the effects of income tax are considered in the economic analysis also affect the relative cost of steam between the second-generation plants and the comparison plants. With other electric utility rate structures in the Ohio area and different treatment of the effect of income tax, the second-generation system becomes the option with the lowest cost of steam in the larger plants. In the smaller plants, the conventional gas turbine cogeneration system remains the lowest cost of steam option—even at the higher electric rates used in this study.

An increase in the size of the plants will tend to benefit the economics of the second-generation system vs. the alternative systems because the second-generation systems have higher capital costs than alternative cogeneration systems. As the plants get larger, capital costs do not increase as much as the output of electricity and steam. For example, the cost of such systems as coal handling and slurry feeding decreases on a pound-of-coal basis as the plant coal flow increases. The effect on capital equipment is the same.

The larger plants in this study appear to be a borderline size where the second-generation system may or may not be economically attractive, depending on the specific costs of a user. As the size increases, the cost scenarios under which the second-generation system results in the lowest cost of steam will become more numerous. There are certainly electric rates in different parts of the country that are as high or higher than 7.60¢/kWh. At these rates the second-generation system can be of economic benefit—even at the 41-MWe size.

During FY92, a detailed report on the study was submitted to METC and approved for publication. The DOE was given the results of the study on September 27, 1990; during this reporting period some additional analyses requested by the DOE were performed, and the report was prepared. The general conclusion of the study is that second-generation systems can be built as small as 7 MWe for industrial cogeneration. Most of the technological concerns for these industrial-size plants are the same as those determined for the utility-size plant—the subject of research and development efforts now under way.

5.2.2 Commercial Plant Update

The low-Btu fuel gas system of a commercial plant will contain relatively large valves to regulate or shut off the flow of fuel to the topping combustors in the event of a plant upset, change of load, or loss of load. An additional system of valves is required to ensure over-speed protection for

the gas turbine. Because of the large inventory of hot, pressurized air in the CPFBC subsystem and piping, merely shutting off the fuel is not sufficient for over-speed protection. The considerable amount of pressurized air and thermal energy that exist in the CPFBC subsystem from the compressor discharge to the topping combustor inlet must be controlled to prevent excessive over-speed of the gas turbine/generator unit and subsequent catastrophic failure.

Because of the relatively large sizes and 1600°F temperature, these valves are not currently marketed by the normal gas turbine valve suppliers. However, conversations with these suppliers during Phase 1 indicated that current technology should be adequate to support their design and manufacture, and a major R&D development effort should not be required.

A plan for quantifying the design requirements of these valves (closure times, maximum permissible leak rates, etc.) was prepared and presented to METC for review/approval. Upon receipt of METC approval, this valve development effort will begin.

Section 6

TASK 6 – PROGRAM MANAGEMENT / REPORTING

Section 6

TASK 6—PROGRAM MANAGEMENT/REPORTING

6.1 OBJECTIVE

The Program Management/Reporting task objectives encompass overall guidance and control of the various test programs so that the greatest amount of information is obtained on time and for the least expenditure.

6.2 STATUS

6.2.1 Technical Progress

During the period beginning November 1991 and ending September 1992, a series of tests has been conducted in the carbonizer to ascertain its performance characteristics. The carbonizer operated for 533 hours in the originally proposed jetting fluidized bed configuration at 37 setpoints. Testing was completed for four different fuels, with 83 tons of feedstock consumed. An exploratory circulating fluidized bed carbonizer test entailing four setpoints and 9.9 tons of two different fuels was also completed. Throughout the program, carbonizer operation was smooth and controlled. Feedstocks were switched "on the fly" from petroleum coke to subbituminous to bituminous coal and from limestone to dolomite without causing any process upsets or problems.

At no point in the carbonizer test program have any tar or oil vapors been found/condensed from the fuel gas. This is a very important finding as their presence might jeopardize barrier filter operation by causing filter blinding and blocking. Freeboard and demister gas samples have been in close agreement, indicating that no shifting in gas composition is occurring in the downstream piping; the high carbonizer operating temperature, relatively long in-bed gas residence time (approximately 8 seconds for the jetting fluidized bed configurations), high pressure, exposure to lime-based sorbents, and method of coal/air feed (coal and air injected at the same point) destroy all tar and oil vapors in the bed.

The solids residue from the process is a char/sorbent mixture containing fine carbon particles. The residue is essentially free flowing and not sticky, except possibly at very low sorbent-to-coal ratios. If caking bituminous coals are operated at low values of 1500°F and sorbent-to-coal mass ratios of 0.15 and lower, a sticky ash condition can develop, leading to particle growth/agglomeration and a loss of fluidization after several hours of operation. These operating conditions are far outside the range of commercial plant interests and pose no risk to successful commercial plant operation.

With regard to ceramic barrier filter operation, one of the major technical uncertainties of the proposed commercial plant was whether the carbonizer-generated fuel gas would contain tar or oil vapors or sticky ash that would blind barrier filters. The fuel gas and residues from the four

different fuels posed no problem to the barrier filter; the filter cakes were easily cleaned, there was little change in filter permeability, and there were no indications of significant particle reentrainment during pulse cleaning. Although the filter operating times were relatively short, these findings are very encouraging. Unfortunately, mechanical- and structural-type problems were experienced with both the cross-flow and candle barrier filters. Analyses are under way to better understand the causes of the problems.

In summary, the carbonizer test program has been very successful. It has demonstrated that carbonizer operation is smooth and controlled, that emissions are lower than were previously estimated for a commercial plant, and that the char/sorbent residue and fuel gas from the process appear compatible with particle-capturing ceramic barrier filters.

Having successfully completed the carbonizer test program, the pilot plant is being modified for circulating pressurized fluidized bed operation. In this next effort, the effectiveness and emissions of circulating pressurized fluidized bed combustion will be investigated in a series of tests in an 8-in. I.D. CPFBC operating with coal and then the carbonizer char/sorbent residue. CPFBC testing, which will also focus on converting the calcium sulfide generated in the carbonizer to calcium sulfate, is planned to commence in February 1993 and be completed by August 1993.

With regard to topping combustor development, no testing occurred during this reporting period. Analysis of previously collected test data has revealed the need to redesign the combustor primary premix rich-burn zone to minimize the conversion of carbonizer fuel gas ammonia to NO_x . In September Westinghouse PGBU was released to proceed with this redesign, and testing of the modified combustor should be completed by the end of 1993.

6.2.2 Management Activities

The following second-generation PFB-related papers were written and presented:

■ Second-Generation PFB Pilot Plant Start-Up	Eighth Annual Pittsburgh Coal Conference	October 1991
■ Second-Generation Pressurized Fluidized Bed Combustion Pilot Plant Construction	Fifth Annual Fluidized Bed Combustion Conference	December 1991
■ Second-Generation Pressurized Fluidized Bed Combustion	Energy and Environment: Transitions in Eastern Europe	April 1992
■ Second-Generation PFBC Systems Research and Development—Phase 2	Gasification and Gas Stream Cleanup Systems Contractors Review Meeting	September 1992
■ Electric Utility Second-Generation Pressurized Fluidized Bed Combustion Plants	EPRI—Application of Fluidized Bed Combustion for Power Generation Utility Conference	September 1992

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