1 of 1

**Second-Generation Pressurized Fluidized Bed
Combustion Plant -- Conceptual Design and
Optimization of a Second-Generation PFB Combustion Plant**

**Annual Report
October 1, 1991 - September 30, 1992**

**A. Robertson
W. Domeracki
R. Newby
A. Rehmat
D. Horazak**

October 1992

Work Performed Under Contract No.: DE-AC21-86MC21023

For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia

By
Foster Wheeler Development Corporation
Livingston, New Jersey

MASTER

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12 Peach Tree Hill Road
Livingston, New Jersey 07039**

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, is to be built 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 (PGBU)
- Westinghouse Electric Corporation, Science & Technology Center (S&TC)

has embarked upon a three-phase 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, hot electrostatic precipitators, 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

Figure 1.1 14-atm Second-Generation PFB Combustion Plant Heat and Material Balance (1600°F Carbonizer Case)

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 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 a carbonizer and a CPFBC, together with their ceramic barrier filters, will be designed, constructed, and tested at a 1-MWe (equivalent) scale as a fully integrated subsystem at the FWDC John Blizzard Research Center in Livingston, New Jersey. This effort will allow component and integrated subsystem performance data to be collected and proposed plant operating modes/philosophies investigated. A topping combustor integrated with a carbonizer and PFB combustor will also be tested in Phase 3, either at FWDC or an alternative site to be identified later. 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 3 SCHEDULE AND OBJECTIVES

Figure 1.2 is the schedule and milestones originally planned for Phase 3. This schedule assumed that Phase 3 would start shortly after the completion of Phase 2. In an effort to expedite Phase 3, work actually began before Phase 2 ended. However, actual construction for Phase 3 cannot begin until Phase 2 testing has been completed. The objective of each of the work

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Milestones:

1. Submit test plan for approval and review conceptual design with DOE.	5. DOE issues release to fabricate.	11. Testing completed
2. Design review meeting and Foster Wheeler requests permission to procure critical items.	6. Pressure vessels received.	12. Data reduction completed.
3. DOE releases procurement of critical items.	7. Computer software package completed.	13. Submit topical reports for approval.
4. Submit design package for approval.	8. Submit test plan for approval.	14. Issue final reports for approval.
	9. Begin testing.	15. DOE approves final report.
	10. Begin preparation of final project report.	

Figure 1.2 Program Schedule—Phase 3

elements shown is discussed in this section. Section 2 of this report contains a discussion of the technical work performed by the team during the period shown on the report cover.

Task 1—Test Plan

Objective: To prepare a detailed test plan for integrated subsystems testing that specifies how the facility will be able to produce data that addresses R&D technology gaps.

Workscope:

- Prepare specifications for an Integrated Subsystems Test Facility.
- Define specific tasks and decision points for Phase 3 activities.
- Estimate cost and schedule of Phase 3 activities.
- Prepare a comprehensive Phase 3 test plan that contains all these elements and submit it to the DOE COTR for approval no later than 60 days after notice to proceed to Phase 3.

Task 2—Design

Objective: To submit a detailed design for the proposed Integrated Subsystems Test Facility.

Subtask 2.1—Conceptual Design:

Workscope:

- Prepare heat and material balances.
- Prepare P&I diagrams.
- Prepare outline drawings for all major equipment.
- Identify interfaces for equipment to be supplied by team members.

Subtask 2.2—Detailed Design

Workscope:

- Prepare thermal, hydraulic, and mechanical designs of all components and systems.

- Prepare assembly and subassembly detailed drawings for all components to be manufactured.
- Prepare test facility arrangement drawings.
- Identify instrumentation, controls, and balance-of-plant equipment.
- Prepare specifications and obtain quotations for equipment to be procured.
- Prepare schedules and cost estimates.
- Prepare and submit a design package to DOE.

Task 3—Construction and Shakedown

Objective: To construct a 1-MWe (equivalent) Integrated Subsystems Test Facility capable of meeting the test objectives of Phase 3 defined in the Task 1—Test Plan and to prepare it for operation.

Subtask 3.1—Construction

- Fabricate components unique to this facility (carbonizer, CPFBC, and topping combustor).
- Procure balance of equipment, instruments, and controls.
- Erect the test facility.

Subtask 3.2—Shakedown

Workscope:

- Calibrate equipment and check to ensure its ability to operate under all modes of operation as defined in Task 1—Test Plan and updated in Task 4.
- Train facility operators.
- Operate facility to verify that performance meets design performance goals.

Task 4—Test Plan Implementation

Objective: To obtain the data defined in the DOE-approved Phase 3, Task 1 Test Plan.

Workscope:

- Revise the Test Plan to incorporate changes.

- Perform all tests in accordance with the approved Test Plan.
- Analyze test data.

Task 5—Phase 3 Topical Report

Objective: To deliver to DOE a Topical Report covering the results of Phase 3 testing.

Workscope:

- Prepare a Topical Report detailing the results of Phase 3 testing and tying together the Test Plan, modifications, execution, and results.
- Submit the report in draft form to the DOE COTR for approval at least 30 days before the completion of Phase 3.

Task 6—Revised Conceptual Design

Objective: To deliver to DOE a Topical Report updating Phase 1, Task 1 results to incorporate the results of Phases 2 and 3

Subtask 6.1—Revisions to Conceptual Design

Workscope:

- Revise the conceptual design to incorporate the results of the component and subsystem testing (Phases 2 and 3).

Subtask 6.2—Conceptual Design Topical Report

Workscope:

- Prepare and deliver a Topical Report on the revised conceptual design.

Task 7—Program Management/Reporting

Objective: To provide overall guidance and control of the test program 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 proposal ground rules (FWDC).

Section 2

TASK 1—TEST PLAN

Section 2

TASK 1--TEST PLAN

2.1 OBJECTIVE

The Task 1 objective is to develop a plan for testing an integrated carbonizer and CPFBC system. In Phase 2 the key components are being tested separately to determine their individual characteristics. In Phase 3 the equipment in the carbonizer flow train and the CPFBC flow train will be operated as an integrated system. Phase 3 testing will fulfill several R&D needs. It will demonstrate the ability of the system to safely and reliably transfer hot solids from the carbonizer and its HGCU to the CPFBC. It will also demonstrate the ability to burn this char and circulate the solids within the CPFBC, hot recycle cyclone, and external FBHE loop at smooth, responsive, and controllable rates. The pilot plant will be operated over a range of excess air levels that will simulate the complete range of commercial plant operation.

2.2 STATUS

In 1991 a Test Plan, together with an update reflecting the use of a coal/water/sorbent slurry feed system was submitted to METC for approval. It details the measurements and the test matrix for the Phase 3 pilot plant.

Figure 2.1 is a schematic diagram of the pilot plant. The Phase 2 coal and sorbent feed systems are reused in Phase 3; however, modifications will be made to the coal and sorbent feed hopper outlets. There will be two outlets on each feed hopper. One outlet can be used for the pressurized pneumatic feeding of sorbent and coal to the carbonizer. This system is used during start-up to load sorbent into the carbonizer bed and then to the CPFBC bed. It is also used when burning lignite. The other feed hopper outlets are used during normal operation (with Pittsburgh No. 8 coal as the fuel) to feed dolomite and coal to the slurry system. Slurry and air are fed to the carbonizer.

The fuel gas generated in the carbonizer is cleaned of particulates in the primary cyclone and cross-flow filter. An alkali getter vessel downstream of the cross-flow filter removes alkalis. A choked-flow orifice reduces the system pressure to essentially ambient. The fuel gas is then cooled in a spray tower, flows through a demister, and flows through a baghouse for final particulate removal. The fuel gas is burned in an incinerator, and the products of combustion go up the stack. The carbonizer and all the equipment in the fuel gas flow stream, with the exception of the alkali getter vessel, is Phase 2 equipment that is reused.

Figure 2.1 Integrated Carbonizer/CPFBC Pilot Plant

The char and spent sorbent from the carbonizer overflow, primary cyclone dip leg, and cross-flow filter are collected in the carbonizer collecting hopper and then transported to the CPFBC. The char is burned in the CPFBC and coal/sorbent slurry can be injected directly into the CPFBC to vary its excess air from a maximum of 185 percent to a minimum of 30 percent. The CPFBC will be a new vessel.

The CPFBC and FBHE comprise a circulating fluidized bed system with the heat load being absorbed in the FBHE water-cooled tube bundle. The heat load varies, depending upon the test matrix operating conditions. The bed height in the FBHE, and hence amount of tube surface immersed in the bed, will be adjusted to obtain the required heat absorption. Flue gas leaving the FBHE cyclone goes to a candle filter for particulate removal. A choked-flow orifice then provides the pressure reduction, and a spray tower cools the flue gas stream. It then passes through a demister and a baghouse. The CPFBC gas stream uses the same stack as the carbonizer gas stream. The flue gas from the CPFBC enters the stack downstream of the incinerator. Most of the equipment in the CPFBC gas train will be new.

Second-generation PFB combustion plants can be designed to operate over a wide range of conditions. There are two basic ways the output of a second-generation PFB combustion plant can be changed. One is to reduce the fuel gas generation rate by reducing the coal and airflow to the carbonizer. This method results in lower system pressure and lower duty in the gas turbine and steam turbine cycles. This type of output change will be referred to as load change, and it is one of the means by which a commercial second-generation PFB combustion plant can load follow.

In addition to this load-changing capability, the second-generation system has the capability for another mode of output variation. Coal can be fired in the CPFBC along with the char from the carbonizer, and the level of this coal firing can be varied. The heat and material balance that results from firing char (without coal) in the CPFBC has relatively high excess-air levels. Therefore, additional coal can be fired in the CPFBC, reducing the excess air level without significantly changing the gas flow rate to the gas turbine. The additional heat input to the CPFBC is transferred to the steam cycle in the FBHE, thereby increasing the duty to the steam cycle. This type of output change will be referred to as a change in excess air level; it is another technique by which a second-generation PFB combustion plant can load follow.

Figure 2.2 illustrates the range of operation that can be achieved in commercial second-generation PFB combustion plants by changing the CPFBC excess air level and topping combustor outlet temperature (fuel gas generation rate). This plot is based on a 1500°F carbonizer. Second-generation PFB combustion plants can be designed to operate anywhere in this envelope. A particular plant can also operate at different points within the envelope, although design considerations would make it impractical for one plant to be capable of operation within the entire envelope.

Figure 2.2 PFBC Excess Air vs. Topping Combustor Outlet Temperature as a Function of Plant Operating Configuration

A test matrix consisting of six separate test series has been developed to fulfill the objectives of the Phase 3 R&D program. The tests are designed to explore the operating envelope of the plant. Process data will be obtained under different operating conditions, and operating experience will be gained when changing from one operating condition to another. The first five tests will be run with Pittsburgh No. 8 coal, and the last test will be run with lignite. The Pittsburgh No. 8 tests will be with a slurry feed, but the lignite tests will be with a dry (pneumatic) feed. Summaries of the test series follow:

- **Test Series A**—Plant operation will be simulated at full load, and the excess air level will be varied from 185 percent (the best efficiency point of the plant with slurry feed) to 30 percent (the minimum excess air point).
- **Test Series B**—The pilot plant will be operated at pressure levels ranging from minimum- to full-load operation of the commercial plant. Two-load change sequences will be executed during this test series. One sequence will be at CPFBC excess-air levels that correspond to best-efficiency-point excess-air levels. The other sequence will be at the minimum excess-air level (30 percent).
- **Test Series C**—The pilot plant uses a choked-flow orifice to simulate the pressure-flow characteristics of the gas turbine. Because of this situation, the reduced-load operating points of Test Series B do not exactly match the commercial plant values. In this test series, a new choked-flow orifice will be installed to more accurately match the commercial plant gas flow rate and pressure characteristics. The excess-air level at this minimum load will then be varied from 185 to 30 percent.
- **Test Series D**—During this test series, the pilot plant will be operated for an extended period at steady-state conditions. The plant will be operated at the full-load best-efficiency point for 10 days. Data taken during this period will indicate any gradual changes in process parameters or the formation of agglomerates.
- **Test Series E**—During this test series, the fuel will be lignite. It is not practical to feed lignite as a slurry because of its already high level of moisture. The lignite and sorbent will be pneumatically fed to the carbonizer. The CPFBC will be fired with only the char produced in the carbonizer. The test will be run at full-load pressures, and the excess-air levels will range from 184 to 167 percent.

The specific operating parameters for Test Series A, B, C, and E are listed in Tables 2.1 through 2.4. Measurement locations in the pilot plant are indicated in Figure 2.3, and described in Table 2.5.

Car			
Test Point	Pressure (psia)	Bed Temperature (°F)	Slurry Flow† (lb/h)
A1	212	1600	824
A2	212	1600	824
A3	212	1600	824
A4	212	1600	824
A5	212	1600	824
A6	212	1600	824
A7	212	1600	824

C					
Test Point	Pressure (psia)	Excess Air (%)	Char/Sorbent* Flow (lb/h)	Slurry Flow† (lb/h)	Pr Air (lb/h)
A1	192	185	323	247	
A2	192	105	323	472	
A3	192	30	323	906	
A4	192	105	323	472	
A5	192	185	323	247	
A6	192	105	323	472	
A7	192	30	323	906	

*Based on a Ca/S feed ratio of 1.75. Values will be adjusted as required to yield desired sulfur-capture

†Nitrogen added just upstream of choked orifice to reduce fuel gas flow.

§Does not include nitrogen purges.

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Table 2.1 Test Series A

Ionizer			
Airflow (lb/h)	Fuel Gas Flow [§] (lb/h)	Char/Sorbent* Flow (lb/h)	Air Inlet Temperature (°C)
953	1454	323	711
953	1454	323	711
953	1454	323	711
953	1454	323	711
953	1454	323	711
953	1454	323	711
953	1454	323	711

FBCE					
Primary Flow (lb/h)	FBHE Secondary Airflow (lb/h)	Bypass Secondary Airflow (lb/h)	Air Temperature (°F)	Flue Gas [§] Flow Rate (lb/h)	FBHE Heat Load (10 ⁶ Btu/h)
690	2142	4750	711	10,000	0.58
664	2142	3584	711	10,000	2.10
549	2142	1327	711	10,000	5.05
664	2142	3584	711	10,000	2.10
690	2142	4750	711	10,000	0.58
664	2142	3584	711	10,000	2.10
549	2142	1327	711	10,000	5.05

efficiencies.

Carbon			
Test Point	Pressure (psia)	Bed Temperature (°F)	Slurry Flow (lb/h)
B1	212	1600	824
B2	212	1600	824
B3	160	1600	567
B4	160	1600	567
B5	150	1600	318
B6	150	1600	318
B7	160	1600	567
B8	160	1600	567
B9	212	1600	824
B10	212	1600	824

CPFI					
Test Point	Pressure (psia)	Excess Air (%)	Char/Sorbent* Flow (lb/h)	Slurry Flow (lb/h)	Prim Airf (lb/h)
B1	192	185	323	247	269
B2	192	30	323	906	554
B3	154	160	222	285	234
B4	154	30	222	767	444
B5	144	160	125	360	218
B6	144	30	125	810	414
B7	154	160	222	285	234
B8	154	30	222	767	444
B9	192	185	323	247	269
B10	192	30	323	906	554

*Based on a Ca/S feed ratio of 1.75.

†Nitrogen added just upstream of choked orifice to reduce fuel gas flow.

§Does not include nitrogen purges.

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Table 2.2 Test Series B

izer			
Airflow (lb/h)	Fuel Gas Flow (lb/h)	Char/Sorbent* Flow (lb/h)	Air Inlet Temperature (°F)
953	1454 [†]	323	711
953	1454 [†]	323	711
655	1000 [†]	222	711
655	1000 [†]	222	711
368	561 [†]	125	711
368	561 [†]	125	711
655	1000 [†]	222	711
655	1000 [†]	222	711
953	1454 [†]	323	711
953	1454 [†]	323	711

C					
ry ow n)	FBHE Secondary Airflow (lb/h)	Bypass Secondary Airflow (lb/h)	Air Temperature (°F)	Flue Gas Flow Rate _s (lb/h)	FBHE Heat Load (10 ⁶ Btu/h)
0)	2142	4750	711	10,000	0.58
3)	2142	1327	711	10,000	5.05
3)	1718	3568	637	8,021	0.65
4)	1718	1060	637	8,021	3.92
3)	1606	3318	615	7,500	0.70
0)	1606	981	615	7,500	3.76
3)	1718	3568	637	8,021	0.65
4)	1718	1060	637	8,021	3.92
0)	2142	4750	711	10,000	0.58
3)	2142	1327	711	10,000	5.05

Carbonization				
Test Point	Pressure (psia)	Bed Temperature (°F)	Slurry Flow (lb/h)	
C1	150	1600	318	
C2	150	1600	318	
C3	150	1600	318	
C4	150	1600	318	
C5	150	1600	318	
C6	150	1600	318	
C7	150	1600	318	

CPFBC					
Test Point	Pressure (psia)	Excess Air (%)	Char/Sorbent* Flow (lb/h)	Slurry Flow (lb/h)	Primary Airflow (lb/h)
C1	144	160	125	470	2664
C2	144	120	125	574	3116
C3	144	70	125	767	3955
C4	144	30	125	1017	5040
C5	144	70	125	767	3955
C6	144	120	125	574	3116
C7	144	160	125	470	2664

*Based on a Ca/S feed ratio of 1.75.

§Does not include nitrogen purges.

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Table 2.3 Test Series C

Airflow (lb/h)	Fuel Gas Flow [§] (lb/h)	Char/Sorbent* Flow (lb/h)	Air Inlet Temperature (°F)
368	561	125	711
368	561	125	711
368	561	125	711
368	561	125	711
368	561	125	711
368	561	125	711
368	561	125	711

	FBHE Secondary Airflow (lb/h)	Bypass Secondary Airflow (lb/h)	Air Temperature (°F)	Flue Gas Flow Rate [§] (lb/h)	FBHE Heat Load (10 ⁶ Btu/h)
	1606	4389	615	9142	0.96
	1606	3848	615	9142	1.67
	1606	2944	615	9142	2.98
	1606	1545	615	9142	4.69
	1606	2944	615	9142	2.98
	1606	3848	615	9142	1.67
	1606	4389	615	9142	0.96

Carbon				
Test Point	Pressure (psia)	Bed Temperature (°F)	Lignite Flow (lb/h)	
E1	212	1600	862	
E2	160	1600	650	
E3	150	1600	610	
E4	160	1600	650	
E5	212	1600	862	

CPFB					
Test Point	Pressure (psia)	Excess Air (%)	Char/Sorbent* Flow (lb/h)	Lignite Flow (lb/h)	Primary Airflow (lb/h)
E1	192	184	362	0	1846
E2	154	171	273	0	1393
E3	144	167	256	0	1307
E4	154	171	273	0	1393
E5	192	184	362	0	1846
E6	154	171	273	0	1393
E7	144	167	256	0	1307

*Based on a Ca/S feed ratio of 1.0.

†Nitrogen added just upstream of choked orifice to reduce fuel gas flow.

§Does not include nitrogen purges.

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Table 2.4 Test Series E

er			
Airflow (lb/h)	Fuel Gas Flow [§] (lb/h)	Char/Sorbent* Flow (lb/h)	Air Inlet Temperature (°F)
948	1448	361	711
716	1098	273	711
671	1025	256	711
716	1098	273	711
948	1448	361	711

y w)	FBHE Secondary Airflow (lb/h)	Bypass Secondary Airflow (lb/h)	Air Temperature (°F)	Flue Gas Flow Rate [†] (lb/h)	FBHE Heat Load (10 ⁶ Btu/h)
	2142	2576	711	6772	0
	1718	1609	711	4878	0
	1606	1455	711	4515	0
	1718	1609	711	4878	0
	2142	2576	711	6772	0
	1718	1609	711	4878	0
	1606	1455	711	4515	0

Figure 2.3 Phase 3 Integrated Carbonizer/CPFBC Pilot Plant

Table 2.5 Phase 3 Integrated Carbonizer/CPFBC Pilot Plant

Description	Stream No.	Flow, lb/h	Temperature, °F	Pressure, atm/ Δ P	Analysis (Chemical/Physical)
Carbonizer					
Air Stream	1,4	■	■	■	
Sorbent	2	■	■	■	□
Coal	3	■	■	■	□
N ₂ Purges	5,12,15,52,53	■	■	■	
Bed T and Δ Ps	6,7		■	■	
Freeboard Gas	8		■	■	□
Char Samples	9,10	□	□		□
Alkali	13,14,48				□
Water Quench	16	■	■		
Tar/Oil	17	□	□		
Demisted Product Gas	11		■	■	
Baghouse Fines	18	□			□
Incinerator Air and CH ₄	19,20	■	■	■	
Stack Gas	21		■		■
Quenched Fuel Gas	36	■	■	■	■
CPFBC and FBHE					
Air Streams	22,25,30,31	■	■	■	
Sorbent	23	■	■	■	□
Coal	24	■	■	■	□
Char Feed to CPFBC	26		■	■	□
Primary and Secondary Zone T and Δ Ps	27,29		■	■	
N ₂ Purges	28,38,39,40	■	■	■	
CaSO ₄ /Ash Product	32	□	□		□
FBHE Cooling Water	33,34	■	■		
FBHE Exit Air	35		■		□
Bypass Secondary Air	31		■	■	
Gas Particulate Samples	41,42	□		■	
Candle Filter Fines	43	□	□		□
Water Quench	44	■	■		
Baghouse Fines	45	□			□
Stack Gas	21		■	■	■
Quenched Flue Gas	46		■	■	■
Circulating Solids	50		■	■	
■ Continuous Measurement □ Intermittent Measurement					

Section 3

TASK 2—DESIGN

Section 3

TASK 2-DESIGN

3.1 OBJECTIVE

The Task 2 objective is to design the pilot plant, including component process vessels, interconnecting piping, and instrumentation.

3.2 PLANT DESCRIPTION

3.2.1 Overview

The general arrangement of the plant is shown in Figures 3.1, 3.2, and 3.3. Most of the equipment will be located in or adjacent to the Phase 2 pilot plant. The roof will be raised to 67 ft in the new bay, where the carbonizer, CPFBC, and FBHE are located. This roof elevation will allow the CPFBC and carbonizer height to be raised above the Phase 2 dimensions. The Phase 3 carbonizer and CPFBC will be 5 and 10 ft taller respectively.

The equipment from Phase 2 is being reused in Phase 3 to the greatest extent possible. The carbonizer train, consisting of the carbonizer and the equipment in the fuel gas stream, will utilize equipment from Phase 2 except for an alkali-removal vessel added downstream of the cross-flow filter. The coal and sorbent silos, conveyors, and feed vessels will be reused from Phase 2 with minor modifications. The CPFBC train will be mostly new equipment with the exception of the FBHE, which will be the Phase 2 FBHE modified for the additional heat-load requirements of Phase 3. The system that transfers char and sorbent from the carbonizer to the CPFBC will consist of all new equipment (this function was not part of the Phase 2 pilot plant). The slurry-coal/water paste feed system is also new.

Estimated pressure and temperatures throughout the pilot plant are shown in Figure 3.4. These values are based on full-load operation with a 1600°F carbonizer.

3.2.2 Carbonizer Train

The arrangement of the equipment in the carbonizer train was shown in Figures 3.1 and 3.2. The carbonizer dimensions are shown in Figure 3.5. This is the same vessel that was used for the carbonizer and CPFBC in the Phase 2 pilot plant. The internal ceramic sleeves added in Phase 2 to modify the vessel from a 10-in.-dia. carbonizer to an 8-in.-dia. CPFBC will be removed so that the vessel can be used as a carbonizer again for Phase 3. A section will also be added to increase the carbonizer bed height by 5 ft, to the expected commercial plant value of approximately 25 ft.

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Figure 3.1 Pilot Plant General Arrangement--Plan View

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ELEVATION B-B

Figure 3.2 Pilot Plant Carbonizer Configuration--Elevation View

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ELEVATION D-D

Figure 3.3 Pilot Plant CPFBC Configuration--Elevation View

COAL - SORBENT -

		1	2	3	4	5	
CARBONIZER	P(psia) / T(F)	218/1600	212/1581	211.9/1541	211.6/1509	211.6/1487	209
CPFBC	P(psia) / T(F)	196/1600	192/1600	192.0/1598	190.6/1589	190.3/1571	187

WATER SLURRY FEED

6	6a	6b	7	8	9	10	11
6/1487	209.6/1471	208.6/1451	208.4/1378	15.7/392	15.6/386	15.4/382	15.3/370
3/1566	————	————	187.2/1562	15.5/390	15.3/384	15.1/380	15.0/370

Figure 3.4 Phase 3 Pressures and Temperatures (Design Point)

Figure 3.5 10-in. Carbonizer Test Unit-Phase 3

The superficial velocity in the carbonizer will be 3 ft/s in the 10-in.-I.D. section under most of the test conditions. The carbonizer heat and material balance at 1600°F with Pittsburgh No. 8 coal is shown in Figure 3.6. A 10-in.-I.D. with a 3 ft/s superficial velocity and the expected conversion ratio results in a fuel gas flow rate of 1454 lb/h at full load. The corresponding char/sorbent flow rate is 323 lb/h. The system for the transfer of char/sorbent to the CPFBC is discussed in Subsection 3.2.3.

The fuel gas produced in the carbonizer is cleaned of particulates with a cyclone followed by a cross-flow filter. Both of these items are Phase 2 components that will be used without modification. The next component in the carbonizer train is the alkali-removal vessel. The gas passes over pellets of emathlite in the alkali getter vessel and reacts with these pellets to remove the alkalis from the fuel gas. An alkali-removal vessel was not part of the Phase 2 system, so it has been designed specifically for the Phase 3 conditions. A more detailed discussion of the alkali-removal vessel design is presented in Subsection 3.2.5.

Downstream of the alkali getter vessel, the fuel gas passes through a U-shaped piping loop that provides penetrations for gas sampling. This piping is reused from Phase 2. The fuel gas then enters the orifice tee, where a choked-flow orifice breaks the system pressure and establishes the pressure vs. gas flow rate characteristics of the carbonizer train. The tee section that houses the orifice is reused from Phase 2.

The spray tower and demister for the fuel gas stream can also be seen in Figure 3.2. Water spray in the cooling tower cools the fuel gas to approximately 400°F before it enters the baghouse. The demister ensures that no water droplets enter the baghouse. The spray tower, demister, and baghouse are all Phase 2 components that are used without modification. The fuel gas then goes to the incinerator, where it is burned before entering the stack. The incinerator and stack are from Phase 2, although modifications will be made to inject the CPFBC flue gas stream into the stack.

3.2.3 Char/Sorbent Transfer System

One of the R&D needs that will be addressed in the Phase 3 program will be the transfer of the char/sorbent residue from the carbonizer to the CPFBC. This system must safely and reliably transfer these solids at controlled rates as the pressure differential between the two systems varies from approximately 6 lb/in² at 50-percent to 20 lb/in² at 100-percent load.

In the proposed design, the char and sorbent particles are collected at one point—the carbonizer collecting hopper—from the bottom and top of the carbonizer bed and from the primary cyclone and the cross-flow filter. The carbonizer overflow from the top of the bed is fed to the carbonizer collecting hopper through an inclined pipe. This arrangement effectively keeps the collecting hopper at the carbonizer freeboard pressure. The primary cyclone and cross-flow filter are thus at a lower pressure than the collecting hopper, and dip-leg sealing (shown in Figure 3.7) is required.

Char Sorbent
Flow Rate = 46.861 lb/h

Char	(32.401 lb/h)
Carbon	23.662
Hydrogen	0.294
Oxygen	0.422
Nitrogen	0.405
Sulfur	0.842
Chlorine	0
Ash	6.776
Moisture	0

HHV (Btu/lb) = 11,188.71
LHV (Btu/lb) = 11,107.62

Tar
Flow Rate = 0 lb/h

Atomic Composition
C (0)
H (0)
O (0)
N (0)
S (0)
Cl (0)

Gas
Flow Rate = 211.264 lb/h,
8.2853 mol/h

	wt%	mol%
CO	5.745	5.23
CO ₂	22.46	13.013
H ₂	0.609	7.703
H ₂ O	16.874	23.883
CH ₄	1.634	2.597
C ₂ H ₆	0.335	0.284
O ₂	0	0
N ₂	49.965	45.479
H ₂ S	0.145	0.109
COS	0.009	0.004
NH ₃	0.09	0.135
HCN	0	0
HCl	0	0
C ₂ H ₄	1.339	1.217
C ₂ H ₂	0	0
C ₃ H ₈	0.349	0.209
C ₄ H ₁₀	0.06	0.026
C ₆ H ₆	0.107	0.035
C ₇ H ₈	0.172	0.048
CH ₃ OH	0	0
C ₆ H ₅ OH	0.108	0.029

HHV (Btu/lb) = 1548.12
LHV (Btu/lb) = 1415.8

Spent Sorbent	(14.459 lb/h)
CaCO ₃	8.176
MgCO ₃	0
CaO	0
MgO	4.086
CaS	1.861
CaSO ₄	0
Inerts	0.336

HHV (Btu/lb) = 732.9192
LHV (Btu/lb) = 732.9192

HHV (Btu/lb) = 7962.50
LHV (Btu/lb) = 7908.43

Pittsburgh Coal
Flow Rate = 100 lb/h (wet)

Moisture-Free	wt%
Carbon	73.79
Hydrogen	4.81
Oxygen	6.47
Nitrogen	1.29
Sulfur	3.07
Chlorine	0
Ash	10.57

Moisture in Coal (wt%) = 35.89
HHV (Btu/lb) = 8492.651 (wet)
LHV (Btu/lb) = 7858.178 (wet)

	lb/h
O ₂	31.97
N ₂	105.294
Moisture	1.121

	wt%
CaCO ₃	54.79075
MgCO ₃	43.00074
Moisture	0.5
Inerts	1.7085

Figure 3.6 14 atm/1600°F Carbonizer Slurry Feed Balance

Figure 3.7 HGCU Dip-Leg and N-Valve/Slide Valve Char/Sorbent Transfer System

The dip leg from the primary cyclone is sealed from the cross-flow filter pressure with a seal pot. The solids collected in the primary cyclone flow through the seal pot and then fall through an inclined pipe into the cross-flow filter dip leg. The mixing of the two solids flow streams will reduce the tendency for the cross-flow filter fines to bridge, because these particles will mix with the coarser particles from the primary cyclone. The pressure differential between the cross-flow filter and the carbonizer collecting hopper is sealed with a J-valve.

Residue drained from the bottom of the bed will be lifted up into the collecting hopper via the N-valve arrangement shown in Figure 3.8.

The carbonizer collecting hopper performs the function of combining the four char/sorbent streams. The storage capacity of the collecting hopper also acts as a buffer so that effects of any upsets in the char/sorbent flow from the carbonizer can be controlled and minimized. At the full load char/sorbent flow rate of 323 lb/h, the carbonizer collecting hopper will contain approximately a 3-hour supply of char/sorbent. A nuclear level device will measure the solids level in the collecting hopper.

The char/sorbent flow rate to the CPFBC is controlled downstream of the carbonizer collecting hopper. A system similar to that shown in Figure 3.7 is proposed for the commercial plant. This arrangement uses a slide valve and an N-valve to control the solids flow rate and dissipate the pressure difference between the carbonizer and the CPFBC. The carbonizer is 6 to 20 psi higher in pressure, depending on the plant load. With the arrangement shown in Figure 3.7, the slide valve position, nitrogen stripping flow rate, and riser fluidizing nitrogen flow rate can be varied to control the solids flow rate and pressure dissipation.

There is concern that, at the relatively low solids flow rates of the pilot plant, the slide valve will need to be throttled to the point where bridging of the solids may occur.

Flow model tests conducted by IGT at the anticipated pilot plant pressures with char/sorbent material obtained from the Phase 2 carbonizer tests should ascertain the performance characteristics of the design shown in Figure 3.7. A test plan describing the flow model tests in detail was approved by DOE METC. Solids flow rates from 50 to 600 lb/h will be tested with different solids mixtures. The solids mixtures will be varied by changing the proportions of carbonizer, cyclone, and cross-flow filter material. Flow model tests have begun and are described in Section 3.3.4.

3.2.4 CPFBC Train

The Phase 3 CPFBC will be a totally new vessel. It has been made larger than the Phase 2 vessel to accommodate the entire char/sorbent residue generated by the carbonizer. The CPFBC vessel is shown in Figure 3.9 along with the FBHE. It has a 13-in. I.D. that yields a 12 ft/s velocity when operated with a flue gas flow rate of 10,000 lb/h at full-load conditions (14 atm/

Figure 3.8 N-Valve Arrangement for Transferring Carbonizer Bed Drains to Char-Collecting Hopper

Figure 3.9 CPFB/FBHE

1600°F). This gas flow rate corresponds to dry feed of Pittsburgh No. 8 coal to the carbonizer, with the resultant char/sorbent residue being fired alone in the CPFBC. When a slurry feed is used, a greater amount of the coal is consumed in the carbonizer; hence the char/sorbent residue flow to the CPFBC is reduced. As a result, slurry will be fed to the CPFBC to make up this shortfall rather than reduce the size of the CPFBC.

Although the capacity of the CPFBC could be reduced with slurry feed, we decided to keep the capacity at the dry feed level for maximum pilot plant operating flexibility—it is capable of both dry and slurry feed operation. In addition, the larger Phase 3 unit allows for a limited analysis of scaling effects (Phase 2 vs. Phase 3).

The Phase 2 FBHE vessel will be modified for the higher heat loads and solids circulation rates required in Phase 3. The maximum FBHE heat duty in Phase 3 is approximately three times the maximum Phase 2 duty. A tube bundle with more surface area will replace the Phase 2 bundle. A larger tube bundle will be used for the ash cooler because of the greater solids throughput in Phase 3. The design solids circulation rate for Phase 2 was 40,000 lb/h. In Phase 3 solids circulation rates of up to 100,000 lb/h are planned. This increase required 8-in.-dia. L- and J-valves instead of the 6-in.-dia. Phase 2 components. The cyclone will also be new because of the higher gas flow and solids circulation rates.

Ceramic candle filters and ceramic cross-flow filters are the two leading types of barrier filters being developed for PFB combustion cycles. Since the cross-flow filter was used with the Phase 2 carbonizer, rather than risk the perception of "marrying" second-generation PFB combustion technology to cross-flow filters, the CPFBC has been provided with a ceramic candle filter. In any event, the pressure vessels of both filter units are large enough to accommodate either technology.

The flue gas exits from the FBHE cyclone and goes to the CPFBC barrier filter for final filtering of particulates. The detailed design of the CPFBC barrier filter is discussed in Subsection 3.2.6. Downstream of the CPFBC barrier filter, the flow goes through a choked-flow orifice, spray tower, and demister. The design and operation of this equipment is the same as for the analogous equipment in the carbonizer train; however, the equipment is larger because of the higher gas flow rate. The CPFBC spray tower and demister are shown in Figure 3.10. After the CPFBC baghouse, the flue gas stream enters the stack downstream of the incinerator. The stack and incinerator are the same as those in Phase 2.

The stack emissions for Phase 3 were calculated, and modeling was performed to predict grade-level emissions concentrations. The results indicate that the Phase 2 already-permitted emissions and ground level concentrations will not be exceeded.

Figure 3.10 CPFBC Spray Tower

3.2.5 Alkali-Removal Vessel

The design philosophy of the alkali-removal vessel is to incorporate as many commercial design features into the Phase 3 unit as are feasible within the constraints of gas flow and space imposed by the test facility. This philosophy ensures that the Phase 3 alkali-removal testing will provide results significant for commercial development. The design will apply gas velocities, general vessel configuration, and pellet sizes similar to those that have been proposed in the Phase 1 commercial design study. Bed depths are limited by pilot plant physical constraints, and the alkali-removal bed life will be relatively short (500 hours at nominal conditions), but compatible with the test program needs. Because the vessel is small, the design of the gas distribution features within the vessel cannot be prototypical of commercial practice. The design basis for the alkali-removal vessel is specified in Table 3.1. The vessel has been designed for an alkali outlet content of less than 50 ppm(v) and particulate entrainment from the alkali getter bed of less than 15 ppm.

Table 3.1 Alkali Removal Vessel Design Basis

Gas inlet temperature, °F	1471
Maximum inlet temperature, °F	1559
Inlet pressure, lb/in ² a	212
Gas molecular weight	26.1
Gas density, lb/ft ³	0.267
Gas mass flow, lb/h	1934
Gas volumetric flow, aft ³ /min	121
Gas environment	Reducing
Particulate	Coal char and partially sulfided lime
Inlet fly ash loading, ppm(w)	15
Maximum inlet alkali, ppm(v)	63 (total Na plus K vapor)
Pellet size, in.	1/4 x 1/2 cylinders
Pellet material	Emathlite

The vessel shell is 0.5-in.-thick carbon steel (Figure 3.11). To insulate the vessel wall from the process temperature, 8 in. of cast refractory material is applied to the inside. Flat heads are placed at both the top and bottom of the vessel for purposes of compactness and to ease pellet loading, unloading, and removal. Gas enters the vessel near the top, through a nozzle with a 3-in. I.D. opening. It passes downward through the packed bed of 1/4 in. $\times$ 1/2 in. pellets and is discharged through the 3-in. outlet nozzle near the base of the vessel. To prevent entrainment of pellets or plugging of the outlet, the outlet nozzle is covered by a high-alloy perforated plate/screen.

The active bed depth is about 50 in. The maximum superficial velocity is 0.85 ft/s, with a minimum gas residence time in the active bed of about 4.9 seconds. The bed life before alkali breakthrough is about 500 hours at maximum flow conditions; however, the bed will be recharged more frequently. Recharging will be done at intervals that will yield information on pellet conversion under different operating conditions.

Thermodynamic estimates were made of the alkali content of the carbonizer products that could result during Phase 3 testing. The carbonizer material balances predicted by FWDC for coal slurry feeding with 1500 and 1700°F carbonizer temperatures were the basis for these estimates. The Pittsburgh No. 8 coal and Plum Run dolomite alkali contents used in the Phase 1 system study were applied in these estimates. In addition to estimating the alkali species and concentrations emitted from the carbonizer, the influence of temperature losses on the alkali species and concentrations as the gas passes farther downstream were also estimated. These estimates are tabulated in Table 3.2 for vapor species (g) and solid and liquid species (c).

With the 1500°F carbonizer, the predicted vapor-phase alkali content in the gas is high in both NaCl and KCl. If the gas temperature falls to 1400°F, the condensation of NaCl is significant, but a large amount of NaCl vapor remains, and KCl is not influenced. If less-than-equilibrium amounts of NaCl actually exist in the gas from the carbonizer, the fractional loss by condensation will be less than shown in the table. Temperature loss to 1300°F results in significant loss of KCl vapor. Similar results are obtained for the 1700°F carbonizer, with vapor losses occurring only when the gas temperature is reduced to 1400°F. Alkali measurements should be taken as close to the carbonizer as possible to get that actual vapor content. The vapor contents are high enough that, even if vapor losses are significant before the alkali-removal vessel, the testing will still be a meaningful process demonstration of the alkali-removal capabilities.

3.2.6 CPFBC Candle Filter

General Specifications. The candle filter design incorporates as many commercial design features into the Phase 3 CPFBC unit as are feasible within the constraints of gas flow and space imposed by the test facility. This philosophy ensures that the Phase 3 filter testing results will be significant

Figure 3.11 Carbonizer Alkali Getter

Table 3.2 Phase 3 Carbonizer Fuel Gas Alkali Content Estimates

Alkali Component ppm(v)	Temperature, °F					
	1700	1600	1500	1400	1300	1200
Coal Slurry Carbonizer at 1500°F						
NaCl			44.4	17.7	4.0	0.7
KCl(g)			17.7	17.7	8.9	1.7
NaCl(c)			0	26.7	40.4	43.7
KCl(c)			0	0	8.8	16.0
NaOH(g)			0	0	0	0
KOH(g)			0	0	0	0
Coal Slurry Carbonizer at 1700°F						
NaCl	39.5	39.6	39.6	17.7	4.0	0.7
KCl(g)	15.8	15.8	15.8	15.8	8.9	1.7
NaCl(c)	0	0	0	21.9	35.6	38.9
KCl(c)	0	0	0	0	6.9	14.1
NaOH(g)	0.1	0	0	0	0	0
KOH(g)	0	0	0	0	0	0

for eventual filter commercialization. The filter conceptual design will apply scaled dimensions for many key components based on earlier detailed designs for filter vessels of similar size and design conditions. The design will incorporate candle filter elements, but the vessel is large enough to accommodate/interchange with cross-flow elements.

The filter system design is based on the 70:30 coal/sorbent/water slurry feed of the Phase 3 plant. Nominal, full-load filter operating conditions, design conditions, and performance to achieve commercial acceptance are presented in Table 3.3.

The features to be incorporated into the design are:

- 1.5-m candle filter elements
- Candles fabricated from Mullite (Coors)
- Individual candle mounts with fail-safe component
- Commercial candle gasket design
- Interchangeable candles and cross-flow elements
- Multiple filter element cleaning from single pulse source

Table 3.3 Full-Load Filter Operating Conditions, Design Conditions, and Performance

Full-Load Operation	
Gas inlet temperature, °F	1571
Inlet pressure	192
Gas molecular weight	29
Gas density, lb/ft ³	0.256
Gas mass flow, lb/h	11,390
Gas volumetric flow, aft ³ /min	742
Gas environment	Oxidizing
Particulate	Coal ash and partially sulfated lime
Fly ash loading, ppm(w)	10,000
Design	
Pressure, psig	270
Shell metal temperature, °F	40 to 650
Insulation type	Ceramic fiber
Cold face temperature, °F	165 (based on 1700°F internal)
Tubesheet design life	9000 hours at 1570°F with 200 start-up cycles
Performance	
Maximum pressure drop, lb/in ²	3
Temperature drop	Minimized
Maximum outlet loading, ppm(w)	15
Pulse cleaning frequency	<3 per hour per pulse valve

- Gas inlet shroud
- Flexible-joint tubesheet
- Blanket insulation with metal liner
- Redundant, automatic switching of pulse valves/on-line maintainability
- Venturi-assisted pulse cleaning
- Tangential gas inlet (for high particulate loadings)
- Split-ring plenum attachment

The design must meet all of the dimensional constraints imposed by the FWDC pilot plant layout, as incorporated into the Phase 3 conceptual design. Candle installation and maintenance must be performed directly above the

vessel with <11-ft headroom. In addition, the vessel must incorporate the pressure-relief philosophy imposed on the carbonizer filter in Phase 2.

A general arrangement drawing of the 6-ft-dia. x 18-ft-tall filter vessel is presented in Figure 3.12. The vessel shell is 0.875-in.-thick carbon steel. The vessel is insulated with 8 in. of refractory blanket material applied to the inside diameter. A metal liner provides an interface between the internal hot gas and the insulation. Fiber insulation reduces the weight of the filter vessel while providing greater insulating value.

Gas enters the vessel tangentially, with a transition from the 6-in.-I.D. inlet pipe to a rectangular slot that fits into the 4-in. shroud gap. The shroud is a metal cylinder that surrounds the bundle of filter elements to protect them from direct impingement by the high-velocity inlet gases and helps to distribute the gas within the vessel. The shroud extends above the chamber that holds the candle attachments and below the candle filter elements. The shroud gap is 4 in. and is not tapered at the base. This shroud design promotes gas flow both downward and upward to dissipate the inlet velocity and eliminates the possibility of fly ash accumulation within the shroud.

The tubesheet is 1.5-in.-thick high-alloy steel plate supported by an expansion web. The tubesheet supports the candle filter assembly. A splitting ring design couples the assembly with the tubesheet. The bundle contains forty-eight 1.5-m candle filter elements, with a nominal face velocity of about 7 ft/min. A truncated cone sits on the pulse-cleaning chamber to prevent the accumulation of large quantities of fly ash on top of the chamber. Four, straight pulse-cleaning lines—each with a venturi-pass through the top of the vessel—are directed into the four compartments of the pulse-cleaning plenum.

Estimated Phase 3 Candle Filter Performance. The Phase 3 candle filter performance has been estimated to select the design face velocity for the filter. The major performance parameters are the filter pressure drop, the pulse-cleaning frequency, and the thickness of the filter cake collected. Since the nature of the filter cake is unknown, two extremes are assessed—a high-permeability cake and a low-permeability cake. The permeabilities have been selected to cover the range that could exist based on past test experience. Figure 3.13 shows the behavior of the high-permeability cake and Figure 3.14, the low-permeability cake as a function of the face velocity. A face velocity of about 5.5 ft/min results in a nominal filter pressure drop of 3 lb/in² and an acceptable trade-off between pulse frequency and cake thickness.

Heat-loss estimates through the 8-in. fiber refractory were generated. The maximum vessel skin temperature is 165°F, assuming a gas temperature in the filter vessel of 1700°F. The maximum temperature loss in the CPFBC gas on passing through the filter vessel is 30°F.

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Figure 3.12 CPFBC Candle Filter

Figure 3.13 Phase 3 CPFBC Candle Filter High-Permeability Cake

Figure 3.14 Phase 3 CPFBC Candle Filter Low-Permeability Cake

3.2.7 Slurry System

A schematic of the proposed slurry/paste system is shown in Figure 3.15. Coal and sorbent, purchased crushed and screened with the proper size distribution, will be stored in the Phase 2 silos and transferred to their respective lock and feed hoppers. A side discharge nozzle will be added near the bottom of each of the two feed hoppers. The existing center discharge nozzle of each vessel will be used with the feed hoppers pressurized to pneumatically feed coal and sorbent to the carbonizer. During start-up, sorbent will be pneumatically injected into the carbonizer to build a bed that will overflow into the carbonizer-to-CPFBC char/sorbent transfer system and eventually build a bed in the CPFBC and FBHE.

The side discharge nozzles will feed the slurry system. These outlets will discharge to weigh-belt feeders that control the coal and sorbent mixture ratios. The coal and sorbent will then be transported to a blender/mixing tank; based on the solids flow rates, water will be added by a metering pump. Different systems are being investigated to transfer the coal and sorbent from the weigh feeders to the mixing tank. An important factor in the choice of this equipment is the available space, and layouts are being made to investigate the feasibility of using either a "Z" belt conveyor or a screw conveyor/bucket elevator arrangement.

A conical screw mixer will mix the slurry/paste, keep it agitated, and provide a 30-minute supply of slurry (based on maximum facility flow rate). A progressive cavity pump will maintain slurry flow in the lines leading to the injection pumps as well as recirculate material to the conical mixing tank. A density meter in this line will keep operating personnel apprised of the condition of the slurry. Both diaphragm and piston pumps are being investigated as candidates for injecting the slurry into the carbonizer and CPFBC.

A preliminary layout of the slurry feed system equipment has been developed, and it is shown in Figure 3.16. In this layout a "Z" belt conveyor transports the coal and sorbent from the weigh feeders to the mixing tank.

3.3 ACCOMPLISHMENTS

3.3.1 Pilot Plant

During this reporting period, the detailed design of the tower required to support and enclose the Phase 3 components and the detailed design of all of FWDC's Phase 3 pressure vessels continued to completion. Instrumentation nozzle locations were checked against structural steel locations to ensure that all instrumentation could be inserted and withdrawn without interferences. The final drawings were resubmitted to bidding vendors for best-and-final offers.

Figure 3.15 Schematic of Slurry Feed System

Figure 3.16 Slurry/Paste Feed System
Layout (Preliminary)

In December 1991 the Livingston Planning Board granted FWDC a variance to construct the 67-ft tall tower required for Phase 3. An environmental assessment of the impact of expanding the Phase 2 Pilot Plant for Phase 3 testing, prepared by Foster Wheeler, showed that expansion would have no significant impact on the environment. In March 1992 the New Jersey Department of Environmental Protection and Energy issued FWDC a permit allowing construction and operation of the Phase 3 Pilot Plant. With the permit in hand, FWDC issued purchase orders for Phase 3 long-lead items (i.e., the tower structural steel and pressure vessels).

In September Foster Wheeler received release from the DOE to proceed with the design and supply of a slurry/paste feed system for the plant. The coal/water paste feed nozzle required in the carbonizer and CPFBC was designed and is shown in Figure 3.17. The nozzle is modeled after the one successfully used in the 12-in. sq., 20-atm Leatherhead PFB in 1984 (DOE Tests 21 through 25). The nozzle exterior and tip are water cooled; the tip will be inserted into the carbonizer and CPFBC via a 6-in. refractory plug. Figures 3.18 and 3.19 show the paste feed nozzle locations; "clean-out" nozzles are provided in close proximity to them.

Preliminary layouts of the small piping systems (e.g., instrument air, nitrogen, water, steam) that will be required for the plant were begun. To meet the additional air requirements of the Phase 3 pilot plant, two air compressors will be rented and positioned behind the laboratory building. A preliminary piping layout from the compressors to the building, with the units sitting on an existing concrete pad, is shown in Figure 3.20. The feasibility of an alternative "closer-in" arrangement (Figure 3.21) is also being investigated.

3.3.2 Carbonizer Alkali-Removal Vessel

The alkali removal vessel preliminary design has been completed by W-STC and has been reviewed by FWDC. Earlier, Figure 3.11 showed the general vessel arrangement drawing. Alkali pellets will be removed from the vessel by a shop vacuum; W-STC will specify an appropriate vacuum system. Methods to sample the bed contents during shutdown periods will also be needed, and either vacuum withdrawal of a portion of the bed or manual sampling will be used. Based on past testing experience, the emathlite pellet bed will probably not be difficult to remove; but should agglomeration occur, the pellet bed will be broken up by rodding.

Some simple testing was conducted at W-STC to confirm that the vacuum system will be effective—emathlite pellets were vacuumed from an elevation simulating the height in the alkali removal vessel.

The vessel insulation has been specified. Detailed fabrication drawings are being prepared for the vessel. Details of the perforated plate/screen that covers the gas outlet nozzle have been completed.

Figure 3.17 Coal/Water Paste Feed Nozzle

Figure 3.18 Coal/Water Paste Feed Nozzle Location-Carbonizer

Figure 3.19 Coal/Water Paste Feed Nozzle Location-CPFBC

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Figure 3.20 Preliminary Air Compressor Layout

Figure 3.21 Alternative Air Compressor Arrangement

3.3.3 CPFBC Candle Filter

The CPFBC candle filter system design is nearly complete. It has involved an iterative process with conceptual designs, preliminary designs, and final detailed designs being sent to FWDC for review, recommended modifications, and approval to ensure that the system boundaries are consistent with the plant layout. W-STC will review all of the P&IDs, the component designs and their associated installation and maintenance requirements (lifting points, clearances, weights), and the selected vendors with FWDC before placing orders for equipment.

The status of the design of the major components of this system:

- Pressure vessel
- Internals (plenum, liners, shroud, pulse pipes, eductors)
- Tubesheet
- Candles and gaskets
- Compressor skid
- Pulse skid

is described in the following subsections.

Vessel. An outline drawing of the filter vessel (Figure 3.22) shows major vessel dimensions and lifting clearances required for the removal of internals. The vessel design details and fabrication drawings are being completed based on the pressure-relief nozzle design, tubesheet details, identification of vessel instrument nozzle locations, and detailed consideration of the tangential gas inlet.

The design criteria for the pressure-relief nozzle on the candle filter vessel were reviewed, and two alternative design procedures followed to size the relief nozzle. The preliminary design for the pressure-relief nozzle on the candle filter vessel has been completed and approved by FWDC. FWDC recommended a pressure-relief nozzle location on the vessel head. This location was checked by W-STC and is acceptable. The pressure-relief system design procedure and calculations are being documented by W-STC for FWDC review. Detailing of the preliminary pressure-relief system design is under way based on vendor dimensions provided for the pressure-relief valve and for the rupture disk assembly. Bids have been received for both. All instrument and access nozzles on the vessel have been identified, located, and sized.

Figure 3.22 CPFBC Candle Filter Maintenance Clearances

Internals. The detailed drawings for the candle cluster structure, the pulse pipes and eductors, vessel liners, and shroud are being completed based on W-STC's internal review. Materials of construction have been selected.

A conceptual design of a frame to hold the candle filter internals during maintenance has been completed, and a preliminary design is being prepared by W-STC for review by FWDC.

Tubesheet. Alternative tubesheet materials have been surveyed with respect to physical properties relating to tubesheet life, availability in standard thicknesses, and costs. Alloys 333, 310, 253MA, and 601 are being considered. The tubesheet design and life calculations have been completed based on Type 310 properties—the worst material with regard to physical properties. The design meets the life design specification of 9000 hours with 200 cycles. Further design evaluation has been conducted to provide the best transient and steady-state temperature profiles through the tubesheet structure to achieve the longest tubesheet life. The detailed tubesheet design and the insulating metal cover design have been completed, including the addition of some instrumentation accesses that pass through the tubesheet ring (Figure 3.23). The optimum tubesheet insulation design with respect to tubesheet life has been developed. Documentation for the procedures followed for the tubesheet design is being prepared. Bids have been received and a fabricator selected. The tubesheet order will be placed following approval by FWDC to begin major equipment purchase.

Candles and Gaskets. Coors mullite candles and associated gaskets will be specified for the filter, based on parallel development and testing of the filter elements and gaskets. The order for the candle filter elements will be placed when a confident position with respect to their performance has been established.

Compressor Skid. Because of its size and the limited building space available, the compressor skid will be located on an outside pad. The compressor dryer and accumulation tank must be inside, in a heated area. The compressor, dryer, and primary accumulation tank have been specified, and vendor bids and vendor information have been compiled. W-STC has given FWDC information on the selected compressor components for their review.

Pulse Skid. The pulse skid design strategy has been resolved, and a preliminary layout design has been developed (Figure 3.24). The layout fits on a skid with a cross-section of 2 ft x 12.5 ft; because of lifting constraints, the maximum envelope height is 5 ft. The skid is lifted into place completely assembled with the exception of the accumulation tank. This tank will be lifted into place separately and welded to the pulse piping. The completely assembled skid weight is 1500 to 2000 lb. The pressure relief on the accumulation tank is nominally specified to be 700 psig, 156 sft³/min, with an outlet size of 2 in. Further evaluation of the accumulation tank pressure relief is needed, and the vent conditions will be specified by FWDC. A detailed design of the pulse skid will not be generated since the skid will

Figure 3.23 CFBBC Candle Filter Tubesheet

Figure 3.24 Pulse Cleaning Skid for CPFBC Candle Filter

be assembled by W-STC in line with the conceptual layout constraints. Primary and secondary accumulation tank specifications were sent out for bids, and bids have been received.

The candle filter pulse system preliminary P&ID is shown in Figure 3.25; the system includes the compressor skid and the pulse skid components. In addition, a pulse control system specification was developed and sent out for bids to several control companies. The control system has been specified and five bids to provide the system were received by W-STC. Vendor selection was completed, and an order was placed with Great Lakes Control. The control system components have been received from Great Lakes Control.

3.3.4 Cold Flow Model N-Valve Tests

In Phase 1 of this project, we proposed collecting commercial second-generation PFB combustion plant carbonizer char/sorbent residue in a central hopper and transferring it to the CPFBC via a system utilizing a slide valve, a nonmechanical N-valve, and a vertical line. A similar arrangement (Figure 3.7) has been proposed for the Phase 3 Pilot Plant. The solids from the carbonizer char-collecting hopper to the CPFBC must be transferred safely and reliably at controlled rates over a range of operating conditions. In Phase 3 these operating conditions include transfer rates between 100 and 500 lb/h, system pressures of 100 and 200 psig, and differential pressures of 6 to 20 psi.

Using a slide valve to control solids flow rate and dissipate some differential pressure has certain advantages. These include simplicity of piping design and straightforward solids flow control. To minimize risks of bridging and blocking, the pipe size for draining solids in the carbonizer pilot plant has been limited to a 4-in. minimum. Because of the relatively small flow rates involved (100 to 600 lb/h), we thought that an attempt to induce a 10-psi or higher pressure drop by throttling a 4-in.-dia. slide valve might, itself, cause blocking. As a result, IGT has constructed a flow model replica of the transfer piping and is conducting tests to determine its performance characteristics.

The test system (Figure 3.26) consists of a pressurized feed vessel, the pressure-reduction piping configurations, and a pressurized receiver vessel. The system is extensively instrumented to allow the determination of pressure drops and solids and gas flow rates for analysis and interpretation of the results of the tests.

An existing 3-ft-dia. vessel (9 ft long) was used as the feed vessel. This vessel, which has a 4-in. solids outlet nozzle, simulates the pilot plant char-collecting hopper. Solids flow from the feed vessel into the transfer piping and proceed to an 18-in.-dia. receiver vessel approximately 9 ft long. The receiver vessel was modified to simulate the CPFBC. The pressures in each vessel are maintained by pressure regulators acting on the gas (nitrogen) flowing into the system for the feed vessel and from the system for the receiver vessel.

Figure 3.25 Process Instrumentation Drawing for CPFBC Candle Filter

Figure 3.26 Slide-Valve/Fluidized Riser Configuration (C₁) as Constructed

A flow test consists of loading the feed vessel, transferring the solids from the feed vessel through the pressure-reduction piping, and collecting the solids in the receiver. After a test, the solids from the receiver vessel are transferred back into the feed vessel.

The solids flow rate through the pressure-reduction piping is controlled by a 4-in.-dia. V-ported slide valve below the pressure vessel. The diameter of the solids transfer piping in this configuration is 4 in. After flowing through the slide valve, the solids flow down a 45-deg sloping line into a fluidized riser. The solids exit from the top of the riser and flow through a downward-sloping 3-in.-dia. pipe and a second slide valve and into the receiver vessel. The amounts of riser aeration gas and stripping gas are varied independently to determine the effect, if any, of changing these rates on the solids flow rate and the slide-valve pressure drop.

The system is extensively instrumented with differential-pressure transmitters (Δ PTs) distributed around the test loop. The outputs from all the Δ PTs are recorded for detailed analysis and interpretation following the tests. All gas flows and aeration rates to and from the system are measured by rotameters.

The solids flow rate through the system is determined by measuring the rate at which the level in the receiver rises. As the solids flow into the receiver they form a packed bed of material at the bottom of the vessel. Aeration gas added to the receiver passes upward through the solids. This gas flow generates a pressure drop through the solids. One Δ PT determines the pressure drop per unit length through the solids. The rate of solids transfer through the system is then determined by measuring the rate at which the solids level in the receiver (i.e., the pressure drop across the receiver) rises vs. time. Several Δ PTs placed in series along the length of the receiver are used to monitor the solids flow rate during each test. All Δ PT taps are in the straight-wall section of the receiver.

The material to be tested first was supplied by FWDC in three separate fractions, designated CA (carbonizer ash), PA (primary cyclone ash), and SA (secondary cyclone ash). These three materials were mixed to the proportions supplied by FWDC of 32 CA: 41 PA: 27 SA (wt%) and loaded into the feed hopper. The mixture did not flow freely and was prone to segregation during handling.

Early shakedown tests were conducted at a nominal system pressure of 100 psig and a hopper-to-receiver differential pressure of 10 to 15 psi. Solids flow was often intermittent, caused by material bridging either at the small slide valve openings or in the lower section of the riser. During the short period of sustained operation, a solids transfer rate of approximately 2000 lb/h was measured in the receiver.

Based on the results of these shakedown tests, several operating changes and modifications were made to improve operation and the control of material transfer. To keep material from bridging in the riser, the gas velocity in the riser must be high enough to fluidize the largest particles in

the mixture (about 1.0 ft/s). To measure these higher gas flow rates, both the riser aeration and exit gas rotameters were replaced with larger-capacity units. Higher exit gas rates caused larger amounts of fines to be carried to the exit gas filter, increasing filter pressure drop and varying receiver pressure. A cyclone was therefore installed between the receiver and the filter to reduce the rate of fines buildup in the filter.

Increased superficial gas velocity in the riser solved the problem of solids bridging, but required another operating change. At higher gas velocities, the mixture of solids segregated in the riser at low transfer rates. The finer material was quickly conveyed up the riser and into the receiver, while the coarser, heavier material accumulated in the riser. Eventually, the larger solids reached a steady-state concentration in the riser and were carried over into the receiver with the fines. The receiver aeration was apparently enough to fluidize the initial fines reaching the receiver, but any coarse solids entering the receiver would not be fluidized. They would sink through the fines, below the measurement zone, and not be accounted for by ΔP measurements. To overcome this problem, the direction of gas flow in the receiver was changed to downward to ensure the measurement of all of the material in a packed-bed mode, and a porous metal filter was installed inside the bottom of the receiver, as shown in Figure 3.27. A rotameter and manual control valve were also added to adjust the exit gas flow rate for reasonable values of $\Delta P/L$.

Tests conducted after completing these modifications were much more successful. Solids transfer rate measurement by packed-bed ΔP worked very well. At slide valve openings of less than about 1.5-in. the flow of solids was usually intermittent. The test conditions at the lowest repeatable solids transfer rate were:

- Hopper pressure: 100 psig
- ΔP between the vessels: 9 psi
- Superficial gas velocity in the riser: 0.84 ft/s
- Standpipe ΔP : ≈ 5 psi (or 1.23 psi/ft)
- Slide valve ΔP : ≈ 1 psi
- Slide valve opening: ≈ 1.5 -in.
- Riser $\Delta P/L$: 0.16 to 0.12 psi/ft (from bottom to top)
- Measured solids transfer rates: 1920 and 2045 lb/h

When the 4-in.-dia. slide valve was opened far enough to allow a continuous flow of solids through it, the flow rate was consistently much higher than the required 100 to 600 lb/h.

Figure 3.27 Modified Slide Valve/Fluidized Riser Configuration

The ΔP across the slide valve cannot be controlled independently, but can be affected by changing other sources of ΔP in the transfer line. By using the second slide valve in this configuration—a 3-in. V-port slide valve on the receiver inlet nozzle (Figure 3.26)—both the standpipe and the riser should operate in a packed-bed mode. This change would lower the slide valve ΔP and the rate of airflow through it, and allow the control of lower solids transfer rates.

To test this alternative transfer method, the 4-in. slide valve was opened all the way with no imposed ΔP between the vessels and only minimal aeration to the riser. The ΔP between the vessels was gradually increased to 10 psid, with only a small amount of material entering the bottom of the riser. Adding aeration to the bottom of the riser would cause some additional material to surge into the riser, but the result was not very effective. The ΔP between the vessels was increased to roughly 20 psid, yet the solids in the 4-in.-dia. lateral would not move freely into the riser.

An aeration tap was added to the 4-in.-dia. lateral section a short distance from the riser to induce solids flow around the 135-deg turn. This aeration location proved much more effective in terms of causing solids to flow into the lower riser, but did not control the flow smoothly, as an L-valve would. As the solids filled the riser, the measured $\Delta P/L$ in each section exceeded 1 psi/ft, even though the solids had not reached the top of the riser and were therefore not physically restricted from flowing upward. This type of behavior indicated that it is not feasible to transfer this material in packed-bed flow up the riser.

IGT therefore proposed an alternative solids-transfer configuration. It utilizes a restricted pipe discharge (RPD) standpipe, a gas/solids disengager, an L-valve, and a fluidized riser. The redesigned portion of this system is shown in Figure 3.28. Most of the ΔP between the vessels is dissipated in a packed bed of solids in the RPD standpipe, and the L-valve controls the solids transfer rate.

Initial testing of this configuration, conducted at system pressures of approximately 100 psig and imposed ΔP s of 10 to 20 psi, was fairly successful. Steady solids transfer rates between 150 and 900 lb/h were measured, but actual control of the transfer rate was poor.

The reduction in pipe diameter from 4 in. to 2 in. between the disengaging cross and the standpipe was the controlling restriction affecting the solids transfer rate and the mode of operation in the standpipe. The 4-in.-dia. section of the transfer line between the hopper and the a 2-in.-dia. standpipe was therefore replaced with a 2-in. pipe, as shown in Figure 3.29. The purpose of these modifications was to remove the fixed-flow restrictions mentioned and to allow the L-valve to control the solids transfer rate.

Figure 3.28 Initial RPD-Standpipe/L-Valve Configuration

Figure 3.29 Modified 2-in.-dia. RPD Standpipe/L-Valve Configuration

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Several tests were then conducted at a nominal system pressure of 100 psig and a ΔP between the vessels of approximately 10 psid. With the valve in the standpipe bypass line opened slightly, a significant amount of L-valve aeration was required to initiate solids flow. However, this aeration rate had to be reduced quickly to prevent the flow of solids from increasing uncontrollably.

The placement of the gas/solids disengaging cross above the 45-deg lateral and out of the solids flow path may be largely responsible for this behavior. If the gas traveling down with the solids cannot move upward through the packed material in and below the disengaging cross quickly enough, more and more gas will flow down the standpipe to the L-valve. This increasing net flow of gas through the L-valve would result in an ever-increasing solids transfer rate, limited only by the sustainable ΔP between the vessels and the amount of material in the hopper.

The tests apparatus is therefore being modified to place the 2-in. gas/solids disengaging across just below the 45-deg lateral section, at the start of the standpipe; and testing will resume in October 1992.

Section 4

TASKS 3, 4, 5, AND 6

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Section 4

TASKS 3, 4, 5, AND 6

Task 3—Construction and Shakedown

No work scheduled or performed.

Task 4—Test Plan Implementation

No work scheduled or performed.

Task 5—Phase 3 Topical Report

No work scheduled or performed.

Task 6—Revised Conceptual Design

No work scheduled or performed.

Section 5

TASK 7—PROGRAM MANAGEMENT/ REPORTING

Section 5

TASK 7—PROGRAM MANAGEMENT/REPORTING

5.1 OBJECTIVE

The Program Management/Reporting task objectives encompass overall guidance and control of the test program.

5.2 STATUS

The goal of the design effort is to have vessels delivered to the site so that Phase 3 construction can begin as soon as Phase 2 testing has been completed. Phase 2 testing is expected to be completed in May 1993.

The Phase 3 effort has been moving smoothly, no major technical problems have been encountered or are foreseen, and long-lead delivery items have been ordered. There should be no difficulty in mobilizing the construction activity in May and breaking ground for the Phase 3 Pilot Plant in June 1993, when Phase 2 testing is expected to end.

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