HOT GAS CLEANUP USING CERAMIC CROSS FLOW MEMBRANE FILTERS

Final Report

By
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Morgantown Energy Technology Center
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By
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Pittsburgh, Pennsylvania

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1. EXECUTIVE SUMMARY

The single unresolved technical issue in the commercialization of pressurized fluid-bed combustion (PFBC) for electric power production is the hot gas cleaning problem. In this technology, high-temperature and -pressure (HTHP), dust-laden flue gases from the combustor must be cleaned enough to reduce expansion turbine blade erosion to an economically acceptable level. Additionally, the level of particulate emission must be compatible with the New Source Performance Standards (NSPS) for environmental acceptability. The Department of Energy (DOE) has sponsored a wide range of research and development programs directed at the solution of this problem. These programs were divided into two classifications, one dealing with more advanced concepts where testing was to be done at relatively large scale and a second group of less advanced, novel concepts where the testing was to be carried out at a bench scale. The cross-flow ceramic membrane filter program described in this report is a member of the small-scale, novel concept group.

This report covers the progress of the program from October 1979 to October 1982 and includes a project summary, conclusions, and recommendations section and, finally, a task-by-task description of the actual test program.

Section 3 contains the project task reports and is therefore the heart of this report. It consists of the following subsections:

- 3.1 Systems Preparation - a description of the work involved in setting up the high-temperature and -pressure test facility, along with its capabilities
- 3.2 Cross-Flow Filter Supply - describes the arrangements under which a filter supplier was contracted for this work and the entire set of filters supplied and tested in this program

- 3.3 Test Program Results - a very long section that presents the actual HTHP test results. This section is presented in three levels of detail: a brief summary with a tabular summary of the experimental results; a more detailed account of this program consisting of a chronological compilation of the monthly project summaries; and, finally, a very detailed accounting of the month-by-month technical progress of the work, including all of the data generated.
- 3.4 Commercial Module Design - presents current best thoughts on how the cross-flow filter would be embodied in a commercial module.
- 3.5 Power Plant Economic Analysis - presents the results of a plant economic study performed for us by Burns & Roe, Inc., for the purpose of estimating the impact on power cost of implementing the cross-flow filter concept. Comparisons with an all cyclone reference plant are made.
- 3.6 Alkali Metal Removal - describes the early work carried out on this program in an attempt to incorporate both alkali and particulate removal in the same unit operation. This also contains a brief update of the considerable progress recently made in parallel alkali gettering programs that are being carried out at Westinghouse.
- 3.7 Field Testing - The original program allowed for slip-stream testing of the filter unit at an appropriate coal burning pilot PFBC. This work has been completed and is reported in Appendix E.

2. PROJECT SUMMARY, CONCLUSIONS, AND RECOMMENDATIONS

2.1 SUMMARY

A high-temperature and -pressure (HTHP), bench-scale experimental program was carried out on a novel porous ceramic filter to be applied in advanced coal conversion technologies such as pressurized fluid-bed combustion (PFBC) and coal gasification. This work has shown that the ceramic cross-flow filter can easily both meet the turbine tolerance criteria and satisfy the New Source Performance Standards (NSPS). A conceptual design for a commercial-scale unit has been carried out and then integrated into an existing economic study for comparison with an all-cyclone PFBC reference plant. This work indicated that the cross-flow system has the potential to reduce the specific hot gas cleaning costs to about 55% of those associated with the all-cyclone reference design and thereby to result in a significant net decrease in the levelized busbar power cost.

The filters under consideration have a unique configuration that allows a very high surface area per unit volume, so that the resulting filter systems are very compact and are physically no larger than the stage of cyclones that would usually precede them in a hot gas cleaning application.

In all, a total of seven different filter types were investigated, with nominal pore sizes ranging from 10 to 144 μm and filter flat thicknesses from 1 to 3 mm. Approximately 450 hours of actual test time were logged during a series of 48 different test periods. Most of the testing was carried out at a temperature of 1088 K (1500°F), at a system pressure of 1114 kPa (150 psig) and with an inlet dust concentration ranging from 2 to 6 g/Nm³ (1555 to 4666 ppm). The range of filter

velocities investigated was from 1.6 to 5.4 m/min, with most of the testing carried out at about 1.7 m/min.

At filter velocities around 2 m/min, the more porous elements gave pressure drops of about 1 to 2 kPa (4 to 8 in. wg) while the less porous filters had resistances that were typically 6 to 8 kPa (24 to 32 in. wg).

We found that under these conditions the filters were essentially absolute barriers to the redispersed PFBC test dust used: measured overall efficiencies were 99.99+%.

We experienced initial difficulty in obtaining adequate cleaning with a simple pulse-jet blowback system, but careful measurements of the magnitude and distribution of the pressure rise at the clean side of the filter during a blowback pulse allowed us to make adjustments in the nozzle position and pulse reservoir pressure so that successful on-line cleaning techniques were defined. Adequate filter cleaning could generally be achieved through the use of a 2069 to 3551 kPa (300 to 515 psia) pulse through a 0.8 cm nozzle lasting for 0.2 to 0.5 seconds.

Subsequent to the discovery of proper blowback conditions, a long duration test was carried out on an element with 100 μm pore size and 1.5 mm thick filter flats. During this test 80 hours of essentially continuous operation were achieved. Operating at 1.7 m/min with an inlet concentration of 5.04 g/Nm^3 (3,900 ppm) of dust, the filter settled into virtually stable operation within a few hours. In the conditioned state the base-line pressure drop was 1 kPa (4 in. wg) and would rise to the blowback trigger pressure drop of 3 kPa (12 in. wg) over a constant 45 minute period.

The only recurrent experimental difficulty experienced was a tendency for some of the filters to delaminate at one of the bonded layer interfaces. Dense, low-porosity elements were much more prone to failures of this type. Very large pore size elements, on the other hand, appeared to become too weak and to fail across filter flats. Working with the Coors Porcelain Co. (the filter manufacturer), we made

significant gains in filter element design and fabrication techniques, which have minimized this problem.

As part of this program, we investigated the possibility that alkali metal removal could be incorporated into a cross-flow filter system by injecting a powder of sorbent material into the gas stream ahead of the filter. Early experiments with the best-known sorbents (such as activated bauxite) indicated that the gettering reaction rates were so slow that sufficient contact time could not be achieved in this application. Additionally, the observed capacities of the materials tested implied poor utilization of sorbent material. More recently, in the course of related programs, sorbents have been identified that have very much improved alkali capacities, but the kinetic constraints still require sorbent thicknesses greater than could be considered in this type of filter.

Field testing at an actual coal-burning PFBC pilot plant was carried out at Argonne National Laboratories. The results of this testing are reported in Appendix E.

2.2 CONCLUSIONS

Conclusions about the cross-flow ceramic filter that can be drawn from this work are as follows:

- The filter is capable of essentially absolute filtration and more than meets operating and environmental constraints.
- The "conditioned" filter resistance to flow is very low compared to other filter systems and inertial devices, being typically 1 kPa (4 in. wg) at a filter velocity of 2 m/min.
- Over the range of filtration velocities tested (1.6 to 5.4 m/min) the filters did not appear to become plugged or "blinded."
- The filters could be cleaned on line with a simple pulse jet of reverse flow.

- Even though fabrication considerations constrain the overall element to a relatively small size, many elements can be configured into a module of very large filter area. Using this technique, a filter unit consisting of 18 modules can be housed in a pressure vessel roughly 5 m in diameter and 9 m tall and can process 170,000 m³/hr (100,000 acfm) of flue gas.
- Although filter failures by delamination at bonded joints did occur in some of the test elements, Coors personnel feel that relatively minor changes in fabrication procedures can eliminate this difficulty.
- Because of the compact nature of the cross-flow filter system, its capital costs are low relative to cyclone systems that have to rely on conventional baghouse units to meet NSPS. In the study conducted by Burns & Roe, Inc., the specific hot gas cleaning cost for the cross-flow system was determined to be \$69/kW net, while for the comparable all-cyclone case it was \$125/kW net.
- Because of the physical and chemical stability of the ceramic filter systems, we feel that this technology can easily be transferred to other high-temperature and -pressure hot gas cleaning applications, such as fluid-bed coal gasification.
- At this early point in development, we see no technical barriers to the successful implementation of the cross-flow technology.

2.3 RECOMMENDATIONS

Based on the very successful results of this test program, we suggest a sensible course of further development work in the following set of tasks.

2.3.1 Element Mechanical Design

The cross-flow design tested in this program was originally intended for cross-flow heat exchangers and consequently not intended to resist the pressure forces exerted by the pulse cleaning technique. We are confident that a task directed at improving the bonding between layers of the cross-flow filter would result in vastly improved mechanical properties for the element. Specific areas of investigation would include a parametric study of firing, temperature, and element

composition; the use of glassy bonding agents; and an improvement in the mechanical design of the joint that exists at layer interfaces. We would recommend that these studies be carried out on a simply shaped test specimen that is amenable to standard physical testing rather than on a multilayer filter. As part of this effort, techniques for improved mounting and sealing of the elements in a module should be investigated.

2.3.2 Bench-Scale Confirmation

Following this development, optimized test elements should be fabricated and tested in a well-controlled test facility, such as the one developed for this program, so that actual filtration characteristics could be determined and compared with previous experience.

2.3.3 Scale-up Parameters for Multielement Module

Subsequent to confirming the performance of a single-element device, a task should be initiated that would include the design and testing of a single, multielement module. Because a single module would consist of many (20 to 50) elements, it would be capable of processing very large flows and would require testing at a facility much larger than the available bench-scale facility. The Westinghouse pilot-scale High-Pressure and -Temperature Particulate Control Test Facility at the Synthetic Fuels Division would be suitable for this scale of testing. At this site work could be carried out in developing cleaning techniques for multielement modules as well as in confirming the mechanical design of the module. The product of this testing would be ready for testing at a large pilot facility such as the Curtiss-Wright facility in Wood-Ridge, New Jersey, or the Grimethorpe PFBC in the United Kingdom.

3. PROJECT TASKS

3.1 TASK 1 - HIGH-TEMPERATURE AND -PRESSURE TEST SYSTEM DESIGN AND CONSTRUCTION

The first step in carrying out the proposed high-temperature and -pressure filtration testing was to prepare a test passage for the work. This task consisted of the following four primary subtasks:

1. Determination of an appropriate laboratory space and subsequent construction of that laboratory
2. Design, procurement, fabrication, and installation of the actual test passage hardware and instrumentation
3. Preparation of the required operating and safety documentation and subsequent safety review
4. Modifications to the system as indicated by early run experience.

3.1.1 Laboratory Site Preparation

At the time the proposal for this work was prepared, what was thought to be suitable laboratory space was identified. The divisional operation safety administrator, however, later decided that none of the existing available sites was appropriate for this type of work. Many options were carefully examined, and eventually Westinghouse capital was used to build new space (adjoining a combustion laboratory). This new laboratory was to be dedicated to research in the area of hot gas cleaning. Consequently, a great deal of unanticipated planning and meeting with plant facilities personnel was required. Appropriate approvals within Westinghouse had to be obtained, as did authorization from state and local governments. Architects and consultants were contracted with, and construction was started in mid-July 1980.

Drawings of the new space are shown in Figures 3.1-1 to 3.1-3.

Construction was completed, with all facilities installed and shakedown tests begun in February 1981.

3.1.2 Test Passage Design, Fabrication, and Installation

Design of the filter test system was begun as soon as a filter supply commitment was obtained from Coors Porcelain Company. The safety committee required that, for the design basis, all the hot gas vessels be code-stamped ASME Section I (fired pressure vessels), and the ambient vessels such as the dust injector housing be stamped ASME Section VIII (an unfired vessel).

Figure 3.1-4 presents a process and instrumentation diagram for the entire test loop and indicates the major subsystems. A very brief description of the operation of the passage follows.

Air is compressed to 1793 kPa (250 psig) in a 29.84 kW (40 hp) reciprocating compressor and flows through a refrigerated dryer and filter into one of two parallel mass flow measuring sections. From the flow measuring and control section, the airflow is split into two streams, one for transporting particulates from the dust feeder and the other for primary combustion air. These streams rejoin in the combustor section of the test piping proper. Fuel for operation is supplied from 45 kg (100 lb) tanks of propane, which are heated in a water bath to a temperature of 319 K (115°F) in order to provide vapor pressure high enough to cause flow through heat-traced lines into the pressurized combustor. There it is mixed with air and burned. The temperature of the hot dusty air leaving the combustor is sensed and used to regulate the propane flow for temperature control. The hot dusty gas flows past an inlet sampling port before entering the filter housing. The filter housing is separated into two sections, the dirty side (in which the filter hangs) and the clean side (into which the filtered gas flows). These sections are separated by a plenum, which is part of the filter element mounting assembly. Air can flow from either of these two

Figure 3.1-1 Floor Plans for New HTHP Test Laboratory

Figure 3.1-2 Facilities Layout for HTHF Laboratory

Figure 3.1-3 Systems Layout for HTHP Laboratory

Figure 3.1-4 Filter Test System Schematic

sections into the exhaust lines. From the clean side the filtered air flows past the outlet sampling port and then into the water-jacketed exhaust lines. From the dirty side of the filter housing, air can flow through a bypass line that provides a momentary flow path around the filter during blowback cleaning when the bypass valve is opened. Both paths join in a single, water-jacketed exhaust line that leads to a back-pressure control valve used to establish passage pressure. The flow is then exhausted through a roof-mounted cyclone.

Table 3.1-1 presents a summary of the test loop operating parameters

Table 3.1-1
TEST PASSAGE PARAMETERS

Airflow	0-4.66 kg/min (10.3 lbm/min)
Pressure	0-1793 kPa (250 psig)
Temperature	294-1253 K (70-1800°F)
Filter Housing	25 cm dia. x 46 cm long (10 in dia. x 18 in long)
Dust Feed	0-10,000 ppm typical

Figure 3.1-5 presents a photograph of the test passage.

A very detailed description of the test passage and its operation and control is provided in Appendix A.

3.1.3 Operating and Safety Documentation

Before the test loop was approved for operation, considerable safety documentation had to be prepared and submitted to the Westinghouse R&D Safety Review Committee. This committee subsequently held a formal review of the material. The documentation consisted of:

1. Ceramic Filter Test System Description (Appendix A)
2. Ceramic Filter Test System Operating Procedure Document (Appendix B)
3. Failure Assessment for Ceramic Filter Test System (Appendix C).

Figure 3.1-5 High-Temperature and -Pressure Filter Test Facility

3.1.4 System Modifications

By March 1981 the test loop was installed, all approvals had been obtained, and shakedown runs were under way at temperature and pressure. Initially, several bothersome difficulties with the flow control loop were worked through with the aid of a local Westinghouse Division (Industry Systems Division). The next difficulty encountered arose from the poor combustion characteristics of propane air in our system at elevated pressure. Figures 3.1-6 through 3.1-8 show carbon deposits that occurred when the system was fired at pressures greater than about 203 kPa (2 atm). The accumulated soot eventually deflected the flame, so it impinged directly on the Inconel combustor lines and caused it to melt, as shown in Figure 3.1-8. This difficulty was discussed with personnel in the combustion section, and on their recommendation a small amount of "premix" air was blended with the propane in the flameholder nozzle. This simple procedure vastly improved the quality and stability of the flame and solved the problem.

In subsequent tests, the system was operated at temperature and pressure with dust injection. Early operating experience indicated that some minor modifications should be made. These include the following:

- Mill a set of grooves in the stationary dust injector plate. This would allow more uniform airflow through the dust injector system and thereby reduce the pulse effect caused by the holes in the rotating dust delivery plate.
- Replace large eductor in the dust injection system with a Swagelock tee.
- Install differential pressure gauges on the dust feed lines. These sense dust flow as "bumps" in the differential pressure. Plugging can be detected rapidly by a steady pressure drop.
- Install a set of ball valves in the dust injector lines that allows each line to be vented to the atmosphere while either the passage or dust injector side is isolated. This allows the lines to be blown out quickly and easily.

Figure 3.1-6 Soot Accumulation on Flameholder

Figure 3.1-7 Carbon Deposits in Combustor

Figure 3.1-8 Damaged combustor Liner

- Extend the rear dust feed line into the reduced cross-sectional area of the combustor to a point beside the flameholder. This is an attempt to ensure that all the dust fed from the dust injector is delivered to the filter.

With these modifications to the dust injector accomplished, reliable dust feed was possible, but we observed that the actual dust delivery rates were usually about twice those expected, based on rotational speed, feed plate hole spacing, and dimension. This discrepancy was attributed to dust "sneaking" between the fixed and rotating feed plates. A simple fix was attempted that consisted of gluing a thin pad of felt to the stationary plate on which the rotating plate turned. This effectively blocked or sealed all delivery holes in the moving plate except the one aligned with the fixed plate through-hole. Figure 3.1-9 is the calibration curve obtained subsequently and indicates the effectiveness of the fix. In the higher feed rates (above 2 g/min) the rate correlates very well with the calculated rate if one allows for somewhat less than 100% filling of the delivery holes. Typical dust feed rates were on the order of 2 to 4 g/min.

Figure 3.1-9 Dust Feed System Calibration Curve

During a later stage of the first filter tests, we noticed that the indicated filter pressure drop no longer fell when the bypass valve was opened as it had early in the test. When the system was shut down and cooled overnight, the exhaust bypass line was opened for inspection and was found to be plugged solid with large chunks of agglomerated dust. Figure 3.1-10 indicates the plugged region. We had anticipated that dust would blow out of the system during blowback and periods when the bypass line was opened, but this did not occur. An additional problem appeared when we discovered that the pressure relief line for the filter housing was also plugged, rendering it ineffective. For both operational and safety reasons the system was shut down until modifications of this section could be completed.

To address this difficulty the modification shown in Figure 3.1-11 was developed. During blowback the bypass line is opened and the large clumps of dust expelled from the filter surface fall down past the bypass tee and into the catch pot. The catch pot has been sized so that more than a full day of testing can be carried out before it needs to be emptied. This is accomplished by removing the pot from the system by the union provided and emptying it through any of the several penetrations provided.

In addition to providing the catch pot, the tee section to the pressure relief system has been rotated 90° to a vertical orientation so that residual dust and condensate would not accumulate in the rupture disc line.

The final addition was the inclusion of the drain tee shown at the right side of Figure 3.1-11. Past experience indicated that the water vapor formed from the combustion of propane was likely to condense in the long run of water-jacketed exhaust pipe and run backward into the lowest part of the exhaust system. This was possibly a large factor in the plugging problems previously observed. The new system provided the ability to drain this condensate off as required.

Figure 3.1-10 Filter Test System Schematic

Figure 3.1-11 Exhaust Modification

This modification greatly reduced the exhaust line plugging problem and made it possible to carry out day-long, uninterrupted tests. Experience has shown that plugging of the bypass line does occur gradually, especially at low flow rates, and this line must be cleaned after about 10 to 12 hours of dust feed time.

3.2 CROSS-FLOW FILTER SUPPLY

Subsequent to the award of this contract, Westinghouse was notified by W. R. Grace Co. that they would be unable to participate in this development program as a subcontracted supplier of the filter elements as had been planned. Since the entire program hinged on the supply and testing of cross-flow filter elements, Westinghouse did not feel it was prudent to commit any significant funds until an alternative course of action could be defined. Our technical monitor concurred in this decision. Efforts initially centered on identifying organizations that might have similar technology in hand or the capability of developing it in a short period of time.

Two potential suppliers were identified and approached to determine their interest in the program and their estimate of the level of effort required to carry out the original scope of work envisioned in the W. R. Grace subcontract. The alternative suppliers identified were G.T.E. Sylvania and Coors Porcelain Company. In both organizations technical personnel expressed interest in participating in the program, and both organizations eventually obtained approval for the proposed effort from their programing and planning personnel.

After the two interested suppliers were identified and their proposals to participate in the program had been received, Westinghouse evaluated them. With the DOE technical monitor's concurrence, Coors was selected, to some extent on the basis of technical merits, although both suppliers' materials had some potential advantages. Figure 3.2-1 is a photograph of the various materials. Coors was in a position to respond to our schedule and thus became the primary candidate.

Figure 3.2-1 Ceramic Cross-Flow Filter Elements

Having made this decision, we then had a problem in modifying the contract to substitute Coors for W. R. Grace, since the actual responsibility for management of this contract was in the process of being transferred from DOE Germantown to DOE Morgantown. To deal with this problem, a purchasing arrangement was negotiated with Coors whereby small lots of test specimens could be purchased at a unit cost that would not require formal contract modification through DOE. By pursuing this course of action, Coors rapidly became actively involved in the program. The initial four test elements were ordered and fabrication began. These initial test elements were to be fabricated to the following approximate specifications:

Body - Alumina-silicate

Rib height - 0.38 cm

Rib spacing - 1.52 cm

<u>Element No.</u>	<u>Pore Size, μm</u>	<u>Back Wall (Filter Surface) Thickness, cm</u>
1	10	0.10
2	10	0.15
3	100	0.10
4	100	0.15

Over the course of the program Coors provided a total of seven different types of filters. These are tabulated in Table 3.2-1, along with some qualitative comments concerning the qualities of each filter type.

The first samples received from Coors were WH-1 and WH-2, and both were received as $15.24 \times 15.24 \times 15.24$ cm ($6 \times 6 \times 6$ in) blocks of material. Figure 3.2-2 is a photograph of pieces of these materials cut to size for testing. These materials were intended to be identical except that WH-1 had a nominal 100 μm pore size and WH-2 had a 10 μm pore size. As the photograph shows, the WH-1 specimen had apparently experienced a firing that caused some sagging and deformation in the lower layers of the element. This particular filter turned out to

WH1

WH2

Figure 3,2-2 Coors Cross-Flow Filter Samples WH1 and WH2

Table 3.2-1

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CROSS-FLOW FILTER PROPERTIES

Filter ID	Element Size, cm	Nominal Pore Size μm	Filter Thickness, cm	Comments	Qualitative Performance
WH 1	15.24 $\times$ 15.24 $\times$ 15.24	100	0.15	Block cut into 3 test specimens each with about 0.11 m^2 (1.2 ft^2) of available filter area	Excellent filter qualities with good mechanical properties
WH 2	15.24 $\times$ 15.24 $\times$ 15.24	10	0.15	Same as WH-1	Poor cleaning and low resistance to delamination failure
WH 3	15.24 $\times$ 15.24 $\times$ 15.24	10	0.10	Thinner filter flats caused some fabrication difficulties 0.11 m^2 (1.2 ft^2)	Same as WH 2
WH 4	15.24 $\times$ 15.24 $\times$ 5.08	10	0.15	Two elements fabricated to fit test assembly incorporated design improvements to reduce delamination failures. Were supposed to have $100 \mu\text{m}$ but found to have $10 \mu\text{m}$ pore size	Same as WH-2
WH 5	15.24 $\times$ 15.24 $\times$ 5.08	144	0.15	Porosity created by mixing rubber particles in green ceramic prior to firing	Excellent filter qualities but poor mechanical strength and low resistance to breaking across filter flats
WH 6	15.24 $\times$ 15.24 $\times$ 5.08	144	0.30	Same as WH-5 but thicker filter flats for more resistance to breaking	Not significantly different from WH-5
WH 7	15.24 $\times$ 15.24 $\times$ 5.08	100	0.15	Wider perimeter for better bonding between layers and more resistance to delamination 0.068 m^2 (0.73 ft^2)	Limited life testing but excellent filter qualities and apparently very good mechanical properties

have very good resistance to delamination, which is probably related to its firing conditions. The second element, WH-2, was much denser and apparently bonded less well, since failures by delamination occurred within a day on both WH-2 elements tested, as described in Section 3.3. Test element WH-3 (Figure 3.2-3) was thought to be 100 μm in pore size, like WH-1 except for a 0.10 cm (40 mil) back web (or filter flat) instead of a 0.15 cm (60 mil). When it was tested, however, its pressure drop characteristics were found to be more like the 10 μm materials tested. Subsequent investigation by Coors revealed that both WH-3 and WH-4 had been fabricated from 10 μm material instead of the intended 100 μm plates. Test elements from WH-3 suffered from the same type of delamination failures as those from WH-2.

We attempted to address the bonding problem with test elements of WH-4. Westinghouse provided the sketch shown in Figure 3.2-4. The intention was to have Coors fabricate elements that would fit the filter holder assembly directly without having to cut elements from larger blocks. This allowed Coors to incorporate a wider perimeter around the clean-side flats and at the dirty-side edges, which should have promoted better bonding between layers and, thus, better delamination resistance. Figure 3.2-5 is a photograph of the element as delivered. As the figure shows, bonding of the 10 μm base medium was not good. These elements failed by delamination on the second day of testing.

Test elements WH-5 and WH-6 (shown in Figure 3.2-6) were fabricated from the same pressed plates of green ceramic and, consequently, had the same porosity and pore size (nominally 144 μm). The 0.15 cm (60 mil) WH-5 material showed greatly improved resistance to delamination, but after a considerable amount of very promising test time it failed by breaking across several filter flats. WH-6, with its thicker 0.30 cm (1.0/2.0 mil) filter flats was an attempt to resist this failure, made at the cost of increased filter pressure drop. Subsequent testing of the thicker filters did not indicate any improvement; we observed the same type of failure during testing.

WH3

WH4

Figure 3.2-3 Coors Cross-Flow Filter Samples WH3 and WH4

Figure 3.2-4 New Element Design

Figure 3.2-5 New Filter Medium Fabricated by Coors: Approximately 0.12 m^2 (1.31 ft^2) Filter Area ($10 \text{ }\mu\text{m}$ Pore Size) 1.5 mm (60 mil) Thick Back Web

WH5

WH6

Figure 3.2-6 Coors Cross-Flow Filter Samples WH5 and WH6

The final element tested in the current program was sample WH7. This element had a nominal pore size of 100 μm and a filter thickness of 0.15 cm (60 mils), much like the original material, WH1. This element (shown in Figure 3.2-7) incorporated all the fabrication improvements developed through the program. Although test time on this element has been limited because of schedule and financial constraints, this filter does appear both to be an excellent filter and yet to retain good mechanical properties.

3.3 HIGH-TEMPERATURE AND -PRESSURE CROSS-FLOW CERAMIC FILTER TEST RESULTS

3.3.1 Introduction

The high-temperature and -pressure filtration work carried out under this contract is presented here in three levels of detail. First, a summary overview of test results is presented in compact form, reviewing filter types, test totals, and ranges of operation. The summary overview presents a table of all test results obtained. This table includes comments for each test sequence and is accompanied by a brief discussion of the test results for each filter series.

The overview section is followed by a summary section, which is comprised of a chronological set of monthly summaries of results and progress. The final section presented is a very detailed description of the test program itself, including all the data generated and reported through the course of the test program.

3.3.2 Overview

In carrying out the experimental phase of the cross-flow filter program, we examined several different filter types. These filters were fabricated by the Coors Porcelain Company and supplied to Westinghouse for testing. Pore sizes ranged from a nominal 10 to 144 μm . Filter flat thicknesses tested included 1.0, 1.5, and 3.0 mm (40, 60, and 120 mils). Virtually all testing was conducted at a total pressure of 1114 kPa (150 psig). Early testing was carried out at 813 K (1000°F), but most of the testing was done at a system temperature of 1088 K (1500°F). The

Figure 3.2-7 Coors Cross-Flow Filter Sample WH-7

test dust employed was redispersed PFBC ash from either the Exxon miniplant or from the Curtiss-Wright test rig. This material had an average size of about 8 to 10 μm . Tests were conducted at dust loadings ranging from 1.93 to 13.4 g/Nm^3 (1,500 to 10,400 ppm) but most frequently were in the range of 3.9 to 6.5 g/Nm^3 (3,000 to 5,000 ppm). Filter velocities examined ranged from 1.7 to 5.2 m/min (5.7 to 17.2 ft/min), but most of the testing was carried out in the 1.5 to 3.7 m/min (5 to 12 ft/min) range. Typically, the pressure drop across a large pore size test element rose from a steady base of 1 kPa (4 in. wg) to 3 kPa (12 in. wg) over a 30 to 60 minute period and was returned to the base pressure drop by means of a simple high-pressure pulse from a 2.12 dm^3 (0.075 ft^3) pulse accumulator pressurized to 2512 to 2800 kPa (350 to 400 psig). The blowback nozzle employed was a simple 0.95 cm (3/8 in) tube centered above the clean gas exit. The less porous (10 μm) elements behaved much the same except that they were cycled between pressure drops of 6 to 10 kPa (24 to 40 in. wg) and 2.5 to 6 kPa (10 to 24 in. wg). Overall collection efficiencies for all filters were exceptionally high when no flaws were present, typically 99.9+ %.

Elements that failed mechanically during the test usually showed an easily detectable decrease in efficiency to the range of 99.0 to 99.5%. Failures generally occurred by a simple delamination of layers of the filter element and appeared to be due to mechanical stresses imposed by the metallic filter mounting assembly during rapid heat-up or cool-down. The less porous, small-pore-size materials typically seemed to be more subject to this type of failure, and the larger-pore-size materials were resistant to it.

A total of 48 high-temperature tests were carried out on 14 different test elements of 7 different types. This effort represented approximately 450 hours of actual high-temperature and -pressure test time.

Table 3.3.1 presents a brief summary of test program totals and typical test parameters. Table 3.3-2 presents a complete summary of the test program results and comments.

Table 3.3-1
TEST PROGRAM SUMMARY

Element Types	7
Total Elements Tested	14
Pore Size, μm	10, 100, 144
Filter Flat Thickness, mils/mm	40, 60, 120/ 1, 1.5, 3
Total Number of HTHP Tests	48
Total HTHP Test Time, hr	450
Typical Base-Line Pressure Drop, kPa/in. wg, at Filter Velocity of 1.8 m/min (6 ft/min)	
• Large pore	1.0/4
• Small pore	3-5/12-20
Typical Overall Efficiency on Redispersed PFBC Ash, %	
• No delamination	99.99
• With delamination	99.5-99.9
Typical Dust Loading, $\text{g}/\text{Nm}^3/\text{ppm}$	3.9-6.5/3,000-5,000

3.3.3 Test Program Summaries

This section presents a slightly more detailed account of observed test results and progress through the test program.

During March 1981, the actual filter test program began with a series of five tests conducted at elevated temperatures and pressures, including two tests in which a filter element was exposed to high-temperature and -pressure test conditions. In the conduct of these experiments, several operational problems and instrumentation difficulties typical of starting a relatively complex system were encountered and successfully dealt with. Passage operation proceeded in the following sequence of base test conditions:

Table 3.3-2

SUMMARY OF CERAMIC CROSS-FLOW FILTER TEST PROGRAM

Dwg. 560XC69

Filter Sample No	Test Number	Temp. °F (°C)	Pressure psig (atm)	Concentration gr/scf (ppm)	Filter Velocity ft/min (m/min)	Blowback Pressure psig/atm	Steady Cycle Time min	Blowback Trigger Pressure Drop/ Base Press. Drop In. wg. (kPa)	Overall Efficiency %	Comments
IWH 1	3.25.81	1500 (815)	150 (11)	0	6.15 (1.87)	0	—	—	—	First shutdown run at full operating temperature and pressure with filter element installed but no dust injection. Carbon deposition in combustor led to shutdown.
IWH 2	3.27.81	1500 (815)	150 (11)	0	6.76 (2.06)	0	—	—	—	Carbon deposition problem solved by use of air premix in propane burner. Examination of filter after test showed it to be in good condition, no dust injection during this test.
IWH 1	4.15.81	1000 (540)	150 (11)	2.1 (3.700)	6.76 (2.06)	175-240/ 13-17	75	5.5/3 (1.4/ 75)	99-99+	First HTHP test with dust injection. Some difficulty with dust injector but built cake and tried various blowback pressures for cleaning. Limited to maximum of 240 psig blowback pressure at this point. Very high efficiencies and good cleanability observed. Filter removed for inspection and appeared to be in good condition. Several minor improvements made in dust feed system.
IWH 1	4.23.81	1000 (540)	150 (11)	5.8 (10.400)	5.6 (1.7)	200-240/ 14.3-17	(2.7)	15/4 (3.75/1)	99-99+	First multicycle test included seven filter/clean cycles over 150 minutes of filtration time. Investigated effect of blowback time and pressure as well as operating parameters for dust injector. Inspection of the filter revealed incomplete cleaning of bottom half of filter and virtually no cleaning at top half. Serious plugging of bypass line and bursting disk indicated a modification to system was required before more testing could be carried out.
IWH 1	6.15.81	1000 (540)	150 (11)	2.46 (4.358)	5.7 (1.7)	235/ 16.7	5	12/4 (3/1)	99-99+	Steady but relatively short cycle times were achieved using 235 psig blowback pulse. Rapid initial rise in pressure drop trace indicated incomplete cleaning of significant areas of the filter. Confirmed on posttest inspection that top half of filter not cleaned. Reinstalled filter without cleaning.
IWH 1	6.18.81	1000 (540)	150 (11)	2.21 (3.940)	5.7 (1.7)	235/ 16.7	2-3	15/12 (3.75/4)	99-91	Investigated effect of higher trigger pressure drops on filter performance. Ran at 12, 15, and 20 in. wg. pressure drops and able to establish steady cycle at all points. Inspection indicated incomplete cleaning at top part of filter with reasonably good cleaning at bottom. Filter vacuumed clean and reinstalled. Modified blowback system to allow higher pulse pressure.
IWH 1	6.26.81	1000 (540)	150 (11)	3.5 (6.250)	5.7 (1.7)	260-300/ 18.3-21	6	12/4 (3/1)	99-99+	Test looked at effect of higher pressure cleaning pulses. As pulses were increased from 235 to 260 to 300 psig noticeable improvements in cleaning were achieved. Gradual shortening of cycles indicated cleaning still incomplete. Posttest examination of filter showed improved cleaning but still poor at top of filter.
IWH 1	7.2.81	1000 (540)	150 (11)	2.56 (4.537)	5.7 (1.7)	300/21	7	12/9 (3/1.75)	99-94	Installed differential pressure transducers and high speed recorders to quantify effects of pulse pressure on cleaning. Ran test at 300 psig pulse and got steady but relatively short cycles. Good cleaning in bottom portion of filter but poor at top.
IWH 1	7.15.81	1000 (540)	150 (11)	2.3 (4.075)	5.7 (1.7)	300-400/ 21-28	—	—	99-99	Run cut short because of dust plug in pressure drop taps, tried pulses of 300 and 400 psig. Filter found to be very clean at bottom and cleaner at top - some nonuniformity however.
IWH 1	7.31.81	1000 (540)	75-150 (6-11)	0	—	200-500/ 14-34	—	—	—	Ran test with no dust but differential pressure transducers in top and bottom of filter clean side. Ran pulse tests from 200-500 psig and measured clean side pressure rises that occurred.
IWH 1	8.5.81	1000 (540)	75-150 (6-11)	1.34 (2.375)	5.7 (1.7)	200-500/ 14-34	35	12/4 (3/1)	99-35	High-pressure pulse tests revealed suction effect at top of filter - moved blowback nozzle out of clean side plenum for more uniform cleaning in latter tests.
IWH 1	8.12-13.81	1000 (540)	150 (11)	1.8 (3.233)	5.7 (1.7)	500/34	50-60	12/4 (3/1)	98-35	Long cycle time and lower efficiency indicated gasket leak due to high-pressure pulses.
IWH 1	8.18-21.81	1000 (540)	150 (11)	1.34 (2.375)	5.7 (1.7)	500/34	40-50	12/5 (3/1.25)	99-95	Element IWH 1 with new gasket and sealing cement. First test with blowback nozzle out of clean side plenum for better cleaning. Four day test with excellent performance and no flaws detected.
IWH 1	8.27-28.81	1000 (540)	150 (11)	1.30 (2.306)	5.7 (1.7)	500/34	40	12/6 (3/1.5)	99-46	Uniform cleaning but small leak due to gasket blow out.
IWH 1	9.1-2.81	1000 (540)	150 (11)	3.5-5.3 (6.200-9.300)	5.7 (1.7)	500/34	25-50	12/5 (3/1.25)	99-91-98-62	Very high dust loadings caused bypass plugging during second day and impaired cleaning. Low efficiency suspected result of poor seal.
IWH 1	9.8-10.81	1000 (540)	150 (11)	3.4-4.4 (6.000-7.700)	5.7 (1.7)	500/34	60	12/5 (3/1.25)	98-84-99-52	Element checked for leak between test days and none found. Checked and found some system associated potential leaks and made indicated repairs.
IWH 1	9.17.81	1000 (540)	150 (11)	5.7 (10.100)	5.7 (1.7)	500/34	60	12/6 (3/1.5)	99-85	Ruptured dust feed line limited test to 90 min. Observed first crack in filter at front face easily fixed by cementing front row of channels.

Table 3.3-2 (Continued)

Dwg. 5603C71

Filter Sample No.	Test Number	Temp °F (°C)	Pressure psig (atm)	Concentration gr/scf (ppm)	Filter Velocity ft/min (m/min)	Blowback Pressure psig/atm	Steady Cycle Time min	Blowback Trigger Pressure Drop/ Base Press. Drop In. wg. (kPa)	Overall Efficiency %	Comments
1WH 1	9 28. 81	1000 (500)	150 (11)	5.519 730)	5.7 (1.7)	500/34	13	12/10 (3/2.5)	99.91 98.10	Bypass line plug caused "on line" cleaning to be somewhat less effective than usual. Decreasing efficiency indicated worsening leak. Filter examination revealed delamination at plane of previous flaw. Element 1WH 1 accumulated 35 thermal cycles. 500 blowback pulses. 100 hr. operation and 13 mountings and dismountings.
2WH 1	9 29-30 81	1000 (500)	150 (11)	3.516 200)	5.7 (1.7)	400/27.7	70	12/2 (3/ .5)	99.95	Second element cut from same block as first element. Excellent performance over first 6 1/2 hr of test.
2WH 1	10. 7 81	1000 (500)	150 (11)	2.314 000)	5.7 (1.7)	400/27.7	90	12/3 (3/ .75)	99.89	Improvements in dust feeder kept loadings lower and lengthened cycle time. Continued success with lower pressure cleaning pulse.
2WH 1	10. 8 81	1000 (500)	150 (11)	2.614 530)	5.7 (1.7)	400/27.7	100	12/3 (3/ .75)	99.70	Slightly longer cycle times and lower efficiency indicated small constant leak.
2WH 1	10 12. 81	1500 (815)	150 (11)	2.915 128)	5.7 (1.7)	400/27.7	210	12/3 (3/ .75)	99.01	First test at 1500°F. Thermal expansion of liner and of clamping assembly suspected of causing delamination and crack across filter flats. Installed titanium hold-down studs to match coefficients of expansion and prevent loosening of element.
1WH 3	10 15-16 81	1500 (815)	150 (11)	3.315 900)	6.3 (1.9)	400/27.7	90	15/8 (3.75/2)	99.70- 99.48	New element 40 mils thick and supposed to be 100 µm pore size (later found to be 10 µm). Gradual drop in efficiency indicated leak. Thermal growth of inlet liner deformed filter holder assembly and poor gasketing procedures caused element to be loose. Element failed by delamination.
2WH 3	10. 30. 81	1500 (815)	150 (11)	3.315 900)	6.3 (1.9)	400/27.7	70-90	12/7 (3/1.75)	99.97	Modifications to passage liners to accommodate thermal growth worked well. New 3M gasketing material used in conjunction with a phosphate ceramic cement also worked well.
2WH 3	11 3-5 81	1500 (815)	150 (11)	3.816 750)	5.7 (1.7)	400-500/ 27.7 34		12/9 (3/2.25)	99.95- 99.86	Continuation of 10. 30. 81. Very high efficiencies first two cycles with marked decrease during third day. Examination revealed delamination failure at end plate.
1WH 2	11 11. 81	1500 (815)	150 (11)	3.315 900)	5.7 (1.7)	500/34	30	34/28 (8.5/7)	99.62	New element 60 mil thick. 10 µm pore size had higher pressure drop. Increasing penetration throughout test led to examination and discovery of start of delamination. Flaw patched with ceramic adhesive and put back into test system.
1WH 2	11 16. 81	1500 (815)	150 (11)	—	—	500/34	—	—	—	Repair of filter unsuccessful as element delaminated at beginning of test.
2WH 2	11 17 18. 81	1500 (815)	150 (11)	3.315 900)	6.5 (2)	350/24.8	6-12	40/35 (10/3.75)	99.91 98.63	New piece of 60 mil 10 µm material tested well first day but low efficiency second day indicated delamination failure.
1WH 1	11 23 24. 81	1500 (815)	150 (11)	3.616 500)	5.2 (1.6)	350/24.8	0		99.77 99.25	First WH 4 element 10 µm pore size with 60 mil (1.5 mm) thickness. Element made to fit test holder and had 1.31 ft ² (12 m ²) area. Did not appear to be bonded well and failed second day.
3WH 1	11 30-12. 4. 81	1500 (815)	150 (11)	2.213 900)	5.7 (1.7)	350/24.8	45	12/4 (3/1)	99.94	Long duration test of original 60 mil 100 µm material. Accumulated 80 hr. of filter time over five days with 14 thermal cycles. Filter performance very consistent over entire test period with high efficiency.
3WH 1	12 9-10. 81	1500 (815)	150 (11)	1.512 670)	11.4 (3.5)	350/24.8	30-40	18/8 (4.5/2)	99.89	Continued testing of 3WH 1 at twice previous filter velocity showed predictable and excellent performance.
3WH 1	12 16. 81	1500 (815)	150 (11)	2.213 950)	11.4 (3.5)	350/24.8	30	24/8 (6/2)	99.77	Continued high velocity test at higher dust loading. Cleaning remained excellent but first indication of minor leak.
3WH 1	12 18. 81	1500 (815)	150 (11)	1.112 060)	11.4 (3.5)	350/24.8	30	24/8 (6/2)	99.81- 99.61	Intended to be repeat of previous runs but dust load measured low. Most probably an error in in that determination since run was virtually the same as previous test in all other respects.
3WH 1	1. 5-6. 82	1500 (815)	150 (11)	2.314 100)	11.4 (3.4)	400/28.2	20	24/10 (6/2.5) 24/12 (6/3)	99.74 99.81	Pushed to yet higher velocities and still able to maintain stable cleaning of filter. Minor leak does not worsen.
3WH 1	1. 14. 82	1500 (815)	150 (11)	1.612 850) 1.312 360)	14.2 (4.3) 17.2 (5.2)	400/28.2	12	24/15 (6/3.75) 24/21 (6/4.25)	99.80	Pushed to very high velocities (17.2 ft/min) where stability of cycle was questionable. Return to lower velocities indicated some filter area had been plugged. Slightly higher pressure series of blowbacks at end of test returned previous base line pressure drop.

Table 3.3-2 (Continued)

Filter Sample	Test Number	Temp °F (°C)	Pressure psig (atm)	Concentration g/sch (ppm)	Filter Velocity ft/min (m/min)	Blowback Pressure psig/atm	Steady Cycle Time min	Blowback Trigger Pressure Drop/ Base Press. Drop In wg. (kPa)	Overall Efficiency %	Comments
2WH 1	1 15.82	1500 (815)	150 (11)	1 612 850	5.7 (1.7) 11.4 (3.4) 17.6 (5.4) 14.1 (3.4)	400/24.2	15 11 25 2	12/6 (3/1.5) 24/15 (66/3.75) 24/22 (6/5.5)	99.85	Stepped through velocities 5.7 11.4 17.6 14.1 ft/min. Performance of filter good except at two highest velocities where base pressure drop was very near max. trigger pressure drop. Small leak observed to remain constant. Filter examined after 120 hr. 602 blowback pulses and 25 thermal cycles and found to have small crack (partial delamination at the clamped end.
1WH 5	1 22 28.82	1500 (815)	150 (11)	1 7 2 (3 000 3 500)	5.7 (1.7)	350/24.8	110	12/2 (3/1.5)	99.99 ⁺	Four-day test of new filter element 60 mil 144 um pore size. Material very low pressure drop high efficiency and good cleaning characteristics over this 70 hr test period.
1WH 5	2 1 3.82	1500 (815)	150 (11)	1 512 700	5.7 (1.7)	350/24.8	130	12/2 (3/1.5)	99.99 ⁺	Continuation of previous test with same excellent performance after 51 hr testing.
1WH 5	2 5-10.82	1500 (815)	150 (11)	1 412 400	11.4 (3.5)	350/24.8	30	24/10 (6/4)	99.98 99.76	Doubled filter velocity to 11.4 ft/min and retained excellent repeatable performance through this two day test. Small penetration revealed during second day led to examination of filter. Both break across filter flat and delamination observed on removal of element from holder.
2WH 5	2 18.82	1500 (815)	150 (11)	9411 670	17.2 (5.2)	350 400 (24.8 26.7)	30	24/5 (6/1.25)	99.91	Very high velocity (17.2 ft/min) test of new piece of 144 um pore size material. Performed very well with good repeatable cleaning and high efficiency. Small penetration presumed to come from cracks in front sealing face where Arema cement was missed. No major failure but removed and sent to Coors for examination of front face cracks and crumbling that occurred on removal from filter holder.
2WH 5	3 4 9.82	1500 (815)	150 (11)	8711 550	17.2 (5.2)	350/24.8	30-40	24/2 12 (6/5-3)	99.99 ⁺ 99.56	Third piece of 144 um pore size material tested at very high filter velocities (17.2 ft/min) over two day period. Small drop in efficiency second day led to examination which revealed same type of breaks that had been observed in 1WH 5 and 2WH 5.
1WH 6	3 16.82	1500 (815)	150 (11)	1 111 900	17.2 (5.2)	350/24.8	50	24/11 (6/4.25)	91.17	New element fabricated from same material as previous element however filter flat 120 mils thick instead of 60 mils for added strength. Joints were not reinforced with phosphate adhesive at Coors suggestion. Element failed during first day of test in same fashion WH 5 elements had.
2WH 5	3 19-23.82	1500 (815)	150 (11)	2 3 3 (3 400 5 800)	17.2 (5.2) 11 (3.3) 5.7 (1.7)	350/24.8	2 8 12 24	24/18 (6/4.5) 24/12 (6/3) 24/5 (6/1.25)	99.99	New piece of 120 mil material run at high medium and normal filter velocities. First day of two day test very good performance. Second day cleaning remained good but efficiency fell off examination revealed delamination break across filter flat and front sealing face crumbling.
1WH 7	7 28.82	1500 (815)	150 (11)	2 614 700	11 (3.35)	250-350/ 18 24.8)	80	12/4 (3/1)	99.86	New material 60 mil 100 um pore size with improved bonding surfaces and machined sealing and clamping surfaces. Operated using new start up procedure that minimized thermal effects during heat up and shutdown. Typical long conditioning cycles during first exposure to dust.
1WH 7	7 30.82	1500 (815)	150 (11)	2 314 120	11 (3.35)	—	150	12/4 (3/1)	99.71	Combustion system difficulty caused early shutdown during this test. Filter was not blown back. Drop in efficiency was noted throughout tests.
1WH 7	8 2.82	1500 (815)	150 (11)	2 714 740	11 (3.35)	350/24.8	150	12/4 (3/1)	99.60	Unexplainably long cycle times and increasing penetration were indications of some difficulty. Filter was removed and found to be loose in clamping assembly but no failures were observed in element. Major source of leak found to be a warped clean/dirty plenum in filter housing. Placed extra gasket material at warped spot and reassembled unit for test.
1WH 7	8 5.82	1500 (815)	150 (11)	2 814 970	11 (3.35)	350/24.8	20-25	12/8 (3/2)	99.96	More reasonable cycle times and high efficiency indicate system leak had been fixed during this run. Filter performed very well.
1WH 7	8 9.82	1500 (815)	150 (11)	2 414 281	11 (3.35)	350/24.8	25	12/8 (3/2)	99.76	Filter performance continued to be excellent except for somewhat lower than expected efficiency. Filter removed after test and examined. No flaws were observed in the filter element suspect reoccurrence of system associated leak.

- 128 kg/hr, 811 K, 0-297 kPa (280 lbm/hr, 1000°F, 0-40 psig)
- 103 kg/hr, 811 K, 0-1114 kPa (225 lbm/hr, 1000°F, 0-150 psig)
- 91 kg/hr, 950 K, 0-743 kPa (200 lbm/hr, 1250°F, 0-100 psig)
- 46 kg/hr, 811 K, 0-1114 kPa (100 lbm/hr, 1000°F, 0-150 psig) - with filter
- 50 kg/hr, 811-1089 K, 0-1114 kPa (110 lbm/hr, 1000-1500°F, 0-150 psig) - with filter

During these tests, very low pressure drops across the clean filter (1WH-1) were observed [on the order of 0.25 kPa (1 in. wg) at face velocities of 1.2 to 1.5 m/min (4 to 5 ft/min)].

A check of Coors's literature for filters of similar pore size indicated that these pressure drops were consistent with established standards for clean ceramic filter bodies. The filter element and the mounting and sealing assemblies were all inspected subsequent to each high-temperature exposure and found to be in excellent condition.

During the next month, the first two high-temperature and -pressure filtration runs were completed, using redispersed PFBC fly ash. Both tests were conducted at a temperature of 813 K (1000°F) and a system pressure of 1114 kPa (150 psig). The air mass flow was set at 57 kg/hr (125 lbm air/hr), which yielded a filter face velocity of 1.7 m/min (5.6 ft/min). Clean filter pressure drops were on the order of 6 kPa (2.5 in. wg) and rose at a rate of about 0.1 kPa/min (0.4 in. wg/min) when exposed to a particulate loading of 13.4 g/Nm³ (10,400 ppm). The filters were cleaned by a pulse jet system when the pressure drop reached 3.7 kPa (15 in. wg). Pulses from a reservoir pressure of 1722 kPa (240 psig) lasting for one second were sufficient to return the filter pressure drop to a constant base of 1.25 kPa (5 in. wg). The rate of pressure drop rise was very much higher for all cycles after the initial cleaning, 0.75 kPa/min (3 in. wg/min). This was an indication that the pulse cleaning was effective for only a fraction of the filter area, which would "heal" rapidly upon subsequent exposure to dust flow. This was confirmed by visual examination of the

filter element in posttest inspections. These inspections showed that the lower half of the filter element was cleaned more effectively than the upper three inches. We believe this was due to the lack of jet expansion in the upper part of the filter.

An operating difficulty forced the system down for modifications to the exhaust system. The bypass line became plugged with dust expelled from the filter during blowback, which led to poor cleaning and, more importantly, disabled the pressure relief line for the system. We anticipated that the system modification would require about three weeks to implement.

During this time, we devised a mechanical means of closing the clean gas plenum for better cleaning, if required. This was an effort to improve the poor cleaning observed in previous tests. In parallel with this effort, we expended increased effort in the theoretical modeling of the filter cleaning process, since this appeared to be the central issue in the success of the concept.

During June, approximately ten hours of filter operation were accumulated over the course of three tests. The tests were all carried out at a temperature of 811 K (1000°F), pressure of 1114 kPa (150 psig), and filter velocity of 1.7 m/min (5.7 ft/min). Inlet dust concentrations ranged from 5.17 to 8.1 g/Nm³ (4000 to 6250 ppm), and average outlet concentration ranged from 0.0003 to 0.0008 g/Nm³ (0.2 to 0.6 ppm), indicating filter efficiencies in the 99.9 to 99.99% range. When we started with a clean filter, the pressure drop generally rose from about 1 kPa (4 in. wg) to 3 kPa (12 in. wg) in about an hour. One-second cleaning pulses of 1692 and 1854 kPa (235 and 260 psig) were not enough to prevent gradual shortening of the cleaning cycle and slow increase of the base-line pressure drop. This indicated incomplete filter cleaning, which we confirmed visually. At the end of the last test a cleaning regimen consisting of 2127 kPa (300 psig) pulses into the 1114 kPa (150 psig) filter ambient proved substantially better. Further, 2127 kPa (300 psig) pulses into a 780 kPa (100 psig) filter ambient proved even

more effective and were successful in returning the filter pressure drop to its original 1 to 1.12 kPa (4 to 4.5 in. wg) value. Subsequent examination of the filter body indicated substantially better cleaning.

Since we felt that progress was being made in understanding and improving the blowback process, we kept our efforts focused on this area. Work proceeded with the same filter element so that we could understand the results of the cleaning parameters, such as pulse pressure, duration, and volume. In parallel, we continued effort in the modeling task. The numerical model as it then existed correctly predicted preferential cleaning of the bottom of the filter and a gradual instability in the cleaning process. At this time we installed a differential pressure transducer across the filter in order to quantify the results of altered cleaning techniques.

Efforts in July concentrated on improving the understanding and performance of the blowback cleaning process. The filter system was equipped with a differential pressure transducer and a high-speed strip chart recorder that allowed the pressure rise on the clean side of the filter to be recorded during blowback cycles of varying pulse pressure, duration, and volume.

The initial tests were conducted without dust feed at 1114 kPa (150 psig) and ambient temperatures using 0.5 to 1 second pulses from a pulse reservoir charged to pressures of 1449 to 2127 kPa (200 to 300 psig). These tests showed a strong relation between the pressure rise at the bottom of the filter and the initial pulse pressure. Pressure rises ranged from 3.45 to 19 kPa (0.5 to 2.8 psi) over the range of blowback pressures tested.

A subsequent test was conducted at conditions of 811 K (1000°F), 1114 kPa (150 psig), airflow of 58 kg/hr (125 lbm/hr), average inlet dust concentration 5.86 g/Nm³ (4537 ppm). The cleaning pulse time employed was 1 second, with a pulse pressure of 2127 kPa (300 psig). The cleaning performance as reflected in filtration cycle time was much improved over previous tests using lower blowback pressures, but we did

observe a gradual shortening of cycle time. Visual inspection of the filter indicated good cleaning at the filter bottom and poor cleaning at the top. The measured overall efficiency of the filter was 99.94%, with an average outlet concentration of only 0.003 g/Nm^3 (2.0 ppm).

Minor modifications were made to the blowback system to allow investigation of pressures higher than 2127 kPa (300 psig). Subsequently, a high-pressure and -temperature run with dust was made in which pulses up to 2806 kPa (400 psig) were used. Unfortunately, plugging of the pressure tap lines during the test made reliable measurements of pressure drop and pressure rise impossible. Posttest inspection confirmed a more complete cleaning of the filter lower half but little improvement in cleaning the upper half.

The final test of this period was carried out at temperature and pressure but with no dust injection. Another differential pressure transducer was obtained, and high side taps were placed 2.54 cm (1 in) from the top and bottom of the clean side of the filter element in an attempt to understand the nonuniform cleaning observed. A wide range of blowback parameters were investigated including the following:

pulse time: 0.5, 1.0 s.

pulse pressure: 1449 to 3444 kPa (200 to 500 psig)

filter ambient pressure: 608,780, 1114 kPa (75, 100, 150 psig).

At high pulse pressures and lower passage pressures, the bottom of the clean-side plenum reached peak differential pressures of up to 62 kPa (9 psi). Although the differential pressures at the top of the filter (near the blowback nozzle) were consistently lower and occurred more slowly than at the filter bottom, the difference measured between them did not seem to explain the observed gradient in cleaning. In fact, we subsequently discovered that the pressure "rise" observed at the top of the filter was actually negative, the suction resulting from a venturi effect.

During the month of August a great deal of progress was made in quantifying the blowback parameters required for steady filter operation. Blowback tests were conducted at 811 K (1000°F) and at system pressures of 608, 790, and 1117 kPa (75, 100, and 150 psig), with pulse pressures from 1449 to 3475 kPa (200 to 500 psig). Using two differential pressure transducers located at the filter top and bottom in the clean side, we again measured the transient pressure behavior of the various pulses. The cleaning pulses caused a pressure rise of up to 34.5 kPa (5 psi) at the filter bottom, but there was a pressure drop of nearly the same magnitude at the filter top near the blowback nozzle. This phenomenon was apparently caused by an eductor effect of the blowback jet in the clean-side plenum. This effect on cleaning was confirmed visually by posttest examination of the filter cake distribution. As expected, the top of this filter was plugged and showed no sign of cleaning, while the bottom half of the filter had been cleaned very well. A simple remedy to the problem proved effective. We merely positioned the blowback nozzle about 3.8 cm (1.5 in) above the cleaned gas outlet in the clean-side plenum. Subsequent transient pulse measurements and visual observations of the posttest dust cake distributions confirmed the effectiveness of this procedure.

Prior to this modification 13.3 hours of testing at 811 K (1000°F), 1114 kPa (150 psig) had been accumulated. Dust inlet loadings were typically 2.58 to 3.23 g/Nm³ (2000 to 2500 ppm), with a filter velocity of 1.7 m/min (5.7 ft/min). Cycle times for filter pressure drop to rise from about 1.25 to 3 kPa (5 to 12 in. wg) typically started at about 1 hour and gradually shortened throughout this test period as the upper half of the filter element became more and more plugged. Subsequent to the blowback nozzle modification, we accumulated about 22 hours of testing at the above conditions. The pressure drop vs time behavior of the system was much more stable under these conditions. A series of three, 0.5 s pulses from a 3475 kPa (500 psig) source was enough to maintain a stable cycle time of about 40 to 60 minutes and base-line pressure drop of about 1.25 kPa (5 in. wg) over the test period.

During this testing period a sealing problem emerged as a result of the more energetic blowback regimen in use. Several times the fibrous material used to seat the ceramic element in the metallic holder was punctured during blowback.

The main goal of the work carried out in September was to understand the problem of the apparent leaks in the filter system that resulted in typical efficiency degradations of 99.9+% to 98-99%. Another goal was to gain further experience with operating the filter and system at high temperature and pressure.

During August and throughout September, some measurable penetration was observed. Analysis of the problem led to the conclusion that the source could have been from any of four sources:

- Leaks around the filter/holder sealing gaskets
- Dust blown around the bypass loop and into the clean-side exit during blowback
- Leaks in the clean-side/dirty-side chamber gasket
- Defects or holes in the filter element itself.

Experiments conducted over the course of the month led us to believe that at various times all four mechanisms had contributed to the penetration.

The initially observed leaks were almost certainly caused by failure of the filter element sealing gasket material at high blowback pressures. Improved gasket materials, the use of ceramic cements, and improved clamping techniques minimized this problem.

The combined eductor effect and momentary overpressurization of the dirty side during blowback can lead to dust movement down the clean-side exit. By sampling only through blowback sequences and by using a labeled dust, we confirmed this mechanism but found it to be only a minor source of leaking, about 0.2 to 0.5 mg sample filter weight gain per blowback sequence, or a degradation of 0.006% in efficiency per blowback sequence.

The potential for penetration of the clean-side/dirty-side filter chamber existed because of damage to the insulation around the filter chamber liner and to the gasketing material in that location. Although significant air leaks were detected, large amounts of dust could probably not have penetrated at this point. Appropriate repairs were made, and subsequently routine inspection of this area was instituted.

Near the end of the month, an actual filter leak in the element 1WH-1 was detected and determined to have arisen from a simple delamination of one layer of ribs from a filter flat. This failure occurred after approximately 35 thermal cycles, 500 blowback pulses, 100 hours of operation at temperature and pressure, and 13 mountings and dismountings. While this type of failure was of concern, consultation with Coors indicated several methods by which to improve the resistance to failure in this mode, such as providing a better mechanical joint and using high-temperature adhesives and higher firing temperatures.

After the filter failure was detected, a new element (2WH-1) was cut from the same original ceramic body and was installed and tested at 811 K (1000°F), 1114 kPa (150 psig), with an inlet concentration of 8.27 g/Nm³ (6400 ppm) and a filter velocity of 1.7 m/s (5.7 ft/min). The test lasted a total of 380 minutes and spanned two separate days. The measured overall efficiencies were 99.97 and 99.94% for the two days, indicating that virtually all sources of penetration had been eliminated during these runs. Blowback cycles consisted of 2806 kPa (400 psig) pulses for 0.5 s, and these were sufficient to maintain cycle times at about 1 hour, during which time the filter differential pressure would rise from about 0.25 kPa (1 in. wg) to 3 kPa (12 in. wg). In total, about 20 hours of actual filter operation at pressure and temperature were accumulated in September.

During October a total of six high-temperature and -pressure runs were made, two at 811 K (1000°F) and four at the full system temperature of 1088 K (1500°F). Tests at the higher temperatures indicated the need for a few minor modifications to the test passage

liner system and the filter clamping and sealing assembly. The installation of thermal expansion bellows in the liners and the use of titanium hold-down studs that match the thermal expansion of the ceramic filters have reduced the system-associated test problems that arose with the higher operating temperatures.

During the first two tests at a temperature of 811 K (1000°F) the performance of the 100 μm pore size, 1.5 mm (60 mil) thick filter (2WH-1) deteriorated slightly from 99.90% to 99.70%, indicative of a small leak. During a subsequent test of the filter at 1088 K (1500°F), the efficiency declined markedly to 98.60%, even though the pressure drop and cleaning characteristics of the filter remained quite good, with a roughly 200 minute cycle time from a steady base line of about 0.75 kPa (3 in. wg) to 2.5 kPa (10 in. wg) at a filter velocity of 1.7 m/min and an inlet loading of 6.6 g/Nm³ (5128 ppm). Subsequent examination of the filter revealed that the filter had become quite loose in the holder and had failed because of delamination of the bottom plate and cracking of the next two layers of the filter.

A sample of what was thought to be 100 μm pore size, 1.0 mm (40 mil) filter medium (1WH-3) was cut to size and installed for testing at 1088 K (1500°F). Over two consecutive test days the filter efficiency was relatively constant at 99.70 to 99.49%, which by past experience indicated a small leak. We discovered that this medium had a somewhat higher pressure drop than the thicker, 1.5 mm (60 mil) medium used previously. Its clean base-line resistance was about 2 kPa (8 in. wg), compared to about 0.75 kPa (3 in. wg) for the thicker material at a filter velocity of 1.7 m/min (5.7 ft/min). Posttest inspection of the filter revealed another delamination, but thermal expansion problems with the system's liner and poor clamping practice left the exact cause of the failure in doubt. We later determined that this material actually had 10 μm pores, which explained the higher than expected pressure drop.

The final run of October was made with a new piece of this same material (2WH-3) and appeared to be successful, with efficiencies in the 99.97% range. The liner modifications were implemented prior to this test, and both the clamping and liner expansion problems seemed to have been corrected.

During November a series of five tests lasting from one to three days were carried out on the cross-flow filter system. Nominal conditions for all tests were $T = 1088 \text{ K}$ (1500°F), $P = 1114 \text{ kPa}$ (150 psig), $V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min), with an inlet dust loading of approximately 7.75 g/Nm^3 (6000 ppm). All five tests eventually ended with a filter failure when one or more layers of the cross-flow medium delaminated from adjacent layers. The first and last tests were conducted on what was thought to be $100 \text{ }\mu\text{m}$ pore size by 1.0 and 1.5 mm (40 and 60 mil) back web (i.e., 2WH-3 and 1WH-4). The clean filter pressure drops measured were much higher than anticipated from previous experience, with 1.5 mm (60 mil) material ranging from 2 to 2.75 kPa (8 to 11 in. wg) instead of the previously observed 0.12 to 0.25 kPa (0.5 to 1 in. wg). This ambiguity was clarified when Coors discovered that the actual pore size was $10 \text{ }\mu\text{m}$ instead of the specified $100 \text{ }\mu\text{m}$. The remaining three tests were conducted on two pieces of cross-flow material cut from the same block of 1.5 mm (60 mil) back web and $10 \text{ }\mu\text{m}$ pore size, that is elements 1WH-1 and 2WH-1.

The delamination failures appeared to occur because of poor bonding of the $10 \text{ }\mu\text{m}$ material, with consequent joint failures during thermal cycling and the imposition of high-pressure stresses caused by the high blowback pressures required to clean these high-pressure drop materials.

New $100 \text{ }\mu\text{m}$ by 1.5 mm (60 mil) back web media were ordered, and we anticipated that a higher firing temperature and lower pressure pulse cleaning would improve filter durability.

During the first week of December a long-duration, around-the-clock test was carried out using the last piece of filter element,

3WH-1, which was 1.5 mm (60 mils) thick, with a nominal pore size of 100 μm . Over 80 hours of actual filtration test time were accumulated at 1088 K (1500°F), 1114 kPa (150 psig), with a filter velocity of 1.7 m/min (5.7 ft/min). The pressure drop cycle became stable after a brief conditioning period and remained unchanged for the duration of the test. With an inlet dust concentration of 5.2 g/Nm³ (4000 ppm), this cycle started at a base-line pressure drop of 1 kPa (4 in. wg) and rose to the trigger level of 3 kPa (12 in. wg) over a time period of about 45 to 60 minutes. The filter was successfully cleaned using a 0.5 s pulse of N₂ from a 2512 kPa (350 psig) source. The overall efficiency measured throughout the test was constant at 99.94% and represented a penetration level below our limits of reliable detection. Because of periodic plugging in the bypass line and the need to change propane tanks, the run was interrupted for about 2-hour intervals 14 times over the test period. During each of these short periods of downtime, the filter was thermally cycled from 1088 K (1500°F) to about 477 K (400°F).

Subsequent to this long duration test, three more tests were carried out at twice the filtration velocity, 3.5 m/min (11.4 ft/min). Throughout these tests the pressure drop cycles remained stable with the same cleaning regimen. The blowback trigger pressure drop was increased to 6 kPa (24 in. wg) without any observable reduction in filter cleanability. During the second of these three high-velocity runs, some certain penetration of dust was observed and continued throughout the third test. The overall efficiencies for these tests were 99.77 and 99.61%, down from the previous values of 99.94 and 99.89%. The exact source of this small amount of penetration had not been determined at that time.

During January 1982, a total of eight, one-day tests were carried out using two ceramic filter elements, 3WH-1 and 1-WH5. Four tests were carried out using the same piece (3WH-1) of 100 μm , 1.5 mm (60 mil) filter element used in the previous month's long-duration test. Then four more tests were conducted on a new, more porous, filter element (1WH-5) Coors supplied. This element had a nominal maximum pore

size of 144 μm , with a 1.5 mm (60 mil) backweb. All testing was carried out at 1088 K (1500°F) and at a total pressure of 1114 kPa (150 psig).

The objective of the first four tests was to investigate that filter's performance at higher filter velocities. During these tests the face velocity was increased from the 1.7 to 3.4 to 4.3 and, finally, to 5.2 m/min (5.7, 11.4, 14.2, 17.2 ft/min). During these tests we attempted to maintain the dust concentration at its previous nominal level of 5.2 g/Nm³ (4000 ppm), but in some tests the dust feed system could not quite keep up with the increased airflow, with the result that some inlet loadings dropped to values as low as 3.0 g/Nm³ (2300 ppm). Perfectly stable and predictable cycles with acceptable cleaning were easily maintained for all velocities except 5.2 m/min (17.2 ft/min). At this highest velocity the base-line pressure drop was 5 to 5.5 kPa (20 to 22 in. wg), and the pressure drop instrumentation used was spanned only to 6.25 kPa (25 in. wg), so the cycle times were very short (1 to 2 min). This situation made conclusions about filter cleaning at this very high velocity uncertain, although there was indication that the cleaning under these conditions was incomplete. Throughout these tests the overall filter efficiency remained virtually constant at 99.8%, indicating a small but constant leak. Visual inspection of the filter after the testing had been completed revealed a small crack and partial delamination of the 3WH-1 filter element, which is presumed to be the source of the observed penetration.

This element had accumulated over 120 hours of filter test time at 1088 K (1500°F) and 1114 kPa (150 psig). It had been blown back (pulse cleaned) 602 times and thermally cycled from ambient to 1088 K (1500°F) and back down a total of 25 times.

The next (more porous) element, 1WH-5, was tested at the more conservative velocity of 1.7 m/min (5.7 ft/min), for a total of 21 hours during January. This filter performed very well, with a base-line pressure drop typically about 0.5 kPa (2 in. wg). Cleaning was accomplished with a simple pulse from a 2127 to 2462 kPa (300 to 350 psig)

reservoir. Virtually no penetration was measured as efficiencies exceeded 99.99%. Subsequent testing of this element was aimed at higher-velocity operation.

Testing carried out in February centered on work with the new filter material supplied by Coors (WH-5). This large-pore-size material was achieved using a technique not employed in any of the less porous elements supplied to date. This method consisted of blending rubber particles in the green ceramic mix, creating the desired pore size and porosity when the material was fired.

The first element tested (1WH-5) performed exceptionally well during a series of tests to establish a base line. These tests were carried out at 1088 K (1500°F), 1114 kPa (150 psig), at a face velocity of 1.7 m/min (5.7 ft/min). With a dust loading of 3.4 to 4.5 g/Nm³ (2600 to 500 ppm), the typical filter cycle from a steady base of 0.5 kPa (2 in. wg) to the trigger pressure drop of 4 kPa (12 in. wg) took about 1.5 hours. Successful cleaning was achieved with a reverse flow pulse from a 2462 kPa (350 psig) reservoir. Over 50 hours of test time were logged at these conditions during a period of six separate tests days. Throughout this period the measured overall efficiency was always greater than 99.99%. Subsequently, tests were run at twice this velocity 3.5 m/min (11.4 ft/min). After a period of time, stable pressure drop cycles were established and maintained that rose from a base pressure drop of 2 to 2.5 kPa (8 to 10 in. wg) to the trigger pressure drop of 6 kPa (24 in. wg) over about a 0.5-hour period, with the dust loading at 3.2 to 4.5 g/Nm³ (2500 to 3500 ppm). During the second day of these high-velocity tests, a significant penetration of dust was observed as the efficiency fell to 99.76%. Examination of the filter element revealed a series of breaks in it, some of which occurred across the filter flats, rather than the simple delaminations that had been observed previously.

A second piece of this same material (2WH-5) was prepared for testing and was used at a very high filter velocity, 5.2 m/min

(17.2 ft/min). After only one cycle of conditioning, the filter settled into a stable, repeatable, filter cycle in which the pressure drop rose from a base line of about 1.5 kPa (6 in. wg) to the trigger point of 6 kPa (24 in. wg) over about a 30 minute period when the inlet dust loading was 2.2 g/Nm^3 (1673 ppm). A very small leak was observed, lowering the measured efficiency to 99.91%, so the filter was removed for examination. A few minor cracks were observed at a joint where preparatory gluing had been missed. Additionally, some crumbling at the clean face/gasket interface was observed, although that did not appear to be the source of the minor leak. Both of the most recently tested filter elements (1WH-5 and 2WH-5) were shipped to Coors for examination and analysis.

During March testing continued on very porous filter elements having a pore size of $144 \text{ }\mu\text{m}$. The first testing was a continuation of work started the previous month with filter elements having a backwall (filter flat) thickness of 1.5 mm (60 mils). The testing later in the month centered on elements of the same material but of twice the filter flat thickness 3 mm (120 mils). All testing was carried out at 1088 K (1500°F) and at 1114 kPa (150 psig).

The first test of this period was carried out with the third piece of the $144 \text{ }\mu\text{m}$, 1.5 mm (60 mil) filter medium (3WH-5). Because of the low resistance to flow, this two-day test was carried out at 5.2 m/min (17.2 ft/min). During the first day of testing the base-line pressure drop rose slowly from an initial clean value of 0.37 kPa (1.5 in. wg) to a constant value of 1.25 kPa (5 in. wg). At an inlet concentration of 0.98 g/Nm^3 (1736 ppm) the cycle time was about 35 minutes from the base-line pressure drop to the cleaning setpoint of 6 kPa (24 in. wg). Virtually no penetration was observed during the first day. During the second day of testing with these elements at similar conditions, the cycle time shortened to 20 to 25 minutes as the base-line pressure drop rose to 2.5 kPa (10 in. wg). The efficiency dropped to 99.56% and the penetration increased throughout the test. Posttest examination of the filter revealed that it had failed in a

manner similar to that of the previous two elements of this type, 1 and 2WH-5.

The remaining two tests were carried out on elements of the same material but with twice the filter thickness (1WH-6 and 2WH-6). The first of these tests was carried out at 5.2 m/min (17.2 ft/min) as before, but no preparatory gluing was used because Coors was uncertain of the effect of using the adhesive. The pressure drop and cleanability of the filter were excellent and predictable from the thinner filter experience. Significant penetration was measured from the start, however, as evidenced by an average overall efficiency of only 91.17%. Posttest examination of the filter (1WH-6) revealed that it had broken in several places in a fashion similar to that of the previous tests with the thinner wall material. Subsequent to this test, a second element (2WH-6) was prepared in the usual way by reinforcing joints and sealing the element in the filter holder. Tests were carried out at velocities of 1.7, 3.4, and 5.6 m/min (5.6, 11.4, and 17.2 ft/min) over a period of two days. During the first day of testing, the filter was exposed to very high dust loadings ranging from 4.4 to 7.5 g/Nm³ (3400 to 5800 ppm). At the high velocity and dust concentration, cycle times were very short, only 2 to 3 minutes, but at the intermediate and lower velocities acceptable, constant cycles were established and maintained. Throughout this day the filter efficiency was constant at 99.99+%. The second day of testing was confined to the lower two velocities, and, as before, stable cycles were established and maintained. From the start of the second day, some measurable penetration was obtained. At this lower velocity an average efficiency of 99.70% was obtained, and this declined to 99.41% for the 3.4 m/min (11.4 ft/min) run. Examination of the filter after the test indicated that it had broken in a manner similar to that of all the previously tested, high-porosity, 144 μ m filters.

After we consulted with Coors, they began formulating a new batch of material to provide one more set of filters with properties similar to the original material tested (WH-1), but the elements were to

incorporate all the fabrication and firing improvements that had been developed throughout this program.

No actual filters were tested during the eleventh quarter, as the last set of filter elements was still being fabricated at Coors. We had been concerned that in some way the recent mechanical failures of the test filters may have been due to thermal effects during start-up and shutdown. During start-up, for example, stable combustion could be maintained only with air temperatures at about 811 K (1000°F), so both the filter element and the metallic holder were exposed to high rates of heating. Conversely, on shutdown past procedures called for maintaining purge flow after flameout in the combustor, with consequently high cooling rates. In an attempt to mitigate the start-up effects, a dilution air line was plumbed into the inlet sample port. The combustor could then be operated stably at high temperature, while the downstream dilution ensured cooler initial temperatures in the filter chamber. A few preliminary experiments were carried out to try the new system, and it worked well. The initial filter temperature rise could be limited to 422 to 478 K (300 to 400°F) and the subsequent rate of temperature increase could be controlled at will. Additionally, a new shutdown procedure was initiated whereby both the propane and airflow could be shut simultaneously so that the filter and holder temperature would decay very slowly by heat loss through the chamber walls. We anticipated that these procedures would maximize the filter element resistance to mechanical failure in the final test series.

The final set of four filter elements (WH-7) was received late in July. These filters were a nominal 1.5 mm (60 mil) thick filter flat with a 100 μ m pore size. In an attempt to avoid the delamination problem, we made some changes in these filters. We fabricated a wide 1.27 cm (0.5 in) bonding edge around the perimeter of the filter. All faces of the 5.1 $\times$ 15.2 $\times$ 15.2 cm (2 $\times$ 6 $\times$ 6 in) element also had been machined to provide square, flat surfaces. Incorporation of the wide bonding perimeter resulted in a reduced effective filter area for these elements. An area of about 662 cm² (0.713 ft²) is estimated per

element. When examined, two of the four filters were found to be flawed. Coors is currently firing two replacement filters; the remaining two filters were prepared for testing.

A final series of five, one-day-long tests were conducted on the WH-7 filter element Coors supplied. Test conditions were held constant throughout the test series at 1088 K (1500°F), 1114 kPa (150 psig), filter velocity of 3.35 m/min (11 ft/min), and a dust concentration from 5.2 to 6.5 g/Nm³ (4000 to 5000 ppm). The trigger pressure drop at which blowback was initiated was 3 kPa (12 in. wg). The blowback pulse was provided by a 2512 kPa (350 psig) reservoir pulsing through a 0.86 cm (0.34 in) nozzle for 0.5 seconds.

During the first tests the "conditioning" period seemed to last longer than expected, as reflected in unusually long cycle times. The third day test, for example, began with a cycle time of 150 minutes, during which time the pressure drop rose from 1 to 3 kPa (4 to 12 in. wg). Additionally, considerable dust penetration was observed as the overall efficiency over the first three days fell from 99.86 to 99.71 to 99.60%. Subsequent examination of the filter revealed no cracks or obvious leaks, but there was a significant leak between the dirty-side and clean-side chambers caused by a warp in the plenum. This difficulty was addressed by adding extra gasketing material in the warped area and by being more careful about twisting down the plenum plate studs.

The next two tests progressed in an expected manner. The cycle times became steady at 20 to 25 minutes, and the base pressure drop leveled out at about 2 kPa (8 in. wg). Additionally, the overall efficiencies measured were higher, 99.96% the first day and 99.76% the second. When the filter was removed from the test unit for the final time, we again found it in excellent condition, with good and uniform cleaning. Some dust, however, was evidently leaking past the plenum, as we had observed in the previous tests.

During this test series the filter element WH-7 was operated for a total of 21 filtration hours and pulsed a total of 70 times. The

system was thermally cycled five times and the filter dismounted two times with no mechanical failure. While a longer test period would be desirable for endurance testing, we felt that the WH-7 filters were a significant improvement over any of the filter materials tested recently and represented a definite step forward in element design and fabrication.

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3.3.4 Detailed Test Program Results

The following section presents a detailed compilation of the evolution of the test program as excerpted from monthly progress reports.

By the end of March, 1981, most of the system shakedown had been completed and the first test filter elements had been installed. The first test element (LWH-1) was cut to size [5 x 15 x 15 cm roughly (2 x 6 x 6 in.)] and installed in the filter element housing. We estimate the free filter area offered by this element to be 0.11 m^2 (1.2 ft^2). The element chosen was the 100 μm pore size with a 1.5 mm (60 mil) thick membrane.

Test 3.25.81 was carried out at a mass flow of 46 kg/hr (100 lbm/hr). First a temperature of 811 K (1000°F) was established, and then system pressure was raised to 1114 kPa (150 psig). As in previous shakedown tests, the flame became quite luminous at high pressure. We decided that the newly installed air premix would not be used in this test in the hope that the oil filter would solve the soot problem experienced during previous shakedown runs. The ceramic filter pressure drop rose to about 0.75 kPa (3 in. wg) during this period, even though the filter bypass line was open. Passage temperature was then increased to 1088 K (1500°F) and steady operation maintained. At this time the filter pressure drop decreased to a level that we could not accurately measure. Just prior to shutdown there appeared to be some obstruction in the combustor section around the flameholder. Inspection of the combustion section the following day revealed that large chunks of carbon had been formed in the area around the flameholder. The deposits were so large that the normal flame pattern had been distorted,

causing direct impingement on the combustor liner and subsequently melting about a 3.8 cm (1.05 in) diameter region in the Inconel wall. Although the wall section bulged, it was still serviceable. Additionally, cooling air had not been provided for the forward dust injector nozzle and it had burned off. The filter element housing was opened to inspect the filter since there had been an unexplainable decrease in filter pressure drop when the system temperature was raised from 811 to 1088 K (1000 to 1500°F). Aside from some carbon dust, the filter appeared to be in good shape and fiber frax seals seemed to be intact. The element still mounted in its holder is shown in Figure 3.3-1.

Figure 3.3-1 Filter Element in Mounting

A reasonably large volume of water was discovered in the bottom of the bypass line, the lowest point in the piping. This water was removed, and plans were made to allow on-line blowdown in future runs. At the lower temperatures the accumulated water in the bypass line may have effectively sealed the bypass, even though the valve was open, thereby forcing flow through the filter element and creating the observed pressure differential. As the system temperature was increased, this

water seal may have been vaporized or blown over, opening the bypass line and resulting in the observed lower filter pressure drop.

Test 3.27.81 was conducted to try the air/propane premix option for prevention of carbon deposition. Mass flow for these tests was set at about 50 kg/hr (110 lbm/hr). As in previous runs, a stable passage temperature of 811 K (1000°F) was established, and then the pressure was incremented. By adjusting the premix air, we could maintain a steady blue flame at pressures up to about 371 kPa (50 psig). Above that pressure the flame became a bright yellow-white but remained well defined and stable. Passage pressure was brought up to 1114 kPa (150 psig), and the temperature was brought to 1088 K (1500°F). Stable operation was easily maintained, and the flame looked much improved. The pressure drop behavior of the filter was much the same as in the previous test. During 811 K (1000°F) operation, a small pressure drop of about 0.5 kPa (2 in. wg) was observed. As before, it disappeared, however, at about the same time as the temperature was increased to 1088 K (1500°F). Subsequent inspection of the combustor revealed no carbon buildup of any significance. The filter housing was again opened and the filter inspected. As before, it seemed in good shape.

While the outlet section of the filter housing was off, a quick pressure drop flow experiment was conducted at ambient pressure and temperature. The filter element was replaced in the filter housing, and the high (dirty)-side pressure tap was connected to a manometer. The low side of the manometer was left unattached, open to the atmosphere, as was the clean side of the filter. The pressure drop vs filter velocity curve shown in Figure 3.3-2 was generated. The clean filter is remarkably permeable, giving a pressure drop of only 0.25 kPa (1 in. wg) at a filter velocity of 2.44 m/min (8 ft/min). This resistance to flow was checked against literature Coors had published for filters of similar pore size and was found to be in general agreement.

During April the first two actual filtration tests were carried out using PFBC dust. In the first test (4.15.81) the system was first

Curve 727760-A

Figure 3.3-2 Pressure Drop vs Filter Velocity for Ceramic Filter 1WH-1

brought to a temperature of 811 K (1000°F) at a mass flow of 57 kg/hr (125 lb/hr) and stabilized. Subsequently, the system pressure was raised to 1114 kPa (150 psig). Prior to initiating dust injection, we forced the air through the filter by closing the bypass valve. A filter pressure drop of 1.61 kPa (6.5 in. wg) was noted, and then the filter was blown back twice, once for 100 milliseconds (ms) at 1367 kPa (175 psig), then for 300 ms. The pressure drop declined to 0.75 kPa (3 in. wg) but gradually rose to a steady 1.5 kPa (6 in. wg).

At this point both dust feeding and isokinetic sampling of the filter outlet were started. The dust feed motor was set at 3.2 rpm, or approximately 3.5 g/min. Within five minutes the dust feed line had plugged, and efforts to blow it out were unsuccessful. At this point the first dust feed line was shut and the second feed line opened to bleed down the pressure in the dust injector. During this period considerable dust was blown into the system, even though the metering

system was off. This dust could be seen through the combustor sight port and is believed to have been the dust trapped in the common line that fed both dust feed lines. During this bleed downtime from 1479 to 1114 kPa (200 to 150 psig), the filter pressure drop rose from 0.87 to 1.5 kPa (3.5 to 6 in. wg). Three cleaning pulses of 1307, 1479, and 1753 kPa (175, 200, and 240 psig) for 500 and 1000 ms were administered, and the pressure drop fell to 0.75 kPa (3 in. wg). Subsequently, dust feed was started again through the second dust feed line and appeared to proceed smoothly. "Puffs" of dust could be observed in the combustor as the discrete "packets" of dust were delivered by the dust injector. This dust eventually filled the rear portion of the combustor entrance and blocked the viewing port. Dust feed proceeded without incident for about 75 minutes, during which time the pressure drop across the filter rose slowly from 0.75 kPa (3 in. wg) to 1125 kPa (5 in. wg). At this time the dust feed was shut off and the system blown back with two pulses. The first pulse was 1479 kPa (200 psig) for 1000 ms, and the second was 1753 kPa (240 psig) for 500 ms. The first pulse did not accomplish any ΔP recovery, but the second pulse reduced the pressure drop back to the original 0.75 kPa (3 in. wg). The system was then shut down by normal procedures and readied for inspection.

The various systems were opened the following day for inspection.

Dust Injector System

The dust feed line that introduces dust ahead of the combustor flame was found to be plugged, for no apparent reason, at the fitting that penetrates the combustor. About one-third to one-half of the 1120 g of PFBC ash originally placed in the dust hopper had been fed, corresponding to an approximate dust feed ratio of 3 to 5 g/min, or a concentration about 6.5 g/Nm^3 (5000 ppm). When the combustor section was opened, however, a large amount of dust was found to have accumulated in the large, cross-sectional area at the rear of the combustor, where the dust was introduced. Because it was difficult to measure this material,

the actual dust feed rate to the filter was very uncertain, but surely less than the 3 to 5 g/min anticipated. This difficulty was eliminated by extending the dust feed tube into the smaller 5 cm (2 in) section of the combustor where the velocity is much higher.

Filter Element

The filter element housing was opened and the actual filter block removed for inspection. Figure 3.3-3 is a photograph of the element in as-removed condition. A uniform layer of dust approximately 1.6 mm (0.06 in) thick covered most of the outer surfaces. Inlet holes near the top of this element were almost closed by this layer, while holes at the filter bottom were for the most part open. Most channels had an accumulation of dust in them of roughly the same depth as the outside surfaces. In one region in the center of the element the dust appeared to have shed from the surface in relatively large, 1.9 cm (0.75 in) flakes. The dust was not sticky or hard but retained its fluffy, soft, cohesive nature. It was easily removed by running a vacuum cleaner down a section, as indicated in Figure 3.3-3. The outlet, or clean, side of the filter looked very clean, with only the smallest trace of dust on the clean-side surface.

The absence of dust in the outlet was confirmed by the outlet sample. Visual inspection of the outlet total filter revealed that some soot had been collected, but no dust was visible. The total mass collected on the filter was found to be 1.65 mg, which corresponds to an average outlet concentration of 0.0006 g/Nm^3 (0.48 ppm) soot plus dust. The filter was vacuumed clean and remounted in the system.

Several minor system modifications were made while the facility was apart, including:

- Installing flowmeters in the air/propane premix line and in the Fireye purge line
- Installing a new flameholder that was intended to give better combustion.

As Removed

Vacuumed Section

Clean Side Surfaces

Figure 3.3-3 Cross-Flow Element after First HTHP Test

Having accomplished these changes, we tried a second run. At the start we could not seem to keep the new flameholder lighted. Ignition would proceed as normal, but we were unable to maintain a stable flame. After several tries, the old flameholder was installed and the system worked. By this time it was too late to follow through with a "dust" run, so this run was used as a system test run. The passage was brought to a flow of 57 kg/hr of (125 lbm/hr), 811 K (1000°F), and 1114 kPa (150 psig). After checking for leaks and proper operation of the instrumentation, we shut the system down and prepared for a subsequent dust injection run.

For the third run (Test 4.23.81) 1200 g of Exxon PFBC ash was loaded into the dust injector. The system was brought to a flow of 57 kg/hr (125 lb/hr), 811 K (1000°F), and 1114 kPa (150 psig). The dust injector and total filter isokinetic outlet sampler were started simultaneously. A record of the pressure drop vs time for this test is presented in Figure 3.3-4. Dust was initially fed through the line that introduces the dust downstream of the combustor. As in the last run, this line plugged within 5 minutes. The line was successfully opened using the new blowdown system, but the feed continued to be erratic because of plugging. After about the first hour of feeding dust, the feed line was switched so that the other dust feed line was used. At this point ($t = 60-65$ min) the dust feed went smoothly and continued that way for the remainder of the test. The pressure drop had then reached about 2.25 kPa (9 in. wg) from an original level of 0.5 kPa (2 in. wg). While dust was being fed through the second line, the pressure drop rose smoothly from 2.25 to 3.75 kPa (9 to 15 in. wg) (65 to 80 min). At this point the dust feed was shut off and we attempted filter blowback.

The filter blowback sequence requires that the filter bypass line be open prior to initiating a pulse. We noted that when we opened the bypass valve, the pressure across the filter did not drop, indicating a possible high resistance to flow (plug) in the bypass line. In spite of this, a sequence of pulses were administered to the filter as follows:

Figure 3.3-4 Pressure Drop vs Time for Cross-Flow Filter Experiment.
 Test 4.23.81.

- 1479 kPa (200 psig), 500 ms, no change
- 1479 kPa (200 psig), 1000 ms, no change
- 1651 kPa (225 psig), 500 ms, 3.75 to 3.5 kPa ($\Delta P = 15$ to 15 in. wg)
- 1651 kPa (225 psig), 1000 ms, no change
- 1651 kPa (225 psig), 2000 ms, [$\Delta P = 3.5$ to 3.25 kPa (14 to 13 in. wg)]
- 1702 kPa (240 psig), 1000 ms, [$\Delta P = 3.25$ to 2.5 kPa (13 to 10 in. wg)]
- 1602 kPa (240 psig), 1000 ms, [$\Delta P = 2.5$ to 1.6 kPa (10 to 6.5 in. wg)]
- 1702 kPa (240 psig), 1000 ms, [$\Delta P = 1.6$ to 1.5 kPa (6.5 to 6 in. wg)].

With the bypass closed again the pressure drop steadied out at a level of 1.5 kPa (6 in. wg), and dust feeding was restarted.

The differential pressure rise during this cycle was very much more rapid than previously, going from 1.5 to 3.75 kPa (6 to 15 in. wg) in about 3 minutes. The dust feed was shut off and the bypass checked for blockage. When the bypass was opened, the filter pressure drop fell as it should and is shown as the * at $t = 103$. Two pulses were then administered, both at a pressure of 1702 kPa (240 psig) for 1000 ms. The first dropped the filter ΔP from 3.75 to 1.75 kPa (15 to 7 in. wg) and the second to 1.25 kPa (5 in. wg). Dust feed was then resumed and, as before, the pressure drop rose rapidly. For a brief period in this cycle the dust injector vibrator was shut off, as is reflected in a slowing of the rate of rise in the ΔP vs t curve. At 3.75 kPa (15 in. wg) the dust was shut off and, as before, two cleaning pulses were administered, lowering the pressure drop to 1.25 kPa (5 in. wg). This cycle was followed by two more nearly identical cycles, where the pressure drop rose from 1.25 to 3.75 kPa (5 to 15 in. wg) in about 3 to 5 minutes and was then returned to 1.25 kPa (5 in. wg) with a single pulse of 1702 kPa (240 psig) for 1000 ms. During the following two cycles the dust injector vibrator was shut off, resulting in a marked decrease in dust feed rate, as reflected in the lower slope of the ΔP vs t curve. The

time to reach 3.75 kPa (15 in. wg) was about 8 minutes during these cycles. The first blowback was successful at 1702 kPa (240 psig) and 1000 ms. The second blowback, however, was unsuccessful, as there was preliminary indication that the bypass line had plugged. Repeated blowback attempts could not get the filter ΔP below 3.25 kPa (13 in. wg), and the pressure drop remained the same with the bypass line open or closed. At this time the system was shut down for inspection.

The dust injector vessel was opened, and 750 g of dust were found to have been fed to the system, corresponding to feed rates of about 12 g/min with the vibrator on and 4 g/min with it off. The average filter inlet concentration was calculated at 13.4 g/Nm³ (10,400 ppm).

The filter housing was opened and the filter element removed. Both inlet sides of the filter element are shown in Figure 3.3-5. As before, most of the outside surface of the filter had about a 1.6 to 3.2 mm (0.06 to 0.13 in) layer of dust, with approximately the same amount on the filtering surface inside the channels. The variation of cleaning from the top to the bottom of the filter was even more pronounced in this case, with the bottom half of the element cleaned to some extent, while the upper part was virtually covered over with dust. As before, the dust layer was friable and easily removed.

The observation of partial cleaning of the filter surface is consistent with the observed pressure drop behavior; that is, a very slow rate of pressure drop rise, starting with the clean filter, which occurs as the dust cake builds uniformly. When the filter was pulsed for cleaning, only a relatively small fraction of the dust was dislodged. This lowered the base pressure drop to the previous low value, but since virtually all of the subsequent flow was concentrated in the cleaned spot, it "healed" or plugged very rapidly, leading to the observed overall pressure drop behavior. This suggested the possibility that more rigorous cleaning cycles may have been necessary to obtain efficient usage of the filter surface.

Figure 3.3-5 Cross-Flow Element after Second HTHP Test 4.23.81

A considerable amount of soot was mixed in with the ash on the filter surface, indicating that the combustion was not as efficient as in previous runs. We believe this was because the dust injector line was placed about even with the combustor flameholder, and since a large fraction of the total airflow came through the dust injector and did not mix in the actual flame region, it left a rich, "smoky" flame.

The clean side of the filter was again free of significant dust, and the outlet sample indicated an average outlet loading of about 0.004 g/Nm^3 (2.8 ppm), most of which appeared to be soot. We estimated on the basis of these measurements that the overall efficiency was greater than 99.99 +%.

When the exhaust bypass line was opened for inspection, we found it to be plugged solid with large chunks of agglomerated dust. Figure 3.3-6 indicates the plugged region. We had anticipated that dust would blow out of the system during blowback and periods when the bypass line was opened, but this was not the case. An additional problem occurred when we discovered that the pressure relief line for the filter housing was also plugged, rendering it ineffective. For both operational and safety reasons the system was shut down until this section could be modified.

By mid-June all the modifications to the filter bypass system had been installed and pressure checked, thereby enabling filter testing to resume. Concurrent with the bypass modifications, attempts had been made to reduce the combustor inlet cross-sectional area so that both dust injector air and combustion air could mix prior to entering the flameholder. This was an attempt to improve the propane combustion and reduce the amount of soot formed. The insert used did not work well because it tended to blow the flame out at higher mass flows. Test 6.15.81 was the first conducted with the modified bypass system. It was carried out at a temperature of 811 K (1000°F), pressure at 1114 kPa (150 psig), and a mass flow of 57 kg air/hr (125 lbm air/hr). This yielded an average filter velocity of 1.74 m/min (5.7 ft/min). The

Figure 3.3-6 Filter Test System Schematic

duration of the actual filtering was 150 minutes, during which time 542 g of PFBC ash was fed to the filter. The average inlet concentration of the dust was a relatively high 5.6 g/Nm^3 (4358 ppm). The expected loadings at the filter in an actual PFBC plant might be expected to be lower by a factor of 20 to 50 if this filter system followed an efficient cyclone train. Testing at these low concentrations, however, would take a prohibitive amount of time. Figure 3.3-7 presents the filter pressure drop vs time history of the test. As experienced in previous tests, the cycle time was relatively long for the initially clean filter, taking about 65 minutes to rise from 1 kPa (4 in. wg) to 3 kPa (12 in. wg). At this time, one pulse from a 2.1 dm^3 (0.075 ft^3) blowback reservoir pressurized to 1692 kPa (235 psig) was administered for 1 second. As indicated in Figure 3.3-7 the pressure drop recovery was essentially complete. The initial rate of pressure drop increase on all subsequent cycles was very much greater than the rate experienced in the initially clean filter, indicative of incomplete cleaning. The cycle time to reach 3 kPa (12 in. wg) consequently became shorter over the next several cycles. In addition, the freshly blown-back filter pressure drop began slowly to increase. At the end of the third cycle, the pulsing sequence was modified to include four pulses in sequence, although the second through fourth pulses were only marginally effective. At the end of the test day it appeared that a reasonably stable cycle had been established, as indicated by the last three cycles of Figure 3.3-7. We decided to inspect the filter element and then replace it in the system without any external cleaning and thereby pick up testing where it had been stopped. The element was removed for inspection, and its condition is shown in Figure 3.3-8. The photograph indicates that the most effective area of cleaning was at the bottom of the filter nearest to the clean gas plenum, and that the area cleaned least was at the top of the filter, farthest from the clean gas plenum. These observations are consistent with the current concept that the blowback jet does not expand and raise the pressure at the top of the filter. It was later found to be possible for the blowback jet to actually create a low pressure at the

Figure 3.3-7 Pressure Drop vs Time for Ceramic Cross-Flow Filter Test 6.15.81
 Test Conditions: $T = 811 \text{ K (1000}^\circ\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $C = 5.6 \text{ g/Nm}^3 \text{ (4358 ppm)}$
 $V_{\text{filter}} = 5.7 \text{ ft/min (1.74 m/min)}$

Figure 3.3-8 Ceramic Filter Element after Seven Cycles and 16.7 atm Pulse Cleaning at 811 K (1000°F) and 1114 kPa (150 psig) Test 6.15.81.

filter top by inducing flow in this region as an eductor does. The outlet sample was weighed, and we determined that the average outlet concentration was 0.0002 g/Nm^3 (0.17 ppm). The filter efficiency was calculated at 99.99+%.

The filter was reinstalled without external cleaning, and a second "continuation" test was initiated. The test conditions of Test 6.18.81 were identical with the previous test, except that the dust feed was later found to be 10% lower, giving an average inlet concentration of 6.1 g/Nm^3 (3940 ppm). Cumulative dust feed time was 2.07 hours, during which a total of 463 g of dust was fed to the filter. The pressure drop history of this test is presented in Figure 3.3-9. The pressure drop at the onset of the test was much higher than it was at the end of the previous test, 3.4 vs 1.75 kPa (12.5 vs 7 in. wg). This may have been due to absorption of moisture in the dust cake during start-up. Three pulses of 1692 kPa (235 psig) were sufficient to reduce the filter pressure drop to about 2 kPa (8 in. wg). Over the period of the next hour [seven cycles to 3 kPa (12 in. wg)] the cycle time gradually shortened from about 3.75 to 2.5 minutes, and the base-line pressure drop rose from 2.1 to 2.5 kPa (8.5 to 10 in. wg). The cleaning sequence during this time consisted of four pulses in series from the 1692 kPa (235 psig) pulse tank. The pressure drop was then allowed to rise to 3.75 kPa (15 in. wg) over the next 17 cycles. The initial cycle time increased to about 13 minutes for the rise from 2.5 to 3.75 kPa (10 to 15 in. wg), but gradually decreased to about 2.5 minutes to go from 3.1 to 3.75 kPa (12.5 to 15 in. wg) during the seventeenth cycle. During the final two cycles of the test day, the pressure drop was allowed to climb to 5 kPa (20 in. wg). As with previous tests, the cycle time appeared to become shorter, and the base-line pressure drop increased to 3.4 kPa (13.5 in. wg), although time did not allow us to observe more than two cycles. By the end of this test, the pressure drop observed on bypass had increased from about 0.5 to 1 kPa (2 to 4 in. wg), indicating the possibility of a slow plugging of the bypass line.

Figure 3.3-9 Pressure Drop vs Time (continuation test) Test 6.18.81
 Test Conditions: $T = 811 \text{ K}$ (1000°F) (538°C)
 $C = 5.1 \text{ g/Nm}^3$ (3940 ppm)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_{\text{filter}} = 5.7 \text{ ft/min}$ (1.74 m/min)

The filter element 1WH-1 was removed from the test unit for inspection and is shown in Figure 3.3-10. The same general features existed as before, cleanest at the bottom, near the plenum, and dirtiest at the top, away from the plenum. The filter was overall considerably more loaded with dust. The outlet sample obtained from this test indicated an overall filter efficiency of 99.91%.

The filter element was vacuumed clean and reinstalled in the filter housing. Minor maintenance was carried out on the system, which was prepared for the next filter test. The main objective of test 6.26.81 was to investigate methods for improving cleaning performance. As part of this effort, the blowback reservoir was attached to a nitrogen tank so that pulse pressures could be increased over the 1692 kPa (235 psig) limit imposed by the compressor used in previous tests. The system was again brought to 811 K (1000°F), 1114 kPa (150 psig), and a mass flow of air at 57 kg/hr (125 lbm/hr). The average dust concentration was later calculated as 8.1 g/Nm³ (6250 ppm). As Figure 3.3-11 indicates, the initial pressure drop rise was relatively slow and constant, rising from about 0.6 to 3 kPa (2.5 to 12 in. wg) over the first 48 minutes. At this point a series of four pulses of 1854 kPa (260 psig) were administered from the reservoir, which had been pressurized using the nitrogen tank. Each 1-second pulse has been estimated to deliver 0.014 Nm³ (0.51 scf) of nitrogen. Pressure drop recovery is fairly stable to about 1.25 kPa (5 in. wg) over the first several cycles, as shown in Figure 3.3-11. As in previous runs, however, the cycle times shortened, and the base-line pressure eventually began to creep up, reaching 2.5 kPa (10 in. wg) by the twelfth cycle. During this cycle the dust feed rate slowed, and we discovered that the dust feeder was unexpectedly empty. At this point the dust feeder was valved out of the system, and the pulse reservoir was brought to 2127 kPa (300 psig). A series of four, 1-second pulses returned the base-line pressure drop from 2.5 to 1.8 kPa (10 to 7.25 in. wg). A repeat of this sequence brought the pressure drop down a little more to 1.68 kPa (6.75 in. wg). At this point the passage pressure was reduced

Figure 3.3-10 Ceramic Filter Element after Seven Hours of Testing at 811 K (1000°F), 1114 kPa (150 psig), Cleaning Pulse of 1692 kPa (235 psig). Test 6.18.81

Figure 3.3-11 Pressure Drop vs Time for Ceramic Filter 1WH-1
 Test Conditions: $T = 811 \text{ K}$ (1000°F)
 $P = 1114 \text{ kPa}$ (150 psig)
 $C = 8.1 \text{ g/Nm}^3$ (6250 ppm)
 $V_{\text{filter}} = 5.7 \text{ ft/min}$ (1.7 m/min)
 Test 6.26.81

from 1114 to 750 kPa (150 to 100 psig), and the mass flow was reduced in proportion to maintain a constant differential pressure. A third series of 2127 kPa (300 psig) pulses was then administered, and the filter pressure drop was reduced to about 1 kPa (4 in. wg). Subsequently, both passage flow and pressure were returned to their previous set points. The pressure drop remained constant at 1 kPa (4 in. wg). The system was then shut down and prepared for inspection. The outlet sample from this test indicated an average outlet concentration of dust plus soot of 0.008 g/Nm^3 (0.6 ppm), or 99.99%.

The filter element was removed and is shown in Figure 3.3-12. The higher cleaning pulse pressure and the lower filter ambient resulted in substantially better cleaning of the filter element. The bottom half of the filter was cleaned very well, and only the uppermost rows, furthest away from the clean gas plenum, were plugged. These results were most encouraging, and we arranged to increase the blowback pulse pressure and volume.

Inspection of the bypass line revealed that a gradual buildup of dust had occurred and it was necessary to clean out the line. Arrangements were made to obtain a differential pressure transducer to measure the pressure differential between the dirty side of the filter and the inside of the filter plenum during blowback. With this additional instrumentation we anticipated that different pulsing techniques could be quantified and that these data would be useful for the modeling effort of the cleaning process.

During July, several runs were carried out using a differential pressure transducer to characterize the pressure rise in the filter clean-side plenum during blowback. For the initial test, with the pressure transducer in place, the passage was pressurized and a flow of 57 kg/hr (125 lbm/hr) was kept, but the temperature of the system was maintained at room ambient.

Pulses were administered by pressurizing a 2.12 dm^3 (0.075 ft^3) pulse tank with a nitrogen bottle and subsequently opening a 0.95 cm

Figure 3.3-12 Ceramic Filter after Test at 811 K (1000°F), 1114 kPa (150 psig), and Pulse Cleaning at 2127 kPa (300 psig) and then Same Pulse at Filter Ambient of 780 kPa (100 psig), Test 6.26.81

(3/8 in) solenoid valve in a 0.95 cm (3/8 in) tube that served as the blowback nozzle. Pulse pressures from 1449 to 2127 kPa (200 to 300 psig) were used, exhausting into the filter element ambient, which was maintained at pressures from 780 to 1114 kPa (100 to 150 psig). The consequent pressure differential rise between the clean-side plenum and the dirty side was sensed by placing one pressure tap at a position near the bottom of the filter element and another in the dirty-side of the filter element housing. These taps were brought through the pressure vessel and connected to a differential pressure transducer and strip chart recorder. Figure 3.3-13 presents some typical results of this first test. The 1449 kPa (200 psig) pulse into the 1114 kPa (150 psig) filter did not result in a very large pressure differential across the filter, peaking at about 3.4 kPa (0.5 psi). Additionally, with the blowback geometry being used, the pressure rise was only significant for about the first 0.2 s. The remaining 0.8 s was essentially useless. The pulse pressure was boosted to the maximum value available from the nitrogen regulator, 2127 kPa (300 psig), and the system was then pulsed. With the filter at 1114 kPa (150 psig) the peak pulse height increased to about 13.1 kPa, (1.9 psi), and the time of positive pressure differential increased slightly. A final series of pulses were administered at 2127 kPa (300 psig) into the filter after its pressure had been lowered to 780 kPa (100 psig). As Figure 3.3-13 indicates, this gave a peak pressure rise of 19.3 kPa (2.8 psi), but essentially all the last half of the 1.0 s pulse was of no use for cleaning.

Test number 7.2.81 was carried out at elevated pressure and temperature with dust injection. The nominal test conditions were as follows:

T = 811 K (1000°F)
P = 1114 kPa (150 psig)
Airflow = 57 kg/hr (125 lbm/hr)
V filter = 1.7 m/min (5.7 ft/min).

Curve 729226-A

Test Conditions Test 7. 1. 81
P = 150 psig (11 atm)
T = 70°F (21°C)

Figure 3.3-13 Pressure Rise at Filter Bottom during Blowback

During this run a total of 981 g as dust were fed to the filter, and an average inlet concentration of 5.86 g/Nm^3 (4537 ppm) was maintained. Isokinetic total mass filter outlet samples were taken over the duration of the test and indicated an average outlet concentration of 0.0025 g/Nm^3 (2.0 ppm), yielding an overall efficiency of 99.94%. The pressure drop-time behavior, shown in Figure 3.3-14, in general, was similar to past runs, with the cleaning cycle gradually shortening over the test period, except that the shortening of the cycle time was much slower this time, probably because of the higher cleaning pulse pressure used. Throughout this test the pulse pressure used was 2127 kPa (300 psi), which was the most available from the nitrogen bottle pressure regulator on hand. As Figure 3.3-14 shows, the initial filter cycle began at a pressure drop of about 1.25 kPa (5 in. wg) and rose to 3 kPa (12 in. wg) over the first 40 minutes. After 18 cycles (270 min), the initial pressure loss across had risen to 2.4 kPa (9.5 in. wg), and the time to reach 3 kPa (12 in. wg) had diminished to about 5 minutes. The cleaning sequence between filter cycles consisted of four consecutive pulses from a 2.12 dm^3 (0.075 ft^3) tank charged to 2127 kPa (300 psig). For the final cleaning episode of this test, the passage pressure was reduced to 780 kPa (100 psig) and the mass flow was proportionately reduced so as to maintain constant filter velocity. Under the reduced pressure condition the filter cleaning was significantly more effective as the initial pressure loss was reduced from 2.4 to 1.75 kPa (9.5 to about 7 in. wg), which was only slightly above the starting pressure. The last filter cycle shown in Figure 3.3-14 indicates that this opened the cycle time from 5 to 15 minutes.

During this test the differential pressure transient across the filter was monitored during blowback using a differential pressure transducer. As before, the high-side probe [0.64 cm (1/4 in)] tube was located about 2.54 cm (1 in) from the bottom of the filter element in the clean-side plenum, and the low-pressure tap was inserted in the dirty side of the filter element housing. Figure 3.3-15 presents the traces obtained during the two different blowback conditions.

Figure 3.3-14 Pressure Drop vs Time for Test 7.2.81 with Filter Element 1WH-1

As indicated, the 2127 kPa (300 psig) pulse into 1114 kPa (150 psig) ambient caused about a 15.9 kPa (2.3 psi) peak differential, which decayed to essentially nothing within 0.5 second. With the filter ambient reduced to 780 kPa (100 psig), this pulse resulted in a 17.9 kPa (2.6 psi) peak differential, which decayed somewhat more slowly. It was encouraging to note that the relatively small difference between the two pulse traces shown in Figure 3.3-15 resulted in the substantially better cleaning cycle shown at the end of Figure 3.3-14.

The filter housing was opened and the filter element removed for inspection. Figure 3.3-16 presents a photograph of both sides of the element still mounted in its holder. The bottom half of the element appears to have been cleaned quite well, while the upper region of the filter is essentially plugged. The overall appearance of the filter was substantially better than that obtained using 1854 kPa (260 psig) pulses and led us to expect that cleaning methods that gave more spatially uniform and higher differential pressures during blowback would lead to more adequate cleaning.

The failure of the filter differential pressure to fall to low values when the bypass line was opened during tests was evidence that the bypass line was plugged to some degree. The bypass line was removed and inspected and found to have accumulated a good deal of dust (Figure 3.3-17), so that was cleaned and reassembled.

In preparation for the next test, the blowback system was equipped with a larger, 1.9 cm (0.75 in) solenoid valve and a high-pressure regulator for the nitrogen bottle. Additionally, two more pulse tanks were fabricated and pressure tested. These tanks are four and nine times the volume of the tank used to date.

Test number 7.15.81 was carried out at the usual conditions of 811 K (1000°F), 1114 kPa (150 psig), and a flow of 57 kg/min (125 lbm/hr). During this test 241 g of dust were fed, and a continuous outlet sample revealed an average outlet concentration of only 0.00013 g/Nm³ (0.1 ppm). Difficulty occurred during this test when the filter pressure drop trace

Figure 3.3-15 Pressure Rise at Filter Bottom during HTHP Test 7.2.81

Figure 3.3-16 Filter Element 1WH-1 after Test 7.2.81

Figure 3.3-17 Bypass Line Plug - Test 7.2.81

did not behave consistently. The pressure drop recorder wandered around, rising and falling whether dust was being fed or not. This test was cut short because of the spurious pressure drop readings, but at the end of the run the higher-pressure blowback system was tried. First, a series of four 1 s pulses at 2127 kPa (300 psig) were given, followed by four similar 0.5 s pulses, and finally by four 0.33 s, 2806 kPa (400 psig) pulses. The last series of pulses appeared to affect the pressure drop readings significantly, but we did not consider these readings reliable. During these blowback cycles the differential pressure transient at the top of the filter was monitored. Figure 3.3-18 presents the traces of the pressure rise measurements for the three blowback conditions. The measured peak pulses were, in general, about 13.8 kPa (2 psi) and somewhat broader than the pulses measured at the bottom of the filter. These results were placed in question, however, when the system was opened for inspection. We determined that the dirty-side pressure drop tap had

Curve 729227-A

Test Conditions for Test 7.15.81

P = 150 psig (11 atm)

T = 1000 °F (538°C)

Pulse 1 sec at 300 psig

Passage at 150 psig

Pulse .5 sec at 300 psig
Passage at 150 psig

Pulse .5 sec at 400 psig
Passage at 150 psig

Figure 3.3-18 Pressure Rise at Top of Filter during Blowback

become plugged with dust. Since this tap was used as the high-side tap for filtering pressure drop measurements and the low-side tap for blowback recording, we were uncertain how valid the results shown on Figure 3.3-18 were.

When the filter element was removed and inspected, it appeared to be very clean in the bottom half (Figure 3.3-19), with the usual plugging at the top. The comparison with previous tests is somewhat clouded because this test lasted a relatively short time, with only about 25% of the usual amount of dust being fed.

The last run of July (7.31.81) was conducted at temperature and system pressure but with no dust injection. A series of 1 s and then 0.5 s pulses at initial pressures of 1449, 1873, 2127, 2462, 2806, 3140, 3475 kPa (200, 250, 300, 350, 400, 450, and 500 psig) were administered. This series was then repeated at passage pressures of 608 and 780 kPa (75 and 100 psig). Prior to this test the single 0.64 cm (0.25 in) high-side tap had been replaced with two 0.32 cm (0.125 in) taps located 2.5 cm (1 in) from the filter top and bottom. Both were referenced to a low-side tap placed on the filter vessel housing inlet. Two differential pressure transducers were used with a two-pen strip chart recorder to examine the pressure rise at the filter top and bottom. Figure 3.3-20 presents three traces typical of the group. The pressure peak measured in the upper part of the filter was somewhat lower than in the bottom, and the pressure ramp was much less steep. At the higher pulse pressure, substantial pressure differentials could be maintained over most of the pulse duration. For pulse pressures of 3475 kPa (500 psig) into a 608 kPa (75 psig) ambient pressure, differentials were as high as 55.2 kPa (8 psi).

A bothersome issue at this time was the relatively small peak pressure differential between the filter top and bottom. It seemed unlikely that the small difference measured could explain the dramatic difference in cleaning observed. In subsequent testing we discovered that the pressure differential readings at the filter top were actually negative and were caused by the eductor effect of the blowback nozzle.

Figure 3.3-19 Filter Element after Test 7.15.81

Figure 3.3-20 Pressure Rise at the Bottom and Top of Filter, Common Low Side Referenced to Passage Inlet, Test 7.31.8. Conditions: $T = 811 \text{ K}$ (1000°F)

During August a great deal of effort was expended on improving the blowback cleaning technique, and in the course of ten actual days of high-temperature and -pressure testing, we made good progress.

During the previous month we obtained the blowback differential pressure rise data of two transducers connected to the clean side plenum of the filter. One transducer was located near the top of the filter and one near the bottom. Both were referenced to a common low side located back at the passage inlet. There was some concern about the location of the low side tap, since it was some distance from the actual filter and the tap did not penetrate the hot gas liner system. There was, thus, the potential for capacitive effects on the transient (~ 0.5 s) pressure pulse that was being measured. To resolve this question the experiments were repeated with the low-side tap placed in the dirty side of the filter chamber, right next to the filter element. A series of blowback cycles were carried out at 811 K (1000°F) and passage pressures of 608, 780, and 1114 kPa (75, 100, and 150 psig). Blowback pulse of 0.55 s was provided from a 2.1 dm^3 (0.075 ft^3) tank pressurized from 1449 to 3475 kPa (200 to 500 psig). The transient pressure pulse at the filter top and bottom were recorded, and the peak pressures are presented in Figures 3.3-21 and 22. The bottom peak pressures were similar to those measured previously, rising linearly to about 34 kPa (5 psi) with the 3475 kPa (500 psi) pulse. The top differential pressures measured, however, were totally different.

In this experiment we discovered that the upper portion of the clean side of the filter element actually experiences a negative pressure (relative to the dirty side) during blowback. This negative pressure is on the same order as the positive rise at the bottom of the filter, shown in Figure 3.3-21, and was apparently caused by the eductor effect of the 0.95 cm (0.4 in) blowback nozzle, which was placed about 2.54 cm (1 in) into the clean side of the filter plenum. In retrospect, this was most likely the cause of the nonuniform filter cleaning that had been observed to date. Figure 3.3-23 presents one of the pressure transients we recorded during these blowback tests.

Figure 3.5-21 Peak Pressure Rise at Top of Filter - Referenced to Dirty Side of Filter - 8.5.81
Test conditions: $T = 811 \text{ K (1000}^\circ\text{F)}$

Figure 3.3-22 Peak Pressure Rise at Bottom of Filter - Referenced to Dirty Side of Filter -
8.8.81 Test conditions $T = 811 \text{ K}$ (1000°F)

Blowback Nozzle Positioned inside Filter Plenum

Blowback Pulse Pressure 3545 kPa (500 psig; 35 atm)

Figure 3.3-23 Differential Pressure Rise at Top and Bottom of Cross-Flow Filter Referenced to Dirty Side of Filter, Test 8.5.81. Conditions: $T = 811\text{ K}$ (1000°F), $P = 1114\text{ kPa}$ (150 psig)

At the end of the blowback pulse tests, we began dust injection and continued it for the short period remaining in the test day. We began again and continued throughout the following day. Test parameters were $T = 811 \text{ K}$ (1000°F), $P = 1114 \text{ kPa}$ (150 psig), filter velocity 1.7 m/min (5.7 ft/min), average inlet concentration 3.07 g/Nm^3 (2377 ppm). The measured average overall filter efficiency was 99.35% over the 400 minutes of testing accumulated. The pressure drop vs time history of the test is presented in Figure 3.3-24. The base-line pressure drop was maintained at a fairly low, constant level of about 0.75 to 1 kPa (3 or 4 in. wg), but the cycle time to reach 3 kPa (12 in. wg) gradually diminished from about an hour to a half-hour during the 7.5 cycles of the test.

At the end of test 8.5.81, the system was disassembled and the filter element removed for observation. Figure 3.3-25 is a photograph that confirms the suction measured at the filter top during cleaning pulses. The filter was very clean at the bottom where the pressure rise was measured as about 34.5 kPa (5 psi), but the dust cake was untouched at the top where the suction pressures were observed. We hypothesized that the gradual shortening of the filter cycle time was due to slow continual blinding of the upper portion of the filter element.

Although the overall average efficiency was high (99.35%), there was some evidence of a dust leak. Inspection of the fibrous seal around the filter element revealed what was suspected to be a leak near the element top, as shown in Figure 3.3-26. Inspection of the clean side channels, however, gave no indication of an actual leak in the element. Given the long series of rigorous blowback pulses prior to the start of dust injection, we did not feel the apparent blowout was unreasonable. Inspection of the filter chamber (dirty side) gave some insight into the effectiveness of blowback during this test. As shown in Figure 3.3-27 during a cleaning pulse, the dust was expelled from the filter with enough velocity and force to strike against the chamber wall and stick there.

Figure 3.3-24 Pressure Drop vs Time for Ceramic Cross-Flow Filter. Test 8.5.81 Conditions:
 $T = 811 \text{ K (1000}^\circ\text{F)}$, $P = 1114 \text{ kPa (150 psig)}$

Figure 3.3-25 Posttest Condition Showing Cleaning Nonuniformity, test 8.6.81

Figure 3.3-26 Filter Seal Leak Observed in Test 8.6.81.

Figure 3.3-27 Filter Chamber Dust Deposit after Test 8.6.81 Showing Vigorous Action of Blowback Cleaning

Normal maintenance was performed on this system, and the filter element was resealed within a new fiber frax gasket. The system was prepared for another run. Test 8.12.81 extended over two test days and was carried out at the same operating conditions as previous tests. During this test the measured inlet concentration was 4.1 g/Nm^3 (3233 ppm) and the overall average efficiency was only 98.35%. The pressure drop vs time record of the test is presented in Figure 3.3-28. Rather low, base-line pressure drop values were observed [about 0.75 kPa (3 in. wg)]. With a 0.5 s pulse at 3475 kPa (500 psig), however, the lengthening cycle times indicated the probability of a substantial leak. This was visually confirmed by inspection of the isokinetic outlet sample. The test was terminated at this point and the filter removed for inspection. Figure 3.3-29 presents a photograph of the filter exterior, and, as in previous runs, the bottom was cleaner than the top. The light spot above the test number tag turned out to be a result of another hole in the sealing gasket. Figure 3.3-30 is a close-up view showing the roughly 2 cm (0.75 in) portion of the failed gasket. There was also indication (Figure 3.3-29) that the seal had failed at the filter top in the same location as in the previous test.

In preparation for the next test, a new fiber frax gasket was cut and placed in the filter system. This time an alundum cement was spread over the exterior of the gasket material in an attempt to strengthen the seal. In addition to performing the usual posttest maintenance, we altered the blowback nozzle position in an attempt to gain more uniform pressure distribution on blowback. Using the photograph of Figure 3.3-25 as a visual guide for estimating the jet expansion, we raised the pulse nozzle about 6.35 cm (2.5 in) to a position roughly 3.8 cm (1.5 in) above the clean-side plenum, as shown in Figure 3.3-31.

Tests 8.18.81 through 8.21.81 were carried out over a period of four consecutive days during which nearly 1000 minutes (16.7 hr) of filter time was accumulated. The test conditions were, as before, 881 K (1000°F), 1114 kPa (150 psig), and a filter velocity of 1.7 m/min 5.7 ft/min).

Filter 3.3-28 Pressure Drop vs Time for Ceramic Cross-Flow Filter Test 8.12.81 - 8.13.81

Conditions: $T = 811 \text{ K} (1000^\circ\text{F})$
 $P = 1114 \text{ kPa} (150 \text{ psig})$
 $V_{\text{filter}} = 1.7 \text{ m/min} (5.7 \text{ f/min})$
 $C = 4.2 \text{ g/Nm}^3 (3233 \text{ ppm})$
 $\eta = 98.35\%$

Figure 3.3-29 Posttest Filter Conditions 8.12.81

Figure 3.3-30 Gasket Failures during Test 8.12.81

Figure 3.3-31 New Blowback Nozzle Placement to Improve Cleaning Uniformity

The measured average inlet concentration was 3.07 g/Nm^3 (2375 ppm), and a total of nearly 2 kg of dust were fed to the system. The measured overall efficiency was found to be 99.98% over the first two-day period and 99.29% over the second. Figure 3.3-32 presents the pressure drop history of the test sequence. An apparently stable filter cycle was established after the first cycle or two, with the pressure drop rising from a base of about 1.35 kPa (5 in. wg) to 3 kPa (12 in. wg) over about a 40-to-45 minute time span. The only anomalous cycle was the next to last (from 815 to 915 min), which was somewhat longer than the others. This was the first cycle of the last test day, and the relatively flat portions of the cycle are probably attributable to variations in the dust feed rate. This stable performance was maintained using three, consecutive, 0.5 s pulses from the 3475 kPa (500 psig) tank. A typical blowback pressure rise transient is shown in Figure 3.3-33. The new placement of the blowback nozzle resulted in a much more uniform pressure rise distribution, with the bottom peak reaching about 38 kPa (5.5 psi) and the top at a positive 31 kPa (4.5 psi). This filter element was subsequently removed for inspection and is shown in Figure 3.3-34. Visual examination of the filter confirmed the more uniform pulse distribution, as the dust accumulation appeared to be virtually the same across the filter surface, with the top channels being no more plugged than the bottom. The overall cleanliness of the filter appeared to be somewhat poorer than generally observed at the filter bottom in previous tests. This was probably due to a small reduction of the peak pulse pressure [from about 55 to 38 kPa (8 to 5.5 psi)] that resulted when the blowback nozzle was moved out of the clean-side duct and placed above the plenum outlet. The accumulation on the outside of the filter surface, however, has little effect on the filter pressure drop, since the actual filtering surface is inside the filter body. It is for this reason that the base-line pressure drops measured were as low as shown in Figure 3.3-32.

Figure 3.3-32 Pressure Drop vs Time for Ceramic Cross-Flow Filter

Test 8.18.81

Conditions: Temperature = 811 K (1000°F)

Pressure = 1114 kPa (150 psig)

$V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min)

$C = 3.1 \text{ g/Nm}^3$ (2375 ppm)

$\eta = 99.98\%$

Curve 729701-A

Blowback Nozzle Positioned above Filter Plenum

Blowback Pulse Pressure 3545 kPa (500 psig, 35 atm)

Figure 3.3-33 Differential Pressure Rise at Top and Bottom of Cross-Flow Filter Referenced to Dirty Side of Filter Test 8.18.81. Conditions: $T = 811\text{ K}$ (1000°F), $P = 1114\text{ kPa}$ (150 psig)

Figure 3.3-34 Posttest Condition of Filter Element after Test 8.18.81
Showing More Uniform Cleaning

Prior to the next test sequence the bypass and exhaust lines were opened and flushed clean with water because they had accumulated a considerable amount of dust. Additionally, a dust shield was placed over the dirty side differential pressure tap since that 0.63 cm (0.25 in) tube has a tendency to plug after numerous pressure surges from the blowback system.

The filter element was put back into the system without any external cleaning or adjustment to the seals. Test 8.27.81 was carried out at the same temperature, pressure, and flow conditions as the previous tests. The average dust feed concentration was calculated to be 3.0 g/Nm^3 over the roughly 5-hour test period. Although the overall average efficiency was calculated to be 99.46%, it was evident that the filter was leaking. For this reason the test was stopped after the nine steady cycles shown in Figure 3.3-35, which presents the pressure drop vs time history for the run. The apparently steady-state nature of the cycles was maintained by 3475 kPa (500 psig) pulses for blowback. A typical trace of the pressure rise at the filter top and bottom is presented in Figure 3.3-36. When the system cooled, the filter element was removed for inspection and appeared as shown in Figure 3.3-37. As in the last test, the dust removal appeared to be uniform across the filter surface as a result of the more spatially uniform pulse of Figure 3.3-36. Close examination of the filter revealed the cause of the observed leak. A small section of this fiber frax gasket had been blown out during blowback, as shown in Figure 3.3-38. The subsequent penetration of dust explains the lower efficiency observed.

In an attempt to remedy the sealing problem, the filter element holder was machined out and enlarged a bit to seat the ceramic element more easily. A new gasket was cut, the element was seated and twisted down, and then a bead of ceramic cement was placed around the lip of the seal in an attempt to prevent the localized breakthrough observed to date. At this point, we anticipated that these efforts directed at the sealing problem would prove successful.

Figure 3.3-35 Pressure Drop vs Time for Ceramic-Flow Filter, Test 8.27.81

Conditions: $T = 811 \text{ K} (1000^{\circ}\text{F})$
 $C = 3.0 \text{ g/Nm}^3 (2306 \text{ ppm})$
 $P = 1114 \text{ kPa} (150 \text{ psig})$
 $V_{\text{filter}} = 1.7 \text{ m/min} (5.7 \text{ ft/min})$
 $\eta = 99.46\%$

Curve 729702-A

Figure 3.3-36 Pressure Rise in Clean Plenum during Blowback Pulse
 Test 8.27.81. Test Conditions: $T = 811 \text{ K} (1000^{\circ}\text{F})$,
 $P = 1114 \text{ kPa} (150 \text{ psig})$, $P_{\text{Blowback}} = 3475 \text{ kPa} (500 \text{ psig})$

Figure 3.3-37 Posttest (8.27.81) Condition of Filter Showing Uniform Cleaning

Figure 3.3-38 Seal Failure of Test 8.27.81

Work during September centered on determining the causes of the dust penetration observed in the filter system, as well as on accumulating a considerable amount of high-pressure and -temperature run-time experience.

The initial test of the month (9.1.81) was carried out over two consecutive days, during which over 6 hours of dust feeding time were realized. The test conditions were similar to those of all recent work:

$$T = 811 \text{ K (1000}^{\circ}\text{F)}$$

$$P = 1114 \text{ kPa (150 psig)}$$

$$\text{Airflow} = 57.5 \text{ kg/hr (125 lbm/hr)}$$

$$\text{Filter velocity} = 1.7 \text{ m/min (5.7 ft/min)}$$

$$\begin{aligned} \text{Avg. inlet dust loading} &= 8.07 \text{ g/Nm}^3 \text{ (6246 ppm)(9/1)} \\ &= 12 \text{ g/Nm}^3 \text{ (9264 ppm) (9/2)}. \end{aligned}$$

Figure 3.3-39 presents the pressure drop/time curve for the test run. The cleaning pulses used were as in previous runs, 0.5 s at 3475 kPa (500 psig). The curve shows that during the first day a fairly steady base-line pressure drop of 1.25 to 1.5 kPa (5 to 6 in. wg) was maintained over the three, 50-to-60-minute-long cycles that were run. During this run the observed overall efficiency was very high, measured at 99.91%. The bypass line was evidently becoming plugged, because when the system was switched to bypass prior to the filter being cleaned, the pressure drop would not fall from the 3 kPa (12 in. wg) to the usual 0.25 to 0.5 kPa (1 to 2 in. wg) observed when the bypass is newly cleaned. The cleaning pulses appeared still to be effective, in spite of not actually being "off-line" cleaning, so the test was continued. The plugging problem was probably in part due to the high dust loading generated by the dust injector. Posttest weighing of the dust indicated that the dust loading was 8.07 g/Nm^3 (6246 ppm) and that 963 g of dust had been fed over 163 minutes. This rate of nearly 6 g/min is about twice the usual desired test rate. The test was terminated when the dust injector ran dry about 2 hours sooner than anticipated.

Figure 3.3-39 - Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 9.1.81

Conditions: $T = 811 \text{ K } (1000^\circ\text{F})$

$P = 1114 \text{ kPa } (150 \text{ psig})$

$V_{\text{filter}} = 1.7 \text{ m/min } (5.7 \text{ ft/min})$

The dust feeder was reloaded, and the test sequence was continued the following day. The dust accumulated in the bypass line apparently became more impervious overnight because attempts to get the bypass pressure drop down were futile. Even with the filter element essentially on line all the time, the blowback pulses were fairly effective, initially returning the system pressure drop from 3 to 2 kPa (12 to 8 in. wg). As the test proceeded, however, it appeared that the base-line pressure drop had increased a bit to 2.25 kPa (9 in. wg). During this run the dust feed became even higher than on the previous day and was measured at 8.8 g/min, leading to a loading to the filter of 12 g/Nm³ (9264 ppm). The cause of the unexpectedly high feed rates was later determined, and provisions for better control were implemented. During this test several samples were taken, and it became apparent from the very first sample that the system was leaking dust. The overall average efficiency for the run had dropped to 98.62%. The initial sample was taken over the first 51 minutes of the test and included the first blowback sequence at about $t = 205$ in Figure 3.3-39. A total of 72 mg was measured, a clear indication that something was wrong. A second sample was taken only through the second blowback sequence, during which time and dust feed was off. This led to the collection of 5.3 mg and is an indication that the penetration was high during or just subsequent to a blowback episode. A third sample was taken for 26 minutes during the third cycle ($t = 255-281$, Figure 3.3-39) and included no blowback sequence. This sample had 36.5 mg on it, indicating a rate of 1.4 mg/min. The final sample spanned the last 80 minutes of the test and included two blowback cycles. This sample had about 154 mg accumulated for an average rate of 1.93 mg/min, indicating a worsening condition. At the final cycle of the day, no blowback was performed, in the hope that dust tracks on the clean-side plenum would aid in locating the leak. Figure 3.3-40 is a photograph of the filter when it presents 3 kPa (12 in. wg) resistance at 1.7 m/min (5.7 ft/min). A shot of the newly vacuumed element is provided for comparison. Examination of the clean side unfortunately revealed no clue as to where the leak

As Removed from Test Rig with No Blowback

Cleaned Element

Figure 3.3-40 Dust Deposit on Filter That Had 3 kPa Pressure Drop at 811 K, 1115 kPa, and Filter Velocity of 1.7 m/min.

was occurring. There were several spots around the element seal that appeared to be potential leaks, as shown in the close-ups of Figure 3.3-41. The filter was removed from the mount, cleaned, and replaced in the filter holder after the seating face of the filter had been squared up using a diamond cut-off wheel. This procedure was carried out because we noticed that the seating face had up to a 0.32 cm (0.125 in) wave in it, causing uneven clamping pressure on the gasket. This time the gasket between the mount and the filter was made from two layers of a dense, 0.3 cm (0.125 in) thick alumina fiber paper or mat rather than the previously used blanket material. As before, a ceramic cement was pasted around the outer lip, as shown in Figure 3.3-40, and the unit was replaced in the system for subsequent testing.

Test 9.8.81 was carried out at the same base operating conditions as the previous tests, except that the dust loading was reduced somewhat to about 7.75 g/Nm^3 (6000 ppm). Figure 3.3-42 presents the pressure drop data for this test and the next one, 9.10.81. As the figure indicates, a nice, stable-appearing cycle was maintained with the cycle from 1.25 to 3 kPa (5 to 12 in. wg) taking about 50 to 60 minutes. Cleaning pulses were as before, 0.5 s at 3475 kPa (500 psig). Only one sample was taken during this test, and it included all four blowback episodes shown on the chart. Inspection of the outlet filter after the test indicated a large leak, since 380 mg of dust were present. This is an average rate of 1.4 mg/min to the sampler and corresponds to an average overall efficiency of only 98.84%.

When the system was disassembled the following day and the filter element removed for inspection, we could see no conclusive signs of leaks. The filter element and holder were placed in a plastic bag, and air loaded with carbon black was blown through the filter assembly. A moist piece of white filter paper was held against the two rows of clean-side exit channels in an attempt to locate any channel that was leaking. After the entire dirty side of the filter was caked with carbon black, the filter paper was removed and we could see no trace of carbon black. Next, the moist filter paper was taped over the

Figure 3.3-41 Potential Leak Areas in Filter Seal after Testing

Figure 3.3-42 Pressure Drop vs Time for Ceramic Cross-Flow, Test 9.9.81

Conditions: $T = 811 \text{ K } (1000^\circ\text{F})$
 $P = 1114 \text{ kPa } (150 \text{ psig})$
 $V_{\text{filter}} = 1.7 \text{ m/min } (5.7 \text{ ft/min})$

rectangular exit of the clean-side plenum and the experiment was repeated, but there was still no trace of penetration.

At this point we concluded that if the leak was associated with the filter element or sealing system, it had to depend on temperature. Such a leak could occur because of the differential thermal expansion between the stainless clamping rods and the ceramic element itself. At these temperatures a "loosening" of about 0.08 cm (0.03 in) was estimated, but we thought that the soft gasket at both ends of the filter could accommodate this slack.

The other alternative was a "systems"-associated leak from the dirty side to the clean side. Two possible mechanisms were: a simple leak from the dirty side through the seal between the filter element support plate (Figure 3.3-43) or an eduction of air and dust from the bypass and exit during blowback. During blowback the eductor effect created by the blowback nozzle causes a momentary low pressure on the clean side of the filter, and it is possible to envision dusty air being pumped around the loop and drawn down the clean air exit section, as indicated.

In an attempt to prove this possibility, several grams of magnetite powder were placed in the clean gas outlet elbow prior to the next test.

The element was cleaned and replaced in the system for the short test 9.10.81 shown in Figure 3.3-42. The system was brought to the usual operating flow, temperature, and pressure, but with no dust injection. An initial sample was taken for 10 minutes with the system in bypass to get a background reading. This sample weighed less than 0.05 mg, which is the lower limit on the balance for these conditions. Next, the system was given a series of four, 0.5 s blowback pulses at 3475 kPa (500 psig), and the sampler was left on during this 10-minute sequence. A very small amount of black powder (0.13 mg) was collected, but the small amount precluded an easy magnetic confirmation that the dust was the magnetite placed in the outlet elbow. Posttest observation

Figure 3.3-43 Possible Sources of "System" - Associated Dust Leaks

indicated that certainly the bulk of the magnetite powder remained in the outlet elbow and that the blowback surge is not capable of dislodging large dust deposits and pumping them into the clean side. It is, reasonable, however, to expect that some of the dilute phase suspended dust could arrive at the clean side by this mechanism. At that point the system had to be shut down for a short time because the propane tank was empty. Two 10-minute samples taken upon restart confirmed the very low background dust levels, so dust feed was initiated and one 1-hour cycle was carried out. During the first 15 minutes of the cycle, the dust injector vibrator was left off in an attempt to keep the dust feed rate down. As shown in Figure 3.3-42, unfortunately very little, if any, dust was fed without the vibrator on, so it was started at about $t = 15$. The sample taken spanned the entire cycle but did not include the blowback cycle at the end. This sample collected 35.2 mg over 57 minutes for an average sample rate of 0.62 mg/min. This gave a reasonably high efficiency of 99.52% but nonetheless indicated a degree of particle penetration that should not have been there.

The small amount (if any) of magnetite on the outlet filter samples suggested that the clean/dirty seal should be checked for a leak. The filter vessel top flange was removed, and fine powder was placed around the clean/dirty sealing circumference and flow through the filter was begun. At a pressure drop of 3.75 kPa (15 in. wg) a large air leak was observed in the region above the inlet to the filter liner shown in Figure 3.3-43. The size of the leak indicated that it was potentially a dust leak as well as an air leak at higher filter differential pressures. The filter chamber Inconel liner was removed, and we saw that the inlet liner had pulled away from the filter chamber liner (because of thermal cycling), leaving about a 1.25 cm (0.5 in) gap. The inlet dusty air had eroded a sizable amount of the insulation above the inlet, as shown in Figure 3.3-44. This low-resistance path could have led to the observed penetration. The inlet liner was fitted with a slip joint and mechanically fixed to the chamber liner to prevent

Figure 3.3-44 Erosion of Filter Chamber Insulation and Potential Dust Sealing Problem

subsequent liner gaps. The insulation was repacked, and a new gasket was installed in the filter plate/vessel seal.

Test 9.17.81 was short because after about 1.5 hours of testing one of the dust feed lines eroded through and caused an emergency shut-down. The pressure drop curve from the shortened run is shown in Figure 3.3-45. Several visual checks of the sample filter paper indicated some degree of penetration during the test. The final filter weight was 22.9 mg, collected over 99 minutes of sampling, or 0.23 mg/min. This figure corresponded to a respectable 99.85% overall efficiency for the filter, but it was not as good as the 99.9+% efficiencies achieved in earlier testing.

When the filter was examined closely after the test, a small crack was observed in one end cross barrier, as shown in Figure 3.3-46. These are the only ribs that, if separated from the filter surface above, will result in a dust leak. This is because the clean-air plenum is on the other side, rather than just another cross channel. This problem was easily solved by cementing this first row of channels shut, as shown in Figure 3.3-46. At that point the separation at the clamped end of the filter in the same plane was unnoticed: this defect is clearly shown in Figure 3.3-46.

Test 9.23.81 was carried out at the usual test conditions of 811 K (1000°F), 1114 kPa (150 psig) and $V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min). When we started the passage, we noticed that the bypass line was plugged, evidently irreversibly, since cleaning pulses were ineffective in lowering the bypass pressure drop from 1.9 kPa (7.75 in. wg). We decided to carry through with the test anyway, since we had had reasonable luck in cleaning with a plugged bypass in the past. The pressure drop/time curve for this run is presented in Figure 3.3-47. Our inability to get the filter off line by opening the bypass is reflected in the relatively short cycle times experienced, although these were in part due to the high dust loading of 12.6 g/Nm^3 (9729 ppm). In spite of the test system, the filter was cleaned in a stable fashion and cycled

Figure 3.3-45 - Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 9.17.81

Conditions: $T = 811 \text{ K}$ (1000°F)
 $C = 13.1 \text{ g/Nm}^3$ (10.094 ppm)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min)
 $\eta = 99.85\%$

Separation of End Rib from Filter Surface and Resultant Leak

Filter with First Channels Cemented - Note Undetected Separation at Clamped End

Figure 3.3-46 Filter Element after 9.17.81 Test and Prior to Next Test 9.23.81

Figure 3.3-47 - Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 9.233.81

Conditions: $T = 811 \text{ K } (1000^{\circ}\text{F})$
 $C = 12.6 \text{ g/Nm}^3 \text{ (9729 ppm)}$
 $P = 1114 \text{ kPa (150 psig) (11 atm)}$
 $V_{\text{filter}} = 1.7 \text{ m/min (5.7 ft/min)}$
 $\eta = 99.27\%$

between 2.25 to 2.5 (9 or 10 in. wg) and 3 kPa (12 in. wg) over the 14 cycles of the test. The posttest appearance of the filter was somewhat dirtier than usual but quite good, considering that it had been cleaned with essentially no bypass (Figure 3.3-48).

The average overall efficiency of the filter was measured at 99.27%, but the rate of dust penetration increased throughout the test. A total of six samples were taken over the course of the test, and the sample filter rates observed were in time sequence 0.08, 0.07, 0.14, 0.32, 1.17, and 1.75 mg/min. This deterioration in efficiency from roughly 99.91 to 98.10% was clear evidence of a worsening problem. Subsequent examination of the filter revealed that the element had delaminated along the plane shown in Figure 3.3-46 and was held together only at the newly cemented and repaired end. This element was removed and was sent to Coors for examination. A new element (2WH-1) was cut from the same block. This filter was mounted in the system and prepared for testing.

Test 9.29.81 was carried out over two successive days at the usual test conditions, as indicated in Figure 3.3-49. This figure shows that the initial pressure drop across the filter was extremely low at 0.25 to 0.375 kPa (1 to 1.5 in. wg). With an average inlet dust load of 9 g/Nm^3 (6972 ppm), the initial cycle took about 100 minutes to reach 3 kPa (12 in wg). Subsequent cycles were somewhat shorter, as expected, but were in the 60-to-80-minute range. Blowback during the first cycle consisted of a single 2806 kPa (400 psig) pulse, which dropped the pressure differential from 3 to 0.3 kPa (12 to 1.25 in wg). Since the previously employed 3475 kPa (500 psig) pulse was not required, we did not use it throughout the remainder of the test. Over the first day of testing the average overall efficiency was measured at 99.97%. Sample filter weight gain rates did not change significantly over the duration of the first test and averaged about 0.04 mg/min. Testing was picked up the following day, as indicated in Figure 3.3-49, with very similar results. The average dust feed concentration was slightly lower the second day at 7.5 g/Nm^3 (5801 ppm), as was the overall efficiency, which

Figure 3.3-48 Cross-Flow Filter after Test in Which On-Line Cleaning Was Employed

Figure 3.3-49 - Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 9.26.81

Conditions: $T = 811 \text{ K}$ (1000°F)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min)
 Element 2WH-1

was measured at 99.94%. This variation in efficiency, however, was probably within experimental error. The base-line pressure drop appeared steady at about 0.5 kPa (2 in. wg) and was maintained by 2806 kPa (400 psig) pulses. The success of this test was encouraging, since both seals and filter survived thermal cycling and pulse cleaning.

At the beginning of October we decided that some time should be devoted to up-grading the dust injector so that more predictable delivery rates could be maintained. We had noticed that the actual dust delivery rates were usually about twice the rates expected on the basis of rotational speed, feed plate hole spacing, and dimension. This discrepancy was attributed to dust "sneaking" between the fixed and rotating feed plates. A simple fix was tried that consisted of gluing a thin pad of felt to the stationary plate on which the rotating plate turned. This effectively blocked or sealed all delivery holes in the moving plate except the one aligned with the fixed plate through-hole. Figure 3.3-50 is the calibration curve obtained subsequent to this modification and indicates the effectiveness of the remedy. In the higher feed rates (above 2 g/min) the rate correlates very well with the calculated rate if one allows for somewhat less than 100% filling of the delivery holes. Typical dust feed rates are on the order of 2 to 4 g/min for these filter tests, so the system feed rate was then both predictable and controllable. We then resumed testing.

The filter element under examination was the second piece of the 0.15 cm (60 mil) membrane, 100 μ m pore size material (2WH-1) that had accumulated about 600 minutes of operating time during the 9.29.81 and 9.30.81 tests.

Test 10.7.81 was carried out at the same conditions as the previous two tests. The modified dust injector provided a uniform feed of 3.8 g/min, corresponding to an inlet concentration of 5.23 g/Nm³ (4026 ppm). Blowback during this test consisted of a 0.5 s pulse from a 2.122 (0.075 ft³) tank pressurized to 2856 kPa (400 psig). Figures 3.3-51 and -52 present the results of this test, which was conducted over

Figure 3.3-50 Dust Feed System Calibration Curve

two consecutive days. The cycles ran from about 0.75 to 3 kPa (3 to 12 in. wg) over a 1.5 to 2 hour period. Samples taken over the first day of testing showed dust penetration rates of 0.034, 0.08, and 0.08 mg/min, which indicated no problem with the filter and yielded an overall efficiency of 99.89%.

The system was restarted the following day for a continuation test. The system was brought to the same T, P, and flow conditions as on the previous day. The overall average inlet concentration for this test was determined to be 5.85 g/Nm^3 (4529 ppm). To determine the outlet size distribution a 5.5 hour outlet sample was taken using a cascade impactor. During this time a total of 51.03 mg of dust was collected, which corresponded to an average overall efficiency of 99.70%. The size distribution presented in Figure 3.3-53. Both the relatively low efficiency and the outlet size distribution (which is very similar to the inlet) indicated a small leak was in the system. The pressure drop curve presented in Figure 3.3-52 was very similar to

Figure 3.3-51 Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 10-7-81

Conditions: $T = 811 \text{ K } (1000^{\circ} \text{ F})$
 $P = 1114 \text{ kPa } (150 \text{ psig})$
 $V_{\text{filter}} = 1.7 \text{ m/min } (5.7 \text{ ft/min})$
 $C = 5.23 \text{ g/Nm}^3 \text{ (4028 ppm)}$
 $\eta = 99.89\%$

Figure 3.3-52 Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 10-7-81, Continued

Conditions: $T = 811 \text{ K (1000}^{\circ}\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $V_{\text{filter}} = 1.7 \text{ m/min (5.17 ft/min)}$
 $C = 5.88 \text{ g/Nm}^3 \text{ (4029 ppm)}$
 $\eta = 99.70\%$

Figure 3.3-53 - Size Distribution of Outlet Dust as Determined by Impactor

that of the previous day and indicated that the leak was not large at this point. After this test the system was shut down for a day of maintenance and preparation for the next run.

We decided the next test would be carried out at full temperature. Accordingly, test 10.12.81 was carried out at 1088 K (1500°F) and 1114 kPa (150 psig). In order that filter velocities be kept comparable, the mass flow was reduced to 42.4 kg/hr (93 lbm air/hr). As the system was brought to temperature, a great deal of care was taken to examine various parts of the facility for hot spots. None were found after a few hours of stable operation, so dust feeding was initiated. With a feed dust concentration of 6.6 g/Nm³ (5128 ppm) the pressure drop rose slowly from about 0.75 kPa (3 in. wg) to 3 kPa (12 in. wg) over a period of about 3.5 hours, as shown in Figure 3.3-54. The overall efficiency of the filter during this period was 99.1%, but the three roughly hour-long samples taken during this test showed dust penetration rates of 0.35, 0.77, and 1.35 mg/min at the sampler

Figure 3.3-54 - Pressure Drop vs Time for Ceramic Cross-Flow Filter, test 10-21-81.

Conditions: $T = 1088$ (1500°F)
 $P = 1114$ kPa (150 psig)
 $V_{\text{filter}} = 1.7$ m/min (5.7 ft/min)
 $C = 6.66$ g/Nm³ (5128 ppm)
 $\eta = 99.01\%$

When this test day was over, the system was shut down and disassembled to inspect the filter.

When the filter chamber was opened we had some difficulty in removing the filter element holder because the inlet liner had grown by thermal expansion and pushed against the back of the element holder. When the filter was removed, two types of failure were observed, as shown in Figure 3.3-55. The first was a small hole for dust penetration. The second was delamination of the filter top plate, as well as fracture of the next two lower plates at the corner. The delamination was attributed to differential thermal expansion on heating or cooling and, primarily, on poor bonding of the rib/filter flat surface. A possible explanation for the fracture is the differential expansion between the filter element and the stainless steel hold-down clamps. On heating, the hold-down rods grow nearly twice as much as the filter element (about 2 mm), loosening the element. Because the rods are only pinned, they sag under their own weight (Figure 3.3-56). When cooled, the rods contract and tend to force the filter element upward in the holder, causing a load to be exerted on the element where it pushes against the top edge of the holder recess and resulting in the cracking shown in the close-up of Figure 3.3-56.

At this point only one piece of the original three pieces of 1.5 mm (60 mil) filter medium (WH-1) was left, and we were reluctant to put it in the system until we were fairly certain we knew we could avoid further cracking problems. We decided to test what we thought was the 1.0 mm (40 mil), 100 μ m medium (1WH-3).

This material looked denser and had a smoother surface. It was also slightly smaller than the 15 cm (6 in) cubes used previously, which necessitated making an Inconel shim insert for the filter holder. The bonding between ribs and filter flats looked poorer than the previous two elements tested. The element was mounted in the filter holder, and this time titanium hold-down rods were used because their coefficient of thermal expansion is much closer to that of the ceramic filter. We,

Small Crack Failure

Delamination of base plate and fracture in filter element corner

Figure 3.3-55 Filter Element Failure (2WH-1)

Hold down clamp loosened and in sagging position

Close up of cracked layers of filter surface

Figure 3.3-56 Filter Element Failure (2WH-1)

thus, tried to avoid the loosening problem observed during previous thermal cycling.

Test 10.15.81 was carried out at a temperature of 1088 K (1500°F) at 1114 kPa (150 psig), with an average inlet concentration of 7.6 g/Nm^3 (5917 ppm). The mass flow for this test was set at the previous rate of 42 kg/hr (93 lbm air/hr). Since the element was slightly smaller than the previous two elements tested, the actual filter velocity was a little higher, 1.9 m/min (6.3 ft/min), rather than the usual 1.7 m/min (5.7 ft/min). Even if allowance was made for the slightly higher velocity, the pressure drop of this 1.0 mm (40 mil) filter was higher than the previously tested 1.5 mm (60 mil) material. Conversation with Coors revealed that a slightly different mixture of materials was used in formulating the 1.0 mm (40 mil) material, possibly explaining the observed flow resistance differences. Clean filter pressure drops for the new material were about 2 kPa (8 in. wg) at 1.9 m/min, which would be about 1.8 kPa (7.25 in. wg) at 1.7 m/min, while the previous materials' clean pressure drops were typically about 0.5 kPa (2 in. wg) for these conditions.

The results of this two-day, high-temperature and -pressure test are shown in Figure 3.3-57. The pressure drop rose from a base line of about 2 kPa (8 in. wg) to the blowback set point of 3.7 kPa (15 in. wg) over about a 1.5 hour cycle time. Cleaning pulses consisted of 0.5 s pulses of 2135 kPa (400 psig) air, which gave rise to about a 48 kPa (7 psi) pressure rise in the clean-side plenum as measured by differential pressure transducers. Samples of the outlet taken during this test day indicated sample filter rates of 0.11 and 0.64 mg/min, which yielded an overall efficiency of 99.70% for the day and indicated the probability of a small leak.

Testing was continued the next day at similar conditions. This time the inlet concentration was determined to be 7.7 g/Nm^3 (5944 ppm), and the overall efficiency decreased to 99.48%. A series of 1-hour

Figure 3.3-57 Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 10-15-81. Using 40 ml (1 mm), 100 μm element

Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $V_{\text{filter}} = 1.9 \text{ m/min (6.3 ft/min)}$
 $C = 7.68 \text{ g/Nm}^3 \text{ (5917 ppm)}$
 $\eta = 99.70\%$

outlet samples indicated sample filter rates of 0.33, 0.41, 0.47, and 0.73 mg/min, so we decided to inspect the filter.

The following working day the filter housing was opened and we tried to remove the filter holder and filter. We had a great deal of difficulty in removing the assembly because the inlet liner had expanded and pushed the filter chamber liner into the back of the filter element holder, denting it and locking it in place, as shown in Figure 3.3-58. The slip fit joints in the liner system had become stuck and were unable to accommodate the thermal expansion experienced. When the filter element was finally retrieved, we noted that the top hold-down clamp was very loose, with a gap of about 0.32 cm (0.125 in) between the bar and retaining nuts, as shown in Figure 3.3-59.

When the clamp was removed, we noted that the filter had delaminated at both sides of one layer, also as shown in Figure 3.3-59. The very loose condition has since been traced to the unusually thick fiber frax pad used between the clamp and the ceramic. This material retains its resiliency at temperatures up to 811 K (1000°F) but seems to "heat set" at 1088 K (1500°F). With the observed very loose clamping on the filter and rather rough handling required to extricate the filter holder from the chamber because of the inlet liner push, we could not say with much certainty how the delamination occurred.

At that point it was obviously time to change the method of accommodating thermal movement in the liner system. We inserted two lined bellows sections in the inlet liner (one on each side of the inlet sample port) and then fastened the liner at the combustor, the sample port, and at the filter chamber liner.

When the modification of the test passage liner system had been completed, the second piece of 100 μ m, 1.0 mm (40 mil) material (2WH-3) was placed in the filter holder assembly. Prior to this, we had "patched" and reinforced the filter with a high-temperature, phosphate-based, ceramic cement (Aremco Ceramabond 503). A new gasketing material for seating the filter in the holder was also used this time. It is

Figure 3.3-58 Damage to Back of Filter Mounting Caused by Inlet Liner Thermal Expansion

Hold-Down Clamp Very Loose after Test

Delamination Failure

Figure 3.3-59 Posttest Condition of 40 ml (1.5 mm) 100 μ m Ceramic Filter (1WH-3)

Figure 3.3-60 Pressure Drop vs Time for Ceramic Cross-Flow Filter, Test 10-30-81

Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $V_{\text{filter}} = 1.9 \text{ m/min (6.3 ft/min)}$
 Element 2WH-3

made by 3M under the name of Interam, and it expands somewhat on heating. We expected it to help in sealing.

Test 10.0.88 was carried out at virtually the same conditions as the previous two tests, and the results of this run are presented in Figure 3.3-60. The pressure drop performance of the filter is similar to that of previous tests on this material, with the pressure drop rising from 1.75 to 2.5 kPa (7 to 10 in. wg) over roughly 1 to 5 hour periods. During this time we measured very high efficiencies, with an overall test average of 99.97%. Weight gain rates recorded on the outlet samples were 0.024, 0.077, and 0.020 mg/min and indicated no deterioration of performance with time. As before, successful blowbacks consisted of 0.5 s pulses from a 2735 kPa (400 psig) reservoir. After this 4 hour run the system was shut down for inspection of the filter and clamping assembly and for modifications of the liner. The liner modifications were found to be successful, with no evidence of bending or other deformations.

The filter, as shown in Figure 3.3-61, was in excellent condition and still firmly clamped in place. Since the entire system appeared to be in good shape, the filter was placed (uncleaned) back into the test apparatus for further testing.

Test 11.3-5.81 was carried out on element 2WH-3 at 1088 K (1500°F), 1114 kPa (150 psig), and a filter velocity of 1.7 m/min (5.7 ft/min). The inlet dust concentration averaged 8.7 g/Nm³ (6754 ppm) over the three-day test period. The measured overall efficiencies were 99.95, 99.95, and 99.86%. During the final period of the run, the outlet sample rate of dust collection increased from a previously steady rate of 0.02 to 0.05 g/min to about 0.16 mg/min, indicating the start of significant leaking through the cross-flow filter. Figure 3.3-62 presents the pressure drop/time history of this test. Pulses of 2806 and 3475 kPa (400 and 500 psig) were used for cleaning the filter, and reasonably stable cycles were maintained between about 2.75 and 3kPa (9 and 12 in. wg). When a marked increase

Figure 3.3-61 Posttest Condition of 100 μ m Pore Size, 1.0 mm (40 mil) Thick Ceramic Filter (2WH-3)

Figure 3.3-62 - Cross-Flow Ceramic Filter Test 1.0 mm (40 mil) Back Web Thickness, 10 μm Pore Size, Test 11.3 - 5.81

Conditions: $T = 1088 \text{ K (1500}^\circ\text{F)}$
 $V_{\text{filter}} = 1.7 \text{ m/min (5.7 ft/min)}$
 $P = 1114 \text{ kPa (150 psig)}$
 Element 2WH-3

in outlet concentration was observed the test was terminated and the filter removed for observation. Once the element had been removed from the system, we could see that the leak had occurred near the bottom plate of the filter (seen in the bottom right-hand corner of Figure 3.3-63). When the element was unclamped and removed, a complete delamination was observed. This delamination was similar to the one observed at the end of test 10.15.81, except that this element had been thermally cycled three times and used for a total of about 17 hours, as opposed to the previous failure, which had occurred with poor clamping in the first day of testing.

Both elements had been cut from the same block of material, which Coors reported to be of 100 μm pore size and 1.0 mm (40 mil) back web. The high-pressure drop across the material, relative to the previously tested 100 μm 1.5 mm (60 mil) material, could not be explained until a check of the Coors records revealed that the material actually had a 10 μm pore size instead of the reported 100 μm .

The smaller pore size explained both the higher filtering pressure drop and the higher pulse pressure required for cleaning. We felt that this higher pulse pressure cleaning and the relatively poor bonding between filter ribs and flats were the underlying causes of the failures by delamination in the WH-3 test elements.

To prepare for the next test, an element was cut from the 10 μm , 1.5 mm (60 mil) sample block that had been supplied by Coors in their first shipment (1WH-2). As with the previously tested 10 μm , 1.0 mm (40 mil) element, cracks between laminations and the sealed end were reinforced with ceramic cement. After the cement had cured, the element was mounted in the filter holder and placed in the test rig.

Test 11.11.81 was carried at 1088 K (1500°F) and 1114 kPa (150 psig) and at a filter velocity of 1.7 m/min (5.7 ft/min) and inlet concentration of 7.6 g/Nm³ (5921 ppm). The measured overall efficiency was 99.62%, and the outlet sample dust accumulation rate increased from 0.12 mg/min to a final value of 0.7 mg/min, indicating a significant

As Removed from Test Rig

Delamination Occurred near End Plate after Removal from Filter Holder

Figure 3.3-o3 Failure of 10 μ m Pore Size, 1 mm (40 mil) Membrane Filter after Test Days (11.3 - 5.81) and 10.30.81 at 1088 K (1500°F), 1114 kPa (150 psig)

Figure 3.3-64 Cross-Flow Filter Test 11.11.81 - 10 μ m Pore Size
1.5 mm thick (60 mil).

Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$
 $V_{\text{filter}} = 5.7 \text{ ft/min}$
 $\gamma = 99.62\%$
 $P = 1114 \text{ kPa (150 psig)}$
 $C = 7.6 \text{ g/m}^3 \text{ (5921 ppm)}$

leak. The pressure drop/time chart for this run is shown in Figure 3.3-64. The first cycle is indicative of the completely clean filter pressure drop of about 2.5 kPa (10 in. wg). This base line increased to about 3 or 3.25 kPa (12 or 13 in. wg) in subsequent cycles when cleaning pulses of 0.5 s from the 3475 kPa (500 psig) pulse tank were used. When the filter was removed from the test rig for examination, the start of a delamination was clearly visible, as shown in Figure 3.3-65. Since this was the only flaw observed, we tried to repair the element by patching it with ceramic cement. When the repair was finished, the element was replaced in the test rig for another trial. In test 11.16.81 the system was brought to the same conditions as the previous test. As usual, just prior to starting dust feed, we pulse-cleaned the filter with a single 3475 kPa (500 psig) pulse. The filter pressure drop fell from 5.5 kPa (22 in. wg) to 0.3 kPa (1.2 in. wg). The assumption that the filter had broken was confirmed when 7 minutes of dust feed produced no increase in pressure drop and the outlet sample filter was overloaded. Figure 3.3-66 indicates that not only had the repaired joint failed, but a clean-side/dirty-side rib had broken, leaving a large hole in the filter.

The next test was carried out using the remaining new piece of the 10 μ m, 1.5 mm (60 mil) 2WH-2. As before, ceramic cement was used to seal and reinforce the new element.

Test 11.17-18.81 was carried out at the same conditions as the several previous tests except that a slightly higher mass flow resulted in a slightly higher filter velocity than before, 2 m/min (6.5 ft/min) instead of the usual 1.7 m/min (5.7 ft/min). The pressure drop time chart for this run is shown in Figure 3.3-67 and shows the base-line pressure drop increase from about 5.5 kPa (22 in. wg) to a fairly steady value of about 8.7 kPa (35 in. wg) after 4 or 5 cycles with a blowback set point at 10 kPa (40 in. wg). Cleaning during the first day was accomplished using 0.5 s pulses at 2512 kPa (350 psig). The overall efficiency measured was quite high, 99.91% in spite of the high frequency of blowback.

Filter as Removed from Test Rig

Close-up of Crack Where Bottom Plate Has Started to Delaminate

Figure 3.3-65 Cross-Flow Ceramic Filter after Test 11.11.81 -
10 μ m Pore Size 1.5 mm (60 mil) Thick, 1WH-2

Damaged Clean/Dirty Side Rib

Delamination of Repaired Bottom Plate

Figure 3.3-66 Cross-Flow Filter after Test 11.16.81 - 10 μ m Pore Size
1.5 mm (60 mil) Thick, 1WH-2

Figure 3.3-67 Cross-Flow Ceramic Filter 10 μm Pore Size, 1 mm thick (60 mil), Test 11.17.81

$T = 1088 \text{ K} (1500^{\circ}\text{F})$

$V_{\text{filter}} = 2 \text{ m/min} (6.5 \text{ ft/min})$

$P = 1114 \text{ kPa} (150 \text{ psig})$

2WH-2

Testing was resumed the next day, but several things appeared somewhat different. The base pressure drop had fallen to about 7.0 kPa (28 in. wg), the cycle time lengthened to about 20 to 25 minutes, and the outlet sample dust rates increased from previous values of about 0.06 mg/min to 0.20 initially, and finally to 1.7 mg/min. This all indicated that a significant leak was present at the start of the second day of testing and was rapidly worsening. This was reflected in the day-long average overall efficiency of 98.63%. When the filter element was removed, two delaminations were visible, as shown in Figure 3.3-68. As with previous failures, the delamination appeared to be a result of poor bonding that failed under thermal cycling and blowback stresses.

By this time we had received new elements from Coors that had been fabricated to our dimensions. The WH-4 elements as received are shown in Figure 3.3-69. They offered somewhat more filter area, with the rib spacing planned for the test holder (compare to Figure 3.3-68). They also had a wide contact area at the outside of the long channels, which was intended to improve the unit's strength. When we examined the elements, we became concerned about the possibility of delamination failures like the ones observed before because of what appeared to be poor bonding and already existing cracks at the rib/flat joints. We also learned later that these elements were made from 10 μ m pore size plates instead of the 100 μ m pore size called for in the order.

Nevertheless, the element LWH-4 was prepared for testing and placed in the test rig. Test 11.23-24.81 was carried out at conditions similar to the other tests of this period except that the filter velocity was slightly lower than usual, 1.6 m/min (5.2 ft/min). The pressure drop/time record of this test is presented in Figure 3.3-70. The pressure drop behavior of this element was very similar to the element that had just been tested when a correction for filter velocity was made. As before, the base pressure drop increased as the filter was "conditioned". During the first day of testing the outlet sample dust arrival rates were measured at 0.04, 0.09, 0.15, 0.20, and 0.21 mg/min. These converted to an all-day average efficiency of 99.77% and indicated

As Removed from Test Rig

Delamination Observed on Removal from Filter Holder

Figure 3.3-68 Posttest Condition (11.17-18.81) of Failed Cross-Flow Filter 10 μ m Pore Size, 1.5 mm (60 mil) Back Web, 2WH-2

Figure 3.3-69 - New Filter Medium Fabricated by Coors, Approximately 0.12 m^2 (1.31 ft^2) Filter Area, $10 \text{ }\mu\text{m}$ Pore Size, 1.5 mm (60 mil) Thick Back Web, 1WH-4

Figure 3.3-70 Cross-Flow Filter Test (11.23-24.81), 10 μm Pore Size, 1.5 mm (60 mil) Back Web

$T = 1088 \text{ K (1500}^\circ\text{F)}$

$V_{\text{filter}} = 1.6 \text{ m/min (5.2 ft/min)}$

$P = 1114 \text{ kPa (150 psig)}$

IWH-4

indicated the likelihood of a small leak. As in the test prior to this one, the second day was quite different, with the base pressure drop decreasing to about 5 kPa (20 in. wg), the cycle times lengthening, and the outlet sample rates going to 0.28, 0.33, 0.40, and 1.61 mg/min. Again, it was clear that the new element had failed, and removal of the element confirmed this. As shown in Figure 3.3-71, the filter had delaminated in two places. This confirmed the initial assessment of the potential for delamination.

At this time, Coors realized the mistake in the pore size of the elements provided and began preparing filter elements that actually were 100 μm in pore size in an attempt to duplicate the original 100 μm filter element supplied for use. In the interim, a long-term test was performed on the last piece of the original 100 μm pore size, 1.5 mm (60 mil) back web material, i.e., 3WH-1.

All the testing done in December was carried out using the last remaining third of the original ceramic block tested, 3WH-1. This element was nominally 1.5 mm (60 mil) back web with a 100 μm pore size. The element's back (blind) side was sealed with Aremco 503 cement, as were all accessible joint surfaces. The element was then placed in the test rig for a long-duration test.

Test 11.30-12.4.81 was a long-duration test with the new cross-flow filter element. Shifts were scheduled to work around the clock for one full work week, and during this time we accumulated a little more than 80 hours of actual on-line filter time. Because the bypass line for the filter system gradually accumulates dust, regular short periods of shutdown were required to clean the lines. These occurred about two or three times per 24 hour period and lasted on the average of 2 to 2.5 hours. In all, the filter was thermally cycled from 1088 K (1500°F) to about 477 K (400°F) a total of 14 times for system maintenance during the test period.

Test conditions for the entire long-duration test were as follows:

As Removed from Test Rig

Taken from Filter holder

Figure 3.3-71 Failure by Delamination of Cross-Flow Filter (10 μ m Pore Size) 1.5 mm (60 mil) Back Web, 1WH-4

Temperature = 1088 K (1500°F)
Pressure = 1114 kPa (150 psig)
 $V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min)
 $C_{\text{dust}} = 5.0 \text{ g/Nm}^3$ (3860 ppm)
 $\Delta P_{\text{trigger}} = 3 \text{ kPa}$ (12 in. wg).

The entire test pressure drop/time history is presented in Figure 3.3-72. The system maintenance down time has been excluded from the plot. It is interesting to note the very long initial cycle time (4.5 hours) followed by several subsequent cycles that shortened as the filter "conditioned". In the course of about 5 or 6 cycles (10 or 11 hours) the filter was completely conditioned and reached a steady operating state that cycled between a pressure drop of 1.25 to 3 kPa (5 to 12 in. wg) over about a 45 to 60 minute period. This steady behavior continued for the duration of the test, as we had hoped.

During most of the test period the cleaning procedure consisted of from one to three 0.5 s pulses of N_2 from an 2.12 l (0.075 ft³) accumulator pressurized to 2600 kPa (350 psig). The blowback line and nozzle was a simple 0.95 cm (3/8 in) tube with an equivalent length of about 1.8 m (6 ft). The average dust feed concentration was calculated over several periods of time and found to be reasonably constant at about 4.98 g/Nm³ (3860 ppm). The total amount of dust fed was measured at 12.6 kg (28 lbm). The dust penetration throughout the test period was very low, probably below our ability to accurately determine it. The average efficiency over the first 21 hours of sampling was calculated as 99.95%, over the next 45.5 hours 99.94%, and over the final 11 hours 99.94% again.

Completion of this test gave us renewed confidence that the more porous filters of this type can be successfully cleaned over extended periods of time without physically damaging the element and that very high removal efficiencies can be realized at modest pressure drops.

Having successfully completed a long-duration test, we decided to explore filter performance at higher face velocities. The first

Figure 3.3-72 Pressure drop vs Time for Ceramic Cross-Flow Filter, Test 11.30-12.4.81

Conditions: $T = 1088 \text{ K} (1500^{\circ}\text{F})$
 $P = 1114 \text{ kPa} (150 \text{ psig})$
 $V_{\text{filter}} = 1.7 \text{ m/min} (5.7 \text{ ft/min})$
 $C = 5.0 \text{ g/Nm}^3 (3860 \text{ ppm})$
 $\eta = 99.94\%$

Figure 3.3-72 (cont'd)

Figure 3.3-72 (cont'd)

Figure 3.3-72 (cont'd)

Figure 3.3-72 (cont'd)

Figure 3.3-72 (cont'd)

Figure 3.3-72 (cont'd)

attempts to run at higher filter velocities had to be stopped because of minor problems with the combustor and the propane control valves, and a control valve on the compressor.

After these minor repairs were completed test 12-10.81 was carried out at 1088 K (1500°F), 1114 kPa (150 psig), with an average dust inlet concentration of 3.45 g/Nm^3 (2671 ppm). For this test the system airflow was set at 84.4 kg/hr (186 lbm/hr) to yield a filter velocity of 3.5 m/min (11.4 ft/min). Figure 3.3-73 presents the pressure drop/time history of this run. As the figure indicates, a nice, stable filtration cycle was immediately established, running from 2 or 2.25 kPa (8 or 9 in. wg) up to a trigger point of 4.5 kPa (18 in. wg) over a 30 to 40 minute period. A steady base-line pressure drop of about 2 kPa (8 in. wg) was consistent with the 1 kPa (4 in. wg) base line maintained in the long-term run, where the filter velocity was half this value. The measured value of the filter efficiency dropped slightly to 99.89%, but most of this apparent penetration was probably due to difficulty with the first of four outlet samples taken. The first sample was taken over the first 49 minutes of the test and weighed 8.5 mg for an average rate of 0.17 mg/min. The three subsequent, hour-long samples had an average rate of 0.03 mg/min.

The next test, 12.16.81, was run at similar conditions, but the dust injector rate was turned up in an attempt to get the dust loading up to about the 6.2 g/Nm^3 (4000 ppm) level, where the long-term run had been conducted.

The increased dust flow rate turned out about right, as the average dust loading was calculated at 5.09 g/Nm^3 (3944 ppm). Temperature, pressure, and filter velocity were as before. The results of this test are summarized in Figure 3.3-74. As in the previous test, steady, repeatable cycles were immediately established. These cycles ran from 2 kPa (8 in. wg) to the trigger point of 6 kPa (24 in. wg) over a period of about 30 minutes. The overall efficiency decreased to an average value of 99.77%. The first and third samples seemed out of line,

Figure 3.3-73 Test 12.10.81 Cross-Flow Filter Test with 100 μm Pore Size 1.5 mm (60 mil) Wall at High Filter Velocity:

$T = 1088 \text{ K } (1500^{\circ}\text{F})$

$C = 3.47 \text{ g/Nm}^3 \text{ (2671 ppm)}$

$P = 1114 \text{ kPa (150 psig)}$

$\eta = 99.89\%$

$V_{\text{filter}} = 3.5 \text{ m/min (11.4 ft/min)}, 3\text{WH-1}$

Figure 3.3-74 Test 12.16.81 Cross-Flow Filter Test with 100 μm Pore Size 1.5 mm (60 mil) Wall at High Filter Velocity and High Dust Concentrations:

$T = 1088 \text{ K } (1500^{\circ}\text{F})$

$C = 5.12 \text{ g/Nm}^3 \text{ (3945 ppm)}$

$P = 1114 \text{ kPa (150 psig)}$

$\eta = 99.77\%$

$V_{\text{filter}} = 3.5 \text{ m/min (11.4 ft/min), 3WH-1}$

corresponding to 0.25 and 0.38 mg/min, while the others were about 0.11 mg/min. Nevertheless, examination of the filters indicated that for the first time some visible amount of dust was penetrating the system. The final test of this period (12.18.81) was an attempt to repeat the previous test. Temperature, mass flow, pressure, and dust feed rate were all reset to their prior values, but for some reason the recorded dust feed rate was lower than anticipated by a factor of two. This led to the calculation of an inlet concentration level of 2.66 g/Nm^3 (2062 ppm). This level is inconsistent with the pressure drop traces presented in Figure 3.3-75, and we feel that most likely an error was made in measuring or recording the mass of dust fed to the system. The pressure drop traces were virtually the same as for the previous test, as would have been expected from the initial set-up of the test parameters. The efficiency based on the recorded low dust feed rate was 99.61%. If the previous dust feed rate were assumed, the efficiency would increase to 99.81%, which would again have been consistent with the prior test. As in the last test, a small but visible amount of dust was collected at the outlet sampler at a rate between 0.15 and 0.20 mg/min, which seemed to indicate that significant amounts of dust were penetrating the system. The source of the small leak could have been either the filter or the gasketing in the system. In any case, we decided that testing should continue without removing the filter for inspection.

Test 1.5-6-8.82 was the initial three-day test sequence of January, and the pressure drop history of the series is presented in Figure 3.3-76. During the first day of testing the system was brought to a flow of 84.4 kg/hr (186 lbm/hr) at 1114 kPa (150 psig) and a temperature of 1088 K (1500°F). This yielded a nominal filter face velocity of 3.5 m/min (11.4 ft/min). With a dust concentration of about 5.3 g/Nm^3 (4100 ppm) the pressure drop cycle began at 2.5 kPa (10 in. wg) and rose to the trigger pressure drop of 6 kPa (24 in. wg) over a period of about 20 minutes. The filter was cleaned using a 0.5 s, 2857 kPa (400 psig) pulse that returned the pressure drop consistently to 2.5 kPa (10 in. wg). After four cycles the system temperature began to drop, and we

Figure 3.3-75 Test 12.18.81 Cross-Flow Filter with 100 μm Pore Size 1.5 mm (60 mil) Wall
High Velocity Test:

$T = 1088 \text{ K } (1500^{\circ}\text{F})$

$C = 2.68 \text{ g/Nm}^3 \text{ (2062 ppm)}$

$P = 1114 \text{ kPa (150 psig)}$

$\eta = 99.61\%$

$V_{\text{filter}} = 3.5 \text{ m/min (11.4 ft/min), 3WH-1}$

discovered that the propane tank was nearly empty. The remainder of the test day was devoted to shutting the system down and changing the fuel system supply. The next test day the system was brought to the same conditions, and approximately 6 hours of test time were achieved, during which 12 stable cycles were carried out, again ranging from 2.5 to 6 kPa remainder of the test day was devoted to shutting the system down and changing the fuel system supply. The next test day the system was brought to the same conditions, and approximately 6 hours of test time were achieved, during which 12 stable cycles were carried out, again ranging from 2.5 to 6 kPa (10 to 24 in. wg) over roughly 20 minutes. At this point a mass balance on the system indicated that 1.6 kg of dust had been fed, yielding an average inlet concentration of 5.28 g/Nm^3 (4090 ppm) and an outlet concentration of 0.014 g/Nm^3 (10.63 ppm), for an overall average efficiency of 99.73%. While a penetration of this amount of dust is still well within the goals of the program, it was a fairly certain indication of a minor leak in the system. Since the penetration was virtually the same as noted in the final December run, we concluded that whatever leak existed was not getting worse.

Before removing the filter for visual examination, we decided to push on to yet higher velocities. Consequently, the third day of this test series was operated at the same temperature and pressure but at an increased mass flow of 105.3 kg/hr (232 lbm/hr), which yielded a filter face velocity of 4.3 m/min (14.2 ft/min). Attempts to go to higher mass flows during start-up were unsuccessful because of the size of the trim in the flow control valve. Higher flows could have been achieved, but we would have had to shut down the system and lose the test day. As the bottom chart of Figure 3.3-76 indicates, stable pressure cycles ranging from about 3.25 kPa (13 in. wg) to 6 kPa (24 in. wg) were achieved and maintained throughout most of the test day. During the final 1.5 hours there appeared to be a slight increase in base-line pressure drop to about 3.72 kPa (15 in. wg), but this was an artifact of the gradual plugging of the bypass line rather than a change caused by filter performance. As in previous tests, cleaning was accomplished using a

Figure 3.3-76 Test 1.5-6-8.21 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1.5 mm (60 mil) Thick, 100 μm Pore Size, 3WH-1

Conditions: $T = 1088 \text{ K} (1500^{\circ}\text{F})$
 $C = 3.47 \text{ g/Nm}^3 (2671 \text{ ppm})$
 $P = 1114 \text{ kPa} (150 \text{ psig})$
 $V_{\text{filter}} = 3.4 \text{ m/min} (11.4 \text{ ft/min})$
 $V_{\text{filter}} = 4.3 \text{ m/min} (14.2 \text{ ft/min})$

2857 kPa (400 psig) pulse of 0.5 s. The overall filter efficiency measured during this run was 99.81%, with an inlet loading of 2.93 g/Nm^3 (2269 ppm), which was consistent with the previous days of testing.

For the next test sequence (test 1.14.82) the trim in the flow control valve was changed to allow delivery of higher flows to the passage. The passage initially was brought to 1000 K (1500⁰R) and 1114 kPa (150 psig) and a mass flow of 105.3 kg/hr (232 lbm/hr), which was identical to the conditions of the previous test. The filter was then run through three cycles at this flow, 4.3 m/min (14.2 ft/min) to establish a base line or reference to the previous test day, as shown in Figure 3.3-77. Having established that the filter was performing as it had in the previous run, the mass flow was increased to 127 kg/hr (280 lbm/hr), yielding a filter velocity of 5.2 m/min (17.2 ft/min). At this flow the base-line pressure drop increased to an initial value of 4.5 kPa (18 in. wg) from 3.75 kPa (15 in. wg). In subsequent cycles the base-line pressure drop increased to a fairly steady value of about 5.25 kPa (21 in. wg). The cycle times had to be short because the differential pressure transmitter was spanned to 6.25 kPa (25 in. wg), which meant blowing the system down at 6 kPa (24 in. wg), leaving only a 0.75 or 1.0 kPa (3 or 4 in. wg) swing for filtration. Consequently, the cycle times were short (on the order of 5 min) as shown in Figure 3.3-77. During this portion of the run 33 cycles were carried out over a period of about 160 minutes. After these 33 cycles the mass flow was again decreased to yield a face velocity of 4.3 m/min (14.2 ft/min), and 13 additional cycles were carried out. The base-line pressure drop for these cycles was at about 3.75 kPa (15 in. wg) and remained essentially constant over the remainder of the run, indicating that the very high velocities and dust loadings of the previous interval had caused some plugging in the filter that could not be removed with the usual 0.5 s, 2857 kPa (400 psig) pulse. At the end of the test day a series of 10 blowback pulses were given to the filter at a slightly higher pressure 3201 kPa (450 psig). These pulses were successful in driving the base-line pressure down to its previous value of 3.75 kPa (15.3 in. wg), which was an indication

Figure 3.3-77 Test 1.14.82 Pressure Drop vs Time for Cross-Flow Ceramic Filter 1.5 mm (60 mil) thick, 100 μ m Pore Size

Conditions: $T = 1088$ K (1500 $^{\circ}$ F)
 $P = 1114$ kPa (150 psig)
 $V_{\text{filter}} = 4.3\text{--}5.2$ m/min (14.2–17.2 ft/min)
 $\eta = 99.80\%$
 $C = 3.7$ to 3.04 g/Nm 3 (2844–2357 ppm)

that there had been no irreversible plugging at the 5.2 m/min (17.2 ft/min) velocity. Posttest mass balance indicated that the inlet dust loading was 3.67 g/Nm^3 (2844 ppm) and that, in spite of the high velocity and frequent and vigorous blowback regimen, the overall efficiency of the filter was still 99.80%.

One final run was conducted with filter element 3W11 in which the filter velocity was stepped from 1.7 to 3.4 to 5.4 m/min (5.7 to 11.4 to 17.6 to 14.1 ft/min) and back down. Test 1.15.82 was carried out at the same temperature and pressure that previous runs had been and is shown in Figure 3.3-78. For the first six cycles the filter velocity was maintained at 1.7 m/min (5.7 ft/min) and the pressure drop cycled from about 1.5 to 3 kPa (6 to 12 in. wg) over about a 20 minute period. The velocity was then doubled to 3.4 m/min (11.4 ft/min), and stable cycles from 3.5 to 6 kPa (14 to 24 in. wg) were established and repeated four times. The velocity was then increased to 5.4 m/min (17.6 ft/min), but the cycle times to the trigger pressure drop of 6 kPa (24 in. wg) were so short that it was impractical to operate at these conditions.

As a result the velocity was lowered to 4.30 m/min (14.1 ft/min) for three cycles. These cycles started at 5.25 kPa (21 in. wg) and ran up to 6 kPa (24 in. wg) in about 1 minute. Following this, the velocity was dropped to 3.4 m/min (11.4 ft/min) for two cycles between 4 and 6 kPa (16 and 24 in. wg), and finally to the last cycle at 1.7 m/min (5.7 ft/min). The last cycle ranged from 2 kPa (8 in. wg) to 3 kPa (12 in. wg) over an 8 minute time span, indicating (by comparison to the first six cycles of the day) that some ground had been lost because of the high filter velocities tried. We determined that the overall efficiency remained at 99.85%, where it had been for the past several tests, and that the overall average inlet concentration had been about 3.68 g/Nm^3 (2850 ppm).

At this point we decided to open the filter housing for a visual inspection of the filter element. This was the first time in nearly two months that the filter had been removed. In this period it had been operated over 120 hours at temperature and pressure, had been blown back

Figure 3.3-78 Test 1.15.82 Pressure vs Time for Ceramic Cross-Flow Filter 1.5 mm (60 mil) Thick, 100 μ m Pore Size

Conditions: $T = 1088 \text{ K} (1500^{\circ}\text{F})$
 $P = 1114 \text{ kPa} (150 \text{ psig})$
 $V_{\text{filter}} = 1.7 \text{ to } 5.25 \text{ m/min} (15.7 \text{ to } 17.6 \text{ ft/min})$
 $\eta = 99.89\%$
 $C_{\text{avg}} = 3.7 \text{ g/Nm}^3 (2850 \text{ ppm}), 3\text{WH-1}$

Figure 3.3-79 100 μ m 1.5 mm (60 mil) Test Filter after 120 hr Use at 1088 K (1500°F), 1114 kPa (150 psig) and Filter Velocities Ranging from 1.7 to 5.4 m/min (5.7 to 17.6 ft/min), 25 Thermal Cycles, and 600 Blowback Sequences, Element 3WH-1

602 times, and had been thermally cycled from ambient to 1088 K (1500°F) and back down approximately 25 times. In addition, the last several test sequences we had pushed the filter velocities exceptionally high, in order to explore the filters' capabilities. Figure 3.3-79 is a photograph of the element in the as-removed condition. The exterior face of the filter was fairly heavily coated but not entirely covered. For some reason a large lump of soft dust had formed on the top of the element, as shown in Figure 3.3-79. Figure 3.3-80 is a photograph of the element with the exterior surface dust removed. One can get a reasonable feel for the degree to which the channels had become blocked with harder to remove dust cake. Nothing was sintered or hard, and all the resident dust could easily be blown out of the channels. There was some indication of minor variation in the cleaning from top (right side) to bottom (left side), with the lower channels being slightly cleaner. The probable source of the minor leak in the system was discovered to be a delamination of the bottom plate on the filter, as indicated in Figure 3.3-81. This crack did not extend down through the filter to the clean side face and thus did not represent a major leak in the system.

By this time Coors had delivered a new set of four filter elements (WH-5) for testing. These elements were more porous than any of the previously tested materials, with a maximum nominal pore size of 144 μm and a backweb thickness of 1.5 mm (60 mil). The available filter surface area in the mounted 5.1 x 15.2 x 15.2 cm (2 x 6 x 6 in) element was estimated to be nearly 1250 cm^2 (200 in^2). The first new element (1WH-5) was placed in the system, and low pressure and temperature flow vs pressure drop measurements were conducted. The flow resistance for this clean, unused material was very low, typified by a pressure drop of 0.22 kPa (0.9 in. wg) at a filter velocity of 4.4 m/min (14.3 ft/min).

During the last part of this report period a series of four, day-long tests were carried out with this new material at the usual temperature and pressure. For test 1.22-26-27-28.82 a filter velocity of 1.7 m/min (5.7 ft/min) and a trigger pressure drop of 3 kPa (12 in. wg) were chosen so as to provide a good basis for comparison with the

Figure 3.3-80 100 μ m 1.5 mm (60 mil) Test Filter after 120 hr Use with External Dust Removed:
Note Apparent Better Cleaning in Channels at Bottom of Filter (left side), 3WH-1

Figure 3.3-81 Delamination Failure that Diminished Filter Efficiencies from 99.9+ to 99.8% for 100 μm , 1.5 mm (60 mil) Test Filter after 120 hr Use at 1088 K (1500°F), 1114 kPa (150 psig), and Filter Velocities from 1.7 to 5.4 m/min (5.7 to 17.6 ft/min), 25 Thermal Cycles, and 600 Blowback Sequences, Element 3WH-1

Figure 3.3-82 Test 1.22-26.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter 1.5 mm (60 mil) Thick, 140 μm Pore Size

Conditions: $T = 1088 \text{ K}$ (1500°F)

$P = 1114 \text{ kPa}$ (150 psig)

$V_{\text{filter}} = 1.7 \text{ m/min}$ (5.7 ft/min), 1WH-5

previously tested filter materials. The filter conditioning cycles shown in Figure 3.3-82 are extremely long, with this velocity and an average inlet loading of 3.87 to 4.52 g/Nm^3 (3000 to 3500 ppm). The first cycle took 3 hours to go from about 0.25 to 3 kPa (1 to 12 in. wg). After eight cycles the base-line pressure drop had increased only to about 0.5 kPa (2 in. wg), and the cycle time was still nearly 2 hours. During this test period we cleaned the filter with 1 to 2 pulses from a pulse reservoir charged to 2127 to 1462 kPa (300 to 350 psig). Throughout these tests virtually no penetration of dust was observed, and the overall efficiencies measured were typically 99.99%.

We were encouraged by these results, since they represented the first reasonably successful filter medium obtained since the very first shipment of 100 μm , 1.5 mm (60 mil) medium, WH-1.

Early in February, testing consisted of a continuation of runs from the previous month. Test 2.1-3.82 was carried out at the same conditions in a continued effort to establish a base line for performance. Figure 3.3-83 presents the time-pressure drop history for the two 15 hour test days. As in the previous tests, the cycle time was very long, with typical base-line drops of 0.5 kPa (2 in. wg) climbing to the trigger pressure drop of 3 kPa (12 in. wg) over a 2 to 2.5 hour period, with an inlet dust concentration of 3.36 to 3.61 g/Nm^3 (2600 to 2800 ppm) and a face velocity of 1.7 m/min (5.7 ft/min). Blowback cleaning during this test consisted of one or two 0.5 s pulses from a 2.12 ℓ (0.075 ft^3) reservoir charged to 2431 kPa (350 psig) and discharged through a 0.95 cm (3/8 in) tube that served as the nozzle. As indicated, very high efficiencies, exceeding 99.99%, were maintained throughout the test period. At this point better than 51 hours had been accumulated on the filter element and excellent stable filter behavior had been demonstrated. We decided to explore higher filtering velocities.

To this end test 2.5-10.82 was carried out at a filter face velocity of 3.5 m/min (11.4 ft/min), with dust loadings from 3.1 to 4.38 g/Nm^3 (2400 to 3400 ppm). The trigger pressure drop was also

Figure 3.3-83 Test (2.1-3.82) Pressure Drop vs Time for Ceramic Cross-Flow Filter 1.5 mm (60 mil) Thick, 144 μm Pore Size

Test Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$

$P = 1114 \text{ kPa (150 psi)}$

$V_{\text{filter}} = 1.7 \text{ m/min (5.7 ft/min)}$, Element 1WH-5

Figure 3.3-83 (cont'd)

Figure 3.3-84 Test 2.5-10.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter 1.5 mm (60 mil) Thick, 144 μm Pore Size

Test Conditions: $T = 1088$ (1500°F)

$P = 1114 \text{ kPa}$ (150 psig)

$V_{\text{filter}} = 3.5 \text{ m/min}$ (11.4 ft/min), 1WH-5

Figure 3.3-84 (cont'd)

doubled from 3 to 6 kPa (12 to 24 in. wg). As Figure 3.3-84 shows, the initial cycle consisted of a rise from about 1 to 6 kPa (4 to 24 in. wg) over a 100 minute period. Throughout the day this cycle time gradually believed deterioration was caused by a combination of three things: trouble with the dust injector, incomplete cleaning by the blowback schedule, and the usual (system-associated) rise in base pressure drop caused by gradual plugging of the exhaust lines over the course of the day. About midway through the first test day we had some trouble with the dust injector, when a vibrator inside the pressure vessel failed. This caused some downtime while temporary repairs were made and introduced some uncertainty about the dust feed rate upon restart because the rate is determined by initial and final weights and the assumption of constant feed rate, which may not have been valid for this run. Differentiating between incomplete cleaning and bypass line plugging is often difficult, but a feel for the situation can be obtained by looking at the last cycle of the first day and the first cycle of the second day (at time equal to about 3900 min). The base pressure drop at the end of the first day was about 2.5 kPa (10 in. wg). With the system down, the bypass and exhaust lines were cleaned, but the filter was undisturbed. On subsequent start-up, the base pressure drop was back down to about 1.25 kPa (5 in. wg).

The very steep rise in pressure drop at the start of the cycle is an indication of incomplete or "spotty" cleaning. As these spots of very clean surface are rapidly covered by dust, the filter flow begins to distribute itself uniformly over the filter, and there is a lower but constant rate of pressure drop rise. Cycles at the end of the second day had become reasonably constant, starting at 2.5 kPa (10 in. wg) and reaching the trigger point in about 25 minutes.

During these tests the first sign of measurable penetration occurred with an efficiency over the first day of 99.98% and a significant decline during the second day of 99.76%. Because of this decline in efficiency, the filter was removed from the test rig. Figure 3.3-85 shows the element LWH-5 in its as-removed state. There was a slight

As removed

Sealed side intact with no cracks.

Figure 3.3-85 Ceramic Cross-Flow Filter after 75 hr Test at 1088 K (1500°F), 1114 kPa (150 psig) at Filter Velocities of 1.7 to 3.4 m/min (5.7 to 11.4 ft/min), LWH-5

variation in filter cleaning, as usual, from top to bottom. Figure 3.3-85 also shows the sealed back side of the filter, which is an area prone to cracking; there was no failure at this surface. The element was vacuumed clean so the entire surface could be examined. After this some cracks were observed at the top of the element near the front sealing face. These cracks are shown in the top picture of Figure 3.3-86 before the element was removed from the element holder. After the filter had been removed from the holding fixture, several major breaks were evident, as shown in the lower portion of Figure 3.3-86. Almost all prior failures had been simple delaminations, but in this case breaks occurred across the filter flats at several locations. Additionally, the surface that pressed into the sealing gasket appeared to crumble in such a way that, when the element was removed from the holder, several small pieces of the filter material in contact with the gasket stuck to it. Failure of this nature probably indicated that the method used to introduce the increased pore size, or the increased pore size itself, had diminished the strength of the body to an unacceptable level.

A second piece of this material (2WH-5) was prepared for testing and installed in the test rig. Test 2.18.82 was carried out at the usual operating conditions except that the filtering velocity was very high, 5.2 m/min (17.2 ft/min). Figure 3.3-87 presents the results from this run. The first cycle was, as usual, quite long as the new filter became conditioned. Subsequent cycles were remarkably stable for these extreme conditions of velocity and dust loading [2.16 g/Nm^3 (1673 ppm)]. A stable base pressure drop of about 1.25 kPa (5 in. wg) was established, as was a regular cycle time of 25 to 30 minutes, for the rise in pressure drop to reach the trigger pressure drop of 6 kPa (24 in. wg). The initial cleaning pulses were 0.5 s at 2431 kPa (350 psig) and were raised to 2705 kPa (400 psig) about midway through the test. The only difficulty experienced during the test was a slight penetration of dust that corresponded to an outlet loading of about 0.0026 g/Nm^3 (2 ppm). This leak results in an overall average filter efficiency of 99.91% but was enough for us to remove the filter for examination.

Cracks observed while filter is still mounted.

Breakage observed after filter was removed from holder.

Figure 3.3-86 Ceramic Cross-Flow Filter Failure after 75 hr Testing
1WH-5

Figure 3.3-87 Test 2.18.22 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1.5 mm (60 mil) Thick, 144 μm Pore Size
 Conditions: $T = 1088 \text{ K}$ (1500°F)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_{\text{Filter}} = 5.2 \text{ m/min}$ (17.2 ft/min)
 $C = 2.14 \text{ g/Nm}^3$ (1673 ppm)
 $\eta = 99.91\%$

Figure 3.3-88 is a photograph of the filter element in the as-removed condition, which appears to be good. After cleaning and very close inspection, however, the potential source of the observed leaks was apparent. These sites are shown in Figure 3.3-89 and consist of small cracks through the filter flat at the sealed end of the filter. These cracks were noticeable at only a few points, where a fillet of Aremco adhesive had inadvertently been omitted in the preparation of the filter. Also, the crumbling behavior of the filter surface in contact with the gasket that had been observed in the previous test occurred again. This phenomenon is shown in Figure 3.3-90, which shows a small area of the front "clean face" that chipped or crumbled off when removed from the gasketed surface.

Both of the most recently tested filter elements were prepared and shipped back to Coors for their examination.

The first test conducted in March was carried out on the third of four identical filter elements supplied by Coors (3 WH-5). These elements were fabricated to give a 140 μm pore size, 1.5 mm (60 mil) backwall, and an overall filter surface area of 0.093 m^2 (1.3 ft^2) in a 5 × 15.2 × 15.2 cm (2 × 6 × 6 in) element. All testing in this period was carried out at a temperature of 1088 K (1500°F) and at a pressure of 1114 kPa (150 psig).

Test 3.4-82 was carried out over two test days for a total of about 16 actual test hours. The filter velocity chosen for this test was very high, 5.2 m/min (17 ft/min), and the inlet dust concentration varied from 1.76 to 2.24 g/Nm³ (1360 to 1736 ppm) over the two test days. Figure 3.3-91 presents the pressure drop vs time history for this run. As indicated by this initial cycle, the resistance of the clean filter was very low at about 0.3 kPa (1.2 in. wg) and 5.2 m/min (17 ft/min). The pressure drop was allowed to rise to 6 kPa (24 in. wg) before the cleaning cycle was initiated. This consisted of a 0.5 s pulse from a 2431 kPa (350 psig) reservoir through a 0.95 cm (0.375 in) nozzle. As is the usual case, the first cycle, which lasted almost

Figure 3.3-88 As-Removed Condition of 140 μm , 1.5 mm (60 mil)
Filter Unit after Test 2.18.82, 2WH-5

Figure 3.3-89 Cross-Flow Filter after Test 2.18.82 Showing Cracks
in 3 Layers at Sealed End Where Filter of Cement Was
Missed 2WH-5

Figure 3.3-90 Clean Face of Ceramic Cross-Flow Filter after Test 2.18.82, Showing How Sections of Filter Medium Crumbled Where Compressed into Gasket, 2VH-5

Figure 3.3-91 Pressure Drop vs Time for Ceramic Cross-Flow Filter 1.5 mm (60 mil) Thick, 140 μm Pore
 Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $V_{\text{Filter}} = 5.2 \text{ m/min (17 ft/min)}$

2 hours, was considerably longer than subsequent cycles, which lasted about 30 to 40 minutes, during the first day. During this day the average inlet loading was measured at 2.24 g/Nm^3 (1736 ppm). The base-line pressure drop increased from about 0.5 kPa (2 in. wg) to about 1 kPa (4 in. wg) at the end of the day some 5 hours later. There was virtually no penetration of dust during the first day of this test and overall efficiencies were measured at 99.99+%. The second day of testing began with cycles very similar to the cycles observed at the end of the first day. After about three hours there was a significant decrease in cycle time with an increase in the base-line pressure drop from about 1.25 kPa (5 in. wg) to about 2.5 kPa (10 in. wg). Over the remainder of the test the base-line pressure drop increased slowly to about 3 kPa (12 in. wg), where it appeared to be stable. A much longer test period would be required, however, to confirm the absence of blinding. During this second day the inlet dust concentration was 1.76 g/Nm^3 (1360 ppm), and the average overall efficiency was measured at 99.56%, with measured outlet loadings in the 0.0052 to 0.0155 g/Nm^3 (4 to 12 ppm) range.

While the measured efficiency was still quite high, past experience indicated that a failure of some nature had occurred, so the filter was removed for inspection. Figures 3.3-92 through -94 show what we discovered. Figure 3.3-93 shows the condition of the element just as it was removed from the test system. The external appearance was good, with no obvious plugging or maldistribution of cleaning. It shows a major break in the top plate of the filter, discovered after the element had been vacuumed but left in the filter holder. Figure 3.3-94 indicates that, in addition to the top plate break, a section below this plate had failed similarly. As in previous tests, we noticed that some crumbling of the clean face occurred, although we believe this occurred when the element was "rocked" out of its seal. Since this failure was similar to that of two previous specimens (1 WH-5 and 2 WH-5), we decided to test the most recently received material from Coors.

Figure 3.3-92 Test (3.4-9.82) As-Removed Appearance of Filter
after 18 Hours of Service, 3WH-5

Crack in Top Plate of Filter

Side View of Crack

Figure 3.3-93 Test (3.4-9.82) Failures Observed in Filter prior to Removal from Filter Holder, 3WH-5

Failure of Top Plate and Second Row Down

Note Crumbling of Clean Face at Gasket Interface

Figure 3.3-94 Test (3.4-9.82) - Photo Showing Filter Failures When Removed from Filter Holder, 3WH-5

The new filters (WH-6) were made from the same pressed plates as the previous element, so they had identical pore sizes and distributions. The primary difference was that the new element had a backwall (filtering flat) that was twice as thick as that of the previous elements [3 mm (120 mils)]. Additionally, more flat surface was allowed for at the sealed end of the filter to provide better resistance to delamination. One further change was implemented during this test, the omission of the usual reinforcement of joints with ceramic adhesive. In our discussions with Coors, they had expressed some concern about the use of these adhesives and recommended that we try at least one test without them.

Test 3.16.82 was carried out at a filter velocity of 5.2 m/min (17 ft/min) with an inlet dust loading of 2.47 g/Nm^3 (1910 ppm). The pressure drop history is shown in Figure 3.3-95. As expected, the base pressure drop for the new material was 1 kPa (4 in. wg), or twice that observed in the previous test, because of the doubled thickness. This initial cycle was essentially the same as that with the thinner filter, rising from the base pressure drop to the trigger pressure drop, 6 kPa (24 in. wg), in just less than 2 hours. Using the same 2431 kPa (350 psig) cleaning pulse returned the pressure drop to 1.25 kPa (5 in. wg) initially. This base-line pressure rose gradually to about 3.75 kPa (11 in. wg) at the end of the test day some 11 hours later. The primary difficulty experienced during the run was the significant penetration observed at the onset, which increased over the test period. A sample taken over the first cycle indicated an outlet concentration of 0.0336 g/Nm^3 (26 ppm). This increased steadily over the next 4 hours to an outlet concentration of about 0.284 to 0.323 g/Nm^3 (220 to 250 ppm), where it remained for the rest of the test. Following this test period the filter was removed for examination. Figure 3.3-96 shows the as-removed and the cleaned condition of the filter. Several major breaks occurred in the lower portion of the filter, which led to the very poor efficiency. This failure was very similar to that experienced with the thinner backwall material. There was also some indication of poorer cleaning at the top of the filter, as the upper

Figure 3.3-95 Test 3.16.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 3 mm Thick (120 mil), 144 μm Pore Size
 Conditions: Temperature 1088 K (1500 $^{\circ}\text{F}$)
 Pressure = 1114 kPa (150 psig)
 $V_{\text{Filter}} = 5.2 \text{ m/min}$ (17 ft/min)
 $C = 2.47 \text{ g/Nm}^3$ (1910 ppm)
 $\eta = 91.17\%$

As Removed

Vacuumed

Figure 3.3-96 Test (3.16.82) Failures Observed with Thick Membrane Cross-Flow Filter, 1WH-6

several channels were nearly plugged shut. No crumbling was observed at the clean face gasket interface, an indication that although the ceramic adhesive improves the resistance to delamination failure, it would appear to cause the clean-face crumbling observed at filter removal.

To follow up on this conjecture, a second 3 mm (120 mil) back-wall element (2WH-6) was prepared for test in the usual way. This included sealing the filter joints with the ceramic cement and then sealing the filter into the holder with the same adhesive.

Test 3.19-23.82 was carried out over filtration velocities ranging from 1.7 to 5.2 m/min (5.7 to 17.2 ft/min). The measured dust concentrations were unusually high during the first day of this two-day test, ranging from approximately 4.39 to 7.49 g/Nm³ (3400 to 5800 ppm). This led to unusually short cycle times, as shown in Figure 3.3-97. The clean filter base pressure drop at 5.2 m/min (17.2 ft/min) was 1.5 kPa (6 in. wg). The pressure drop rose to 6 kPa (24 in. wg) in about 35 minutes at an estimated inlet loading of 4.39 g/Nm³ (3400 ppm). Subsequent cycles shortened from about 8 minutes to 2 minutes as the base pressure drop gradually rose from 3.25 kPa (13 in. wg) to 4.5 kPa (18 in. wg). Outlet concentrations measured during this time were very low, averaging less than 0.0013 g/Nm³ (1 ppm). After 4 hours of operation in this mode, the air mass flow was reduced to a filter velocity of 3.4 m/min (11 ft/min). The base pressure drop came down to 2.75 kPa (11 in. wg), and the cycle time to the trigger pressure loss extended out to about 10 minutes, with an estimated inlet loading of 6.85 g/Nm³ (5300 ppm). The system was operated in a stable condition for ten cycles over 2.5 hours, during which time virtually no dust penetration was observed. During the final 4 hours of this test day, the velocity was reduced to 1.7 m/min (5.7 ft/min). An absolutely steady operating cycle was immediately established, running from 1.4 kPa (5.5 in. wg) to 6 kPa (24 in. wg) over about 0.5-hour time increments. The dust loading was estimated to be 7.49 g/Nm³ (5800 ppm) during this period and, as before, there was virtually no penetration of the filter. All blowback

Figure 3.3-97 Test (3.19-23.82) Pressure Drop vs Time for Ceramic Cross-Flow Filter, 3 mm (120 mil) Thick, 140 μm Pore Size
 Conditions: $T = 1088 \text{ K (1500}^\circ\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $\eta = 99.99\% \text{ (3/19), } 99.53\% \text{ (3/23), Element 2WH-6}$

Figure 3.3-97 Continued

pulses prior to $t = 150$ minutes were from a reservoir charged to 2431 kPa (350 psig), while all subsequent pulses were at 2800 kPa (400 psig).

During the second test day of this run, we experienced some difficulty when the dust injector lines plugged, so the dust concentration during the first two cycles is not precisely known. We estimated it at about 3.77 g/Nm^3 (2917 ppm) for the first half of the run. The filter velocity during these first two cycles was set at 1.7 m/min (5.7 ft/min). The longer cycle time reflects the somewhat lower dust concentration and the fact that the bypass line had been cleaned between runs, resulting in a lower base pressure drop, about 0.62 kPa (2.5 in. wg), than in the previous day's 1.25 kPa (5 in. wg) base. Some measurable penetration of dust was observed from the onset of the second day of testing. Outlet dust concentrations averaged about 0.011 g/Nm^3 (8.7 ppm), leading to an average efficiency of 99.70%. After approximately 4 hours of operation at 1.7 m/min (5.7 ft/min) the velocity was doubled to 3.4 m/min (11.4 ft/min), while the dust feed rate remained almost constant. Over the remaining 4.5 hours of testing, the average inlet concentration was 1.95 g/Nm^3 (1507 ppm), and the efficiency declined to 99.41%. The initial cycle at the higher velocity took about an hour to rise from a pressure drop of 0.5 kPa (2 in. wg) to 6 kPa (24 in. wg), while subsequent cycles shortened to a stable cycle of 25 minutes from 2.5 kPa (10 in. wg) to the same trigger pressure loss.

Upon completion of this test, the filter was removed from the system for inspection. Figure 3.3-98 shows the condition of the filter in its as-removed condition. Some degree of nonuniform cleaning had occurred, with the least cleaning in the upper-outer corner of the filter, as would be expected from previous experience. This nonuniformity may be accentuated because the thicker wall elements have their channel opening diminished in height by the amount of increased backwall thickness. As indicated in Figure 3.3-98, when the element was vacuumed one major crack was noticed. Figure 3.3-99 indicates that the break extended across the sealed end and, in fact, had led to delamination of the entire element. Figure 3.3-100 is included to demonstrate the

As Removed

Vacuumed

Figure 3.3-98 Test (3.19-23.82) Failure Observed prior to Removal from Filter Holder, 2WH-6

Delamination and Filter Flat Failure

Failure at Sealed End

Figure 3.3-99 Test (3.19-23.82) Failures Observed upon Removal from Filter Holder, 2WH-6

Figure 3.3-100 Test 3.19-23.82. Photo Showing Manner in which Clean Face of Filter Crumbles, 2WH-6

degree to which this particular element's front face had crumbled when it was removed from the filter holder. We believe that the practice of clamping the wet, glued-up element in the holder, sealing the element in with additional glue, and then finally curing the cement in place causes bonding between the metal holder and the ceramic filter. The adhesive-to-metal bond is strong enough to cause damage under differential thermal expansion and/or the forces used to remove the element from the holder. The use of a mold release agent for the filter element holder would probably allow a good, dust-tight seal to be formed without bonding the filter to the metal parts.

No actual filter testing was done during the next two and a half months because the last set of filter elements (WH-7) was still being fabricated at Coors. We had been concerned that in some part the recent mechanical failures of the test filters may have been due to thermal effects during start-up and shutdown. During start-up, for example, stable combustion could only be maintained with air temperatures at about 811 K (1000°F), so both the filter element and the metallic holder were exposed to high rates of heating. Conversely on shutdown, past procedures called for the maintenance of purge flow after flameout in the combustor with the consequence of high cooling rates. In an attempt to mitigate the start-up effects, we plumbed a dilution air line into the inlet sample port. The combustor could then be operated stably at high temperature while the downstream dilution ensured cooler initial temperatures in the filter chamber. A few preliminary experiments were carried out to try the new system, and it appeared to work well. The initial filter temperature rise could be limited to 422 to 478 K (300 to 400°F), and the subsequent rate of temperature increase could also be controlled at will. A new shutdown procedure was initiated in which both the propane and airflow were shut simultaneously so that the filter holder temperature decayed very slowly by heat loss through the chamber walls. We anticipated that these procedures would maximize the filter element resistance to mechanical failure in the final test series.

The final series of testing resumed at the end of July and extended to the first nine days of August. During this time a series of five single-day experiments were carried out on the test element most recently provided to Westinghouse, 1WH-7. The intent of this final series was to demonstrate that with the experience gained throughout the test program a mechanically sound filter element could be fabricated and employed at high temperature and pressure.

The test conditions chosen for these tests were as follows:

Temperature = 1088 K (1500°F)

Pressure = 1114 kPa (150 psig)

Dust Concentration = 5.17 to 6.46 g/Nm³ (4000-5000 ppm)

Filter Velocity = 3.35 m/min (11 ft/min)

Blow Back = 2512 kPa (350 psig)/0.5 s.

During these tests the new start-up and shutdown procedure was employed in order to minimize transient thermal effects between the filter and the metallic mounting assembly.

Test 7.28.82 was the filter 1WH-7's first exposure to dust and, as is typical, the first cycle was very long. This cycle consisted of a rise in pressure drop from about 1 kPa (4 in. wg) to the trigger pressure drop of 3 kPa (12 in. wg) over a 145 minute period, for an average rate of pressure drop rise of 0.014 kPa/min (0.055 in. wg/min). As the filter became conditioned, subsequent cycles shortened, the second consisting of a rise from 1.1 kPa (4.5 in. wg) to trigger in 93 minutes or 0.020 kPa/min (0.081 in. wg/min). The test ended part way through the third cycle, but the base pressure drop remained at 1.1 kPa (4.5 in. wg), and the rate of rise was 0.019 kPa/min (0.078 in. wg/min), indicating a fairly stable condition. The dust concentration was measured at 6.04 g/Nm³ (4676 ppm) for this test and the overall efficiency at 99.86%. The pressure drop trace for this run is presented in Figure 3.3-101.

Figure 3.3-101 Test 7.28.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1WH-7 (100 μ m Pore Size, 1.5 mm (60 mil) Thick
 Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $V_F = 3.35 \text{ m/min (11.0 ft/min)}$

Test 7.30.82 was merely a continuation of the previous test. During this test we noted that the first cycle had lengthened considerably, taking nearly 150 minutes to rise from 1 kPa (4 in. wg) to 3 kPa (12 in. wg). In addition, the overall efficiency dropped to 99.71%, an indication that something was probably wrong. Just prior to blowback at the end of the first cycle there was difficulty with the fuel supply, which caused a passage shutdown. This minor problem was fixed, and we decided to carry out one more run before opening the filter vessel and examining the element. Figure 3.3-102 is a pressure drop trace of this test day.

Test 8.2.82 was a continuation test, and, again, the first cycle lasted longer than expected, taking 150 minutes to go from 3 kPa (4 to 12 in. wg). Cleaning was accomplished using the usual 2512 kPa (350 psig) pulse for 0.5 s. The second cycle was somewhat shorter, lasting about 100 minutes and starting from 1.25 kPa (5 in. wg). The pressure drop trace from this run is in Figure 3.3-103. During this test the overall efficiency declined to 99.60%, and considerable penetration was evident. We decided, therefore, to remove the filter element for examination.

When the element was removed for examination it appeared to be in excellent shape, as shown in Figure 3.3-104. Cleaning appeared to be complete and uniform and, despite the fact that the bottom end of the clamping bar was fairly loose, there was no indication of filter failure. The element was cleaned for closer inspection and, as shown in Figure 3.3-104, no failure was observed. Finally, the clamping bar was removed prior to being tightened so that the clamped end, a common site of failure, could be examined. As Figure 3.3-104 shows, there was no evident flaw. We then turned our attention to the test vessel for an explanation of the penetration. Figure 3.3-105 is a photograph of the ceramic fiber gasket used to seal the dirty side of the filter chamber and the clean side. It was evident from the dust tracks that a considerable leak had occurred in two places. The mating surface of the clean-side/dirty-side plenum is shown in Figure 3.3-105 and also gives evidence of the poor seal. When the plenum plate was examined for

Figure 3.3-102 Test 7.30.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1WH-7,
 100 μm Pore Size, 1.5 mm (60 mil) Thick
 Conditions: $T = 1088 \text{ K}$ (1500°F)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_F = 3.35 \text{ m/min}$ (11.0 ft/min)
 $\eta = 99.71\%$
 $C = 5.33 \text{ g/Nm}^3$ (4.124 ppm)

Figure 3.3-103 Test 8.2.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1WH-7, 100 μm Pore Size, 1.5 mm (60 mil) Thick
 Conditions: $T = 1088 \text{ K}$ (1500°F)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_F = 3.35 \text{ m/min}$ (11.0 ft/min)
 $\eta_F = 99.60\%$
 $C = 5.86 \text{ g/Nm}^3$ (4787 ppm)

Figure 3 3-104 Test Element LWH-7 after Three Days of Testing in Tests 7. '8.82 to 8.2.82

Inlet tra ke gasket

Corresponding tra ke o plenum plate

Wall plate

Figure 3.3-105 Probable Source of Penetration in Tests 7.28.82 to 8.2.82 Was Warped Filter Mounting/Plenum Assembly

flatness we found it to be warped almost 3.2 mm (0.125 in) in the area of the leak, as shown in Figure 3.3-105.

Since it was impossible to remove the filter element from the mounting unit without risking damage to the element, we could not mechanically straighten the plate. We decided to try to remedy the leak by adding extra gasketing material in the warped area and to tighten the clamping studs carefully when we reinstalled the filter.

Test 8.5.82 picked up testing after the filter was reinstalled. As shown in Figure 3.3-106, the cycle times were much shorter (as expected), and the overall efficiency was again very high, 99.96%. The clean base-line pressure drop started at 1.6 kPa (6.5 in. wg) and leveled out after a few cycles at about 2 kPa (8 in. wg). Cycle times increased from an initial 35 minutes to a steady 20 to 25 minutes after only 3 or 4 cycles.

Before the next test (8.9.82) the bypass line was cleaned and, as is usual, resulted in a slight increase in cycle time for the first cycle or two of the test. This test was a repeat of the previous test, and, as is indicated by Figure 3.3-107, we obtained essentially identical results. Base-line pressure drops stabilized at about 2 to 2.25 kPa (8 to 9 in. wg) and cycle times were about 20 to 25 min. The efficiency decreased slightly to 99.76% during the second day. When the filter was removed for the last time, a leak path had again been established through the plenum/gasket interface. The filter, however, was in excellent shape, showing very good and uniform cleaning (see Figure 3.3-108). The filter element was successfully removed from the mounting without damage and was stored for possible future use.

This element, 1WH-7, experienced 21 hours of filtration time, was pulse cleaned 70 times, and experienced 5 thermal cycles without damage. While this is not conclusive proof of acceptable life, it does represent a significant improvement over the most recently tested elements.

Figure 3.3-106 Test 8.5.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1WH-7, 100 μm Pore Size, 1.5 mm (60 mil) Thick
 Conditions: $T = 1088 \text{ K (1500}^{\circ}\text{F)}$
 $P = 1114 \text{ kPa (150 psig)}$
 $V_F = 3.35 \text{ m/min (11.0 ft/min)}$
 $\eta = 99.96\%$
 $C = 6.42 \text{ g/Nm}^3 \text{ (4968 ppm)}$

Figure 3.3- 107 Test 8.9.82 Pressure Drop vs Time for Ceramic Cross-Flow Filter, 1WH-7, 100 μm Pore Size, 1.5 mm (60 mil) Thick
 Conditions: $T = 1088 \text{ K}$ (1500°F)
 $P = 1114 \text{ kPa}$ (150 psig)
 $V_F = 3.35 \text{ m/min}$ (11.0 ft/min)
 $\eta = 99.76\%$
 $C = 5.53 \text{ g/Nm}^3$ (4281 ppm)

As removed

After vacuumed

Figure 3.3-108 Cross Flow Filter Condition after Testing Test Element 1WH-7

3.4 COMMERCIAL MODULE DESIGN

3.4.1 Summary

The preliminary conceptual design of a full-scale, cross-flow filter module has been completed on the basis of the experimental results of the test program. In addition to the module design, specific application to the particular PFBC system chosen for the economic study task of this project has been analyzed.

On the basis of fabrication constraints, the individual filter elements comprising a module are $30.5 \times 30.5 \times 10.2$ cm ($12 \times 12 \times 4$ in). Each module incorporates 40 such elements (10 to a side) in a rectangular array, so that the overall module dimensions are roughly 4.9 m (16 ft) long by 0.6 m (2 ft) square. Each module is capable of processing 100 to 168 m^3/min (3500 to 6000 acfm) of flue gas.

Module cost estimates have been developed with Burns & Roe, and the estimated installed cost of a module is \$14,230, or $\$2.23/\text{m}^2$ ($\$24/\text{ft}^2$) of filter area.

In order to facilitate the overall plant economics study we have developed a concept for the installation of many filter modules into a pressure vessel unit suitable for insertion in a chosen PFBC reference plant design. These units are typically shop fabricated, 4.3 to 4.9 m (14 to 16 ft) diameter by 11 m (35 ft) tall vessels containing 15 to 20 modules. Nominal flow per unit is on the order of 2,832 m^3/min (100,000 acfm). An installed cost for such units has been derived and is estimated at about \$687,500/unit ($\$6.89/\text{acfm}$) or \$8 to 10/ m^2 ($\90 to 108/ ft^2) of installed filter area.

3.4.2 Module Design Results

The preliminary conceptual design of a commercial module utilizing the cross-flow filter concept has been completed. Our first efforts centered on attempts to devise a configuration for mounting the individual elements in an array or module of useful scale. We finally

Figure 3.4-1 Preliminary Filter Module Design, Nominal Capacity is 100-168 m³/min (3500-6000 acfm)

settled on a configuration in which many filter elements are mounted on the sides of a square conduit, as shown in Figure 3.4-1. On the basis of experimental data gathered in this program and the fabrication constraints imposed, we have chosen the filter element geometry and operating conditions, which we review here. The overall dimensions of the elements are $30.5 \times 30.5 \times 10.2$ cm ($12 \times 12 \times 4$ in). The filtering membrane thickness will be 1.5 mm (0.060 in) thick and have an average pore size of 100 μ m. The rib height will be 3.8 mm (0.15 in) and be spaced 13.5 mm (0.53 in) apart. This configuration will yield about 1.38 m^2 (14.85 ft^2) of filter area per element. As indicated, a typical module will be about 4.6 m (15 ft) long and carry about 40 elements. At conservative filter velocities of 1.8 to 3.05 m/min (6 to 10 ft/min) each module will handle from 100 to 168 m^3/min (3500 to 6000 acfm).

The module layout and pressure vessel design were chosen to be compatible with the Burns & Roe economic studies and are shown in Figure 3.4-2. This vessel houses 18 modules and is capable of filtering a nominal flow of 2832 m^3/min (100,000 acfm). It is obvious that slightly larger (field-fabricated) vessels could process much greater flows.

The actual vessel design calls for a 4.92 m (16 ft, 1.75 in) by 10.21 m (33 ft 6 in) long carbon steel vessel lined with 0.15 m (6 in) of refractory. Inlet and cleaned gas outlet nozzles are 1.83 m (6.5 ft) and corresponding velocities are about 15.3 m/s (50 ft/sec). All metallic internals are fabricated from RA 333 for high-temperature strength and creep resistance. As shown in Section B-B of Figure 3.4-2, the filter modules are supported by channels that run across the 4.11 m (13.5 ft) support plenum. The actual plenum floor itself, then, can be fabricated from fairly light plate, since its main function is to serve as a clean-/dirty-side barrier rather than as a support for the filter modules. The suspended conical section comprising the upper part of the plenum assembly carries the static load of the filter system and can accommodate thermal expansion and contraction.

Figure 3.4-2 Cross-Flow Filter Unit - Shop-Fabricated Unit, Nominal Capacity (2832 m³/min 100,000 action)

A blowback pulse pipe 7.6 cm (3 in) is required for each of the filter modules. These blowback pulse pipes penetrate the vessel head and are then manifolded into a single header to the blowback pulse reservoir. The experimental program indicated that the proper choice of cleaning parameters is crucial to acceptable filter operation. For this reason considerable time and effort have been expended on an analysis of the blowback process and its scale-up. A design procedure for scaling blowback systems has been developed for this effort and is presented in Appendix D.

On the basis of preliminary CXF (cross-flow filter) module performance and fabrication data gathered by Westinghouse and Burns & Roe, the CXF/HGC system-related capital cost estimates were developed for the Burns & Roe plant economic analysis task.

Table 3.4-1 presents an installed cost summary of the CXF units and attendant major blowback system components for the 2-cyclone and 1-cyclone CXF/HGC systems.

Table 3.4-2 presents a summary of CXF module component weight and installed cost for the nominally 4.88 m (16 ft) high module, consisting of 40 elements mounted on a common plenum, shown in Figure 3.4-1.

Table 3.4-3 presents a summary of CXF unit housing weight and installed cost of the nominal 4.8 m (16 ft) diameter by 9.75 m (32 ft) long unit, shown in Figure 3.4-2, which contains 18 modules.

Table 3.4-4 presents a summary of CXF unit housing weight and installed cost for the nominal 4.3 m (14 ft) diameter by 9.75 m (32 ft) long unit, which contains 14 modules. The 4.3 m (14 ft) diameter unit closely approximates the 4.8 m (16 ft) diameter unit shown in Figure 3.4-2. These cost summaries indicate that installed cost can be scaled very closely by the square of the vessel diameter.

Table 3.4-1

CXF/HGC SYSTEM INSTALLED COST SUMMARY FOR A NOMINAL 500 MW STEAM-COOLED
PFBC COMBINED-CYCLE PLANT (January 1980 \$1000)

<u>HGC System Configuration</u>	<u>2-Cyclone/CXF</u>	<u>1-Cyclone/CXF</u>
CXF Units:		
CXF Modules (128)	1,821,440	1,821,440
Pressure Vessels (8)	<u>3,273,760</u>	<u>3,273,760</u>
Subtotal	5,095,200	5,095,200
Blow-back System:		
Compressors	224,000	118,000
Interconnecting Piping	36,000	33,000
Heaters	100,000	100,000
Accumulators	32,800	32,800
316 SS Pulse Air Manifold	16,000	16,000
Miscellaneous Insulation	<u>12,800</u>	<u>12,800</u>
Subtotal	<u>421,600</u>	<u>312,600</u>
Total	5,516,800	5,407,800
Specific CXF/HGC Cost ¹ , \$/ft ²	90 to 108	88 to 106

(1) 51200 to 61440 ft² of filter surface area

Table 3.4-2

CXF MODULE WEIGHT AND COST SUMMARY

<u>Component</u>	<u>Material</u>	<u>Weight (lbs)</u>	<u>Cost (January 1980 dollars)</u>
Filter Elements ¹ (40)	Ceramic	500	4,760
Plenum	RA333	344	6,500
Clamps (40)	RA333	163	2,000
Mounting Hardware	Ti Alloy	<u>115</u>	<u>600</u>
Subtotal		1,122	13,860
Module Assembly and Installation			<u>370</u>
Total			14,230

(1) 10 to 12 ft² of filter surface area/element

Table 3.4-3

CXF UNIT PRESSURE VESSEL HOUSING WEIGHT AND INSTALLED COST SUMMARY
FOR 18 MODULE VESSEL PBF COMBUSTOR CXF/HGC TRAIN

<u>Component</u>	<u>Material</u>	<u>Weight</u>	<u>Installed Cost (January 1980 dollars)</u>
Pressure Vessel	CS	100,000	230,100
1 In. Lining	Refractory	16,000	7,200
5 In. Lining	Insulation	23,100	17,300
Plenum Assembly	RA333	6,300	167,700
Pulse Air Pipes (18)	RA333	1,300	22,140
Solenoid Valves (18)	SS	<u>720</u>	<u>16,920</u>
		147,420	461,360

Table 3.4-4

CXF UNIT PRESSURE VESSEL HOUSING WEIGHT AND INSTALLED COST SUMMARY
FOR 14 MODULE VESSEL PFB COMBUSTOR CXF/HGC TRAIN

<u>Component</u>	<u>Material</u>	<u>Weight (lbs)</u>	<u>Installed Cost (January 1980 dollars)</u>
Pressure Vessel	CS	75,000	173,700
1 In. Lining	Refractory	13,200	5,900
5 In. Lining	Insulation	19,100	14,300
Plenum Assembly	RA333	5,000	132,800
Pulse Air Pipes (14)	RA333	1,000	17,200
Solenoid Valves (14)	SS	<u>560</u>	<u>13,160</u>
		113,860	357,080

3.5 POWER PLANT ECONOMIC ANALYSIS

3.5.1 Introduction

In order to assess the economic impact of successful implementation of the cross-flow filter (CXF) technology in PFBC systems, Burns & Roe, Inc., were contracted to perform an economic analysis of the ceramic cross-flow filter system and to integrate these results into an existing PFBC plant study. Westinghouse and Foster Wheeler had just completed a PFBC plant study under the sponsorship of DOE/ANL at this time, so a decision was made to use it as the base plant design.

This reference plant design is a nominally 500 MW steam-cooled, pressurized, fluidized-bed combustion power plant. It consists of two gas turbine and one steam turbine modules. Each gas turbine module has a PFB combustor/boiler module consisting of two vessels. Stack-gas coolers are used for recovering heat from the gas turbine exhausts.

Burns & Roe completed a study in which the base PFBC system incorporated the following hot gas cleaning (HGC) systems:

- Three stages of hot cyclones and a conventional baghouse (the reference case)
- Two stages of cyclones with the cross-flow ceramic filter system and no baghouse
- One stage of cyclone and the ceramic cross-flow filter system with no baghouse.

The original study also incorporated a conventional, pulverized-coal unit with flue gas desulfurization for a comparison with current and conventional technology. In the interim, some question has arisen over the choice of several critical design and operating parameters used in the reference PFBC study. This, in turn, has made comparisons with conventional PCF costs controversial. Ultimately, this has served only to cloud the issue under investigation here - namely, the comparison of the cross-flow filter-based systems with other "standard" methods for

dealing with hot gas cleaning. The original Burns & Roe study has been modified somewhat to emphasize the relative, rather than the absolute, costs, i.e., those associated with the use of the cross-flow filter system in a PFBC plant as compared to a more conventional, all cyclone system followed by a standard, low-temperature baghouse.

The conclusions we have drawn from the Burns & Roe study are summarized below.

3.5.1.1 Simplified Plant Layout

One of the assets of the cross-flow filter system is the extremely compact nature of the filter, which serves to reduce the complexity of the plant as well as the amount of expensive hot gas piping. This point is well illustrated by comparison of the reference PFBC system with the preferred cross-flow system. The reference system (three stages of cyclone with a baghouse) occupies an area roughly $143 \text{ m} \times 175 \text{ m}$ ($470 \text{ ft} \times 575 \text{ ft}$) and includes 24 cyclone vessels, each roughly 5.8 m (19 ft) in diameter and 7.6 m (25 ft) tall. The reference plant also includes two conventional baghouses for final filtration prior to exhausting up the stack. The one-cyclone, cross-flow filter system is situated in an area $143 \text{ m} \times 123 \text{ m}$ ($470 \text{ ft} \times 405 \text{ ft}$) and includes only 8 cyclone vessels and 8 cross-flow filter vessels, all of roughly the same dimensions. The cross-flow filter plant also requires no additional final cleaning system and totally eliminates the need for a baghouse. The particular plant design chosen does not make optimal use of the compact nature of the cross-flow system, as it would be possible to house all of the filter area required for each 123 MW gas turbine in a single field-fabricated vessel approximately 7.6 m (25 ft) in diameter and thereby reduce the total number of filter vessels from 8 to 2.

3.5.1.2 Significant Reduction in Cost of Hot Gas Cleaning

The Burns & Roe study presents a calculation that determines the specific hot gas cleaning system costs. This calculation indicates that the reference PFBC specific hot gas cleaning cost is \$125/kW net, while the one-cyclone cross-flow filter system is only \$69/kW net, representing a 45% reduction. This amount eventually translates into a 4.4% reduction in the levelized busbar power cost for the cross-flow system.

In the above analysis the operating and maintenance costs for the hot gas cleanup system are reached by taking a fixed fraction of operating and maintenance (O&M) labor costs. This amount is generally adequate for the purposes of estimating a final levelized cost of electricity (COE).

We have attempted to examine more closely the trade-off between the cost due to turbine blade erosion (which is experienced in the all-cyclone reference plant) and the cost of ceramic cross-flow filter replacement (due to possible degradation with time). To carry out this calculation, two components of COE were calculated for each system: the original hot-gas-cleaning system capital cost and the capital replacement cost. For the reference PFBC plant the capital replacement costs considered are the cost of turbine blade replacement and 2-to-4-year bag life for the bag filter system. The turbine blade replacement costs have been based on an optimistic 2-year life with a cost of \$42/kW per change-out. For the cross-flow system the capital replacement costs have been based on a yearly change-out of all the filter media. We believe this to be a very conservative estimate and expect substantially better life for the filter elements. Table 3.5-1 presents the results of these calculations. One can see that the estimated COE for hot gas cleaning in the one cyclone/CXF/HGC is less than half that for the reference system.

Table 3.5-1
COMPARISON OF COE

	Ref PFBC	1 Cyclone/CXF/HGC
Capital Equipment,		
\$	66,588,000	36,850,000
\$/kW	125	69
COE, mills/kWh	3.68	2.02
Capital Replacement,		
\$/yr	2,997,000	733,440
\$/kW-yr	5.64	1.37
COE, mills/kWh	1.74	0.42
Total COE Contribution,		
mills/kWh	5.42	2.44

3.5.2 EVALUATION OF CERAMIC CROSS-FLOW
FILTRATION FOR PFBC HOT GAS CLEANUP

34-RP-58253-A

SYSTEM DESCRIPTION AND ECONOMIC ANALYSIS

June 1982

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ABSTRACT

Westinghouse Research and Development Center (WRDC) has engaged Burns and Roe, Inc. (B&R) to provide a preliminary technical and economic evaluation of 500 MW steam-cooled pressurized fluidized-bed combustion (PFBC) combined cycle power systems, based on a reference plant developed from an existing conceptual design study, and attendant modifications to accommodate ceramic cross-flow filter (CXF) hot gas cleanup systems (HGC).

The following PFBC/HGC concepts are evaluated:

- PFBC plant utilizing three cyclonic stages and a baghouse
- PFBC plant utilizing two cyclonic stages and a ceramic cross-flow filter
- PFBC plant utilizing one cyclonic stage and a ceramic cross-flow filter

This report presents the results of the economic evaluation of the above PFBC plants burning eastern coal and utilizing an all-cyclone and combination of cyclone/CXF/HGC system. The analysis includes the development of the capital cost estimates resulting from the equipment and operational modifications attendant to the implementation of the CXF/HGC system concepts considered. The results of the economic analyses presented are based on estimates of plant capital and operating costs, reported in January 1980 dollars. Estimates for the busbar power cost include a breakdown of the fixed and variable constituents and are projected on a 30-year levelized cost basis. In addition, the design criteria and economic assumptions forming the basis of the evaluation, and presented in this report, have been jointly agreed upon by WRDC and B&R.

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Section 1

INTRODUCTION AND SUMMARY

Background

The steam-cooled pressurized fluidized-bed combustion (PFBC) combined cycle power plant concept offers the potential for significant improvement in thermal cycle efficiency over that obtainable from conventional pulverized coal-fired (PCF) electrical power generating plants with flue gas desulfurization (FGD). The PFBC concept is designed to burn noncompliance coal in an environmentally acceptable manner, consistent with the current environmental standards promulgated by the EPA in the June 1979 New Source Performance Standards (NSPS), and operates in combination with a gas turbine topping cycle and a steam bottoming cycle. In order to realize the high cycle efficiency potential of the PFBC combined cycle concept and to prevent or minimize corrosion and erosion of internal gas turbine components, an effective hot gas cleanup (HGC) system is required. The candidate HGC system must be economic and capable of effectively removing ash and spent sorbent particles from the PFB combustor effluent gas stream prior to its introduction into the gas turbine expander inlet. The ceramic cross-flow filtration (CXF) option evaluated in this study is considered to be a potentially effective HGC system alternative adaptable to the combined cycle power system concepts. The CXF/HGC system operates in combination with cyclones and does not require the inclusion of an electrostatic precipitator or baghouse filter to achieve compliance with current (June 1979) NSPS requirements for particulate emission. The CXF/HGC concept offers the potential to achieve compliance with possible future inhaleable particulate standards and/or a more stringent opacity standard. The ceramic cross-flow filter is a near perfect filter with regard to submicron particulate.

Objective

The objective of this study is to provide an economic evaluation of the effect of two alternative cyclone/CXF/HGC system concepts on a 500 MW steam-cooled reference PFBC combined cycle

power plant system burning eastern coal and utilizing a three-cyclone/HGC system requiring a baghouse filter.

Scope

The steam-cooled PFBC combined cycle power plant system analyzed in this study is based on an existing conceptual design study, modified to suit the requirements and format of the criteria document.

Estimates are derived for the power plant systems listed below for plant capital costs, operation and maintenance costs, levelized busbar power costs, plant heat rates, raw material consumption rates, and spent solids production rates.

The following plant concepts are evaluated:

- PFBC plant utilizing three cyclonic stages and a baghouse
- PFBC plant utilizing two cyclonic stages and a CXF stage
- PFBC plant utilizing one cyclonic stage and a CXF stage

The analyses employ the EPRI economic premises and bases as presented in "EPRI Economic Premises for Electric Power Generating Plants" (5/19/80). Costs are reported in January 1980 dollars. The plant location is taken to be mid-continental U.S.A. Where detailed cost estimates are not already available, Class I (simplified) cost estimates are derived. The power plants considered in the study consist of a nominal 500 MW unit, excluding electrical switchyard facilities, and are designed to meet the following June 1979 NSPS:

Emission Limits

SO ₂	90% removal for eastern bituminous
NO _x as NO ₂	0.6 lbs/10 ⁶ Btu for eastern bituminous
Total Particulate	0.03 lbs/10 ⁶ Btu

Report Format

This report is presented in the following format:

- Summary of the results of the economic evaluations (Section 1) of the reference, 1-cyclone/CXF/HGC and 2-cyclone/CXF/HGC PFBC plants burning eastern coal.
- Economic and design criteria (Section 2) which form the basis of the analytical procedure.
- Conceptual plant descriptions, based on available existing studies, for the steam-cooled reference (Section 3), 2-cyclone/CXF/HGC (Section 4) and 1-cyclone/CXF/HGC (Section 5). Sections 3, 4 and 5 include the economic evaluation of their respective PFBC plants for eastern coal operation.
- Description of the CXF module and CXF unit, including principles of operation and attendant cost estimates are presented in Appendix A.

Summary of Results

Exhibit 1-1 presents a summary of the operating and design parameters of the nominal 500 MW steam-cooled, reference, 2-cyclone/CXF/HGC and 1-cyclone/CXF/HGC PFBC plants considered in this study. The resulting net plant heat rate for the PFBC plants considered are as follows:

EXHIBIT 1-1

OPERATING AND DESIGN PARAMETERS SUMMARY FOR NOMINAL 500 MW
STEAM-COOLED PFBC POWER PLANTS
WITH A Ca/S = 1.3

	REFERENCE PFBC PLANT	2-CYCLONE/CXF/HGC PFBC PLANT	1-CYCLONE/CXF/HGC PFBC PLANT
POWER OUTPUT			
Gross Steam Turbine Output, kW	421,000	421,000	421,000
Gross Gas Turbine Output, kW	123,000	123,000	123,000
Gross Output, kW	544,000	544,000	544,000
Auxiliary Power, kW	13,000	10,415	7,000
Net Plant Output, kW	531,000	533,585	536,300
Net Plant Heat Rate, Btu/kWh	8,684	8,642	8,598
PFBC COMBUSTOR			
Number	4	4	4
Type	Cylindrical, 2-Bed Horizontal	Cylindrical, 2-Bed Horizontal	Cylindrical, 2-Bed Horizontal
Total Air Flow, 10 ⁶ lb/h	5.25	5.25	5.25
Air Inlet Temperature, °F	544	544	544
Air Outlet Temperature, °F	1636	1636	1636
PFBC Operating Pressure, psig	98	98	98
Bed Depth, ft	9	9	9
Bed Temperature, °F	1686	1686	1686
Combustor Size (Evaporator Duty)	36' Dia. x 100' L	36' Dia. x 100' L	36' Dia. x 100' L
Combustor Size (SH/RH Duty)	36' Dia. x 115' L	36' Dia. x 115' L	36' Dia. x 115' L
Hot Gas Cleanup System Type	3-Stage Cyclone/Baghouse	2-Stage Cyclone/CXF	1-Stage Cyclone/CXF
Superficial Velocity, ft/sec	4.5	4.5	4.5
GAS TURBINE			
Number	2	2	2
Type	Commercially Available Utility Gas Turbine	Commercially Available Utility Gas Turbine	Commercially Available Utility Gas Turbine
Compressor Pressure Ratio	8	8	8
Total Compressor Air Inlet Flow, 10 ⁶ lb/h	5.47	5.47	5.47
Compressor Outlet Temperature, °F	544	544	544
Turbine Inlet Conditions, psig/°F	87.2/1600	87.2/1600	87.2/1600
Turbine Outlet Temperature, °F	923	923	923
Exhaust Gas Waste Heat Recovery	Feedwater Heating	Feedwater Heating	Feedwater Heating
STEAM CYCLE			
Main Steam Flow, 10 ⁶ lb/h	1.18	1.18	1.18
Main Steam Conditions, psig/°F	2400/1000	2400/1000	2400/1000
Reheat Steam Flow, 10 ⁶ lb/h	1.16	1.16	1.16
Reheat Steam Conditions, psig/°F	530/1000	530/1000	530/1000
Condenser Pressure, in Hg Abs	2.0	2.0	2.0
Feedwater Heating Stages	2-LP, D/A & Exhaust Gases	2-LP, D/A & Exhaust Gases	2-LP, D/A & Exhaust Gases
EMISSIONS			
Total Stack Flow, 10 ⁶ lb/h	5.83	5.83	5.83
Stack Gas Temperature, °F	280	280	280
Sulfur Removal, %	90	90	90
Particulate, lb/10 ⁶ Btu	0.004	Near 0.0	0.004
NO _x , lb/10 ⁶ Btu	0.25	0.25	0.25

<u>Plant Type</u>	<u>Net Plant Heat Rate (NPHR) Btu/kWh</u>
Steam-Cooled Reference PFBC	8684
Steam-Cooled 2-Cyclone/CXF/HGC PFBC	8642
Steam-Cooled 1-Cyclone/CXF/HGC PFBC	8598

The preceding indicate only about a 1% improvement in NPHR for the 1 cyclone/CXF/HGC system over the reference PFBC concept.

A summary of plant operating costs and busbar power costs, resulting from the economic evaluation of the nominal 500 MW steam-cooled PFBC plants developed for this study, all burning eastern coal, are presented in Exhibit 1-2. The resulting levelized busbar power costs for the plants considered are as follows:

<u>Plant Type</u>	<u>Levelized Busbar Power Cost mills/kWh</u>
Steam-Cooled Reference PFBC	70.14
Steam-Cooled 2-Cyclone/CXF/HGC PFBC	68.30
Steam-Cooled 1-Cyclone/CXF/HGC PFBC	67.02

This tabulation indicates that just the incorporation of the 1-cyclone/CXF/HGC system alone can yield a 4.4% decrease in the levelized busbar power cost over the reference PFBC plant concept.

EXHIBIT 1-2

CAPITAL INVESTMENT AND
REVENUE REQUIREMENTS SUMMARY FOR 500 MW POWER PLANTS
BURNING EASTERN COAL

PARAMETER	STEAM-COOLED REFERENCE PFBC POWER PLANT	STEAM-COOLED 1-CYCLONE/CXY/HQC PFBC POWER PLANT	STEAM-COOLED 1-CYCLONE/CXY/HQC PFBC POWER PLANT
<u>Criteria</u>			
SO ₂ Emission Std.	90%	90%	90%
Ca/S, mole ratio	1.1	1.1	1.1
<u>Operating Conditions</u>			
Gross Power Output, MW	544.000	544.000	544.000
Auxiliary Power, MW	13.000	10.415	7.700
Net Power Output, MW	531.000	533.585	536.300
Net Plant Heat Rate, Btu/kWh	8,684	8,642	8,598
<u>Materials/Consumption/Production</u>			
Coal Consumption, ton/h	228.27	228.27	228.27
Dolomite Consumption, ton/h	67.71	67.71	67.71
Solid Wastes, wet, ton/h	0.0	0.0	0.0
Solid Wastes, dry, ton/h	91.90	91.90	91.90
Water Consumption, GPM	4,550	4,550	4,550
<u>Capital Investment (January 1980 dollars)</u>			
Capital Cost, \$ millions	552.1	524.9	507.8
Capital Cost, \$/net kW	1,040	984	947
10-Yr Levelized Fixed Charges, \$/kW-yr	187.2	177.1	170.4
<u>10-Yr Levelized Costs (1980-2009)</u>			
Dolomite, mills/kWh	2.64	2.64	2.63
Waste Disposal, mills/kWh	1.47	1.46	1.45
Water, mills/kWh	0.44	0.43	0.43
Other Consumables, mills/kWh	0.28	0.0	0.0
Total Variable O&M, mills/kWh	4.83	4.53	4.51
Fixed O&M, mills/kWh	6.65	6.62	6.58
Fixed Charges, mills/kWh	30.33	28.88	27.80
Subtotal, mills/kWh	42.01	40.11	39.17
Fuel Cost, mills/kWh	28.11	27.99	27.85
Busbar Power Cost, mills/kWh	70.14	68.10	67.02

Results are based on plant located in mid-continental U.S. area burning eastern coal with a HHV of 10,100 Btu/lb, 4% sulfur and a 70% capacity factor. All costs are based on January 1980 dollars.

Exhibit 1-3 presents a summary of total direct costs, presented by FERC account, and total capital requirements used in the economic analyses of each plant reported. The reference PFBC plant costs presented are developed from cost data available from existing studies. The 2-cyclone/CXF/HGC and 1-cyclone/CXF/HGC PFBC plant costs presented are based on modifying the previously developed reference PFBC plant costs to accommodate a CXF based HGC system. A summary of the cost modifications applicable to each PFBC plant considered in the analyses is included in the various respective PFBC plant description sections. The costs of the various modifications are included in the appropriate FERC account shown in Exhibit 1-3. The major modifications to the PFBC reference plant required for the implementation of the 2-cyclone/CXF/HGC system, aside from the installation of the requisite number of CXF units and a blow-back system, includes the attendant deletion of the tertiary stage multicyclone and the baghouse filter, a slight reduction in support structure, and a slight reduction in hot gas piping. The major modifications to the reference PFBC plant required for the implementation of the 1-cyclone/CXF/HGC system, aside from the installation of the requisite number of CXF units and a blow-back system, similar to, but less costly than the 2-cyclone/CXF/HGC blow-back system, includes the attendant deletion of both the secondary and tertiary multicyclones and baghouse filter, a modest reduction in support structures and foundation, and a significant reduction in hot gas piping. The resulting total capital required for the various PFBC plants and the comparable reference PCF/FGD plant are as follows:

<u>Plant Type</u>	<u>Total Capital Requirement \$/kW net</u>
Steam-Cooled Reference PFBC	1040
Steam-Cooled 2-Cyclone/CXF/HGC PFBC	984
Steam-Cooled 1-Cyclone/CXF/HGC PFBC	947

EXHIBIT 1-3

TOTAL CAPITAL REQUIREMENTS SUMMARY FOR NOMINAL 500 MW POWER PLANTS (January 1980 \$1000)

FERC Acct	Description	STEAM-COOLED REFERENCE PFBC POWER PLANT	STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT	STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT
310	Land and Land Rights	\$ 2,420	\$ 2,420	\$ 2,420
311	Structures and Improvements	39,626	39,511	38,807
312	Boiler Plant Equipment	211,668	192,543	181,930
314	Turbogenerator Units	90,100	90,100	90,100
315	Accessory Electrical Equipment	25,140	25,140	25,140
316	Misc. Power Plant Equipment	3,300	3,300	3,300
353	Station Equipment	3,400	3,400	3,400
	Total Direct Costs (Excluding Land)	373,234	353,994	342,777
	Undistributed Costs	22,394	21,240	20,560
	Process Capital	355,628	375,234	363,337
	General Facilities	0	0	0
	Engineering and Home Office Fees	37,323	35,399	34,688
	Subtotal	432,951	410,633	397,905
	Project Contingency	64,943	61,595	59,626
	Process Contingency	12,637	12,452	11,402
	Sales Tax	0	0	0
	Total Plant Investment	510,531	484,680	468,533
	Total Plant Investment at Startup	530,163	503,118	486,550
	Royalty Allowance	0	0	0
	Preproduction Costs	13,407	12,890	12,567
	Inventory Capital	6,293	6,293	6,293
	Initial Catalyst and Chemicals Charge	0	0	0
	Land	2,420	2,420	2,420
	Total Capital Requirements	552,283	524,921	507,830
	Total Capital Required, \$/kW net	1,040	984	947
	Net Plant Output, MWe	531	534	536

Note: Costs are based on January 1980 startup for a nominal 500 MWe plants located in a mid-continental U.S. area burning eastern coal with a HHV of 10,100 Btu/lb and 4% sulfur.

These computations indicate that the incorporation of a 1 cyclone/CXF/HGC system in a PFBC system has the potential to reduce the total capital requirement by about 9% over an all cyclone/baghouse reference PFBC plant.

Exhibit 1-4 presents a cost distribution summary for FERC Account 312, showing the cost contribution of the major elements of the HGC system, and the balance of boiler plant (combustor, and coal, dolomite and waste handling, etc.) and balance of plant costs for each PFBC plant. These results indicate that the CXF/HGC systems have a dramatic effect on the specific HGC system cost. The 2-cyclone and 1-cyclone-based CXF systems offer 29% and 45% reductions in the specific HGC system cost (i.e., \$89 and 69/kW net vs. \$125/kW net for the reference PFBC plant).

For the steam-cooled reference PFBC plant, the HGC system accounts for about 31% of boiler plant cost and about 18% of total plant costs. The cost of the reference PFBC plant HGC system is not an unreasonable fraction of boiler plant and total direct cost, being comparable to, but somewhat higher than, the corresponding system cost fractions for a typical PCF/FGD plant, which are about 26% of boiler plant cost and about 16% of total direct costs. However, the reference PFBC plant specific HGC system cost, the ratio of total HGC system direct capital cost to net plant output, estimated to be \$125/kW net, is about 37% higher than a typical PCF/FGD plant specific HGC system cost, which is estimated to be \$91/kW net, due to the higher boiler plant capital cost requirements associated with the reference PFBC plant.

For the steam-cooled 2-cyclone/CXF/HGC PFBC plant, the HGC system accounts for about 25% of boiler plant cost and about 13% of total direct costs and results in a specific HGC system cost estimated to be \$89/kW net. The cost parameters, fraction of boiler plant and total direct cost, and specific HGC system cost,

EXHIBIT 1-4

SUMMARY OF HGC SYSTEM DIRECT CAPITAL COST
FOR NOMINAL 500 MW PFBC POWER PLANTS
(January 1980 \$1000)

	<u>Reference PFBC Plant</u>	<u>2-Cyclone/CXF/HGC PFBC Plant</u>	<u>1-Cyclone/CXF/HGC PFBC Plant</u>
Primary Multicyclones	9,800	9,800	9,800
Secondary Multicyclones	7,900	7,900	-
Tertiary Multicyclones	6,920	-	-
Baghouse Filter	17,700	-	-
CXF Units	-	5,095	5,095
Blow-Back System	-	422	313
Refractory Lined Piping	19,026	19,004	16,519
Metal Lined Piping	<u>5,242</u>	<u>5,242</u>	<u>5,123</u>
TOTAL HGC SYSTEM	66,588	47,463	36,850
Balance of Boiler Plant	<u>145,080</u>	<u>145,080</u>	<u>145,080</u>
TOTAL FERC ACCOUNT 312	211,668	192,543	181,930
Balance of Plant	<u>161,566</u>	<u>161,451</u>	<u>160,747</u>
TOTAL DIRECT COST (Excluding Land)	373,234	353,994	342,677
Specific HGC System Cost, \$/kW net	125	89	69

for the 2-cyclone/CXF/HGC PFBC plant present a significant improvement over the corresponding cost parameters for the reference PFBC plant. The 2-cyclone/CXF/HGC system cost parameters, are comparable to, and slightly lower than, the corresponding HGC system costs associated with the typical PCF/FGD plant.

For the steam-cooled 1-cyclone/CXF/HGC PFBC plant, the HGC system accounts for about 20% of boiler plant costs and about 11% of total plant direct costs and results in a specific HGC system cost estimated to be \$69/kW net. The cost parameters for the 1-cyclone/CXF/HGC PFBC plant present a further improvement, over the reference PFBC plant's corresponding cost parameters, than those cost improvements projected for the 2-cyclone/CXF/HGC plant. The 1-cyclone/CXF/HGC system cost parameters are also significantly lower than those corresponding a typical PCF/FGD plant.

The CXF module and CXF unit cost estimates, reported in Appendix A, were developed based on data provided by WRDC for the preliminary design of a 16 ft. long module, comprised of 40 ceramic cross-flow filter elements, having 10 to 12 ft² of surface per filter element, and a nominal 16 ft. diameter by 34 ft. long shop fabricated CXF unit, housing 18 modules.

A review of cost development details, presented in Appendix A, indicates the following:

- Approximately 34% of the cost of a CXF module is attributable to the cost of filtration elements
- Approximately 36% of the cost of a CXF unit is attributable to the cost of CXF modules
- The specific CXF/HGC cost, the ratio of CXF/HGC system cost to total installed filter element surface area, is estimated to be in the range of 88 to 108 dollars per ft²

System power losses and/or the reduction in plant output due to the pressure drop across the CXF units, and the power required for the blow-back systems are estimated to be the following:

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<u>Steam-Cooled PFBC Plant</u>	<u>CXF Unit @ 0.75 psi</u>	<u>Blow-Back System Heaters</u>	<u>Blow-Back System Compressors</u>	<u>Total</u>
2-Cyclone/CXF/HGC, kW	1680	45	180	1905
1-Cyclone/CXF/HGC, kW	1660	25	65	1750

A complete tabulation of estimated PFBC plant auxiliary power requirement, including system losses expressed as auxiliary power (for convenience), are presented in the respective plant descriptions sections.

Conclusions

The results presented in this report are intended to provide a preliminary estimate of plant capital, operating, and busbar power costs, and attendant HGC system related costs for the steam-cooled reference PFBC combined cycle concept and two adaptations employing the first generation 2-cyclone/CXF/HGC and 1-cyclone/CXF/HGC systems considered in this study. A review of the economic evaluation indicates that the implementation of a CXF based HGC system, in the 500 MW steam-cooled reference PFBC combined cycle plant dramatically reduces plant complexity, auxiliary power, total capital requirement and levelized busbar power cost.

Of the two CXF based HGC system options investigated, the 1-cyclone/CXF/HGC system is superior to the 2-cyclone/CXF/HGC system in the areas of reduced plant complexity, auxiliary power, capital required and busbar power cost. These advantages not only result from the direct savings brought about by the deletion of the second stage multicyclones, but also by the attendant increase in allowable filtration cleaning cycle period. The gas cleaning cycle duration advantage of the 1-cyclone/CXF/HGC PFBC plant results from the enhanced performance of the ceramic cross-flow filter elements. The increased concentration of coarser particles in the particulate laden gas stream, resulting from the deletion of second stage multicyclones, tends to delay

the onset of filter plugging by the finer particulate, as the deposited coarser particulate itself tends to act somewhat as a filter. The net result is an increase in filtration capacity for the 1-cyclone/HGC system, as compared to the 2-cyclone/HGC system.

Although beyond the scope of the economic evaluation conducted, it is expected that for a nominal 250 MW steam-cooled PFBC plant utilizing a CXF based HGC system, both capital cost required and levelized busbar power cost would be reduced below the corresponding cost levels of a 250 MW PCF/FGD plant, due to the basic modularity of the reference PFBC plant design.

Section 2

ECONOMIC AND DESIGN CRITERIA

Plant Location

The plant site is located in the mid-continental U.S.A. with available raw makeup and condenser cooling water, utilities and ground transportation.

Plant Size

The PFBC plant, sized to utilize existing commercial gas turbine modules, consists of a single nominal 500 MWe unit.

Steam Conditions

The PCF thermal cycle has seven feedwater heaters and turbine throttle steam conditions of 2400 psig and 1000^oF, and reheat to 1000^oF.

The PCF turbine generator is guaranteed for a 568,215 kW gross output at 2.4 inches Hg back pressure and 0 percent makeup. The maximum capability of this turbine generator is 620,180 kW with valves wide open and 5 percent overpressure at the throttle. The turbine is a tandem compound four flow machine with 30 inch last stage blade length and a speed of 3600 rpm.

Plant Life

The plant life is 30 years.

Plant Construction Startup and Completion Dates

The plant construction period is 5 years, with construction startup on January 1, 1975 and construction completion on December 31, 1979. Plant operation begins on January 1, 1980.

Other Plant Factors

The reference PFBC plant utilizes an all cyclone HGC system and a baghouse filter, while alternative PFBC plants utilize a combination of cyclone/CXF/HGC systems.

Coal Source and Analysis

The analysis of the coal selected for this study is presented in Exhibit 2-1. The eastern bituminous coal analysis is from the Illinois reserves. The coal ash analysis is presented in Exhibit 2-2.

Limestone and Dolomite Analyses

The analyses of the limestone and dolomite sorbents selected for mass and energy balances and equipment sizing used in this study are presented in Exhibit 2-3.

Environmental Standards

The environmental standards correspond to the June 1979 New Source Performance Standards:

	<u>Emission Limits</u>
SO ₂	90% removal for eastern bituminous
NO _x as NO ₂	0.6 lbs/10 ⁶ Btu for eastern bituminous
Total Particulate	0.03 lbs/10 ⁶ Btu

EXHIBIT 2-1
COAL ANALYSES

Proximate Analysis, %

Moisture (as received)	12.0
Volatile Matter	33.0
Fixed Carbon	39.0
Ash	<u>16.0</u>
	100.0

Ultimate Analysis, %

Ash	16.0
Sulfur	4.0
Hydrogen	3.7
Carbon	57.5
Nitrogen	.9
Chlorine	.1
Oxygen	5.8
Moisture	<u>12.0</u>
	100.0
Higher Heating Value, Btu/lb	10,100

EXHIBIT 2-2
COAL ASH ANALYSES

<u>Constituent</u>		
Silica	- SiO_2	45.0%
Ferric Oxide	- Fe_2O_3	20.0
Alumina	- Al_2O_3	18.0
Titania	- TiO_2	1.0
Calcium Oxide	- CaO	7.0
Magnesium Oxide	- MgO	1.0
Sulphur Trioxide	- SO_3	3.5
Potassium Oxide	- K_2O	1.9
Sodium Oxide	- Na_2O	0.6
Phos. Pentoxide	- P_2O_5	0.2
Miscellaneous		<u>1.8</u>
		100.0%

EXHIBIT 2-3
DESIGN LIMESTONE AND DOLOMITE ANALYSES

	<u>Limestone</u>	<u>Dolomite</u>
CaCO_3	90.25%	53.9%
MgCO_3	-	41.4%
H_2O	5.0 %	1.0%
SiO_2 + Inerts	<u>4.75%</u>	<u>3.7%</u>
Total	100.0 %	100.0%

Economic Analysis

The economic analysis is based on the procedures, unit costs and escalation and inflation rates presented in the EPRI Economic Premises (Ref. 2-1).

Total Plant Investment

The determination of total plant investment (in January 1980 dollars) is illustrated below:

Process Capital	A
General Facilities	B
Engineering and Home Office Fees	C
Project Contingency	D
Process Contingency	E
Sales Tax	F
<hr/>	
Total Plant Investment	$TPI = A + B + C + D + E + F$

Process Capital

Process (Onsite) Capital Costs include both Total Direct Construction Costs and Undistributed Construction Costs.

Total Direct Construction Costs for all onsite processing units will be determined in January 1980 dollars.

Undistributed Construction Costs include the costs of onsite temporary facilities, construction equipment and construction services, taken to be 6% of the total direct construction costs.

The interface between the power generating plant and the distribution system will be at the high voltage bushings of the step-up transformers. Therefore, switchyard costs are not included in the plant capital cost estimate.

General Facilities

General Facilities (offsite) Costs such as roads, office buildings, shops, laboratories, chemical and by-product storage systems are included in the onsite capital costs. No allowances will be made for any off-site or general facilities beyond those included in the plant facilities.

Engineering and Home Office

Engineering design and construction management costs are taken to be 10% of total direct construction costs.

Project Contingency

A capital cost contingency factor is added. This is a project contingency factor that is intended to cover additional equipment that might result from a more detailed design of a definitive project at an actual site. The project contingency is 15 percent for all plant systems and components and is applied to items A, B and C above.

Process Contingency

This is an additional capital cost contingency applied to unproven technology in an effort to quantify the uncertainty in the design and cost of the commercial scale equipment. This contingency is used for comparing an unproven process with a commercially proven process, and is taken as 10%. This factor is applied only to the costs of the PFB combustor assembly, coal and dolomite preparation and injection system, and ash removal system, HGC system, and the PCF wet FGD system.

Sales Tax

Sales tax are included in items A and B.

Total Plant Investment at Startup

The term TPI* is defined as the total plant investment at startup and includes escalation and allowance for funds during construction. The allowance for funds during construction is based on compounding expenditures at the annual interest rate over the plant construction expenditure schedule.

The calculation of TPI* is based on the following conditions:

- a. Five year construction period
- b. Equal expenditures (E_i) each year, i.e.

$$E_1 = E_2 = E_3 = E_4 = E_5 = \frac{TPI}{5}$$
- c. Annual interest rate, i , equals 8%
- d. Annual escalation rate, e , equals 6%.

E_1	E_2	E_3	E_4	E_5	
					TPI*

$$TPI^* = \frac{E_1}{(1.06)^4} (1.08)^4 + \frac{E_2}{(1.06)^3} (1.08)^3 + \frac{E_3}{(1.06)^2} (1.08)^2 + \frac{E_4}{(1.06)} (1.08) + E_5$$

$$TPI^* = \frac{TPI}{5} \left(\frac{1.08}{1.06} \right)^4 + \frac{TPI}{5} \left(\frac{1.08}{1.06} \right)^3 + \frac{TPI}{5} \left(\frac{1.08}{1.06} \right)^2 + \frac{TPI}{5} \left(\frac{1.08}{1.06} \right) + \frac{TPI}{5}$$

$$TPI^* = TPI \left[\frac{1 + \left(\frac{1.08}{1.06} \right)^1 + \left(\frac{1.08}{1.06} \right)^2 + \left(\frac{1.08}{1.06} \right)^3 + \left(\frac{1.08}{1.06} \right)^4}{5} \right]$$

$$TPI^* = 1.0385 \text{ TPI}$$

Total Capital Requirement

The total capital requirement includes all capital necessary to complete the entire project. These items include:

Total Plant Investment at Startup	TPI*
Royalty Allowance	G
Preproduction Costs	H
Inventory Capital	I
Initial Catalyst and Chemicals Charge	J
Land	X
Total Capital Requirement	TCR = TPI* + G + H + I + J + X

Royalty Allowance

It is assumed that all royalties are paid by the equipment manufacturers and are included in the equipment cost.

Preproduction Costs

Preproduction costs include the following:

- a. One month's variable operation costs at full capacity excluding coal. Variable costs are for catalysts and other chemicals and utilities.
- b. One month's fixed costs. Fixed costs include operating and maintenance labor, administrative and support labor, and maintenance materials.
- c. 2% of total plant investment. (This charge covers expected changes and modifications to equipment that will be needed to bring the plant up to full capacity).
- d. 25% of full capacity fuel cost for one month. This charge covers inefficient operation that occurs during the startup period.

Inventory Capital

The value of inventories of fuel, other consumables and byproducts is capitalized and included in the inventory capital account. The inventory capital is estimated as follows:

- a. One month's supply of fuel based on full capacity operation.
- b. One month's supply of other consumables (excluding water) based on full capacity operation.

Initial Catalyst and Chemicals Charge

The initial cost of any catalyst or chemicals that are contained in the process equipment (but not in storage, which is covered in inventory capital) is to be included.

Land

The cost of land acquisition and land rights is assumed to be \$5,500/acre, in January 1980 dollars.

Operating Costs

Coal

The unit cost of the eastern coal delivered to the plant is \$31.31/ton (\$1.55/10⁶ Btu) in January 1980 dollars.

Limestone

The unit cost of the limestone delivered to the plant is \$11.00/ton, in January 1980 dollars.

Dolomite

The unit cost of the dolomite delivered to the plant is \$11.00/ton, in January 1980 dollars.

Waste Disposal

Unit disposal cost for dry, inert waste is \$4.50/ton, in January 1980 dollars. Unit disposal cost for wet sludge waste is \$9.40/ton, on a dry basis, in January 1980 dollars.

Untreated River Water

Untreated river water cost is \$0.45/1000 gallons, in January 1980 dollars.

Waste Water

Waste water streams are treated to meet discharge requirements with the water-treating equipment costs charged to the plant.

Sulfur Byproduct

No credit is taken for any byproduct sulfur produced.

Operating and Maintenance Labor

Annual operating and maintenance costs are estimated based on the personnel required per job category and rate consistent with the job classification. Rates will include both direct labor charge and payroll burden.

Administrative and Support Labor

Administrative and support labor costs are 30% of operating and maintenance labor costs.

General and Administrative Expense

General and administrative expenses are not included in annual revenue requirements.

Property Taxes and Insurance

Property taxes and insurance costs are 2.0% per year of total capital requirement and are included in the levelized fixed charge rate.

Product Cost Basis

Capacity Factor

The capacity factor is 70%, based on guaranteed turbine-generator output.

Operating Costs

Operating costs are presented on a 30-year levelized basis using 1980 to 2009 levelized prices. For costs other than fuel, first-year costs are multiplied by 1.886. This corresponds to 6 percent escalation and 10 percent cost of capital. Coal first year costs are multiplied by 2.09 (6.8 percent escalation, 10 percent cost of capital).

Cost of Capital

The fixed charges (income taxes, interest on debt, return on equity, depreciation and property taxes and insurance) are computed on a levelized basis with a 10% discount rate. The discount rate is based on the average cost of money. Using this basis, the fixed charges are 18.0% per year of the Total Capital Requirement. The investment factors that form the basis for the 18.0% per year capital charge are shown below:

Depreciation	Straight Line
Tax Life	20 years
Book Life	30 years
Debt/Equity Ratio	50/50
Bond Interest	8%/yr
Bond Life	30 years
Preferred Stock Ratio	15%
Preferred Stock Cost	8.5%/yr
Common Stock Ratio	35%
Common Stock Cost	13.5%/yr
Return on Equity After Taxes	12%/yr
Weighted Cost of Capital	10%/yr
Federal and State Income Tax Rate	50%
Property Tax and Insurance	2%
Escalation Rate	Not included
Investment Tax Credit	Not included
Iowa Type S ₁ Retirement Dispersion	

The fixed charge is based on the Total Capital Requirement with working capital treated the same as depreciable capital.

Power Cost Basis

The busbar power cost, in mills/kWh, is calculated using the 30-year levelized fixed charges, O&M charges, and fuel charges.

An example format is presented in Exhibit 2-4.

EXHIBIT 2-4

EXAMPLE FORMAT FOR CAPITAL INVESTMENT AND REVENUE REQUIREMENT SUMMARY TABLES

TITLE

PLANT CAPACITY, FUEL TYPE, ETC.

Capital Investment (Jan. 1980 \$ and Startup)

	<u>\$/kW</u>
Process Capital ^{a)}	A
General Facilities	B
Engineering and Home Office Fees	C
Project Contingency	D
Process Contingency	E
Sales Tax ^{b)}	F

Total Plant Investment in Jan. 1980 \$
(overnight construction)

$$TPI = A+B+C+D+E+F$$

Total Plant Investment for Jan. 1980
Startup

$$TPI^* = TPI \times (\text{Adj. Factor})^c)$$

Royalty Allowance	G
Preproduction Costs	H
Inventory Capital	I
Initial Catalyst and Chemicals Charge	J
Land	X

Total Capital Requirement (Jan. 1980
Startup)

$$TCR = TPI^*+G+H+I+J+X$$

Fixed Operating Costs (First Year)

	<u>\$/kW-yr</u>
Operating Labor	L
Maintenance Labor	M
Maintenance Materials	N
Administrative and Support Labor	P
Total Fixed O&M First Year	$\frac{P}{0.3} = 0.3 (L+M)$ $FOM = L+M+N+P$

a) A more detailed breakdown will be shown using FERC code of accounts.

b) Sales tax will be included in A and B.

c) For adjustment factor see page 2-8.

EXHIBIT 2-4 (Continued)

Variable Operating Cost Excluding Fuel
(First Year)

mills/kWh

Water

Q

Chemicals

R

Other Consumables

S

Waste Disposal

T

Variable Maintenance (if known)

U

Total Variable (excluding fuel, 1st year) VOM = Q+R+S+T+U

Byproduct Credits (First Year)

V

Fuel Cost (First Year)

W

30-Year Levelized O&M Costs

Equations

30-year Levelized Fixed O&M

LFOM = 1.886 x (FOM) \$/kW-yr

30-year Levelized Variable O&M (excl. fuel) LVOM = 1.886 x (VOM) mills/kWh

30-year Levelized Byproduct Credit

LB = Z* x V mills/kWh

30-year Levelized Fuel

LFU = Y* x W mills/kWh

30-Year Levelized Fixed Charges (Capital)

FC = 0.18 x (TCR) \$/kW-yr

30-Year Levelized Busbar Cost of Power at
Levelized-Capacity Factor (CF)

$$\text{Power Cost} = \frac{(\text{FC} + \text{LFOM})(1000 \text{ mills}/\$)}{(\text{CF}) \times (8760 \text{ hr/year})} + \text{LVOM} - \text{LB} + \text{LFU}$$

*Y is the 30-year Levelization Factor for fuel and Z is the 30-year Levelization Factor for byproducts.

Section 3

500 MW STEAM-COOLED REFERENCE PFBC COMBINED CYCLE POWER PLANT

SYSTEM DESCRIPTION

The nominal 500 MW steam-cooled reference PFBC combined cycle power plant concept burning eastern coal, presented in this study, is based on the conceptual design developed for the DAFFS (Design of Advanced Fossil Fuel Systems) study, which Argonne National Laboratories (ANL) conducted for DOE. The design was originally developed jointly by Westinghouse Electric Corporation/Foster Wheeler Development Co., under ANL Contract No. 31-109-38-6308 (Ref. 3-1 and 3-2), and Bechtel Group Inc., under ANL Contract No. 37-109-38-6212 (Ref. 3-3). This plant concept was modified by Burns and Roe, Inc. to comply with the economic and design criteria presented in Section 2 of this report.

The nominal 500 MW steam cooled reference PFBC combined cycle power plant simplified flow schematic is presented in Exhibit 3-1. The coal and dolomite handling and PFBC system preliminary general arrangement is shown in Exhibit 3-2, and the preliminary plant hot gas piping arrangement is shown in Exhibit 3-3. The plant comprises two PFBC/GT (gas turbine) modules, 123 MW total, supplying steam to a common subcritical steam turbine, rated at 421 MW, and includes a baghouse filter required to meet the current NSPS particulate emission standard. Each PFBC/GT module consists of a commercially available gas turbogenerator, a pair of dedicated complementary PFB combustor modules, designated as vessel A (saturation duty) and vessel B (superheat/reheat duty), and a three stage cyclonic HGC system. PFBC solid wastes generated by the plant are depressurized, cooled and stored in on-site waste storage silos prior to ultimate disposal off-site.

EXHIBIT 3-1

SIMPLIFIED FLOW SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED REFERENCE PFBC
COMBINED CYCLE POWER PLANT (REF. 3-3)

EXHIBIT 3-2

PRELIMINARY GENERAL ARRANGEMENT FOR A
 NOMINAL 500 MW STEAM-COOLED REFERENCE PFBC
 COMBINED CYCLE POWER PLANT (REF. 3-1)

Note: Dimensions of Vessels are Specific to Vessel A. Vessel B Arrangement will be Conceptually Similar

0 20
ft.

EXHIBIT 3-3

PRELIMINARY HOT GAS PIPING ARRANGEMENT FOR A
NOMINAL 500 MW STEAM-COOLED REFERENCE PFBC COMBINED CYCLE POWER PLANT (REF. 3-1)

Air is delivered to each pair of complementary PFB combustor modules by their dedicated gas turbine driven compressor. The capacity of the compressor is sufficient to maintain an air equivalence ratio of approximately 1.5 (i.e., 50% excess air) in each PFB combustor during operation at rated conditions. The high air equivalence ratio, in conjunction with bed operating conditions, insures almost complete carbon burnout (approximately 99% combustion efficiency), thereby negating the need for an ash recycle system. Coal and dolomite are fed into each PFB combustor in predetermined proportions to achieve the required degree (90%) of sulfur retention. Steam is generated in PFB combustor vessel A, superheated in vessel B, and reheated, after expansion through the steam turbine, in vessel B, through in-bed and water-wall type tubes. Flue gas from each complementary PFB combustor is ducted through the HGC system and routed to the common dedicated gas turbine. Gas turbine exhaust waste heat is recovered by the feedwater system through an economizer. Spent bed material waste heat recovery is accomplished in a fluidizing cooler appendage, located within each PFBC module, using a portion of the inlet combustion air stream as the coolant. The cooled spent bed material is then depressurized through a lockhopper arrangement and cooled to near ambient temperature by a circulating water cooled screw conveyor. The sensible heat in the fine spent solids collected by the three cyclone stages is utilized for feedwater heating, after depressurization, using helical screw type solids coolers.

Cyclone HGC System

The steam-cooled reference PFBC plant HGC system utilizes three stages of cyclonic particulate removal for gas turbine protection. Each PFB combustor bed is coupled to a first stage multicyclone module. Each first stage multicyclone module is coupled to a second stage module, which is in turn coupled to a third stage module. Exhibit 3-4 presents a schematic diagram of the multicyclone HGC module adopted for this study, and includes an accompanying tabulation of module

EXHIBIT 3-4

MULTICYCLONE MODULE SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED PFBC COMBINED CYCLE
POWER PLANT (REF. 3-1)

dimension and quantity of cyclones. Each multicyclone module comprises a large pressure vessel, internally lined with both an abrasive resistance layer and a thermal insulation layer, housing the requisite number of individual cyclones fabricated of a suitable high temperature service metal coated to minimize oxidation, corrosion and erosion. The individual cyclones are supported within the module housing by a reinforced gas seal plate, which separates the particulate laden inlet gas stream from the cleaner outlet gas stream. The seal plate and the module upper dome section forms the module upper plenum, where the individual cyclone outlet gas streams are combined. The interconnecting hot gas piping, between PFB combustor and multicyclone modules, and between multicyclone modules, is sized for a nominal gas velocity of 50 fps and is fabricated as a two component castable refractory lined carbon steel pipe. The outer layer of refractory consists of 3 in. of thermally insulating material, similar to H-W Lightweight Castable 20, and the inner layer consists of 3 in. abrasion/erosion resistant material, similar to H-W Harcast ER Castable. The hot gas piping between third stage multicyclone modules and turbine expander inlet is sized for a nominal velocity of 100 fps and is fabricated as a single component castable refractory lined carbon steel pipe, with a 6 in. thick layer of thermally insulating material, similar to H-W Lightweight Castable (Low Iron), and a $\frac{1}{4}$ in. thick Haynes 188 internal liner.

The level of particulate removal effected by the steam-cooled reference PFBC plant all cyclone HGC system, while adequate for gas turbine protection, is not sufficient to meet the current EPA particulate standard ($0.03 \text{ lbs}/10^6 \text{ Btu}$), thereby, necessitating the use of a baghouse prior to flue gas discharge to the stack.

Primary Cyclonic HGC Stage

The fine particulate laden effluent gas stream from each PFB combustor bed is ducted into a multicyclone module containing the requisite number of high efficiency tangential entry primary cyclones, where the relatively coarser particles are separated from the gas stream by centrifugal action. There are nine primary cyclones per 18 ft. diameter x 36 ft. high primary multicyclone module in each A vessel PFB combustor train, with each cyclone designed to operate with an inlet gas velocity of approximately 88 fps. Correspondingly, there are twelve cyclones per 20 ft. diameter x 36 ft. high primary module in each B train, with each cyclone designed to operate with an inlet gas velocity of approximately 86 fps. Each primary cyclone is about 4 ft. in diameter and operates with a pressure drop estimated to be about 0.7 psi. The inlet and outlet particulate loading from each primary cyclone is estimated to be about 17442 and 1233 ppm, respectively, based on scaling the performance data presented in reference 3-1 by the ratio of original to reference plant waste generation rates. The outlet gas stream from each primary cyclone is combined in the upper plenum section of the module. Particulate collected by each cyclone in the primary module is discharged to a common standpipe, via dipleg connections, depressurized, cooled and pneumatically transported to the fine waste storage silo.

Secondary Cyclonic HGC Stage

The effluent gas stream from each primary multicyclone module upper plenum is ducted into a secondary module, containing the requisite number of high efficiency tangential entry secondary cyclones, which forms the second stage of the HGC system. There are fourteen secondary cyclones per 18 ft. diameter x 24 ft high secondary multicyclone module in each A vessel PFB combustor train, with each cyclone designed to

operate with an inlet velocity of approximately 108 fps. Correspondingly, there are eighteen cyclones per 20 ft. diameter x 24 ft. high secondary module in each B train, with each cyclone designed to operate with an inlet velocity of approximately 102 fps. Each secondary cyclone is about 3 ft. in diameter and operates with a pressure drop estimated to be about 1.1 psi. The outlet particulate loading for each secondary cyclone is estimated to be about 529 ppm. based on scaling the performance data presented in reference 3-1. The outlet gas stream from each secondary cyclone is combined in the upper plenum section of the module. Particulate collected by each cyclone in the secondary module is combined, depressurized, cooled and pneumatically transported to the fine waste storage silo.

Tertiary Cyclonic HGC Stage

The effluent gas stream from each secondary multicyclone module upper plenum is ducted into a tertiary module, containing the requisite number of high efficiency tangential entry tertiary cyclones, which forms the third stage of the HGC system. There are 25 secondary cyclones per 18 ft. diameter x 16 ft. high tertiary multicyclone module in each A vessel PFB combustor train, with each cyclone designed to operate with an inlet velocity of approximately 127 fps. Correspondingly, there are 33 cyclones per 20 ft. diameter x 16 ft. high tertiary module in each B train, with each cyclone designed to operate with an inlet velocity of approximately 125 fps. Each tertiary cyclone is about 2 ft. in diameter and operates with a pressure drop estimated to be about 1.6 psi. The outlet particulate loading for each tertiary cyclone is estimated to be about 307 ppm, based on scaling the data presented in reference 3-1. The outlet gas stream from each tertiary cyclone is combined in the upper plenum section of the module. The outlet gas stream from each pair of tertiary multicyclone modules in each

PFB combustor train is combined into a single line and then routed to the gas turbine expander inlet. Particulate collected by each cyclone in the tertiary module is combined, depressurized, cooled and pneumatically transported to the fine waste storage silo.

Operating and Design Parameters

The complementary steam-cooled reference PFB combustor modules, vessels A and B, each consist of two beds housed in an insulated horizontal cylindrical pressure vessel, each approximately 36 ft. in diameter and 100 ft. and 115 ft. in length, respectively, designed to combust coal and dolomite in an environmentally acceptable manner. Each PFB combustor operates at a bed temperature of 1686°F and a bed pressure of 98 psig, with a superficial velocity of 4.5 fps and a bed depth of approximately 9 ft. Compressed air required for combustion and fluidization is supplied to the PFB combustor at 544°F by the gas turbine driven air compressor, which has an 8:1 compression ratio. The cleaned PFB combustor exhaust gas stream, at a temperature of 1600°F, is ducted into its dedicated gas turbine. The gas turbine drives the air compressor and also generates 61.5 MW of electricity. The exhaust from the gas turbine, at a temperature of 923°F, is directed into the economizer inlet header and used to pre-heat boiler feedwater. The steam generated in the PFB combustor is used to drive a subcritical steam turbine which has a gross output of 421 MW under plant operating conditions of 100% load and 2 in. Hg condenser back pressure. The steam turbine is supplied with main steam at 2400 psig and 1000°F and reheat stream at 1000°F. The steam-cooled reference PFBC plant operating and design parameters are shown in Exhibit 3-5 for the eastern coal investigated in this study.

Exhibit 3-6 presents the material balance schematic with all material streams expressed in pph.

Exhibit 3-7 presents the auxiliary power requirements.

EXHIBIT 3-5

OPERATING AND DESIGN PARAMETERS FOR A NOMINAL 500 MW
STEAM-COOLED REFERENCE PFBC POWER PLANT
WITH A Ca/S = 1.3

POWER OUTPUT

Gross Steam Turbine Output, kW	421,000
Gross Gas Turbine Output, kW	123,000
Gross Output, kW	544,000
Auxiliary Power, kW	13,000
Net Plant Output, kW	531,000
Net Plant Heat Rate, Btu/kWh	8,684

PFBC COMBUSTOR

Number	4
Type	Cylindrical, 2-Bed Horizontal
Total Air Flow, 10 ⁶ lb/h	5.25
Air Inlet Temperature, °F	544
Air Outlet Temperature, °F	1636
PFBC Operating Pressure, psig	98
Bed Depth, ft	9
Bed Temperature, °F	1686
Combustor Size (Evaporator Duty)	36' Dia. x 100' L
Combustor Size (SH/RH Duty)	36' Dia. x 115' L
Hot Gas Cleanup System Type	3-Stage Cyclone/Baghouse
Superficial Velocity, ft/sec	4.5

GAS TURBINE

Number	2
Type	Commercially Available Utility Gas Turbine
Compressor Pressure Ratio	8
Total Compressor Air Inlet Flow, 10 ⁶ lb/h	5.47
Compressor Outlet Temperature, °F	544
Turbine Inlet Conditions, psig/°F	87.2/1600
Turbine Outlet Temperature, °F	923
Exhaust Gas Waste Heat Recovery	Feedwater Heating

STEAM CYCLE

Main Steam Flow, 10 ⁶ lb/h	1.18
Main Steam Conditions, psig/°F	2400/1000
Reheat Steam Flow, 10 ⁶ lb/h	1.16
Reheat Steam Conditions, psig/°F	530/1000
Condenser Pressure, in Hg Abs	2.0
Feedwater Heating Stages	2-LP, D/A & Exhaust Gases

EMISSIONS

Total Stack Flow, 10 ⁶ lb/h	5.83
Stack Gas Temperature, °F	280
Sulfur Removal, %	90
Particulate, lb/10 ⁶ Btu	0.004
NO _x , lb/10 ⁶ Btu	0.25

EXHIBIT 3-6
MATERIAL BALANCE SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED REFERENCE PFBC COMBINED CYCLE POWER PLANT

EXHIBIT 3-7

AUXILIARY POWER REQUIREMENTS FOR A
NOMINAL 500 MW STEAM-COOLED REFERENCE
PFBC COMBINED CYCLE POWER PLANT (REF 3-3)

Coal and Dolomite Handling and Feed	1300
Waste Solids' Handling	600
Condensate Pumps	200
Circulating Water Pumps	3000
Service Water Pumps	200
Cooling Tower Fans	2100
Transformer Losses, Lighting & Miscellaneous	3000
Gas Turbine Auxiliary Load	100
Steam Turbine Auxiliary Load	400
PFB Related Auxiliary Equipment	1100
Miscellaneous	<u>300</u>
TOTAL	13000 kW

ECONOMIC ANALYSES

The economic evaluation of the nominal 500 MW steam-cooled reference PFBC combined cycle power plant, combusting eastern coal, was conducted in conformance with the criteria presented in Section 2.

The resulting capital cost estimates and total capital requirements for the 500 MW plant are presented in Exhibit 3-8 by FERC (Federal Energy Regulatory Commission) code of account number, expressed in January 1980 dollars. Exhibit 3-9 presents a summary of the modifications applied to the indicated original cost to arrive at the capital costs used in the economic analysis of the steam-cooled reference PFBC plant, utilizing a three stage cyclonic HGC system.

Process contingency costs are based on the cost of unproven equipment: PFBC boiler, ash removal systems, coal and dolomite Preparation and injection systems, and HGC cleanup system. The costs of these components are shown in Exhibit 3-10.

Annual operating and maintenance labor costs are based on the estimated requirements presented in Exhibit 3-11. These estimates are derived by modifying the reference PCF/FGD plant O&M cost data, presented in Exhibit 6-12, to accommodate the steam-cooled reference PFBC plant O&M requirements.

Estimated preproduction and inventory capital requirements are presented in Exhibits 3-12 and 3-13, respectively.

The resulting plant operating costs for the nominal 500 MW steam-cooled reference PFBC combined cycle plant are presented in Exhibit 3-14.

The capital and operating cost results for the nominal 500 MW steam-cooled reference PFBC plant, presented in Exhibits 3-8 and 3-14, respectively, indicate a total capital requirement

EXHIBIT 3-8

TOTAL CAPITAL REQUIREMENTS FOR A
NOMINAL 500 MW STEAM-COOLED REFERENCE
PFBC POWER PLANT
(January 1980 \$1000)

<u>FERC</u> <u>Acct</u>	<u>Description</u>	
310	Land and Land Rights	\$ 2,420
311	Structures and Improvements	39,626
312	Boiler Plant Equipment	211,668
314	Turbogenerator Units	90,100
315	Accessory Electrical Equipment	25,140
316	Misc. Power Plant Equipment	3,300
353	Station Equipment	3,400
	Total Direct Costs (Excluding Land)	<u>373,234</u>
	Undistributed Costs	22,394
	Process Capital	<u>395,628</u>
	General Facilities	0
	Engineering and Home Office Fees	37,323
	Subtotal	<u>432,951</u>
	Project Contingency	64,943
	Process Contingency	12,637
	Sales Tax	0
	Total Plant Investment	<u>510,531</u>
	Total Plant Investment at Startup	<u>530,163</u>
	Royalty Allowance	0
	Preproduction Costs	13,407
	Inventory Capital	6,293
	Initial Catalyst and Chemicals Charge	0
	Land	2,420
	Total Capital Requirements	<u>552,283</u>
	Total Capital Required, \$/kW net	1,040
	Net Plant Output, MWe	531

Note: Costs are based on January 1980 startup for a nominal 500 MW steam-cooled reference PFBC plant located in a mid-continental U.S. area. The eastern coal-fired plant burns coal with a HHV of 10,100 Btu/lb and 4% sulfur.

EXHIBIT 3-9

NOMINAL 500 MW STEAM-COOLED REFERENCE PFBC POWER PLANT
MODIFICATIONS TO ORIGINAL CAPITAL COSTS

FERC Account	Original Cost Estimate (Ref. 3-1) (<u>\$1000</u>)	Modification	Revised Cost Estimate (<u>January 1980 \$1000</u>)
310 Land	3,400	• Off-site disposal area of 260 acres deleted; plant site increased to 440 acres, from 350 acres, to accommodate limited on-site disposal	2,420
311 Structures and Improvements	58,590	• Based on \$6.3 million difference in FERC Account 311 for a PFBC plant and a PCF/FGD plant (Ref. 3-1 and 3-4, respectively)	39,626
312 Boiler Plant Equipment			
PFB Combustors	37,900		
PFB Combustor Solids Feeding	35,000		
PFB Combustor Solids Removal	4,500		
Primary Multicyclone	9,880		
Secondary Multicyclone	7,900		
Tertiary Multicyclone	6,920		
Multicyclone Solids Removal	1,000		
Refractory Lined Pipe	39,400	• Revised cost estimate	19,026
Refractory & Metal Lined Pipe	12,100	• Revised cost estimate	5,242
Baghouse Filter	17,700		
Balance of Account 312	<u>66,600</u>		
	238,900		211,668
314 Turbogenerator Units	90,100	• None	90,100
315 Accessory Electrical Equipment	25,140	• None	25,140
316 Miscellaneous Power Plant Equipment	3,300	• None	3,300
353 Station Equipment	<u>3,400</u>	• None	<u>3,400</u>
TOTAL (Excluding Land)	419,430		373,234

EXHIBIT 3-10

500 MW STEAM-COOLED REFERENCE PFBC POWER PLANT
COSTS OF ADVANCED TECHNOLOGY EQUIPMENT TO WHICH
PROCESS CONTINGENCY IS APPLIED
(January 1980 \$1000)

FERC ACCT 312

PrB Combustor/Boiler	37,900
PFB Spent Solids Removal System	4,500
Coal and Dolomite Preparation & Injection Systems	35,000
Hot Gas Cleanup System:	
1st Stage Multicyclones	9,880
2nd Stage Multicyclones	7,900
3rd Stage Multicyclones	6,920
Refractory Lined Large Hot Gas Piping	19,026
Metal Lined Large Hot Gas Piping	<u>5,242</u>
TOTAL	126,368

ESTIMATE OF ANNUAL OPERATION AND MAINTENANCE LABOR COST (January 1980 Dollars)
FOR A NOMINAL 500 MW STEAM-COOLED REFERENCE PFBC POWER PLANT

<u>Classification</u>	<u>Annual*</u> <u>Salary Per Capita</u>	<u>No. of Person.</u>	<u>Annual Salary</u>
<u>Operation</u>			
Operation Supervisor	39,600	1	39,600
Shift Supervisor	39,300	6	217,800
Control Room Operator	33,000	6	198,000
Asst. Control Room Operator	26,400	6	158,400
Turbine Operator	33,000	12	396,000
Aux. Equipment Operator	26,400	12	316,800
Boiler Operator	26,400	6	158,400
Coal and Ash Crew	24,200	10	242,000
Computer Specialist	28,000	1	28,000
Environmental Specialist	28,000	1	28,000
Results Engineer	26,400	1	26,400
Assistant Results Engineer	22,500	4	90,000
Chemical Supervisor	36,300	1	36,300
Chemists	26,400	4	105,600
		<u>71</u>	<u>2,041,300</u>
<u>Maintenance</u>			
Maintenance Supervisor	39,600	1	39,600
Mechanical Maint. Foreman	34,100	1	34,100
Mechanic	26,400	16	422,400
Mechanic's Helper	20,900	16	334,400
Machinist	26,400	2	52,800
Welder	26,400	3	79,200
Carpenter	24,200	2	48,400
Bricklayer	20,900	1	20,900
Electr. Maint. Foreman	34,100	1	34,100
Electrician	26,400	8	211,200
Electrician's Helper	20,900	8	167,200
I & C Foreman	34,100	1	34,100
I & C Repairman	28,600	5	143,000
Building Maint. Foreman	24,200	1	24,200
Plumber	20,900	1	20,900
Laborers	13,200	16	211,200
		<u>83</u>	<u>1,877,700</u>
		154	3,919,000
<u>Average Hourly Rates (\$/HR)</u>			
Operating			13.83
Maintenance			10.88
Total			12.23

*Annual salaries include both direct labor charges and payroll burden.

Notes: 1) Man-power estimates are based on a single unit 544 MWe steam-cooled PFBC power generating plant.

$$2) \text{ Average Hourly Rate} = \frac{\text{Total Annual Cost (\$)}}{\text{Number of Person.} \times 2080 \text{ working hrs/year}}$$

EXHIBIT 3-12

PREPRODUCTION COSTS FOR A NOMINAL 500 MW
STEAM-COOLED REFERENCE PFBC POWER PLANT

ITEM	QUANTITY	UNIT COST	CONVERSION FACTOR	COST ¹
Preproduction Cost				
Operating Labor	\$2,041,300	See Note 2	1/12	170,100
Maintenance Labor	\$1,877,700	See Note 2	1/12	156,500
Administrative Labor	\$1,175,700	See Note 3	1/12	98,000
Maintenance Material	\$6,384,200	See Note 4	1/12	532,000
Waste Disposal	91.90 TPH	\$ <u>4.50</u> Ton	730 h	301,900
Dolomite	67.71 TPH	\$ <u>11.00</u> Ton	730 h	543,700
Water	4550 GPM	\$ <u>0.45</u> 1000 Gal	730 h	89,700
Coal	228.27 TPH	\$ <u>31.31</u> Ton	<u>730</u> 4 h	1,304,400
2% TPI				<u>10,210,600</u>
Total Preproduction Cost				13,406,900

- Note:
1. All Costs are in January 1980 dollars
 2. Exhibit 3-11 presents annual operating and maintenance labor
 3. Administrative costs are 30% of O&M labor costs
 4. Maintenance materials taken as 3.4 x maintenance labor

EXHIBIT 3-13

INVENTORY CAPITAL FOR A NOMINAL 500 MW
STEAM-COOLED REFERENCE PFBC POWER PLANT

Inventory Capital Material	Quantity	Rate	Conversion Factor	Cost ¹
Coal	228.27 TPH	<u>\$31.31</u> Ton	730 h	5,217,400
Dolomite	67.71 TPH	<u>\$11.00</u> Ton	730 h	543,700
Maintenance Material				<u>532,000</u>
TOTAL				6,293,100

Note: 1. All costs are in January 1980 dollars

EXHIBIT 3-14

CAPITAL INVESTMENT AND
REVENUE REQUIREMENTS FOR A 500 MW
STEAM-COOLED REFERENCE PFBC POWER PLANT

PARAMETER	EASTERN COAL
<u>Criteria</u>	
SO ₂ Emission Std.	90%
Ca/S, mole ratio	1.3
<u>Operating Conditions</u>	
Gross Power Output, MW	544.000
Auxiliary Power, MW	13.000
Net Power Output, MW	531.000
Net Plant Heat Rate, Btu/kWh	8,684
<u>Materials/Consumption/Production</u>	
Coal Consumption, ton/h	228.27
Dolomite Consumption, ton/h	67.71
Solid Wastes, wet, ton/h	0.0
Solid Wastes, dry, ton/h	91.90
Water Consumption, GPM	4,550
<u>Capital Investment (January 1980 dollars)</u>	
Capital Cost, \$ millions	552.3
Capital Cost, \$/net kW	1,040
30-Yr Levelized Fixed Charges, \$/kW-yr	187.2
<u>30-Yr Levelized Costs (1980-2009)</u>	
Dolomite, mills/kWh	2.64
Waste Disposal, mills/kWh	1.47
Water, mills/kWh	0.44
Other Consumables, mills/kWh	0.28
Total Variable O&M, mills/kWh	4.83
Fixed O&M, mills/kWh	6.65
Fixed Charges, mills/kWh	30.53
Subtotal, mills/kWh	42.01
Fuel Cost, mills/kWh	28.13
Busbar Power Cost, mills/kWh	70.14

Results are based on plant located in mid-continental U.S. area burning eastern coal with a HHV of 10,100 Btu/lb, 4% sulfur and a 70% capacity factor. All costs are based on January 1980 dollars.

of \$1040/kW net and a busbar power cost of 70.14 mills/kWh. Compared to the reference PCF/FGD plant costs, presented in Exhibits 6-8 and 6-9, the steam-cooled reference PFBC plant is about 24.0% more expensive in total capital requirement and about 4.3% more expensive in busbar power cost.

Section 4

500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC COMBINED CYCLE POWER PLANT

The nominal 500 MW steam-cooled 2-cyclone/CXF/HGC/ PFBC combined cycle power plant concept burning eastern coal, presented in the study, is based on the reference conceptual plant design presented in Section 3. The reference plant design was modified by Burns and Roe, in the course of developing the 2-cyclone/CXF concept, to accommodate a Westinghouse R&D specified HGC system comprising 2 cyclonic stages and a CXF stage. The 2-cyclone/CXF/HGC system concept is designed to provide adequate gas turbine particulate protection and to achieve compliance with current EPA particulate ($0.03 \text{ lbs}/10^6 \text{ Btu}$) and opacity standards, and to offer the potential to achieve compliance with possible future inhaleable particulate standards and/or a more stringent opacity standard. The ceramic cross-flow filter is a near perfect filter with regard to submicron particulate.

The nominal 500 MW steam-cooled 2-cyclone/CXF/HGC PFBC combined cycle power plant simplified flow schematic is presented in Exhibit 4-1. The coal and dolomite handling and PFBC system preliminary general arrangement is shown in Exhibit 4-2, and the preliminary plant hot gas piping arrangement is shown in Exhibit 4-3. The plant comprises two PFBC/GT modules, 123 MW total, supplying steam to a common subcritical steam turbine, rated at 421 MW, and does not require a baghouse filter to achieve compliance with the current NSPS particulate emission standard. Each PFBC/GT module consists of a commercially available gas turbogenerator, a pair of dedicated complementary PFB combustor modules, designated as vessel A (saturation duty) and vessel B (superheat/reheat duty), and a HGC system comprising two cyclonic stages followed by a third CXF stage. PFBC solid wastes generated by the plant are depressurized, cooled and stored in on-site waste storage silos prior to ultimate disposal off-site.

EXHIBIT 4-1

SIMPLIFIED FLOW SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC
PFBC COMBINED CYCLE POWER PLANT

EXHIBIT 4-2

PRELIMINARY GENERAL ARRANGEMENT FOR A NOMINAL 500 MW
STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC COMBINED CYCLE POWER PLANT

Note: Dimensions of Vessels are Specific to Vessel A. Vessel B Arrangement will be Conceptually Similar

0 20
ft.

EXHIBIT 4-3

PRELIMINARY HOT GAS PIPING ARRANGEMENT FOR A NOMINAL 500 MW STEAM-COOLED
2-CYCLONE/CXF/HGC PFBC COMBINED CYCLE POWER PLANT

Air is delivered to each pair of complementary PFB combustor modules by their dedicated gas turbine driven compressor. The capacity of the compressor is sufficient to maintain an air equivalence ratio of approximately 1.5 (i.e., 50% excess air) in each PFB combustor during operation at rated conditions. The high air equivalence ratio, in conjunction with bed operating conditions, insures almost complete carbon burnout (approximately 99% combustion efficiency) thereby negating the need for an ash recycle system. Coal and dolomite are fed into each PFB combustor in predetermined proportions to achieve the required degree (90%) of sulfur retention. Steam is generated in PFB combustor vessel A, superheated in vessel B, and reheated, after expansion through the steam turbine, in vessel B, through in-bed and water-wall type tubes. Flue gas from each complementary PFB combustor is ducted through the 2-cyclone/CXF/HGC system and routed to the common dedicated gas turbine. Gas turbine exhaust waste heat is recovered by the feedwater system through an economizer. Spent bed material waste heat recovery is accomplished in a fluidizing cooler appendage, located within each PFBC module, using a portion of the inlet combustion air stream as the coolant. The cooled spent bed material is then depressurized through a lockhopper arrangement and cooled to near ambient temperature by a circulating water cooled screw conveyor. The sensible heat in the fine spent solids collected by the two cyclone stages and the CXF stage is utilized for feedwater heating, after depressurization, using helical screw type solid coolers.

2-Cyclone/CXF/HGC System

The steam-cooled 2-cyclone/CXF/HGC PFBC plant utilizes two stages of cyclonic and one stage of CXF particulate removal for gas turbine protection. Each PFB combustor bed is coupled to a first stage multicyclone module. Each first stage multicyclone module is coupled to a second stage module, which is in turn coupled to a third stage CXF unit.

Exhibit 4-4 presents a schematic diagram of the multicyclone HGC module adopted for this study, and includes an accompanying tabulation of module dimension and quantity of cyclones. Each multicyclone module comprises a large pressure vessel, internally lined with both an abrasive resistance layer and a thermal insulation layer, housing the requisite number of individual cyclones fabricated of a suitable high temperature service metal coated to minimize oxidation, corrosion and erosion. The individual cyclones are supported within the module housing by a reinforced gas seal plate, which separates the particulate laden inlet gas stream from the cleaner outlet gas stream. The seal plate and the module upper dome section forms the module upper plenum, where the individual cyclone outlet gas streams are combined.

Exhibit 4-5 presents a simplified diagram of the CXF module adopted for this study. Exhibit 4-6 presents a simplified diagram of the CXF unit, and includes an accompanying tabulation of unit dimensions and quantity of modules for both the A and B vessel PFB combustor HGC trains. Each CXF unit comprises a large pressure vessel, internally lined with a 1-inch abrasive resistance layer and a 5-inch thermal insulation layer, housing the requisite number of CXF modules. The individual CXF modules are suspended within the unit housing by support channels. The support channels also reinforce the gas seal plate, which separates the particulate laden inlet gas stream from the near particulate free outlet gas stream. The gas deflector plate, designed to eliminate the effects of unimpeded direct inlet gas impingement on the operation of the CXF modules adjacent to the inlet, is suspended from the reinforced seal plate. Both the channels and seal plate are supported by the plenum skirt attached to the inside of the dome of the CXF unit. The seal plate, skirt and upper dome section forms the CXF unit upper plenum, where the individual module outlet gas streams are combined. A detailed description of the CXF module and CXF unit is presented in Appendix A.

EXHIBIT 4-4

MULTICYCLONE MODULE SCHEMATIC FOR A
 NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC COMBINED CYCLE
 POWER PLANT

EXHIBIT 4-5
CERAMIC CROSS-FLOW FILTRATION MODULE

EXHIBIT 4-6

CERAMIC CROSS-FLOW FILTRATION UNIT

The interconnecting hot gas piping, between PFB combustor and multicyclone modules, between multicyclone modules, and between multicyclone modules and CXF units is sized for a nominal gas velocity of 50 fps and is fabricated as a two component castable refractory lined carbon steel pipe. The outer layer of refractory consists of 3 in. of a thermally insulating material, similar to H-W Lightweight Castable 20, and the inner layer consists of 3 in. of an abrasion/erosion resistant material, similar to H-W Harcast ER Castable. The hot gas piping between third stage CXF units and turbine expander inlet is sized for a nominal velocity of 100 fps and is fabricated as a single component castable refractory lined carbon steel pipe, with a 6 in. thick layer of thermally insulating material, similar to H-W Lightweight Castable (Low Iron), and a $\frac{1}{4}$ " thick Haynes 188 internal liner.

Primary Cyclonic HGC Stage

The fine particulate laden effluent gas stream from each PFB combustor bed is ducted into a multicyclone module containing the requisite number of high efficiency tangential entry primary cyclones, where the relatively coarser particles are separated from the gas stream by centrifugal action. There are nine primary cyclones per 18 ft. diameter x 36 ft. high primary multicyclone module in each A vessel PFB combustor train, with each cyclone designed to operate with an inlet gas velocity of approximately 88 fps. Correspondingly, there are twelve cyclones per 20 ft. diameter x 36 ft. high primary module in each B train, with each cyclone designed to operate with an inlet gas velocity of approximately 86 fps. Each primary cyclone is about 4 ft. in diameter and operates with a pressure drop estimated to be about 0.7 psi. As previously discussed in Section 3, the inlet and outlet particulate loading from each primary cyclone is estimated to be about 17,442 and 1233 ppm, respectively, based on scaling the performance data presented in reference 3-1. The outlet gas stream from each primary cyclone is combined in the upper plenum section of the module. Particulate collected by each cyclone in the primary module is discharged to a common standpipe, via dipleg connections,

depressurized, cooled and pneumatically transported to the fine waste storage silo.

Secondary Cyclonic HGC Stage

The effluent gas stream from each primary multicyclone module upper plenum is ducted into a secondary module, containing the requisite number of high efficiency tangential entry secondary cyclones, which forms the second stage of the HGC system. There are fourteen secondary cyclones per 18 ft. diameter x 24 ft. high secondary multicyclone module in each A vessel PFB combustor train, with each cyclone designed to operate with an inlet velocity of approximately 108 fps. Correspondingly, there are eighteen cyclones per 20 ft. diameter x 24 ft. high secondary module in each B train, with each cyclone designed to operate with an inlet velocity of approximately 102 fps. Each secondary cyclone is about 3 feet in diameter and operates with a pressure drop estimated to be about 1.1 psi. As previously discussed in Section 3, the outlet particulate loading for each secondary cyclone is estimated to be about 529 ppm. The outlet gas stream from each secondary cyclone is combined in the upper plenum section of the module. Particulate collected by each cyclone in the secondary module is combined, depressurized, cooled and pneumatically transported to the fine waste storage silo.

Tertiary Ceramic Cross-Flow Filtration HGC Stage

The effluent gas stream from each secondary multicyclone module upper plenum is ducted into a tertiary CXF unit, containing the requisite number of high efficiency CXF modules, which forms the third stage of the HGC system. There are fourteen CXF modules per 13.5 ft. diameter x 31.5 ft. high CXF unit in each A vessel PFB combustor train. Correspondingly, there are eighteen CXF modules per 16 ft. diameter x 33.5 ft. high CXF unit in each B train. Each CXF module is about 16 ft. in height and each CXF unit operates with an average pressure drop estimated to be about 0.75 psi. The outlet particulate loading for each

?

CXF unit is estimated to be about 0 ppm. The outlet gas stream from each CXF module is combined in the upper plenum section of the CXF unit. The outlet gas stream from each pair of CXF units in each PFBC train is combined into a single line and then routed to the gas turbine expander inlet. Particulate collected by each module in a tertiary stage CXF unit is collected during the periodic on line filter cleaning (backwashing) phase of the gas cleaning cycle. The collected particulate is discharged through the bottom of the CXF unit and depressurized through a lockhopper arrangement, cooled and pneumatically transported to the fine waste storage silo.

Blow-Back Air System

The blow-back air system is required to provide the high temperature, high pressure air necessary for ceramic cross-flow filter backwashing. The system consists of a bank of 3 motor driven reciprocating compressors, located in the plant compressor building, delivering high pressure air, via the interconnecting piping, to each dedicated compressed air accumulator located above each CXF unit. Each compressor is sized to deliver one-half of the plant's blow-back air requirement based on each CXF module operating on a 35 minute gas filtration cycle, requiring 44.1 pps of air at 45 atmospheres and 500°F, for a duration of 0.1 seconds. The combination of 3 compressors, each equipped with reed valves will enable the blow-back system to supply backwash air during partial load plant operation, provide for a standby compressor during full load plant operation, and provide additional compressed air capacity for possible startup or maintenance considerations. The compressors are designed to deliver air at 45 atmospheres pressure and near ambient temperature to each accumulator. Each accumulator is designed as a thermally insulated horizontal cylindrical tank, about 2 ft. in diameter x 12 ft. long, having an internal capacity of about 35 cubic feet. The air inlet to each accumulator is equipped with a Chromalox type in-line electric air heater, housed in a nominally ½ ft. diameter x 4 ft. long, 316 stainless steel tube. The inlet air heater is equipped with an SCR type controller and is designed to heat up the incoming compressed air to 500°F. In order to offset the ambient heat loss

and maintain the air in the accumulator at 500°F, an electric heater cable is installed around the cylindrical section of the accumulator, between the outer wall and the 3 in. thermal insulation blanket. The operation of the heater cable is controlled by a thermostatic type controller. The accumulator outlet line is fabricated as a 3 in. 316 stainless steel manifold, routed around the top of the CXF unit and connected to each pulse air pipe solenoid valve. During the backwashing phase of the gas filtration cycle, the solenoid valves are operated in sequence to permit hot compressed air to be admitted to the pulse air line of each CXF module in its turn.

Operating and Design Parameters

The complementary steam-cooled PFB combustor modules, vessels A and B, each consist of two beds housed in an insulated horizontal cylindrical pressure vessel, each approximately 36 ft. in diameter and 100 ft. and 115 ft. in length, respectively, designed to combust coal and dolomite in an environmentally acceptable manner. Each PFB combustor operates at a bed temperature of 1686°F and a bed pressure of 98 psig, with a superficial velocity of 4.5 fps and a bed depth of approximately 9 ft. Compressed air required for combustion and fluidization is supplied to the PFB combustor at 544°F by the gas turbine driven air compressor, which has an 8:1 compression ratio. The cleaned PFB combustor exhaust gas stream, at a temperature of 1600°F, is ducted into its dedicated gas turbine. The gas turbine drives the air compressor and also generates 61.5 MW of electricity. The exhaust from the gas turbine, at a temperature of 923°F, is directed into the economizer inlet header and used to pre-heat boiler feedwater. The steam generated in the PFB combustor is used to drive a subcritical steam turbine which has a gross output of 421 MW under plant operating conditions of 100% load and 2 in. Hg condenser back pressure. The steam turbine is

supplied with main steam at 2400 psig and 1000°F and reheat stream at 1000°F. The steam-cooled 2-cyclone/CXF/HGC PFBC plant operating and design parameters are shown in Exhibit 4-7 for the eastern coal investigated in this study.

Exhibit 4-8 presents the material balance schematic, with all material streams expressed in pph.

Exhibit 4-9 presents the auxiliary power requirements. System power losses incurred due to the implementation of the specified 2-cyclone/CXF/HGC system and the attendant deletion of the third stage multicyclones and baghouse filter from the reference PFBC plant concept, are reported, for the sake of convenience, as auxiliary power requirements, rather than an increase or decrease in gas and/or steam turbogenerator output. The auxiliary power requirements shown under "Equipment Modification" include the increase or decrease in power requirement resulting from the addition or deletion of the equipment indicated. The requirements shown under "Operational Modifications" include the increase or decrease in system power losses attributable to flow through the HGC system equipment indicated, calculated as the product of pressure drop and average volumetric gas flow rate.

ECONOMIC ANALYSES

The economic evaluation of the nominal 500 MW steam-cooled 2-cyclone/CXF/HGC PFBC combined cycle power plant, combusting eastern coal, was conducted in conformance with the criteria presented in Section 2.

The resulting capital cost estimates and total capital requirements for the 500 MW plant are presented in Exhibit 4-10 by FERC (Federal Energy Regulatory Commission) code of account number, expressed in January 1980 dollars. Exhibit 4-11 presents a summary of the modifications applied to the indicated cost to arrive at the capital costs used in the economic analysis of the steam-cooled PFBC plant, utilizing a 2-cyclone/CXF/HGC system.

EXHIBIT 4-7

OPERATING AND DESIGN PARAMETERS FOR A NOMINAL 500 MW
STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT
WITH A Ca/S = 1.3

POWER OUTPUT

Gross Steam Turbine Output, kW	421,000
Gross Gas Turbine Output, kW	123,000
Gross Output, kW	544,000
Auxiliary Power, kW	10,415
Net Plant Output, kW	533,585
Net Plant Heat Rate, Btu/kWh	8,642

PFBC COMBUSTOR

Number	4
Type	Cylindrical, 2-Bed Horizontal
Total Air Flow, 10 ⁶ lb/h	5.25
Air Inlet Temperature, °F	544
Air Outlet Temperature, °F	1636
PFBC Operating Pressure, psig	98
Bed Depth, ft	9
Bed Temperature, °F	1686
Combustor Size (Evaporator Duty)	36' Dia. x 100' L
Combustor Size (SH/RH Duty)	36' Dia. x 115' L
Hot Gas Cleanup System Type	2-Stage Cyclone/CXF
Superficial Velocity, ft/sec	4.5

GAS TURBINE

Number	2
Type	Commercially Available Utility Gas Turbine
Compressor Pressure Ratio	8
Total Compressor Air Inlet Flow, 10 ⁶ lb/h	5.47
Compressor Outlet Temperature, °F	544
Turbine Inlet Conditions, psig/°F	87.2/1600
Turbine Outlet Temperature, °F	923
Exhaust Gas Waste Heat Recovery	Feedwater Heating

STEAM CYCLE

Main Steam Flow, 10 ⁶ lb/h	1.18
Main Steam Conditions, psig/°F	2400/1000
Reheat Steam Flow, 10 ⁶ lb/h	1.16
Reheat Steam Conditions, psig/°F	530/1000
Condenser Pressure, in Hg Abs	2.0
Feedwater Heating Stages	2-LP, D/A & Exhaust Gases

EMISSIONS

Total Stack Flow, 10 ⁶ lb/h	5.83
Stack Gas Temperature, °F	280
Sulfur Removal, %	90
Particulate, lb/10 ⁶ Btu	Near 0.0
NO _x , lb/10 ⁶ Btu	0.25

EXHIBIT 4-8

MATERIAL BALANCE SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC
COMBINED CYCLE POWER PLANT

EXHIBIT 4-9

AUXILIARY POWER REQUIREMENTS FOR A
NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC
PFBC COMBINED CYCLE POWER PLANT

Coal and Dolomite Handling and Feed	1300
Waste Solids Handling	600
Condensate Pumps	200
Circulating Water Pumps	3000
Service Water Pumps	200
Cooling Tower Fans	2100
Transformer Losses, Lighting & Miscellaneous	3000
Gas Turbine Auxiliary Load	100
Steam Turbine Auxiliary Load	400
PFBC Related Auxiliary Equipment	1100
Miscellaneous	<u>300</u>
Reference PFBC Plant Subtotal (Section 3)	13000 kW
Equipment Modifications:	
Blow-back System Compressors (added)	180
Blow-back System Heaters (added)	45
Baghouse Filter Reverse Air Fans (deleted)	<u>(85)</u>
Subtotal	140 kW
Operational Modifications:	
CXF Units @ $\Delta p = 0.75$ psi (added)	1680
Baghouse Filter @ $\Delta p = 0.14$ psi (deleted)	(825)
Third Stage Multicyclone @ $\Delta p = 1.6$ psi (deleted)	<u>(3580)</u>
Subtotal	<u>(2725)</u> kW
Total	10415 kW

EXHIBIT 4-10

TOTAL CAPITAL REQUIREMENTS FOR A
NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC
PFBC POWER PLANT
(January 1980 \$1000)

<u>FERC</u> <u>Acct</u>	<u>Description</u>	
310	Land and Land Rights	\$ 2,420
311	Structures and Improvements	39,511
312	Boiler Plant Equipment	192,543
314	Turbogenerator Units	90,100
315	Accessory Electrical Equipment	25,140
316	Misc. Power Plant Equipment	3,300
353	Station Equipment	3,400
Total Direct Costs (Excluding Land)		<u>353,994</u>
Undistributed Costs		21,240
Process Capital		<u>375,234</u>
General Facilities		0
Engineering and Home Office Fees		35,399
Subtotal		<u>410,633</u>
Project Contingency		61,595
Process Contingency		12,452
Sales Tax		0
Total Plant Investment		<u>484,680</u>
Total Plant Investment at Startup		<u>503,318</u>
Royalty Allowance		0
Preproduction Costs		12,890
Inventory Capital		6,293
Initial Catalyst and Chemicals Charge		0
Land		2,420
Total Capital Requirements		<u>524,921</u>
Total Capital Required, \$/kW net		984
Net Plant Output, MWe		534

Note: Costs are based on January 1980 startup for a nominal 500 MW steam-cooled reference PFBC plant located in a mid-continental U.S. area. The eastern coal-fired plant burns coal with a HHV of 10,100 Btu/lb and 4% sulfur.

EXHIBIT 4-11

NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT
MODIFICATIONS TO REFERENCE PFBC PLANT CAPITAL COSTS

FERC Account	Reference PFBC Plant Cost Estimate (Section 3) (\$1000)	Modification	Cyclone/CXF/HGC PFBC Plant Cost Estimate (January 1980 \$1000)
310 Land	2,420	• None	2,420
311 Structures and Improvements	39,626	• Structural adjustment resulting from deletion of third stage multicyclones and addition of third stage CXF Units	39,511
312 Boiler Plant Equipment			
PFB Combustors	37,900		
PFB Combustor Solids Feeding	35,000		
PFB Combustor Solids Removal	4,500		
Primary Multicyclone	9,880		
Secondary Multicyclone	7,900		
Tertiary Multicyclone	6,920	• Third Stage Multicyclone Deletion	0
Multicyclone Solids Removal	1,000		
Refractory Lined Pipe	19,026	• Reduction in Piping	19,004
Refractory & Metal Lined Pipe	5,242		
Baghouse Filter	17,700	• Baghouse Filter Deletion	0
Balance of Account 312	66,600	• Addition of CXF Units	5,095
		• Addition of Blow-Back System	422
	211,668		192,543
314 Turbogenerator Units	90,100	• None	90,100
315 Accessory Electrical Equipment	25,140	• None	25,140
316 Miscellaneous Power Plant Equipment	3,300	• None	3,300
353 Station Equipment	3,400	• None	3,400
TOTAL (Excluding Land)	373,234		353,994

Process contingency costs are based on the cost of unproven equipment: PFBC boiler, ash removal systems, coal and dolomite preparation and injection systems, and HGC cleanup system. The costs of these components are shown in Exhibit 4-12.

Annual operating and maintenance labor costs are based on the estimated requirements presented in Exhibit 4-13. These estimates are derived by modifying the reference PCF/FGD plant O&M cost data, presented in Exhibit 6-12, to accommodate the steam-cooled 2-cyclone/CXF/HGC PFBC plant O&M requirements.

Estimated preproduction and inventory capital requirements are presented in Exhibits 4-14 and 4-15, respectively.

The resulting plant operating costs for the nominal 500 MW steam-cooled 2-cyclone/CXF/HGC PFBC combined cycle plant are presented in Exhibit 4-16.

The capital and operating cost results for the nominal 500 MW steam-cooled 2-cyclone/CXF/HGC PFBC plant, presented in Exhibits 4-10 and 4-16, respectively, indicate a total capital requirement of \$984/kW net and a busbar power cost of 68.30 mills/kWh. Compared to the reference PCF/FGD plant costs, presented in Exhibit 6-8 and 6-9, the steam-cooled 2-cyclone/CXF/HGC PFBC plant is about 17.3% more expensive in total capital requirement and about 1.6% more expensive in busbar power cost.

EXHIBIT 4-12

500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT
COSTS OF ADVANCED TECHNOLOGY EQUIPMENT TO WHICH
PROCESS CONTINGENCY IS APPLIED
(January 1980 \$1000)

FERC ACCT 312

PFB Combustor/Boiler	37,900
PFB Spent Solids Removal System	4,500
Coal and Dolomite Preparation & Injection Systems	35,000
Hot Gas Cleanup System:	
1st Stage Multicyclones	9,880
2nd Stage Multicyclones	7,900
3rd Stage CXF Units	5,095
Refractory Lined Large Hot Gas Piping	19,004
Metal Lined Large Hot Gas Piping	<u>5,242</u>
TOTAL	124,521

EXHIBIT 4-13

ESTIMATE OF ANNUAL OPERATION AND MAINTENANCE LABOR COST (JANUARY 1980 DOLLARS)
FOR A NOMINAL 500 MW STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT

Classification	Annual* Salary Per Capita	No. of Person.	Annual Salary
<u>Operation</u>			
Operation Supervisor	39,600	1	39,600
Shift Supervisor	39,300	6	217,800
Control Room Operator	33,000	6	198,000
Asst. Control Room Operator	26,400	6	158,400
Turbine Operator	33,000	12	396,000
Aux. Equipment Operator	26,400	12	316,800
Boiler Operator	26,400	6	158,400
Coal and Ash Crew	24,200	10	242,000
Computer Specialist	28,000	1	28,000
Environmental Specialist	28,000	1	28,000
Results Engineer	26,400	1	26,400
Assistant Results Engineer	22,500	4	90,000
Chemical Supervisor	36,300	1	36,300
Chemists	26,400	4	105,600
		<u>71</u>	<u>2,041,300</u>
<u>Maintenance</u>			
Maintenance Supervisor	39,600	1	39,600
Mechanical Maint. Foreman	34,100	1	34,100
Mechanic	26,400	16	422,400
Mechanic's Helper	20,900	16	334,400
Machinist	26,400	2	52,800
Welder	26,400	3	79,200
Carpenter	24,200	2	48,400
Bricklayer	20,900	1	20,900
Electr. Maint. Foreman	34,100	1	34,100
Electrician	26,400	8	211,200
Electrician's Helper	20,900	8	167,200
I & C Foreman	34,100	1	34,100
I & C Repairman	28,600	5	143,000
Building Maint. Foreman	24,200	1	24,200
Plumber	20,900	1	20,900
Laborers	13,200	<u>16</u>	<u>211,200</u>
		<u>83</u>	<u>1,877,700</u>
		154	3,919,000
<u>Average Hourly Rates (\$/HR)</u>			
Operating			13.83
Maintenance			10.88
Total			12.23

*Annual salaries include both direct labor charges and payroll burden.

Notes: 1) Man-power estimates are based on a single unit 544 MWe steam-cooled PFBC power generating plant.

$$2) \text{ Average Hourly Rate} = \frac{\text{Total Annual Cost (\$)}}{\text{Number of Person.} \times 2080 \text{ working hrs/year}}$$

EXHIBIT 4-14

PREPRODUCTION COSTS FOR A NOMINAL 500 MW
STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT

ITEM	QUANTITY	UNIT COST	CONVERSION FACTOR	COST ¹
Preproduction Cost				
Operating Labor	\$2,041,300	See Note 2	1/12	170,100
Maintenance Labor	\$1,877,700	See Note 2	1/12	156,500
Administrative Labor	\$1,175,700	See Note 3	1/12	98,000
Maintenance Material	\$6,384,200	See Note 4	1/12	532,000
Waste Disposal	91.90 TPH	\$ <u>4.50</u> Ton	730 h	301,900
Dolomite	67.71 TPH	\$ <u>11.00</u> Ton	730 h	543,700
Water	4550 GPM	\$ <u>0.45</u> 1000 Gal	730 h	89,700
Coal	228.27 TPH	\$ <u>31.31</u> Ton	$\frac{730}{4}$ h	1,304,400
2% TPI				<u>9,693,600</u>
Total Preproduction Cost				12,889,900

- Note:
1. All Costs are in January 1980 dollars
 2. Exhibit 3-11 presents annual operating and maintenance labor
 3. Administrative costs are 30% of O&M labor costs
 4. Maintenance materials taken as 3.4 x maintenance labor

EXHIBIT 4-15

INVENTORY CAPITAL FOR A NOMINAL 500 MW
STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT

Inventory Capital Material	Quantity	Rate	Conversion Factor	Cost ¹
Coal	228.27 TPH	<u>\$31.31</u> Ton	730 h	5,217,400
Dolomite	67.71 TPH	<u>\$11.00</u> Ton	730 h	543,700
Maintenance Material				<u>532,000</u>
TOTAL				6,293,100

Note: 1. All costs are in January 1980 dollars

EXHIBIT 4-16

CAPITAL INVESTMENT AND
REVENUE REQUIREMENTS FOR A 500 MW
STEAM-COOLED 2-CYCLONE/CXF/HGC PFBC POWER PLANT

PARAMETER	EASTERN COAL
<u>Criteria</u>	
SO ₂ Emission Std.	90%
Ca/S, mole ratio	1.3
<u>Operating Conditions</u>	
Gross Power Output, MW	544.000
Auxiliary Power, MW	10.415
Net Power Output, MW	533.585
Net Plant Heat Rate, Btu/kWh	8,642
<u>Materials/Consumption/Production</u>	
Coal Consumption, ton/h	228.27
Dolomite Consumption, ton/h	67.71
Solid Wastes, wet, ton/h	0.0
Solid Wastes, dry, ton/h	91.90
Water Consumption, GPM	4,550
<u>Capital Investment (January 1980 dollars)</u>	
Capital Cost, \$ millions	524.9
Capital Cost, \$/net kW	984
30-Yr Levelized Fixed Charges, \$/kW-yr	177.1
<u>30-Yr Levelized Costs (1980-2009)</u>	
Dolomite, mills/kWh	2.64
Waste Disposal, mills/kWh	1.46
Water, mills/kWh	0.43
Other Consumables, mills/kWh	0.0
Total Variable O&M, mills/kWh	0.28
Fixed O&M, mills/kWh	4.81
Fixed Charges, mills/kWh	6.62
Subtotal, mills/kWh	28.88
Fuel Cost, mills/kWh	40.31
Busbar Power Cost, mills/kWh	27.99
	68.30

Results are based on plant located in mid-continental U.S. area burning eastern coal with a HHV of 10,100 Btu/lb, 4% sulfur and a 70% capacity factor. All costs are based on January 1980 dollars.

Section 5

500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC COMBINED CYCLE POWER PLANT

The nominal 500 MW steam-cooled 1-cyclone/CXF/HGC PFBC combined cycle power plant concept burning eastern coal, presented in the study, is based on the reference conceptual plant design presented in Section 3. The reference plant design was modified by Burns and Roe, in the course of developing the 1-cyclone/CXF concept, to accommodate a Westinghouse R&D specified HGC system comprising 1 cyclonic stage and a CXF stage. The 1-cyclone/CXF/HGC system concept is designed to provide adequate gas turbine particulate protection and to achieve compliance with current EPA particulate ($0.03 \text{ lbs}/10^6 \text{ Btu}$) and opacity standards, and to offer the potential to achieve compliance with possible future inhaleable particulate standards and/or a more stringent opacity standard. The ceramic cross-flow filter is a near perfect filter with regard to submicron particulate.

The nominal 500 MW steam-cooled 1-cyclone/CXF/HGC PFBC combined cycle power plant simplified flow schematic is presented in Exhibit 5-1. The coal and dolomite handling and PFBC system preliminary general arrangement is shown in Exhibit 5-2, and the preliminary plant hot gas piping arrangement is shown in Exhibit 5-3. The plant comprises two PFBC/GT modules, 123 MW total, supplying steam to a common subcritical steam turbine, rated at 421 MW, and does not require a baghouse filter to achieve compliance with the current NSPS particulate emission standard. Each PFBC/GT module consists of a commercially available gas turbogenerator, a pair of dedicated complementary PFB combustor modules, designated as vessel A (saturation duty) and vessel B (superheat/reheat duty), and a HGC system comprising one cyclonic stage followed by a CXF stage. PFBC solid wastes generated by the plant are depressurized, cooled and stored in on-site waste storage silos prior to ultimate disposal off-site.

EXHIBIT 5-1

SIMPLIFIED FLOW SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC
PFBC COMBINED CYCLE POWER PLANT

EXHIBIT 5-2

PRELIMINARY GENERAL ARRANGEMENT FOR A NOMINAL 500 MW
STEAM-COOLED 1-CYCLONE/CXF PFBC COMBINED CYCLE PLANT

Note: Dimensions of Vessels are Specific to Vessel A. Vessel B Arrangement will be Conceptually Similar

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FL.

EXHIBIT 5-3

PRELIMINARY HOT GAS PIPING ARRANGEMENT FOR A NOMINAL 500 MW STEAM-COOLED
1-CYCLONE/CXF/HGC PFBC COMBINED CYCLE POWER PLANT

The combustion and steam generation processes for the complementary PFB combustor modules of the 1-cyclone/CXF/HGC PFBC plant are essentially identical to those presented in Section 4, for the 2-cyclone/CXF/HGC plant concept. Similarly, flue gas from each complementary PFB combustor is ducted through the 2-cyclone/CXF/HGC system and routed to the common dedicated gas turbine. Gas turbine exhaust waste heat recovery and spent bed material waste heat recovery is accomplished as described in Section 4. Similarly, the sensible heat in the fine spent solids collected by the cyclone stage and the CXF stage is utilized for feedwater heating, after depressurization, using helical screw type solid coolers.

1-Cyclone/CXF/HGC System

The steam-cooled 1-cyclone/CXF/HGC PFBC plant utilizes one stage of cyclonic and one stage of CXF particulate removal for gas turbine protection. Each PFB combustor bed is coupled to a first stage multicyclone module. Each first stage multicyclone module is coupled to a second stage CXF unit.

Exhibit 5-4 presents a schematic diagram of the multicyclone HGC module adopted for this study, and includes an accompanying tabulation of module dimension and quantity of cyclones. The multicyclone HGC module is described in Section 4.

Exhibit 5-5 presents a simplified diagram of the CXF module adopted for this study. Exhibit 5-6 presents a simplified diagram of the CXF unit, and includes an accompanying tabulation of unit dimensions and quantity of modules for both the A and B vessel PFB combustor HGC trains. The CXF module and CXF unit are described in Section 4 and Appendix A.

The interconnecting hot gas piping, between PFB combustor and multicyclone modules, and between multicyclone modules and CXF units is sized for a nominal gas velocity of 50 fps and is fabricated as a two component castable refractory lined carbon

EXHIBIT 5-4

MULTICYCLONE MODULE SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC COMBINED CYCLE
POWER PLANT

EXHIBIT 5-5
CERAMIC CROSS-FLOW FILTRATION MODULE

EXHIBIT 5-6
CERAMIC CROSS-FLOW FILTRATION UNIT

steel pipe, as described in Section 4. The hot gas piping between third stage CXF units and turbine expander inlet is sized for a nominal velocity of 100 fps and is fabricated as a single component castable refractory lined carbon steel pipe, as described in Section 4.

Primary Cyclonic HGC Stage

The fine particulate laden effluent gas stream from each PFB combustor bed is ducted into a multicyclone module containing the requisite number of high efficiency tangential entry primary cyclones, where the relatively coarser particles are separated from the gas stream by centrifugal action. There are nine primary cyclones per 18 ft. diameter x 36 ft. high primary multicyclone module in each A vessel PFB combustor train, with each cyclone designed to operate with an inlet gas velocity of approximately 88 fps. Correspondingly, there are twelve cyclones per 20 ft. diameter x 36 ft. high primary module in each B train, with each cyclone designed to operate with an inlet gas velocity of approximately 86 fps. Each primary cyclone is about 4 ft. in diameter and operates with a pressure drop estimated to be about 0.7 psi. As previously discussed in Section 3, the inlet and outlet particulate loading from each primary cyclone is estimated to be about 17,442 and 1233 ppm, respectively, based on scaling the performance data presented in reference 3-1. The outlet gas stream from each primary cyclone is combined in the upper plenum section of the module. Particulate collected by each cyclone in the primary module is discharged to a common standpipe, via dipleg connections, depressurized, cooled and pneumatically transported to the fine waste storage silo.

Secondary Ceramic Cross-Flow Filtration HGC Stage

The effluent gas stream from each primary multicyclone module upper plenum is ducted into a secondary CXF unit, containing the requisite number of high efficiency CXF modules, which forms the second stage of the HGC system. There are fourteen

CXF modules per 13.5 ft. diameter x 31.5 ft. high CXF unit in each A vessel PFB combustor train. Correspondingly, there are eighteen CXF modules per 16 ft. diameter x 33.5 ft. high CXF unit in each B train. Each CXF module is about 16 ft. in height and each CXF unit operates with an average pressure drop estimated to be about 0.75 psi. The outlet particulate loading for each CXF unit is estimated to be about 0 ppm. The outlet gas stream from each CXF module is combined in the upper plenum section of the CXF unit. The outlet gas stream from each pair of CXF units in each PFBC train is combined into a single line and then routed to the gas turbine expander inlet. Particulate collected by each module in a secondary stage CXF unit is collected during the periodic on line filter cleaning (backwashing) phase of the gas cleaning cycle. The collected particulate is depressurized, cooled and pneumatically transported to the fine waste storage silo, as described in Section 4.

Blow-Back Air System

The blow-back air system is required to provide the high temperature, high pressure air necessary for ceramic cross-flow filter backwashing. The system consists of a bank of 3 motor driven reciprocating compressors, located in the plant compressor building, delivering high pressure air, via the interconnecting piping, to each dedicated compressed air accumulator located above each CXF unit. Each compressor is sized to deliver one-half of the plant's blow-back air requirement based on each CXF module operating on a 95 minute gas filtration cycle, requiring 44.1 pps of air at 45 atmospheres and 500°F, for a duration of 0.1 seconds. The combination of 3 compressors, each equipped with reed valves will enable the blow-back system to supply backwash air during partial load plant operation, provide for a standby compressor during full load plant operation, and provide additional compressed air capacity for possible startup or maintenance considerations. The compressors are designed to deliver air at 45 atmospheres pressure and near ambient temperature to each accumulator. The design features of the accumulator and the related CXF backwashing operation are described in Section 4.

Operating and Design Parameters

The general operating and design parameters applicable to the PFB combustor modules and gas and steam turbogenerators of the 1-cyclone/CXF/HGC PFBC plant concept are taken to be identical to those presented for the 2-cyclone/CXF/HGC plant concept presented in Section 4. Exhibit 5-7 presents the resulting steam-cooled 1-cyclone/CXF/HGC PFBC plant operating and design parameters for the eastern coal investigated in this study.

Exhibit 5-8 presents the material balance schematic, with all material streams expressed in pph.

Exhibit 5-9 presents the auxiliary power requirements. System power losses incurred due to the implementation of the specified 1-cyclone/CXF/HGC system and the attendant deletion of the second and third stage multicyclones and baghouse filter from the reference PFBC plant concept, are reported, for the sake of convenience, as auxiliary power requirements, rather than an increase or decrease in gas and/or steam turbogenerator output. The auxiliary power requirements shown under "Equipment Modification" include the increase or decrease in power requirement resulting from the addition or deletion of the equipment indicated. The requirements shown under "Operational Modifications" include the increase or decrease in system power losses attributable to flow through the HGC system equipment indicated, calculated as the product of pressure drop and average volumetric gas flow rate.

ECONOMIC ANALYSES

The economic evaluation of the nominal 500 MW steam-cooled 1-cyclone/CXF/HGC PFBC combined cycle power plant, combusting eastern coal, was conducted in conformance with the criteria presented in Section 2.

EXHIBIT 5-7

OPERATING AND DESIGN PARAMETERS FOR A NOMINAL 500 MW
STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT
WITH A Ca/S = 1.3

POWER OUTPUT

Gross Steam Turbine Output, kW	421,000
Gross Gas Turbine Output, kW	123,000
Gross Output, kW	544,000
Auxiliary Power, kW	7,700
Net Plant Output, kW	536,300
Net Plant Heat Rate, Btu/kWh	8.598

PFBC COMBUSTOR

Number	4
Type	Cylindrical, 2-Bed Horizontal
Total Air Flow, 10^6 lb/h	5.25
Air Inlet Temperature, °F	544
Air Outlet Temperature, °F	1636
PFBC Operating Pressure, psig	98
Bed Depth, ft	9
Bed Temperature, °F	1686
Combustor Size (Evaporator Duty)	36' Dia. x 100' L
Combustor Size (SH/RH Duty)	36' Dia. x 115' L
Hot Gas Cleanup System Type	1-Stage Cyclone/CXF
Superficial Velocity, ft/sec	4.5

GAS TURBINE

Number	2
Type	Commercially Available Utility Gas Turbine
Compressor Pressure Ratio	8
Total Compressor Air Inlet Flow, 10^6 lb/h	5.47
Compressor Outlet Temperature, °F	544
Turbine Inlet Conditions, psig/°F	87.2/1600
Turbine Outlet Temperature, °F	923
Exhaust Gas Waste Heat Recovery	Feedwater Heating

STEAM CYCLE

Main Steam Flow, 10^6 lb/h	1.18
Main Steam Conditions, psig/°F	2400/1000
Reheat Steam Flow, 10^6 lb/h	1.16
Reheat Steam Conditions, psig/°F	530/1000
Condenser Pressure, in Hg Abs	2.0
Feedwater Heating Stages	2-LP, D/A & Exhaust Gases

EMISSIONS

Total Stack Flow, 10^6 lb/h	5.83
Stack Gas Temperature, °F	280
Sulfur Removal, %	90
Particulate, lb/ 10^6 Btu	0.004
NO _x , lb/ 10^6 Btu	0.25

EXHIBIT 5-8

MATERIAL BALANCE SCHEMATIC FOR A
NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC
COMBINED CYCLE POWER PLANT

EXHIBIT 5-9

AUXILIARY POWER REQUIREMENTS FOR A NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC COMBINED CYCLE POWER PLANT

Coal and Dolomite Handling and Feed	1300
Waste Solids Handling	600
Condensate Pumps	200
Circulating Water Pumps	3000
Service Water Pumps	200
Cooling Tower Fans	2100
Transformer Losses, Lighting & Miscellaneous	3000
Gas Turbine Auxiliary Load	100
Steam Turbine Auxiliary Load	400
PFB Related Auxiliary Equipment	1100
Miscellaneous	<u>300</u>
Reference PFBC Plant Subtotal (Section 3)	13000 kW

Equipment Modifications:

Blow-back System Compressors (added)	65
Blow-back System Heaters (added)	25
Baghouse Filter Reverse Air Fans (deleted)	<u>(85)</u>
Subtotal	5 kW

Operational Modifications:

CXF Units @ $\Delta p = 0.75$ psi (added)	1660
Baghouse Filter @ $\Delta p = 0.14$ psi (deleted)	(825)
Second Stage Multicyclone @ $\Delta p = 1.1$ psi (deleted)	(2440)
Third Stage Multicyclone @ $\Delta p = 1.6$ psi (deleted)	(3580)
Compressed Combustion Air Line (52'/train reduction)	(50)
Refractory Lined Hot Gas Piping (90'/train reduction)	(65)
Metal Lined Hot Gas Piping (5'/train reduction)	<u>(5)</u>
Subtotal	<u>(5305)</u> kW
Total	7700 kW

The resulting capital cost estimates and total capital requirements for the 500 MW plant are presented in Exhibit 5-10 by FERC (Federal Energy Regulatory Commission) code of account number, expressed in January 1980 dollars. Exhibit 5-11 presents a summary of the modifications applied to the indicated cost to arrive at the capital costs used in the economic analysis of the steam-cooled PFBC plant, utilizing a 1-cyclone/CXF/HGC system.

Process contingency costs are based on the cost of unproven equipment: PFBC boiler, ash removal systems, coal and dolomite preparation and injection systems, and HGC cleanup system. The costs of these components are shown in Exhibit 5-12.

Annual operating and maintenance labor costs are based on the estimated requirements presented in Exhibit 5-13. These estimates are derived by modifying the reference PCF/FGD plant O&M cost data, presented in Exhibit 6-12, to accommodate the steam-cooled 1-cyclone/CXF/HGC PFBC plant O&M requirements.

Estimated preproduction and inventory capital requirements are presented in Exhibits 5-14 and 5-15, respectively.

The resulting plant operating costs for the nominal 500 MW steam-cooled 1-cyclone/CXF/HGC PFBC combined cycle plant are presented in Exhibit 5-16.

The capital and operating cost results for the nominal 500 MW steam-cooled 1-cyclone/CXF/HGC PFBC plant, presented in Exhibits 5-10 and 5-16, respectively, indicate a total capital requirement of \$947/kW net and a busbar power cost of 67.02 mills/kWh. Compared to the reference PCF/FGD plant costs, presented in Exhibit 6-8 and 6-9, the steam-cooled 1-cyclone/CXF/HGC PFBC plant is about 12.9% more expensive in total capital requirement and about 0.3% less expensive in busbar power cost.

EXHIBIT 5-10

TOTAL CAPITAL REQUIREMENTS FOR A
NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC
PFBC POWER PLANT
(January 1980 \$1000)

<u>FERC Acct</u>	<u>Description</u>	
310	Land and Land Rights	\$ 2,420
311	Structures and Improvements	38,807
312	Boiler Plant Equipment	181,930
314	Turbogenerator Units	90,100
315	Accessory Electrical Equipment	25,140
316	Misc. Power Plant Equipment	3,300
353	Station Equipment	3,400
	Total Direct Costs (Excluding Land)	<u>342,677</u>
	Undistributed Costs	20,560
	Process Capital	<u>363,237</u>
	General Facilities	0
	Engineering and Home Office Fees	34,268
	Subtotal	<u>397,505</u>
	Project Contingency	59,626
	Process Contingency	11,402
	Sales Tax	0
	Total Plant Investment	<u>468,533</u>
	Total Plant Investment at Startup	<u>486,550</u>
	Royalty Allowance	0
	Preproduction Costs	12,567
	Inventory Capital	6,293
	Initial Catalyst and Chemicals Charge	0
	Land	2,420
	Total Capital Requirements	<u>507,830</u>
	Total Capital Required, \$/kW net	947
	Net Plant Output, MWe	536

Note: Costs are based on January 1980 startup for a nominal 500 MW steam-cooled reference PFBC plant located in a mid-continental U.S. area. The eastern coal-fired plant burns coal with a HHV of 10,100 Btu/lb and 4% sulfur.

EXHIBIT 5-11

NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT
MODIFICATIONS TO REFERENCE PFBC PLANT CAPITAL COSTS

FERC Account	Reference PFBC Plant Cost Estimate (Section 3) (\$1000)	Modification	1-Cyclone/CXF/HGC PFBC Plant Cost Estimate (January 1980 \$1000)
310 Land	2,420	• None	2,420
311 Structures and Improvements	39,626	• Structural adjustment resulting from deletion of second and third stage multicyclones and addition of second stage CXF Units	38,807
312 Boiler Plant Equipment			
PFB Combustors	37,900		
PFB Combustor Solids Feeding	35,000		
PFB Combustor Solids Removal	4,500		
Primary Multicyclone	9,880		
Secondary Multicyclone	7,900	• Second Stage Multicyclone Deletion	0
Tertiary Multicyclone	6,920	• Third Stage Multicyclone Deletion	0
Multicyclone Solids Removal	1,000		
Refractory Lined Pipe	19,026	• Reduction in Piping	16,519
Refractory & Metal Lined Pipe	5,242	• Reduction in Piping	5,500
Baghouse Filter	17,700	• Baghouse Filter Deletion	0
Balance of Account 312	66,600	• Additions of CXF Units	5,095
		• Addition of Blow-Back System	313
	<u>211,668</u>		<u>181,930</u>
314 Turbogenerator Units	90,100	• None	90,100
315 Accessory Electrical Equipment	25,140	• None	24,140
316 Miscellaneous Power Plant Equipment	3,300	• None	3,300
353 Station Equipment	<u>3,400</u>	• None	<u>3,400</u>
TOTAL (Excluding Land)	373,234		347,677

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EXHIBIT 5-12

500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT
COSTS OF ADVANCED TECHNOLOGY EQUIPMENT TO WHICH
PROCESS CONTINGENCY IS APPLIED
(January 1980 \$1000)

FERC ACCT 312

PFB Combustor/Boiler	37,900
PFB Spent Solids Removal System	4,500
Coal and Dolomite Preparation & Injection Systems	35,000
Hot Gas Cleanup System:	
1st Stage Multicyclones	9,880
2nd Stage CXF Units	5,095
Refractory Lined Large Hot Gas Piping	16,519
Metal Lined Large Hot Gas Piping	<u>5,123</u>
TOTAL	114,017

EXHIBIT 5-13

ESTIMATE OF ANNUAL OPERATION AND MAINTENANCE LABOR COST (January 1980 Dollars)
FOR A NOMINAL 500 MW STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT

Classification	Annual* Salary Per Capita	No. of Person.	Annual Salary
<u>Operation</u>			
Operation Supervisor	39,600	1	39,600
Shift Supervisor	39,300	6	217,800
Control Room Operator	33,000	6	198,000
Asst. Control Room Operator	26,400	6	158,400
Turbine Operator	33,000	12	396,000
Aux. Equipment Operator	26,400	12	316,800
Boiler Operator	26,400	6	158,400
Coal and Ash Crew	24,200	10	242,000
Computer Specialist	28,000	1	28,000
Environmental Specialist	28,000	1	28,000
Results Engineer	26,400	1	26,400
Assistant Results Engineer	22,500	4	90,000
Chemical Supervisor	36,300	1	36,300
Chemists	26,400	4	105,600
		<u>71</u>	<u>2,041,300</u>
<u>Maintenance</u>			
Maintenance Supervisor	39,600	1	39,600
Mechanical Maint. Foreman	34,100	1	34,100
Mechanic	26,400	16	422,400
Mechanic's Helper	20,900	16	334,400
Machinist	26,400	2	52,800
Welder	26,400	3	79,200
Carpenter	24,200	2	48,400
Bricklayer	20,900	1	20,900
Electr. Maint. Foreman	34,100	1	34,100
Electrician	26,400	8	211,200
Electrician's Helper	20,900	8	167,200
I & C Foreman	34,100	1	34,100
I & C Repairman	28,600	5	143,000
Building Maint. Foreman	24,200	1	24,200
Plumber	20,900	1	20,900
Laborers	13,200	<u>16</u>	<u>211,200</u>
		<u>83</u>	<u>1,877,700</u>
		154	3,919,000
<u>Average Hourly Rates (\$/HR)</u>			
Operating			13.83
Maintenance			10.88
Total			12.23

*Annual salaries include both direct labor charges and payroll burden.

Notes: 1) Man-power estimates are based on a single unit 544 MWe steam-cooled PFBC power generating plant.

$$2) \text{ Average Hourly Rate} = \frac{\text{Total Annual Cost (\$)}}{\text{Number of Person.} \times 2080 \text{ working hrs/year}}$$

EXHIBIT 5-14

PREPRODUCTION COSTS FOR A NOMINAL 500 MW
STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT

ITEM	QUANTITY	UNIT COST	CONVERSION FACTOR	COST ¹
Preproduction Cost				
Operating Labor	\$2,041,300	See Note 2	1/12	170,100
Maintenance Labor	\$1,877,700	See Note 2	1/12	156,500
Administrative Labor	\$1,175,700	See Note 3	1/12	98,000
Maintenance Material	\$6,384,200	See Note 4	1/12	532,000
Waste Disposal	91.90 TPH	\$ <u>4.50</u> Ton	730 h	301,900
Dolomite	67.71 TPH	\$ <u>11.00</u> Ton	730 h	543,700
Water	4550 GPM	\$ <u>0.45</u> 1000 Gal	730 h	89,700
Coal	228.27 TPH	\$ <u>31.31</u> Ton	<u>730</u> 4 h	1,304,400
2% TPI				<u>9,370,700</u>
Total Preproduction Cost				12,567,000

- Note:
1. All Costs are in January 1980 dollars
 2. Exhibit 3-11 presents annual operating and maintenance labor
 3. Administrative costs are 30% of O&M labor costs
 4. Maintenance materials taken as 3.4 x maintenance labor

EXHIBIT 5-15

INVENTORY CAPITAL FOR A NOMINAL 500 MW
STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT

Inventory Capital Material	Quantity	Rate	Conversion Factor	Cost ¹
Coal	228.27 TPH	<u>\$31.31</u> Ton	730 h	5,217,400
Dolomite	67.71 TPH	<u>\$11.00</u> Ton	730 h	543,700
Maintenance Material				<u>532,000</u>
TOTAL				6,293,100

Note: 1. All costs are in January 1980 dollars

EXHIBIT 5-16

CAPITAL INVESTMENT AND
REVENUE REQUIREMENTS FOR A 500 MW
STEAM-COOLED 1-CYCLONE/CXF/HGC PFBC POWER PLANT

PARAMETER	EASTERN COAL
<u>Criteria</u>	
SO ₂ Emission Std.	90%
Ca/S, mole ratio	1.3
<u>Operating Conditions</u>	
Gross Power Output, MW	544.000
Auxiliary Power, MW	7.700
Net Power Output, MW	536.300
Net Plant Heat Rate, Btu/kWh	8,598
<u>Materials/Consumption/Production</u>	
Coal Consumption, ton/h	228.27
Dolomite Consumption, ton/h	67.71
Solid Wastes, wet, ton/h	0.0
Solid Wastes, dry, ton/h	91.90
Water Consumption, GPM	4,550
<u>Capital Investment (January 1980 dollars)</u>	
Capital Cost, \$ millions	507.8
Capital Cost, \$/net kW	947
30-Yr Levelized Fixed Charges, \$/kW-yr	170.4
<u>30-Yr Levelized Costs (1980-2009)</u>	
Dolomite, mills/kWh	2.63
Waste Disposal, mills/kWh	1.45
Water, mills/kWh	0.43
Other Consumables, mills/kWh	0.0
Total Variable O&M, mills/kWh	0.28
Fixed O&M, mills/kWh	4.79
Fixed Charges, mills/kWh	6.58
Subtotal, mills/kWh	27.80
Fuel Cost, mills/kWh	39.17
Busbar Power Cost, mills/kWh	27.85
	67.02

Results are based on plant located in mid-continental U.S. area burning eastern coal with a HHV of 10,100 Btu/lb, 4% sulfur and a 70% capacity factor. All costs are based on January 1980 dollars.

Section 7

REFERENCES

- 2-1 "EPRI Economic Premises for Electric Power Generating Plants," Reference Date: May 19, 1980.
- 3-1 "PFB Contractor Services for the Design of Advanced Fossil Fuel Systems - Task 1: Steam-Cooled PFB Reference Plant Design," Contract No. ANL-31-109-38-6308, J.R. Hamm, Westinghouse Electric Corporation/Foster Wheeler Development Co., October 30, 1981.
- 3-2 "PFB Contractor Services for the Design of Advanced Fossil Fuel Systems - Task 2: Steam-Cooled PFB Reference Plant Costs," Contract No. ANL-31-109-38-6308, J.R. Hamm, Westinghouse Electric Corporation/Foster Wheeler Development Co., October 30, 1981.
- 3-3 "A-E Services for the Design of Advanced Fossil Fuel Systems - Steam-Cooled Pressurized Fluidized Bed Combined Cycle Power Plant, Final Report," Contract No. ANL-37-109-38-6212, Bechtel Group, Inc., December 1981.
- 3-4 "Coal-Fired Power Plant Capital Cost Estimates," EPRI PE-1865, TPS 78-810, Final Report, May 1981, Prepared by Bechtel Power Corporation, San Francisco, California.
- 6-1 "Economic Evaluation Techniques, Results, and Computer Modeling for Flue Gas Desulfurization", R.L. Torstrick, L.J. Henson, and S.V. Tomlinson, Emission Control Development Projects, Tennessee Valley Authority, Muscle Shoals, Alabama, EPA Flue Gas Desulfurization Symposium, November 1977.

Section A1

PRINCIPLES OF OPERATION

The Westinghouse R&D Center ceramic cross-flow filtration (CXF) concept, herein described, is an outgrowth of the current DOE project (Contract No. DE-AC21-79ET15491) for the development of a rigid porous ceramic membrane filter of unique construction. The current DOE project is an extension of earlier work performed by WRDC for EPA (Contract No. 68-02-1887/Report EPA-600/2-77-207), in which a cross-flow ceramic membrane filter was found to give essentially 100 percent efficiencies on sub-micron dust at modest pressure drops. The previously tested cross-flow filter elements, having the geometry shown in Exhibit A-1, were fabricated in the following manner:

- A thin flat layer of gas-permeable ceramic material is laid out
- A layer of corrugations is then placed on the flat layer
- A second flat layer is placed on top of the corrugations
- Finally, a second layer of corrugations is placed on the second flat layer, oriented perpendicularly to the first corrugations

This sequence was repeated over and over, to a maximum of about six inches or 50 to 85 layers. The element was then fired and the composite became a monolith. The currently used alternative route to a similar structure is to extrude (or machine) thin flat sheets with corduroy-like ribs on one side, and then stack the sheets with alternating rib orientation, as shown in Exhibit A-2. Again, the green ceramic mix material is laid up and then fired to form a unitized structure.

A-1

EXHIBIT A-1

CROSS-FLOW CERAMIC MEMBRANE FILTER
(Courtesy of Westinghouse R&D Center)

EXHIBIT A-2

NEW FILTER MEDIA FABRICATED BY COORS APPROXIMATELY
1.31 FT² (.12 m²) FILTER AREA (10 μ PORE SIZE)
60 mil (1.5 mm) THICK BACK WEB
(Courtesy of Westinghouse R&D Center)

These materials can be utilized as filters as shown in Exhibit A-1. With one end sealed, dusty gas can enter the element corrugations running parallel to the sealed face. The gas permeates the flat sheets, which separate the layers of corrugation, depositing the dust on the higher pressure side. The cleaned gas then flows out of the element through the corrugations (oriented perpendicular to the inlet channels). Accumulated dust within the filter can be blown out periodically (backwashing) by a burst of high pressure blow-back air from the clean side. This air flow back through the membrane filter lifts the cake off the wall and transports it out of the element. An alternative method is to initiate a high velocity through the dirty side channels to sweep accumulated dust from the wall.

The primary attraction of this configuration is the ability to pack a great deal of filter area in a very compact module; a most important consideration for pressurized high-temperature systems. The current estimate of the filter element dimensions are 12" x 12" x 4" overall with 0.060" thick filter flats. These dimensions are primarily set by constraints imposed by shrinkage during the firing step of fabrication. The separate ^{inlet} ribs are .15" high by 0.0625" thick and separate 0.458" wide channels. This configuration leads to 10-12 ft² of filter surface per element. At conservative filter velocities of 6-10 ft/min, each element can process ^{60 to 120} fl20 actual ft³/min. With a nominal pore size of 140 ^μ, filtration rates in this range lead to "conditioned" baseline pressure drops on the order of 5 to 10 in. Wg.

For full scale PFBC application, filtration modules, consisting of individual filter elements mounted on a common plenum are housed in a thermally insulated pressure vessel and integrated into the hot gas cleanup (HGC) system.

Exhibit A-3 depicts a CXF module composed of 40 filter elements.

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EXHIBIT A-3

CERAMIC CROSS-FLOW FILTRATION MODULE CONFIGURATION

Exhibit A-4 depicts the CXF unit utilized in each B vessel PFB combustor train for both the 1-cyclone/CXF/HGC and 2-cyclone/CXF/HGC steam-cooled PFBC plant concepts presented in Sections 4 and 5, respectively. Each of these CXF units comprises 18 modules housed in a nominally 16' diameter by 34' long thermally insulated pressure vessel, capable of handling over 100,000 acfm of flue gas. Correspondingly, each complementary A vessel PFB combustor train utilizes CXF units comprised of 14 modules housed in a nominally 14' diameter by 32' long insulated pressure vessel.

Periodic on line cleaning of the filter elements is accomplished by administering a burst of high pressure blow-back air down the clean side plenum of each module, via the pulse air pipe which penetrates the dome of the dome of the pressure vessel, as shown in Exhibit A-4. For a case in which the inlet dust concentration is 2000 ppm (typical of a primary cyclone outlet) the cycle time for blow-back is about 1-1/2 hr. At this time a single 0.1 sec. burst of about 2 kg of air from a 45 atm reservoir is used to clean one module. While one module is momentarily in the cleaning phase of its filtration cycle, the remaining modules continue to operate but with a slightly increased filter velocity due to the attendant momentary increase in filtration duty.

Based on preliminary CXF module performance and fabrication data provided by WRDC, CXF/HGC system related capital cost estimates were developed for the economic analyses of the 2-cyclone/CXF/HGC and 1-cyclone/CXF/HGC PFBC plants presented in Section 4 and 5, respectively.

Exhibit A-5 presents an installed cost summary of the CXF units and attendant major blow-back system components for the 2-cyclone and 1-cyclone/CXF/HGC systems.

Exhibit A-6 presents a summary of CXF module component weight and installed cost for the nominally 16' high module, consisting of 40 elements mounted on a common plenum, shown in Exhibit A-3.

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EXHIBIT A-4

NOMINAL 16 FT OD CERAMIC CROSS-FLOW FILTRATION UNIT ARRANGEMENT

EXHIBIT A-5

CXF/HGC SYSTEM INSTALLED COST SUMMARY FOR A NOMINAL 500 MW STEAM-COOLED PFBC COMBINED CYCLE PLANT (January 1980 \$1000)

<u>HGC System Configuration</u>	<u>2-Cyclone/CXF</u>	<u>1-Cyclone/CXF</u>
CXF Units:		
CXF Modules (128)	1,821,440	1,821,440
Pressure Vessels (8)	<u>3,273,760</u>	<u>3,273,760</u>
Subtotal	5,095,200	5,095,200
Blow-back System:		
Compressors	224,000	118,000
Interconnecting Piping	36,000	33,000
Heaters	100,000	100,000
Accumulators	32,800	32,800
316 SS Pulse Air Manifold	16,000	16,000
Miscellaneous Insulation	<u>12,800</u>	<u>12,800</u>
Subtotal	<u>421,600</u>	<u>312,600</u>
Total	5,516,800	5,407,800
Specific CXF/HGC Cost ¹ , \$/ft ²	90 to 108	88 to 106

(1) 51200 to 61440 ft² of filter surface area

EXHIBIT A-6

CXF MODULE WEIGHT AND COST SUMMARY

<u>Component</u>	<u>Material</u>	<u>Weight (lbs)</u>	<u>Cost (January 1980 dollars)</u>
Filter Elements ¹ (40)	Ceramic	500	4,760
Plenum	RA333	344	6,500
Clamps (40)	RA333	163	2,000
Mounting Hardware	Ti Alloy	<u>115</u>	<u>600</u>
Subtotal		1,122	13,860
Module Assembly and Installation			<u>370</u>
Total			14,230

(1) 10 to 12 ft² of filter surface area/element

Exhibit A-7 presents a summary of CXF unit housing weight and installed cost for the nominal 16' diameter by 32' long unit, shown in Exhibit A-4, which contains 18 modules and is utilized in each B vessel PFB combustor CXF/HGC train.

Exhibit A-8 presents a summary of CXF unit housing weight and installed cost for the nominal 14' diameter by 32' long unit, which contains 14 modules and is utilized in each A vessel PFB combustor CXF/HGC train. The 14' diameter unit closely approximates the 16' diameter unit shown in Exhibit A-4.

EXHIBIT A-7

CXF UNIT PRESSURE VESSEL HOUSING WEIGHT AND INSTALLED COST SUMMARY
FOR THE "B" VESSEL PFB COMBUSTOR CXF/HGC TRAIN

<u>Component</u>	<u>Material</u>	<u>Weight</u>	<u>Installed Cost (January 1980 dollars)</u>
Pressure Vessel	CS	100,000	230,100
1 In. Lining	Refractory	16,000	7,200
5 In. Lining	Insulation	23,100	17,300
Plenum Assembly	RA333	6,300	167,700
Pulse Air Pipes (18)	RA333	1,300	22,140
Solenoid Valves (18)	SS	<u>720</u>	<u>16,920</u>
		147,420	461,360

EXHIBIT A-8

CXF UNIT PRESSURE VESSEL HOUSING WEIGHT AND INSTALLED COST SUMMARY
FOR THE "A" VESSEL PFB COMBUSTOR CXF/HGC TRAIN

<u>Component</u>	<u>Material</u>	<u>Weight (lbs)</u>	<u>Installed Cost (January 1980 dollars)</u>
Pressure Vessel	CS	75,000	173,700
1 In. Lining	Refractory	13,200	5,900
5 In. Lining	Insulation	19,100	14,300
Plenum Assembly	RA333	5,000	132,800
Pulse Air Pipes (14)	RA333	1,000	17,200
Solenoid Valves (14)	SS	<u>560</u>	<u>13,160</u>
		113,860	357,080

3.6 ALKALI METAL REMOVAL

3.6.1 Objectives

The principal objectives in alkali metal removal are to:

- Characterize the background alkali concentration in prospective alkali getter materials, such as the Exxon 3⁰ cyclone PFBC ash and activated bauxite. In addition to these getters, the aluminum- and silicon-rich ceramic filter medium is included in this evaluation because of its potential gettering ability.
- Assess the potential release of the background alkali content from the prospective getter or filter materials as a function of time and reaction temperature. As a general premise, low, nonvolatile alkali media are essential for minimizing the concentration of released background alkali during the gettering reaction.
- Evaluate the ability of the Exxon 3⁰ cyclone PFBC ash, activated bauxite, and the ceramic filter medium to sorb sodium from sodium chloride (NaCl)-laden gases at reaction temperatures and pressures typifying hot-gas cleanup conditions. In addition to gettering capacity, the ceramic filter medium will be evaluated on its ability to withstand any deleterious effects that the reaction conditions or alkali have on the physical properties of the filter medium.

The following discussion is divided into three major sections that include the test facility (3.6.2), material characterization (3.6.3), and alkali gettering (3.6.4). In the test facility section we have provided a description of the bench-scale, high-pressure/high-temperature reactor, and the double-balance thermogravimetric analysis (TGA) system. While both units are frequently used in the alkali gettering evaluation, the bench-scale reactor served as the principal unit in this study. In the material characterization section, the physical and chemical composition of the Exxon 3⁰ cyclone PFBC ash, activated bauxite, and ceramic filter medium are defined, and the potential release of background alkali at the defined reaction conditions of 108 and 1200 K (1500 and 1700°F) at 1114 kPa (150 psig) total pressure is evaluated. In the alkali gettering discussion, data are presented that demonstrate the effectiveness of the Exxon 3⁰ cyclone PFBC ash, activated bauxite, and ceramic filter medium in reacting with

and retaining sodium. The alkali content within the various phases of the reacted materials is quantified by atomic absorption analysis. The following discussion summarizes additional alkali gettering data and the system performance projections generated in parallel, continuing alkali studies.

3.6.2 Test Facility

The principal test facility used in the alkali metal removal evaluation was the bench-scale, high-temperature and -pressure, horizontal flow reactor shown in Figure 3.6-1. This reactor consisted basically of two sections. The first was a gas preheater in which the carrier gas (argon and nitrogen) was heated and flowed over a boat containing molten alkali salt. By controlling temperature and gas flow, we could regulate the concentration of alkali vapor to the desired levels. In order to obtain a practical alkali flux from the boat, temperatures somewhat higher than the anticipated range of sorption temperatures were required in the preheat section. An extended section joining the preheat and reaction portions of the system was, therefore, provided, allowing a gentle moderation of gas temperature to the desired reaction temperature. Under these conditions the potential for alkali vapor to condense into a fog of small droplets was minimized.

The reaction zone was the final length of constant temperature tube into which the sample material was inserted. The Exxon 3⁰ cyclone PFBC ash and/or activated bauxite was pressed into 0.03 cm (0.125 in) thick × 1.3 cm (0.5 in) diameter discs and placed on ceramic filter discs of similar dimensions. The entire sample was then loaded into a quartz holder and inserted into the reactor perpendicular to the gas flow. The sample plug dimensions ensured permeation by the alkali-laden reactant gases of the ash and/or activated bauxite and the ceramic filter sample.

During the heating-up period, the ash getter and ceramic material was isolated from reactant gas by a ball valve in the cooled

Figure 3.6-1 Bench-Scale Reactor for the Alkali Metal Removal Evaluation

exit section. When the desired temperature profile was established in the reactor, the ball valve was opened and the plug advanced by a push rod that penetrated the pressure barrier by means of a Conax fitting. Reacted samples were withdrawn from the system in the reverse manner, which allowed the unit to be maintained at pressure and temperature through several runs.

A process flow diagram and a functional schematic of the system, instrumentation, and controls are presented as Figures 3.6-2 and 3.6-3. The bench-scale reactor had been reviewed by our Stored Energy Committee and approved for test operations. As part of the approval procedure, formal instructions were issued for normal and emergency conditions, together with reactor vessel purge and vacuum procedures for operating with combustible, toxic gases.

Subsequent to system fabrication and check-out, calibration runs using NaCl were carried out in the bench-scale system. The results obtained from five preliminary runs with NaCl indicated that the bulk gas downstream of the boat was supersaturated in all runs because the salt source was maintained at temperatures in excess of that of the downstream sorbent position. Test conditions for these runs were 1114 kPa (150 psig) pressure, with bulk gas temperatures ranging from 1003 to 1123 K (1346 to 1562^oF). In order to rectify this situation, additional calibration runs were conducted using a modified temperature distribution along the length of the reactor. The modified temperature distribution was such that the temperature at the platinum boat containing NaCl was the lowest in the system. Furthermore, the temperature of the bulk gas approaching the boat was set to approach the boat temperature. This latter temperature was maintained slightly higher to reduce the possibility that supersaturation would take place. Using this operating technique, we were able to obtain experimentally, on a reproducible basis, a NaCl concentration in the bulk gas stream that was in reasonable agreement with the vapor pressure curves for NaCl and its dimer. To compare the experimental measurements and the vapor pressure curve, we measured the boat temperature. On the

Figure 3.6-2 High-Temperature and -Pressure Alkali Removal Apparatus (Flow Schematic)

Figure 3.6-3 High-Temperature and -Pressure Alkali Removal Test Apparatus (Electrical Schematic)

basis of the above, a 1093 K (1508⁰F) boat temperature was selected for our initial runs with ash, activated bauxite, and ceramic filter sample in the reactor. The amount corresponded to a bulk gas alkali concentration of 0.1 g/Nm³ (80 ppm) NaCl for a bulk gas flow of 0.02 dm³/s (0.5 scfm) through the reactor. On the basis of the Argonne work,¹ this bulk gas concentration appeared to exceed, by at least one order of magnitude, that required for saturation of the pellet within the 1 hour reaction period studied.

The two-balance TGA system shown in Figures 3.6-4 and 3.6-5 had been designed in order to study effectively the rate of the alkali removal or gettering reaction under DOE Contracts DE-AC21-78ET11026² and DE-AC21-81MC16372.³ This system permits us to obtain precise data for both the weight gain of a small cylindrical [0.3 to 1.0 cm (0.125 to 0.375 in) diameter] pellet of the ash* or getter material and the quantity of alkali evaporated from the lower pan, which is directly related to the concentration of alkali in the bulk gas stream. This information has permitted a single-pellet model to be formulated in terms of mass transfer, chemical reaction, and heat transfer resistances. Figures 3.6-6 and 3.6-7 are schematics of the TGA gas flow and the furnace electrical and control systems as approved by the Westinghouse Stored Energy Committee. Figure 3.6-8 shows the laboratory support frame.

3.6.3 Material Characterization

3.6.3.1 Background Alkali Analysis for the Exxon 3⁰ Cyclone PFBC Ash, Activated Bauxite, and Ceramic Filter Medium

This section reviews the chemical analyses and particle size distribution data obtained for the initial, unreacted Exxon 3⁰ cyclone PFBC ash, activated bauxite, and ceramic filter materials prior to their exposure to alkali.

*Gasifier, boiler, or PFBC ash materials.

Figure 3.6-4 Two-Balance System for Kinetic Studies of Gettering Reactions

Figure 3.6-5 TGA Data Acquisition and Control System

Figure 3.6-6 Dual-Balance Thermogravimetric Analysis System (Flow Schematic)

Figure 3.6-7 Dual-Balance Thermogravimetric Analysis System (Reactor Vessel Electrical Schematic)

Figure 3.6-8 Dual-Balance Thermogravimetric Analysis System (Support Frame)

The particle size distribution for the Exxon 3⁰ cyclone PFBC ash was obtained by Coulter counter analysis and for activated bauxite, by sieve analysis (Table 3.6-1 and Figure 3.6-9). The physical morphology of the getter materials (PFBC ash and activated bauxite) and of the ceramic filter materials are shown in the scanning electron micrographs (SEM) of Figures 3.6-10 through 3.6-13. Energy dispersive X-ray analysis (EDAX) of select locations within each material is also presented.

The SEM area scan (Figure 3.6-10-A) of the Exxon 3⁰ cyclone PFBC ash illustrates the formation of fine ash agglomerates. At higher resolution (Figure 3.6-10-B-2.4) the agglomerate clusters are shown to contain localized regions rich in aluminum and silicon, with calcium, sulfur, magnesium, potassium, iron, and titanium present in minor concentrations, in addition to areas rich in either silicon and calcium (Figure 3.6-10-B-3) or in iron (Figure 3.6-10-B-1).

The micrograph of activated bauxite (Figure 3.6-11-A) indicates that in the as-received material* particles range from less than 6 μm to larger than 1 mm. EDAX confirms the presence of aluminum as the major constituent (Figure 3.6-11-B-1), but silicon and titanium can be detected at trace levels along the larger bauxite particle surfaces.

The porous Coors ceramic filter 427 as shown in cross-section (Figure 3.6-12-A) consists of an aluminum and silicon complex with trace concentrations of calcium, titanium, and iron present in weblike formations (Figure 3.6-12-B-1), in addition to flatter areas consisting only of silicon (Figure 3.6-12-B-2). The silicon-rich phase appears to predominate within the external edge of the filter (Figure 3.6-12-C-1), extending to depths of 200 μm .

Under the same magnification (Figure 3.6-13-A) the Coors ceramic filter 485 appears to be more porous than filter 427, consisting of flat

* Activated bauxite (20/60 Porocel HA 6% Volatile Material)

Table 3.6-1

PARTICLE SIZE DETERMINATIONS OF GETTER MATERIALS

As-Received Activated Bauxite Particle Size Determination by Sieve Analysis

Size	Weight Percent	Cumulative Weight Percent
> 2.36 mm	0	-
1180 - 2360 μm	10.2	10.2
500 - 1180 μm	46.9	57.1
250 - 500 μm	33.6	90.7
125 - 250 μm	5.4	96.1
63 - 125 μm	2.2	98.3
45 - 63 μm	1.3	99.6
< 45 μm	0.4	100.0

Coulter Counter Size Determination

Size, μm	Exxon 3° Cyclone PFBC Ash		Activated Bauxite	
	Percent	Cumulative Percent	Percent	Cumulative Percent
40.3	-	-	-	-
32.0	-	-	16.1	16.1
25.4	-	-	9.1	25.2
20.2	1.5	1.5	7.8	33.0
16.0	0.9	2.4	6.9	39.9
12.7	1.9	4.3	7.7	47.6
10.08	2.7	7.0	7.1	54.7
8.00	6.8	13.8	6.7	61.4
6.35	10.6	24.4	5.6	67.0
5.04	14.8	39.2	5.3	72.3
4.00	15.8	55.0	5.5	77.8
3.17	15.4	70.4	6.4	84.2
2.52	13.0	83.4	5.0	89.2
2.00	10.0	93.4	5.2	94.4
1.59	6.6	100.0	5.6	100.0
1.26	-	-	-	-

Figure 3.6-9 Particle Size Distribution of Exxon 3° Cyclone PFBC Ash and Activated Bauxite by Coulter Counter Determination

regimes rich in aluminum, silicon, and magnesium (Figure 3.6-13-B-1) that surround "knobbed" aluminum formations (Figure 3.6-13-B-2). The SEM mounting medium is distinctly evident in many of the filled filter pores (Figure 3.6-13-B-3), permitting only weak silicon signals to be detected from underlying surfaces. A definitive gradient cannot be observed in either the morphology or the composition of the filter 485 material.

Chemical analysis by emission spectroscopy and atomic absorption of the background elemental concentration within the getters and ceramic filters, and the SEM/EDAX and Coulter counter results, are summarized in Tables 3.6-2 through 3.6-5. The emission spectroscopy data semiquantitatively illustrates the variation in elemental concentration in the getter and filter materials (Table 3.6-2). Aluminum and silicon are once again detected as major constituents enriched in the filters;

FIG B, SPOT 1

FIG B, SPOT 2

FIG B, SPOT 3

FIG B, SPOT 4

Figure 3.6-10

Exxon 3° Cyclone PFBC Ash

FIG. B, SPOT 1

Figure 3.6-11 Activated Bauxite

Figure 3.6-12 Coors Ceramic Filter 427

Figure 3.6-13

Coors Ceramic Filter 485

Table 3.6-2

EMISSION SPECTROSCOPY ANALYSIS OF EXXON 3° CYCLONE ACTIVATED BAUXITE,
PFBC ASH, AND CERAMIC FILTER MATERIALS, wt%*

Element	Activated Bauxite	3° Cyclone Exxon Ash	Ceramic Filter #427	Ceramic Filter #485
Al	<10	<10	<10	<10
Ag	<0.001	<0.001	<0.001	<0.001
B	<0.001	0.05	0.01	0.005
Ba	0.002	0.05	0.02	0.02
Be	<0.0005	0.0005	<0.0005	<0.0005
Bi	<0.001	<0.001	<0.003	<0.003
Ca	0.003	>1	>1	>>1
Cd	<0.01	<0.01	<0.01	<0.01
Co	<0.001	0.001	<0.002	<0.002
Cr	0.02	0.02	0.003	0.003
Cu	<0.001	0.02	0.005	0.003
Fe	0.5	>1	0.3	0.3
Ga	0.008	0.003	0.005	0.005
Ge	<0.01	<0.01	<0.01	<0.01
K	0.008	>>1	0.5	0.5
Li	<0.003	0.007	0.007	0.006
Mg	0.001	>>1	>1	>>1
Mn	0.003	0.02	0.003	0.002
Mo	<0.001	0.004	<0.001	<0.001
Na	0.01	>1	0.2	0.2
Nb	0.005	<0.003	<0.005	<0.005
Ni	<0.003	0.02	0.003	0.003
Pb	0.002	0.007	<0.003	<0.003
Sb	<0.02	<0.02	<0.02	<0.02
Si	>1	>>1	>>1	>>1
Sn	0.003	<0.003	<0.003	<0.003
Sr	0.003	0.02	0.006	0.006
Ti	>1	1	0.5	0.5
V	0.02	0.02	0.004	0.004
W	<0.01	<0.01	<0.01	<0.01
Zn	<0.005	0.03	<0.003	<0.003
Zr	0.04	0.009	0.007	0.004

* Average of three samples.

Table 3.6-3
EXXON 3⁰ CYCLONE PFBC ASH CHARACTERIZATION

Material	Alkali Characterization, wt %						Total Carbon (wt %)	Emission Spectroscopy (> 1 wt %)	SEM/EDAX	Coulter Counter
	Soluble*		Insoluble		Total					
	Na	K	Na	K	Na	K				
PFBC Ash (1)	0.34	0.58	0.37	1.15	0.71	1.73	0.27	Majors: Ca, Fe, K, Mg, Na, Si, Ti, Al	Agglomerate fines rich in Al, Si with Ca, S, Mg, K, Ti	Fines less than 25.4 μm $d_p = 3.4 \mu\text{m}$ $\sigma = 1.78$
PFBC Ash (2)	<0.29	0.51	0.42	1.29	0.71	1.80	0.24			
PFBC Ash (3)	<0.30	0.52	0.38	1.17	<0.68	1.69	0.25		Present as minor elements	

* Water and acid soluble fractions

Table 3.6-4
ACTIVATED BAUXITE CHARACTERIZATION

Material	Alkali Characterization, wt %						Emission Spectroscopy (≥ 1 wt %)	SEM/EDAX	Coulter Counter
	Soluble*		Insoluble		Total				
	Na	K	Na	K	Na	K			
Activated Bauxite (1)	<0.07	0.06	0.06	0.05	<0.13	<0.11	Majors Si Ti Al	Particles Ranging from 6 μ m 1mm rich in Al Si Ti	Particle size of the as received material is too large for adequate coulter determination. Material was pulverized for 3.5 min in a shatter box and coulter counter analysis indicates that the resulting fines are less than 40.3 μ m almost equivalent to the PFBC ash size range $d_p = 10 \mu$ m $\sigma = 3.65$
(2)	0.07	0.06	0.09	0.05	<0.16	<0.11			
(3)	<0.07	<0.06	0.07	0.05	0.14	0.11			

* Water and acid soluble fractions

Table 3.6-5

FILTER CHARACTERIZATION

Material	Alkali Characterization, wt %						Emission Spectroscopy (> 1 wt %)	SEM/EDAX	
	Soluble*		Insoluble		Total				
	Na	K	Na	K	Na	K			
Ceramic Filter 427	(1)	<0.11	<0.06	0.10	0.12	<0.21	<0.18	Majors: Ca, Mg, Si, Al	Porous Al, Si structure with Ca, Ti, Fe present in trace concentrations
	(2)	<0.11	<0.06	0.12	0.13	<0.23	<0.19		
	(3)	<0.09	<0.07	0.14	0.14	<0.23	<0.21		
Ceramic Filter 485	(1)	0.15	0.08	0.05	0.19	0.20	0.27	Majors: Ca, Mg, Si, Al	Porous Al, Si Mg structure
	(2)	0.15	0.08	0.05	0.15	0.20	0.23		
	(3)	0.15	0.08	0.05	0.18	0.20	0.26		

* Water and acid soluble fractions

titanium in the activated bauxite; and calcium, iron, magnesium, and the alkalis in the Exxon 3^o cyclone PFBC ash.

In order to quantitatively determine the background alkali concentration in the getters and filters, three samples of each material were subjected to atomic absorption analysis by the following procedure. Each material is finely ground, leached with distilled water, and the sodium and potassium concentration in the filtrate determined. The alkali in this fraction is referred to as water soluble alkali. The solid residue that remains is then treated on a 1:1 volumetric basis with hydrochloric acid (HCl), heated for one hour at ~ 368 K (~ 203°F), filtered, and the alkali content determined and reported as the acid soluble phase. The remaining residue is then heated in alcohol, followed by the addition of hydrogen fluoride (HF) to the solution, and heated for one hour. Perchloric acid (HClO₄) is then added to the sample and the entire mixture brought to fuming. The alkali concentration determined in this fraction is designated as the acid insoluble alkali.

The atomic absorption analysis shows the background alkali concentration in activated bauxite to be almost equally divided between the soluble (both water and acid soluble) and insoluble phases, implying that approximately one-half of the total alkali content can be stripped from the aluminate lattice through the leaching procedure (Table 3.6-6). The alkali insoluble content, which is expected to have been bound within the aluminate matrix, either as ions embedded between the single or double octahedral aluminum sheets of the bauxite crystal or ionically substituted for hydrogen in the hydroxyl groups surrounding the aluminum ion in the octahedra, remains intact during the water and acid leaching. It is only through the addition of HClO_4 and HF that these alkalis are totally released from the aluminate lattice.

Unlike the equivalent partitioning of alkali between the soluble and insoluble phases in bauxite, the atomic absorption data of the Exxon 3⁰ cyclone PFBC ash indicate a considerably higher background alkali content of the insoluble complex (55 to 70% of the total background content). The acid soluble fraction contains 30 to 40% of the total alkali content, while the water soluble alkalis are present at 1 to 3% levels.

The atomic absorption data of the ceramic filters indicate that sodium exists preferentially within the acid insoluble matrix of filter 427 and within the acid soluble phase of 485, while potassium predominates within the acid insoluble fraction of both filter materials.

From a review of Table 3.6-6, it is apparent that the Exxon 3⁰ cyclone PFBC ash content has the highest background alkali of all the getter and filter materials considered. It is also evident that the Exxon 3⁰ cyclone PFBC ash soluble alkali content exceeds the total alkali content reported for bauxite or for the ceramic filters. When the Exxon 3⁰ cyclone ash is heated at PFBC temperatures and pressures, the soluble alkalis may either volatilize or be driven into the insoluble aluminosilicate matrix.

Table 3.6-6

CONCENTRATION OF ALKALI IN GETTER AND FILTER MATERIALS, wt%*

<u>Material</u>	<u>Sodium</u>				<u>Potassium</u>			
	<u>Water Soluble</u>	<u>Acid Soluble</u>	<u>Acid Insoluble</u>	<u>Total</u>	<u>Water Soluble</u>	<u>Acid Soluble</u>	<u>Acid Insoluble</u>	<u>Total</u>
Activated Bauxite	<0.02	<0.05	0.073	<0.143	<0.01	<0.05	<0.05	<0.110
3° Cyclone Exxon Ash	<0.02	0.290	0.390	<0.700	0.017	0.520	1.230	1.740
Ceramic Filter #427	<0.02	0.083	0.120	<0.223	0.013	<0.05	0.130	<0.193
Ceramic Filter #485	0.050	0.107	0.050	0.207	0.030	0.050	0.173	0.253

* Average of three samples.

3.6.3.2 Potential Alkali Release

In order to be considered as a candidate getter, any material that contains a substantial background alkali concentration must demonstrate minimal alkali release at reaction conditions.

The Exxon 3⁰ cyclone PFBC ash was, therefore, subjected to temperatures of 1088 and 1200 K (1500 and 1700⁰F) in argon at 1114 kPa (150 psig) pressure for 0.5, 1, and 4 hours. Table 3.6-7 summarizes the concentration of the alkali that results in the top, middle, and bottom layers of an ~ 40 mg Exxon 3⁰ cyclone PFBC ash sample plug. The background sodium and potassium did not volatilize at either temperature over the 4-hour test duration. Figure 3.6-14 illustrates the repartitioning of the alkali within the water soluble, acid soluble, and acid insoluble phases of the Exxon 3⁰ cyclone PFBC ash as a function of time and exposure temperature. Note that approximately 2.3 to 5.3 wt% of the background sodium and 2.4 to 3.9 wt% of the background potassium exist in the water soluble ash fraction, possibly as either an alkali chloride, carbonate, or water glass silicate or aluminate, while 42 to 54 wt% of the ash sodium, and 29 to 43 wt% of the ash potassium exists as acid soluble phases that may include the feldspar and nepheline structures. Approximately 43 to 55 wt% of the ash sodium and 59 to 69 wt% of the ash potassium exist within the insoluble aluminosilicate PFBC ash lattice. With the exception of the 0.5 hour data point, the alkali generally tends to migrate from the acid soluble to the acid insoluble ash fraction as a function of time, particularly in the lower 1088 K (1500⁰F) temperature exposure.

3.6.4 Alkali Gettering

3.6.4.1 Exxon 3⁰ Cyclone PFBC Ash, Activated Bauxite, and Ceramic Filter Medium Sodium Gettering

Preliminary data were obtained for the sodium getting of ~ 0.3 cm (0.125 in) Exxon 3⁰ cyclone PFBC ash cakes supported on ceramic filter discs exposed at temperatures of 108 and 1200 K (1500 and 1700⁰F)

Table 3.6-7

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EXXON 3⁰ CYCLONE PFBC ASH ALKALI RELEASE DATA

Alkali Characterization

Material	Exposure Time (Hrs)	Temp., K (°F)	Gauge Pressure (Atm)	Gas Composition	Velocity (cm/min)	Soluble		Insoluble		Total	
						Na	K	Na	K	Na	K
PFBC Ash	1/2	1088 (1500)	10	Ar	15	0.40	0.70	0.33	1.03	0.73	1.73
						0.39	0.71	0.33	0.94	0.72	1.65
						0.37	0.75	0.37	1.11	0.74	1.86
	1	1088 (1500)	10	Ar	15	0.37	0.70	0.35	1.10	0.72	1.80
						0.34	0.59	0.42	1.28	0.76	1.87
						0.39	0.71	0.35	1.09	0.74	1.80
	4	1200 (1700)	10	Ar	15	0.48	0.85	0.29	0.87	0.77	1.72
						0.38	0.75	0.34	0.94	0.72	1.69
						0.38	0.83	0.29	0.95	0.67	1.78
	1/2	1200 (1700)	10	Ar	15	0.34	0.63	0.38	1.13	0.72	1.76
						0.36	0.68	0.36	1.11	0.72	1.79
						0.36	0.65	0.38	1.16	0.74	1.81
	1	1200 (1700)	10	Ar	15	0.35	0.62	0.38	1.13	0.73	1.75
						0.31	0.53	0.41	1.28	0.72	1.81
						0.32	0.53	0.41	1.22	0.73	1.75
	4	1200 (1700)	10	Ar	15	0.36	0.69	0.34	1.13	0.70	1.82
						0.37	0.73	0.38	1.11	0.75	1.84
						0.39	0.74	0.34	1.06	0.73	1.80

Figure 3.6-14 Averaged Repartitioning of Sodium and Potassium within Water Soluble (WS) Acid Soluble (S) and Insoluble (I) Phases of Exxon 3° Cyclone Ash as a Function of Time

and 1013 kPa (136 psig) gauge pressure to an argon gas environment 0.1 g/Nm^3 (80 ppm) NaCl (Table 3.6-8). Comparison of the background alkali levels within the unreacted PFBC ash with atomic absorption data for the 1088 K (1500°F) gettered ash cake indicates that alkali gettering increases as a function of time (4.4-fold increase of total Na after 1 hour, 8.5-fold increase of total Na after 4 hours). The water soluble phase forms as the major gettered phase in the PFBC ash cake, with the formation of secondary acid soluble phases and minor contributions to the total gettered alkali concentration from the acid insoluble constituents.

Previous experimental programs have investigated the gettering ability of gasifier ash in similar alkali-laden environments (Table 3.6-9).² Comparison of the background alkali content of the Exxon 3° cyclone PFBC ash and gasifier ash materials indicates that both sodium and potassium are present at higher concentrations within the PFBC ash material. Comparison of the gettering ability of the PFBC ash cake and

Table 3.6-8

IXXON 3⁰ CYCLONE PFBC ASH GETTERING CAPACITY

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Material	Cake Thickness cm (in)	Exposure Time (Hrs)	Temp °C (°F)	Gauge Pressure (Atm)	Gas Composition	Velocity (cm/min)	Alkali Characterization wt %							
							Water Soluble		Acid Soluble		Acid Insoluble		Total	
							Na	K	Na	K	Na	K	Na	K
PFBC Ash (Average Background)	—	—	—	—	—	—	< 0.02	0.017	0.29	0.52	0.39	1.23	< 0.70	1.77
Ceramic Filter (Average Background)	—	—	—	—	—	—	0.05	0.03	0.107	0.05	0.05	0.173	0.207	0.253
PFBC Ash	0.31 (1/8)	1	816 (1500)	10	80 ppm Na/Ar	30	1.81	0.08	0.88	0.05	0.48	0.06	3.17	0.19
							0.64	0.009	0.91	0.03	0.52	0.03	2.07	0.069
							2.99	0.16	0.65	0.08	0.39	0.1	4.03	0.34
Ceramic Filter	—	1	816 (1500)	10	80 ppm Na/Ar	30	0.85	0.01	0.18	0.02	0.25	0.03	1.28	0.06
							0.89	0.03	0.15	0.02	0.23	0.05	1.27	0.10
PFBC Ash	0.31 (1/8)	4	816 (1500)	10	80 ppm Na/Ar	30	6.52	0.31	0.50	0.02	0.26	0.02	7.28	0.35
							5.11	0.28	0.57	0.01	0.11	0.02	5.79	0.31
							3.58	0.42	0.71	0.01	0.39	0.02	4.68	0.45
Ceramic Filter	—	4	816 (1500)	10	80 ppm Na/Ar	30	6.95	0.02	0.18	0.01	0.31	0.03	7.44	0.06
							0.29	0.005	0.16	0.01	0.31	0.02	0.76	0.035
PFBC Ash	0.31 (1/8)	1	927 (1700)	10	80 ppm Na/Ar	30	0.19	0.008	0.58	0.01	0.95	0.01	1.72	0.028
Ceramic Filter	—	1	927 (1700)	10	80 ppm Na/Ar	30	0.08	0.003	0.13	0.01	0.30	0.02	0.51	0.033
							0.87	0.01	0.15	0.02	0.38	0.03	1.40	0.06

Table 3.6-9

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GASIFIER ASH GETTERING CAPACITY

Material	Sample Size	Exposure Time (Hrs)	Getter Temp. °C (°F)	Gauge Pressure (ATM)	Gas Composition	Gas Flow Rate (scfm)	Alkali Characterization, wt %					
							Acid Soluble		Acid Insoluble		Total	
							Na	K	Na	K	Na	K
Gasifier Ash (Background)	—	—	—	—	—	—	0.12	0.11	0.25	1.42	0.37	1.53
							0.05	0.11	0.26	1.38	0.31	1.49
							0.06	0.12	0.27	1.35	0.33	1.47
							0.06	0.11	0.25	1.38	0.31	1.49
							0.07	0.15	0.22	1.34	0.29	1.49
							0.07	0.14	0.22	1.21	0.29	1.35
Gasifier Ash	1/8" Pellets	1	845 (1553)	10	80 ppm Na/Ar	0.5	0.23	0.09	0.37	0.90	1.10	0.99
							0.15	0.10	0.55	0.94	0.70	1.04
							0.19	0.07	0.74	0.79	0.93	0.86
							0.13	0.09	0.65	0.90	0.78	0.99
Gasifier Ash	1/8" Pellets	4	850 (1562)	10	80 ppm Na/Ar	0.5	0.44	0.15	2.16	1.18	2.50	1.33
							0.64	0.10	2.12	0.73	2.76	0.83
							1.47	0.03	3.05	0.12	4.52	0.15
							0.62	0.05	2.12	0.39	2.74	0.44
Gasifier Ash	1/8" Pellets	1	900 (1652)	10	80 ppm Na/Ar	0.5	0.17	0.09	1.00	0.85	1.17	0.94
							0.15	0.05	0.88	0.95	1.03	1.00
							0.18	0.06	1.22	0.86	1.40	0.92

gasifier ash pellet shows that instead of a 4.4-fold increase of Na over background resulting at 1088 K (1500°F) in the PFBC cake, a 2.8-fold increase of Na results within the gasifier ash pellet after 1 hour at 1118 K (1553°F). An 8.5- and 9.9-fold increase in Na over background is noted in the PFBC ash cake and gasifier ash pellet at the respective temperatures after four hours. In contrast to the 1088 K (1500°F) PFBC ash, the predominant gettered phase formed after 1 and 4 hours in the 1118 K (1553°F) gasifier ash pellet is that of the acid insoluble constituent. The gasifier ash is expected to have experienced temperatures within the PDU higher than those experienced by the Exxon 3° cyclone PFBC ash. As a result, the aluminosilicate lattice may have been sufficiently disoriented in the gasifier ash, producing a greater number of active sites that promote reaction within the insoluble claylike lattice. This conclusion is supported by X-ray diffraction data that identify the amorphous phase as the principal constituent of the gasifier ash, while calcium is the major constituent in the Exxon 3° cyclone PFBC ash, with quartz present only in minor concentrations (Table 3.6-10).

When the Exxon 3° cyclone PFBC ash reacts with gaseous NaCl at temperatures of 1200 K (1700°F) for 1 hour, there is only a 2.5-fold Na increase over background. The 4.4-fold Na increase at 1088 K (1500°F) after 1 hour, when compared to the 1200 K (1700°F) data, shows enrichment of sodium in the insoluble fraction, with lower alkali gettering by the water and acid soluble fractions at the higher reaction temperature. Sodium is apparently being driven into the claylike lattice, with alkali volatilizing simultaneously from the water and acid soluble phases, resulting in an overall lower gettered alkali content at higher reaction temperatures. The lower reaction temperature may have permitted condensation of gaseous NaCl along the ash-cake surface, promoting reactions of the alkali with the acid soluble lattice.

The sodium gettering capacities of the Exxon 3° cyclone PFBC ash cake and the PDU gasifier ash pellets are summarized in Table 3.6-11. A general trend relating gettering of the ash forms (powder vs pellet) as

Table 3.6-10

ELEMENTAL ANALYSIS OF ASH FROM AFBC, PFBC, AND GASIFICATION PROCESSES

Element	AFBC PROCESSES				PFBC PROCESSES						GASIFICATION PROCESSES	
	MERC Bed ppm	MERC Secondary Cyclone ppm	MERC Bag Filter ppm	CPU Baghouse ppm	Leatherhead 1 st Ash ppm	Leatherhead 2 nd Ash ppm	Exxon (67) 3 rd Cyclone Ash ppm	Exxon (73) Res ppm	Exxon (73) 2 nd Cyclone ppm	Exxon (73) 3 rd Cyclone ppm	Pittsburgh Ash Wt %	Minnehaha Char Wt %
Al	Major	Major	Major	Major	3.8%	9.94%	25.4%	1.33%	3%	5.89%	Major	Major
Ag	<1	<1	<1	<1	<1	<1	<1	1	1	1	<0.001	0.001
As												
B	50	500	500	500	200	200	300	200	200	200	0.003	0.004
Ba							500					
Be	10	10	10		2	10	15	1	2	5	0.0005	0.0005
Br	<10	<10	<10	<10	<1	<1	<1	<1	<1	<1	<0.001	0.001
Ca	39%	18%	16%	16%	14.56%	3.84%	7.2%	22.56%	18.48%	16.24%	>1.0	0.1
Cd	<3	<3	<3	<3	<33	<33	3	<33	33	<33	<0.003	<0.003
Co	5	15	15	15	10	20	10	<5	10	10	0.001	0.001
Cr	60	100	100	100	120	190	200	150	150	350	0.003	0.002
Cu	30	100	100	100	20	40	30	10	10	20	0.003	0.001
Fe	Major	Major	Major	Major	5.7%	8.93%	5.53%	2.53%	4.6%	3.66%	>1.0	>1.0
Ga					10	20		3	10	10	0.001	<0.001
Ge					10	30		20	10	20	<0.001	<0.001
Hg												
K					0.83%	2.3%	1.16%	0.19%	0.42%	1.01%	2.485	0.434
Mg	>1000	>1000	>1000	8.7%	8.16%	2.2%	3.84%	12.33%	11.18%	7.72%	0.05	0.01
Mn	190	220	280	220	200	200	330	100	100	100	0.02	0.002
Mo	30	10	10	50	<30	<30	30	<30	<30	<30	0.001	0.001
Na	>1000	>1000	>1000	>1000	0.14%	0.39%	0.59%	0.06%	0.14%	0.29%	0.104	0.064
Ni	40	70	70	80	40	80	150	25	25	100	0.002	0.001
Pb	<10	100	130	50	60	80	30	30	50	60	<0.001	<0.001
Sb	<33	<33	<33	<33	<50	50	33	50	50	<50	<0.003	<0.003
Se												
Si	Major	Major	Major	Major	8.03%	19.21%	15.4%	3.21%	5.91%	9.63%	>1.0	>1.0
Sn	<10	<10	<10	<10	<10	<10	<5	30	<10	10	<0.003	0.003
Sr	500	1000	1000	100	100	150		150	150	300	0.01	0.004
Ti	>1000	>1000	>1000	>1000	>1000	>1000	>1000	300	1000	1000	0.2	0.07
V	40	100	80	60	60	150	200	40	50	90	0.006	0.003
Zn	<33	100	100	150	50	50	500	<50	<50	50	<0.003	<0.003
Zr					100	100	500	<10	30	50	<0.03	0.01
Compound												
CaSO ₄	Major	Major			Major	Minor		Major	Major	Major		
CaO		Low Minor	Trace					Possible Trace				
CaCO ₃					Low Minor				Trace	Trace		
MgO					Minor			High Minor	High Minor	Minor		
SiO ₂ (Quartz)	Major	Major			Minor	Major			Minor	Low Minor	Secondary	Secondary
α-Fe ₂ O ₃	Minor	Minor			Trace	Low Minor			Trace			
FeS												Minor
Spinel						Low Minor						
Amorphous											Major	Major

a function of temperature and time is not apparent from the limited data. Gettering by both materials after 1 hour indicates that less than a 5-fold increase over background sodium results, while reaction after 4 hours in similar gas environments shows a less than 10-fold sodium increase over background in both materials. Plotting the average gettered alkali concentration as a function of time for the Exxon 3⁰ cyclone PFBC ash and gasifier ash (Figure 3.6-15) shows that the gettering capacity is dependent primarily on the material's physical and chemical composition.

As previously demonstrated with the gasifier ash,² the gettering of sodium by Exxon 3⁰ cyclone PFBC ash enhances the release of potassium bound initially in the ash. Although the scatter in the current data does not reveal a definite trend, potassium appears to be released from the insoluble phase initially in an ionic exchange with sodium with the recapture of released insoluble potassium by the soluble fractions as the 1088 K (1500°F) reaction time is extended to 4 hours. During sodium gettering at 1200 K (1700°F), 98% of the background potassium was released from the PFBC ash, with depletion occurring from all phases.

The ceramic filters used to support the Exxon 3⁰ cyclone PFBC ash cake act also as sodium gettering agents. The SEM/EDAX (Figures 3.6-12 and 3.6-13) and atomic absorption analyses (Table 3.6-5) of the ceramic filter materials indicate that the high silicon and aluminum matrix is virtually free of alkali. Characterization of the ceramic filter material after exposure to argon gases containing approximately 0.1 g/Nm³ (80 ppm) NaCl at 1114 kPa (150 psig) total pressure for 1 and 4 hour periods at reaction temperatures of 1088 K (1500°F) indicates that alkali gettering results primarily as a condensed water soluble phase (Table 3.6-12). Only a slight increase in the sodium acid soluble phase is detected over the 4 hour reaction period, while a 10- to 16-fold increase in the amorphous acid insoluble sodium content is detected over background.

Table 6.3-11

RATIO OF GETTERED SODIUM TO BACKGROUND SODIUM IN EXXON 3^o CYCLONE
PFBC ASH CAKES AND GASIFIER ASH PELLETS

Material	Time, hr	Temperature, K(°F)	Ratio
Exxon 3 ^o Cyclone	1	1089 (1500)	4.4
PFBC Ash Cake	4	1089 (1500)	8.5
	1	1200 (1700)	2.4
PDU Gasifier	1	1118 (1553)	2.8
Ash Pellets	4	1123 (1562)	9.9
	1	1173 (1652)	3.8

Figure 3.6-15 -- Average Sodium Concentration in Getter Material as a
Function of Time

Table 3.6-12

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CFRAMEIC FILTER GETTERING CAPACITY

Alkali Characterization

Material	Exposure Time (Hrs)	Temp. K (°F)	Gauge Pressure (Atm)	Gas Composition	Velocity (cm/min)	Water Soluble		Acid Soluble		Acid Insoluble		Total	
						Na	K	Na	K	Na	K	Na	K
Ceramic Filter	NA	NA	NA	NA	NA	<0.01	<0.01	0.09	0.02	<0.01	0.17	<0.11	<0.20
						<0.01	<0.01	0.10	0.02	<0.01	0.16	<0.12	0.19
						<0.01	<0.01	0.09	0.02	<0.01	0.18	<0.11	<0.21
	1	1088 (1500)	10	80 ppm Na/Ar	30	8.09	0.03	0.15	0.01	0.13	0.04	8.37	0.08
						7.50	0.05	0.13	0.02	0.09	0.05	7.72	0.12
						7.43	0.06	0.14	0.02	0.09	0.05	7.66	0.13
	4	1088 (1500)	10	80 ppm Na/Ar	30	2.98	0.01	0.17	0.02	0.14	0.04	3.29	0.07
						3.00	0.01	0.14	0.01	0.20	0.05	3.34	0.07
						3.09	0.05	0.13	0.02	0.13	0.08	3.35	0.15
	1	1200 (1700)	10	80 ppm Na/Ar	30	0.35	<0.01	0.23	0.02	0.22	0.02	0.80	<0.05
						0.19	<0.01	0.24	0.02	0.18	0.02	0.61	<0.05
						0.32	<0.01	0.25	0.01	0.22	0.02	0.79	<0.04
	4	1200 (1700)	10	80 ppm Na/Ar	30	0.02	<0.01	0.22	<0.01	0.19	<0.01	1.23	<0.03
						0.02	<0.01	0.24	<0.01	0.10	<0.01	1.06	<0.03
						0.02	<0.01	0.28	<0.01	0.14	<0.01	1.24	<0.03

NA = Not Applicable

Under similar test conditions the total gettering capacity of the filter material decreases as the reaction temperature is increased to 1200 K (1700°F). Although at the higher temperature NaCl is expected to remain in the gas phase, minimizing the concentration of condensed sodium in the water soluble fraction along the filter surface, reaction of the alkali with the insoluble filter lattice increases (20-fold increase after 1 hour, 100-fold increase after 4 hours over background). An approximate 3-fold increase is also detected over the background content. From these data, migration of sodium from the water soluble into the acid soluble and finally into the acid insoluble phase appears to depend on both time and reaction temperature.

As sodium is gettered from the gas stream by the ceramic filter, potassium is simultaneously released from the getter lattice. At 1289 K (1500°F) approximately one-third of the background insoluble potassium remains within the insoluble fraction after sodium gettering, while one-third is repartitioned into the water soluble phase and the remainder released into the gas stream. Once a minimum concentration of insoluble potassium has been achieved within the filter material at a particular reaction temperature, it no longer repartitions into the various alkali phases or volatilizes from the lattice as a function of reaction time. This phenomenon is in contrast to the water soluble gettered alkali fraction activity, which continues to volatilize from the filter lattice as a function of time.

Exposure of the ceramic filter to 1200 K (1700°F) further depletes the insoluble potassium concentration relative to background content. Recapture of the released insoluble potassium into either the water soluble or the acid soluble fraction does not occur at 1200 K (1700°F). The release of background potassium coupled with sodium gettering in the ceramic filter material generally increases as a function of both time and reaction temperature.

A final series of twelve test runs were conducted in the bench-scale reactor (Figure 3.6-1) using 0.3 cm (0.125 in) cylindrical

activated bauxite or Exxon 3^o cyclone PFBC ash pellets without the ceramic filter support. Each getter material was exposed to gaseous sodium chloride for 1, 3, and 6 hours at getter temperatures of 1123 and 1153 K (1562 and 1616^oF), with the NaCl source temperature at 1088 and 1103 K (1499 and 1526^oF). This corresponds to gaseous NaCl concentrations of approximately 0.005 and 0.15 g/Nm³ (11 and 115 ppm) at a total system pressure of 1114 kPa (150 psig).

Activated bauxite, which initially contains a slightly higher background level of sodium than of potassium (0.07 vs 0.05 wt%), principally in the acid insoluble, claylike fraction, getters sodium within the water and acid soluble and the acid insoluble phases at reaction temperatures of 1123 and 1153 K (1562 and 1616^oF) (Table 3.6-13). The concentration of sodium in the water soluble phase is consistently maintained at an average of 0.07 to 0.08 wt% within the 1123 K (1562^oF) activated bauxite pellets throughout the 1 to 6 hour gaseous NaCl exposures. The water soluble phase concentration varies widely, however, <0.01 to 0.23 wt% in the 1153 K (1616^oF) pellets.

The concentration of sodium in the acid soluble and insoluble phases of the 1123 K (1562^oF) activated bauxite pellets continuously increases with exposure time from about 0.27 to 0.53 wt% (Figure 3.6-16). Approximately 73% of the total gettered sodium content is partitioned within the acid insoluble lattice after 1 hour and is maintained at this level throughout the remainder of the gettering reaction. Although activated bauxite has the ability to getter 0.72 wt% sodium after 6 hours at 1123 K (1562^oF), the data indicate that the principal gettering reaction of sodium is with the insoluble activated bauxite lattice.

Within the first 3 hours of gaseous NaCl exposure at 1153 K (1616^oF), the concentration of sodium in all three phases of activated bauxite has increased to concentrations greater than those obtained in the 1123 K (1562^oF) reaction. The atomic absorption data indicate that the principal gettering reaction at 1153 K (1616^oF) is once again through reaction of the alkali with the insoluble lattice (Figure 3.6-17).

Table 3.6-13

SUMMARY OF BACKGROUND AND BENCH-SCALE EXPERIMENTAL GETTERING
DATA FOR REACTIONS OF ACTIVATED BAUXITE AND EXXON 3⁰ CYCLONE PFBC
ASH WITH GASEOUS SODIUM CHLORIDE

T. T. PARAMETER										ATOMIC ABSORPT. UNIT									
EST. NO.	MATERIAL	CLAS. COMPOSITION	AN. ON. AS. AC. N.	PR. LAZ. A. N.	PART. S. C. SIZE (MICRONS)	T. ER. TEMP. (°C)	HAL. TEMP. (°C)	EXPOS. RE. TIME (HRS)	WATER VOLUME (ML)	WATER VOLUME (ML)	ACTIV. VOLUME (ML)	ACTIV. VOLUME (ML)	INSOLUBLE (ML)	INSOLUBLE (ML)	TOTAL (ML)	TOTAL (ML)	TOTAL (ML)	TOTAL (ML)	TOTAL (ML)
BACKGROUND DATA																			
	BAUXITE								< 0.02	< 0.01	< 0.04	< 0.05	0.26	< 0.04	0.26	< 0.04	< 0.04	< 0.04	< 0.04
	PFBC								< 0.02	< 0.01	< 0.04	< 0.05	0.09	< 0.04	0.09	< 0.04	< 0.04	< 0.04	< 0.04
	ASP								< 0.02	< 0.01	< 0.04	< 0.05	0.17	< 0.04	0.17	< 0.04	< 0.04	< 0.04	< 0.04
PRELIMINARY BENCH-SCALE DATA - TEMPERATURE DEPENDENCE																			
1A	BAUXITE (20 g)	AA	0.5	11	1/8	1123	1082	1	0.09	< 0.03	0.04	< 0.02	0.31	< 0.03	0.31	< 0.03	< 0.03	< 0.03	< 0.03
B	BAUXITE (10 g)	AA	0.5	11	1/8	1123	1082	1	0.07	< 0.03	0.02	< 0.02	0.28	< 0.03	0.28	< 0.03	< 0.03	< 0.03	< 0.03
1C	BAUXITE (20 g)	AA	0.5	1	1/8	1123	1082	1	0.07	< 0.03	0.02	< 0.02	0.31	< 0.03	0.31	< 0.03	< 0.03	< 0.03	< 0.03
2A	BAUXITE (10 g)	AP	0.5	11	1/8	1123	1082	1	0.06	< 0.03	0.04	< 0.02	0.31	< 0.03	0.31	< 0.03	< 0.03	< 0.03	< 0.03
B	BAUXITE (10 g)	AA	0.5	11	1/8	1123	1082	1	0.04	< 0.03	0.10	< 0.02	0.31	< 0.03	0.31	< 0.03	< 0.03	< 0.03	< 0.03
2C	BAUXITE (20 g)	AA	0.5	1	1/8	1123	1082	1	0.07	< 0.03	0.02	< 0.02	0.31	< 0.03	0.31	< 0.03	< 0.03	< 0.03	< 0.03
3A	BAUXITE (10 g)	AP	0.5	1	1/8	1123	1082	1	0.09	< 0.03	0.11	< 0.02	0.32	< 0.03	0.32	< 0.03	< 0.03	< 0.03	< 0.03
1B	BAUXITE (10 g)	AA	0.5	1	1/8	1123	1082	1	0.07	< 0.03	0.07	< 0.02	0.32	< 0.03	0.32	< 0.03	< 0.03	< 0.03	< 0.03
1	BAUXITE (10 g)	AA	0.5	1	1/8	1123	1082	1	0.06	< 0.03	0.12	< 0.02	0.31	< 0.03	0.31	< 0.03	< 0.03	< 0.03	< 0.03
4A	BAUXITE (10 g)	AA	0.5	11	1/8	1153	1103	1	0.08	< 0.03	0.12	< 0.02	0.32	< 0.03	0.32	< 0.03	< 0.03	< 0.03	< 0.03
4B	BAUXITE (20 g)	AA	0.5	11	1/8	1153	1103	1	0.18	< 0.03	0.15	< 0.02	0.45	< 0.03	0.45	< 0.03	< 0.03	< 0.03	< 0.03
4	BAUXITE (20 g)	AA	0.5	1	1/8	1153	1103	1	0.05	< 0.03	0.13	< 0.02	0.46	< 0.03	0.46	< 0.03	< 0.03	< 0.03	< 0.03
5A	BAUXITE (10 g)	AA	0.5	11	1/8	1153	1103	1	0.04	< 0.03	0.13	< 0.02	0.46	< 0.03	0.46	< 0.03	< 0.03	< 0.03	< 0.03
5B	BAUXITE (10 g)	AA	0.5	11	1/8	1153	1103	1	0.13	< 0.03	0.29	< 0.02	0.46	< 0.03	0.46	< 0.03	< 0.03	< 0.03	< 0.03
5	BAUXITE (10 g)	AA	0.5	11	1/8	1153	1103	1	0.04	< 0.03	0.11	< 0.02	0.46	< 0.03	0.46	< 0.03	< 0.03	< 0.03	< 0.03
6A	BAUXITE (20 g)	AA	0.5	11	1/8	1153	1103	1	0.21	< 0.03	0.04	< 0.02	0.35	< 0.03	0.35	< 0.03	< 0.03	< 0.03	< 0.03
6B	BAUXITE (20 g)	AA	0.5	11	1/8	1153	1103	1	< 0.01	< 0.03	< 0.02	< 0.02	0.35	< 0.03	0.35	< 0.03	< 0.03	< 0.03	< 0.03
6C	BAUXITE (10 g)	AA	0.5	11	1/8	1153	1103	1	< 0.01	< 0.03	< 0.02	< 0.02	0.07	< 0.03	0.07	< 0.03	< 0.03	< 0.03	< 0.03
6D	BAUXITE (20 g)	AA	0.5	11	1/8	1153	1103	1	0.02	< 0.03	< 0.02	< 0.02	0.08	< 0.03	0.10	< 0.03	< 0.03	< 0.03	< 0.03
6E	BAUXITE (10 g)	AA	0.5	11	1/8	1153	1103	1	0.11	< 0.03	< 0.02	< 0.02	0.17	< 0.03	0.17	< 0.03	< 0.03	< 0.03	< 0.03
7A	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.12	< 0.03	1.70	0.12	0.8	0.09	2.61	< 0.03	< 0.03	< 0.03	< 0.03
7B	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.89	0.03	1.28	0.04	1.67	0.04	3.64	< 0.03	< 0.03	< 0.03	< 0.03
7C	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.08	< 0.03	1.13	0.29	0.74	0.29	1.95	< 0.03	< 0.03	< 0.03	< 0.03
8A	ASH (10 g)	AP	0.5	11	1/8	1123	1082	1	0.08	< 0.03	1.36	< 0.03	1.03	< 0.03	2.39	< 0.03	< 0.03	< 0.03	< 0.03
8B	ASH (10 g)	AP	0.5	11	1/8	1123	1082	1	0.07	< 0.03	0.98	< 0.03	1.59	< 0.03	2.57	< 0.03	< 0.03	< 0.03	< 0.03
8C	ASH (10 g)	AP	0.5	11	1/8	1123	1082	1	0.09	< 0.03	1.68	0.03	1.17	< 0.03	2.85	< 0.03	< 0.03	< 0.03	< 0.03
9A	ASH (10 g)	AA	0.5	11	1/8	1123	1082	1	5.34	0.03	4.31	0.03	0.39	0.03	10.04	< 0.03	< 0.03	< 0.03	< 0.03
9B	ASH (10 g)	AA	0.5	11	1/8	1123	1082	1	16.28	0.03	4.62	0.02	0.40	< 0.03	21.30	< 0.03	< 0.03	< 0.03	< 0.03
9C	ASH (10 g)	AA	0.5	11	1/8	1123	1082	1	4.78	0.04	4.66	0.03	0.42	< 0.03	19.86	< 0.03	< 0.03	< 0.03	< 0.03
10A	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.52	< 0.03	1.04	0.07	0.41	0.13	2.97	< 0.03	< 0.03	< 0.03	< 0.03
10B	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.05	< 0.03	0.97	0.19	0.88	0.52	1.94	< 0.03	< 0.03	< 0.03	< 0.03
10C	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.11	< 0.03	1.33	0.13	0.74	0.26	2.36	< 0.03	< 0.03	< 0.03	< 0.03
11A	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.32	< 0.03	4.03	< 0.03	1.1	< 0.03	5.46	< 0.03	< 0.03	< 0.03	< 0.03
11B	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	0.82	< 0.03	1.92	< 0.03	1.60	< 0.03	4.34	< 0.03	< 0.03	< 0.03	< 0.03
11C	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	1.34	0.04	3.59	0.04	0.70	< 0.03	7.85	< 0.03	< 0.03	< 0.03	< 0.03
12A	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	2.34	0.07	1.26	0.03	0.36	< 0.03	3.99	< 0.03	< 0.03	< 0.03	< 0.03
12B	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	12.75	< 0.03	1.38	0.02	0.87	< 0.03	14.58	< 0.03	< 0.03	< 0.03	< 0.03
12C	ASH (10 g)	AA	0.5	11	1/8	1153	1103	1	2.0	< 0.03	1.83	0.02	0.71	< 0.03	7.64	< 0.03	< 0.03	< 0.03	< 0.03

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Dug 5596C.62

Figure 3.6-16 Partitioning of Sodium in the 1123 K (1562°F) Getter Reaction

Figure 3.6-17 Partitioning of Sodium in the 1153 K (1616°F) Getter Reaction

A maximum of 0.50 wt% insoluble sodium is established within the pellet after 1 hour of reaction at 1153 K (1616°F), comparable to the concentration detected in the 1123 K (1562°F) insoluble activated bauxite phase after 6 hours of alkali exposure. This level is maintained after 3 hours of exposure to the 0.1 g/Nm³ (80 ppm) NaCl, 1153 K (1616°F) gases, tentatively indicating that maximum saturation or available active sites for sodium within the insoluble activated bauxite lattice has been achieved. Once the insoluble sites have been saturated, gettering continues to occur within the acid soluble fraction of the getter material. The sodium concentration within the soluble phase is enhanced in the 1153 K (1616°F) vs 1123 K (1562°F) reaction, because of saturation of the insoluble active sites at the higher temperatures. Chemical analysis of the 1123 K (1562°F) reacted Exxon 3° cyclone PFBC ash indicates that gettering begins through interaction of sodium with both the feldspar and aluminosilicate lattice (Figure 3.6-16). Saturation of the available insoluble sites within the PFBC ash appears to have occurred after 3 or 4 hours at a concentration of about 1.2 wt%, prior to what appears to be the transformation of gettered alkali from the insoluble aluminosilicate phase into the acid soluble fraction. Saturation of the acid soluble phase appears also to have been achieved within the first few hours of the reaction. With release from the insoluble fraction, possibly through a structural rearrangement within the PFBC ash lattice, an increased rate is established for gettering by the acid soluble complex, as indicated in Figure 3.6.16.

Gettering of 15 wt% sodium takes place within the water soluble phase after 6 hours of reaction. Although this represents 75% of the total gettered sodium in the PFBC material at 1123 K (1562°F), saturation of the 0.3 cm (0.125 in) pellets was not yet established. As the water soluble alkali phase accumulates in the pellet, both acid soluble and insoluble reactions are suppressed.

Atomic absorption analysis of PFBC ash pellets exposed to 0.1 g/Nm³ (80 ppm) NaCl at 1153 (1616°F) indicates a gettering reaction pattern similar to that of the 1123 K (1562°F) reaction (Figure 3.6-17).

Gettering begins in the acid soluble and insoluble phase but after extended exposure periods shifts principally to the water soluble phase. Saturation appears to have been achieved within the insoluble ash lattice after 1 to 3 hours of reaction at concentrations of 1 to 1.6 wt%, with sodium maintained as a stable phase as a function of time. Similarly, saturation of the active sites within the acid soluble fraction appears to have occurred after 3 hours of reaction at concentrations of 2 to 4 wt%. As both the insoluble and acid soluble sites attain saturation, additional sodium gettering is predominantly through the water soluble fraction. After 6 hours of exposure to the alkali-laden gas at 1153 K (1616°F), 75% of the total gettered sodium content of the 0.3 cm (0.125 in) PFBC ash pellet is within the water soluble fraction, and the pellet has reached very high levels of sodium, on the order of 14 to 17 wt%.

Table 3.6-14 indicates the percent of background potassium released from the Exxon 3⁰ cyclone PFBC ash material as a function of reaction temperature and time during sodium gettering. Release is predominantly from potassium bound within the insoluble aluminosilicate structure and, secondarily, from the potassium acid soluble or feldspar complexes. Both reaction temperatures indicate virtually 100% release of the background potassium content after 6 hours in the sodium environment.

Evolution of potassium from the Exxon 3⁰ cyclone PFBC ash insoluble phase in the presence of sodium has been thought to be an ionic exchange mechanism. The ratio of sodium gettering within the soluble lattice to potassium released from the same lattice indicates that for every atom of potassium released, one atom of sodium is repositioned within the PFBC ash lattice (Table 3.6-15). The simple ionic exchange between the alkalis is not evident in the water or acid soluble phases.

An important feature of the Exxon 3⁰ cyclone PFBC ash experiments was the fact that background potassium monitored throughout the

Table 3.6-14
POTASSIUM RELEASE FROM EXXON 3⁰ CYCLONE PFBC ASH

Reaction Conditions		Time, hr	Percent K Release	Weight Percent K
Temp., K (°F)				
1123	(1562)	1	82%	(1.43)
1123	(1562)	3	99%	(1.72)
1123	(1562)	6	96%	(1.67)
1153	(1616)	1	71%	(1.24)
1153	(1616)	3	98%	(1.71)
1153	(1616)	6	98%	(1.71)

Table 3.6-15
RATIO OF SODIUM GETTERING TO POTASSIUM RELEASED IN EXXON 3⁰ PFBC ASH

Reaction Conditions		Time, hr	Percent Insoluble Na Gained	Percent Insoluble K Released	$\frac{\text{Na}}{\text{K}}$
Temp., K (°F)					
1123	(1562)	1	0.68	1.06	0.64
1123	(1562)	3	0.87	1.20	0.72
1153	(1616)	1	0.62	0.90	0.68
1153	(1616)	3	0.75	1.20	0.62
1153	(1616)	6	0.75*	1.20	0.63

radius of Na⁺ = 0.97⁰A

radius of K⁺ = 1.33⁰A

Na⁺/K⁺ = 0.73

tests indicated that virtually all the potassium was lost during sodium sorption. We believe this is a solid indication that the concept of using the Exxon 3^o cyclone PFBC ash as a gettering agent will not work. Background levels of potassium in the bauxite were so low that rerelease of potassium could not be determined accurately.

Although the gettering work supported by this program is essentially complete with the data and discussion presented above, other Westinghouse programs have continued to conduct alkali gettering experiments in both the bench-scale and TGA apparatus. These studies have included the reaction of activated bauxite with gaseous potassium chloride (KCl), the performance of alternative getter material, and the effects of water vapor on gettering reactivity. The kinetic data available were used to generate a model that estimates the size requirements of a full-scale unit and to provide projections of system performance. The results of these efforts are presented in the following section.

3.6.4.2 Comparison of Sodium and Potassium Gettering in Activated Bauxite

Bench-scale studies³ conducted under DOE Contract DE-AC21-81MC16372 exposed 0.3 cm (0.125 in) cylindrical activated bauxite pellets to an argon environment containing 0.1 g/Nm³ (80 ppm) KCl for the duration of 1, 3, and 6 hours at temperatures of 1123 and 1153 K (1562 and 1616^oF) and 1114 kPa (11 atm) total pressure. As shown in Figures 3.6-18 and 3.6-19, the basic differences in the 1123 and 1153 K (1562 and 1616^oF) gettering reactions of Na and K are that:

- On the basis of the total gettered concentration, the reactivity of K with activated bauxite is much greater than that of Na.
- Potassium at 1123 K (1562^oF) appears to be released from the insoluble lattice as a function of time, while Na remains chemically bound within the fraction. The roles of Na and K appear to be interchanged as temperature is increased to 1153 K (1616^oF).
- At 1123 K (1562^oF) potassium may be released from the insoluble bauxite lattice and recaptured within the acid soluble fraction, while Na does not appear to be involved in either a release or a

Getter Temperature: 850°C (1123K)
 Gas Composition: 80 ppmv NaCl or KCl/Ar
 Gas Velocity: 0.5 scfm (1039 m/hr)
 Pressure: 11 atm (1114 kPa)

Figure 3.6-18 Partitioning of Sodium and Potassium in Activated Bauxite at 1123 K (1562°F)

Figure 3.6-19 Partitioning of Sodium and Potassium in Activated Bauxite at 1153 K (1616°F)

recapture mechanism. At 1153 K (1616°F) Na release is apparent in all phases as the exposure period increases, but no evidence of a recapture mechanism is apparent.

- The gettering reactivity within the nepheline or feldspar fractions of the acid soluble phase prefers K to Na at both temperatures.
- The water soluble gettering reaction favors the capture of K over Na at both temperatures, possibly through a physical adsorption process.

The gettering of NaCl by activated bauxite can be enhanced by a factor of 2.8 in the water soluble phase and by 1.7 in the combined acid soluble-insoluble phase at 1123 K (1562°F) and 1114 kPa (150 psig) when 4.6 mol% water vapor is present in the alkali-laden environment. Preliminary chlorine analysis of the water soluble fraction indicates that NaCl is not the principal resulting getterred phase in the water soluble fraction. Emission spectroscopic data define the principal water soluble constituents as sodium, aluminum, and silicon, the chemical structure of which has not yet been defined.

3.6.4.3 Alternative Getter Material Performance

In search of a more suitable getter material, nine (20 total candidates) additional generic types of materials were reacted with NaCl in argon at 1123 K (1562°F) for 6 hours at 1114 kPa (150 psig) and then analyzed by atomic absorption to determine the resulting concentration of Na getterred in the water soluble and acid soluble-insoluble phases. These materials ranged from pure alumina (Al_2O_3) and silica (SiO_2) to aluminates, silicates, and aluminosilicates containing various quantities of calcium, magnesium, and iron. Table 3.6-16 identifies and summarizes the alternative getter material performance.

In order for any material to be considered as a more suitable getter than activated bauxite, chemical reaction must remove alkali at a higher getterred concentration within either the acid soluble or insoluble fraction while releasing minimal levels of background contaminants. The suggestion that the chemical reaction route is

Table 3.6-16

Dwg. 5600C21

Conditions: Temperature: 1123K (1562°F)
 Pressure: 11 atm
 Duration: 6 hrs

SUMMARY OF ALTERNATIVE GETTER PERFORMANCE

Alternative Getter Material	General Composition	Background Total Alkali		Gettered Alkali Concentration					
		Na, wt %	K, wt %	Water Soluble Na, wt %	Acid Soluble - Insoluble Na, wt %	Total Na, wt %	Water Soluble K, wt %	Acid Soluble - Insoluble K, wt %	Total K, wt %
Activated Bauxite	SiO_2 Al_2O_3 1.2 % Fe_2O_3	0.015	0.016	0.99	1.67	2.66	< 0.03	< 0.15	< 0.18
Activated Bauxite	SiO_2 Al_2O_3 1.38 % Fe_2O_3	< 0.06	< 0.06	0.68	0.94	1.62	< 0.03	< 0.03	< 0.09
Activated Bauxite	SiO_2 Al_2O_3 2.66 % Fe_2O_3	< 0.06	0.06	0.58	1.25	1.83	< 0.03	< 0.03	< 0.09
Activated Bauxite	SiO_2 Al_2O_3 7.28 % Fe_2O_3	0.09	0.12	0.52	1.66	2.18	< 0.03	< 0.06	< 0.09
PFBC Ash	SiO_2 Al_2O_3	0.68	1.74	15.47	4.93	20.40	0.03	0.03	0.07
Olivine	$(\text{Mg Fe})_2(\text{SiO}_4)$	< 0.01	< 0.01	0.13	0.74	0.87	< 0.06	< 0.03	< 0.10
Vansil [®]	CaSiO_3	0.24	0.28	1.06	7.50	8.56	< 0.09	< 0.04	< 0.13
Novacite [®]	SiO_2	< 0.01	0.04	8.93	1.1	10.03	< 0.09	0.09	< 0.18
Ebony Novacite [®]	SiO_2	0.06	2.20	0.51	5.83	6.34	< 0.1	0.06	< 0.18
Emtal Platy Talc	MgO SiO_2	< 0.01	< 0.01	4.02	2.20	6.22	< 0.06	< 0.03	< 0.10
New York State Talc	MgO SiO_2	0.15	0.09	0.17	1.20	1.37	< 0.03	< 0.03	< 0.06
Iron Ore	Fe_3O_4	< 0.01	0.06	0.16	0.27	0.43	< 0.03	< 0.06	< 0.09
Activated Alumina (spheres)	Al_2O_3	0.22	< 0.01	15.29	0.14	15.43	0.11	< 0.03	< 0.14
Activated Alumina (pellets)	Al_2O_3	0.18	< 0.01	1.65	0.41	2.06	< 0.03	< 0.03	< 0.06
Emathlite	Ca Mg Al SiO_2	0.09	0.83	0.06	3.50	3.56	< 0.03	0.11	< 0.14
Zeolite (70 μm)	$(\text{Na}_2 \text{ K}_2 \text{ Ca, B}) \cdot (\text{Al SiO}_2)_n \cdot x\text{H}_2\text{O}$	0.30	< 0.01	0.11	0.31	0.42	< 0.06	< 0.03	< 0.09
Zeolite (spheres)	$(\text{Na}_2 \text{ K}_2 \text{ Ca, B}) \cdot (\text{Al SiO}_2)_n \cdot x\text{H}_2\text{O}$	2.60	0.30	0.31	4.40	4.71	< 0.1	< 0.10	< 0.20
Zeolite (60-80 mesh)	$(\text{Na}_2 \text{ K}_2 \text{ Ca, B}) \cdot (\text{Al SiO}_2)_n \cdot x\text{H}_2\text{O}$	1.7	0.40	0.57	6.4	6.97	< 0.03	0.05	< 0.08
Diatomaceous Earth-Calcined Diatomite	SiO_2 Al_2O_3	2.40	0.30	0.19	2.60	2.79	< 0.13	< 0.26	< 0.39
Diatomaceous Earth-Celite Diatomite	SiO_2 Al_2O_3	0.46	0.62	14.50	1.70	16.20	0.06	0.11	0.17
Diatomaceous Earth-Synthetic Calcium Silicate	SiO_2 Al_2O_3	0.38	0.35	3.64	2.16	5.80	0.11	< 0.23	< 0.34

desirable is based on the possibility that desorption of the water soluble alkali would occur as the concentration of alkali in the fuel gas decreases. Such a variation, for example, could result from a change and/or variation in the feedstock.

Although a higher Na gettering activity is evident in Vansil,[®] Ebony Novacite,[®] zeolite, Emathlite, and diatomaceous earth than in activated bauxite, some potassium is subsequently released during the gettering reaction. Pretreatment of these potential getters, reducing background potassium, must therefore be considered.

Variation in the trace contaminant concentrations has often been thought to contribute to activating alkali gettering. The atomic absorption analyses in Table 3.6-16 indicate that activated bauxite containing 1.2 wt% Fe_2O_3 enhances sodium gettering in both water soluble and acid soluble-insoluble phases. Increasing the Fe_2O_3 concentration to 1.38, 2.66, and 7.28 wt% in activated bauxite tends to reduce water soluble sodium gettering while increasing sodium gettering in the acid soluble-insoluble fraction. The total gettered sodium content, however, does not exceed that obtained in the 1.2 wt% Fe_2O_3 activated bauxite. Table 3.6-16 also demonstrates the gettering of iron as Fe_3O_4 . Although as a gettering material Fe_3O_4 has one of the lowest sodium gettering capacities, 62% of the reacted alkali was shown to be partitioned within the insoluble fraction. Note that the insoluble fraction of the activated bauxite is also the preferred reacted phase, which suggests that a direct correlation between the increased gettered acid soluble-insoluble sodium and increase in the iron content exists.

3.6.4.4 Getter Reaction Stoichiometry and Proposed Reaction Mechanism

As indicated in Table 3.6-17, a total of 19 tests were involved in our preliminary TGA studies.³⁻⁵ These tests served two purposes: to develop a satisfactory operating procedure and, more importantly, to obtain kinetic data on the five getter materials tested in the bench-scale unit that showed relatively high capacities and reaction rates.

Table 3.6.17

SUMMARY OF PRELIMINARY TGA RUNS^a

Test No.	Material original wt, mg	Pressure, atm.	Getter Temperature, K (°F)	Run Duration, hr	NaCl Concentration, ppmv	Getter Weight Gain, mg
1	Bauxite (360)	1	1169 (1645)	2.0	6.3	0.86
2 A ^b	Bauxite (390)	1	1171 (1648)	1.75	7.2	0.74
B		1	1170 (1647)	1.75	11.5	0.75
C		1	1173 (1652)	1.87	14.5	0.58
D		1	1214 (1726)	2.7	17.3	0.80
E		10	1215 (1728)	2.0	4.5	0.27
3	Bauxite (406)	1	1175 (1656)	4.0	10.5	1.49
4	Bauxite (434)	1	1175 (1656)	93	0.65	1.42
5	Shattered bauxite (198)	1	1175 (1656)	15	5.1	1.33
6	Boiler ash (180)	1	1175 (1656)	27	5	1.57
7	Gasifier ash (210)	1	1175 (1656)	16.5	10.1	0.5
8	Gasifier ash	1	1176 (1657)	42+	8.7	>8 ^c
9 A	Char ash (225)	1	1175 (1656)	140	9.5	>7 ^{d,e}
B		1	1173 (1652)	180	0	-3.2
10 A	Bauxite (208)	1	1124 (1564)	71	21	0.05 ^f
B		1	1128 (1571)	18	8	~2.5 ^e
11	Bauxite (213)	1	1123 (1562)	18	4	4.2

Notes: a. Normal flows: 5 ℓ pm N₂ carrier gas, 1.25 ℓ pm He purge gas.

b. Test numbers with alphabetic designations indicate continued exposure of the same pellet.

c. Run conducted at low (0.12 ℓ pm) purge helium flow, salt found on hangdown wires after run.

d. Salt found on hangdown wires after run.

e. Resume normal flows.

f. Run made with 3 ℓ pm N₂, 5 ℓ pm He.

Table 3.6-17 (Continued)

Test No.	Material, Original wt, mg	Pressure, atm	Getter Temperature, K (^o F)		Run Duration, hr	Mean NaCl Concentration, ppm	Getter Weight Gain, mg
12	Bauxite (208)	1	1123	(1562)	19	11.5	4.0
13	Bauxite (203)	1	1174	(1654)	22	9.4	2.7
14	Bauxite (201.7)	1	1123	(1562)	44	3.3	3.2
15	Bauxite (212.4)	1	1173	(1652)	46	3.4	2.2
16	Bauxite (205.8)	1	1123	(1562)	96	0.79	1.8
17	Bauxite (217.8)	1	1173	(1652)	92	0.78	1.4
18	Bauxite (201.7)	1	1123	(1562)	43 ^(a)	12	4.4
					30 ^(d)	0	-1.1
19	Bauxite (197)	1	1173	(1652)	48 ^(a)	11	3.4
					70 ^(d)	0	-0.3

(a) = Adsorption test

(d) = Desorption test with same pellet

Results from the TGA tests, however, indicated lower performance levels on all the materials tested.

The difference in performance between the two units led to the speculation that other mechanisms might be present. One possible explanation was that the pellets may sinter in the TGA during start-up. To examine this possibility, the surface area of 573 K (572°F) dried samples of each candidate getter tested was measured, as well as samples heated for 48 hours at 1173 K (1652°F). The measurements, summarized in Table 3.6-18, show that surface area loss can indeed occur, but apparently not enough to explain the reactivity decline in ash materials. Interestingly, the bauxite showed both higher surface area and surface stability at 1173 K (1652°F).

At this point we attempted to interpret the observed TGA weight gain in terms of an assumed reaction stoichiometry. To accomplish this objective, a number of factors had to be considered. These included atomic absorption analyses of reacted and unreacted pellets, statistical analyses to determine the precision of the atomic absorption data, precision of the observed TGA weight, and assumptions with respect to reaction stoichiometry. Because of the preliminary nature of the TGA tests and other factors, we determined that only four of the tests (3 through 7) were suitable for this analysis. While the details of the procedure involved can be found in reference 6, the assumptions made with respect to stoichiometry will be mentioned here. These were as follows:

Reaction (1) was the fastest of the series and proceeded to the extent indicated by the atomic absorption:

Table 3.6-19
SURFACE AREAS OF HEATED GETTER MATERIALS

Material	573 K (572°F), 3 hr	1173 K (1652°F), 48 hr
Char Ash	1.1 m ² /g	0.39 m ² /g
Gasifier Ash	9.8	5.4
Boiler Ash	12	0.33
Bauxite	38	44
Shattered Bauxite	26	11
Kaolin	0.42	0.10

We assumed that the extent of reaction (2) was governed by the quantity of water soluble sodium in the pellet:

For the purpose of our calculations, we also assumed that the extent of reaction (3) was governed by the quantity of sodium remaining, taking into account any background sodium after reactions (1) and (2) were completed.

Using the above stoichiometry, we calculated net weight gains on the basis of atomic absorption sodium and potassium levels and then compared them to actual TGA recorded weight gains.

With the possible exception of run 3, the observed and calculated weight gains (Table 3.6-19) would appear to agree when the precision of the atomic absorption data and TGA weight measurements are taken into account. On the basis of this preliminary assessment, we have concluded that the proposed stoichiometry approaches actual reaction conditions, but additional refinements are required.

Table 3.6-19
COMPARISON OF CALCULATED AND EXPERIMENTALLY OBSERVED TGA WEIGHT GAINS

Run No.	Material	Calculated Weight Gain, mg	Observed Weight Gain, mg
TGA-3 ^a	Activated bauxite	0.82	1.49
TGA-5 ^a	Activated bauxite	1.23	1.33
TGA-6	Keystone boiler ash	1.86	1.57
TGA-7	PDU gasifier ash	0.70	0.50

^aCalculations for activated bauxite are based on a 15% weight loss during heat-up caused by hydrated water.

Since the optimization of the TGA operating procedure was completed, eight tests have been made on one of the leading candidates -- activated bauxite -- to gain insight into the adsorption mechanism by surveying the effects of temperature and gas-phase salt concentration and by observing the degree of reversibility of the adsorption. The data are summarized in Table 3.6-17, test 12 through 19. Figure 3.6-20 shows pellet weight gain vs accumulated salt exposure (product of concentration and time) for the first 6 tests of this series. Although in some cases saturation of the getter pellet had not been completed, the ultimate capacity is clearly higher at 1123 K (1562°F) than at 1173 K (1616°F) at each salt partial pressure level and at higher salt partial pressure at each temperature. These results are consistent with reversible surface adsorption mechanisms.

Tests 18 and 19 in Table 3.6-17 examined the degree of reversibility of the adsorption process. Each consisted of an adsorption portion with salt exposure, as in the previous tests, and a desorption portion in which the salt concentration in the reaction gas was essentially zero. The adsorption results agree well with the

Figure 3.6-20 Activated Bauxite Gettering TGA Data

corresponding previous results at a similar time of exposure, as for instance in the comparison of runs 12 and 18 shown in Figure 3.6-20. Importantly, the desorption appears to be not completely reversible. Transfer of salt from the gaseous state to the adsorbed reversible and irreversible states may be by competitive parallel or series mechanisms. In a hot gas cleaning system it would be better to maximize irreversibly bound sodium to preclude discharge during periods of fluctuating inlet salt concentration. On this basis, these results suggest that the capacity advantages of operating at lower temperatures are reduced.

3.6.4.5 System Performance Projections

Reactor sizing has proceeded through two stages of analysis. In the first an irreversible first-order adsorption kinetic mechanism was assumed, while in the second reversibility was assumed, based on evidence accumulated in this and other programs. In each analysis the

data available at that point were used to define kinetic expressions that were, in turn, used to obtain preliminary estimates of reactor size.

Irreversible Adsorption. An irreversible first-order adsorption rate expression is:

$$\frac{dw}{dt} = k_o p e^{-E/RT} (w_s - w), \quad (4)$$

where

w = g Na per g original pellet weight

t = time, hr

k_o = rate constant ($\mu \text{ atm} \cdot \text{hr}$)⁻¹

p = partial pressure of NaCl, $\mu \text{ atm}$

E = activation energy, cal/mol

R = ideal gas constant, cal/mol $\cdot$ K

T = temperature, K

w_s = saturation, w .

This expression was integrated and expanded as a first-order Taylor series around $t = 3$ hr and $T = 1150$ K (1610°F) to get a linear expression capable of being curve fitted⁷ with available data.

Westinghouse bench-scale reactor data for kaolin, gasifier ash, and char ash at 0.1 g/Nm³ (80 ppm) NaCl at 1114 kPa (150 psig) and 1123 to 1172 K (1562 to 1616°F)^{2,7,8} were used, as were ANL data^{9,10} on activated bauxite at 1153 K (1616°F), 101.3 kPa (1 atm), and 0.1 g/Nm³ (80 ppm) NaCl. The resulting parameters were:

	<u>k_o</u>	<u>(E/R)</u>	<u>w_s</u>
Kaolin	5.79 × 10 ⁻⁹	-12300	0.010
Gasifier ash	3.53 × 10 ¹¹	41400	0.188
Char ash	1.85 × 10 ¹²	43700	0.176
Bauxite	3.08 × 10 ⁻³	0*	0.009

The negative E value of the kaolin cannot be an activation energy, but represents the lower rates observed at higher temperatures, suggesting a physical adsorption mechanism.

A preliminary study showed these data not to be film diffusion limited.² Furthermore, the rates observed have been so low that pore diffusion also was not limiting.⁷ This means that a chemical reaction expression for the rate is appropriate. In addition, reduction of particle size cannot result in a significantly faster rate, a conclusion that will be valid at lower salt concentration if the reaction is truly first order in the gas-phase salt partial pressure.¹¹

Equation (4) can be used to calculate the design equation for an entrained flow reactor, in which getter is injected into the gas stream as a dust.⁷ The design equation is as follows:

$$\frac{C_{out}}{C_{in}} = \frac{e^{-bkL/UM}}{1 + \frac{MC_{in}}{b} (1 - e^{-bkL/UM})}, \quad (5)$$

in which

C_{out} = outlet concentration in gas-phase, mol Na/m³

C_{in} = inlet concentration in gas-phase, mol Na/m³

L = reactor length, m

b = w_sG - MC

k = k_o e^{-E/RT} RT

*Temperature dependence of rate not determined.

- G = getter loading in gas stream, g/m^3
 M = atomic weight of Na
 R = ideal gas constant ($82 \text{ m}^3 \cdot \text{atm/mol} \cdot \text{K}$)
 U = superficial gas velocity, m/hr.

With this expression an entrained flow reactor was shown to be impractical⁷ because of the very slow rates of the adsorption process with any of the above getters.

An alternative reaction is the batch packed bed in which both gas and solid phase salt concentration vary with time and position. This technique was modeled⁷ by modification of the method of Do and Weiland¹² for catalyst beds. The design equation is:

$$L_p = \frac{U}{k} \left[\ln (R - 1) + \frac{k C_{in} t M}{w_s \rho_B} \right], \quad (6)$$

where

- L_p = length of packed bed reactor, m
 ρ_B = Packed bed bulk density, g/m^3
 $R = C_{in}/C_{out}$.

This expression gives the length of reactor required as a function of time and outlet concentration. Figures 3.6-21 and 3.6-22 show bed depth and required bed life for activated bauxite and gasifier ash packed-bed reactors at a specific set of conditions. Kaolin and char ash are similar, respectively. Also shown are the fractional (relative to ultimate capacity) utilizations of getter at the conclusion of bed life. For short, useful bed life, the required depth depends primarily on kinetics; for long, useful bed life, the required bed depth depends primarily on getter capacity. These asymptotes are shown in the figures. The dependence is reasonable, since the reactor at any time will have a saturated region followed by a short transition region in which reaction occurs, followed by an unreacted region. For short beds

the transition region will occupy most of the bed, while for tall beds it will occupy a small fraction of the bed height.

From Figures 3.6-21 and 3.6-22 we see that shallow beds are useful only if replaced frequently. A deep bed (~ 10 m for bauxite, 1 m for ash) can operate for a year between replenishments. The lower ash depth is due to the higher capacity. Unfortunately, much of this sodium adsorption is accompanied by desorption of potassium, so the ash is not a candidate. What is needed is a potassium-free material of similar sodium capacity.

Reversible Adsorption. Some of the bonds formed during gettering could be weak, and previously adsorbed salt could be released in a hot, salt-free, gaseous environment. Such a reversible mechanism is suspected on the basis of some of the TGA and batch reactor test runs discussed above. Data reported by ANL¹⁰ and General Electric (GE)¹³ also support this possibility. The reversible mechanism can be treated either by a Langmuir (single layer, physical adsorption or chemisorption) or a BET (multilayer physical adsorption) model.

The rate equation in either case is

$$\frac{dw}{dt} = \frac{k_1 p}{w_s} (w_s - w) , \quad (7)$$

where

k_1 = forward rate constant (g Na/g sorbent·u-atm·hr). With either isotherm w_s is now a variable.

The Langmuir isotherm gives

$$w_s = \frac{a}{1 + \frac{K}{p}} , \quad (8)$$

where

a = maximum capacity (as $p \rightarrow \infty$), g Na/g sorbent

K = ratio of reverse to forward rate constants.

Figure 3.6-21 Bed Depth/Utilization vs Life (Bauxite)

Figure 3.6-22 Bed Depth/Utilization vs Life (Gasifier Ash)

The constants k_1 , a , and K were determined to be 0.5×10^{-4} , 0.0056, and 0.25, respectively, by trial and error analysis of Westinghouse bauxite data, including the batch reactor data for 1, 3, and 6 hour exposure at 1123 and 1153 K (1562 and 1616°F) to 880 μ -atm NaCl and the data of TGA runs 1 through 5 at 0.65 to 17 μ -atm NaCl and 1173 to 1213 K (1652 to 1724°F). The TGA curves were divided into a discrete number of points corresponding to different exposure times for analysis. Temperature dependences of k_1 and K could not be discerned because of data scatter.

If the BET isotherm, rather than the Langmuir, is used, equation (7) still applies, but w_s is now given by:

$$\frac{p}{w_s (p_o - p)} = \frac{1}{b w_m} + \frac{(b - 1)}{b w_m} \frac{p}{p_o}, \quad (9)$$

where

p_o = equilibrium vapor pressure of the adsorbed species, μ atm

b = constant

w_m = w for monolayer coverage.

The constants b and w_m were determined from regression analysis using the Westinghouse and GE data to be 3085 and 0.0050, respectively, while the same value of k_1 is used.

The above two reversible models were applied to the source Westinghouse data and to published ANL and GE data. ANL data points,¹⁴ covering temperatures of 1073 and 1153 K (1472 to 1616°F), atmospheric pressure, salt partial pressures of 30 to 150 μ -atm, exposures of 1 to 15 hours, and various gas velocities and packed-bed depths, were checked. Since the ANL data resulted from packed beds, a numerical version of the packed-bed model presented above was used to test the

above expression. The GE data¹³ covered temperatures of 1073 and 1173 K (1472 and 1616°F), atmospheric pressure, salt partial pressures of 0 to 170 μ -atm, and exposure times of 0 to 250 hours. Average absolute values of errors in the fit of models to data were:

Data Source	Model	
	Langmuir	BET
Westinghouse	18%	28%
ANL	18%	10%
GE	41%	41%

The packed-bed reactor model of equation (6) was used to examine the size reactor needed, based on these kinetics and the conditions used in the irreversible kinetics study above. However, k is redefined as:

$$k = k_1 RT \rho_B / M \quad (10)$$

because of the difference between equations (4) and (7).

The model is valid only if w_s is constant, which, for the reversible adsorption mechanisms, it is not. It was therefore applied by breaking the exposure time down into small time increments, during each of which the bed was broken down into small layers. Each layer for each time increment was assumed to have a constant w_s .

The results indicate that to prevent breakthrough after 6000 hours, a 26 m bed is needed, while to prevent breakthrough at 4000 hours, a 17 m bed is needed. Using the irreversible kinetic model, we had projected that a 10 m deep bed would be required to achieve the 4000 hour criterion. The difference is due solely to the new estimation of capacity (0.0056 vs 0.0090 g Na/g sorbent at 80 μ -atm NaCl) and is not significantly affected by reversibility itself, because, in a deep bed, most of the bed is either completely saturated and exposed to bulk inlet level concentrations of alkali, thus preventing desorption, or

completely unsaturated and exposed to no salt. Only in a thin layer is the actual reaction occurring and influenced by variable capacity. Of more concern would be desorption of previously adsorbed alkali during any fluctuations in inlet loading. This would cause migration of the alkali to downstream regions of the bed and would lead eventually to premature breakthrough. For a similar reason a fluidized-bed contactor would also be inappropriate.

Ten additional getter material screening tests were conducted on the TGA.¹⁵ Tables 3.6-20 and 3.6-21 summarize the test NaCl concentration, temperature, duration, measured weight gain, and resulting alkali concentration within the various gettered phases. These data indicate the enhanced gettering ability of Ebony Novacite,[®] Emathlite, and diatomaceous earth in comparison with activated bauxite, and also that the presence of water vapor in the reactant gases further promotes the alkali gettering reaction. Because of their high gettering capacity, these materials serve as candidate getters in hot gas cleaning systems.

3.6.5 Summary

Quantitative analysis by atomic absorption has indicated that the Exxon 3^o cyclone PFBC ash contains < 0.70 wt% sodium and 1.77 wt% potassium, while activated bauxite contains < 0.14 wt% sodium and < 0.11 wt% potassium. The background alkali content of the ceramic filter was determined at levels of < 0.22 wt% sodium and < 0.19 wt% potassium in the 427 filter medium and at < 0.21 wt% sodium and 0.25 wt% potassium in the 485 filter medium. Both the PFBC ash and ceramic filter medium are rich in silicon and aluminum, which are considered to be the essential elemental constituents for the gettering reaction.

Because of the relatively high alkali background content in the Exxon 3^o cyclone PFBC ash, initial release studies were conducted prior to exposing the PFBC ash to an alkali-laden environment. The background alkali concentration in the PFBC ash was retained within the ash lattice at the reaction conditions typifying the hot gas cleanup system.

Table 3.6-20
COMPARISON OF TGA DATA WITH ATOMIC ABSORPTION RESULTS

Run	Material	Concentration, ppmv	Temperature, K (°F)	Length, hr	Wt Gain, mg	% Wt Gain	Wt % Na by Atomic Absor.
20	Vansil	13.4	1173 (1652)	26	0.3	0.144	0.30
21	Vansil	12	1073 (1472)	24	0.3	0.160	0.34
22	Ebony Novacite ^(R) a	13.9 - 5.7	1123 (1562)	165	2.3	0.833	3.56
23	Ebony Novacite ^(R) b	12.1 - 3.1	1123 (1562)	267	14.5	4.71	5.61
24	Zeolite	12.0 - 9.6	1123 (1562)	50	0.8	0.465	2.92
25	Emathlite	11.0 - 3.7	1123 (1562)	284	10.8	5.48	5.45
26	Diatomaceous Earth	9.7 - 4.2	1123 (1562)	300	6.5	4.67	4.59
27	Novacite ^(R)	11.2 - 5.0	1123 (1562)	283	4.9	2.49	2.15
28	Activated Bauxite ^c	22.0 - 6.6	1123 (1562)	257	16.1	7.87	6.61
29	Activated Bauxite ^c	24.0 - 8.1	1123 (1562)	113	12.5	6.11	5.26
30	Emathlite ^c	17.0 - 7.0	1123 (1562)	117	20.8	9.52	10.66 ^d (3.81)

^aEffect masked by carbon oxidation.

^bPrebaked to remove volatiles.

^cWater vapor present in carrier gas at 4.0 mol %.

^dGlass outer shell (interior of the pellet).

Table 3.6-21
DETAILED ATOMIC ABSORPTION RESULTS FOR ALTERNATIVE GETTERS

Run	Material	Background Total Alkali		Water Soluble Wt % Na	Reacted Getter Analysis		
		Wt % Na	Wt % K		Insoluble Wt % Na	Water Soluble Wt % K	Insoluble Wt % K
20	Vansil ^(R)	0.24	0.028	NA ^a	0.30	NA	< 0.05
21	Vansil ^(R)	0.24	0.028	NA	0.34	NA	< 0.05
22	Ebony Novacite ^(R)	0.057	2.2	0.047	3.51	< 0.01	< 0.02
23	Ebony Novacite ^(R)	0.057	2.2	0.028	5.58	< 0.01	< 0.02
24	Zeolite	1.7	0.40	0.010	2.91	< 0.01	0.18
25	Emathlite	0.09	0.83	0.058	5.39	< 0.01	< 0.02
26	Diatomaceous Earth	0.46	0.62	0.095	4.49	< 0.01	< 0.02
27	Novacite ^(R)	< 0.01	0.044	0.22	1.93	< 0.01	< 0.02
28	Activated Bauxite	0.07	< 0.05	3.10	3.51	< 0.05	< 0.15
29	Activated Bauxite	0.07	< 0.05	1.64	3.62	< 0.05	< 0.15
30	Emathlite	0.09	0.83	0.69 ^b (0.16)	9.97 (3.65)	< 0.06 (0.06)	< 0.14 (0.14)

^aNA -- not analyzed and therefore included in the insoluble fraction.

^bGlass outer shell (interior of the pellet).

Gettering of gaseous alkali by the Exxon 3^o cyclone PFBC ash, activated bauxite, and the ceramic filter medium was enhanced as a function of time, and at lower reaction temperatures. There was an average increase of 2.39 wt% total sodium over background in the 1088 K (1500^oF), 1 hour, 1114 kPa (150 psig) pressure gettered PFBC ash, while an average of 5.22 wt% total sodium resulted after 4 hours at these reaction conditions. Gettering of the PFBC ash was reduced to 1.02 wt% total sodium as the 1 hour reaction temperature was increased to 1200 K (1700^oF) and 1114 kPa (150 psig) pressure. In these tests the ceramic filter supporting the PFBC ash cakes gettered 1.07 and 3.89 wt% sodium after 1 and 4 hours of exposure to 0.1 g/Nm³ (80 ppm) gaseous NaCl at 1088 K (1500^oF) and 1114 kPa (150 psig) pressure. At 1200 K (1700^oF) and after 1 hour of reaction, the concentration of sodium increased only 0.75 wt% in the filter medium.

During sodium gettering, background potassium in these materials is released simultaneously, implying that an ionic exchange mechanism may be occurring. As a gettering medium, the ceramic filter, when exposed to the alkali-laden environment, is able to react with sodium, gettering an average of 4.88 wt% sodium after 4 hours at 1088 K (1500^oF) and 1114 kPa (150 psig) pressure, and an average of 1.07 wt% sodium after 4 hours at 1200 K (1700^oF) and comparable system pressure. The ceramic filter appears to maintain its structural stability throughout the 4-hour alkali exposure.

At the termination of a 6-hour exposure of activated bauxite to 0.1 g/Nm³ (80 ppm) NaCl in argon gas at 1123 K (1562^oF) and 1114 kPa (150 psig) pressure, an average of 0.65 wt% sodium results over background, while at 1153 K (1616^oF) only 0.2 wt% sodium is gettered.

These data indicate that the gettering reactions and mechanisms apparently are directly affected by the reaction temperature and depend primarily on the material physical and chemical structure and composition.

Because of the limited rate, capacity, and other problems of potassium released in these materials, we have directed part of our efforts to evaluating new candidate getter materials. The preliminary results obtained from the bench-scale unit have been encouraging. Several of the getters -- Vansil,[®] Ebony Novacite,[®] zeolite, Emathlite, and diatomaceous earth -- had higher gettering capacities than activated bauxite. The presence of water vapor in the reactant gases tends to promote even further the alkali gettering capacity in these materials.

On the basis of available data and modeling efforts, we concluded that long contact times must be provided for effective gettering, so deep packed beds appear to be the best of the various contacting devices (packed beds, entrained flow reactors, and fluidized beds) considered. With such beds, bed heights of about 10 m for activated bauxite and about 1 m for ash would be required, if we assume annual replacement on the basis of bed saturation capacity. Superficially, it would appear that an ash depth of 1 m would be satisfactory. Unfortunately, much of this sodium adsorption is accompanied by the desorption of potassium. The ash, therefore, is not an acceptable candidate. One concern with activated bauxite is the reversibility of the reaction. Desorption of previously absorbed alkali, for example, could occur during fluctuations in inlet loading. As a result, it may be more prudent to operate with this material at a higher temperature to encourage formation of insoluble alkali even at a reduced capacity. This penalty may be more than compensated for by higher temperature operation of the hot gas cleaning system.

3.7 FIELD TESTING

At the beginning of this project, we planned to carry out tests of the cross-flow filter at an actual coal-burning, pressurized, fluid-bed combustor. The primary motivation for such testing was concern that redispersed PFBC ash in a high-pressure and -temperature test rig might not properly simulate the sticky, adhesive nature of ash created in an actual coal-burning test rig.

To this end several possible test sites were identified early in the program and reviewed with both test site and DOE personnel. From these efforts the small-scale test PFBC under construction at ANL was identified as being the most suitable for these purposes.

The subsequent detailed design of the filter housing pressure vessel and its inlet and outlet sections were coordinated with the ANL hot gas piping, so that, upon completion of bench-scale testing at Westinghouse, the filter, filter housing, and inlet and outlet sections could be shipped to ANL and be integrated into their system with a minimum of cost and effort.

As the current test program evolved and progressed toward completion, the funding required for operation of the ANL test facility became uncertain, and no commitment for this testing was possible. Candidate filter media and the associated test hardware were stored and ready at the time, and when operating funds for the ANL system became available testing proceeded. The results of this field testing appear in Appendix E.

4. REFERENCES

1. Lee, S. H. D., and I. Johnson, Removal of Gaseous Alkali Metal Compounds from Hot Flue Gas by Particulate Sorbents, presented at the 24th Annual International Gas Turbine Conference, San Diego, CA, March 11-15, 1979.
2. Mulik, P. R., D. F. Ciliberti, M. A. Alvin, M. M. Ahmed, D. M. Bachovchin, and D. L. Keairns, Simultaneous High-Temperature Removal of Alkali and Particulates in a Pressurized Gasification System, Final Technical Progress Report prepared for DOE by Westinghouse R&D Center, Pittsburgh, PA, October 1978-October 1980, Contract No. DE-AC21-78ET11026.
3. Mulik, P. R., M. A. Alvin, and D. M. Bachovchin, Simultaneous High-Temperature Removal of Alkali and Particulates in a Pressurized Gasification System, Annual Technical Progress Report prepared for DOE by Westinghouse R&D Center, Pittsburgh, PA, April 1981-March 1982, Contract No. DE-AC21-81MC16372.
4. Mulik, P. R., M. A. Alvin, and D. M. Bachovchin, High-Temperature Removal of Alkali Vapors in Hot Gas Cleaning Systems, presented at the DOE/Morgantown Energy Technology Center Second Annual Contractor's Meeting on Contaminant Control in Hot Coal-Derived Gas Streams, February 17-19, 1982.
5. Mulik, P. R., M. A. Alvin, and D. M. Bachovchin, Simultaneous High-Temperature Removal of Alkali and Particulates in Pressurized Gasification Systems, January 15, 1982, Monthly Progress Letter, Westinghouse R&D Center, Pittsburgh, PA, Contract No. DE-AC21-81MC16372.
6. Mulik, P. R., M. A. Alvin, and D. M. Bachovchin, Simultaneous High-Temperature Removal of Alkali and Particulates in a Pressurized Gasification System, Third Quarterly Technical Progress Report, October 1981-December 1981, Westinghouse R&D Center, Pittsburgh, PA, Contract No. DE-AC21-81ET16372.
7. Mulik, P. R., M. A. Alvin, and D. M. Bachovchin, Simultaneous High-Temperature Removal of Alkali and Particulates in a Pressurized Gasification System, First Quarterly Technical Progress Report, April 1981-June 1981, Westinghouse R&D Center, Pittsburgh, PA, Contract No. DE-AC21-81ET16372.

8. Mulik, P. R., et al., High Temperature Removal of Alkali and Particulates in a Pressurized Gasification System, ASME Paper No. 81-GT-67, Gas Turbine Conference, March 1981, Houston, TX.
9. Lee, S. H. D., W. M. Swift, and I. Johnson, Screening of Granular Sorbents for the Removal of Gaseous Alkali Metal Compounds from Hot Flue Gas, Report prepared for DOE, Argonne National Laboratory, Argonne, IL, November 1979, ANL/CEN/FE-79-17.
10. Lee, S. H. D., and I. Johnson, Removal of Gaseous Alkali Metal Compounds from Hot Flue Gas by Particulate Sorbents, J. Eng. Power, 102; April 1980: 397-402.
11. Carberry, J. J., Chemical and Catalytic Reaction Engineering, New York: McGraw-Hill Book Company; 1976, p. 222.
12. Do, D. D., and R. H. Weiland, Fixed Bed Reactors with Catalyst Poisoning. Chem. Eng. Sci., 36; 1981: 97-104.
13. Boericke, R., et al., Alkali Control in PFB Combustion Systems, presented at DOE Contractors Meeting on High-Temperature and High-Pressure Particulate and Alkali Control in Coal Combustion Process Streams, Morgantown, WV, February 1981.
14. Lee, S. H. D., W. M. Swift, and I. Johnson, Diatomaceous Earth and Activated Bauxite Used as Granular Sorbents for the Removal of Sodium Chloride Vapor from Hot Flue Gas, Report prepared for DOE, Argonne National Laboratory, Argonne, IL, January 1980, ANL/CEN/FE/79-18.
15. Mulik, P. R., M. A. Alvin, and D. M. Bachovchin, High-Temperature Removal of Alkali and Particulates in a Pressurized Gasification System, Monthly Progress Letters for June 16, 1982 and July 19, 1982, Westinghouse R&D Center, Pittsburgh, PA, Contract No. DE-AC21-81ET16372.

APPENDIX A
CERAMIC FILTER TEST SYSTEM DESCRIPTION DOCUMENT

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1. INTRODUCTION

The program for which this test passage is being constructed is one of five bench-scale programs that DOE is currently sponsoring in an attempt to resolve the hot gas particulate and alkali removal problems associated with pressurized fluid combustion of coal. The program as it currently stands is about an 18-month effort, during which time about \$500,000 will be expended.

We proposed specifically to advance the use of cross-flow porous ceramic membrane filters to the point where large-scale pilot testing can be undertaken to confirm commercial feasibility.

Conventional inertial methods for particulate removal are generally effective for particles larger than about 10 μm in diameter; however, other methods must be employed for smaller material. Filtration - which uses the capture mechanisms appropriate to these fine contaminants - is one such technique.

The cross-flow ceramic membrane filter is a significant new development in hot gas filtration and warrants additional investigation. As shown in Figure A1.1, this particular configuration of filter medium offers many unique attributes. In the course of preliminary testing done by Westinghouse, the following characteristics were determined experimentally:

- Essentially absolute filtration for $< 10 \mu\text{m}$ dust
- Compact configuration, about 492 m^2 of area per m^3 volume ($150 \text{ ft}^2/\text{ft}^3$)
- Modest pressure drops in operation, similar to those of conventional bag filters
- Temperature and thermal shock resistance
- Corrosion resistance for virtually all coal conversion technologies.

Figure A1.1 Schematic of Large-Scale Configuration for Ceramic Membrane Filter Unit

We believe these operational advantages form a strong basis for establishing technical feasibility. In addition, compact, economical devices are inherent to this filter configuration because the material has an enormous filter-area-to-volume ratio (on the order of ten times that of conventional bag filters). Coupled with recent developments in module configuration (shown in Figure A1.1), a strong case can be made for economic feasibility.

2. BACKGROUND

In mid-1975, Westinghouse was awarded an EPA contract to look at thin porous ceramics as a possible technique for fine particulate filtration in hot gases.

A portion of the effort was directed toward looking at various available forms of ceramic materials for use as hot gas filters. During this work, the concept of using cross-flow ceramic blocks for hot gas filtration was evolved and tested.

Cross-flow elements are constructed in the following sequence (Figure A2.1):

- A thin flat layer of gas-permeable ceramic material is laid out.
- A layer of corrugations is then placed on the flat layer.
- A second flat layer is placed on top of the corrugations.
- Finally, a second layer of corrugations is placed on the second flat layer, oriented perpendicularly to the first corrugations.

This sequence is repeated over and over, to a maximum of about six inches or 50 to 85 layers. The element is then fired, and the composite becomes a monolith. An alternative route to a similar structure is to extrude thin flat sheets with corduroylike ribs on one side and then stack the sheets with alternating rib orientation. Again, the green ceramic mix material is laid up and then fired to form a unitized structure.

Figure A2.1 Cross-Flow Ceramic Membrane Filter

These materials can be used as filters as shown in Figure A2.1. With one end of the filter sealed, dusty gas can enter the element corrugations running parallel to the sealed face. The gas permeates the flat sheets, separating layers of corrugations and depositing the dust on the higher-pressure side. The cleaned gas then flows out of the element through the corrugations (oriented perpendicular to the inlet channels). Accumulated dust within the filter can be blown out periodically by a burst of high-pressure air from the clean side. This air flows back through the membrane filter, lifts the cake off the wall, and carries it out of the element. An alternative method is to send high velocity air through the dirty-side channels to sweep accumulated dust from the wall. One of the primary attractions of this configuration is the ability to pack a great deal of filter area in a very compact module: a most important consideration for pressurized high-temperature systems.

3. MAJOR SUBSYSTEMS DESCRIPTIONS

An overall schematic of the test passage and its major instrumentation and control loops is presented in Figure A3.1. The test facility is located in the newly constructed laboratory area, 301-1C13, shown in Figure A3.2.

3.1 Combustor/Temperature Control

3.1.1 Fuel Supply

The choice of heat input for the test system was between resistive electric heating and direct combustion. Resistive heating was found to have several drawbacks for this particular application. Commercially available heaters are not considered to be very reliable at the temperatures required [1144 K (1600°F)] when used with dusty air that may contain alkali metal vapors. Additionally, since the maximum heat input is about 65 kW, approximately 1419 dm² (2,200 in²) of heater surface would be required, resulting in a rather cumbersome system. Direct combustion of a fuel in the system offers a reliable, compact solution to these difficulties. Burning a fuel such as propane directly in the flow systems requires a flow of only 0.76 dm³/s (1.6 scfm) at the maximum usage. The main disadvantage to the use of a combustion system is the additional flame safety instrumentation required, along with provision for a safe fuel supply system. Since a great deal of experience in dealing with combustion systems such as the one required here was available from our combustion group, this option was chosen. Propane was chosen as fuel for the system because relatively large amounts of it can be purchased in standard tanks. A standard 1F tank of propane contains 45 kg (100 lb) of propane (enough for about 10 hours of run time at the maximum usage). For comparison, a 1F tank of methane contains only about 0.7 kg (0.16 lb) of methane. This problem of very large storage requirements is common to most gaseous fuels. Additionally, propane can be brought to system pressure with simple thermostatic control, whereas pressurization of facility-supplied natural gas would

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Figure A3.1 Filter Test System Schematic

Figure A3.2 Ceram Facility

require an extensive, complicated, natural gas compressor. Alternative liquid fuels such as alcohol would require a boiler, liquid metering systems, pressure control, atomizing systems. On consultation with the combustion personnel, the propane system was selected.

A functional schematic of the propane system is shown in Figure A3.3. Under normal operating conditions propane must be supplied at pressures of approximately 1076 kPa (150 psig). From Figure A3.4 one can see that the propane supply must be warmed slightly to raise its vapor pressure above 1076 kPa (150 psig). We anticipate that propane supply temperatures of about 300 to 326 K (100 to 100°F) will be sufficient to create adequate supply pressure. This warming will be accomplished by immersing the propane tank in an electrically heated water bath. The water bath contains roughly 189 dm³/s (50 gal) of water and provides a very large thermal mass. This allows temperature control [by TC(1)] to ± 0.5 to 1°K (1 or 2°F) without difficulty. As an additional high-temperature safety control, TC(2), a high-temperature switch, has been installed to provide redundancy in the high-bath-temperature control. The propane tank is housed in a totally enclosed hood ventilated by an explosion-proof fan capable of exhausting 944 dm³/s (200 cfm) at 2.5 kPa (10 in. wg). All electrical contactors have been mounted outside the hood, while all tubing connections (with the exception of the final one) are inside the hood. Pressure relief for the system is provided by PRV(1), which vents to the roof of the building at pressures above 2869 kPa (400 psig). Even at boiling water temperatures the internal propane pressure is only about 4734 kPa (660 psig), which is well within the capability of the supply tank. Propane vapor flows from the tank through a tank-mounted shut-off valve [PTV(1)], through an additional shut-off valve [CV(3)], and through a check valve [SV(1)] to a normally closed explosion-proof solenoid valve [SV(1)]. During combustion, vapor flows through SV(1) to a temperature control valve [TCV(1)] that regulates the passage temperature by control of the propane flow rate. From TCV(1) propane flows directly to the combustor section of the test passage. The tube between the propane

Fuel Supply Loop				
Qty	Description	Manufacturer	Symbol	Remarks
1	Shutoff Valve 1/4" - 1/2" NPT - Std. Sill.	Whitely	PTV (1)	Mounted on Tank
1	Solenoid Valve Normally-Closed 1/4" NPT 1/8" Orifice .35 CV	ASCO	SV (1)	N.C. When not Testing Shuts Off On • Loss of Flame • High Temp. Air Flow Thru Comb • Receiver Low Pressure • High Cooling Water Temp
1	Temperature Control Valve-1/4" NPT Std. Sill. Body Air to Open CV-.0081-.32	Research Control Valve	TCV (1)	
2	Check Valve		CK (1) & (2)	
1	Thermocouple Type K Chromel-Alumel	Marlin	TE (1)	2300°F Range
1	Thermocouple Amplifier Temperature Transmitter 1-5 VDC Output	Veritrac	TT (1)	
1	Temperature Recorder Controller 1 Pen 3 Mode 1-5 VDC Input 4-20 mA Output	Veritrac	TRC (1)	Auto/Manual
1	Current/Pressure Transducer Input 4-20 mA Output 3-15 psi	Moore	I/P (2)	
1	Water Bath Temp Controller	Omega	TC (1)	Controls Water Bath Temp at 110 to 145°F
1	Hi Limit Control	Chromalox	TC (2)	High Limit
1	Pressure Relief Valve	Matheson	PRV (1)	Relieves at 400 psi Vents to Roof
1	Manual Shut Off Valve	Whitely	CV (3)	N.C. When Not in Operation
1	Heater Coil	Chromalox		240 V AC, 5 kW
1	Water Bath	Plastic Piping Systems		50 Gallon Capacity Polypropylene Const
1	Exh Fan w/Exp.-Proof Motor	Acme Model #QW9A8		Min 200 CFM @ 10" SP
1	Flowmeter 0 - 100 %	Brooks	FM (2)	Propane Premix
1	Flowmeter 0 - 100 %	Brooks	FM (3)	Fireye Purge
2	Control Valve	Hoke	PPV (1) FPV (1)	Propane Premix Fireye Purge Control

Figure A3.3 Functional Schematic of Propane System

Figure A3.4 Propane Vapor Pressure vs Temperature

tank and the passage is traced with heating tape to prevent condensation of propane in the lines.

3.1.2 Combustor Design

The combustor design consists of the simple flameholder shown in Figures A3.5 and A3.6. Propane flows radially out through the eight [1.78 mm (0.070 in) dia.] holes located on the turbulent side of the flameholder cone. Ignition is provided by a standard oil furnace igniter and transformer. The combustor is centered in the combustor liner/holder assembly, shown in Figure A3.6, which is a section of Inconel pipe. This section is wrapped with fiber frax insulation and attached to the pressure containment for the combustor, shown in Figure A3.7. This section is 15 cm (6 in) 304 SS pipe that has been code stamped. Penetrations through the pressure shell and liner sections have been included for the flame-sensing safety equipment.

3.1.3 Flame Safety Control

Combustion safety controls are summarized in Figure A3.8. The first safety control discussed is for flameout condition. In order to prevent a situation where propane is being fed to the passage but not being combusted, a standard flame safety control has been provided. The flame detector is supplied by Fireye and senses the infrared radiation from the flame through a quartz lens. The lens is purged with air to cool it and keep it clean. Within 2-1/2 seconds of loss of flame, the Fireye control unit opens the circuit that holds the propane solenoid valve [SV(1)] open. With SV(1) closed, no additional propane can flow into the passage without returning to the proper start-up sequence. If a flameout occurs at the maximum expected flow of propane [about 5.5 kg/hr (12 lbm/hr)], the total flow of propane into the passage during the 2-1/2 second shut-off period would be about 0.004 kg (0.009 lbm) of propane. At maximum propane flow the passage conditions would be 1144 K (1600°F) and 1138 kPa (165 psia). If the propane admitted to the system after a flameout were ignited and if it were assumed that no flow would

Figure A3.5 Hot Gas Cleaning System - Burner and Dust Nozzle Assembly

Combustion Control Loop				
Qty	Description	Manufacturer	Symbol	Remarks
1	Pressure Control Range 0-350 psi Deadband 2-8 psi Low Pressure	United Electric	PSL (1)	Shuts Off Propane on Loss of Passage Pressure
1	Solenoid Valve Normally Closed 1/4" NPT 1/8" Orifice 35 CV	ASCO	SV (1)	Shuts Off Propane Flow On Any of Several Safety Signals
1	Ignitor 6000 Volts	Bendix		Normal Oil Furnace Ignitor
1	Ignition Transformer Pri. Volts 120 Amps 1.5 Sec. Volts 6000 Amps 0.020	Bendix Industry	Ign. Tx	
1	Safe Guard Controls Flame Proven Indication Flame Scanner Infrared	Fireye		Shuts Off Propane on Loss of Flame
1	Thermocouple Type K Chromel-Alumel	Marlin	TE (2)	Senses High Passage Temp. and Shuts Off Propane
1	Thermocouple Type J Iron-Constantan	Marlin	TE (3)	Senses High Water Temp and Shuts Off Propane
1	Temperature Indicator Controller Type K	Omega	TSH (1)	Passage High Temp Switch
1	Temperature Indicator Controller Type J	Omega	TSH (2)	Cooling Water High Temperature Switch
1	Isolation Valve	Whitey	TV (9)	Isolates CGA from Passage
1	Combustible Gas Analyzer	Bacharach	CGA	Detects Combustible Gas in Passage

Figure A3.8 Combustion Safety Control Loop

be allowed from the passage, the maximum pressure rise in the system would occur. Calculations indicate the maximum pressure rise would be from 1138 to 1827 kPa (165 to 265 psia) for these conditions. This is well within the system's pressure capability.

A second occurrence that will shut the combustion system down is a high-temperature condition as sensed by thermocouple TE(2) and the temperature indicating controller, TSH(1). Should the indicated temperature exceed the chosen set point [usually 1144 K (1600°F)], TSH(1) closes SV(1) and combustion ceases.

A third condition that will stop propane flow is a high-temperature reading from the exhaust system cooling water. Thermocouple TE(3) and indicator controller TSH(2) monitor the cooling jacket water temperature for the exhaust system. Should the water temperature exceed the chosen set point [usually 352 K (175°F)], TSH(2) will close SV(1) and stop combustion.

A fourth system upset that will automatically close SV(1) is a loss of system pressure as sensed by a pressure switch, PSL(1), which is mounted on the compressor receiver.

The fifth method of abnormal combustion interruption is by activation of three panic buttons located at the control panel and at each of the two laboratory exits.

3.1.4 Temperature Control

Temperature control for the system is effected as indicated in Figure A3.9. The operation is as follows.

System temperature is sensed by thermocouple TE(1), whose output is amplified and transmitted by TT(1) to a temperature recorder/controller. In automatic mode the controller compares the measured temperature with the desired set point and opens or closes the propane flow control valve, TCV(1), accordingly. Since TCV(1) is a pneumatically operated valve, a current-to-pressure transducer [I/P(2)] is provided.

Fuel Supply Loop				
Qty	Description	Manufacturer	Symbol	Remarks
1	Shutoff Valve 1/4 - 73 CV-Stn STL	Whitely	PTV (1)	Mounted on Tank
1	Solenoid Valve Normally-Closed 1/4 NPT 1/8" Orifice .35 CV	ASCO	SV (1)	N.C. When not Testing Shuts Off On • Loss of Flame • High Temp. Air Flow Thru Comb • Receiver Low Pressure • High Cooling Water Temp
1	Temperature Control Valve-1/4 NPT Stn STL Body Air to Open CV-0081-32	Research Control Valve	TCV (1)	
2	Check Valve		CK (1) & (2)	
1	Thermocouple Type K Chromel-Alumel	Marlin	TE (1)	2300°F Range
1	Thermocouple Amplifier Temperature Transmitter 1-5VDC Output	Veritrac	TT (1)	
1	Temperature Recorder Controller 1 Pen 3 Mode 1-5 VDC Input 4-20 mA Output	Veritrac	TRC (1)	Auto/Manual
1	Current/Pressure Transducer Input 4-20 mA Output 3-15 psi	Moore	I/P (2)	
1	Water Bath Temp Controller	Omega	TC (1)	Controls Water Bath Temp at 110 to 145°
1	Hi Limit Control	Chromalox	TC (2)	High Limit
1	Pressure Relief Valve	Matheson	PRV (1)	Relieves at 400 ps Vents to Roo
1	Manual Shut Off Valve	Whitely	CV (3)	N.C. When Not in Operation
1	Heater Coil	Chromalox		240 V AC 5 in
1	Water Bath	Plastic Piping Systems		50 Gallon Capacity Polypropylene Cons
1	Exh Fan w/Exp -Proof Motor	Acme Model #949AB		Min. 200 CFM @ 10" SP
1	Flowmeter 0 - 100 %	Brooks	FM (2)	Propane Premix
1	Flowmeter 0 - 100 %	Brooks	FM (3)	Fireve Purge
2	Control Valve	Hoke	PPV (1) FPV (1)	Propane Premix Fireve Purge Control

Figure A3.9 Temperature Control System

In manual control TCV(1) can be operated from the control panel to give any desired position. The recording of actual passage temperature occurs continuously, regardless of which mode is currently in use.

3.2 High-Pressure Air Supply and Control

3.2.1 Compressor

The high-pressure air supplied to the test passage is provided by an Ingersoll-Rand reciprocating compressor whose maximum capacity is $60.4 \text{ dm}^3/\text{s}$ (128 scfm) at a discharge pressure of 1793 kPa (250 psig). As shown in Figure A3.10, the compressor system is a skid-mounted package consisting of the compressor, a 29.84 kW (40 HP) electric motor, and a 2.83 m^3 (10 ft^3) air receiver.

3.2.2 Dryer/Filter System

To remove moisture and any oil in the air from the compressor, a Hankison Model 80100 air dryer has been installed. This device is a refrigerated system that will give at least a 283 K (50°F) dew point when the compressor flow is a maximum. Oil and condensate are removed by a separator/filter assembly that has an automatic blowdown feature.

3.2.3 Flow Control and Measurement

Airflow control and measurement is accomplished by the system shown schematically in Figure A3.11. Mass flow measurements are made through the use of standard orifice plates and pressure taps. Since the normal range of turndown for a single orifice plate is only slightly more than 3:1, two plates have been installed in parallel. These allow about a 10:1 turndown in flow measurement by simply switching from one orifice plate to the other through the use of GV(1) and GV(2). Air from the compressor/dryer system flows through the chosen orifice plate, and the differential pressure across the plate is monitored by a differential pressure transmitter PDP(1). The upstream pressure is simultaneously monitored by PT(2). These two signals are fed to a mass flow

DIMENSIONS
Type 40 Packaged Air Compressor
Models 25 Thru 100

Form AP 0276 A
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Ingersoll-Rand
AIR POWER COMPRESSOR DIVISION

Figure A3.10 Compressor System

Flow Control Loop				
Qty	Description	Manufacturer	Symbol	Remarks
1	Pressure Regulator		PR (1)	
2	Gate Valve 1.0"	Smith	GV (1) (2)	Isolates Unused Orifice Plate
4	Toggle Valve 1/4"	Whitely	TV (1) (2) (3) (4) (10)	
2	Orifice Flanges 1.0" Weldneck 300 lb Sch 40	Daniel Ind.	FE (1) (2)	
2	Orifice Plates	Daniel Ind.		2 Plates in System at Any One Time to Allow 10-1 Turn Down in Flow
1	Capsuletic Differential Pressure Gauge 0-100" WC	Dwyer	DPG (1)	
1	Differential Pressure Transmitter 0-250" WC 4-20 mA Output	Veritrak	PDT (1)	
1	Flow Recorder Controller 1 Pen 2 Mode 4-20 mA Input & Output	Veritrak	FRC (1)	
1	Current/Pressure Transducer Input 4-20 mA Output 3-15 psi	Moore	I/P (1)	
1	Flow Control Valve Air to Close 1.0" NPT Input 3-15 psi CV Range .04 to 2.3	Masonellan	FCV (1)	
1	Pressure Transmitter 0-400 psig 4-20 MA Output	Veritrak	PT (2)	
1	Pressure Transmitter	Veritrak	PT (2)	Signals Orifice Plate Pressure to FRC (1)

Figure A3.11 Flow Control Loop

controller/recorder that uses the pressure and differential pressure measurements to derive the total mass flow. The controller then adjusts the flow control valve FCV(1) through the current/pressure transducer [I/P(1)] to match the input set point.

3.2.4 Back-Pressure Control

The system pressure control loop is shown in Figure A3.12. The system consists of a pressure tap at the filter housing inlet that is connected to a pressure transmitter, PT(1). The signal from PT(1) is fed to an indicating pressure controller, PIC(1), which operates the pressure control valve, PCV(1), through a current-to-pressure transducer [I/P(3)]. System pressure is indicated and recorded at the control panel. The back-pressure control valve, PCV(1), is a Masoneilan Micropak high-performance control valve rated at 10,758 kPa (1500 psig). The continuous operation temperature limit is 617 K (650°F). For this reason the back-pressure valve is located some 7.3 m (24 ft) away from the test passage outlet, and this entire length is water-cooled to drop the gas temperature.

3.2.5 Overpressure Relief

To protect the system from being overpressurized by the compressor, bursting discs are provided at two points in the system. The first bursting disc is located in the bypass line at the bottom of the filter housing. This disc is vented vertically up through the building roof. The second disc is located on the dust injection pressure vessel and protects this vessel from overpressure. Both discs are sized to accommodate the maximum compressor flow. The discs are supplied by Fike and are 1.3 cm (1/2 in) nickel discs set to relieve at pressures in excess of 2152 kPa (300 psig).

Additional pressure relief is provided for by a pressure relief valve supplied on the compressor air receiver. This valve is set to relieve at pressures in excess of 1972 kPa (275 psig).

3.3 Pressure Containment Vessels

All the stainless steel piping that comprises the test passage has been fabricated according to ASME code Sect I for a design pressure of 1793 kPa (250 psig) and is stamped as indicated in the assembly drawing of Figure A3.13. The fabrication records and testing reports provided by the coded shop are appended to this section.

3.3.1 Combustor Section

The section of piping that contains the combustor is shown in Figure A3.14. This section has penetrations for

- Air inlet
- Dust inlet
- Igniter electrical lead through
- Flame detection device.

3.3.2 Inlet Sampling Section

A 1.2 m (4 ft) section of pipe is provided in the system for the injected dust and air to mix and be sampled. (Figure A3.15). This section has penetrations for:

- Temperature control thermocouple TE(1)
- High-temperature shutdown thermocouple TE(2)
- Dust-sampling port.

Hot dusty gases are confined to the 3.8 cm (1-1/2 in) stainless tube that forms the liner for this section and is shown in Figure A3.16. Insulation is provided by packing the space between the liner and pressure containment with fiber frax.

Figure A3.13 Hot Gas Cleanup System Assembly

Figure A3.14 Hot Gas Cleaning System-Combustor Section

Figure A3.15 Hot Gas Cleaning System-Straight Sampling Weldment

Figure A3.16 Hot Gas Cleaning System-Liner Weldment

3.3.3 Filter Housing

The actual filter housing is formed by a section of 25.4 cm (10 in) pipe with a cap at the bottom and a weld neck flange at the top, as shown in Figure A3.17. Wall temperatures are kept low through the use of an Inconel liner and fiber frax insulation, as indicated in the assembly drawing. Penetrations in this vessel include:

- Bypass flow
- Total pressure tap
- Two differential pressure taps
- Combustible gas analyzer tap
- Filter blowback inlet.

3.3.4 Outlet Sampling Section

To ensure that the outlet sample is representative of the gas penetrating the filter system, both temperature and pressure must be maintained to the point of sampling. For this reason the insulated outlet sampling section shown in Figure A3.18 has been provided. This section is lined and insulated in the same way as the inlet sampling section. The only penetration in this section is the indicated outlet sampling port.

3.3.5 Exhaust System Piping

The exhaust piping system serves the purpose of ducting the system exhaust to the roof and of cooling the gases before they flow through the back-pressure control valve. An isometric drawing of the piping system is shown in Figure A3.19. The exhaust system consists of two legs, one attached to the outlet sampling section and the other attached to the filter housing vessel. During normal filtration the bypass line from the bottom of the filter housing is closed by BBV(1) and flow is through the filter, up through the outlet sampling section, and into the exhaust piping. During the brief intervals when the filter

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Figure A3.18 Hot Gas Cleaning System-Outlet Sampling Section

Figure A3.19 301-1C13 Ceram Facility - Test Passage Exhaust System

element is being cleaned by a pulse of air from the clean side, flow is diverted around the filter by opening BBV(1). This serves the additional purpose of sweeping the dust expelled from the filter out of the system. When blowback is completed, BBV(1) is closed and flow and filtration resumed. The bypass valve BBV(1) is a Masoneilan Complex valve rated at 4003 kPa (600 psig) at 672 K (750°F). Water cooling will be provided for this valve. The actual exhaust system consists of 2.5 cm (1 in) stainless steel pipe centered inside 5.1 cm (2 in) stainless steel pipe with cooling water flow in the annular region.

This system detailed design was prepared by a consulting engineer, James Wolfe, who also helped prepare a very detailed bid specification package for outside fabricators. The following test procedure is part of the bid specification package:

TESTING

In cooperation with Westinghouse personnel and in accordance with the test specifications on Drawings 3SK-316 and 3SK-324, test the exhaust system as follows.

All hydro tests will be witnessed and verified by Westinghouse personnel. To expedite fabrication, Westinghouse personnel will travel to the mechanical contractors shop to witness the tests on Sections 1 and 2.

However, the final test will be conducted on site, after total assembly of the exhaust system, with the air compressor delivering the maximum operating pressure of 1793 kPa (250 psig).

The exhaust system has two (2) sections that will require hydro testing. Drawing 3SK-324 indicates the extent of each of these two (2) sections and how they will be isolated for containment of the test pressure for the hydro test.

Section 1 will require three (3) separate hydro tests.

Two (2) of the hydro tests will be performed at 2690 kPa (375 psig) on the same exhaust line piping.

- The first test will be made after the exhaust line piping is completed but before the cooling water jacket is installed.
- If this test is successful, the mechanical contractor will then complete Section 1 by installing the cooling water jacket.
- The second test will be made on the same exhaust line piping after the cooling water jacket has been installed.

The third hydro test will be performed at 717 kPa (100 psig) on the cooling water jacket.

Section 2 will also require three (3) separate hydro tests.

Two (2) of the hydro tests will be performed at 2690 kPa (375 psig) on the same exhaust line piping.

- The first test will be made after the exhaust line piping is completed but before the cooling water jacket is installed.
- If this test is successful, the mechanical contractor will then complete Section 2 by installing the cooling water jacket.
- The second test will be made on the same exhaust line piping after the cooling water jacket has been installed.

The third hydro test will be performed at 717 kPa (100 psig) on the cooling water jacket.

In the event that any section of the exhaust system being hydro tested does not meet the test specifications, all leaking joints will be located and repaired and then the system will be retested. This procedure will be followed until the system meets the test specification. At that time, the Westinghouse representative must be present to witness and verify that the test is successful.

After the exhaust system is completely assembled at the Westinghouse R&D site, the total assembly, including valves, flanges, and rupture disc, will have a final test performed with the air

compressor supplying an operating pressure of 1793 kPa (250 psig) as the test pressure.

During this final test all flanged, union, and threaded joints in the exhaust system will be checked with a leak-tek solution to verify the integrity of the exhaust system.

All leaking joints will be repaired and then the final test for the total assembly will be performed again.

After the air has expanded atmospheric pressure across the back-pressure valve, PCV(1), it flows into a vertical section of 10 cm (4 in) pipe through the roof to the small, highly efficient cyclone dust collector shown in Figure A3.20. This device will minimize the small amount of test dust vented to the atmosphere.

3.3.6 Dust Injection System

In order to test filters in the test passage, fine test dusts must be metered and dispersed into the pressurized test passage. A flow schematic for accomplishing this is shown in Figure A3.21. A fraction of the air from the compressor is split off and directed through the dust injector vessel by manipulation of GBV(1) and GBV(2). The flow through the dust injector is measured by a rotameter FM(1). Inside the dust injector dust is mechanically metered into an air ejector that is operated by the air through FM(1). The dust is dispersed by this high-velocity air and subsequently transported to the test passage through either BV(1) or BV(2). The path through BV(1) will cause the dust to pass through the combustor and flame, while BV(2) routes the dust upstream of the flame and avoids the possibility of momentary high temperatures for the dust particles.

The mechanical metering system is shown in Figure A3.22.

The test dust is placed in a hopper (Item A) inside the pressure chamber. A variable-speed motor rotates a plate at the bottom of the dust hopper. This plate (Item B) has a fixed number of through-

Figure A3.20 Cyclone

Dust Injection Loop				
Qty	Description	Manufacturer	Symbol	Remarks
2	Globe Valves 1.0" NPT 3/4" Dia. Port Forged Steel 800 psi at 850°F	Smith	GBV (1),(2)	
2	Ball Valves 1/4" NPT 7/16 Dia. Port Forged Steel 800 psi at 850°F	Smith	BV (1),(2)	
1	Pressure Gauge 0-300 psi Range	Trerice	PG (1)	
1	Flowmeter Range 1.5-15 scfm at 200 psig-200°F Range .93-9.3 scfm at 70 psig-200°F	Brooks	FM (1)	Calibration Data for Both Conditions
1	Rupture Disc 1/2" Dia. 300 psig Bursting Rate Greater Than Compressor Max. Output	Fike		Nickel Construction
1	1/4" Whitey Valve	Whitey	MV (1)	Bleed Down Valve For Dust Injection

Figure A3.21 Dust Injection Loop

Figure A3.22 Dust Injection System Assembly

holes at a given radius that fill with the test dust as it rotates through the hopper. The rotating plate is supported from below by a stationary plate that has a single through-hole at the same radius as the rotating plate. The hopper is shaped so that only this stationary hole position is outside the dust hopper, as seen in the plan view. As the rotating holes travel through the dust hopper they fill with dust and are scraped level as they pass through a felt pad dust seal just prior to lining up with the through-hole in the stationary plate. As the holes line up, the dust falls through the stationary plate into the air ejector below (Item C). Here high-pressure air disperses the individual pellets of dust and transports the dust to the test passage. For very adhesive dusts, provisions have been made for "puffing" the dust out of the rotating hole down into the ejector. This is accomplished by keying the puffer nozzle (Item D) to an external solenoid through the cam-and-microswitch assembly shown as Item E.

Dust can be reloaded through a 5.1 cm (2 in) ball valve located on the top blind flange. This allows long-term runs to be carried out without having to remove the rather cumbersome 41 cm (16 in) blind flange at each dust change.

The actual pressure vessel that houses the dust injector system was fabricated in a coded shop according to the requirements of ASME Section VIII Div. I. The vessel was designed for working pressures of 1793 kPa (250 psig) and has been stamped.

APPENDIX B
CERAMIC FILTER TEST SYSTEM OPERATING PROCEDURE DOCUMENT

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OP 1 - GENERAL OPERATING PROCEDURE

1.1 Start-up

- 1.1.1 Check proper valve position for all valves as per "Start-up Valve Position Checklist". OP 2.
- 1.1.2 Energize control panel with "power on" switch.
- 1.1.3 Check proper settings for all instrumentation in control panel as per "Start-up/Shutdown Instrumentation Checklist" OP 3.
- 1.1.4 Activate propane supply system as per "Fuel Supply Loop Start-up/Shutdown Procedure" OP 4.
- 1.1.5 Check house instrument air supply pressure of FCV(1), BBV(1), PCV(1), TCV(1).
- 1.1.6 With panel controls in manual mode stroke FCV(1), BBV(1), PCV(1), TCV(1), and visually confirm valve operation.
- 1.1.7 Start compressor/dryer system as per "Compressor/Dryer Start-up Procedure" and bring to pressure designated in daily test plan OP 5.
- 1.1.8 Open GV(4) and confirm flow of recirculating water in exhaust cooling lines.
- 1.1.9 Open GV(3).
- 1.1.10 In manual mode initiate flow through the passage with mass flow controller. Set initial flow at 0.25 kg/min (0.5 lbm air/min).

- 1.1.11 Initiate combustor operation as per "Combustor Light-off Procedure" OP 6.
- 1.1.12 Having established satisfactory combustor operation, increase mass flow, pressure, and temperature in increments so as to roughly maintain velocity in the passage.
- 1.1.13 Establish that flow, temperature, and pressure are those called for in daily test plan.
- 1.1.14 Initiate filter test by closure of BBV(1) and direction of flow through the filter media.

NOTE: Should a flameout occur it is mandatory to return passage to ambient pressure prior to light-off.

- 1.1.15 Initiate dust feed to test system as per "Dust Injection Start-up/Shutdown Procedure" OP 7.

1.2 Shutdown

- 1.2.1 Shutdown dust injection system as per "Dust Injector Start-up/Shutdown Procedure" OP 7.
- 1.2.2 Open BBV(1).
- 1.2.3 To initiate normal shutdown close TCV(1). Upon loss of flame the flame safety circuitry will close SV(1) OP 6.
- 1.2.4 Shutdown fuel supply loop as per "Fuel Supply Start-up/Shutdown Procedure" OP 4.
- 1.2.5 Alternately lower system pressure and mass flow in steps that roughly maintain passage velocity. Allow ambient pressure flow

at about 0.25 kg/min (0.5 lbm/min) until internal passage temperature is below 533 K (500°F).

- 1.2.6 Shutdown compressor/dryer system as per "Compressor/Dryer Start-up/Shutdown Procedure" OP 5.
- 1.2.7 Return control panel to shutdown condition as per "Instrumentation Start-up/Shutdown" Procedure OP 3.
- 1.2.8 Return all valves on "Start-up/Shutdown" checklist to indicated position OP 2.

Figure B1 Filter Test System Schematic

OP 2 - START-UP/SHUTDOWN VALVE POSITION CHECKLIST

<u>Valve</u>	<u>Position</u>	<u>Function</u>
GV(1)	0	Allows flow through orifice 1.
TV(0)	0	Isolates compressor outlet pressure transmitter
TV(1)	0	Orifice differential pressure tap
TV(2)	0	Orifice differential pressure tap
GV(2)	C	Isolates orifice plate not in use
TV(3)	C	Isolates pressure taps not in use
TV(4)	C	Isolates pressure taps not in use
FCV(1)	C	Flow control valve/operated from control panel
GBV(1)	C	Dust injector airflow control
GBV(2)	0	Dust injector/passage airflow split control
BV(1)	C	Dust feed to passage/through flame
BV(2)	C	Dust feed to passage/in front of flame
TV(5)	0	Pressure tap for system back-pressure control
TV(6)	0	Filter pressure differential tap
TV(7)	0	Filter pressure differential tap

TV(8)	C	Blowback pulse
SV(2)	O	Normally open solenoid closes to isolate DP cell during blowback
SV(3)	O	Normally open solenoid closes to isolate DP cell during blowback
SV(4)	C	Opens for filter blowback
GV(3)	C	Feed high pressure air for blowback
GV(4)	C	Feeds house recirc water to exhaust cooling pipe
TV(9)	C	Combustible gas analyzer tap
BBV(1)	O	Shunts flow through filter bypass loop
PTV(1)	C	Propane tank valve
CV(3)	C	Propane feed line valve
SV(1)	C	Normally closed propane solenoid
TCV(1)	C	Temperature control valve/operated from control panel
HIV(1)	O	House instrument air

OP 3 - INSTRUMENTATION START-UP/SHUTDOWN CHECKLIST

3.1 PASSAGE CONTROL/RECORDING

3.1.1 Temperature Controller - Indicate and set to 0

3.1.1.1 Flow Controller - Indicate and set to 0

3.1.1.2 System Pressure - Indicate and set to 0

3.1.1.3 Pressure and Pressure Drop Recorder - Indicate 0

3.1.2 PASSAGE STATUS

3.1.2.1 Passage Flow - Indicate 0 lbm/air min.

3.1.2.2 Passage Pressure - Indicates 0 psig

3.1.2.3 Filter Delta P - Indicates 0 in. wg

3.1.2.4 Passage Temperature - Indicates ambient

- o Set point indicates high-temperature limit specified in daily test plan

3.1.2.5 Cooling Water Temperature - Indicates ambient

- o Set point indicates 175°F

3.1.2.6 Over Temp Light - Off

3.1.2.7 PSI Low Light - Off

3.1.2.8 Over Temp Light (Cooling) - Off

3.1.3 COMBUSTION CONTROL

3.1.3.1 Purge Light - Off

3.1.3.2 Reset Light - Off

3.1.3.3 Ignition Light - Off

3.1.3.4 Flame On Light - Off

3.1.3.5 Flame Off Light - Off

3.1.4 DUST INJECTOR CONTROL

- 3.1.4.1 Motor Switch and Light - Off
- 3.1.4.2 Vibrator Switch and Light - Off
- 3.1.4.3 Pulse Switch - Off
- 3.1.4.4 Observation Light Switch - Off

3.1.5 FILTER BLOWBACK CONTROL

- 3.1.5.1 Bypass Valve Light and Switch On and Open
- 3.1.5.2 Blowback Timer - Set to pulse time called for in daily test plan
- 3.1.5.3 Pulse Light - Off

3.1.6 FUEL SUPPLY STATUS

- 3.1.6.1 Propane Temperature - Ambient Temp.
- 3.1.6.2 Feed Line Temperature - Ambient Temp.

OP 4 - FUEL SUPPLY LOOP

Elevated passage temperatures are achieved by direct combustion of propane gas in the system. During high-pressure passage operation, the propane supply needs to be warmed slightly in order to get the vapor pressure of the propane in the supply tank above the passage operating pressure. The vapor temperature is maintained from the tank to the combustor by means of heating tape. Propane is supplied in the 1F tank size which is normally charged to pressures of 20,685 kPa (3000 psi), which correspond to a propane temperature of $> 373 \text{ K}$ (212°F). DOT regulations require that the tanks be supplied with a pressure relief valve set for 2869 kPa (400 psig) [347 K (165°F)]. This pressure relief valve is vented through the ventilated hood to the roof. Ventilation of the enclosed hood area is provided by a $94 \text{ dm}^3/\text{s}$ (200 cfm) explosion proof fan. All contacts and relays for the water bath heater system are external to the enclosed area. Water bath temperature is controlled by an Omega temperature controller and redundancy is provided for with a Chromolox high limit control.

FUEL SUPPLY LOOP START-UP/SHUTDOWN

4.1 Start-up

- 4.1.1 Start exhaust fan.
- 4.1.2 Fill bath with water to within 15 cm (6 in) of top.
- 4.1.3 Set TC(1) to set point as per daily test plan.
- 4.1.4 Set high limit control TC(2) to 333 K (140°F).
- 4.1.5 Switch on heating tape power regulator to 339 K (150°F).
- 4.1.6 Open propane tank PTV(1).
- 4.1.7 Switch on water bath circulation pump.

4.2 Shutdown

4.2.1 Close PTV(1).

4.2.2 Switch off circulation pump.

4.2.3 Shut heater power off.

4.2.4 Shut heating tape power off.

4.2.5 Close CV(3).

4.2.6 Check to see SV(1) and TCV(1) are shut.

4.2.7 Shut off exhaust fan.

Fuel Supply Loop				
Qty	Description	Manufacturer	Symbol	Remarks
1	Shut Off Valve 1/4" - 73 CV-Stn. Stil	Whitely	PTV (1)	Mounted on Tank
1	Solenoid Valve Normally-Closed 1/4" NPT 1/8" Orifice 35 CV	ASCO	SV (1)	N.C. When Not Testing Shuts Off On: • Loss of Flame • High Temp. Air Flow Thru Comb • Receiver Low Pressure • High Cooling Water Temp
1	Temperature Control Valve-1/4" NPT Stn. Stil Body Air to Open CV-0081-32	Research Control Valve	TCV (1)	
2	Check Valve		CK (1) 6 (2)	
1	Thermocouple Type K Chromel-Alumel	Marlin	TE (1)	2500°F Range
1	Thermocouple Amplifier Temperature Transmitter 1-5 VDC Output	Veritrac	TT (1)	
1	Temperature Recorder Controller 1 Pen 3 Mode 1-5 VDC Input 4-20 mA Output	Veritrac	TRC (1)	Auto/Manual
1	Current/Pressure Transducer Input 4-20 mA Output 3-15 psi	Moore	I/P (2)	
1	Water Bath Temp Controller	Omega	TC (1)	Controls Water Bath Temp at 110 to 145°F
1	Hi Limit Control	Chromalox	TC (2)	High Lin
1	Pressure Relief Valve	Matheson	PRV (1)	Relieves at 400 ps Vents to Roo
1	Manual Shut Off Valve	Whitely	CV (3)	N.C. When Not in Operation
1	Heater Coil	Chromalox		240 V AC 5 kw
1	Water Bath	Plastic Piping Systems		50 Gallon Capacity Polypropylene Cons
1	Exh Fan w/Exp -Proof Motor	Acme Model FPM-2AE		Min. 200 CFM @ 1/2" SF
1	Flowmeter 0-100 %	Brooks	FM (2)	Propane Premix
1	Flowmeter 0-100 %	Brooks	FM (3)	Fireve Purge
2	Control Valve	Hoke	PPV (1) FPV (1)	Propane Premix Fireve Purge Control

Figure B2 Schematic of Propane System

OP 5 - COMPRESSOR/AIR SUPPLY

High-pressure air is supplied to the test passage by an Ingersoll Rand reciprocating compressor. This machine is equipped with a 29.9 kW (40 hp) electric motor drive and is air cooled. Pressure oscillations are damped by a $283 \text{ dm}^3/\text{s}$ ($10 \text{ ft}^3/\text{s}$) receiver which is skid mounted with the compressor package. Moisture and oil are removed from the air supply by a refrigerated air dryer/filter which will lower the dew point to 283 K (50°F).

5.1 Compressor Start-up

- 5.1.1 Check oil level and fill if required.
- 5.1.2 Check main breaker - off.
- 5.1.3 Check control switch - off.
- 5.1.4 Intercooler and receiver drain valves - closed.
- 5.1.5 Turn main breaker - on.
- 5.1.6 To start compressor turn control switch to "hand" position.
- 5.1.7 Open scale trap bleed for 20 seconds.
- 5.1.8 With compressor under load open intercooler bleed cock for 20 seconds.
- 5.1.9 With air receiver at operating pressure switch air dryer on and note that both green and red control lamps are lit. If red lamp does not go off within 15 minutes, check manual for operating problems.

5.2 Compressor Shutdown

- 5.2.1 Switch control to "off" position.
- 5.2.2 Switch main breaker to "off" position

5.2.3 Ensure GC(1) & GV(2) are closed.

5.2.4 Bleed down receiver pressure and then close bleed valve.

5.2.5 Switch off air dryer.

DIMENSIONS
Type 40 Packaged Air Compressor
Models 25 Thru 100

Form AP 0276 A
June 1977
Supersedes First Edition

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Ingersoll-Rand
AIR POWER COMPRESSOR DIVISION

Figure B3 -- Schematic of Air Compressor

OP 6 - COMBUSTOR LIGHT-OFF/SHUTDOWN PROCEDURE

6.1 Open CV(3).

6.1.1 Open TV(9) and adjust flow through combustible gas analyzer to $0.83 \text{ dm}^3/\text{s}$ (1.75 scfm) as measured by local rotameter.

6.1.2 Depress purge period start pushbutton (Line 2) and observe that the yellow purge cycle indicator (Line 6) illuminates. This indicates that the purge period cycle has started, and the purge period timer (PPT Line 7) should begin to time. (On Delay 1 to 10 min).

NOTE: During the purge period cycle, the operator should begin to flow process air through the test passage to purge any combustible gas that may be present. Observe the combustible gas indicator (Line 22) to ensure that the system has been thoroughly purged of combustible gases.

6.1.3 At the completion of the purge period cycle, the yellow purge cycle indicator (Line 6) should go out (N.C. PPT Contact Line 6 Opens), and the green ignition start O.K. indicator (Line 9) should come on (N.O. PPT Contact Line 10 Closes).

NOTE: Before continuing, ensure that the combustible gas indicator (Line 22) is not on. If it is, depress the purge period reset button (Line 2), then depress the purge period start pushbutton (Line 2) again and continue to flow air through the system until the combustible gas indicator (Line 22) goes out.

6.1.4 Depress the ignition start pushbutton (Line 4 and 10) FSO Relay (Line 4) is energized. (FSO N.O. Contacts Lines 5, 10 and 15 Close) Ignition Transformer (Line 12) is turned on. Power is also applied

to the control circuit of the fireye. (IS N.O. Contact Line 15 and FSO N.O. Contact Line 15) RL Relay (Line 17) in the fireye is energized and closes N.O. RL3 Contact (Line 15) energizing MBT (Line 15) and opens propane supply solenoid valve (Line 16).

6.1.5 If the flame is detected, RF2 (Line 14) in the fireye closes. MBP (Line 14) is energized and MBP N.O. Contact (Line 16) seals around IS Contact (Line 15) maintaining power through the fireye system to the propane solenoid valve. (SV-Line 16) MBP N.O. Contact (Line 8) seals around MBT Contact (Line 7) maintaining PPT (Line 7) in the energized state. MBT N.C. Contact (Line 10) will time open, disconnecting the ignition transformer (Line 12) and shutting off the ignition start indicator (Line 9), and the flame off indicator (Line 18) should go out. (RF3 N.C. Contact Opens) and the flame-on indicator (Line 19) should come on. (RF4 N.O. Contact Closes).

6.1.6 If the flame is not detected, MBT N.C. Contact (Line 10) opens, disconnecting the ignition transformer (Line 12) and resets system into a purge cycle. Process air should again be passed through the combustor to purge out any combustible gas that may be present. Monitor combustible gas indicator, wait for a complete purge cycle and the ignition start O.K. indicator to come on before trying to ignite again.

6.1.7 Once flame is established, close TV(9) and isolate combustible gas analyzer from passage.

OP 7 - DUST INJECTION

To meter and disperse test dusts to the filter system, a high-pressure dust injection system has been assembled. Within the pressure containment an electric, variable-speed-gear motor meters dust to an internal air ejector which disperses the dust and transports it to the test passage.

7.1 Dust Injector Start-up

- 7.1.1 With passage operating and compressor operating open GBV(1) and pressurized injector vessel.
- 7.1.2 Open either BV(1) or BV(2) as per daily test plan.
- 7.1.3 Adjust flow through dust injector by throttling GBV(1) and GBV(2) and setting flow meter FM(1) to value set in daily test plan.
- 7.1.4 Start gear motor and set to speed called for in daily test plan.

7.2 Dust Injection Shutdown

- 7.2.1 Shut off gear motor.
- 7.2.2 Shut off GBV(1).
- 7.2.3 Shut off BV(1) and BV(2).
- 7.2.4 Bleed down vessel pressure through MV(1).

7.3 Dust Hopper Refill

- 7.3.1 Shut off gear motor.
- 7.3.2 Close GBV(1).
- 7.3.3 Close BV(1) and BV(2).

- 7.3.4 Open MV(1) and bleed dust injection vessel pressure to 0 psig.
- 7.3.5 Confirm no internal pressure on PG(1).
- 7.3.6 Unlock fill port ball valve on top flange.
- 7.3.7 Open fill port valve and charge required amount of dust into internal hopper.
- 7.3.8 Close fill port valve and lock.
- 7.3.9 Close MV(1).
- 7.3.10 Open GBV(1) and bring vessel pressure back up.
- 7.3.11 Open either BV(1) and BV(2) and adjust flow.
- 7.3.12 Start gear motor.

APPENDIX C
FAILURE ASSESSMENT FOR CERAMIC FILTER TEST SYSTEM

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1. GENERAL FACILITIES FAILURES

1.1 Power Loss

The system is designed so that loss of electric power will cause all systems to fail safe:

- o Fuel Supply - SV(1) and TCV(1) will close
- o Air Supply - compressor will shut down
 - flow control valve FCV(1) will close
 - back pressure will be relieved as PCV(1) opens.

1.2 Control Air Supply

The control valves in the system are air operated and will fail to the safe position upon loss of supply pressure:

- o Fuel Supply - TCV(1) closes and shuts fuel flow
- o Air Supply - FCV(1) closes and stops airflow from compressor
- o Pressure Control - PCV(1) opens and relieves passage pressure.

1.3 Recirculating Cooling Water Failure

The exhaust system is cooled by means of jacketed pipe cooled by recirculating water. Should this flow stop while the passage is hot, the water in the jacket will warm and be sensed by TSH(2), an indicating temperature switch that is mounted in the panel. When the TSH(2) set point is exceeded, the safety circuitry automatically shuts off fuel flow.

2. FUEL SUPPLY SYSTEM

2.1 Water Bath Temperature Control Failure

The propane water bath temperature is controlled by an indicating temperature controller mounted in the control panel. Normally the system is quite stable because of the large thermal mass of the 227 dm³

Figure C1 Filter Test System Schematic

(60 gal) of water. Full power heating rates are only about $10^{\circ}\text{F}/30\text{ min.}$ Should the primary controller TC(1) fail, a second, redundant system is included by placing of TC(2), a high-temperature switch, in the system. Once TC(2) has tripped it must be reset manually prior to continuing to heat.

2.2 Propane Overpressure Protection

The propane bottles are supplied with a tank shut-off valve that has an integral pressure relief valve. This valve is set to relieve at 3227 kPa (450 psig) as per DOT and CGA regulations and is installed on all volatile liquids in lieu of a fusible plug. A second pressure relief valve PRV(1) is included in the system and is to be set at 2152 kPa (300 psig). Should propane be vented through the pressure relief valves, it is vented through a pipe that carries the propane through the roof of the building.

2.3 Ventilation Failure

To ensure that small leaks of propane do not result in an accumulation of propane in the hood, the hood is continuously ventilated with a $94.4\text{ dm}^3/\text{s}$ (200 cfm) explosion-proof fan. Fan and air inlet are positioned low in the hood so that the lower part of the hood is continuously swept by air. A pressure switch will be mounted in the fan exhaust line which will indicate loss of exhaust airflow by switching a red warning light on.

2.4 Fuel Shut-Off Valve Failure

The main fuel flow shut-off capability is provided by SV(1), which is a normally closed, explosion-proof valve. Should this valve stick open, a second system will shut off propane flow. This is accomplished by plumbing a 3-way solenoid valve in the supply air for the temperature control valve TCV(1). When power to SV(1) is cut, it also activates the solenoid valve on TCV(1) in such a way as to bleed

off the control air supply. This causes TCV(1) to return to its normally closed position.

3. PASSAGE TEMPERATURE CONTROL

3.1 Thermocouple Failure

Passage temperature is controlled by sensing temperature with thermocouple TE(1). If TE(1) fails, the control system perceives this as a high temperature and closes TCV(1). As a back-up TE(2) is positioned next to TE(1) and is wired to a control-panel-mounted, temperature-indicating switch, TSH(1). When a reading exceeds the high-temperature set point of TSH(1), the safety circuitry cuts power to the fuel supply shut-off valve, SV(1), closing it and TCV(1) simultaneously.

3.2 Control Valve Failure

As mentioned previously, the temperature control valve fails to the closed position. Should it be in an uncontrolled open position, the high temperature shutdown circuit (TSH(2)) will shut off the supply solenoid valve SV(1).

4. FLAME-SENSING SYSTEM

A standard combustion control has been incorporated for proving a flame at the combustor. This serves as a necessary requirement for propane flow during normal operation.

4.1 Fireye System during Normal Operation

During normal operation infrared radiation from the propane is sensed by a Fireye flame safety instrument. An optical path between the sensor and the flameholder has been incorporated into the combustion section. A relay [which opens SV(1)] is held closed by the photo cell output. When the loss of flame is sensed, SV(1) is closed within 2.5 seconds. Worst case calculations have been carried out that

indicate that even at the maximum expected propane flow, not enough propane can enter the system in the 2.5 second shut-off period to cause more than a 690 kPa (100 psi) rise in pressure. This rise from the maximum normal 1076 to 1793 kPa (150 psig to 250 psig) is well within the pressure capability of the system.

4.2 Start-up Interlocks

During start-up propane flow is required prior to proven flame in order to get the system going. The operation procedure document provided a detailed description of the start-up safety circuitry, which is condensed here.

- A timed purge cycle (1-10 min) is required before being able to start ignition in the system.
- Completion of the purge cycle enables the operator to initiate propane flow and igniter operation simultaneously. A second timer allows both spark and propane flow for only a set time, ~3-4 seconds. If a flame is not proved by the Fireye in this time, the fuel flow and igniter are shut off, and the system is automatically placed in the timed purge mode. This purge period must again be satisfied before starting the ignition sequence.

5. PASSAGE PRESSURE CONTROL

The passage and all components have been designed for working pressures of 1793 kPa (250 psig), although normal operation will be at 1076 kPa (150 psig).

5.1 Back-Pressure Control Valve Failure

Passage pressure is maintained by the back-pressure control valve PCV(1), which fails to the open position upon loss of power or control air. Should this valve or the exhaust line become plugged, the passage would rise to the compressor high-pressure set point ~1757 kPa (245 psig). This is well within the system's rated pressure capacity. As flow stopped, the propane supply would be shut down by either loss of flame or the high-temperature control and/or switch. The compressor

receiver is equipped with a pressure relief valve set at 1936 kPa (270 psig). Additionally the passage is fitted with a 1.3 cm (0.5 in) bursting disc which is plumbed to exhaust through the roof as passage pressures exceed 1793 kPa (250 psig).

6. DUST INJECTION SYSTEM FAILURE

6.1 Overpressure Protection

The major protection from overpressure in the system is the intrinsic design and fabrication techniques. All 2151 kPa (300 psig) fittings used have pressure ratings around 5020 kPa (700 psig). The pressurizing unit can only produce 1793 kPa (250 psig) flows. However, in addition to this safety factor a bursting disc and vent sized to relieve full compressor flow has been installed on the dust injector. This disc is rated at 2152 kPa (300 psig).

6.2 Accidental Venting during Dust Refill

We anticipate that dust will have to be loaded into the dust injector on a regular schedule. In order to avoid having to remove the 20 bolts securing the top flange each time, a ball valve arrangement has been incorporated in the design to allow test dusts to be loaded into the depressurized unit. To reduce the possibility of accidentally opening this ball valve while the system is pressurized, a pressure gauge has been mounted adjacent to the valve handle and a lock installed on the handle.

APPENDIX D

CERAMIC FILTER BACKFLUSH MODEL

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1. SUMMARY

The cross-flow ceramic filter is being developed as a final particulate removal device for gas streams from pressurized fluidized-bed combustion (PFBC) processes. During operation of the ceramic filter a dust cake is formed on the filter and this cake must be periodically removed by imposing a force on the cake greater than and opposite to the adhesion force binding the cake to the filter. This is done by occasionally raising the pressure on the clean side of the filter to a level higher than the normal system operating pressure by backflushing with high pressure gas from a storage tank into the clean side plenum. The backflushing gas entrains additional gas from the clean side environment into the clean side plenum, and if gas losses from the plenum through the opening at the top of the plenum and through the filter itself are kept lower than the total rate of gas introduction the pressure will rise to the necessary level for filter cake removal.

This dynamic and complex situation is modeled mathematically using transient mass, momentum, and energy balances applied to the various parts of the system. Procedures for characterizing filter and cake resistance to gas flow, and for predicting the necessary reverse pressure drop for cake removal are developed. Adiabatic, compressible flow relationships are used to relate backflush gas instantaneous flow to blowback system design. A momentum balance and turbulent jet analysis are used to estimate the rate of entrainment of additional gas by the motive backflush gas. The various sub-models are tied together in a numerical model that treats each blowback event as a step-wise transient process,

with pressures and gas flows varying with both time and location until such point as either the pressure on the clean side of the filter is high enough to dislodge the cake or the pressure begins to fall. In using the model all system design and operating parameters are specified and the model predicts what minimum pulse tank pressure is needed to clean the filter. Parameters used to calibrate the model were derived from actual ceramic filter laboratory tests at high temperature and pressure.

The model was used to evaluate performance of a scaled-up ceramic filter module for several design options including location after primary or secondary cyclones, and system operating pressure. It was found that a low resistance nozzle system is needed and that best performance results from using a large cross-section clean side plenum and by allowing a thick dust cake to accumulate between pulses. For example, the model was run to determine the blowback parameters for the full scale module used by Burns and Roe in the economic study of this filtering concept. It was determined that the following combination of parameters would give successful cleaning:

Given

System Pressure	703 kPa (87 psig)
System Temperature	1173 K (1650°F)
Inlet Loading	1200 PPM
Face Velocity	3 m/min (10 ft/min)

The model calculated

Plenum cross-sectional area	0.052 m ² (81 in ²)
Cycle Duration	1.6 hr
Nozzle Diameter	0.076 m (3 in)
Blowback Tank Pressure	3.5 MPa (35 atm)
Blowback gas requirement per pulse	3.4 kg (7.5 lbm)

An important unresolved question was identified, that being the effect of dust particle size on the adhesion force between cake and filter.

2. INTRODUCTION

In pressurized fluidized bed combustion (PFBC) the fuel gas stream must be cleaned of particulates prior to entering a gas turbine. One or two stages of cyclone dust collectors are used to remove much of the larger particles but further cleaning to remove the remaining fine dust is necessary. Positive filtration for this purpose is preferred, and one approach that is being investigated is the use of cross-flow ceramic membrane filter. This device consists of closely packed, thin, low- porosity ceramic wafers through which the gas is forced. This particular type of ceramic filter offers the advantages of compactness, excellent filtration efficiency, low pressure drops, temperature and thermal shock resistance, and corrosion resistance.

A filter cake collects during operation and periodically this cake must be removed. The cake is held to the filter surface by an adhesion force that must be overcome. This can be achieved by imposing a reverse flow of gas periodically for short times to push the cake off and allow it to settle to the vessel's floor for subsequent removal. A gas at higher than system pressure is used to cause this reverse flow. This blowback gas cannot be plumbed directly to the clean side of the filter without obstructing the normal flow of gas in the opposite direction unless filter duplication and the use of large high-temperature, high-pressure, and corrosion-resistant diverter and shut-off valves is possible. Consequently separated flow is used to pump up the gas pressure on the clean side of the plenum in excess of the system pressure and thereby cause cleaning. A high velocity blowback gas jet is delivered periodically into the clean side plenum from a high-pressure gas tank through a solenoid-valved nozzle. The blowback gas entrains surrounding gas to aid in its role of pumping up the pressure on the clean side of the filter. At the same time gas is lost through

the filter and out the top of the plenum. This dynamic situation must be clearly understood so we can design blowback systems such that the rate of gas delivery to the plenum will exceed the rate of loss causing the pressure in the clean side plenum to rise suitably.

A dynamic model of the ceramic filter blowback system has been developed to aid in the examination of the interactions among design parameters and to help in the specification of these parameters for scaleup. This model treats the pressures and gas flows in the blowback system as time-dependent variables during a single pulse for a given geometry, initial conditions, and cake thickness and properties.

In the following sections an outline of how blowback works is given. Then procedures for estimating filter and cake resistances and the necessary reverse flow pressure drop; and mass, momentum and energy balances for various parts of the blowback system are developed. The numerical solution procedure is outlined. Finally the model is used to examine scale-up of laboratory test data to a commercial scale filter module for a specific PFBC application.

3. OVERVIEW

Figure 1 shows the ceramic filter blowback system as represented in this model, along with pressure notation for later reference. Dust laden gas enters the dirty side of the vessel and flows through the ceramic filter elements to the clean side of the vessel, being stripped of particulate as it does so. Only one element is depicted in Figure 1 although there would be multiple elements per module in a real system. The dirty gas pressure (P_f) is assumed to be a constant throughout both the filtration and cleaning cycles. The clean side plenum pressure (P_4) falls as a dust cake accumulates until some predetermined pressure drop ($P_f - P_4$) is reached, at which point the blowback portion of the cycle is begun. The clean side vessel pressure (P_0) is assumed to equal P_4 during the filtration portion of the cycle.

Prior to initiation of the blowback cycle, a large tank, the pulse gas accumulator, is filled with gas at some relatively high pressure (P_{bf}). The blowback cycle is begun by quickly opening the solenoid valve between the accumulator and the clean side plenum, causing pulse gas to rush into the plenum at a rate determined by frictional resistance in the nozzle line and by the pressures at either end, limited in the extreme of low resistance to critical flow. It is assumed that the rate of discharge of gas from the accumulator is much higher than the rate of re-supply of gas to the accumulator, so the latter is neglected. The pressure in the accumulator falls from P_{bf} with time, and its time-dependent value is represented by P_1 .

High velocity gas leaves the nozzle and enters the plenum at P_2 . This gas forms a turbulent jet that entrains additional gas from the clean side vessel according to the momentum conservation requirement. Together the gas from these two sources and from the dirty side (while

Figure D1 - Schematic of Ceramic Filter Blowback System Showing Pressure Notation

$P_f > P_4$) causes the pressure in the plenum to rise until such point as there is a reverse flow of gas out through the filter elements that is from the clean side to the dirty chamber. This reverse gas flow exerts a drag force on the filter cake that gradually increases as the flow and P_4 increase. When the drag force reaches a certain level the cake is separated and the filter is assumed to have been cleaned.

It cannot be assumed that the pressure P_4 will necessarily rise to a level such that the cake is removed. In fact P_4 will rise only so long as the volumetric rate of flow of gas into the clean side plenum exceeds the rate of flow out of it. Both the rates of motive gas flow and the dependent entrained gas flow decline with time as the accumulator tank pressure falls. In addition, at some point P_4 will be high enough that gas is lost into the clean side vessel rather than entrained from it. (It is assumed that P_o remains at the initial value of P_4 throughout the pulse.) Similarly, eventually P_4 will exceed P_f and gas will be lost through the filter. Thus P_4 will reach a maximum at some point during a pulse and then begin to fall. If the cake has not been separated by this time, the cleaning pulse will be ineffective. An additional complication is added by the fact that if relatively cold blowback gas is used, it will cause a cooling of the gas in the clean side plenum and consequently less of a pressure rise than would have been realized had hot blowback gas been used.

It is clear that the dynamic situation is a complex one that can only be analyzed by transient mass, momentum, and energy balances on the system. An analytical solution is impractical so a numerical solution has been devised, whereby the pressures, flows, and temperatures at each point in the system are calculated for finite time increments from the initial conditions. Below are presented the models of the individual phenomena that constitute the overall model, then the overall model solution procedure.

4. MODEL DEVELOPMENT

4.1 RESISTANCE OF FILTER WALL TO FLOW

The filter pores are assumed to be so small that low Reynolds number flow through them occurs. In that case

$$\Delta P_w = R_w \mu v \quad (1)$$

where ΔP_w = pressure drop of gas across filter, Pa

R_w = resistance coefficient, m^{-1}

μ = gas viscosity, kg/m.s

v = superficial gas velocity through filter, m/s

In Equation 1, R_w is characteristic of the filter material, geometry, and manufacture, and is independent of gas properties, so it can be determined from a laboratory flow test on the filter in question and used in scale-up evaluation. Such a resistance measurement should be made on a filter which had previously been caked with appropriate dust, but which has been backflush-cleaned prior to resistance measurement. This is because such a filter will have retained some dust and will have a greater flow resistance than a never-dirty filter.

Equation 1 applies to flow in either direction.

4.2 RESISTANCE OF FILTER CAKE TO FLOW

The interparticle channels in the filter cake are also assumed to be small enough that low Reynolds number flow occurs, so

$$\Delta P_c = R_c \mu v h / d_p^2 \quad (2)$$

where ΔP_c = pressure drop across filter cake, Pa
 R_c = resistance coefficient, dimensionless
 h = thickness of cake, m
 $\bar{d}_p$ = dust particle size, m

The dust cake thickness is

$$h = t U_o W / \rho_b$$

where t = time into filtration cycle (since last successful blowback), s
 U_o = filtration face velocity, m/s
 W = solids loading in gas, kg/m³
 ρ_b = cake bulk density, kg/m³

R_c can be measured with a laboratory test using a particular filter and dust with Equations 2 and 3, in which case ΔP_c is the pressure drop across the filter at some point into the filtration cycle in excess of the pressure drop across the just cleaned filter, ΔP_w .

Equation 2 can be used for flow in either direction. For filtration $v = U_o$, while for blowback $v \neq U_o$ "t" is the time between blowbacks.

In Equation 2 the mean particle size is

$$\bar{d}_p = 1 / \sum_i (x_i / d_{p_i}) \quad (4)$$

where x_i = weight fraction of dust particles in a small size range
 d_{p_i} = mean size in that range, m

4.3 CAKE BREAKAGE PRESSURE DROP

When P_4 is greater than P_f (Figure 1) during blowback, part of this pressure differential will be exerted on the filter wall (Eq. 1) and part as drag on the filter cake (Eq. 2). The cake is attached to

the filter surface by the forces of adhesion. When the drag force on the cake exceeds the cohesion force, the cake will be broken off its retaining filter surface.

The magnitude of this adhesion force can be estimated from experimental data by measuring the minimum pressure drop across the filter (clean side plenum to dirty side vessel) that is required to cause cleaning and estimating (with the help of Eqs. 1 and 2) the portion of this that is across the cake rather than the filter wall:

$$\Delta P_b = \Delta P_{tot} \times \frac{R_c h / d_p^2}{R_c h / d_p^2 + R_w} \quad (5)$$

where ΔP_b = pressure drop across cake needed to remove cake during backflush, Pa

ΔP_{tot} = pressure drop measured across filter and cake required to remove cake, Pa

Equation 5 provides a method of estimating the pressure drop required to remove a cake based on experimental data, however, this ΔP may have to be corrected if the dust is different in size. Zimon¹ shows that the dependence of adhesive force on particle size may be proportional, inverse, or nonexistent depending on conditions, definitions, and method of measurement. The consensus, however, is that a single particle is attached to a plane surface by the force²

$$F_a = B d_p \quad (6)$$

where F_a = single particle adhesion force, N

B = breakage constant, N/m

d_p = particle diameter, m

The total force holding N particles on a surface is then

$$F = \sum_i N_i B d_{p_i} \quad (7)$$

where N_i = number of particles of size, d_{p_i}
 d_{p_i} = diameter of particle, m

The total area occupied by these particles is approximately

$$A = \sum_i N_i d_{p_i}^2 \quad (8)$$

Combining Eqs. 7 and 8 we see that

$$\Delta P_b = \frac{F}{A} = B/d_p' \quad (9)$$

where

$$d_p' = \frac{\sum_i N_i d_{p_i}^2}{\sum_i N_i d_{p_i}} \quad (10)$$

Since

$$w_i \propto N_i d_{p_i}^3 \quad (11)$$

$$d_p' = \frac{\sum_i (w_i/d_{p_i})}{\sum_i (w_i/d_{p_i}^2)} \quad (12)$$

where w_i = weight fraction of particles of size d_{p_i}

We therefore define a breakage constant B by Eqs. 9, 5, and 12. This constant can be used to scale up to other conditions and particle sizes.

Equation 9 indicates that the cake separation pressure drop will be inversely proportional to particle size. We cannot be confident that this will be the case based upon many reported discrepancies.¹ Intuitively it might be felt that at small particle sizes the cake approaches a plane in geometry and that therefore the adhesive force per unit area becomes independent of particle size. In this study we have examined the implications of both possibilities -- inverse proportionality and independence of force with regard to dust size. A

ceramic filter study at relevant conditions using dusts of various particle sizes in separate tests is needed to conclusively resolve this important question.

4.4 BACKFLUSH GAS FLOW

The solenoid valve is assumed to open completely and instantly at the start of a backflush cleaning cycle. Flow from the pulse accumulator into the clean side plenum is then assumed to occur at a rate given by the steady-state adiabatic flow relationship in Perry.³ Here the maximum rate of discharge through a frictionless nozzle is

$$G_{2m} = A_n P_1 \sqrt{M/2.72 RT_1} \quad (13)$$

where G_{2m} = maximum flow from nozzle, kg/s

A_n = nozzle cross-sectional area, m^2

M = molecular weight of gas

R = ideal gas constant, 8314 J/kg-mol:K

P_1 = pressure in accumulator, Pa

T_1 = temperature in accumulator, °K

The true flow, G_2 , is then given as a graphical function (G_2/G_{2m}) of the number of velocity heads lost in the accumulator to nozzle path (N_n), the ratio P_2/P_1 , and the ratio of specific heats, assumed to be 1.4 in the present work. It is first assumed that $P_2 = P_3$ (see Fig. 1), but if the resulting G_2 is greater than critical flow P_2 must be re-estimated as higher than P_3 to get a new, realistic G_2 corresponding to critical flow. The nozzle discharge temperature T_2 (as T_2/T_1) is read from the same charts. For computational purposes the design charts have been transformed into algebraic relationships in this model. Nozzle gas velocity v_2 is then calculated given G_2 , T_2 , P_2 and A_n .

4.5 ENTRAINMENT OF GAS AT NOZZLE

The high velocity gas issuing from the nozzle entrains gas from the clean side vessel into the clean side plenum, thereby causing the pressure in the latter location to rise. The rate of entrainment is

$$G_3 = \rho_3 (A_p - A_n) \sqrt{\frac{2 (P_o - P_3)}{\rho_3 N_c}} \quad (14)$$

where G_3 = rate of flow of gas into plenum from clean side vessel, kg/s

A_p = plenum cross-sectional area of plenum, m^2

P_o = pressure in clean side vessel, Pa

P_3 = pressure at entrance to plenum, Pa

N_c = number of velocity heads lost by gas contracting suddenly to enter plenum

ρ_3 = gas density at entrance to plenum, calculated at P_3, T_3 , kg/m^3

It is possible that P_3 will rise so that gas is expelled from the plenum rather than being entrained into it. In that case:

$$G_3 = - \rho_3 (A_p - A_n) \sqrt{\frac{2 (P_3 - P_o)}{\rho_3 N_e}} \quad (15)$$

where N_e = number of velocity loads lost by gas expanding out of plenum

In the calculations made for this program we have assumed that $N_c = 0.5$ and $N_e = 1.0$. When calculating gas density

$$\begin{aligned} T_3 &= T_o & \text{if } P_3 < P_o \\ T_3 &= T_4 & \text{if } P_3 > P_o \end{aligned} \quad (16)$$

where T_3 = temperature at entrance to plenum, $^{\circ}K$

T_o = temperature in clean side vessel, $^{\circ}K$

(assumed equal to temperature of filtered gas)

T_4 = temperature in clean side plenum, $^{\circ}K$

In addition, we have assumed that the maximum possible value of G_3 is given by the well known empirical expression for expansion of turbulent jets in unconfined environments:⁴

$$\frac{G_{3m}}{G_2} = 0.32 \left(\frac{X}{d_n} \right) \sqrt{\frac{\rho_3}{\rho_2}} \quad (17)$$

where G_{3m} = maximum rate of gas entrainment, kg/s

G_2 = nozzle gas flow, kg/s

X = distance jet travels before striking wall of plenum, m

d_n = nozzle inside diameter, m

ρ_2 = gas density of gas issuing from nozzle, kg/m³

The distance parameter X is calculated using system geometry and an assumed jet half-angle of 5.4°, i.e.,

$$X = \frac{\sqrt{1.27 \frac{A_p}{A_n}} - d_n/2}{\tan 5.4^\circ} \quad (18)$$

It is assumed that the jet strikes the wall above the top filter element. During the development of this model it was found that if the jet strikes the wall part way along an element, the upper part of that element cannot be cleaned.

Entering gas velocity v_3 is determined last from G_3 , $A_p - A_n$, and ρ_3 .

4.6 MOMENTUM BALANCE ON NOZZLE REGION

A mixing zone is assumed to exist at the top of the clean side plenum. Gas enters from the nozzle at rate G_2 and by entrainment at rate G_3 . Gas leaves at the rate $G_2 + G_3$ into the plenum proper. The mixture temperature T is given by an energy balance around the mixing zone (neglecting the kinetic energy of all streams except that of the nozzle gas):

$$(T' - T_2) C_p (G_2 + G_3) = \frac{1}{2} G_2 v_2^2 + G_3 C_p (T_3 - T_2) \quad (19)$$

(if $G_3 > 0$ and $G_2 + G_3 > 0$)

or

$$(T' - T_2) C_p = \frac{1}{2} G_2 v_2^2 \quad (20)$$

(if $G_3 < 0$ and $G_2 + G_3 > 0$)

or

$$T' = T_4 \quad (\text{if } G_2 + G_3 < 0) \quad (21)$$

where T' = temperature of gas leaving zone, entering plenum proper, K

T_2 = temperature of gas leaving nozzle, K

C_p = gas heat capacity (constant pressure), J/kg.K

v_2 = velocity of gas leaving nozzle, m/s

Then a velocity is defined as

$$v' = \frac{(G_2 + G_3)}{\rho' A_p} \quad (22)$$

where v' = velocity of gas leaving mixing zone and entering plenum proper, m/s

ρ' = gas density defined at T' and P_4 , kg/m³

A momentum balance relates pressures:

$$P_3 (A_p - A_n) = G_2 (v_2 - v') + G_3 (v_3 - v') + P_2 A_n - P_4 A_p \quad (23)$$

where v_3 = velocity of gas entering plenum from clean side vessel, m/s

P_4 = pressure in plenum, Pa

P_2 = pressure of gas issuing from nozzle, Pa

4.7 TANK BLOWDOWN EFFECT

As the pulse proceeds the pressure in the blowback accumulator will fall, effecting the rate of discharge of gas from the nozzle as well as all other pressures and flows in the overall system. An

unsteady-state energy balance on the accumulator is used to model this effect by giving the new tank pressure and temperature after a small time increment during which gas is lost at rate G_2 .

First, from a mass balance we know that

$$\rho_{1n} = \rho_1 - \frac{G_2 \Delta t}{V} \quad (24)$$

where ρ_{1n} = gas density in accumulator at conclusion of short time interval, kg/m^3

ρ_1 = gas density in accumulator at start of time interval, kg/m^3

Δt = length of time increment, s

V = volume of pulse accumulator, m^3

The energy balance on the accumulator is

$$\begin{aligned} \rho_{1n} V [C_p (T_{1n} - T_r) - R T_{1n}] - \rho_1 V [C_p (T_1 - T_r) - R T_1] \\ = - G_2 \Delta t [C_p (T_2 - T_r) + v_2^2 / 2] \end{aligned} \quad (25)$$

where T_r = reference temperature, $^{\circ}\text{K}$

T_{1n} = temperature in accumulator at conclusion of interval, $^{\circ}\text{K}$

T_1 = temperature in accumulator at start of interval, $^{\circ}\text{K}$

Equation (25) can be solved for T_{1n} , which in conjunction with ρ_{1n} can be used to determine P_{1n} , the new tank pressure at the conclusion of the interval.

4.8 PLENUM PRESSURE RISE

Gas enters the plenum at a rate $G_2 + G_3$ (sections 4.4 and 4.5 above) and at temperature T' (section 4.6). Gas leaves the plenum through the filter at a rate:

$$G_{\text{out}} = (P_4 - P_f) A_f \rho_f / \mu_4 R_T \quad (\text{if } P_4 > P_f) \quad (26)$$

or
$$G_{\text{out}} = (P_4 - P_f) A_f \rho_f / \mu_4 R_T \quad (\text{if } P_4 < P_f) \quad (27)$$

where G_{out} = rate of gas flow from clean side plenum to dirty side vessel through filter, kg/s

P_f = pressure of gas on dirty side of filter, Pa

A_f = filter area available on module, m^2

ρ_4 = density of gas evaluated at T_4 , P_4

μ_4 = viscosity of gas evaluated at T_4

ρ_f = density of gas evaluated at T_f , P_f

μ_f = viscosity of gas evaluated at T_f

T_f = temperature of gas on dirty side of filter, K

R_T = total flow resistance of filter and cake, m^{-1}

The flow resistance R_T is (from Eqs. 1 through 3):

$$R_T = R_w + \frac{U_o (t_f + t_{bf})}{\rho_B \frac{d}{p}^2} R_c \quad (28)$$

where t_f = length of filtration portion of cycle, s

t_{bf} = length of backflush portion of cycle, s

The temperature of the outlet gas is

$$\begin{aligned} T_{out} &= T_4 & (\text{if } P_4 > P_f) \\ T_{out} &= T_f & (\text{if } P_4 < P_f) \end{aligned} \quad (29)$$

An unsteady-state energy balance on the plenum now gives

$$\begin{aligned} &[\rho_4 A_p L_p + (G_2 + G_3 - G_{out})\Delta t][C_p(T_{4N} - T_r) - RT_{4N}] \\ &= C_p(T' - T_r)(G_2 + G_3)\Delta t - C_p(T_{out} - T_r)G_{out}\Delta t \\ &+ \rho_4 A_p L_p [C_p(T_4 - T_r)RT_4] \end{aligned} \quad (30)$$

where ρ_4 = density in plenum before time increment, kg/m^3

L_p = plenum length, m

T_{4N} = temperature in plenum after time increment, K
 T_4 = temperature in plenum before time increment, K
 T_{out} = temperature of gas leaving through filter, K

Equation (30) can be solved to give the new temperature in the plenum after a short time increment.

$$P_{4N} = [\rho_4 A_p L_p + (G_2 + G_3 - G_{out})\Delta t] \times \frac{T_{4N} R}{M A_p L_p} \quad (31)$$

where P_{4N} = pressure in plenum after time increment, Pa

5. NUMERICAL MODEL

The model is applied to a single ceramic filter module consisting of any number of elements attached to a single plenum. Table 1 lists the input parameters for the program. All input parameters are converted to consistent SI units (kg, m, s, K) early in the program. The program finds the minimum accumulator tank pressure required to clean the filter given the input parameters of Table 1.

The following is a step-by-step outline of the solution procedure

1. Guess tank initial pressure (P_{bf}) of 1.0 MPa.
2. Increase tank pressure (P_{bf}) by 0.5 MPa.
3. Calculate simple derived parameters-gas density (at T_f , P_f), viscosity, cleaned filter pressure drop (Eq. 1), and the pressure drop across the cake needed to lift the cake from the filter (Eq. 9). In addition the time between backflush cycles is estimated with Eqs. 2 and 3:

$$t_f = \frac{(\Delta P_t - R_w \mu U_o) \bar{d}_p^2 \rho_B}{\mu U_o^2 W R_c} \quad (32)$$

4. At the start of backflushing it is assumed that $P_o = P_3 = P_4 = P_f - \Delta P_t$, $T_4 = T_f$, $T_1 = T_{bf}$, and that $P_1 = P_{bf}$. Also set the current length of the backflush portion of the cycle, t_c , to zero.

5. Increase t_c , the backflush cycle length so far, by a small increment Δt . Guess what P_4 will be at the end of this increment

$$P_{4G} = P_4 + \frac{dP_4}{dt} \times \Delta t \quad (33)$$

Table D1

Input Parameters

<u>Symbol</u>	<u>Definition</u>	<u>Units</u>
U_o	filter face velocity	cm/s
P_f	dirty gas pressure	kPa
T_f	dirty gas temperature	K
M	gas molecular weight	-
W	solids loading	$\frac{g}{m^3}$ (at T_f, P_f)
ρ_B	density of filter cake	kg/m^3
$\bar{d}_p$	mean particle size of dust (Eq. 4)	μm
d_p	mean particle size of dust (Eq. 12)	μm
R_w	resistance coefficient of ceramic filter wall	m^{-1}
R_c	resistance coefficient of cake	-
B	cake breakage coefficient	kg/s^2
A_f	surface area of filter media	m^2
A_p	cross-sectional area of clean side plenum	m^2
L_p	length of clean side plenum	m
d_n	nozzle inside diameter	cm
V	pulse chamber volume	m^3
N_n	number of velocity heads lost by gas in nozzle piping	-
t_{bf}	backflush pulse duration	s
T_{bf}	temperature in pulse chamber at start of pulse	K
ΔP_t	filter pressure drop that activates blowback cycle	kPa

where P_{4G} = guessed value of P_4 at end of increment Δt
 $\frac{dP_4}{dt}$ = rate of change of P_4 in last increment, calculated in step
 17 below. For the first increment in each cycle assume
 this to be zero

6. Determine the average of P_4 and P_{4G} to be used in place of P_4 in the equations for steps 12 to 15 below.

7. Guess the temperature (T_{1G}) and pressure (P_{1G}) of the gas in the backflush accumulator at the end of the time increment to be the same as at the beginning of the increment, namely T_1 and P_1 . Calculate the gas density (ρ_1) in the accumulator at T_1 , P_1 .

8. Determine the average of T_1 and T_{1G} and of P_1 and P_{1G} to be used instead of T_1 and P_1 in step 10 below.

9. Guess the average value of P_3 during the cycle (P_{3G}) to be that of the last increment (or $P_f - \Delta P_t$ for the first increment).

10. Guess $P_2 = P_{3G}$ and calculate G_2 , T_2 , ρ_2 , and v_2 (see section 4.4 above). (Base calculation on average temperature and pressure in accumulator as calculated in step 8.) Compare G_2 to the flow which would occur if critical flow through the nozzle occurred. If G_2 is larger, then increase P_2 and repeat this step. If not, proceed to step 11.

11. Determine T_3 , ρ_3 , G_3 , and v_3 (see section 4.5 above).

12. Determine T' , ρ' , v' (section 4.6 above). (Use the average plenum pressure from step 6 to calculate v' .)

13. Calculate average P_3 in time increment from Eq. 23. (Use the average plenum pressure from step 6 in this calculation.) If P_3 equals the guessed value P_{3G} , proceed to step 14. If not, guess a new P_{3G} and go to step 10.

14. Determine ρ_{1n} , T_{1n} , P_{1n} at end of time increment (section 4.7). Compare P_{1n} with P_{1G} and if they are equal proceed to step 15. If not guess new values for P_{1G} and T_{1G} and proceed to step 8.

15. Set $P_1 = P_{1n}$ and $T_1 = T_{1n}$. Calculate G_{out} and T_{out} according to section 4.8 (using the average value of P_4 from step 6).

16. Determine the pressure (P_{4N}) and temperature (T_{4N}) in the plenum at the conclusion of the time increment. Compare P_{4n} with P_{4G} and if they are equal proceed to step 17. Otherwise guess a new P_{4G} and proceed to step 6.

17. Set P_4 equal to P_{4n} and T_4 equal to T_{4n} . Determine (dP_4/dt) for the just completed time increment.

18. If P_4 has been falling for the past five time increments (and if there have been more than five increments) or if the total length of the current backflush cycle t_c exceeds t_{bf} then declare the current value of the tank initial pressure to be insufficient and proceed to step 2. Otherwise proceed to step 19.

19. If P_4 has risen sufficiently to separate the dust cake from the filter, then declare the current value of the tank initial pressure to be sufficient and proceed to step 20. P_4 is high enough if $P_4 > P_f$ and if

$$P_4 - \left(\frac{R_w \mu_4 G_{out}}{A_f \rho_4} \right) > P_f + \Delta P_t \quad (34)$$

Otherwise proceed to step 5.

20. End.

6. PARAMETER ESTIMATION FROM TEST DATA

Three key parameters were estimated from laboratory test results on the ceramic filter. In a test (1.5 mm backwall, 100 μm pore size crossflow filter) at 1089°K (1500°F) and 1135 kPa (150 psig) with a face velocity of 0.029 m/s (5.7 ft/min), a pressure drop of 1 kPa (4 inches water) was measured across the filter immediately after backflushing at the end of a cycle well into the apparently stable cycle portion of the run. Estimating a viscosity of 4.4×10^{-5} kg/m:s (2.9×10^{-5} lb_m/ft sec), Eq. 1 gives

$$R_W = 7.87 \times 10^8 \text{ m}^{-1} \text{ (} 2.40 \times 10^8 \text{ ft}^{-1} \text{)}$$

In the same test the filter was routinely cleaned after the pressure drop had risen to 3 kPa (12 inches water) so the pressure drop across the cake must have been 2 kPa (8 inches water) at this point. The bulk density of the cake is estimated as 1000 kg/m³ and the mean particle size as 5 μm . Approximately 1 hr was required for the pressure drop to rise to the trigger level after the last cleaning and the dust loading was about 14.5 g/m³ (6.3 gr/acf). Equations (2) and (3) were used with these parameter values to obtain:

$$R_C = 26.1$$

It was found that approximately a 24.1 kPa (3.5 psid) pressure drop across the filter and cake was needed during blowback to separate the cake. From Eq. 5 we find that this corresponds to 16.2 kPa (2.35 psid) across the cake alone. The mean dust particle size from Eq. 12 is 3.2 μm so from Eq. 9

$$B = 0.051 \text{ kg/s}^2 \text{ (} 0.112 \text{ lb}_m/\text{s}^2 \text{)}$$

The model was applied to the test case using these calculated parameters and others listed in Table 2. The result was a prediction that a pulse air accumulator pressure of about 3.5 MPa (35 atm) was necessary to achieve cleaning of the filter cake. This agreed with experimental observation and was interpreted as model verification.

Table D2

Parameter Values

<u>Parameter</u>	<u>Units</u>	<u>Test Case</u>	<u>Scale-up</u>	
			<u>Case I</u>	<u>Case II</u>
U_o	cm/s (ft/min)	2.9 (5.7)	5.08 (10.0)	
P_f	kPa (psig)	1135 (150)	1135 (150)	
T_f	K (°F)	1089 (1500)	1089 (1500)	
M		29	28	
W	g/m^3 (gr/ft ³)	14.5 (6.3)	7 (3.0)	1.75 (0.76)
ρ_B	kg/m^3 (lb _m /ft ³)	1000 (62)	1000 (62)	1000 (62)
$\bar{d}_p$	μ_m	5.0	4.8	1.46
d'_p	μ_m	3.18	2.07	0.93
R_W	m^{-1} (ft ⁻¹)	7.87×10^8 (2.4×10^8)	7.87×10^8 (2.4×10^8)	
R_c	-	26.1	26.1	
B	kg/s^2 (lb _m /s ²)	0.051 (0.112)	0.051 (0.112)	
A_f	m^2 (ft ²)	0.102 (1.1)	52.8 (568)	
A_p	m^2 (in ²)	0.0015 (2.3)	0.023 (36)	
L_p	m (ft)	0.29 (0.95)	3.66 (12)	
d_n	cm (in)	0.95 (0.375)	7.62 (3)	
V	m^3 (ft ³)	0.0021 (0.074)	1.0 (35.3)	
N_n		10	4	
t_{bf}	s	0.5	0.25	
T_{bf}	K (°F)	300 (80)	533 (500)	
ΔP_t	kPa (psid)	3.0 (0.44)	6.9 (1)	

7. SCALE-UP DESIGN PROJECTIONS

The model has been applied to a scale-up situation with parameters listed in Table 2. A higher filtration face velocity, 5.08 cm/s (10 ft/min) has been assumed as well as a larger module. The clean side plenum is 15.2 cm (6 in) square in cross-section and 3.66 m (12 ft) long. To it are attached 40 elements each containing 1.32 m² (14.2 ft²) filter surface. We have assumed that an accumulator volume of 1 m³ (35 ft³) is available per module, and that the gas in this tank is at 533°K (500°F). During initial testing of the model it became clear that in order to get efficient cleaning gas must be delivered rapidly through the nozzle, at a rate larger than it escapes through the filter and through the top of the plenum. This means that a relatively large nozzle diameter is needed, and so we have assumed a diameter of 7.62 cm (3 in). We have estimated that a 4 velocity head loss would occur in a reasonably well designed nozzle system.

As indicated in Table 2, two major cases have been considered. In case I the ceramic filter is placed after the primary cyclone. In case II an additional cyclone has been placed before the ceramic filter. As a result the loading is lower and the mean particle size smaller. Particle sizes in Table 2 have been calculated from size distributions of typical PFBC dusts at these points. The smaller dust size is very important if Eq. 9 applies because a much larger force will have to be imposed on the cake by the blowback system in order to remove the cake.

Model predictions for the two cases and variations are listed in Table 3. For case I we found (in Run A) that an unacceptably high tank pressure, 16.5 MPa (163 atm) was required. However, the required tank pressure could be reduced (Run B) to 5 MPa (49 atm) by increasing the plenum cross-section (A_p) from 15.2 cm (6 in) square to 22.9 cm (9 in)

square. This is because entrainment of more gas from the clean side vessel occurs with the large area. Alternatively (Run C) a higher trigger pressure drop (ΔP_t) can be used to reduce the required tank pressure. The ΔP_t is the value of the pressure drop across the filter that, when reached during the filtration portion of the cycle, causes the blowback cycle to begin. A ΔP_t translates into a longer cycle time and hence a thicker cake. A thicker cake reduces the rate of loss of gas out through the filter during blowback and hence allows the pressure to build up more rapidly across it. As shown in Run D a further, but marginal, improvement can be achieved by doing both these things.

In Runs A-D it was assumed that the pressure drop ΔP_b needed to overcome cake-to-filter adhesion forces was inversely proportional to dust particle size. This is merely a hypothesis at this point as ΔP_b may well be independent of dust size. Runs E and F were made to show what tank pressures are necessary if the latter assumption is true. As can be seen in Table 3 significant reduction in required backflush pressure relative to Runs A and B is predicted. This shows that it is very important to experimentally clarify the nature of adhesion force dependence on dust size for a ceramic filter in a PFBC environment.

Runs G through J show what pressures may be required for case II (Table 2). The dust size is much smaller than for case I, so if adhesion force is inversely proportional to particle size, much more vigorous blowback conditions are required. For the nominal A_p and ΔP_t of Table 2, Run G showed that a tank pressure in excess of 17 MPa (168 atm) is required. If both A_p and ΔP_t are increased as previously discussed then a 6.0 MPa (59 atm) tank pressure is needed. On the other hand, if adhesion force is independent of dust size then the same pressures are needed for case II as for case I (compare Runs I and J with E and F).

A final series of model Runs (K through O) at a lower system pressure, 703 kPa (87 psig) was made, at a variety of combinations of other assumptions as shown in Table 3. The required tank pressure is found to be roughly proportional to system pressure.

8. CONCLUSIONS

A model has been developed that allows the examination of ceramic filter blowback system design parameters and their interactions. Such an understanding is useful in the specification of scaled up system parameters.

A major uncertainty, that of the effect of dust particle size on the adhesion force between cake and filter wall, has been identified. The nature of this relationship will significantly effect extrapolation of the model to dusts other than those on which data is available. In particular it is not clear how a ceramic filter will perform on dusts penetrating a secondary cyclone. An experimental investigation of this question is needed.

Operation of the ceramic filter backflush system is improved when:

- o A larger diameter nozzle is used to allow delivery of more backflush gas in a shorter time.
- o The backflush nozzle piping is designed for low hydraulic resistance.
- o The rate of gas entrainment by the motive gas jet is maximized by using a wider cross-section plenum and reducing the hydraulic resistance to entrained gas flow.
- o A thicker filter cake is allowed to accumulate between pulses (at the cost of increased pressure drop during normal filtered gas flow).
- o The temperature of the motive gas is kept high, so as to minimize cooling of gas already in the plenum before the pulse.
- o The nozzle jet is designed to impact the plenum wall above the top-most filter element. Filter area above the jet impact point will not be cleaned.

REFERENCES

1. Zimon, A. D., Adhesion of Dust and Powder, second edition, New York, Consultants Bureau, 1982.
2. Dennis, R., et al., "Filtration Model for Coal Fly Ash with Glass Fabrics," GCA report to EPA, August 1977 (EPA-600/7- 77-084), p. 303.
3. Chemical Engineers' Handbook, Perry and Chilton, eds., Fifth edition, New York, McGraw-Hill Book Company, 1973, pp. 5-26 to 5-28.
4. Beer, J. M. and N. A. Chigier, Combustion Aerodynamics, London, Applied Science Publishers Ltd., 1972.

APPENDIX E

HOT GAS CLEANUP USING CERAMIC CROSS FLOW MEMBRANE FILTERS

TEST PROGRAM AT ANL'S PRESSURIZED FLUID BED COMBUSTION SYSTEM

Summary

A test program was planned and carried out where a ceramic cross flow filter was installed and operated in the Argonne National Laboratory's small scale pressurized fluid bed combustor. The purpose of this task was to confirm the effectiveness and operability of the cross flow filter in a real coal burning application, in order to compliment the lengthy bench scale development carried out to date.

The goals of this test were to accumulate approximately 100 hours of filtration time at pressures of 7-10 atm and at temperatures of 700-800°C. The filter was to be operated in the range of filter velocity from 1.8 to 3.6 m/min (6-12 ft/min) with blow back trigger pressures of less than 25 kPa (100 in Wg). Cycle times greater than 10 min were expected for particulate loadings of 1,000-5,000 ppm, and expected efficiencies were greater than 99.5%. We judged that if these goals could be met, and if the filter did not deteriorate mechanically, then the test would be 100% successful.

The Westinghouse filter pressure vessel was slightly modified and delivered to ANL for installation in their PFBC system. ANL made some minor modification to their exhaust piping to accommodate the cross flow system and installed it downstream of their first stage hot cyclone. The actual test program was carried out in two separate test sequences, one in August and the other in September. This occurred because of an operations problem with the PFBC system during the August test.

The results of the program are summarized below:

- o 100% filtration efficiency demonstrated throughout the test period.

- o Total filtration test time of about 90 hours, 20 hr at a filter velocity of 3.8 m/min and 70 hr at 1.9 m/min.
- o Stable operation of the filter at inlet loadings typically in the range of 400 to 1000 ppm but with upsets to loadings as high as 60,000 ppm.
- o Stable base line pressure drop in the range of 1.5 to 3.5 kPa (6 to 14 in Wg) (depending on velocity).
- o Cycle time on the order of 1 hr for steady operation between 1.75 and 3 kPa (7-12 in Wg) pressure drop at a velocity of 1.9 m/min (6.3 ft/min) and with loadings in the range of 300-900 ppm.
- o Successful demonstration of simple pulse jet cleaning.

These results generally exceeded our initial goals and expectations for the filter. The single disappointment of the test program was the fact that when the filter element was removed from the mounting assembly at the end of the test, it was found to have delaminated in a manner similar to that observed in other tests. This test experience reinforces our previous contention that the filter offers great potential for hot gas cleaning applications, but that some well directed effort at improvement in the bonding of the structure is necessary.

TASK 5
ANL Field Test Report

Preparation and ANL/Westinghouse Interface

At the end of March, 1983 Westinghouse received a DOE directive indicating that field testing of the cross flow filter should proceed. This testing was to be carried out at the Argonne National Laboratory's small-scale pressurized fluid bed combustion test loop. In response Westinghouse provided DOE the following test sequence outline after having reviewed the parameters with ANL.

Operating Test Time

Our goal was to accumulate 100 hrs of test time on the filter. Preferably this would be accomplished in a continuous run, but we felt that the quality of the testing would not be compromised if shutdowns occurred in the test sequence.

Operating Conditions

- o Temperatures - A range of 700 to 750°C was expected to be available.
- o Pressure - 6-7 atm
- o Filtration Velocity - 1.8 m/min (6 ft/min) was to be the base condition and we expected to operate 50 hrs at this velocity. The remaining 50 hrs we would attempt to operate at 3.6 m/min (12 ft/min).
- o Inlet Loading - We expected the inlet loading to range from 1,000 to 5,000 ppm and have a size distribution typical of the first stage cyclone exit.

Filter Element

We intended to use the same filter element that we used during the last several HTHP laboratory runs since it had already been conditioned and since we were able to recover it from the system without damage. This element is a 100 μ pore size, 1.5 mm thick wall specimen.

Sampling

A great deal of previous experience indicated that the cross flow is essentially absolute, and outlet sampling essentially serves the purpose of failure detection. We intended to take total mass filter samples at the cross flow outlet on a continuous basis. Because of the relatively high loading at the filter inlet, sampling was somewhat more difficult. We intended to do total mass loading sampling at the inlet as frequently as is experimentally possible. Size distribution data was to be obtained by Coulter Counter analysis of this catch.

The criteria by which we would judge the degree of success for this test program were established from consideration of turbine tolerance requirements, environmental constraints and economic analysis of capital and operating costs for PFBC applications. In our estimate the testing would be 100% successful if:

- o We could operate at filter velocities of 1.8 m/min with stable pressure drop cycles.
- o The average pressure drop was less than 25 kPa.
- o The cycle time at loadings between 1000 and 5000 ppm was greater than 10 min.
- o The overall efficiency was greater than 99.5%.
- o The post-test examination of the element reveals no significant chemical or physical degradation of the element.

ANL/Westinghouse Interface

Subsequent to the DOE request for an estimate of schedule and cost for this work, a meeting between Westinghouse and ANL personnel was held at ANL to plan the details of mechanically interfacing the Westinghouse cross flow filter vessel with the ANL combustion loop.

The existing test loop is shown schematically in Figure E1. Figures E2, E3, and E4 present photographs of various views of the ANL PFBC test loop prior to installation of the Westinghouse filter system.

After meeting with the team of personnel at Argonne National Laboratory and discussing the details of interfacing the cross flow filter with their test rig, it appeared that the task should be fairly straightforward and would require minimal changes in the existing system as shown in the accompanying sketch (Figure E5).

The agreed upon configuration allowed us the flexibility to operate the filter over virtually the whole range of combustor flows by varying the split between the filter leg and the presently existing ANL test leg.

Interfacing the filter vessel required the fabrication of a new filter inlet/sampling elbow section which connected the ANL primary cyclone to the Westinghouse filter vessel inlet (Figure E6). The present filter vessel required one modification in order to mate-up with the ANL dust removal hopper. This involved the installation of a 2" flanged section at the bottom of the filter vessel in place of the present 1" coupling. Westinghouse arranged for this and the elbow to be fabricated in a coded shop. ANL had to do some scaffolding and a relatively small amount of rearrangement of the 2" exhaust pipe to tie in the flow measuring sections and to route the filter exhaust back into the existing pressure control/exhaust section.

The instrumentation requirements for the testing were minimal in that much of what was required was currently installed at the test rig.

Figure E1 - Simplified Piping Schematic of Combustor Facility

Figure E2 - Argonne National Laboratories PFBC Test Rig

Figure E3 - Argonne National Laboratories PFBC Test Rig

Figure E4 - Argonne National Laboratories PFBC Test Rig

Figure E5 - Westinghouse ceramic cross flow filter test set up at Argonne National Laboratories test rig

Flanges and fittings Carbon Steel
ASTM - A105, ANSI B16.5

- Welds Require Liquid Penetrant Inspection

- Hydrostatic Test at 225 psig
at $72^{\circ}\text{F} \pm 10^{\circ}$

Figure E6 - Adapter Section to Interface Westinghouse Filter Vessel to ANL PFBC Cyclone Outlet

Testing of Cross Flow Filter

The actual testing of the cross flow filter at ANL occurred in two phases, one occurring in August, 1983 and another in September 1983. This happened because of an in-bed heat exchanger failure in the PFBC system during the August test. Testing resumed after repair of the system.

Phase I Testing at ANL

Phase I Summary

During the month of August the preliminary stage of filter testing was carried out at the Argonne National Laboratories' pressurized fluid bed coal combustor. During this test the filter was operated at a temperature of 725°C (1340°F) and a pressure of 9 atm (120 psig). Operating at a filtration velocity of 3.8 m/min (12.4 ft/min) the pressure drop cycle remained essentially constant rising from a base of 3.25 kPa (13 in Wg) to 5 kPa (20 in Wg) over a period of about 30 min. During this period the inlet dust concentration ranged from 400 to 600 ppm. Isokinetic samples from the filter outlet indicated that virtually no dust penetrated the filter system giving rise to 100% collection efficiency. After about 15 hours of relatively steady operation the combustor entered into an upset condition causing the loading entering the filter to increase to levels as high as 15,000-24,000 ppm. In spite of the very high loading, the base line pressure drop remained constant as did the absolute filtration efficiency. Cycle times decreased to about 5 minutes which made manual operation difficult, so the filter velocity was cut in half for the last 3 hours of this run. In addition, the blow back trigger pressure drop was allowed to increase from 5 to 6.5 kPa (20-26 in Wg). These adjustments allowed the cycle time to be expanded back to about 35 min.

After about 20 hours of operation it was determined that the heat exchanger in the fluid bed combustor had failed and a decision was made to terminate the test until repairs could be made. Post test

examination of the filter showed it to be in good condition in spite of the very heavy loadings that it had been exposed to. The filter was replaced in the pressure vessel and the system was made ready for further testing upon completion of repairs to the combustor.

Phase 1 Report

During the first week in August a Westinghouse technician was sent to Argonne National Laboratories to help with final preparation, and to install the cross flow filter in the filter housing. During this period the combustor was operated successfully at pressure and temperature in a short shakedown run. Figure E7 presents a photograph of the filter system installed in the ANL test loop.

Actual testing of the filter began slightly after midnight 8/8/83 with start up of the combustor systems. From midnight to about 6:00 a.m. the system was brought to stable operating conditions with the following nominal operating parameters:

Coal	Sewickley
Sorbent	Tymochtee Dolomite
Coal Feed	16 kg/hr
Sorbent Feed	9 kg/hr
Air Flow	163 kg/hr
Bed Velocity	.9 m/min
Temperature	950°C
Pressure	9 atm

At about 6:00 a.m. the filter was brought on line by opening the flow split control valve. The flow was gradually increased as the filter system heated up. After about 15 minutes the flow was adjusted to its nominal rate which yielded a filter velocity of 3.8 m/min (12.4 ft/min). The measured filter temperature continued to climb at a slow rate and reached a final value of about 725°C. This was about 200°C lower than the bed operating temperature and about 100°C lower

Figure E7 - Photographs of Cross Flow Filter Installed at ANL
Pressurized Fluid Bed Combustor

than we had anticipated. At the test velocity the filter pressure drop in the newly cleaned state was about 2.5 kPa (10 in Wg). This rose to the blow back trigger pressure drop of 4 kPa (16 in Wg) over about a 30 minute period as can be seen in Figure E8 which summarizes the pressure drop/time history of the test. The first few samples taken during this run were ruined because they were wet from condensation. After an hour of sampling the samplers were warm enough so that a good inlet/outlet sample was obtained. The inlet loading was taken over a 15 minute period and indicated a loading into the filter of about 1900 ppm. The outlet samples showed absolutely no trace of ash.

Over the first 3-1/2 hours of operation (5 cycles) the base line pressure drop remained constant. The blow back used consisted of a 0.5 second pulse from a 2.12 liter reservoir which was charged to a pressure of 19 atm with a standard bottle of Nitrogen. During each pulse the forward filter flow was stopped by closing the flow split control valve.

After about four hours of filter operation the coal feed line became plugged, and the system had to be shut down for a short while until the problem was remedied.

Upon restart of the system a slightly higher base line pressure drop was observed changing from about 2.5 to 3.25 kPa (10 to 13 in Wg). We now believe that this was most likely due to an error in reading the flow split flow meter, which was at the low end of the scale due to a somewhat oversized orificeplate. For the next 12 hours the combustor system and filter ran in a nice steady mode. Cycle times for the filter were 30-35 minutes for the pressure drop to rise from the base line of 3.25 kPa (13 in Wg) to the trigger pressure drop of 4 kPa (16 in Wg). The inlet to the filter was sampled periodically and indicated loading in the range of 400 to 700 ppm. The outlet was sampled almost continuously with samples spanning an hour or more. As before, absolutely no dust was observed on the outlet filter.

Fig. E8—Test 8-9.83 Pressure drop vs time for ceramic crossflow filter 1 WH-7 (100 μ pore size, 60 mil (1.5 mm) thick), Conditions: Temperature=1340°F (725°C), Pressure=120 psig (9 atm), $V_{\text{filters}} = 12.4 \text{ ft/min. (3.8 m/min)}$; Efficiency = 100 %

After about 12 hours of operation in this mode we began to notice a sudden rise in inlet loading to the filter. Loadings increased first to 2000 ppm, then to 15,000 ppm and one sample indicated a loading of 24,000 ppm. Concurrently the dust on the inlet sample became much more coarse with large particles of both bed material and unburned coal. Figure E9 presents a photograph of the samples obtained under steady and upset conditions. As a consequence of this high loading the filter cycle time began to decrease (see Figure E8 at about time = 950). Operation of the filter with a cycle time of about five minutes was continued for about two hours in hopes that the system upset would pass. It should be noted that even during this upset, a steady cycle and base line pressure drop was maintained.

After two hours of operation we decided to reduce the filter velocity to one half of its previous value and to raise the trigger pressure drop from 5 to 6.5 kPa (20 to 26 in Wg) in order to lengthen the cycle time. When this was done the base pressure drop decreased to 1.5 kPa (6 in Wg) and the cycle time stretched out to about 30-35 minutes. Operation in this mode was steady and continued for about three more hours. At this point ANL operating personnel determined that there had been a failure of one or more water cooled heat exchanger tubes in the bed. Because of this a decision was made to shut the system down for inspection.

After the bed had cooled the top of the combustor was removed and the tubes were removed. Several of the tubes had been burned off and it was apparent that a significant time would be required to repair them. At this point we decided to open the filter vessel and examine the filter. Figure E10 presents a photo of the "as removed" condition and of the filter after the surface dust had been removed for inspection. The "as removed" condition appeared to be similar to that observed during bench scale testing. We observed no cracks in the filter and the gasketing looked to be in good shape so we simply reinstalled the filter in its vessel for subsequent testing.

Figure E9 - Photograph of Typical Total Filter Samples Taken during Steady and Upset Combustor Conditions at ANL Preliminary Test

As Removed Condition

After Cleaning Surface of Filter for Inspection

Figure E10 - Filter Element 1WH-7 after Preliminary Testing at
ANL PFBC

We did examine some of the dust deposits on and around the filter and found some "lumpy" material in the dust (see Figure E11). This material was fairly hard but could easily be crushed between your fingers. We suspected that they may have been formed during the upset condition when it was possible for fairly large amounts of cooling water to leak into the bed.

A post test meeting with ANL personnel was held to decide what repairs were required and to enumerate minor changes that could enhance testing. At this time a plausible mechanism was provided that explained both the upset condition and the tube failure.

It was decided that the continuation of testing would probably occur after Labor Day.

Phase II Testing at ANL

Summary

Upon completion of the repair of the heat exchanger tubes and several minor systems modifications, a second phase of testing was undertaken at ANL. During this week long phase approximately 70 hours of actual filtration time were accumulated. During the week of testing we experienced one short interruption, (due to coal feed problems) and one long interruption lasting about a day and a half due to difficulties with the heat exchanger. Nominal test conditions were

$T = 725^{\circ}\text{C}$ (1340°F)

$P = 9 \text{ atm}$ (120 psig)

Filter = 1.9 m/min (6.3 ft/min)

Loading - 300-1000 ppm

During the entire test sequence the overall efficiency remained constant at 100% with absolutely no trace of ash penetration. Baseline pressure drop remained steady at 1.75 kPa (7 in Wg) and climbed to the blow back

Powdery dust with lumps in

Lumps separated from powder

Figure E11 - Deposits of Dust Lumps found on and Around Filter 1WH-7
after Preliminary Test at ANL

trigger pressure drop of 3 kPa (12 in Wg) over a cycle time that typically lasted about 1-1/2 hours.

During these tests "on line" blow back was successfully demonstrated. In this mode of operation the forward filtration was "not" stopped by closing the flow split control valve during pulse jet cleaning. This greatly simplified blow back procedure and did not impair in any way the effectiveness of the pulse jet cleaning. Higher temperature operation was also demonstrated during this phase. During the last 10 hour test period we successfully initiated operation of a methane injection system at the filter inlet, and we were able to control the filter temperature at 800°C instead of the previous 725°C.

We observed no unexpected or detrimental effects from the higher temperature. In particular, the cleanability of the filter remained excellent with no evidence of plugging by "sticky" ash.

Subsequent to shut down of the system, the filter was removed for inspection. It was found to have delaminated along two bonded surfaces. The failure was similar to others observed in bench scale simulator tests and indicates a need for the development of a sounder mechanical joint.

Phase II Report

During the week of 9/19/83 to 9/23/83 the second phase of testing was resumed at ANL. In the interim between the August Phase I testing and the Phase II testing several modifications had been implemented including:

- o repair of the heat exchange tubes
- o installation of new flow measuring orifice plates
- o installation of block and zero valves at all differential pressure instruments
- o installation of local gauge for filter differential pressure in addition to existing transducer

- o installation of solenoid valves on filter differential pressure lines and interlock with pulse system
- o repair of orifice thermocouples
- o installation of condensate knock out tubes in steam lines
- o retrace sample lines with heating tape
- o installation of a simple methane injection system for cyclone inlet to allow higher filter inlet temperatures.

Additionally, operating procedures were revised to help to minimize the chance of problems with wet coal which had caused feed interruptions.

With these changes in place start up of the system was begun at midnight 9/18/83 and the system was brought to steady conditions by about 5:00 a.m. 9/19/83. Nominal operating conditions of the combustor were:

Coal	Sewickley
Sorbent	Tymochtee Dolomite
Coal Feed	9 kg/hr
Sorbent Feed	7 kg/hr
Air Flow	160 kg/hr
Bed Velocity	.85 m/sec
Temperature	950°C
Pressure	9 atm

At about 5:00 a.m. the filter was gradually started by opening the flow split control valve on the filter inlet system. The temperature of the filter rose slowly to a steady value of about 725°C. Flow through the filter was adjusted to give a filter velocity of 1.9 m/min (6.3 ft/min) and the baseline pressure drop was observed to be a steady value of about 1.87 kPa (7.5 in wg). With measured nominal loading at the filter inlet of 300-1000 ppm the baseline drop

would rise to the trigger pressure drop of 3 kPa (12 in Wg) over roughly a 1-1/2 hour period (Figure E12). Blow back was accomplished by a 0.5 sec pulse jet from a 2.12 liter pulse reservoir charged to a pressure of 19 atm from a standard Nitrogen bottle. During this time forward flow through the filter was momentarily stopped for blow back. This was accomplished by merely shutting the filter flow split control valve.

About 5 hours into the test the methane injection system was tried. This system simply used a standard high pressure bottle of methane as a source for the injection. The methane was introduced in the refractory lined pipe between the fluid bed and the primary "hot" cyclone. Upon start of the flow of methane the hot cyclone inlet temperature was boosted from about 900°C to a temperature of 1050-1150°C. Unfortunately this did not give a large rise in filter temperature (720 to about 750°C) since the filter flow was a relatively low fraction of the total flow. The heat loss between the cyclone and the filter inlet essentially damped out the higher cyclone inlet temperature. After a fairly short time the methane injection tube became plugged, due to melting at the tip. The methane system was subsequently shut off and the run continued as before.

At this time sketches were prepared for a water cooled methane injection nozzle that was to be mounted right at the filter inlet in order to give precise control of the filter temperature. Materials were located and sent to the machine shop. Since installation of the injection port on the filter inlet required the system to be down, we decided to postpone installation at that time anticipating a forced shutdown for any other operating reason.

Operation in this mode remained steady for about the first 13 hours. Then at about $T = 800$ (Figure E12) an upset in the particulate loading occurred with concurrent measurements going from about 300 ppm to 8,000 to 61,000 to 13,000 and finally back to about 300 ppm over roughly a 1-1/2 hour period. We suspect that the high

Fig E12—Test 9, 19-20, 83 Pressure drop vs time for ceramic crossflow filter L W11-7 (100 μ pore size, 60 mil (1.5 mm) thick), Conditions: Temperature = 1340°F (725°C), Pressure = 120 psig (9 atm), $V_{\text{filter}} = 6.3 \text{ ft}^3/\text{min}$ (1.9 m^3/min); Efficiency = 100%

transient loading occurred as a result of an attempt to artificially increase the filter inlet loading by mixing ash from the primary cyclone in with the sorbent feed. Approximately two hours prior to the high dust loadings we charged the sorbent feed lock hopper with a mixture of 5 parts sorbent to 1 part ash from the primary cyclone. We do not understand why the ash came through the system in such a concentrated form over such a short period. Figure E13 presents the Coulter Counter size analysis of the dust during normal and upset condition. In spite of the tremendously high transient loadings the filter managed to maintain a stable and reasonable cycle.

After the upset subsided, the system returned to a stable mode of operation for about 4-1/2 hours. At this time a plug occurred in the coal feed line which required a brief shut down of the system (Figure E12 at $t = 1375$).

After about 2 hours the system was again on line and continued to operate in a steady fashion for the next 36 hours. During this entire time the inlet loading was relatively constant at 200-500 ppm with only a few occasional high measurements. Steady continuous cycles between 1.75 and 3 kPa (7 and 12 in Wg) were maintained with 100% collection efficiency.

During this period "on line" blow back was initiated (time = 3250, Figure E14). In this mode of operation forward filtering flow through the filter was not stopped for pulse jet cleaning; instead the pulse was administered and the system was immediately back in the filtration mode. This method is much simpler to implement and is greatly preferred since in a large multi-element system it eliminates the need for high temperature dampering of large hot flows. The cleaning effectiveness in this mode was as good as in the off line mode as reflected in the constancy of the cycles from $t = 2900$ to 3600 in Figure E14.

At about $T = 3630$ (Figure E14) the coal/sorbent transport line developed a leak because of erosion at a turn. Attempts were made to

Fig. E13—Particle size dist. for cross flow filter inlet at ANL test (steady & upset condition)

Fig. E14—Test 9. 20-21. 83 Pressure drop vs time for ceramic crossflow filter 1WH-7 (100 μ Pore Size, 60 ml (1.5 mm), thick), Conditions: Temperature = 1340°F (725°C), Pressure = 120 psig (9 atm), V_{filter} = 6.3 ft/min (1.9 m/min); Efficiency = 100 %

replace the damaged section without a complete shut down but after an entire night shift of unsuccessful attempts to restart, it was decided to take this opportunity to install the new methane injection system and to inspect the fluid bed and heat transfer surface. To this end the filter inlet elbow was removed and sent to the shop to have the injection port welded on. The bottom of the bed was dropped and some large clinkers were removed. At this time it was determined that three of the five heat exchange tubes were leaking. These were valved off and effectively taken from the system. Plugs in the remaining two lines were removed and their operation was confirmed. The entire system was reassembled by late Thursday afternoon (9/22/83), and the air preheaters were switched on. Because some of the heating elements had failed, it took a long time for the system to be heated to 400°C (a temperature at which coal injection could be started). By 6:00 a.m. Friday (9/23/83) the system was back to normal operating conditions and the filter was brought back on line for the final 10 hour test segment shown in Figure E15.

The filter system was allowed to equilibrate without methane injection for the first 3 hours of operation in order to establish a base line of operation. Then at about $T = 180$ (Figure E15) the methane injection was initiated. The flow of methane was difficult to control at first and a few brief excursions to temperatures in excess of 950°C occurred until control was established and a constant filter temperature was maintained at 800°C. A decision was made to maintain constant mass flow through the filter so that the filter velocity (and consequently pressure drop) increased slightly. Operation of the filter continued the rest of the day in a smooth steady fashion as shown in Figure E15. There was no evidence that the higher operating temperature of the filter in any way made cleaning more difficult or less effective. At 4:00 p.m. Friday (9/23/83) the system was shut down and left to cool prior to post-test inspection.

As the filter system was disassembled, the absolute filtration determined by outlet sampling was confirmed by the total absence of PFBC

Fig.E15—Test 9. 23. 83 Pressure drop vs time for ceramic crossflow filter 1WH-7 (100 μ pore size, 60 mil (1.5 mm) thick), Conditions: Temperature = 1475°F (800°C), Pressure = 120 psig (9 atm), $V_{\text{filter}} = 6.75 \text{ ft/min}$ (2.06 m/min); Efficiency = 100 %

ash on the clean side liner surfaces as indicated in Figure E16. Upon removal of the filter holder from the pressure vessel housing, nothing abnormal was observed and the element is shown in Figure E17. When some of the exterior dust was removed evidence of a crack was revealed as shown in Figure E17. When the element was removed from the holder, it became apparent that the filter had delaminated in two places as shown in Figure E18. This is somewhat puzzling since similar failures in past testing had resulted in a small but measurable penetration of dust.

This mechanical failure of the element occurred after element 1WH-7 had been exposed to approximately:

- o 110 hours hot filtration
- o 200 blow back pulses
- o 10 thermal cycles
- o 5 mounting/dismountings

We believe that this history is encouraging and that well directed efforts at better joint structure and module design have a high probability of resulting in an attractive hot gas filtration system.

Figure E16 - Clean Side Surfaces of Cross Flow Filter System after
Test at ANL

Figure E17 - As Removed Condition of 1WH-7 Cross Flow Filter after
Test at ANL

Figure E18 - Delamination Failures of Cross Flow Element 1WH-7 after
Test at ANL