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TESTING AND VERIFICATION OF GRANULAR BED FILTERS
FOR THE REMOVAL OF PARTICULATE AND ALKALIES

Fifth Quarterly Report for Period October 1—December 31, 1981

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March 18, 1982

Work Performed Under Contract No. AC21-80ET17093

Westinghouse Electric Corporation
Pittsburgh, Pennsylvania

TECHNICAL INFORMATION CENTER
UNITED STATES DEPARTMENT OF ENERGY

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ABSTRACT

The Westinghouse Electric Corporation with Ducon, Inc. and Burns and Roe, Inc. are conducting a test and evaluation program of a Granular Bed Filter (GBF) for gas cleaning applications in pressurized-fluidized bed combustion processes. This work is funded by DOE PRDA for Exploratory Research, Development, Testing and Evaluation of Systems or Devices for Hot Gas Clean-up.

This report describes the status of the testing of the subpilot scale GBF unit and test results and analysis from the bench scale alkali gettering work.

1. INTRODUCTION

The Westinghouse Electric Corporation with Ducon, Inc., and Burns and Roe, Inc. are conducting a test and evaluation program of the Ducon granular-bed filter (GBF) for gas cleaning applications in Pressurized-Fluidized Bed Combustor (PFBC) Power Plants. The granular bed filter system may serve to remove both particulate and alkali metal compounds.

Figure 1-1 shows a schematic diagram of one element of the sub-pilot-scale Ducon granular bed filter. The element shown consists of four parallel operating filter compartments. Each compartment contains a granular filter-bed through which the ash and dust laden gas pass (Figure 1-1a), depositing the ash and dust particles on the surface of the filter media. With increasing deposits of the particulates, the system pressure drop increases until a point is reached when it is no longer practical or economic to operate. The GBF system is then cleaned by sequentially backflushing each element (Figure 1-1b). For this purpose, a backflush motive air is introduced into an eductor at the outlet (clean-air side) of a filter element. The motive air passes through the eductor (inducing additional clean gas from the clean-air plenum) and into the GBF element housing, up through each filter bed at a velocity and rate sufficient to gently fluidize each bed and dislodge the accumulated ash and dust material. The dislodged ash is elutriated and expelled from the element by passing back through the inlet opening provided at the top of each compartment. Figure 1-1 shows a four bed element. Commercial scale GBF systems would consist of multiple elements each comprised of ten, twelve, or more parallel operating beds.

The major technical objectives of this work are:

GBF ELEMENT - FILTRATION MODE

GBF ELEMENT - CLEANING MODE

Fig. 1-1 - Schematic representation of Ducon GBF element

1. The design and testing of a subpilot scale, multi-element (6) GBF to operate to temperatures of 900°C (1650°F) and pressures to 16 atm., at approximately 500 acfm.
2. Test and analysis to define the design approach and filter operation to accomplish alkali removal in PFBC.
3. The performance, economic and technical evaluation of a GBF hot gas clean-up system as it would be integrated in PFBC power plant conceptual designs.

The work reported herein covers the period October 1, 1981 through December 31, 1981. Prior work on this contract is documented in References 1, 2, 3, and 4.

2. TASK 1.1 - DEFINITION AND PRELIMINARY EVALUATION OF PFBC PLANT CONCEPTS AND HOT GAS CLEAN-UP REQUIREMENTS

A draft copy of the reference PFBC system description write-ups has been prepared by Burns and Roe. This work will be issued as a separate topical report when the technical review has been completed. The report covers the overall plant descriptions, performance and costs.

In this study, three existing PFBC plant designs were modified to conform to the overall criterion and assumptions set forth in this study,⁽¹⁾ and to include a GBF for hot gas particulate clean-up and a fixed granular bed for alkali control. The specification of each plant design has included the following considerations:

1. Mass and energy balance
2. Overall plant performance
3. Sketch of plant layout
4. Estimate of plant costs
5. Conceptual design for GBF
6. Conceptual design for the alkali gettering bed
7. GBF and alkali bed operational characteristics
8. Estimate of turbine maintenance intervals

Submittal of the Burns & Roe report and its review and issue as a topical report completes the Task 1 work scope. A summary of the results from this study are provided in Reference 4.

3. TASK 1.2 - DESIGN AND TESTING OF THE SUBPILOT SCALE DUCON GRANULAR BED FILTER

This task encompasses the specification, design and testing of a multiple-element Ducon GBF pilot scale unit. Testing is presently in progress and is being conducted at the Westinghouse Synthetic Fuels Division's Hot Gas Test Facility at their Test and Development Center located at Madison, PA at the Westinghouse Waltz Mill site. Figure 3-1 shows a schematic diagram of the test facility in which redispersed fly ash (or other dusts) are fed at high temperature and pressure into the test vessel containing the GBF pilot unit. Provisions for gas sampling are available on both the inlet and exit lines of the vessel. Figure 3-2 shows a photograph of the test facility. Figure 3-3 shows a schematic of the GBF test unit with associated instrumentation and controls.

3.1 Specification of the GBF Subpilot Unit Design

A preliminary draft of the design and operating criteria for the proposed GBF pilot unit has been compiled and reported, Reference 1. The report provides the basis for specifying or calculating the major design and operating parameters for the GBF. This work has been updated to include an analysis of the design and operation of the eductor used in the GBF backflush system for high pressures and high temperatures, (Reference 2). These studies have formed the basis for the GBF subpilot test unit design.

3.2 GBF Subpilot Unit Design and Fabrication

Final design of the Ducon subpilot scale GBF unit has been completed. Drawings and description of the test unit assembly were included in Reference 2. The drawing identifications are:

Fig. 3-1—Schematic diagram of W-hot gas clean-up facility

Fig. 3-2 - Hot Gas Clean-up Test Facility

Figure 3-3 - GBF Pilot Unit Backflush System
and Control and Instrumentation
Diagram

Drawing No.	Title
K80-0750-1	● Arrangement - Six Element GBF Pilot Unit Adapted to Existing Westinghouse Pressure Vessel
K80-0750-2	● Assembly and Detail - Granular Bed Filter Element Four Compartment
SK80-0750-1	● Detail-machined Ring with Dished Plenum Plate and Eductor-Granular Bed Filter Pilot Model
SK80-0750-2	● Detail-baffle Plate Granular Bed Filter
SK80-0750-3	● Detail-eductor - Granular Bed Filter
SK80-0750-4	● Detail-dolly - Granular Bed Filter
SK80-0750-5	● Detail-deflector Plate - Granular Bed Filter

The overall design basis for this unit is premised on a flow of 500 acfm at 1650°F (900°C) temperature and operation at 10 atm pressure with a maximum filter pressure drop during the filtration cycle of 5 psi. This filter pressure drop criteria actually translates to a maximum operating pressure differential of 9 psi that will occur when one element is backflushed (i.e., during the cleaning cycle).

The fabrication and assembly of the GBF test unit was completed last report period (Reference 4) and the unit delivered to the Westinghouse SFD-TDC facility site. As reported last quarterly (Reference 4), flexible-braded tubing was incorporated into the backflush lines in lieu of the 3/4 inch diameter tubing originally provided. The revised backflush tubing arrangement is shown in Figure 3-4. Figure 3-5 shows an overall view of the sub-pilot scale GBF test unit. During this report period effort in the program has been to characterize the backflush system and the fluidization of the individual beds in cold flow tests. Results of these studies are reported.

3.3 GBF Cold Flow Backflush Tests and Results

Utilizing the six (6) element, 24-bed subpilot GBF test unit shown in Figure 3-5, a series of ambient flow tests have been conducted to accomplish the following:

Figure 3-4 - Top view of GBF subpilot test unit showing the internal backflush piping lines, pressure tap leads and the applied thermal insulation on the inner portion of the support flange.

Figure 3-5 - View of GBF subpilot test unit being lifted to test facility pressure vessel. View shows applied thermal insulation around the outside of the support flange.

- (a) Develop a correlation between the measured eductor pressure differential and total element mass flow. At test operating conditions, this data will serve as the basis for calculating total backflush flow per GBF element;
- (b) Measure the distributor plate/screen pressure drop as a function of backflush velocity. The characterization of the distributor plate/screen pressure drop is necessary to quantify the backflush flow distribution between parallel operating beds (filter compartments);
- (c) Measure and observe the fluidization characteristics of an operating element with appropriate bed media. These tests provide visual conformation of the fluidization behavior of each bed in the operating element and appropriate test data to develop the fluidization curves, and;
- (d) Measure and observe the elutriation characteristics of the bed media under backflush fluidization conditions. These tests can provide the basis for setting bounds to operate the test unit under cleaning conditions.

Figure 3-6 shows a schematic drawing of the test arrangement and the flow and pressure drop measurements made in these cold flow tests. The four beds (filter compartments) are each identified (A,B,C, and D). Pressure taps are provided across the eductor throat and expansion section. A single pressure tap is also provided in each bed located about 1/4 to 1/2 inch above the distributor plate/screen section. Also, in these tests a pressure tube was inserted into the plenum section of element. With this test instrument arrangement, the following pressure drop measurements could be made:

- ΔP_p : pressure difference across the GBF element
- ΔP_E : pressure drop across the eductor section
- ΔP_S : pressure difference between the plenum and bed pressure tap
- ΔP_B : differential in pressure between two selected operating beds
- ΔP_o : pressure drop across the calibrated orifice in the backflush, motive flow line

In these tests all pressure drop measurements were made using water manometers (vertical and inclined) and/or calibrated pressure differential gauges.

Fig. 3-6—Schematic diagram of GBF test arrangement for cold-flow backflush tests

3.3.1 Correlation of GBF Eductor Flow

Using the test arrangement illustrated in Figure 3-6 a series of test runs were made (without bed media) where the motive flow (m_m) was varied and the corresponding eductor pressure drop (ΔP_E) measured. For these tests, the inlet section of the eductor tube was plugged such that the discharge from the backflush nozzle entered the eductor section but could not induce additional flow from the ambient surroundings. Thus, the only flow entering the venturi section was that discharged from the backflush nozzle (motive flow). This motive flow was measured independently through a calibrated orifice plate located in the motive flow line. Figure 3-7 shows the measured eductor (venturi section) ΔP as a function of the measured motive flow. The solid line is drawn through the data with an assumed square-root pressure drop dependence in accordance with usual venturi-equations:

$$m \sim K \Delta P^{(1/2)} \quad (1)$$

where K is the nozzle coefficient. For a well designed venturi and inlet section, the usual value for K is expected to be 0.98.

For the GBF eductor operating in the backflush mode and with the relatively short inlet section, the K value for the design is determined from the test data. The test represent data points taken on at least two different GBF elements. The data give, from Equation (1), $K = 0.73$. The eductor mass flow on backflush is determined from the correlation

$$M_{E,BF} = 0.2982 \sqrt{\Delta P_E \rho_g} \quad (2)$$

where ΔP_E is the observed eductor pressure drop in inch of H_2O and ρ_g the gas density in (lb/ft^3) .

The eductor flow correlation for backflush given by Equation (2) is not apriori assumed valid for calculating the element flow in the

Fig. 3-7—Correlation of eductor backflush flow

filtration direction. For filtration (which cannot be evaluated in these cold flow tests) a $K = 0.98$ value is assumed.

3.3.2 Eductor Efficiency

Assuming the validity of Equation (2), the efficiency of the eductor section can be evaluated by repeating the above described test series but with the plug removed from the eductor inlet. In this test series, the motive flow discharges through the nozzle into the inlet of the eductor thereby inducing additional flow from the ambient surroundings. The effectiveness of the eductor is determined by the ratio of induced flow to motive flow, i.e., M_g/M_m . Table 3-1 summarizes the calculated eductor effectiveness for several motive flow values for the ambient tests conducted and compares these with calculated values based on the eductor model described in Reference 2. The measured flow ratios ranged from 1.33 to 1.73 while the values for the cold flow tests calculated from the model were less. To account for the discrepancy in the observed and calculated flow ratios, only a relatively small (about 15%) error in either the measured motive or total flow values would need to occur. Induced to motive flow ratios for the PFBC test conditions are also shown for different filter pressure drop and backflush flow pressures. The predicted flow ratios range from 0.52 to 1.0 and depend on the operating pressure drop across the filter unit.

3.3.3 Filter Media Fluidization Characterization

Distributor Plate and Screen Pressure Drop Correlation

The fluidization characteristics of the filter bed are determined by the bed media and distributor plate/screen pressure drop. Initial flow tests were done to measure the distributor plate/screen pressure drop characteristics and these values were compared with the correlation recommended by Zenz (1). For the distributor plate and screen pressure drop, Zenz recommends

$$\Delta p_s = \frac{\rho_g v^2}{(C_{DF})^2 2g_c} \quad (3)$$

Table 3-1

MEASURED EDUCTOR EFFICIENCY AND COMPARISON WITH PFBC TEST CONDITION PROJECTIONS

Cold Flow Tests

Backflush Motive Flow Pressure (psia)	Filter Pressure Drop (in H ₂ O)	Motive Flow, M _m (lb/s)	Induced Flow, M _I (lbs/s)	Ratio M _I /M _M	Ratio Predicted* M _I /M _m
17.7	4.4	.060	0.08	1.33	0.95
23.7	8.0	.078	0.135	1.73	0.90
29.7	9.8	.094	0.159	1.69	0.80

* Reference 1

Predicted at PFBC Test Conditions
(with dust cake)

Backflush Motive Pressure (psia)	Pressure (psia)	Filter Pressure Drop (in H ₂ O)	Motive Flow, M _m (lb/s)	Induced Flow, M _I (lb/s)	Ratio M _I /M _m
220	162	55	.62	.55	0.90
220	162	111	.61	.33	0.55
205	162	55	.53	.53	1.0
205	162	83	.52	.27	.52

where ρ_g is gas density, V the superficial gas velocity, g_c the gravitational constant, F the fractional open area of the distributor plate/screen and C_D the characteristic loss coefficient. For the case where there is no bed media (i.e., screen and distributor plate only), Zenz recommends $C_D = 0.60$. Figure 3-8 compares the measured distributor plate and screen pressure drop characteristics with the Zenz correlation for the range of backflush superficial gas velocities of interest. The test data shown represent backflush pressure drop measurements made on each of the four beds in the test GBF element (Element 5). The solid line is drawn through the test data but with the prescribed velocity square - pressure drop dependence and assuming $C_D = 0.6$. For this solid curve, the F parameter is calculated as $F = .058$. This value for the open area term compares very well with that calculated from the known distributor plate and screen geometry of 0.061 (actually, this value may range from 0.047 to 0.061 depending on edge geometry effects). With bed media in the filter unit, Zenz recommends a value of $C_D = 0.80$ be used in the screen-distributor plate pressure-drop correlation.

GBF Fluidization Characteristics - 620 μ Sand Media

Initial tests were done to characterize the fluidization of a 3 inch deep sand bed. The test arrangement is as previously described and illustrated in Figure 3-6. In these tests, the backflush flow was introduced into the GBF element and allowed to flow through each of the four beds and the indicated pressure drop measurements made. Visual observation of behavior of the sand media was also made. The media is regular foundry grade sand that has been sized sieved to 95.5% between 25 and 35 mesh (620 μ m). Subsequent characterization of the sand media showed the minimum fluidization velocity to be 0.59 ft/s.

Figure 3-9 shows a plot of the recorded pressure drop across the GBF element (ΔP_p) as a function of the total mass flow through the GBF element. Also indicated in Figure 3-9 are the visual observations of the behavior of the sand media occurring at the respective operating points. The different test data points correspond to different test

Curve 731419-8

Fig. 3-8—Correlation of distributor plate/screen pressure drop with flow

Fig. 3-9— Fluidization characteristics — 3 inch deep sand bed, 4 operating beds

runs. From 0 to about 0.055 lb/s mass flow the backflush flow passes through the sand media with a resultant increase in pressure drop but no visible indication of bed fluidization. At an indicated total element mass flow corresponding to an average velocity of about 0.38 ft/s bubbling was observed to initiate in Bed B (see Figure 3-6). It was further evident at this point that there existed a non-uniform distribution of flow between the four operating beds. With a slight increase in the backflush flow, a slight increase in the element pressure drop is recorded and the onset of bubbling in the second bed (Bed D) observed. Further increases in backflush flow resulted in little or no increase in the element pressure drop but the observed intensity of the fluidization in Beds B and D increased but with no indication of fluidization in either Bed A or C. At the 0.082 lb/s mass flow point, sand media was observed to be elutriated from Bed B. Further increases in the backflush flow resulted in initiating fluidization in both Beds A and C but also increased elutriation of sand from Bed B.

Results of these tests showed the possibility of a non-uniform backflush flow distribution between the four operating beds in an element of the GBF unit. In these tests this resulted in loss of some bed media prior to the full fluidization of all the operating beds.

Results of testing with a 1-1/2 inch deep sand bed in a second GBF element (Element No. 5) are shown in Figure 3-10. Shown are the element pressure drop ΔP_p , the pressure drop in Bed B measured from the element plenum to just above the distributor plate and screen, ΔP_S , and the difference in the ΔP_S between Beds A and B (ΔP_B). In these tests, the observed behavior of the beds differed somewhat from the 3 inch deep sand bed case in that the top two beds were at least partially fluidized before the onset of sand media elutriation was observed in the bottom beds. In this figure the pressure drop data is plotted against the total backflush mass flow. As in the previous 3 inch deep bed test case, a relatively significant non-uniform distribution in the backflush flow between beds was indicated. In one test the flow distribution between beds was measured using an Alanor-anemometer by measuring the

Fig. 3-10—Pressure drop characteristics of GFB test unit (element 5) with 1.5 in. sand depth (620 μ m diameter) at 1 atm, 70°F

air velocity at the element exit slots that are located at the top side of each filter compartment. These measurements, although considered very qualitative, indicated differences by factors as high as two and three between the flows issuing from the four beds.

Also included in Figure 3-10 for comparison purposes is the calculated distributor plate and screen pressure drop based on the calculated average superficial gas velocity. In this case the screen pressure drop represents only about 5 to 10% of the total element pressure drop over the test range considered. Thus, small differences between beds in their pressure drop characteristics can result in relatively large differences in the distribution of flow between beds at fluidization conditions.

A third series of sand media tests were done where the bed depth was reduced to 1-1/4 inch. These results are indicated in Figure 3-11. In these test runs a somewhat improved flow distribution is observed in that all four beds are fluidized (although to different degrees) before there is evidence of sand media being elutriated from any of the beds. Three fluidization operating zones are distinguished as identified in the figure. The onset of fluidization was observed at a mass flow of 0.052 lb/s which suggest an average backflush fluidizing velocity of about 0.40 ft/s. This compares reasonably well with the 0.59 ft/s value measured in the single, 3 inch diameter bed test and indicates an improved distribution in the backflush flow between beds. This is achieved because the screen pressure drop is now a larger fraction of the total pressure drop (by the difference in bed heights). This difference is apparently sufficient at the lower backflush flows to permit each of the beds to partially fluidize. Then, operating at the higher velocities the flow distribution is stabilized since the distributor plate/screen pressure drop ($\Delta P_S \sim V^2$) becomes a significant fraction of the total system pressure drop.

The sensitivity of the backflush operating characteristics of a GBF element to the distributor plate/screen pressure drop are

Fig. 3-11—Fluidization characteristics of 1.25 inch deep sand bed - 4 operating beds

illustrated in Figure 3-12. Shown are two bed fluidization curves assuming two different pressure drop curves for the distributor plate/screen. The two different distributor plate/screen curves were selected as representative of the uncertainty in the actual measured values. At the set operating point corresponding to the element ΔP , (horizontal line) one bed (solid line) operates at a backflush velocity of 1.75 ft/s while the second (dotted line) shows a backflush velocity of 2.5 ft/s. Between these two values is the observed elutriation velocity of the sand bed media. This example illustrates how relatively small differences in the characteristics of the screen pressure drop could result in satisfactory operation of one bed while the second bed would be elutriating bed media.

At PFBC operating conditions gas density and viscosity differ from the ambient test conditions described above. At the 11 atm pressure and backflush air temperatures of 850°F (mixed mean) gas density will be about 3.5 times the ambient value. The calculated effect on the fluidization characteristics of the shallow sand bed cases described above (Figure 3-12) is shown in Figure 3-13. Because of the high gas density and gas viscosity differences, the operating parameters of the bed (i.e., onset of fluidization and point of bed media elutriation) shift toward lower characteristics velocities. This in part offsets the beneficial effect that the increased gas density has on the distributor plate and screen pressure drop. Assuming the minimum operating pressure drop for the element as indicated by the horizontal line, again the two beds corresponding to different assumed distributor plate and screen pressure drops operate at different characteristic backflush velocities. At the PFBC test conditions, the velocity for the sand bed-media elutriation is expected to shift to a velocity lower than that observed in the ambient tests. Figure 3-14 compares the fluidization characteristics of the GBF element observed and measured at ambient conditions to that projected for PFBC test conditions. One calculation indicates a value for the elutriation velocity of about 1.1 (ft/s). Although there is considerable uncertainty in calculating the

Fig. 3-12—Effect of differences in screen pressure drop characteristics in parallel operating beds at ambient conditions $T = 70^\circ F$, $P = 1 \text{ atm}$, $620 \mu m$ sand media

Fig. 3-13—Expected effect of differences in screen pressure drop characteristics in parallel operating beds at test operating conditions
 $T = 850^\circ F$ (backflush air) $P = 11$ atm

Fig. 3-14—Comparison of the 620 μm , 1.25 inch deep bed fluidization characteristics at ambient and PFBC test conditions

elutriation condition for the sand media at the PFBC test conditions the overall trend toward the lower operating velocity may represent a major disadvantage in utilizing the 620 μm sand media with the present distributor plate and screen design.

Fluidization Characteristics - Tabular Alumina T-61

In view of the above, an alternate bed material, Tabular Alumina T-61 (8 to 14 mesh) was tested at ambient conditions. The advantage for backflush of the Tabular Alumina material is slightly higher material density and larger diameter media. This should provide higher characteristic operating velocities. Sieve analysis indicates that 97% of the media size falls between 841 and 1680 μm . The minimum fluidization velocity was found to be about 2.2 ft/s.

In the tests conducted 2400 grams of material (approximately 1" bed depth) was loaded into each operating compartment of the GBF test element. In these tests each bed was first tested individually by tapping-off the flow exit slot the other three. Thus all the backflush flow introduced passed through one bed. This provides a direct means of calculating the actual superficial gas velocities in the bed. Results of this testing are shown in Figure 3-15. The four curves correspond to the four different filter beds. Two additional curves are included corresponding to the calculated and measured (indirectly) distributor plate and screen pressure drop. With the alumina material the observed onset of fluidization occurred at a measured velocity of 2.1 ft/s. This compares very closely with the minimum fluidization velocity determined in an independent test on a small 3 inch diameter fluidization test rig. Increasing the backflush flow and superficial gas velocity resulted in increasing the intensity of the bed fluidization until a point was reached at about 4.5 ft/s where the alumina media was observed to begin to be expelled from the element housing. Each bed in the element, when tested, showed approximately identical operating characteristics and velocities. At the onset of bed media fluidization, as observed, the distributor plate and screen pressure drop represents

Fig. 3-15 — Fluidization characteristics of Individual GBF beds — element 5
 tabular alumina, 1 inch bed depth

about 23 percent of the total system pressure drop. This is about a factor of 4 higher than measured in the sand media tests. At the point of bed media elutriation, the distributor plate/screen pressure drop fraction is up to about 50 percent of the total element pressure drop. Thus, it can be expected with the alumina material that a more uniform distribution of the backflush flow between parallel operating beds can be achieved. This is shown by the data given in Figure 3-16 where several test runs were made on the element with all four beds operating in parallel. Visual observation confirmed that all beds could be fully fluidized with no visible loss of bed media. In fact, in these tests with all four beds operating, there was insufficient backflush air capacity with the particular experimental set-up used to reach velocities such that the bed media would be elutriated. Measurements of the air velocities at the exit slot above the respective beds with the Alanor anemometer instrument also confirmed that a more uniform flow distribution resulted. Flow difference between beds seemed to range from about 1.25 to 1.5 compared to 2 to 3 values measured with the sand media.

Based on the observations and measurements at ambient conditions projections of the fluidization characteristics of the bed with the alumina media are made for PFBC test conditions. These results are shown in Figure 3-17. Shown for the PFBC test conditions are projections for the minimum fluidization velocity and estimated (lowest) velocity for which bed media would be expected to be elutriated. The characteristic operating velocities range from about 1.3 ft/s to 2.5 ft/s at the PFBC test conditions compared to the 2.2 to 4.5 ft/s values measured under ambient conditions. This result at PFBC test conditions compares to the 0.5 to 1.1 ft/s range predicted for the sand media. At PFBC test conditions the distributor plate/screen pressure drop represents about 38 to 67% of the total element pressure drop over the range of backflush velocities shown. This should provide sufficient pressure drop to adequately distribute the backflush flow between beds and avoid the possibility of elutriating bed media that could result with poor flow distribution.

Fig. 3-16 — Fluidization characteristics of operating element with 1" deep beds of alumina media

Fig. 3-17 — Comparison of fluidization characteristics of alumina media at ambient and PFBC test conditions

3.3.4 GBF Test Unit Installation

Based on the cold flow backflush tests described above a decision to proceed with the hot-gas test program was made based on utilizing (at least initially) the Tabular Alumina T-61 material. Each of the beds in the six elements unit (24-filter beds) have been charged with 2500 gms of the alumina media and the test unit positioned into the facility pressure vessel, Figures 3-18 and 3-19. Final connections on the backflush piping, and the various pressure tap lines are currently being completed. Hot gas testing of the unit is now scheduled to begin January 1982.

Figure 3-18 - View of GBF subpilot test unit positioned over pressure vessel (not shown). View shows location of the bed pressure taps and thermocouples and pressure leads from the eductor sections.

Figure 3-19 - View of GBF subpilot test unit being lowered into pressure vessel.

4. TASK 1.3 - ALKALI METAL REMOVAL STUDIES

Work on this task is aimed at the evaluation and selection of possible alkali sorbents based on bench scale tests conducted to screen candidate materials and tests to determine their chemical kinetic behavior. Material screening tests are being conducted in the batch reactor apparatus sketched in Figure 4-1. This device is simply a pressure vessel in which air or simulated combustion gases are heated along with a bath of alkali metal chloride to give the desired concentration of contaminant. The alkali-doped gas then flows into a reaction zone and through a "plug" of candidate sorbent pellets which can be held at any desired reaction temperature. Provisions have been made to insert and withdraw the test sorbents while the system is at pressure and temperature for quick turn around. Alkali up-take in these tests is determined from chemical analysis of the test pellets.

Tests to determine the chemical kinetics between the alkali and sorbent are carried out using the double balance, high temperature and pressure TG apparatus shown in Figure 4-2. In this system, a bath of molten alkali chloride suspended from a microbalance is maintained at a desired temperature in the lower region of a pressurized furnace. The indicated weight loss from the bath gives a continuous measurement of the gas phase alkali concentration. Simultaneously, a pellet of sorbent material is suspended from a second balance that is also contained in the bell-housing. The read-out from this second balance continuously records the weight change of the sorbent. This apparatus has been designed and built under DOE sponsorship for a program directed toward gasification processes. This gasification program, however, had been interrupted prior to the shakedown and actual operation of the TG apparatus. Although the DOE alkali-gasification program has been resumed, the initial operation of the TGA has been carried through as

Fig. 4-1—Bench scale reactor for preliminary high pressure-temperature studies of alkali sorption

Figure 4-2 — Double-balance high temperature pressure TGA system.

part of the Westinghouse Category I and Category II hot gas clean-up programs. At present, work on this task is being carried out in coordination with the DOE contract - "High Temperature Removal of Alkali and Particulates in Pressurized Gasification Systems" - Contract DE-AC21-80MC16372.

During this report period, candidate getter qualification and screening tests have continued utilizing both the batch reactor and TGA apparatus. Results from these studies are summarized below. In addition, based on previous TGA test runs conducted as part of Task 3, analysis has been initiated of possible reaction stoichiometry. This analysis is currently being carried through as part of the DOE contract DE-AC21-80MC16372, and results have been reported therein.⁽⁵⁾

4.1 Sorbent Material Screening Tests (Batch Reactor Tests)

Work on the batch reactor unit has focussed on screening tests involving 20 candidate getter materials along with water vapor tests on activated bauxite. In the screening tests, the candidate getters (list in Table 4-1) were exposed at 850°C to approximately 80 ppmV NaCl in argon for 6 hours at 11 atmospheres pressure. In each test run, an activated bauxite pellet was incorporated for comparative purposes. All test runs have now been completed and the exposed getter pellets have been submitted for atomic absorption analysis to determine their water soluble and acid-insoluble content of sodium and potassium. Preliminary results back from analysis has indicated that the Vansil, Ebony Novacite, Emathalite and Diatomaceous Earth all show total Na capacities greater than the activated bauxite. Total sodium uptake ranged from about two to five times the bauxite values for the different materials. These preliminary results await further analysis to define soluble and insoluble fractions which are yet to be completed.

Test conditions for the activated bauxite tests involving water vapor were identical to those for the screening tests described above except for the presence of 4.6 mol % water vapor in the argon carrier

Table 4-1

CANDIDATE GETTER MATERIALS FOR SCREENING TESTS

Alternate Getter Material	General Composition
Activated Bauxite	SiO_2 , Al_2O_3 , 1.2 % Fe_2O_3
Activated Bauxite	SiO_2 , Al_2O_3 , 1.38 % Fe_2O_3
Activated Bauxite	SiO_2 , Al_2O_3 , 2.66 % Fe_2O_3
Activated Bauxite	SiO_2 , Al_2O_3 , 7.28 % Fe_2O_3
PFBC Ash	SiO_2 , Al_2O_3
Olivine	$(\text{Mg}, \text{Fe})_2(\text{SiO}_4)$
Vansil [®]	CaSiO_3
Novacite [®]	SiO_2
Ebony Novacite [®]	SiO_2
Emtal Platy Talc	MgO , SiO_2
New York State Talc	MgO , SiO_2
Iron Ore	Fe_3O_4
Activated Alumina (spheres)	Al_2O_3
Activated Alumina (pellets)	Al_2O_3
Emathalite	Ca , Mg , Al , SiO_2
Zeolite (70 μm)	$(\text{Na}_2, \text{K}_2, \text{Ca}, \text{B}) \cdot (\text{Al}, \text{SiO}_2)_n \cdot x\text{H}_2\text{O}$
Zeolite (spheres)	$(\text{Na}_2, \text{K}_2, \text{Ca}, \text{B}) \cdot (\text{Al}, \text{SiO}_2)_n \cdot x\text{H}_2\text{O}$
Zeolite (60-80 mesh)	$(\text{Na}_2, \text{K}_2, \text{Ca}, \text{B}) \cdot (\text{Al}, \text{SiO}_2)_n \cdot x\text{H}_2\text{O}$
Diatomaceous Earth-Calcined Diatomite	SiO_2 , Al_2O_3
Diatomaceous Earth-Cellite Diatomite	SiO_2 , Al_2O_3
Diatomaceous Earth-Synthetic Calcium Silicate	SiO_2 , Al_2O_3

Conditions: Temperature: 1123K
 Pressure: 11 atm
 Duration: 6 hrs

gas. In these tests, the gettering performance of the activated bauxite exposed to the water vapor enriched gas appeared to be about twice that for the bauxite without water vapor. These results fall in line with similar tests reported by ANL.⁽⁶⁾

The series of batch reactor tests reported above complete the planned test program for screening and qualifying getter materials. Future testing will be limited to the double-balance TGA system.

4.2 TGA Testing

At the end of the last quarter it was suspected that salt contamination of the packed bed preheat zone between TGA test runs, with vaporization of this salt early in the following run, might lead to invalid kinetic data. Capacities should be valid however, since they depend on the salt concentration late in a run. Therefore, at the beginning of this quarter, tests were made in which it was found that this source of error could be eliminated by washing the packed bed with deionized water prior to a test. This step has now been added to the operating procedure.

Also in this quarter an analytical investigation was completed for the TGA unit to predict the interaction between the salt-bearing nitrogen reactant gas (5 lpm) and the helium purge gas used to protect the balance and hangdown assembly from salt contamination. This study has confirmed previous experimental findings in that operating the TGA with a 1.25 lpm He purge gas does not result in the purge gas interfering with the pellet reaction. At 5.0 lpm He flow, the purge gas does sweep down over the reacting pellet possibly resulting in the dilution of the local salt concentration or preventing salt to enter the pellet reaction zone. (See Appendix 1). All subsequent runs are to be made with 1.25 l/min. He purge gas and 5 lpm nitrogen carrier gas.

Eight successful runs were subsequently made in this quarter testing bauxite as a getter material. These are summarized in Table 4-2. All were made at 1 atm total pressure. The first six runs were

Table 4-2

SUMMARY OF ACTIVATED BAUXITE TGA TESTS

Test No.	Pellet Initial Wt, mg	Pressure, atm	Getter Temperature, K	Run Duration, hr	NaCl Concentration, ppmv	Getter Weight Gain, mg
12	208	1	1123	19	11.5	4.0
13	203	1	1174	22	9.4	2.7
14	201.7	1	1123	44	3.3	3.2
15	212.4	1	1173	46	3.4	2.2
16	205.8	1	1123	96	0.79	1.8
17	217.8	1	1173	92	0.78	1.4
18	201.7	1	1123	43 ^A	12	4.4
				30 ^D	0	-1.1
19	197	1	1173	48 ^A	11	3.4
				70 ^D	0	-0.3

A = Adsorption test

D = Desorption test with same pellet.

adsorption tests, and both weight gain and gas-phase salt concentration are given as functions of time in Figures 4-3 to 4-5. Notice that (1) capacity is higher at 850°C than at 900°C at each gas-phase concentration level, and (2) capacity is greater at higher gas-phase salt concentration at each temperature. These results are consistent with the reversible adsorption mechanism which was discussed in the previous quarterly report.⁽⁴⁾

For a first-order reaction, weight gain should depend on the duration of exposure to salt at a given concentration. The product of partial pressure of NaCl and time is, therefore, a useful index of the accumulated exposure of a pellet. Figure 4-6 compares the weight gain of pellets at different conditions from Table 4-2 in response to accumulated exposure.

Two tests were made to see if the adsorption is completely or only partially reversible, and to quantify the rate at which desorption occurs. These tests are also shown in Table 4-2. Each test consisted of an adsorption portion in which a pellet was exposed to relatively high salt concentration until virtually saturated, and a desorption portion in which the temperature of the salt vaporization section of the TGA reactor was suddenly lowered to such a point that virtually no salt was transferred to the gas stream passing the pellet. The weight gain and salt concentration histories of these two runs are shown in Figures 4-7 and 4-8.

The adsorption results agree well with the previous bauxite runs if allowance is made for the greater time of exposure. This is shown in Figure 4-6 where tests 12 and 18 are compared. They also confirm the hypothesized effects of temperature and gas-phase salt concentration on capacity in a reversible mechanism.

The desorption results mean that the adsorption is not completely reversible. Transfer of salt from the gaseous state to the adsorbed reversible and irreversible states may be by competing parallel or consecutive mechanisms. If we assign a lesser value to reversibly

Fig. 4-3 — Results of TGA bauxite tests 12 and 13

Curve 730939-A

Fig. 4-4 Results of TGA bauxite tests 14 and 15

Curve 730938-A

Fig. 4-5 — Results of TGA bauxite tests 16 and 17

Fig. 4-6 - Comparison of getter activity at various conditions

Fig. 4-7 — Results of TGA bauxite run 18

Curve 730940-A

Fig. 4-8— Results of TGA bauxite Run 19

adsorbing salt relative to irreversibly adsorbing it, the capacity advantages of operating at lower temperatures are apparently reduced.

5. TASK 1.4 - PFBC SYSTEM PERFORMANCE

A preliminary set of parametric studies have been conducted to evaluate specific GBF design and operating parameters with respect to their impact on the three PFBC reference plant concepts.⁽³⁾ The impact of the different GBF design and operating parameters is measured by the calculated reduction (or increase) in the net plant output relative to the same plant operating without the GBF. For a given PFBC concept, this basis provides a simple normalized comparison to evaluate the relative effects of selected parameters, and the relative sensitivity of varying one parameter over a range of conditions. Also, between concepts, the analysis has provided a relative basis for determining if a different set of GBF operating and design parameters are more appropriate for one PFBC concept over the next.

During this report period, a preliminary set of economic trade-offs have been completed that compares the all-cyclone PFBC steam cooled and air cooled plants to the reference PFBC plants developed in Task 1. The all-cyclone plant description was developed under an independent Westinghouse study funded by ANL/DOE, Contract ANL 31-109-38-6308. For the purpose of the present study, an economic comparison is made between the costs for particulate gas clean-up between the two plants as a function of the operating characteristics of the GBF unit (face velocity).

Reference PFBC plant designs have been described that assumed two stage cyclone with a GBF gas clean-up train⁽⁴⁾ (work conducted in Task 1). Tables 5-1 and 5-2 provides a summary of the design basis and cost summary for both the reference steam cooled and aircooled plant. For comparison, a design and cost for a conventional coal-fired steam boiler power plant with flue-gas desulfurization (PCF/FGD) is also

Table 5-1 - PFBC-Plant Conceptual Designs

Summary Of Plant Concepts - Design Basis

Design Parameter	Steam Cooled PFBC	Aircooled PFBC	PCF/FGD
Standards	NSPS	NSPS	NSPS
Coal	Eastern	Eastern	Eastern
Net Power (MW)	648	462	509
Comb. Temp (°F)	1750	1650	--
Comp. PR	11.5	7	--
TIT (°F)	1664	1600	
Gas Clean-Up			
• Particulate	2-Stages Cyc. + GBF	2-Stages Cyc. + GBF	Stack Gas (FGD/ESP)
• Alkali	Fixed Bed Absorption	Fixed Bed Absorption	--
Steam Conditions (psig/°F/°F)	2400/1000/1000	835/800/700	2400/1000/1000

Table 5-2 - PFBC-Plant Conceptual Designs

Summary Of Costs/Performance

Costs (Mills/kWh)	Steam Cooled PFBC (648 MW)	Air Cooled PFBC (462 MW)	PCF/FGD (509 MW)
Fixed Charges			
• Solids Feeding & Combustors	3.90	6.62	--
• Hot Gas Clean-Up & Gas Piping	6.00	12.57	4.40
• TG/Comp.	6.57	5.09	5.25
• Heat Recovery/BOP	11.26	14.82	14.98
Subtotal	27.73	39.10	24.63
O&M (Levelized)			
Solid/Waste Disposal	5.40	5.72	5.29
Fixed O&M	6.11	8.81	5.90
Subtotal	11.51	14.53	11.19
Fuel	27.85	28.88	31.33
	(8598 Btu/kWh)	(9005 Btu/kWh)	(9671 Btu/kWh)
Total COE	67.09 Mills/kWh	82.51 Mills/kWh	67.14 Mills/kWh

included. As indicated in Table 5-1, there exists different assumptions between the PFBC concepts concerning combustor and turbine inlet design parameters. Thus comparison between PFBC concepts is not emphasized. Table 5-2 shows a summary of the costs and performance for each plant design. The capital costs have been grouped into four overall accounts that distinguish the hot gas cleanup capital costs from the other plant equipment. The gas cleanup capital costs include costs for the primary and secondary cyclone stages, the GBF units, all interconnecting hot gas piping and the capital cost for the alkali removal beds. The levelized operating and maintenance costs and fuel costs are also provided.

For either PFBC concept, the gas cleanup costs represent about 20 to 30 percent of the total capital equipment costs and 9 to 16 percent of the total plant cost of electricity (COE). With either PFBC concept, the actual GBF capital costs represent about 45 to 50 percent of the total gas cleaning costs. The higher gas cleaning costs associated with the indirect aircooled plant are the result of different assumed combustor operating pressure and temperature and higher plant heat rate. These factors result with the aircooled plant in about twice the volumetric gas flow per kilowatt passing through the gas cleanup system. Both PFBC concepts show higher capital costs than the conventional PCF/FGD but this difference is somewhat offset by the lower fuel cost associated with PFBC plants. The PFBC boiler plant shows a total cost of electricity (COE) equivalent to the conventional coal plant. Costs for the aircooled plant are higher.

In either PFBC concept the capital and operating costs for the actual GBF units and their pressure containment vessels represent a relatively small fraction (4 to 8 percent) of total plant costs if the assumed GBF design and operating conditions are proven technically feasible. The economics of the GBF are determined by the assumed design (materials of construction, geometry, etc) and by its operating filter face velocity. The effect on the gas cleanup COE of the latter parameter (face velocity) is shown in Figure 5.1 for both PFBC concepts.

- Basis:
- Gas Clean-up Train Includes Primary Cyclone Thru to Expander Inlet
 - Costs for Stack Gas Particulate Removal Included if Required to Meet NSPS
 - Costs for Turbine Stator/Rotor Reblading

Fig. 5-1 — Comparison of gas clean-up costs for PFBC plants employing GBF or all-cyclones

Also included in this figure are the gas cleanup costs (excluding alkali removal) for PFBC plants that would use only all-cyclone cleanup.^(7,8) In the all-cyclone cases the gas cleanup costs include stack gas particulate removal to meet NSPS and costs for turbine stator/rotor blade replacement. The all-cyclone case for each PFBC concept is shown for two different turbine-life assumptions. A six month and one year blade life is taken for the steam cooled plant and a one and two year assumption for the indirect aircooled plant. These differences in turbine life assumptions reflect differences in the expected dust loading at the turbine inlet of the respective plants.

For the GBF cases, the assumed design point for filter face velocity is taken as 40 ft/min. As seen from the figure, operating at lower filter velocities significantly increases the gas cleaning costs. The increased number of filter units at the lower velocity along with the added cost and complexity associated with the gas piping act to nearly double the COE at 20 ft/min over the 40 ft/min case. Moving toward higher face velocities above 40 ft/min reduce costs but the effects are less dramatic. Thus, the technical risks of pursuing high filter face velocity design may not be justified in view of the relatively small associated incremental improvement in COE.

At the 40 ft/min nominal design point, the GBF cases are shown to be somewhat lower in costs than the most optimistic all-cyclone case. Lower turbine blade life and/or increased filter face velocities can significantly enhance the incentive for the GBF case over the all-cyclone designs. Real cost differences of 3 or 4 mills/kWhr represent capital expenditures on the order of 35 to 50 million dollars.

6. TASK 1.5 - COMMERCIAL SCALE GBF CONCEPTUAL DESIGN

Initial efforts scheduled in this task have been reported.⁽³⁾
No additional effort had been scheduled during this report period.
Effort on this task will continue when test data on the GBF pilot unit has been obtained.

7. REFERENCES

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APPENDIX 1

TGA SYSTEM-COUNTERFLOW JET PENETRATION INTO A CONFINED STREAM

This appendix describes a brief study made of the flow conditions in a portion of a thermal gravimetric analysis test set-up. The region of interest is shown in outline form in Figure 1. A test sample pellet is suspended by a thin wire which transverses the small tube. A protective blanket of helium flows downward through the small tube.

Nitrogen gas passes upward through the large tube to contact the test sample. The purpose of this study is to determine the conditions under which it can be assumed that the sample is in fact in contact with N_2 rather than He.

Two separate flow conditions were examined as indicated by Table 1. As may be seen from the Reynolds number estimates, the flow should be laminar with no turbulent mixing. Thus the boundary between N_2 and He should be fairly sharp.

It should be possible to describe the flow patterns fairly precisely by numerical integration of the Navier-Stokes equation. However, the magnitude of that task is such that it is probably not justified. Instead conservative approximations have been made using simplifying assumptions.

The helium jet penetrates into the nitrogen tube by virtue of its kinetic energy or velocity head expressed by $\rho V^2/2g$. The penetration is countered 1) by the bouyancy of helium in nitrogen, 2) by the pressure gradient resulting from the viscous wall friction in the nitrogen stream, and 3) by the viscous friction within the helium jet and at its boundaries. The more of these retarding effects included, the more accurate and the less conservative is the result.

The first try included only the buoyancy effect.

At point a, $V_a = 2.0 \bar{V}$ for fully developed laminar flow at the helium tube exit. At point b, $V_b = 0.0$. Letting P represent static pressure

$$P_b - P_a = \frac{\rho_{\text{He}} V_a^2}{2g} - \frac{\rho_{\text{He}} V_b^2}{2g} + \rho_{\text{He}} \Delta Z,$$

where ΔZ is the jet penetration. Also,

$$P_b - P_a = \rho_{N_2} \Delta Z.$$

Therefore

$$\Delta Z = \frac{\rho_{\text{He}} (V_a^2 - V_b^2)}{2g (\rho_{N_2} - \rho_{\text{He}})}.$$

For Case I

$$\Delta Z = 0.0152 \text{ ft} = 0.183 \text{ in.}$$

For Case II

$$\Delta Z = 0.266 \text{ ft} = 2.922 \text{ in.}$$

We note that these results are independent of temperature and pressure to the extent that these two gases can be considered perfect.

A lower limit on ΔZ is probably the value that results in about the same lateral area for helium flow outward as the tube cross section for downward flow at the tube mouth. That is

$$2\pi R_1 \Delta Z_m \approx \pi R_1^2$$

$$\Delta Z_m \approx \frac{R_1}{2} = 0.166 \text{ inch.}$$

Thus any refinement that would substantially reduce the ΔZ estimate for Case I is probably not realistic. For Case II a better estimate was made including the viscous friction in the nitrogen stream.

If we assume fully developed flow in the nitrogen tube in the region below (upstream) of the helium introduction point, then the component of pressure gradient resulting from friction is

$$\frac{dP_{\mu}}{dz} = \frac{8\mu_{N_2} V_{N_2}}{R_3^2 g}$$

$$\text{Then } P_b - P_a = \rho_{N_2} \Delta Z + \frac{dP_{\mu}}{dz} \Delta Z$$

and again

$$P_b - P_a = \frac{\rho_{He} V_a^2}{2g} - \frac{\rho_{He} V_b^2}{2g} + \rho_{He} \Delta Z.$$

Solving for ΔZ

$$\Delta Z = \frac{\frac{\rho_{He}}{2g} (V_a^2 - V_b^2)}{\rho_{N_2} - \rho_{He} + \frac{8\mu_{He} V_{N_2}}{R_3^2 g}}$$

For Case II

$$\Delta Z = 0.184 \text{ ft} = 2.203 \text{ in.}$$

It is doubtful if a solution for Case II that is both credible and less conservative without a major effort in attaining a numerical solution to the Navier-Stokes equation.

Table I
Cases Examined

	Case I	Case II
N ₂ flow reduced to Standard conditions (25°C, 1.0 Atm), l/min.)	5.0	3.0
He flow reduced to Standard conditions, l/min.	1.25	5.0
Pressure, Atm	1.0	1.0
Temperature, °C	900°C	900°C
N ₂ density lb/ft ³	.0181765	.0181765
He density lb/ft ³	.0025963	.0025963
N ₂ viscosity lb/ft/sec	.00002928	.00002928
He viscosity lb/ft/sec	.00003268	.00003268
N ₂ Volumetric flow at actual conditions, ft ³ /sec	.011580	.006948
He Volumetric flow at actual conditions, ft ³ /sec	.002895	.01158
N ₂ Average Velocity upstream, ft/sec	2.1242	1.2745
He Average Velocity upstream, ft/sec	1.21185	4.8474
Reynolds no. in N ₂ passage	109.4	65.6
Reynolds no. in He passage	5.31	21.25

Fig. 1 - Schematic of region of interest