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OAK RIDGE Y-12 PLANT

IMPROVING THE PERFORMANCE OF THE Y-12 FLUIDIZED BED CONTACTORS: FINAL REPORT ON INVESTIGATIONS AT THE UNIVERSITY OF TENNESSEE

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EXECUTIVE SUMMARY

Recent tests at The University of Tennessee in Knoxville (UT) have demonstrated improved fluidization performance in a mockup of the Oak Ridge Y-12 Plant* fluidized bed with a modified gas distributor. Combining the modified distributor with the recently developed fluidization intensity module is expected to result in substantial operational improvements for the fluidized beds in Building 9212 at Y-12. Important additional benefits coming from the improved operation of the Y-12 fluidized beds will be a reduction in the consumption of reactant gases, a reduction in scrubber waste production, and a reduced need for manual cleaning of the particulate removal system. UT tests have also demonstrated that the tapered wall design traditionally used for the Y-12 fluidized beds does not provide a sufficient improvement in fluidization quality to justify the continued use of this design in place of a more conventional straight-wall configuration. The straight-wall design for replacement beds is expected to result in substantial equipment cost savings. Other operating scenarios that could potentially be used to reduce reactant gas consumption and scrubber waste production have also been considered. These scenarios are documented here for future reference.

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1.0 INTRODUCTION

Transformation of uranium tetrafluoride from uranium trioxide is a key operation at Y-12. The transformation is carried out batchwise in two stages of nominal 5.5-in.-diam fluidized beds located in Buildings 9206 and 9212. Reactant gas containing hydrogen (stage 1) or hydrogen fluoride (stage 2) is passed upwards through beds of particulate uranium trioxide or uranium dioxide, respectively. Fluidization quality (that is, the efficiency of contacting between the gas and solids) is of concern because poor quality has detrimental environmental, safety, and health consequences.

Poor fluidization quality can be caused by either excessive or insufficient fluidization gas flow. In the former case, excess gas forms large bubbles or slugs that move rapidly up through the bed, effectively bypassing most of the solids and producing violent surges. This condition results in inefficient use of the reactant gas, excessive scrubber waste production, and fouling of the exit gas filters. Filter fouling is of concern because heavy fouling can require that the filters be cleaned manually, exposing personnel to potential radiation and chemical hazards. Insufficient gas flow, on the other hand, leads to losses in particle agitation and reduces heat transfer such that the bed can agglomerate. Again, the result is excessive reactant gas consumption, an increase in scrubber waste production, and the need for manual cleaning.

In fluidized beds, such as those at Y-12, optimum gas flow is difficult to maintain because normal variations in feed particle size, gas composition, and temperature cause fluidization quality to vary unpredictably. Furthermore, there is no direct way to observe fluidization. Two previous letter reports^{1,2} describe earlier test programs at The University of Tennessee in Knoxville (UT). These earlier test programs resulted in the development of a fluidization intensity module (FIM) for monitoring fluidization intensity by means of fluctuations in bed pressure drop. Development of the FIM has made it possible to monitor fluidization quality continuously and adjust the gas flow to its optimum level either manually or by means of automatic control.

During the second series of the FIM development tests, serious deficiencies were recognized in the current Y-12 fluidized bed design that prevented optimum operation. This report summarizes the final tests at UT, which focused on investigating and correcting these design deficiencies. In addition to reporting on the corrections, we describe other operating scenarios that could potentially be used to reduce reactant gas consumption and scrubber waste production even further.

2.0 EXPERIMENTAL APPARATUS AND PROCEDURES

2.1 GENERAL FACILITIES AND CONDITIONS

Consistent with the previous testing, all of the fluidization tests reported here were conducted at room temperature with compressed air as the fluidizing gas. The tests were conducted in the fluidization and chaotic time series analysis laboratory of the Mechanical and Aerospace Engineering Department at UT. This laboratory is located on the mezzanine level of the Dougherty Engineering Building.

2.2 FLUIDIZATION VESSEL WALLS

The 1:1 scale mockup of the Y-12 tapered bed tested previously is shown in Fig. 1. Two new wall configurations were evaluated in the most recent tests, a modified version of the existing "tapered" Y-12 design and a 5.5-in. straight-wall (untapered) design. For the modified tapered design, the steeply tapered bottom cone (marked in Fig. 1) was removed, and the upper wall taper was extended down to the level of the distributor (Fig. 2). This design is also referred to as the "constant taper design." The straight bed (Fig. 3) was constructed to be the same length as the tapered bed with a constant 5.5-in. inside diameter.

The modified tapered bed was constructed because it had been noted previously that the small diameter of the bottom cone enhances spouting (i.e., channeling of the gas toward the center of the bed as the gas rises from the distributor). While sometimes beneficial for preventing solids agglomeration, spouting is known to cause poor gas-solids contacting. We conjectured that removing the more sharply tapered conical section from the tapered bed and replacing it with an equivalent length tapered like the upper section would reduce the spouting tendency.

The straight bed was constructed to allow a careful comparison of its behavior with that of the tapered wall design. The availability of high-speed data acquisition and more quantitative performance measures allowed reassessment of the benefits of the tapered design.

As in the previous tests, the walls of both beds were constructed from plexiglass to allow unobstructed viewing and videotaping. Two 1/2-in.-wide copper strips were attached to the inside walls with epoxy cement and grounded to remove static electricity.

2.3 FLUIDIZING GAS DISTRIBUTOR

As part of the tapered bed revision, the gas distributor diameter was enlarged to provide a more evenly distributed flow across the bottom of the bed. The original four air injection nozzles (each with 12 1/8-in. dia. holes) were replaced with nine nozzles, each having four 1/16-in. diameter holes. Each nozzle was located on the distributor plate so that it

Fig. 1. Scale mockup (1:1) of the Y-12 tapered fluidized bed.

Fig. 2. Modified tapered bed design in which the bottom cone has been removed to increase the distributor cross-sectional area.

Fig. 3. Straight-wall bed constructed for comparison with the tapered bed.

supplied air to 1/9th of the total cross section. The smaller injection holes resulted in a reduction in the total injection nozzle flow area and an increase in the distributor pressure drop. The increased pressure drop was also expected to reduce the possibility of uneven air distribution through the nozzles and to minimize the dynamic coupling between the bed and fluidizing gas supply. Fig. 4 illustrates this distributor, which is referred to as the "flat" design. This was the only distributor design tested with the modified tapered bed.

The "flat" distributor described above was also tested with the straight-wall bed. As discussed later, tests with both the constant taper bed and the straight bed revealed that the increased distributor area makes it more difficult to discharge the solids from the bed pneumatically. This observation prompted the development of four other distributor designs [Figs. 5, 6(a) and (b), and 7(a) and (b)]: a "top hat" design that extended the flat distributor surface up into the fluidization chamber to the level of a solids drain port on the side wall; an "angle" design in which the distributor surface was sloped toward the drain port to enhance solids drainage; a "valley" design in which the surface was sloped in two directions to channel the solids more narrowly toward the side outlet; and a "conical" design in which the distributor surface was sloped to channel the solids toward the center.

Two implementations of the conical design were evaluated. In the first implementation [Fig. 7(a)], approximately one ninth of the fluidizing air was injected into the center of the cone through a vertical standpipe that penetrated the plenum. Air flowed into the standpipe from the plenum chamber through holes in the standpipe wall. This particular design was constructed to provide an automatically self-regulating source of fluidizing gas to the center of the distributor. When discharge was required, the solids were removed through the standpipe.

A second implementation of the conical distributor [Fig. 7(b)] was considered in which the standpipe was closed off, and a 0.5-in. dia. "snorkel" was inserted through the side wall with the exit positioned 1/4 in. above the bottom of the cone. A portion of the fluidizing air could be injected through the snorkel during normal operation, and the solids could be removed pneumatically via the same route after fluidization.

2.4 BED MATERIAL

As mentioned previously, the UT tests were conducted with surrogate solids in place of uranium dioxide, uranium trioxide, and uranium tetrafluoride. Two size ranges of 316 stainless steel from Nuclear Metals, Inc. (density 8.0 g/cm^3), were used, 35×100 mesh and 18×50 mesh. Sieve analyses for each of the sizes are given in the Appendix. The equivalent Sauter mean diameters were 183 and 565 microns, respectively. The corresponding experimentally measured minimum fluidizing velocities were 0.32 and 2.20 ft/s.

Based on their size and density, the small and large sizes of stainless steel would be expected to fall into two characteristically different fluidization groups (Group B and Group D, respectively) as defined by Geldart.³ Group B particles tend to bubble more smoothly, and solids mixing occurs primarily as the result of convection associated with bubbles. Group D particles tend to slug, and solids move in more piston-like arrangements between slugs. Our experimental observations confirm these expected differences.

Fig. 4. "Flat" gas distributor used with the modified tapered bed and for some of the testing with the straight bed.

Fig. 5. "Top hat" gas distributor used with the straight-wall bed. The central region was elevated to the level of the sidewall discharge to reduce the residual solids left following pneumatic discharge.

Fig. 6(a). "Angle" gas distributor designed to enhance solids flow toward one wall during solids discharge.

Fig. 6(b). "Valley" gas distributor used with the straight-wall bed. The valley was designed to channel solids toward the sidewall outlet.

Fig. 7(a). "Conical" gas distributor used with the straight-wall bed. The conical shape causes the solids to flow toward the apex where they can be removed through the vertical standpipe.

Fig. 7(b). "Conical" gas distributor with the "snorkel" solids discharge line.

2.5 PRESSURE TAPS AND PRESSURE-DROP DATA ACQUISITION SYSTEM

Three 1/8-in. NPT pressure taps were installed in the walls of both the tapered and straight-wall beds at the locations noted in Figs. 2 and 3. The lowest tap was connected to a Schaevitz model P1041-0005 pressure transducer (0-1000 Hz frequency response) to provide a dynamic measurement of the overall pressure drop from this tap to atmosphere. When the FIM was tested, a Kulite sealed gauge model KF 1080 (0-10 psi) transducer was connected to the bottom tap to provide the overall pressure drop signal used by the FIM. The upper two taps were connected in all tests to a MKS (Baratron) Model 223BD transducer (4.5 ms time constant, 35 Hz response frequency) to provide a dynamic differential pressure-drop signal for this region.

The Baratron and Schaevitz transducer signals were passed through Wavetek high- and low-pass filters to remove dc bias and prevent aliasing before recording with a Nicolet 400 digital oscilloscope. The high-pass filter was set for 0.1 Hz, and the low-pass filter was set for 30 Hz. Digital sampling rates were held fixed at 200 samples/s. Typical pressure-drop records from both transducers contained 20,000 contiguous data points.

2.6 HEAT-TRANSFER PROBE

A heat-transfer probe was fabricated to make quantitative evaluations of solids mixing. The principle is that the heat-transfer coefficient is sensitive to solids mixing. As the solids cease to move in the vicinity of probe, a sharp drop in the heat-transfer coefficient occurs, causing the temperature of the probe to rise.

The heat-transfer probe (Fig. 8) was constructed along the lines of one reported in the fluidized bed literature.⁴ The probe is inserted through openings in the wall across the bed from one side to the other, passing through the centerline. A resistance heater along the probe axis produces heat, which is conducted through the copper wall to the solids and gas flowing outside. A thermocouple embedded in the wall measures the wall temperature. Based on the known power supplied to the resistance element, the wall temperature and area, and the bulk temperature of the bed solids and gas, it is possible to determine the overall outside heat-transfer coefficient. The probe accuracy was verified by comparing the observed heat-transfer coefficients obtained for air-only tests (i.e., tests in which no solids were present and various air flows were established) with correlation predictions. The results of this comparison are given in the Appendix.

Fig. 8. Heat-transfer probe used to evaluate solids mixing.

2.7 FIM AND FLOW CONTROL SYSTEM

In its present configuration, the FIM is an analog electronic circuit that processes the output of a pressure transducer to obtain a signal proportional to the degree of fluidization (see Fig. A.1 in the Appendix). The pressure transducer should have a frequency response of at least 40 Hz. The FIM has been designed to accept a 4-20 mA signal, which is converted to 2-10 V in the signal processor. The signal goes through a difference amplifier, which subtracts the 4 mA bias on the original signal and adds a gain of two.

The signal is then processed by a differentiator stage, which increases sensitivity to higher frequencies up to 10 Hz. Frequencies above 10 Hz are attenuated with a two-pole low-pass filter incorporated into the differentiator stage. The overall response of the differentiator stage is such that it has a gain of 1 at 0.23 Hz, 40 at 10 Hz, and 1 again at 420 Hz. An additional two-pole filter further reduces higher frequency response. The overall effect of the two stages is a response of -18 dB at 60 Hz.

After being rectified by an absolute value circuit, the signal is filtered to remove ripple above 0.16 Hz. The signal is next passed through a summing amplifier to add a 1 V bias to make it conform to standard control signals. This signal is sent to a moving segment bar graph array to serve as an indicator of fluidization. The summing amplifier also has an adjustment to allow setting the output gain between 0.667 and 1.333.

For closed-loop operation the control, the signal from the FIM serves as the process signal to the process controller. The process controller (for these tests a Foxboro model 760) is an ordinary PI (proportional, integral) controller set with low gain and considerable integration to minimize response to signal fluctuations that are still present in the signal. The proportional factor was 3500%, and the integral factor was 0.027 min. The output of the controller, which is 4-20 mA in this case, goes to an I to P (current to pressure) converter to interface with the air-operated diaphragm-type control valve. If the fluidization signal is below the setpoint of the controller, the controller will increase its output, thus opening the control valve to allow more fluidizing gas into the bed. Conversely, if the fluidization signal is above the setpoint, the controller will reduce its output to reduce the fluidizing gas flow. The control valve used to regulate fluidizing air flow was a 1/2 in., type OSB75 with "E" trim ($C_v=0.5$) manufactured by Research Controls.

2.8 TEST PROCEDURES AND CONDITIONS

The bulk of this phase of the project consisted of three principal types of tests: fluidization quality tests, solids mixing (heat transfer) tests, and solids discharge tests. Closed-loop testing of the FIM was also conducted after completing the other tests.

In the fluidization quality and mixing tests, the objectives were to determine how fluidization quality and solids mixing, respectively, were affected by changes in the wall or distributor designs. The objective of the solids discharge tests was to evaluate how well the solids could be emptied pneumatically from the bed. The latter evaluation is important to ensure that the minimal residual solids are left behind in the bed after each batch.

Fluidization quality tests were conducted consistent with past tests by measuring the pressure-drop fluctuations at the three wall pressure tap locations described earlier. These measurements were recorded digitally at 200 samples/s for several gas velocities and one or more bed heights with both sizes of stainless steel particles. Video recordings were also made of the bed behavior at each flow condition. A similar procedure was followed for the solids mixing tests, except that heat-transfer measurements were made at the two locations noted in Figs. 1 and 2.

Solids discharge tests were conducted by filling the bed with either size of stainless steel particles and then discharging the solids through the selected port into a pneumatic eductor (Fox Valve Development Corporation, 3/4-in., Series 300-SCE). From the eductor, the solids were conveyed to a cyclone (see Fig. 9). Fluidizing air addition and throttling of the bed exhaust valve were sometimes included to determine their impact on the discharge rate. Video recordings were made for most of these tests.

An eductor from Fox Valve Development Corporation was used because that corporation has done considerable development work to characterize the performance of its eductors for solids conveying. After consulting with Fox Development specialists, we selected a 3/4-in. Series 300 SCE eductor, which is rated at 200 lb/h with 20 acfm of air at 50 psig. This rate seemed to offer a reasonable compromise between particle velocity and total transfer time. To prevent choking of the 18 × 50 mesh stainless steel in vertical flow and saltation in horizontal flow, the minimum air flows in 3/4 in. schedule 40 pipe are estimated to be 6 and 5 CFM, respectively.

The final testing stage involved an evaluation of the FIM performance in open and closed-loop configuration with the straight-wall bed and conical gas distributor. The objective here was to confirm that the FIM still performed satisfactorily when used with improved bed and distributor designs. Open-loop tests were conducted by observing the FIM output as the fluidizing gas flow was increased and decreased manually. During closed-loop tests, the controller setpoint was adjusted to incremental values between relatively severe slugging and minimum fluidization conditions, and the operational stability observed. The quality of the control was evaluated by observing fluctuations in the FIM output signal relative to the setpoint and in the motion of the fluidized solids. Videotapes were also made.

Table 1 summarizes all of the design configurations and major test conditions evaluated. Note that while the general types of testing (i.e., fluidization, solids mixing, discharge, and FIM) had been selected from the beginning, the specific design changes and operating conditions evaluated at each stage were determined at the time based on the most recent information available. We think that this procedure was the most effective approach for reaching the project objectives within the available constraints of time and funding.

In addition to the major tests summarized in Table 1, two impromptu test series were conducted to answer specific technical questions that arose during the project. These impromptu tests were (1) measurement of the average pressure as a function of height above

Fig. 9. Equipment arrangement for solids discharge tests.

the distributor in the tapered and straight-wall beds, and (2) measurement of the plugging tendency of the small stainless steel particles in vertical standpipes.

The pressure measurements were made to provide the information needed to estimate local velocities (i.e., the velocities at different heights above the distributor) in the different beds. They were accomplished by attaching a Baratron pressure transducer to pressure taps at various vertical locations on the fluidized bed (venting one side of the transducer to atmosphere) and recording the pressure signals without any high-pass filtering. Mean pressure values were determined by averaging 10,000 measurements.

The plugging tests were conducted to evaluate the difficulty of refluidizing the small particles once they had settled in a small diameter standpipe. These were made using vertical plexiglass tubes of 5/8-, 3/4-, and 1-in. diameter with air supplied through a 200-mesh screen at the bottom end. Small stainless steel particles were poured into the tubes to depths from 2-12 in., and the air flow gradually increased until the solids were well fluidized or blown from the tube.

Table 1. Summary of major test conditions

5.5 in. dia. straight-wall bed								
Fluidization					Heat transfer		Solids discharge	
	Large stainless steel		Small stainless steel		Large stainless steel	Small stainless steel	Large stainless steel	Small stainless steel
Distributor	Bed height (in.)	No. of runs*	Bed height (in.)	No. of runs*	No. of series**	No. of series**	No. of series**	No. of series**
"Flat"	28.9	6	20.3	8	2	3	2	2
	24.3	6	30.0	8	-	-	-	-
	19.0	6	24.9	8	-	-	-	-
"Top hat"	19.0	4	-	-	-	-	2	-
"Angle"	-	-	-	-	-	-	1	-
"Valley"	-	-	-	-	-	-	2	3
"Conical"	28.0	15	30.0	15	3	4	4	2

Modified constant-taper bed								
Fluidization					Heat transfer		Solids discharge	
	Large stainless steel		Small stainless steel		Large stainless steel	Small stainless steel	Large stainless steel	Small stainless steel
Distributor	Bed height (in.)	No. of runs*	Bed height (in.)	No. of runs*	No. of series**	No. of series**	No. of series**	No. of series**
"Flat"	29.5	6	30.0	6	2	2	3	1
	24.3	6	27.3	5	-	-	-	-
	19.0	6	24.5	5	-	-	-	-

*For fluidization tests, each run consisted of fluidization measurements at one gas flow rate.

**For heat transfer and discharge tests, a series consisted of several measurements for one design configuration.

3.0 DISCUSSION OF RESULTS

3.1 FLUIDIZATION TESTS

As reported earlier,² removal of the vertical solids downcomer results in a significant reduction in slugging intensity. Examples of this reduction are illustrated in Figs. 10(a) and 10(b), which are taken from the previous report. Note that the y-axis measures slugging intensity as it is reflected in σ_0 , the standard deviation of the pressure-drop signal. In this case, the signal used is that coming from the Baratron transducer.

In a similar fashion, Figs. 11-14 illustrate the effects of the various design changes investigated in this study. The y-axis in each of these figures represents the intensity of the pressure fluctuations observed for the specified configurations, with higher values indicating more intense slugging. Here, comparisons are made both in terms of σ_0 , the standard deviation of the pressure-drop, and σ_1 , the standard deviation of the first derivative of the pressure drop. Measurements from both the Schaevitz and Baratron transducers are included.

The x axis in these plots is given in terms of mass flow rate rather than velocity so that all of the designs are compared consistently on a gas throughput basis. That is, we want to know the relative amounts of slugging in the various beds when equal amounts of fluidizing gas is fed to them. Comparisons based on superficial gas velocity can be misleading because of the differing (and, in some cases, variable) diameters of the various configurations.

The legends referring to "Upper holes open" and "Lower holes open" in Figs. 11-14 refer to the conical distributor tests in which the upper or lower zone of purge holes between the standpipe and the plenum were open [see Fig. 7(a)].

The apparent trends in Figs. 11-14 and our simultaneous visual observations of the different bed designs indicate the following.

- Removing the steeply tapered section at the bottom of the tapered bed greatly reduces the severity of slugging, especially for the smaller particles. This slugging reduction, combined with the reduction achieved by removing the vertical downcomer, represents a major improvement in fluidization performance.
- The slugging differences between the modified tapered bed and the straight bed are relatively small. It appears that the fluidization quality differences are not sufficiently large and consistent to justify the continued requirement of using only tapered beds in Y-12.
- The fluidization quality differences among the three configurations of the conical distributor in the straight bed appear to be relatively small also. Issues other than fluidization quality (e.g., heat transfer and discharge performance) must be considered to select which of these designs is best.

Figs. 10(a) (upper) and 10(b) (lower). Comparison of slugging tendencies of stainless steel in the original Y-12 tapered bed design with and without the vertical downcomer. These figures are taken from the previously reported test series (2). In Fig. 10(a), the comparison is for 35×100 mesh particles, and in Fig. 10(b), the comparison is for 18×50 mesh particles. Slugging intensity (y-axis) is based on the standard deviation of the signal from the Baratron transducer indicating "local" pressure drop.

Fig. 11(a) and 11(b). Comparison of the slugging tendencies of the small stainless steel particles in the fluidized bed designs evaluated in this study. In Fig. 11(a), the slugging measure (y-axis) is based on σ_0 , the standard deviation of the as-measured pressure-drop signal from the Baratron transducer. In Fig. 11(b), the slugging measure is based on σ_1 , the standard deviation of the first derivative of the Baratron pressure-drop signal.

Figs. 12(a) and 12(b). Comparison of the slugging tendencies of the small stainless steel particles in the fluidized bed designs evaluated in this study. Slugging measure is based on (a) the standard deviation of the "global" pressure-drop signal from the Schaevitz transducer and (b) the standard deviation of the first derivative of the Schaevitz signal.

Fig. 13(a) and 13(b). Comparison of the slugging tendencies of the large stainless steel particles in the fluidized bed designs evaluated in this study. Slugging measure is based on (a) the standard deviation of the pressure-drop signal from the Baratron transducer and (b) the standard deviation of the first derivative of the Baratron signal.

Figs. 14(a) and 14(b). Comparison of the slugging tendencies of the large stainless steel particles in the fluidized bed designs evaluated in this study. Slugging tendency is measured in terms of (a) the standard deviation of the Schaevitz signal and (b) the standard deviation of the first derivative of the Schaevitz signal.

3.2 SOLIDS DISCHARGE TESTING

With either the modified tapered or straight-wall bed designs, there is no longer a vertical downcomer entering through the top of the bed that can be used for solids removal after processing. Thus, an alternate method must be used for solids discharge. Maintaining appropriate standards of uranium accountability requires that the revised discharge method should not permit any significant solids residue to remain following a reasonable discharge interval. Discharge tests were conducted to identify potential problems with residual solids accumulation. Because the solids discharge problems are essentially equivalent for the modified tapered and straight bed designs, most of the discharge tests were conducted using just the straight bed.

Emptying of the modified tapered bed following fluidization tests was accomplished by discharging the solids through a 3/4-in. port in the side wall near the bottom of the bed. Both the large and small stainless steel particles flowed readily through the port when air was injected through the distributor to partially fluidize the bed. However, once the solids level fell below that of the port, the discharge rate dropped to a trickle. It was evident that significant solids residues (e.g., a few grams) were likely to remain in the bed unless the solids discharge level were lowered to the top of the distributor plate. A test with same distributor (the "flat" design in Fig. 4) in the straight-wall bed gave similar results.

At this point the distributor was modified to the "top hat" design depicted in Fig. 5. The stepped central portion of the distributor brought the lowest solids level up to the side wall outlet level, and solids discharge was significantly improved. A few residual particles were still left on the distributor, however, because there was no slope to drive them directly toward the outlet. These particles left only when their random motions (induced by air flow through the nozzles) brought them near the outlet.

The "angle" and "valley" distributors in Figs. 6(a) and 6(b) were constructed to provide a sloping surface that would tend to enhance solids movement toward the side outlet. Slopes were deliberately kept shallow ($<6^\circ$) to avoid affecting air distribution. The "valley" performed somewhat better than the "angle," and both of these were better than the "top hat." The sloping designs did tend to channel the residual particles toward the outlet, but there was still a tendency for some particles (amounting to a fraction of a gram) to collect at the side of the distributor opposite from the outlet. This tendency appeared to be due to the relatively shallow slopes.

The "conical" distributor shown in Fig. 7 was tried last. In this design, the solids can be removed either by discharging through the bottom central outlet or through the small "snorkel" inserted through the side wall. With either discharge port, the solids removal appeared to be superior to any of the other distributor designs. We attribute this improvement to tendency of the solids to funnel readily down into the bottom of the cone where they can be most effectively conveyed away. One apparent shortcoming of the conical design tested is that there is still a nearly flat region between the wall and the nozzles where a few particles can collect. We believe this problem can be solved by increasing the slope near the wall, as described later.

Additional problems were encountered with the conical distributor when fluidizing the smaller stainless steel particles in the bottom-outlet configuration. These problems were formation of an unfluidized solids zone near the bottom of the standpipe and backflow of solids through the gas inlet holes (purge holes) into the plenum. When the side-mounted snorkel was not used and the vertical standpipe unplugged, solids tended to collect in the standpipe as soon as they were added to the bed. The side holes connecting the standpipe with the plenum were installed to provide a self-regulating air supply that would tend to keep the particles in the standpipe fluidized, but the initial location of these holes near the bottom of the plenum proved unsatisfactory. The pressure drop appeared to be insufficient to provide enough air to fluidize the column of solids. Attempts to refluidize the solids in the standpipe by injecting air through the bottom were also unsuccessful, apparently because the solids had become too tightly packed. During the refluidization attempt, the standpipe pressure increased sufficiently to cause some particles to backflow through the holes into the plenum.

Relocating the holes close to the upper end of the standpipe [as shown in Fig. 7(a)] allowed fluidization of the upper portion of the solids column, but the lower portion was still defluidized. The lower portion could be refluidized by injecting air through the bottom of the standpipe. Particle transfer through the side holes into the plenum was no longer observed.

3.3 SOLIDS MIXING (HEAT-TRANSFER) TESTING

Figures 15 and 16 illustrate the solids mixing and heat-transfer effects of changing from the modified tapered-wall to the straight-wall bed design. From these figures, it appears that the drop in solids mixing and associated heat transfer with decreasing velocity occurs at slightly higher gas flows in the straight bed than in the tapered. This drop is consistent with our expectations because the geometry of the tapered bed is more conducive to vertically uniform fluidization just above minimum fluidization. In the straight-wall bed, the onset of fluidization would be expected to vary more from the bottom to top because there is no taper to counteract the vertical pressure variation. The magnitude of the heat-transfer difference between the tapered and straight-wall designs does not appear significant enough, however, to warrant a preference for the former. The observed differences are nearly within the range of measurement error.

The heat-transfer impact of the various revisions to the solids discharge configuration of the conical distributor appears to be more significant. The results depicted in Fig. 17 indicate that heat transfer and mixing are most negatively affected when the gas flow to the central portion of the bed is restricted. Thus, it is clearly important to have sufficient gas flow through either the standpipe opening or solids discharge "snorkel" during normal operation to "stimulate" the central region of the bed above the distributor. Comparisons with the performance of the "flat" distributor also suggest that more than one ninth of the total fluidizing gas should be injected through the standpipe or the "snorkel" to provide this stimulation. The higher central flow is probably necessary because the gas injection velocity for the conical distributor is lower than it is through the central nozzle of the "flat" distributor.

Fig. 15. Comparison of the heat-transfer characteristics of the modified tapered and straight-wall fluidized bed designs for the small stainless steel particles. Measurements from two different levels (near the distributor and near the middle of the bed height) are included. The "flat" distributor is used in both cases.

Fig. 16. Comparison of the heat transfer characteristics of the modified tapered and straight-wall fluidized bed designs for the large stainless steel particles. Measurements from two different levels (near the distributor and near the middle of the bed height) are included. The "flat" distributor is used in both cases.

Fig. 17. Comparison of the heat transfer characteristics of the flat distributor in the straight and tapered beds with various versions of the conical distributor in the straight bed. All heat transfer measurements were made 3 in. above the distributor.

3.4 OPEN- AND CLOSED-LOOP FIM TESTING

Open- and closed-loop FIM tests were conducted with the straight-wall bed using both the large and small-particle stainless steel. The distributor was the conical design illustrated in Fig. 7, with the side-mounted snorkel removed and air injected through the vertical standpipe via the upper side orifices connecting with the plenum. Operation of the system was evaluated for both solids at static bed heights of 29 and 19 in.

The behavior of the FIM in both open and closed-loop operation was quite satisfactory and consistent with the observations made in previous studies. The FIM output appeared to track well with the observed slugging and/or bubbling intensity, and the closed-loop operation was stable. In general, it appeared that optimum fluidization intensity (i.e., an optimum balance between mixing and minimal slugging) was achieved at fluidizing air flows about 10% above the flow at minimum fluidization (see Table A.1 in the Appendix for the minimum fluidizing velocities of the two sizes of stainless steel particles).

3.5 VERTICAL PRESSURE VARIATION AND STANDPIPE PLUGGING TESTS

Measurements of vertical pressure variation revealed that the average pressure decreased linearly with height above the distributor in both the tapered and straight-wall beds. The results of these measurements are summarized graphically in the Appendix. This result implies that superficial gas velocity also varies linearly with vertical position, and thus local velocities can be determined directly from the overall bed pressure drop and local height.

The standpipe plugging tests revealed that in all cases the small stainless steel particles could be easily refluidized from a completely static condition, even when the standpipe diameter was only 5/8 in. and the depth of solids was 12 in. This result indicates that the refluidization problems encountered with the standpipe in the conical distributor were not due to the inherent plugging tendencies of the solids but simply a lack of sufficient pressure drop over the side inlet gas holes (i.e., the holes connecting the plenum and standpipe). In this case, the refluidization problem with the conical distributor can be solved either by raising the level of the inlet holes (as was done) or increasing the pressure drop over the plenum by decreasing the size of the fluidization nozzle holes. The latter change would probably be made anyway at Y-12 if the standard procedures are modified to routinely use smaller particles.

3.6 UNRESOLVED ISSUES

There remains some uncertainty about whether the bottom or side discharge outlet configuration (with the conical distributor) is more appropriate. Apparent advantages of the bottom-outlet design are that

- there is no need for a side-wall penetration of the reactor vessel to accommodate the discharge line, and

- the fluidizing gas to the vertical standpipe is automatically supplied from the plenum without the need for a separate gas regulation system.

An apparent disadvantage of the bottom-outlet design is that some of the solids (below the gas inlet holes) remain defluidized unless additional purge gas is injected through the conveying line. The height of this defluidized section can be reduced by using a smaller plenum, as shown in Fig. 18. The potential effects of those disadvantages are now unknown, but they should be carefully considered before adopting the bottom-outlet design.

The apparent advantages of the side-wall outlet (short "snorkel") are that

- no significant accumulation of defluidized solids is expected, and
- because purge gas is added externally, it can be adjusted to any desired value, and there is less likelihood that it can become obstructed.

The potential disadvantages are that

- purge airflow is not regulated automatically and must be supplied separately, and
- the solids transfer line protrudes into the fluidized solids, disturbing the local fluidization quality.

Further studies will be needed to select which option is definitely superior.

Fig. 18. Recommended distributor design for plant implementation.

4.0 OTHER POSSIBLE STRATEGIES FOR IMPROVING OPERATION

During the course of this investigation, it was recognized that further improvements in the Y-12 fluidized bed performance could also be obtained by modifying the standard operating procedures. While it was not possible to investigate these further within the budget and schedule constraints of this project, we strongly recommend that the following three procedural modifications be considered for future implementation. We believe that these modifications could be implemented with little or no changes in equipment and could yield substantial further reductions in reactant gas consumption and scrubber waste production.

- **Programmed gas flows.** The essential idea here is to use varying fluidization gas flows during the processing of each batch, allowing optimization of the mixing and reactant requirements. For example, it is recognized that good mixing and heat transfer are most important at the beginning of the fluorination reaction when it is most likely that excess particle heating and sintering will occur. The need for good mixing and heat transfer is much reduced during the "soak" stage when only a last residual fraction of uranium oxide is being converted. Thus, instead of operating the beds at a single level of fluidization for the entire batch processing period, a more optimum strategy would be to begin with a higher fluidization intensity and gradually reduce the setting as conversion increases (e.g., by slowly reducing the FIM setpoint in closed-loop operation). For the soak stage, the fluidization state could probably be maintained just above minimum fluidization to minimize excess reactant gas consumption.
- **Pulsed operation.** This modification would use the fact that mixing and heat transfer can be induced by periodic increases in the fluidizing gas flow (e.g., flow surges every 20 s) while maintaining a much lower flow (reduced fluidization intensity) at other times. The advantage of such operation would be to confine excess gas flow to relatively short intervals, thereby reducing the overall average consumption. The frequency and amplitude of the pulsing could also be varied over the course of each batch consistent with programmed flow operation.
- **Smaller particle size.** As has been widely reported in the fluidized bed literature, fluidization tends to be smoother for the smaller particles (i.e., those classified as belonging to Geldart's Group B).³ This well-known phenomenon also applies to the uranium oxide and tetrafluoride particles used at Y-12. No current statistics are available on the mean and variance of the denitrator particle sizes.

However, it is suspected that the particle sizes of at least some batches are beyond Geldart's Group B into Group D. If the preceding is true and if it is possible to reduce the particle size of the initial solids fed to the fluidized beds (i.e., the size of particles coming from the denitrators), the fluidization should be smoother with more bubbling and less slugging. The smoother fluidization should yield improved gas-solids contacting efficiency and much-reduced reactant gas flows. The latter results because of the large reduction in minimum fluidization velocity that occurs with reduced particle size. For the Y-12 beds, we believe that the optimum particle size will be between 100-300 microns. Particles less than 100 microns may exhibit agglomeration and clumping problems.

5.0 CONCLUSIONS AND RECOMMENDATIONS

Based on the results of this latest study, we conclude the following.

- The steep taper cone at the bottom of the current Y-12 tapered bed has a clear negative impact on fluidization quality. Elimination of this design feature, whether for tapered or straight wall configurations, should improve performance substantially.
- The shallow taper region of the existing bed produces no significant improvement in fluidization quality over straight walls. We do not believe that the difference in performance is sufficient to justify the continued use of the tapered design.
- Of the distributor designs evaluated, the nine-nozzle "conical" type is the best. Because there is a tendency for a small amount of solids to settle on the flat region near the wall during discharge, the design in Fig. 18 is expected to be better because it should eliminate the settling region near the wall.
- The FIM performs satisfactorily with the straight-wall bed and nine-nozzle conical distributor in conjunction with either open- or closed-loop control.

For future upgrades and/or replacements of the Y-12 fluidized beds, we recommend the following:

- replace the existing tapered-wall beds with a straight-wall design having no tapered sections and a conical distributor;
- remove the axial solids discharge line (i.e., "downcomer"), and substitute a side or bottom outlet; and
- conduct further tests to determine clearly whether a side or bottom solids discharge outlet is preferable. In the absence of this information, select a side outlet.

6.0 REFERENCES

1. C. S. Daw, Martin Marietta Energy Systems, Inc., letter to J. A. Hawk, Martin Marietta Energy Systems, Inc., April 17, 1991.
2. C. S. Daw and J. A. Hawk, Martin Marietta Energy Systems, Inc., letter to P. Adamson, Martin Marietta Energy Systems, Inc., December 26, 1991.
3. D. Geldart, "Types of Fluidization," *Powder Technology*, 7, 285-92 (1973).
4. D. Geldart, *Gas Fluidization Technology*, p. 41, John Wiley & Sons, New York, 1986.
5. M. Kumada, I. Mabuchi, S. Kume, Y. Watanabe, and M. Hirata, "Characteristics of Dynamic Behavior and Local Heat Transfer Around Single-Row Tubes Immersed in Floating Particles, Experimental Heat Transfer, Fluid Mechanics and Thermodynamics, pp. 1301-1309 in *Proceedings of the First World Conference, Dubrovnik, Yugoslavia, September 1988*.

APPENDIX

Table A.1. Summary of stainless steel particle properties

Particle density = 8.0 g/cc = 500 lb/ft³

35 × 100 mesh

Sieve analysis

Size (m)	Weight %
500	<99.6
300	<96.8
250	<94.3
180	<61.9
150	<4.5
125	<0.7

Sauter mean size = 183 m

Minimum fluidizing velocity = 0.32 ft/s

18 × 50 mesh

Sieve analysis

Size (m)	Weight %
1180	<100.0
850	<95.0
600	<62.3
500	<23.8
300	<0.8

Sauter mean size = 565 m

Minimum fluidizing velocity = 2.20 ft/s

Note: The particle size distributions and minimum fluidization velocity were reevaluated at the beginning of this study to account for any changes that may have occurred due to attrition, fines loss, and the addition of new material. Slight changes were found from the sizes and minimum fluidization velocities measured previously.

Fig. A.1. Schematic for the Fluidization Intensity Module

Figs. A2(a) and A2(b). Variation of velocity with height in modified tapered bed. Part (a) is for the small particles, and part (b) is for the large particles.

Figs. A.3(a) and A.3(b). Variation of velocity with height in straight-wall bed. Part (a) is for the small particles, and part (b) is for the large particles.

Fig. A.4. Calibration of heat transfer probe in air. Theoretical values are from *Fundamentals of Heat and Mass Transfer*, F. P. Incropera and D. P. DeWitt, 3rd ed., John Wiley & Sons, New York, 1990, pp. 132 and 240.

Figs. A.5(a) and A.5(b). Variation of minimum fluidizing velocity with temperature. Part (a) illustrates the variation for different sizes of stainless steel in air. Part (b) illustrates the variation for 500 micron particles of UO_3 , UO_2 , and stainless steel fluidized by various gases.

Fig. A.6. General Fluidization Classification proposed by Geldart.

Internal Distribution

1. W. J. Bicha	36. M. D. Ohara/Y-12 Development
2. M. L. Brown - DMC	37. Process Development
3. H. R. Butler	38. J. H. Pashley - K-25
4-19. C. S. Daw	39. G. L. Pfennigwerth
20. J. E. Hardy	40. R. M. Schilling
21-31. J. A. Hawk	41. C. E. Sliski
32. T. S. Kress	42. J. F. Thomas
33. J. A. Kreykes	43. Y-12 Central Files
34. D. W. McDonald	44. Y-12 Technical Library
35. G. N. Miller	45. X-10 Library

External Distribution

- 46. J. S. Halow, U.S. Department of Energy, P.O. Box 880, Morgantown, WV, 26504
- 47-48. A. K. Lee, Office of Scientific and Technical Information, P.O. Box 62, Oak Ridge, TN, 37831
- 49-50. Ke Nguyen - Department of Mechanical/Aerospace Engineering, UTK
- 51. Randy Robinette, H&R Technical Associates, 151 Lafayette Drive, Oak Ridge, TN, 37830

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