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ABSTRACT

Liquid-solid fluidization studies at room temperature were initiated to evaluate the severity of instabilities associated with fluidized beds. Factors studied include:

1. The effect of gas bubbles on bed stability
2. The effect of inclining the bed at angles up to 20° from the vertical.
3. Particle size distribution
4. Agglomeration of particles into flocs
5. Concentration range for a stable well defined interface.

It is believed that none of these factors gave effects which would eliminate the consideration of fluidization techniques in the design of nuclear reactor cores.

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SUMMARY

Liquid-solid fluidization studies at room temperature were initiated to evaluate the severity of flow instabilities associated with fluidized beds. In these exploratory tests, bubbling a moderate amount of gas through the fluidized bed completely disrupted the flow pattern. Inclining the cylindrical container at angles up to 20° from the vertical did not appear to affect the stability of the upper interface between the fluidized bed and disengaged water; however, at the larger angles pronounced channeling of the water was observed. Agglomeration was indicated in hindered settling tests even though the mean primary particle diameter was about 15 microns. Since the fluidization and hindered-settling velocities were essentially equal, the agglomeration may have existed in fluidization tests also. None of the above instabilities are believed to be sufficiently serious to eliminate the use of the fluidization technique in the design of nuclear reactor cores.

Based on heat transfer considerations, calculations indicate that ThO_2 particles 20 to 30 microns in diameter would be of interest for large, single-region fluidized-bed reactors. Thorium oxide particles of 150 to 1500 microns diameter would be of interest for the blanket region of two-region breeder reactors.

^a L. D. Felten was a summer student employee with the Research and Analysis Section of REED Division during the summer of 1958.

INTRODUCTION

Nuclear reactors utilizing the fluidized solids technique offer the desirable possibilities of

- (a) A high heat transfer rate
- (b) A configuration on which the fuel is retained in the core region and the moderator is circulated to the heat exchanger, thus diminishing the activation of the components in the external loop system.

Particular applications have been discussed^{17, 27, 28} and possible instabilities and problems have been reviewed.^{29, a}

This report presents the results of experimental investigations, at room temperature, of the fluidization characteristics of aqueous suspensions of ThO_2 spheres. The problems studied were:

- (a) The agglomeration characteristics of ThO_2 particles.
- (b) The concentration range for a well defined, stable, interface.
- (c) Comparison of fluidization velocities with hindered settling velocities to establish a basis for possible use of high temperature settling measurements to determine fluidization velocities.

Several tests were also made in which aqueous suspensions of glass spheres were fluidized in an 11-in. dia by 1-ft high column having a top section which tapered outward. The object was to determine if the conical section would concentrate the solids sufficiently to prevent carry-over.

In addition, calculations of particle diameter and flow rate are presented for several possible configurations of core and blanket region to illustrate the range of particle characteristics of interest for use in nuclear reactors.

^a Possible instabilities and problems as outlined in HRP-58-259 are included in Appendix A of this report.

EXPERIMENTAL

A. Fluidization Characteristics of ThO₂ Particles

The first series of tests were made with 15-micron diameter ThO₂ spheres prepared³⁰ in the 200A loop and refired at 1200°C. Fluidization and settling tests were done in the same 4-in. ID, 3-ft long glass column; a porous stainless steel plate was used as a support plate and water flow distributor. (See Fig. 1) The settling rate of ThO₂ in the 4-in. column was measured by timing the rate of fall of the bed interface, which was observed visually. Although there were always some ThO₂ particles in the water above the interface of the fluidized bed, the dust did not obscure the interface until the volume fraction solids was less than 0.05. For volume fraction solids less than 0.05, a light was used behind the glass column so that the interface could be observed. The superficial fluidizing velocity was calculated from the volumetric flow rate, and the free area of the glass pipe. The total thorium oxide concentration in the column, and thus the L/D ratio was varied by a factor of three (from 424 grams ThO₂ to 1246 grams ThO₂ as noted in Fig. 3). No significant difference in operation was noticed as a consequence of this variance.

During rapid expansion of the fluidized bed, the interface becomes quite turbulent as bubbles of water come through the bed. This situation, however, ceases to exist as a steady state is reached, and the interface becomes calm. At volume fraction solids above 0.10 the interface at the top of the fluidized bed was quite sharp. At lower volume fraction solids the interface became rather indefinite. The thickness of the diffuse region as a function of volume fraction solids is shown in Fig. 2. For volume fraction solids less than 0.025 the interface smeared out and there was pronounced carry-over.

When the fluidization column was operated at solid fractions greater than 0.027, the water leaving the top of the column appeared to be completely clear. However, after 50 - 100 hrs operation, some ThO_2 particles were noticed to have settled out in the exit tube of the column. Hence, even at low water flow rates (and thus high solid fractions)^a some solids are carried over by the water. No estimate was made of the rate of carry-over and the possibility of carry-over of solids adhering to minute gas bubbles was not eliminated.

Flocculation Effects --- To estimate the degree of flocculation, a plot of log settling rate versus volume fraction solids (Fig. 3) was compared to similar empirical curves which were determined by Steinour¹ and by Richardson and Zaki.² Through these comparisons a floc density was calculated for the settling particles, and this density was used in Stokes' Law to calculate an equivalent spherical diameter of the particle flocs. The results are tabulated in Table 1.

Table 1. Effective Diameter of Thoria Flocs

	Particle diameter calculated from settling rate at infinite dilution assuming particle density of 10.	Floc density and diameter calculated from comparison of experimental data with empirical correlations presented in the literature.	
		Steinour	Richardson and Zaki
Floc Density (g/cm^3)	-	8.02	6.61
Floc Diameter (microns)	22	25	28

Apparently some flocculation had occurred since the mean diameter of the "as fired" particles was found to be 15 microns by sedimentation analysis and 14.3 microns by measurement with electron microscope. A discussion of particle size measurement follows later.

^a The range of volume fraction solids studied where a clear stable interface with some carry-over was observed was 0.027 to 0.20.

In order to compare the results for fluidization and hindered settling tests, the fluidization velocity for each run is plotted on Fig. 3 along with the settling rate. The best straight line through the velocity data according to the method of least mean squares is seen to lie quite close to the settling rate curve. This feature is examined more closely in Fig. 4 where the ratio of fluidization velocity to settling rate is plotted versus volume fraction solids. It is seen that this ratio remains quite close to a value of one over the range studied, the most extreme point differing from unity by twelve percent. These data have an estimated standard deviation of 0.0447, and the mean ratio is 1.015. Through the use of the "t" test it is found that within 90% confidence limits the true mean lies between 1.0004 and 1.0293. Thus the fluidization velocity and the hindered settling velocity are essentially identical for this series of experiments. There appeared to be little trend with change in volume fraction solids.

Orientation Effects --- In order to evaluate the effect of column orientation, the column was tilted at various angles up to 20° from vertical, and the ThO_2 was fluidized with the column in these positions. Although the bed interface remained fairly level across the column, there was severe channeling of the water up the higher side of the tilted column. This did cause some rippling of the interface especially at volume fraction solids below 0.05. This upward channeling of water also occurred during settling of the ThO_2 particles. The ThO_2 particles moved downward on the lower side of the column during both fluidization and settling. The channeling was pronounced with the column tilted 5° from vertical, and it could not be determined visually that the situation became any worse as the column was tilted at greater angles. The mean ratio of fluidization velocity to settling rate for all angles studied was greater than unity,

as shown in Fig. 5. As can be seen, there is some indication that the fluidization velocity increased more rapidly than the hindered settling velocity as the volume fraction solids increased. It must be emphasized that this increase in the ratio of fluidization velocity to hindered settling velocity was never greater than 15% even for 20° tilt to the column and 0.20 volume fraction solids, so even if the effect is real, it is nevertheless quite small. None of the data is definitely conclusive on this subject. The spread of data was almost as bad at the different angles as it was with the column vertical. The estimated standard deviations of the data ranged from 0.0318 to 0.0447, which was the value for the vertical column.

Simulated Boiling --- In further efforts to demonstrate column instabilities, air was bubbled through the fluidized bed of thorium oxide. During the first run the bed was at a volume fraction solid of 0.178, and the air entered at a distance 40% of the total bed height below the bed surface. The air rate was set at approximately one bubble per second. Each bubble carried ThO_2 several centimeters above the bed surface as it broke through the interface. Most of this ThO_2 immediately began to settle back into the bed. The bubbling did, however, increase the particle carry-over. At the same bubbling rate but at a lower volume fraction solids the bed interface became more turbulent from the disturbance caused by the breakthrough of air bubbles. As the bubbling rate was increased the interface became much more turbulent until the whole column became filled with ThO_2 , and the interface disappeared completely. Particle carry-over was now extremely high although the column was still "bubbling" and not "slugging" at this point.

Particle Size Distribution --- Two runs were made at solid fractions of 0.040 and 0.032 where samples were drawn from the column with a 25-milliliter pipette after the column had operated for two hours. Samples were taken at 10-centimeter intervals of height with the bottom sample taken 10 centimeters above the diffusion plate.

For both runs the particle size distribution was fairly constant throughout the bed. There was some indication that fine particles were slightly concentrated near the top of the bed.

A later run from which samples were taken after four hours of operation showed a much more marked particle size gradient from top to bottom. In this experiment, three samples were taken: one immediately above the diffusion plate, one from the middle of the bed, and one just below the bed interface. The mean particle size was 18 microns for the top sample, 25 microns for the middle sample and 34 microns at the bottom of the bed.

During the course of the tests the mean particle size of the material retained in the bed was observed to increase. For instance, the mean particle size increased from 17 to 20 microns in samples from the two consecutive two-hour runs. The mean particle size in the third run, which was made after the bubbling experiment in which there was considerably carry-over, was even higher. Apparently the smaller particles were preferentially carried over by the fluidizing water; a sample of carry-over particles was analyzed and found to have a mean diameter of 13 microns.

Concentration Gradient Studies --- Concentration measurements were made from the samples taken from the two-hour runs mentioned above. The measurements were somewhat crude, but there was a definite decreasing gradient from the bottom of the bed to the top. The results of the measurements are shown in Tables 2 and 3.

TABLE 2

Concentration of ThO_2 Samples from Column after Operating for two hours
at Volume Fraction Solids of 0.040 and a Bed Height of 36 cm.

<u>Distance from Base of Bed</u>	<u>Volume Fraction Solids</u>
10 cm	0.032
20 cm	0.029
30 cm	0.027

TABLE 3

Concentration of ThO_2 Samples from Column after Operating for two hours
at Volume Fraction Solids of 0.032 and a Bed Height of 45 cm.

<u>Distance from Base of Bed</u>	<u>Volume Fraction Solids</u>
10 cm	0.0232
20 cm	0.0228
30 cm	0.0221
40 cm	0.0172

Experimental Difficulties and Errors

Several sources of error were encountered during the experiments, e.g.:

- a. Velocity Fluctuations. Water for fluidization was run directly into the column from the house water line. A change in water pressure and thus a change in flow rate was noticed on two or three occasions during measurement of a fluidizing velocity. This would cause an error in the measurement of the fluidizing velocity, and through it, an error in the ratio of fluidizing velocity to settling rate.
- b. Diffuse Interface. The uncertainty of the exact position of the interface at low solid fractions gave rise to error in the measurement of settling rate. The falling interface left behind it so much dust and cloudiness in the water that it was sometimes almost impossible to follow. However, the bed fell through a larger distance during settling from low solid fractions than from high solid fractions, and, therefore, the instantaneous interface position was not required to be as accurate as for the tests at higher solid fraction to get the same precision in settling rate.
- c. Particle Carry-over. Another possible error originated from the carry-over of ThO_2 particles. Carry-over did not appear great enough to affect the results of any single series of runs by itself. During the settling rate runs in the vertical column there was always a known amount of ThO_2 placed into the column. However, for all succeeding runs the amount of ThO_2 in the column was not known exactly, because an unknown amount was lost each time the carry-over

ThO₂ was returned to the column from the catch pots. The amount lost was estimated from known losses when all the ThO₂ was removed from the column and later returned and appropriate corrections were made.

- d. Concentration Analysis. The withdrawal of samples from the column for concentration analysis was also subject to error. The ThO₂ particles settled to the bottom of the sampling pipette and tended to drop out as the pipette was removed from the column. This meant that only a qualitative comparison of concentration measurements could be made.
- e. Particle Size Analysis. Correlation of particle size analysis was always a problem because there was rarely a normal distribution of particles in the samples. The carry-over particles from the catch pot were the only ones which followed a normal distribution of particle size. The particle sizes from the second two-hour run approached a normal distribution, but these still had breaks and considerable deviations from the normal curve. The remaining particle distributions all varied greatly from a normal distribution.

All particle size distributions were determined on a weight basis except the analysis by electron microscope. This latter determination was on a count basis. The formula for changing mean particle diameter on a count basis to mean particle diameter on a weight basis is $D_{pmw} = D_{pmc} e^{3\ln^2 \sigma_c}$ and is derived for a normal distribution. The size distribution analysis on a count basis from the electron microscope resulted in a curve with one major break. In converting to a weight basis the branch containing the heavier particles is probably the most correct

one to use for as a normal distribution. From this branch D_{pmc} is 8.9 microns, and the standard deviation, σ_c , is 1.49. From the formula then, the mean particle diameter on a weight basis, D_{pmw} , is 14.3 microns. If the branch containing only the smaller particles is used, D_{pmc} is 3.9 microns, σ_c is 9.5, and D_{pmw} would be 1550 cm which is obviously incorrect.

The other particle size analyses are based upon Stokes' law. Any particle sticking or flocculation which occurs in spite of the dispersing precautions will cause an error in the size determination. The resulting measurement will be increased because of the larger diameter of the particle floc and will be decreased because of the lower density of the floc from the individual particles. Fortunately these errors are in opposite directions.

B. Preliminary Investigation of Effects of Container Geometry on Fluidization
Characteristics of Glass Beads.

A few qualitative tests were made with glass beads in a one-foot diameter Lucite column with a cone top. This column had an inside diameter of 11 inches up to a height of 14 inches. From this point the column was flared outward at an angle of 29 degrees to an inside diameter of 21 inches. The support plate was of 1-1/2-inch Lucite in which 1/4-inch holes were drilled as closely together as possible. This support plate was backed up by two 325 mesh screens and a piece of pipe insulation covering, all of which served as a water distributor. The water left through a weir which was cut into the top of the column wall and drained into a 30-gallon drum from where it could be pumped back into the column. The glass beads had a density of 2.50 g/cm^3 and ranged in diameter from 61 microns to 147 microns.

During fluidization many glass beads stuck to the side of the column. Agglomeration was quite noticeable, and many clumps of beads would rise to the top of the column where some of them would break up and fall back into the bed, and some of them would carry into the overflow. Beads which were scraped loose from the side of the column were always agglomerated and seemed to remain that way.

The settled bed density of unagglomerated glass beads of the size used here was found to be 1.47 g/cm^3 and corresponds to a volume fraction solids of 0.59. The bed in the column settled to a density of 1.82 g/cm^3 which corresponds to a fraction solids of 0.55. This also indicates a slight amount of agglomeration. Steinour,³¹ in his experiments with glass beads took precautions against agglomeration by adding a slight amount of sodium hexametaphosphate to the fluidizing water.

Operation at low void fractions was accompanied by much channeling. When the bed reached a solid fraction of 0.17, channeling had stopped and the bed appeared quite homogeneous. Mixing of the bed appeared to be quite complete.

The bed was raised into the coned part of the column; the associated volume fraction solid was 0.0881. The beads fell from the coned part of the column, where the upward water velocity was lower, into the cylindrical section of the column. This action set up quite turbulent swirling and mixing patterns in the bed. The cone was quite effective in stopping all particles except agglomerated ones from carrying over.

One difficulty which was encountered with the support plate was that during fluidization (at all bed levels) there were some glass beads which remained unfluidized around the edge at the bottom of the column.

CALCULATION OF PARTICLE DIAMETER REQUIRED FOR VARIOUS
NUCLEAR REACTOR CONFIGURATIONS

A. Single-Region Reactors

Calculations have been made of the particle diameter and concentration required for operation of a 15-ft diameter single-region fluidized-bed reactor at various power levels. The results of these calculations are shown in Fig. 6. Fluidization velocities were obtained by extrapolating the results of hindered settling tests with "Standard Slurry" ThO_2 spheres³³ to 275°C using the Stokes' law coefficients. Values of the power for different diameter vessels may be obtained by multiplying values of the power from Fig. 6 by the square of the ratio of vessel diameters.

B. Multi-Region Blanket of a Homogeneous Reactor

Concentration of the ThO_2 in the blanket of a two-region reactor into a layer immediately adjacent to the core wall affords the possibility^a of decreasing the flux at the wall and increasing the breeding ratio with no increase in the inventory. It was suggested^a that one possible way of accomplishing this would be to fluidize a concentrated suspension in a region about six inches thick immediately adjacent to the core wall; a lower concentration slurry would be present in the outer blanket region. If the concentration in the narrow annular space approached that of the settled bed, the small amount of settling that would take place if the pumps were turned off would not add appreciable quantities of reactivity.^{b, 34} Because of these particular qualities, estimates have been made of the particle sizes, concentrations, and flow rates of interest for a reactor with 3-ft core diameter. One possible configuration is shown in Fig. 7.

^a R. B. Briggs - Personal Communication, Dec. 1957.

^b For a more detailed discussion of the nuclear characteristics of a similar configuration see ORNL CF-58-4-66, April 18, 1958 by Melvin Tobias and T. B. Fowler.

The calculations of required particle diameter as a function of concentration were made for fixed blanket powers of 100, 50 and 10 Mw; for a ΔT of 75°C, and a core diameter of 3 ft, with an annular gap of 6 inches. The power levels and ΔT correspond to velocities in the annular gap of 1.73, 0.86 and 0.173 ft/sec if there are no particles present and if it is assumed that all of the blanket power is generated in the annulus. The concentration in the gap was varied from 0 to 6000^a grams ThO₂/liter.

Required particle diameters are given in Fig. 8, and are based on the experimental results of Richardson and Zaki¹⁸ for sedimentation and fluidization of particulate matter under laminar, transition and turbulent flow conditions. (See Appendix B) The terminal settling velocities were calculated in the manner described by Bosanquet et al.^{36, 37} That the particle size is essentially independent of concentration at high powers and concentrations is due to the almost 8-fold decrease in the drag coefficient of single spheres as the Reynolds number increases from 2×10^5 to 4×10^5 . (The drag coefficient varies less than 50 per cent in the range of Reynolds numbers from 10^3 to 2×10^5). The sharp decrease in drag coefficient is attributed⁵ to the formation of eddies immediately behind the particle which travels with the particle, thus forming a streamlined body consisting of sphere and eddy. It is not clear that a similar situation would prevail in concentrated systems. Results calculated in the manner described in Appendix B are given in Fig. 9. Since it was assumed that there was no power generation in the outer region of the blanket, the free stream velocity^b slurry concentration and the particle diameter can be that required to satisfy the inner region conditions.

^a A concentration of 6000 grams ThO₂/liter corresponds to volume fraction solids of 60% which is essentially the porosity of a packed bed of uniform spheres.

^b The free stream velocity is defined as the velocity in the absence of particulate matter.

Application of the results in Figs. 8 and 9 to the configuration shown in Fig. 7 gives the following typical numbers:

1. Blanket Power, Mw	10	50	100
2. Blanket Concentrations, g ThO ₂ /liter			
a. Inner Annulus	4000	4000	4000
b. Outer Annulus	500	500	500
3. Blanket Flow Velocity, ft/sec			
a. Inner Annulus	0.17	0.86	1.73
b. Outer Annulus	0.11	0.11	0.35
4. Blanket Flow, GPM			
a. Inner Annulus (265°C)	400	2000	4000
b. Outer Annulus (265°C)	1000	1000	3200
5. Particle Diameters			
a. Inner Annulus (cm)	0.018	0.20	1
b. Outer Annulus (microns)	30	30	70

Tobias and Fowler³⁴ have shown that double blanket reactors with an annulus containing concentrated ThO₂ adjacent to the core produce 82 to 84% of their power in the core region with less than 8% of the total blanket power (< 1% of total reactor power) being generated in the more dilute blanket region. Thus for a reactor generating a total of 225 thermal megawatts, about 38 Mw would be generated in the blanket region. This corresponds to a required particle diameter in the inner annulus of 0.15 cm (1500 microns).

If the large particles in the inner annulus are ejected into the outer annulus, they will fall rapidly to the bottom where they will be retained by the grid. These large particles may be fed back onto the inner annulus through the slotted venturi. Any particles smaller than 200 microns will be carried from the inner annular fluidized region and can be removed either by hydroclones, settling chambers or filters. The very high flow rates in the outer annulus are a consequence of the one-foot annular thickness of this region, whether nuclear considerations will allow a thinner region is not known.

Qualitative heat transfer calculations¹ indicate that even for the maximum particle diameter and the 100 Mw blanket power level the temperature difference between the fluid and the surface of the particle is less than 5°F.

Finally, the following statement is cited as a disadvantage of fluidized solids systems by R. D. Morse in Lapples book on "Fluid and Particle Mechanics":³⁵

"Parallel or countercurrent flow cannot be secured in the simple fluidized bed; modifications which permit an approach to these types of action complicate the process, or reduce its other advantages."

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APPENDIX A

I. Liquid-Solid Fluidization Problems

A. High Segregation Rates

It has been observed that when there is a large density difference between the particle and the fluid that there is a tendency for the fluid to separate from the particles and form fluid bubbles and pockets. This has been observed specifically when fluidizing lead shot with water; clearly formed water bubbles rose through the bed to burst at the surface in a splash of lead shot. Zenz² has given the tentative correlation shown in Fig. 10. This shows that, for a ratio of particle to fluid density of 10 - 20, slugging can take place if the volume fraction solids range from 0.02 to 0.30 for particles having diameters less than 0.1 to 0.3 inches. This slugging has been observed in preliminary fluidization tests with ThO₂ spheres, and is characterized by a continuously rolling and wavy surface.

B. Viscosity

Gerald²⁰ has claimed to have observed anomalous viscosity effects in liquid-solid fluidization studies. Little work has been done on this subject directly and it would be extremely difficult to compare the work of different investigators who worked with different liquids.

C. Particle Size Distribution

In liquid-solid fluidization tests with non-uniform particles, the fines have been observed to segregate at the top of the bed.²¹ If the flow rate is in the right range, the particles in the bottom were in a turbulent state whereas the fines at the top were in a laminar condition. Richardson and Zaki¹⁸ report that when this segregated condition occurs the pressure drop

may be calculated by treating each region separately. Lewis and Bowerman¹⁰ make similar recommendations; although their results were taken with material of known-non-uniformity, they made no attempt to demonstrate the validity of such a calculation.

D. Entrainment

Betts²² reported that in fluidization tests with gas, particles greater than 250 microns diameter would be entrained at the free fall velocity of the particle and as the size decreased the operating gas velocity could be appreciably higher than this value. It is not clear from the material presented whether this was due to particle agglomeration for particles below 250 microns or whether some other mechanism was responsible.

E. Particle Agglomeration

With catalyst particles less than 10 microns in dia, cohesion becomes so severe²³ that balling of the solids into spheres of several millimeters in diameter may result. With particles of 25 microns, cohesion still controls^{6, 23} the appearance of the bed but balling is not in evidence. With catalyst of this size, the bed settles down at low superficial velocities and channels form through which the gas passes without intimately coming in contact with the bulk of the solids in the bed. Good fluidization is observed for particles greater than 40 microns.

Agglomeration of 15 micron ThO_2 spheres into clusters having a diameter of 40 microns has been observed in hindered settling tests with "Standard Slurry Spheres."

F. Channeling

Berglin²⁴ has observed that channeling does not cease until the bed has expanded to concentrations well below 0.4 volume fraction solids. Richardson and Zaki report that this channeling in relatively concentrated systems is favored at low liquid velocities by large density differences between the solid and liquid.

G. Location of Heat Source

Gas velocity²⁵ does not seem to exert any effect independent of concentration in the case of a bed heated internally with a "Calrod" heater, whereas it exerts an unquestionable effect of its own in the case of the externally heated unit, whether this is due to the "Calrod" in the center of the unit decreasing the slugging which tends to occur at high solids concentration and low gas velocities is not known.

H. Miscellaneous

Other problems that may cause difficulty include a suitably designed distribution plate, particle attrition, and container erosion by particles.

APPENDIX B

The empirical relations developed by Richardson and Zaki¹ for laminar, transition and turbulent fluidized flow conditions are:

$$U_f = \epsilon^n U_t \quad (1)$$

where

$$n = 4.65 \quad N_{Re} < 0.2 \quad (2)$$

$$n = 4.35 N_{Re}^{-0.03} \quad 0.2 < N_{Re} < 1 \quad (3)$$

$$n = 4.45 N_{Re}^{-0.1} \quad 1 < N_{Re} < 500 \quad (4)$$

$$n = 2.39 \quad N_{Re} > 500 \quad (5)$$

U_f = free stream velocity

(i.e., velocity of fluid immediately above particle interface)

U_t = single particle terminal velocity

ϵ = porosity

The terminal velocity of the individual particles was calculated using the method described in the Chemical Engineers Handbook³ which eliminates the iteration procedure required in transition and turbulent drag coefficient calculations.

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Fig. 1. Experimental Equipment.

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Fig. 2. Thickness of Diffuse Region at Top of Fluidized Bed as a Function of Volume Fraction Solid

Fig. 3. Effect of Concentration on Settling Rate of ThO_2 Spheres in 4-inch Diameter Column

Fig. 4. Effect of Volume Fraction Solids on Ratio of Fluidizing Velocity to Settling Rate (Column Vertical)

Fig. 5. Effect of Inclining Fluidization Column on Ratio of Fluidization Velocity to Settling Velocity

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Fig. 6. Effect of Particle Diameter and Concentration
on Power Output of Fluidized Solids Reactor

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Fig. 7. Multi-Region Fluidized Solid Blanket for
Homogeneous Reactor

Fig. 8. Effect of Concentration and Power Level on
Fluidized Bed Particle Diameter

Fig. 9. Effect of Superficial Velocity on Particle Diameter of Fluidized Solid Bed. (Mean Temperature 265°C)

Fig. 10. Some Reported Dense Phase and Dilute Phase Slugging Limits in Terms of Voidage.
(Graph taken from Zenz, F. A., Petroleum Refiner, 36, No 6, 142 (1957))

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