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**SCALE MODELS OF A FLUIDIZED BED COMBUSTER: SIMPLIFICATION
OF THE SCALING LAWS OVER ALL FLOW REGIMES FROM BUBBLING
TO CIRCULATING FLOW AND THE SCALING OF HEAT TRANSFER**

submitted by

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QUARTERLY REPORT - July 1992

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QUARTERLY REPORT - July 1992

Convective heat transfer and hydrodynamic scaling experiments using glass and plastic particles in scaled beds using a simplified set of scaling parameters have been completed. In these tests the particle diameter to bed length scaling parameter was combined with the Reynolds number based on particle diameter and solid density to define a new dimensionless parameter which is essentially the ratio of superficial velocity to minimum fluidization velocity as defined in the Ergun equation. Further theoretical development of the scaling laws has indicated that a set of scaling laws using the ratio of superficial to minimum fluidization velocity in place of the particle Reynolds number and ratio of particle diameter to bed length may be valid over a wider range of conditions than the viscous limit scaling parameters previously defined. A more complete explanation of the proposed simplified scaling parameters, and their regions of validity is included as Attachment A.

Results of the hydrodynamic scaling experiments show good agreement when using this simplified set of scaling laws and strongly suggest that this new set of parameters is acceptable over the range of conditions tested (fast-fluidization regime). Solid fraction profiles using both glass and plastic in the large and small beds are included as Attachment B. Further attempts were made to scale between materials of different density (glass and plastic) using the viscous limit scaling laws. As was the case when attempting to scale between steel and glass, agreement in solid fraction profiles was poor at the lower velocities and improved somewhat at higher velocities (see Attachment C). This suggests that the ratio of solid to fluid density is an important parameter in the scaling of CFB's and must be included in the scaling laws.

Tests aimed at verifying convective heat transfer scaling between beds which are hydrodynamically similar have been completed. The results of these experiments also show reasonable agreement when the convective heat transfer scaling parameter is kept constant in two hydrodynamically similar beds. In the experiments conducted, this amounted to keeping the Prandtl number based on solid specific heat and gas conductivity and viscosity constant (the rest of the parameter is satisfied through hydrodynamic requirements). Plots of Nusselt number based on particle diameter and gas conductivity versus the non-dimensional convective heat transfer parameter at different heights in beds scaled using the simplified set of scaling parameters are included as Attachment D. In these plots, the Nusselt numbers should be the same for hydrodynamically similar cases. Hydrodynamically similar cases between the large and small bed are depicted in the plots through open or solid symbols (i.e. open circles and solid circles represent the low velocity scaled runs, open squares and solid squares represent the intermediate velocity scaled runs, and X's and *'s represent the highest velocity scaled runs).

As a final investigation into the validity of the simplified scaling laws, a 1/16 scale model of the Studsvick CFB been constructed. The Studsvick CFB has a bed diameter of 0.6 meters and a bed height of 7.6 meters. A verification of the full set of scaling laws has already been completed using this bed (Reference 1). Tests to select steel particles which have a minimum fluidization velocity of 1/4 that of the Studsvick bed material (superficial velocity scales as 1/4) have been completed. Preliminary testing of the bed to identify leaks and operating difficulties has begun. Data acquisition on this bed should be completed by mid-July.

The final phase of the project includes completion of the report. It is anticipated that the report will be completed this quarter.

REFERENCES

1. Glicksman, L.R., and Westphalen, D., Work completed on the hydrodynamic scaling of circulating fluidized beds.

ATTACHMENT A

Scaling Relationship Simplifications

Scaling Relationship Simplifications

Scaling Relationships

The scaling relationships are obtained by non-dimensionalizing the equations of motion for the particles and the fluid in a fluidized bed. Using the bed dimension, L , as a typical length dimension, the non-dimensional factors obtained are (Glicksman, 1988).

$$\frac{u_o^2}{gL}, \quad \frac{\rho_f}{\rho_s}, \quad \frac{\beta L}{\rho_s u_o}, \quad \frac{L_1}{L}, \quad (1)$$

Factors omitted include surface forces on particles due to static charge or Van der Waals forces. Also the influence of the particle coefficient of restitution or friction coefficient on inter-particle forces is omitted. Litka and Glicksman (1985) showed that the friction coefficient and coefficient of restitution have a negligible influence on bubbling beds.

In these expression β is the coefficient of the fluid-to-particle drag force per unit volume expressed as $\beta(u-v)$ where u and v are the fluid and particle velocities, respectively. Note that β in general is not a constant, rather it is determined by the relevant expression for the drag and u is the actual fluid velocity, not the superficial velocity.

When the particles are closely spaced approaching a packed bed, the Ergun equation is appropriate for the drag forces,

$$\frac{\Delta p}{L} = \beta|u-v| = 150 \frac{e(1-e)^2}{e^3} \frac{\mu|u-v|}{(\phi d_p)^2} + \frac{1.75e^2(1-e)}{e^3} \frac{\rho_f|u-v|^2}{\phi d_p} \quad (2)$$

Rearranging,

$$\frac{\beta L}{\rho_s u_o} = 150 \frac{(1-e)^2}{e^2} \frac{\mu L}{\rho_s u_o (\phi d_p)^2} + 1.75 \frac{(1-e)}{e} \frac{\rho_f L}{\rho_s} \frac{|u-v|}{\phi d_p u_o} \quad (3)$$

for the most general case if ϕ , $\frac{d_p}{L}$, $\frac{\rho_f}{\rho_s}$ and $\frac{\rho_f u_o d_p}{\mu}$ are matched between the combustor and the model the dimensionless drag coefficient given by equation 3 will remain the same over all conditions. Substituting these dimensionless parameters for $\frac{\beta L}{\rho_s u_o}$ in the group given in equation 1 leads to the most complete or exact set of scaling relationships. If air at standard pressure and temperature is used to fluidize an exact model of a hot combustor, there is only one unique set of bed variables, L , u_o , ρ_s , d_p , which result in a duplication of all the dimensionless parameters for the hot bed. For example, a cold model of an atmospheric combustor has linear dimensions one-quarter those of the combustor.

To obtain more flexibility in the modeling process and to allow a cold model to simulate a hot combustor more than four times larger, simplifications to the full set of scaling laws must be identified. We will investigate this by first exploring the simplifications which hold for several limiting cases.

Low Reynolds Number

To determine the parameters which influence the dimensionless drag, we will first consider the extremes for the Ergun expression. At low particle Reynolds numbers the Ergun expression can be simplified using only the first term in equation 3.

Thus,

$$\frac{\beta L}{\rho_s u_o} = 150 \frac{(1-e)^2}{e^2} \frac{\mu L}{\rho_s u_o (\phi d_p)^2} \quad \text{for} \quad \frac{\rho_f u_o d_p}{\mu} \rightarrow 0 \quad (4)$$

At the same limit the minimum fluidization velocity can be written as,

$$\frac{\Delta p}{L} = (\rho_s - \rho_f) g (1 - e_{mf}) = 150 \frac{(1 - e_{mf})^2}{e_{mf}^3} \mu \frac{u_{mf}}{(\phi d_p)^2} \quad (5)$$

For gas fluidized beds where $\rho_s - \rho_f$ can be replaced by ρ_s ,

$$u_{mf} = \frac{\rho_s g (1 - e_{mf})}{\left[150 \frac{(1 - e_{mf})^2}{e_{mf}^3} \frac{\mu}{(\phi d_p)^2} \right]} \quad (6)$$

Substituting equation 6 into equation 4,

$$\frac{\beta L}{\rho_s u_o} = \frac{\rho_s g (1 - e)^2}{u_{mf} (1 - e_{mf}) e^2} \frac{e_{mf}^3 L}{\rho_s u_o} \quad (7)$$

and

$$\frac{\beta L}{\rho_s u_o} Fr = \frac{u_o}{u_{mf}} \frac{(1 - e)^2 e_{mf}^3}{e^2 (1 - e_{mf})} \quad (8)$$

Thus, in this limit, maintaining $\frac{u_o}{u_{mf}}$, e_{mf} , and Fr identical between two fluidized bed guarantees that $\beta L / \rho_s u_o$ is also identical. Note if the two models display identical dynamic characteristics then e is a dependent variable whose distribution throughout the bed should be identical for both fluidized beds. This limit should cover a wider range than the viscous limit previously defined by the author, since the inertial term in the fluid momentum equation is retained. Only the drag term is simplified.

High Reynolds Numbers

Consider the limit of high particle Reynolds numbers where the inertial term in the Ergun equations dominates.

In this limit,

$$\frac{\beta L}{\rho_s u_o} = 1.75 \frac{(1-e)}{e} \frac{\rho_f L |u'-v'|}{\rho_s \phi d_p} \quad (9)$$

$$\text{where } u' = u/u_o \text{ and } v' = v/u_o \quad (10)$$

and $u'-v'$ is a dimensionless slip velocity.

The minimum fluidization velocity can be expressed as,

$$\frac{\Delta p}{L} = (\rho_s - \rho_f) g (1-e_{mf}) = \frac{1.75 (1-e_{mf})}{e_{mf}^3} \rho_f \frac{u_{mf}^2}{\phi d_p} \quad (11)$$

rearranging and using ρ_s in place of $\rho_s - \rho_f$,

$$\frac{1.75 (1-e_{mf})}{e_{mf}^3} \frac{\rho_f}{\rho_s \phi d_p} = \frac{g (1-e_{mf})}{u_{mf}^2} \quad (12)$$

Substituting this into equation 9 and multiplying by Fr,

$$Fr \frac{\beta L}{\rho_s u_o} = \frac{u_o^2}{gL} \frac{e_{mf}^3 L |u'-v'| g (1-e)}{e u_{mf}^2} = \frac{u_o^2}{u_{mf}^2} \frac{e_{mf}^3 (1-e) |u'-v'|}{e} \quad (13)$$

At large particle Reynolds numbers, just as at low Reynolds numbers, the dimensionless drag, $\beta L / \rho_s u_o$, is identical when u_o / u_{mf} and Fr are identical. e , u' and v' are dependent dimensionless variables which are identical for two similar fluidized beds.

Low Slip Velocity

Finally, consider the case when the magnitude of the slip velocity between the particles and the gas is close to u_{mf}/e everywhere in the fluidized bed. With the vertical pressure drop equal to the particle weight the following holds for any value of the particle Reynolds number,

$$\frac{\Delta p}{L} = \beta |u-v| = \beta \frac{u_{mf}}{e} = \rho_s g (1-e) \quad (14)$$

$$\beta = \rho_s g (1-e) \frac{e}{u_{mf}} \quad (15)$$

and

$$\frac{\beta L}{\rho_s u_o} Fr = u_o/u_{mf}(1-e)e \text{ for } u'-v' = \frac{1}{e} \frac{u_{mf}}{u_o} \quad (16)$$

Again, when u_o/u_{mf} and Fr are identical for two beds and the slip velocity is close to u_{mf}/e the dimensionless drag coefficient is also identical for two beds.

For all the three limiting cases identified above, similitude can be obtained by maintaining constant values for the dimensionless parameters,

$$\frac{u_o^2}{gL}, \frac{\rho_s}{\rho_f}, \frac{u_o}{u_{mf}}, e_{mf} \quad (17)$$

provided that the forces between the particles and gas can be represented by the Ergun equation or an equivalent expression.

General Case

With the limiting cases bounding high and low Reynolds number limits, it would be reasonable to expect that the same dimensionless parameters apply over the entire range of conditions for the Ergun equation, i.e. at intermediate values of the particle Reynolds number and for slip velocities which span a wide range of values for which β can be represented by the Ergun equation. The validity of the simplified parameters can be checked for the intermediate range of values.

From equation 3,

$$\frac{\beta L}{\rho_s u_o} = \frac{150(1-e)^2}{e^2} \frac{\mu L}{\rho_s u_o (\phi d_p)^2} + \frac{1.75(1-e)}{e} \frac{\rho_f L}{\rho_s} \frac{|u'-v'|}{\phi d_p} \quad (18)$$

Using the Ergun equation to determine u_{mf} ,

$$\frac{\Delta p}{L} = \rho_s g(1-e_{mf}) = 150 \frac{(1-e_{mf})^2}{e_{mf}^3} \frac{\mu u_{mf}}{(\phi d_p)^2} + \frac{1.75(1-e_{mf})}{e_{mf}^3} \frac{\rho_f u_{mf}^2}{\phi d_p} \quad (19)$$

Dividing equation 18 by 19, and rearranging,

$$\left(\frac{\beta L}{\rho_s u_o} \right) \frac{Fr(1-e_{mf})e^2}{\left(\frac{u_o}{u_{mf}} \right) (1-e)^2 e^3} = \left[\frac{1 + \frac{1.75}{150} \frac{\phi_s Re}{(1-e)} e |u'-v'|}{1 + \frac{1.75}{150} \frac{\phi_s Re}{(1-e_{mf})} \frac{u_{mf}}{u_o}} \right] \quad (20)$$

where $Re = \rho_f u_o d_p / \mu$.

It is easy to verify by use of equation 20 that in the three limits defined previously, low Re , high Re , and $u'-v'$ approaching $\frac{1}{e} \frac{u_{mf}}{u_o}$, $\frac{\beta L}{\rho_s u_o}$ remains constant if Fr , e_{mf} , and u_o/u_{mf} are held constant.

For the more general case, figure 1 shows the value of β given by equation 20 relative to β at low Re over a range of conditions when u_o/u_{mf} is 10 and 3, respectively. When u_o/u_{mf} or the slip velocity is high there is a larger variation of dimensionless drag coefficient with Reynolds number. Note that β does not vary with particle Reynolds number when the Reynolds number remains above about 10^3 or below about 10. Figure 2 illustrates the results when u_o/u_{mf} is 1000, a condition approached with very fine particle bubbling beds or circulating beds. In the latter instance the use of the Ergun relationship is questionable except for the dense lower part of the bed. In this figure the slip velocity is $1/50 u_o$ or less, which corresponds to $20 u_{mf}$. This limit on the slip velocity is roughly twice the terminal velocity of large particles.

Of particular concern is the error in the dimensionless drag coefficient when the scale model is designed using the simplified set of scaling rules, equation 17. In contrast, a scale model using the more complete or exact set of dimensionless parameters would include the list of parameters in equation 17 as well as the Reynolds number based on the mean particle diameter.

The simplified scaling parameters allows smaller models to be constructed to simulate a given sized combustor. As the length scale is reduced the superficial velocity must also be reduced to maintain a constant Froude number. The particle diameter must also be reduced to keep u_o/u_{mf} constant. Tables 1 and 2 present calculated results when a given exact condition is scaled with the simplified set of scaling laws for two cases: beds with linear dimensions $1/4$ and $1/16$ of the dimensions of the exact model. Table 1 gives the calculated results for a pressurized combustor at 10 atmospheres and 800°C , while table 2 gives results at atmospheric pressure and 800°C . All of the results are given for models at combustor conditions. The ratios of diameters,

velocities etc. between the exact, correctly scaled, bed and the smaller models also holds when dealing with cold models.¹ In this case the exact, or correctly scaled bed, is the cold bed constructed using the full set of hot to cold scaling parameters.

For an exact model of an atmospheric combustor fluidized by air at standard pressure and temperature, the model dimensions and the particle diameter are one quarter those of the atmospheric combustor and the velocity is one half those of the combustor. The one quarter and one sixteenth scale simplified model would be one sixteenth and one sixty fourth the linear dimensions of the atmospheric combustor. An exact scale model of a PFBC, fluidized at 10 atmospheres, would have linear dimension approximately equal to the pressurized combustor when the model is fluidized by air at one atmosphere. The one quarter and one sixteenth scale simplified model would be one quarter and one sixteenth the linear dimensions of the PFBC, see figure 3.

Note that in tables 1 and 2 all of the Reynolds number based on particle diameter of the one quarter length scale and the one sixteenth length scale bed are a factor of approximately 0.34 and 0.12, respectively, of the Reynolds number of the exact bed. Using these Reynolds number scale factors, the errors in the dimensionless drag coefficient $\frac{\beta L}{\rho u_0}$ using the simplified scaling

models, can be found from figures 1 and 2. These errors are shown on figures 4 and 5 for u/u_m of 10 and 1000, respectively, plotted as a function of Re in the exact scaled bed. For a particle Reynolds number of 1000 or less, which corresponds to pressurized beds with particles of 1mm or less, the error in the drag coefficient with the simplified scaling laws is twenty percent or less for a one quarter length scale. The error is forty percent or less for a one sixteenth length scale. For particles of 0.2mm or less, corresponding to a Reynolds number of 100 or less, the errors in drag coefficient are minimal. When the Ergun equation applies for the drag coefficient, a one quarter scale model based on the simplified scaling laws should be valid for any conditions, while a sixteenth linear scale model should apply for particles with diameters of about 0.2 mm or less provided that particle to particle surface forces do not come into play.

Clusters

In the freeboard of a bubbling bed or in the upper portion of

¹. This can be seen by using the full set of scaling laws to construct cold models of the exact, the 1/4 and the 1/16 scale simplified hot models, respectively.

a circulating bed where particles generally are considered to act in clusters or groups, a similar examination of scaling of the gas to solid drag can be made. Consider all of the particles grouped into clusters with an effective diameter d_c and the clusters occupying a volume fraction ϵ_c of the bed volume. The cluster to gas drag will be represented by a drag coefficient defined by reference to the drag coefficient for a solid sphere of diameter d_c .

β can be written as

$$\beta |\bar{u} - \bar{v}| \frac{\pi d_c^3}{6(1-\epsilon_c)} = \frac{1}{2} \rho_f \frac{\pi d_c^2}{4} |\bar{u} - \bar{v}|^2 C_D \quad (21)$$

where ϵ_c is the volume fraction of clusters

$$\beta = \frac{3}{4} \frac{\rho_f |\bar{u} - \bar{v}| C_D}{d_c} (1 - \epsilon_c) \quad (22)$$

and

$$\frac{\beta L}{\rho_s u_o} = \frac{3}{4} \left(\frac{\rho_f}{\rho_s} \right) |u' - v'| C_D \frac{L}{d_c} (1 - \epsilon_c) \quad (23)$$

If the diameter of individual particles does not influence the cluster drag, then the dimensionless drag, $\frac{\beta L}{\rho_s u_o}$ in equation 23 is a function of C_D when the solid to gas density ratio is held constant between the combustor and the model. We will use the drag coefficient of solid spheres as a first approximation of the relative change of C_D with Reynolds number based on cluster diameter. The drag coefficient of a solid sphere can be closely represented by the empirical expression given by White,

$$C_D = \frac{24}{Re} + \frac{6}{1 + \sqrt{Re}} + 0.4 \quad (24)$$

If the reduced scale models faithfully reproduce the dynamics

of the exact case, then the cluster dimensions should scale directly with the linear dimensions of the bed. Thus a one quarter linear scale model which has a velocity one half that of the exact case will have a cluster Reynolds number one eighth that of the exact bed. From the relationship of C_D with Re we can determine the change of C_D with model scale at a given Reynolds number of the exact bed. Figure 6 shows the shift in C_D for length scale of 1/4, 1/8 and 1/16, respectively, of the exact bed length as a function of the cluster Reynolds number of the exact bed. Also shown on the figure is the typical Reynolds number of an atmospheric combustor with a 0.3 m cluster diameter. For pressurized beds, the increase in Re will be proportional to the pressure level. It is reasonable to take the cluster diameter as the horizontal tube spacing in a bubbling bed with tube banks. In a bubbling bed without tubes, the cluster size, proportional to the bubble size can be much larger. In an open bed the cluster diameter is more difficult to determine, it is reasonable to assume its diameter is proportional to the bed diameter or one order of magnitude smaller. From these considerations, the Reynolds number based on the cluster diameter should be 10^4 or larger in an atmospheric combustor and 10^5 or larger in a pressurized combustor. From figure 6 it can be seen that a one quarter scale or an eight scale model should have drag coefficients similar to the exact bed.

Terminal Velocity of Particles

If the drag coefficient is influenced by the characteristics of individual particles, the detailed particle dynamics of the simplified scaled models must be examined. One criterion is to compare the terminal velocity of particles to u_0 , the superficial gas velocity, when u_0/u_{mf} is kept constant in the scaling process.

The Ergun equation, eqn. 19, can be solved to find u_{mf} ,

$$\rho_f \frac{u_{mf} d_p}{\mu} = \frac{\frac{-(150)(1-e_{mf})^2}{e_{mf}^3} + \sqrt{\left[\frac{150(1-e_{mf})^2}{e_{mf}^3} \right]^2 + \frac{7(1-e_{mf})^2}{e_{mf}^3} \phi^3 Ar}}{3.5(1-e_{mf})\phi} \quad (25)$$

where $Ar = \frac{\rho_s \rho_f d_p^3 g}{\mu^2}$ is the Archimedes number.

The terminal velocity for a particle can be obtained as,

$$\rho_s \frac{\pi d_p^3 g}{6} = \frac{1}{2} \rho_f \frac{\pi d_p^2}{4} C_D u_T^2 \quad (26)$$

Rearranging, and using equation 24 for C_D ,

$$Re_T^2 = \frac{\frac{4}{3} Ar}{\frac{24}{Re_T} + \frac{6}{1 + \sqrt{Re_T}} + 0.4} \quad (27)$$

where $Re_T = \frac{\rho_f u_T d_p}{\mu}$.

Thus, u_{mf} and u_T and their ratio should be a function of the Archimedes number for a given value of ϕ and ϵ_{mf} . Figure 7 illustrates the dependence of u_T/u_{mf} on Ar . Also shown are the values of the Archimedes number for a pressurized combustor at two particle diameters. Associated with the exact model are the corresponding Archimedes number for two models scaled using the simplified relationships at linear dimensions one quarter and one sixteenth the exact dimensions. Results for several different particle diameters are presented in tables 1 and 2 for pressurized beds and atmospheric beds. The errors in u_T/u_{mf} are shown on figures 8 and 9 scaling at two different linear dimensions. Scaling a combustor with comparatively small particles, 0.2 mm or less, gives good agreement for u_T/u_{mf} even at one sixteenth linear scale, while for large particles a linear scale of one fourth gives fair agreement of u_T/u_{mf} .

Conclusions

The full set of dimensionless parameters for scaling can be reduced to the simplified set, $\frac{u_o^2}{gL}$, $\frac{\rho_s}{\rho_f}$, $\frac{u_o}{u_{mf}}$, and geometric similarity in the limits of small and large particle Reynolds number as well as when the slip velocity approaches u_{mf} .

In the more general range of conditions, the simplified form of the scaling relationships give acceptable agreement for the drag coefficient and the ratio of terminal velocity to minimum fluidization velocity when the simplified model has reduced linear dimensions between one quarter and unity. For combustors with

particle diameter of 0.2 mm or less, a simplified scale model with reduced linear dimensions up to one sixteenth the exact model should give acceptable agreement.

If the particle size of the simplified model becomes very small, care must be taken to insure that interparticle forces do not become important. A dimensionless form of Geldart's particle classification will aid in determining the approximate range for the onset of the interparticle forces.

REFERENCES

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2. Litka, T. and L.R. Glicksman, "The Influence of Particle Mechanical Properties on Bubble Characteristics and Solid Mixing in Fluidized Beds," Powder Technology, 42, pp.159-167. 1985.

Table 1 Scaling Pressurized Beds p = 10 atm, T = 800°C								
	d_p mm	u_{mf} m/s	$\frac{u_T}{u_{mf}}$	$\frac{u_T/u_{mf}}{(u_T/u_{mf})_{exact}}$	$Ar = \frac{gd_p^3 \rho_p}{\mu^2}$	$\frac{L}{L_{exact}}$	$\frac{u}{u_{exact}}$	$\frac{Re(d_p)}{Re_{exact}}$
exact	1	.39	8.82	1	4.4×10^4	1	1	1
	.61	.193	11.84	1.34	9.93×10^3	1/4	1/2	.31
	.41	.097	16.4	1.86	3.30×10^3	1/16	1/4	.103
exact	0.7	.24	10.7	1	1.5×10^4	1	1	1
	.47	.124	14.6	1.36	4.5×10^3	1/4	1/2	.34
	.33	.064	19.7	1.84	1.57×10^3	1/16	1/4	.12
exact	0.2	.024	29.1	1	350	1	1	1
	.14	.012	36.2	1.24	120	1/4	1/2	.35
	.10	.006	42.2	1.45	43.8	1/16	1/4	.125
exact	.1	.006	42.2	1	43.8	1	1	1
	.07	.003	46.9	1.11	15	1/4	1/2	.35
	.049	.0015	49.7	1.18	5.15	1/16	1/4	.123

The two rows following the exact case present the particle diameter, velocity, etc. for a linear dimension decrease of 1/4 and 1/16, respectively, maintaining constant Froude number, u_o^2/gL and constant u_o/u_{mf} .

$$\rho_p = 2500 \text{ kg/m}^3, \phi = 0.8, e_{mf} = 0.5, Re = \frac{\rho_f u_o d_p}{\mu}$$

Table 2
Scaling Atmospheric Beds
p = 1 atm, T = 800°C

	d_p mm	u_{mf} m/s	$\frac{u_T}{u_{mf}}$	$\frac{u_T/u_{mf}}{(u_T/u_{mf})_{exact}}$	$AR = \frac{gd_p^3 \rho_p}{\mu^2}$	$\frac{L}{L_{exact}}$	$\frac{u}{u_{exact}}$	$\frac{Re(d_p)}{Re_{exact}}$
exact	2	1.65	9.12	1	3.5×10^4	1	1	1
	1.25	.83	12.29	1.34	8.6×10^3	1/4	1/2	.313
	.84	.41	17.04	1.87	2.6×10^3	1/16	1/4	.105
exact	1	.563	14.7	1	4.4×10^3	1	1	1
	.685	.778	20.2	1.37	1.4×10^3	1/4	1/2	.34
	.48	.14	26.8	1.82	484	1/16	1/4	.12
exact	.5	.15	26.1	1	547	1	1	1
	.353	.075	33.1	1.27	192	1/4	1/2	.35
	.25	.038	39.7	1.52	68	1/16	1/4	.125
exact	.2	.024	43.3	1	35	1	1	1
	.14	.012	47.6	1.1	12	1/4	1/2	.35
	.10	.0061	49.9	1.15	4.4	1/16	1/4	.125
exact	.1	.0061	49.9	1	4.4	1	1	1
	.07	.003	51.2	1.03	1.5	1/4	1/2	.35
	.05	.0015	51.7	1.04	.55	1/16	1/4	.125

The two rows following the exact case presents the particle diameter, velocity, etc. for a linear dimension decrease of 1/4 and 1/16, respectively, maintaining constant Froude number, u_o^2/gL and constant u_o/u_{mf} .

$$\rho_p = 2500 \text{ kg/m}^3, \phi = 0.8, e_{mf} = 0.5$$

Drag Coef using Ergun Eqn.

Fig. 1 Ratio of Drag Coefficient to Low Reynolds Number Drag Coefficient, Ergun Equation, u_0/u_{mf} of 10 and 3.

Drag Coef using Ergun Eqn.

$U_0/U_{mf} = 1000$

Fig. 2 Ratio of Drag Coefficient to Low Reynolds Number Drag Coefficient, Ergun Equation, u_0/u_m of 1000.

	P	T(C)	L	U
PFBC	10 atm	800	Lo	Uo
Exact Model	1 atm	20	Lo	Uo
Simplified Model	1 atm	20	Lo/4	Uo/2
Simplified Model	1 atm	20	Lo/16	Uo/4

Fig. 3 Exact and Simplified Models of a Pressurized Combustor.

Error in Drag Coefficient

$$U_0/U_{mf} = 10, U_{slip}/U_0 = 0.3$$

Fig. 4 Error in Drag Coefficient, Ergun Equation, $u_0/u_m = 10$, $u_{slip}/u_0 = 0.3$.

Error in Drag Coefficient

$$U_0/U_{mf} = 1000, U_{slip}/U_0 = 1/50$$

Fig. 5 Error in Drag Coefficient, Ergun Equation, $u/u_{mf} = 1000$, $u_{slip}/u_0 = 1/50$.

Error in Cd for fixed U_0/U_{mf} Cd for Solid Sphere

Fig. 6 Error in Cluster Drag Coefficient for Fixed u_0/u_{mf} , Using C_d for a Solid Sphere.

Fig. 7 Ratio of Terminal Velocity to Minimum Fluidization Velocity vs. Archimedes Number, Examples of Modeling 0.7 mm and 0.2 mm with Simplified Scaling at 10 atm. and 800 °C.

Error in U_t for fixed U_0/U_{mf} $p = 10 \text{ atm}$, $T = 800^\circ\text{C}$

Fig. 8 Error in Terminal Velocity Using Simplified Scaling at 10 atm. and 800 °C.

Error in U_t for fixed U_0/U_{mf} $p = 1 \text{ atm}, T = 800^\circ\text{C}$

Fig. 9 Error in Terminal Velocity Using Simplified Scaling at
1 atm. and 800°C .

ATTACHMENT B

**Solid Fraction Profiles - Small to Large Bed
Glass/Plastic**

Six Figures

Solid Fraction Profiles - Small to Large Bed Glass

Solid Fraction Profiles - Small to Large Bed Glass

Solid Fraction Profiles - Small to Large Bed Glass

Solid Fraction Profiles - Small to Large Bed Plastic

Solid Fraction Profiles - Small to Large Bed Plastic

Solid Fraction Profiles - Small to Large Bed Plastic

ATTACHMENT C

Small Fraction Profiles for Glass and Plastic

Three Figures

Solid Fraction Profiles for Glass and Plastic

Solid Fraction Profiles for Glass and Plastic

Solid Fraction Profiles for Glass and Plastic

ATTACHMENT D

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter

Seven Figures

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter (Bed Height = 41.4%)

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter (Bed Height = 87.5%)

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter (Bed Height = 41.4%)

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter (Bed Height = 64.4%)

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter (Bed Height = 87.5%)

Nusselt Number vs. Non-Dimensional Heat Transfer Parameter

$$[100(1-\epsilon)](U_{op}c_p d_p^2)/(Lk_g)$$

eqn: $ax^{0.5}$, error²:2.59,
a=+2.661E-001

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