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CIRCULATING FLUIDIZATION RESEARCH AT CORNELL: A STUDY OF HYDRODYNAMIC SCALE-UP EFFECTS

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ABSTRACT

Circulating fluidization is a promising technology for designing efficient gas-solid reactors with high material throughputs. Unfortunately, limited hydrodynamic understanding renders scale-up extrapolations both empirical and risky.

At Cornell, a versatile facility has been designed to allow direct and systematic characterization of scale-up effects in vertical risers. Systematic quantification of scale-up may be achieved in a single cold flow riser using fluidization gas mixtures of adjustable density and viscosity. Hydrodynamic analogy between the cold bed and a hot reactor is accomplished by matching all relevant dimensionless parameters. Several choices of gas mixture composition and solid particle properties make the cold flow analogous to that in ideal combustors of several possible sizes.

INTRODUCTION

Circulating fluidization emerged in the 1970's as an attractive means of achieving large gas and solids flow rates in relatively compact vessels, without compromising adequate contacting or reactor efficiency. Initial research on circulating beds (L. Reh in West Germany (1), J. Yerushalmi, et al. in the U.S. (2), L. Strömberg in Sweden (3), L. Youchou and M. Kwauk in China (4)), was quickly followed by major industrial developments (Lurgi, Studsvik, Battelle, Ahlström, etc.).

Invariably, each new industrial design stumbles upon the question of scale-up. Extrapolation of pilot plant data to conditions in full-scale facilities relies mostly on empirical experience. Skepticism often accompanies results obtained in laboratory-scale beds. It is assumed implicitly that a pilot plant must be big enough to reflect conditions in a large industrial facility. Yet it is unclear how big a pilot plant should be to make a larger facility immune to unexpected scale-up effects. Experience shows that fluidized bed scale-up constitutes a large financial risk in a new project, yet it remains more of an art than a science. Clearly, a better understanding of scale-up should provide useful design guidelines for circulating beds.

A cold flow circulating bed is presently under construction at Cornell University. It is designed for direct and systematic quantification of scale-up effects under realistic hydrodynamic conditions. To this end, it operates with controlled gas mixtures which make it analogous to circulating combustors of several sizes. In the next section of this paper, dimensional analysis leads to the concept of binary fluidization mixtures for systematic scale-up experiments in risers; an order-of-magnitude analysis also shows that unambiguous scale-up effects should be observed in our projected experiments. Finally, our circulating bed is described, and a projected research program is outlined.

A COLD FLOW TECHNIQUE FOR SYSTEMATIC INVESTIGATION OF SCALE-UP EFFECTS

The gathering and interpretation of hydrodynamic data from actual circulating bed combustors is particularly difficult. By contrast, cold flow test facilities are convenient for detailed hydrodynamic measurements. However, the use of air as a fluidizing agent raises serious questions about the validity of cold flow data, since gas density and viscosity in the cold bed depart significantly from typical combustor conditions. Moreover, operating conditions achieved in cold research beds have seldom matched typical conditions in industrial combustors. For example, Yerushalmi, et al. (5) have produced solid fluxes in the range $20<G<250\ kg/m^2.sec$, a set of values much larger than those at the Duisburg powerplant (6): $1 < G < 10\ kg/m^2.sec$. Hence, most test facilities have generated data for which interpretation remains perplexing.

Fortunately, from the standpoint of fluid mechanics, dimensional analysis can provide useful scaling laws to reconcile cold test results with the hydrodynamics of "ideal" combustors. In this section, it is shown that the proper choice of gas and solids can make a cold bed analogous to ideal combustors of arbitrary dimensions, thus allowing the systematic study of hydrodynamic scale-up effects in a single cold flow facility.

Glicksman showed that the hydrodynamics of a bubbling bed can be characterized by means of three dimensionless independent parameters (7),

$$Ar = \frac{\rho_s \rho\, d^3 g}{\mu^2} \quad \text{(Archimedes number)},$$

$$Fr = u^2/gd \quad \text{("Froude number")},$$

$$R = \rho/\rho_s \quad \text{(density ratio)}.$$

Table 1 - Summary of scale-up experiments

Gases				Solid			Bed Operations		Analogous combustor diameter
He %	CO_2 %	ρ_1 (kg/m^3)	$\mu_1 \times 10^5$ (kg/m/sec)	type	ρ_{s_1} (kg/m^3)	d_1 (μm)	u_1 (m/sec)	G_1 (kg/m^2/sec)	D_o(m)
99	1	0.18	1.9	silica gel	790	225	1.9→5.7	0.5→5.0	0.23
75	25	0.58	1.8	glass	2550	98	1.3→3.8	1.1→11	0.52
5	95	1.7	1.5	nickel alloy	7700	41	0.8→2.4	2.1→21	1.2

These groups arise from Jackson's governing equations of the separated flow model (8). In this model, Jackson neglected the effect of particle collisions and argued that, even for dense beds, particle drag is more significant than particle-particle interactions. This assumption was designed to simplify considerably the momentum equations for the solid phase. We expect this hypothesis to hold a fortiori for the more dilute mixtures present in circulating beds. In any case, Glicksman's analysis may be derived without writing governing equations (i.e., using Buckingham's Π-theorem). In the absence of other interaction effects (e.g., electrostatic forces), it is therefore independent of any simplifying assumptions leading to the establishment of Jackson's governing equations.

In Glicksman's analysis, velocity, distance and density are made dimensionless using superficial gas velocity u, average particle diameter d, and solid density ρ_s, respectively. Additional dimensionless ratios arise from boundary conditions, namely the bed aspect ratios (e.g., height/diameter) and the ratio of bed diameter to average particle size (L≡D/d). Note that Glicksman's analysis can be extended to circulating beds, provided that a dimensionless solid flux F is added to the list of boundary conditions,

$$F = G/\rho_s u ,$$

where G is the total solid flux in the riser. Archimedes number and density ratio relate to the physical properties of gas and solids. On the other hand, Froude number and dimensionless flux F provide a measure of operating conditions in the bed.

Hydrodynamic similitude between an ideal fluid bed combustor (subscript 0) and a cold bench-scale model (subscript 1) thus requires the following relations among operating conditions:

superficial gas velocity $$u_1/\nu_1^{1/3} = u_o/\nu_o^{1/3} , \quad (1)$$

particle size $$d_1/\nu_1^{2/3} = d_o/\nu_o^{2/3}, \quad (2)$$

particle density $$\rho_{s1}/\rho_1 = \rho_{so}/\rho_o , \quad (3)$$

solid flux $$G_1/\rho_1^{2/3} \cdot \mu_1^{1/3} = G_o/\rho_o^{2/3} \cdot \mu_o^{1/3}, \quad (4)$$

characteristic bed dimension $$D_1/\nu_1^{2/3} = D_o/\nu_o^{2/3} . \quad (5)$$

In this analysis, note that several simplifying assumptions allow gas-solid flows to be treated independently from more intricate combustion effects. Here, cold flow tests are analogous to "ideal combustors" of uniform composition, constant properties and constant temperature. Under these assumptions, combustor characteristics are solely determined by dimensionless hydrodynamic parameters (Ar, Fr, R and F); other parameters related to mixing (Péclet number) or chemical source terms (Damköhler ratios) do not appear in the analysis. The violent mixing characteristics of circulating beds can reasonably justify assumptions of constant properties, so that hydrodynamic effects may be studied independently without jeopardizing the validity of observed scale-up trends.

In a recent study, Nicastro and Glicksman applied the analogy based on eqs. 1, 2, 3 and 5 to simulate bubbling bed combustors (9). To this end, a smaller bench-scale facility was fluidized with air at room temperature (i.e., $\rho_1 \sim 1.2$ kg/m^3, $\mu_1 \sim 1.8.10^{-5}$ kg/m.sec). In agreement with eq. 5, the diameter of the cold bed was about a quarter of the diameter of the combustor being simulated ($D_1/D_0 = (\nu_1/\nu_o)^{2/3} \sim 0.25$). It is encouraging that Nicastro and Glicksman verified the above scaling relationships by comparing the behavior of the hot and cold beds, despite all the simplifying assumptions mentioned above.

Despite the potential rewards resulting from a better understanding of scale-up, no dedicated study has thus far been conducted for circulating beds (10). To this end, Glicksman's experimental method may be improved by replacing air with a mixture of two gases having widely different kinematic viscosities. In this way, the composition of an individual mixture can be adjusted to reach any desired kinematic viscosity in a wide range of possible values. (The absolute viscosity can be inferred with reasonable accuracy from the composition of the mixture using Wilke's semi-empirical formula (11)). By virtue of eq. 5, such binary mixtures can simulate combustion gases in vessels of several possible dimensions. A single test facility is then sufficient to investigate scale-up effects.

Several candidate gases can be used. The broadest span of potential kinematic viscosities is obtained by mixing a heavy gas such as C_3H_8 or CO_2 with a lighter gas such as H_2 or He. Propane or hydrogen raise obvious safety questions. Thus, mixtures of carbon dioxide and helium should be selected. For example, specific mixtures with kinematic viscosities in the range $11.10^{-5} > \nu_1 > 0.9.10^{-5}$ kg/m.sec can be engineered to simulate combustors of diameters larger than our 20 cm test bed ($0.2<D_0<1.2$ m, see Table 1).

A typical industrial combustor may burn particles with a mean diameter d≈250 μm and a density $\rho_s \approx 1500$ kg/m^3 (coal + limestone mixture). In addition, it usually operates at $T_o \approx 1070$°K (i.e., $\rho_o \approx 0.3$ kg/m^3, $\mu_o \approx 4.10^{-5}$ kg/m/sec) with $u_o \approx 2 \rightarrow 6$ m/sec, and $G_o \approx 1 \rightarrow 10$ kg/m^2.sec. Such typical combustor conditions form the basis for the projected scale-up experiments in Table 1.

For complete similitude between reference combustor and laboratory bed, the average size and density of the test particles are specified by eqs. 2 and 3. For the experiments of Table 1, particle mean size and density must therefore lie in the range, $40 \leq d \leq 225$

μm and $790 \leq \rho_s \leq 7700$ kg/m³. In addition, both solid flux and superficial gas velocity are scaled according to eqs. 1 and 4. Finally, note that careful scale-up test require classified powders with similar size distributions relative to the mean. Further, it is desirable that only powders with good sphericity be selected (e.g., atomized metals).

In the next few paragraphs, an order-of-magnitude analysis shows that our projected range of analogous vessel diameters can result in substantially different values of parameters governing gas-solid turbulence. Hence, we anticipate to observe unambiguous scale-up effects in our experiments.

Detailed predictions of the hydrodynamics of dense suspensions are difficult, mainly because of complexities in particle-particle and particle-wall interactions, and the effect of solids on the spectrum of turbulence (12,13). It is therefore crucial to rely on experimental findings for predicting scale-up effects or other hydrodynamic behavior. For the purpose of the present discussion, a simplified description of the flow may suffice to establish whether vessel scale-up bears significant effects on the hydrodynamics. To this end, the influence of particles on turbulence can be ignored. Further, we assume that the suspension is dilute. Finally, our attention is focused on the core of the riser, and details of particle-wall interaction are omitted.

The hydrodynamics of dilute gas-solid suspensions are characterized by a particle-fluid interaction parameter (Soo, (14)),

$$\kappa = \frac{2\langle u_g^2\rangle^{1/2}\tau}{\lambda_l} , \tag{6}$$

where $\langle u_g^2\rangle^{1/2}$ is the rms gas velocity fluctuation, λ_l is the Lagrangian microscale of gas turbulence and τ is the relaxation time for gas→particle momentum transfer (14). Large values of κ correspond to particles lagging behind the gas velocities fluctuations (e.g., heavy particles, or intense gas fluctuations); small values correspond to particles closely following the gas.

For a turbulent flow unaffected by the presence of particles, κ can be estimated in the core of the vessel using correlations for fully-developed turbulent flow in a pipe. In this case, the scale-up trends result mainly from the dependence of turbulence length scales on vessel dimensions. Correlations for pipe flow found in Hinze (15) lead to orders of magnitude for λ_l and $\langle u_g^2\rangle^{1/2}$. For moderate particle Reynolds numbers, τ is obtained using a correction to Stokes' law of drag (14). An order of magnitude for κ can thus be found in the core of the vessel,

$$\kappa \simeq 0.01 \frac{u^{5/4} d^2 \rho_s}{D^{3/4} \nu^{1/4} \mu} = 0.01 \frac{(Fr\ Ar)^{5/8}}{L^{3/4}\ R^{3/8}} . \tag{7}$$

Remarkably, κ is dependent on the dimensionless riser diameter $L = D/d$. According to eq. 7, we therefore expect κ to decrease from 2.7 to 0.7 as our analogous riser is scaled-up from 0.2 to 1.2 m. Several characteristic turbulent flow variables depend upon κ (14,16). Expressed in dimensionless form, they include the rms slip velocity $\langle \Delta V^2\rangle^{1/2}/u$, and the Lagrangian scale of particle-gas interaction L_p/D. As κ decreases from 2.7 to 0.7, $\langle \Delta V^2\rangle^{1/2}/u$ and L_p/D decrease by factors of 2 and 10, respectively. In addition, the ratio of solid to gas diffusivities δ_s/δ_g increases from 0.6 to 0.9. In short, particles are more readily entrained by the gas in larger vessels. Overall, the increased turbulence length scales of larger vessels make solids behave like smaller (or lighter) particles.

Finally, an approximate analysis of particle residence time suggests that average solid hold-up is of the order

$$(1-\bar{\varepsilon}) \sim \frac{G}{\rho_s} \times \frac{1}{u/\bar{\varepsilon} - u_t - cst\ .\ \langle \Delta V^2\rangle^{1/2}} . \tag{8}$$

This eq. indicates that particles lag behind the gas (average velocity $u/\bar{\varepsilon}$) as a result of free-fall (velocity u_t), and the delay caused by turbulent solid recirculation (the term cst . $\langle \Delta V^2\rangle^{1/2}$). The limited information on circulating bed scale-up (Avidan (17)) would confirm the trend suggested by eq. 8. Such trends clearly need experimental quantification. Our cold flow technique can simulate relative combustor diameters in the range $0.2 < D_0 < 1.2$ m. Corresponding values of κ are found in a region where $\langle \Delta V^2\rangle^{1/2}/u$ and L_p/D are most sensitive to changes in κ. Therefore, we can reasonably expect that our cold flow technique will produce unambiguous and observable scale-up effects.

In conclusion, controlled fluidization mixtures of helium and carbon dioxide constitute a realistic environment for circulating bed experiments, and allow direct study of combustor scale-up using a single cold flow facility.

CIRCULATING FLUIDIZATION RESEARCH AT CORNELL

An aluminum test facility is under construction in the Department of Mechanical Engineering at Cornell University. Its main constitutive vessels are sketched in Fig. 1. Aluminum has been selected rather than plexiglas to avoid electrostatic effects in the riser. (Transparency of the riser walls is superfluous for opaque mixtures). The riser (inside diameter: 20cm, total height: 7 m) is composed of 5 tubular sections assembled with flanges, and a 90° disengagement elbow. One or two sections can be removed from this modular design to lower the height over diameter ratio H/D from a maximum of 35 down to 20, thus allowing direct study of vertical scale-up effects. A primary cyclone (minimum estimated efficiency 96%) and a smaller secondary cyclone (efficiency >99%) are connected to the outlet of the riser. Particles are collected in a tubular downcomer composed of three aluminum sections (1 m high), and a shorter section for solid recycle to the riser. Two sections of the downcomer can be removed to match the desired riser height. A sintered butterfly valve is inserted to measure solid flow rates. Incoming solids can be fluidized on top of the closed valve to infer steady-state solid flow rates from variations of the static pressure column of incoming solids. Solid recycle flow is controlled using a gate valve inserted in the tube connecting downcomer and riser. Particles escaping from the secondary cyclone are removed using fabric filters in a closed baghouse. Outlet gas mixtures (He and CO_2) are recycled to a closed atmospheric pressure tank which supplies the facility. A thermal conductivity detector monitors helium and carbon dioxide contents in the tank. Changes in the composition of fluidization gases can thus be corrected. A positive displacement blower (Holmes model HR/21) compresses fluidization gas mixtures to 0.14 bar at 0.4 m³/sec (3260 RPM, 11 HP) and to 0.95 bar at 0.20 m³/sec (2200 RPM, 40 HP). The blower is driven by a 40 HP General Electric DC motor. Blower operations are matched to desired fluidization conditions by adjusting motor angular velocity. Fine tuning is achieved by recycling part of the blower outlet to the tank. As a result, any gas/solid mixture in Table 1 can be fluidized in the riser up to an average solid loading of 20% and/or to a superficial gas velocity well within the pneumatic transport regime. A vortex flow meter is utilized to monitor absolute gas volume flow rates (i.e., to obtain direct measurements of superficial gas velocities).

In the riser, measurements can be performed at selected vertical locations using 17 cm ID ports. In addition, pressure taps are located every 15 cm on the vertical length of the riser. Proven techniques will be used to infer average pressure (static pressure taps), local solid volume fraction (capacitance probe), and local solid flux (isokinetic sample probe). In addition, a new time-of-flight anemometer probe currently under development at Cornell will enable measurements of local time-averaged particle velocity statistics, including $<u_s>$, $<v_s>$, or even $<u_s.v_s>$.

Whereas most experimental investigations of circulating beds have relied on average measurements (e.g., average solid hold-up in a bed cross-section), this research emphasizes the significance of local measurements in gas-solid flows. The debate over the true significance of particle clusters illustrates the need to perform local measurements in a riser. Following the qualitative model and the experimental observations of Yerushalmi, et al. (2), several descriptions of circulating bed hydrodynamics have relied on the behavior of particle clusters (Yang (18,19), Matsen (20,21), Kwauk, et al. (4,22), Grace anf Tuot (23)). The presence of such clusters - and the resulting drag reduction on the solid phase as a whole - would account for average velocity differences ("slip" velocity) between gas and solids in considerable excess of the terminal velocity of individual particles,

$$V = u/\bar{\varepsilon} - G/\rho_s\,(1-\bar{\varepsilon}) >> u_t\ .$$

(V = apparent slip velocity, u = superficial gas velocity, G = total solid flux, $\bar{\varepsilon}$ = average gas voidage, ρ_s= solid density, u_t= terminal velocity). However, as Gajdos and Bierl first pointed out (24), similar "slip" velocities may be explained by a high degree of spatial correlation between local solid flux and volume fraction. The debate over the significance of clusters still awaits conclusive evidence (see e.g., Geldart and Rhodes (25)). Clusters probably exist, but are they primarily responsible for high "slip" velocities? A simple calculation shows that, even if the local difference between gas and solid velocity is of the order of the terminal velocity u_t, reasonable profiles of gas velocity and solid volume fraction can be found that lead to an average slip velocity in considerable excess of u_t. In fact, for profiles such as

$$u_g = (1-r/R)^{1/m},$$

and

$$\varepsilon^n \propto u_g,$$

we find upon direct integration that, for any n and m real and positive,

$$\frac{V}{u_t} > \frac{u}{(1-\bar{\varepsilon})u_t} + 1\ . \qquad (9)$$

and

$$\frac{V}{u_t} > \frac{G}{\rho_s(1-\bar{\varepsilon})u_t} + 1\ . \qquad (10)$$

Typical numbers from Yerushalmi's experiments (FCC catalyst) are $u \approx 3$m/sec, $u_t \approx 0.2$m/sec and $(1-\bar{\varepsilon}) \approx 10\%$. In this case, eq. 9 yields $V/u_t > 150$, despite local relative velocities of order u_t. Remarkably, the qualitative trends suggested by eqs. 9 and 10 are in agreement with Yerushalmi's observations i.e., V/u_t goes up with increasing G and decreasing $(1-\bar{\varepsilon})$. Conceivably, Yerushalmi's observation of $V \gg u_t$ could be attributed to non-uniform cross-sectional profiles in the fast bed as convincingly as to the formation of solid clusters. Conclusive evidence about the true role of clusters is clearly dependent upon measurements of local flow variables. In fact, we note that several workers have already underscored the significance of lateral or vertical profiles (e.g., Gajdos and Bierl (24), Rhodes and Geldart (26), Weinstein, et al. (27), Hartge, et al. (28)). Hence the emphasis of this research program on local diagnostics probes.

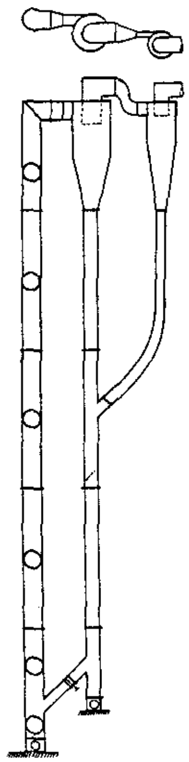

Fig. 1 Circulating bed facility

Commissioning of the circulating bed facility is expected for September 1987. Then, our fluidization research program is expected to meet the following objectives: Investigation of the effects of scale-up and the significance of riser aspect ratio (height/diameter); measurement of several local and average hydrodynamic parameters in the riser (pressure, solid volume fraction, solid flux and velocity vectors); publication of experimental results in dimensionless form for comparison,

$$p/\rho_s u^2 = f(z/d;\ u^2/gd,\ G/\rho_s u;\ D/d,\ H/D);$$

$$(1-\varepsilon) = f(r/d,\ z/d;\ u^2/gd,\ G/\rho_s u;\ D/d,\ H/D).$$

Hence, this research is expected to provide new insight on circulating bed hydrodynamics under scale-up, including e.g., the role of particle clusters, predictions of solid recirculation rates, etc. Note also that our cold flow technique using binary gas mixtures can be applied to several gas-solid reactors (calciners, combustors, cracker risers) in a variety of flow regimes. In the near future, availability of the circulating bed facility will promote additional projects on particle-laden flows. Further research is anticipated on jet penetration, gas backmixing, heat transfer and erosion phenomena. The present design should prove versatile enough to accommodate new research requirements (removable riser sections, convenient diagnostic ports, free access to each vessel, control of a wide range of operating conditions).

CONCLUSIONS

Circulating fluidization is a promising technique for large-scale facilities with high reactant flow rates. While the enonomy of scale dictates careful extrapolation from pilot units to larger plants, it is unfortunate that scale-up effects are still poorly understood.

Hence, a systematic technique to investigate scale-up in a single cold flow facility has been devised. Using gas-solid mixtures of adjustable properties, a single test bed can be made analogous to ideal combustors of several possible dimensions. In this context, a circulating bed facility designed to operate with binary fluidization mixtures will be available in September 1987 in the Department of Mechanical Engineering at Cornell University.

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