Situating the High-Density Circulating Fluidized Bed

John R. Grace, Allan S. Issangya, Dingrong Bai, and Hsiaotao Bi
Dept. of Chemical and Bio-Resource Engineering, University of British Columbia, Vancouver, Canada V6T 1Z4

Jingxu Zhu
Dept. of Chemical and Biochemical Engineering, University of Western Ontario, London, Canada N6A 5B9

Gas−solid flow in high-density circulating fluidized-bed risers differs significantly from that in conventional flow regimes, including fast fluidization and pneumatic conveying. It is shown that when there is no downward flux at the wall at high suspension densities, and hence no core/annulus structure, the flow corresponds to a dense suspension upflow regime. This flow regime is explained, and its key properties given. A lower boundary is also proposed for this new flow regime based on available data.

Introduction

Circulating fluidized bed (CFB) risers have been investigated extensively for the past two decades because of their practical applications, as well as their intrinsic interest. However, the overwhelming majority of such work has been conducted at net solids fluxes, G_s, less than 100 kg/m^2·s, at superficial gas velocities, U, between about 2 and 8 m/s, and at overall volumetric solids concentrations, C, less than about 0.1 (Zhu and Bi, 1995). While these conditions are relevant to CFB combustion, much higher solids fluxes and holdups are encountered in CFB risers used for solid−catalyzed reactions like fluid catalytic cracking and production of maleic anhydride. In such cases, the G_s is commonly 300 to 1,200 kg/m^2·s, with corresponding C values of 0.1 to 0.25.

We have recently undertaken a series of studies of high-density circulating fluidized-bed (HDCFB) risers, defined here as operations with $G_s > 200$ kg/m^2·s and $C > 0.10$ throughout the entire riser, using a dual-loop CFB system, as proposed by Bi and Zhu (1993). Our results demonstrate that the flow behavior then differs in important respects from that in low-density circulating fluidized-bed (LDCFB) systems. The most striking difference is that in HDCFB operation there is no net downflow of particles at the wall of the column, and hence no core/annulus flow structure, at least in the sense that this term has been widely used in describing and modeling CFB risers. In contrast, LDCFB risers show definite downward flux of particles in wall layers. In fact, this downflow is commonly regarded as the key defining feature of the fast fluidization flow regime, a major flow regime widely recognized in upward gas−solids flows, and the one usually associated with CFB risers. This raises the question of what flow regime to associate with HDCFB operations. It is this question that we address in this article.

Terminology and Flow Regimes in Gas Solids Systems

There is considerable confusion in the literature regarding terminology describing gas−solid flows. It is common for the same term to be used by different authors with totally different, even contradictory, meanings. One such term relevant to this article, "choking," has been treated in detail in an earlier article (Bi et al., 1993), where it is shown that there are three different choking phenomena and mechanisms, which we have labeled Types A (accumulative), B (blower- or standpipe-induced), and C (classic) choking. Once the three different mechanisms leading to solids accumulation or collapse of dilute flow are recognized, it is possible to make predictions and to use the terminology with much greater confidence and clarity. Even the term "high-density circulating fluidized bed," defined above, has been used differently (Wei et al., 1997), to denote systems where the density is relatively high at the bottom, but that otherwise appear to correspond to regular fast-fluidization systems.

Similar confusion exists in the literature over the terms "dense phase flow" and "dense phase conveying." To some

Correspondence concerning this article should be addressed to J. R. Grace.

(such as Leung and Wiles, 1976; Konrad, 1986), these terms denote flow below the Type C choking condition, that is, slug flow (or possibly bubbling or turbulent fluidization), corresponding to average volumetric solids concentrations of 0.35 or more. To others (such as Hirama et al., 1992; Bai et al., 1993; Gupta and Berruti, 1998), the term denotes pneumatic conveying, where virtually all particles travel upward and the volumetric concentration of particles is approximately 0.1 to 0.25. As discussed below, this second meaning is of considerable relevance to the flow conditions of the HDCFB riser.

The term "flow regime" is in common usage in the multiphase-flow literature. It is essential to understand what constitutes a flow regime. Three distinguishing criteria appear to be important for any flow regime:

1. A flow regime must have *distinctive features* of an observable physical nature.

2. A flow regime must also have *distinctive trends*, that is, it covers a range of operating variables within which the variation of dependent variables with respect to independent ones follows trends that differ from trends outside the flow regime.

3. The characteristics of the flow regime should be *capable of being* both *fully developed* (that is, independent of axial distance) and *statistically steady* (that is, independent of the time when averaging is initiated).

In the fluidization literature, all of the key, widely recognized flow regimes (packed-bed flow, bubbling fluidization, slug flow, turbulent fluidization, fast fluidization, and dilute pneumatic conveying) appear to be capable of meeting these criteria (although there is sometimes debate (such as Rhodes and Geldart, 1986) about the extent to which turbulent and fast fluidization reach fully developed flow conditions). In a recent article, Zijerveld et al. (1998) proposed no fewer than ten flow regimes, including several that are clearly entrance-related and never approach fully developed conditions. Such phenomena are not considered separate flow regimes in this article, in much the same way that transition flow in single-phase pipe flow is generally not considered a distinct mode (or flow regime), since it involves intermittency between laminar and turbulent flow, tending to progress from one to the other with distance downstream.

Features of High-Density Riser Flows

HDCFB experimental data are now available in the literature for equipment of a wide range of scales, including units of industrial scale (such as Saxton and Worley, 1970; Schuurmans, 1980; Derouin et al., 1997). Some of the key observations concerning the behavior of gas–solid suspensions in risers under high-density conditions are as follows:

1. Net downflow of particles at the wall, a commonly observed feature of fast fluidized beds, is absent. Solids move upward throughout the entire riser cross section (van Zoonen, 1962; Issangya et al., 1997a, 1998; Karri and Knowlton, 1998).

2. While there is no downflow at the wall, there are still considerable radial gradients in particle density, with higher particle concentrations near the wall than in the interior of the riser (van Zoonen, 1962; Saxton and Worley, 1970; Schuurmans, 1980; Issangya et al., 1997b). While there also do not appear to be clusters, there are certainly substantial fluctuations in local voidage.

3. In view of (1), there is reduced segregation of particles by size (Karri and Knowlton, 1998), and a closer approach to plug flow of both gas and solids (van Zoonen, 1962; Dry and White, 1989; Contractor et al., 1994; Derouin et al., 1997; Liu et al., 1999).

4. Axial profiles of solids concentration become relatively flat, with solids volumetric concentrations, averaged over the cross section, of about 0.1 to 0.25 (Weinstein et al., 1984; Contractor et al., 1994; Issangya et al., 1997a; Karri and Knowlton, 1998).

5. Both statistical and chaotic properties of pressure and local voidage fluctuations show that the high-density behavior differs markedly from that found when the same particles are fluidized in standard gas–solids flow regimes in the same column, even when compared at locations where time-mean voidages are equal (Bai et al., 1999).

Flow Regime Corresponding to High-Density Riser

The high-density riser has a number of useful properties (Zhu and Bi, 1995). High solids throughputs, coupled with low axial dispersion of gas and particles, make dense riser flow advantageous for catalytic reactions where relatively short and uniform residence times of catalyst particles are required, as in catalytic cracking. However, the question of which flow regime is applicable has not been resolved. Table 1 shows how the high-density CFB experimental results of Issangya et al. (1997a,b, 1998) would be categorized according to maps, criteria, and correlations from the literature. There is clearly no unanimity with respect to where such operations belong!

Placement of such operations therefore requires further scrutiny. In a recent article (Bai et al., 1999), we have compared bubbling and turbulent fluidized beds of the same FCC particles in a single column with respect to various statistical and chaotic properties determined from pressure and local voidage data with HDCFB and LDCFB operation. Significant differences were found in all of the measures considered —character of the signals; probability distributions; amplitude spectra; radial distributions of time-mean voidage, standard deviations, and intermittency index; cycle frequency; Hurst exponents; correlation dimensions; Kolmogorov entropy—among the various operations. In particular, HDCFB hydrodynamics were shown to differ substantially from those corresponding to the bubbling and turbulent flow regimes. The question of whether or not the differences between low- and high-density operation were sufficient to warrant distinct flow regimes was not addressed.

Figure 1 compares instantaneous voidage traces obtained using an optical-fiber probe under typical low- and high-density conditions ($C = 0.055$ and 0.194, respectively) at the same six radial positions and height in a 76.2-mm-diameter column for FCC particles ($d_p = 70\ \mu$m, $\rho_p = 1{,}600$ kg/m^3) at nearly the same superficial air velocity (8 m/s), but differing values of the net solids circulation flux, G_s. Full experimental details appear elsewhere (Issangya et al., 1997a; Issangya, 1998). The corresponding probability distributions and power-spectral densities are plotted in Figures 2 and 3, respectively. For the low-density case, that is, operation in a typical fast fluidization mode with downflow of strands and streamers along the outer wall, there is a lower concentration of particles at each

Table 1. Flow Regime for the High-Density Circulating Fluidized-Bed Operation According to Previous Criterion in the Literature

Authors	Criterion or Basis	HDCFB Flow Regime
Kunii and Levenspiel (1969)	$G_s^* > 80$?	Not dense-phase conveying
Leung and Wiles (1976)	Slip velocity, choking criteria	Fast fluidization
Li and Kwauk (1980)	Flow regime diagram	Unmapped territory
Konrad (1988)	$C > 0.1$? Filled with particles at one or more cross section?	Neither dilute nor dense flow
Yang (1988)	Axial profile of cross-section-mean voidage	Dense-phase transport
Karri and Knowlton (1991)	Between classic choking and minimum ΔP point	Fast fluidization
Hirama et al. (1992)	Density profile, regime diagram	Probably dense-phase transport
Bai et al. (1993)	Correlations	Fast fluidization or dilute pnuematic conveying
Bai et al. (1993)	Description of phenomena	Dense-phase conveying
Bai et al. (1996)	Cross-section-mean voidage	Turbulent fluidization
Bi and Grace (1997)	Flow regime map for upward transport; key characteristics of flow regime	Dilute-phase transport or fast fluidization
Gupta and Berruti (1998)	G_s, C, upward flow across entire radius?	Dense-phase transport

radial position examined. In both cases, however, denser structures (such as clusters and streamers) are observed as one moves outward from the axis to the wall, and there is a rather similar spread of amplitudes over a range of frequencies. The character of the signals shown in these three figures is neither sufficiently similar nor so different that one can say unambiguously whether or not the corresponding flow regimes are the same.

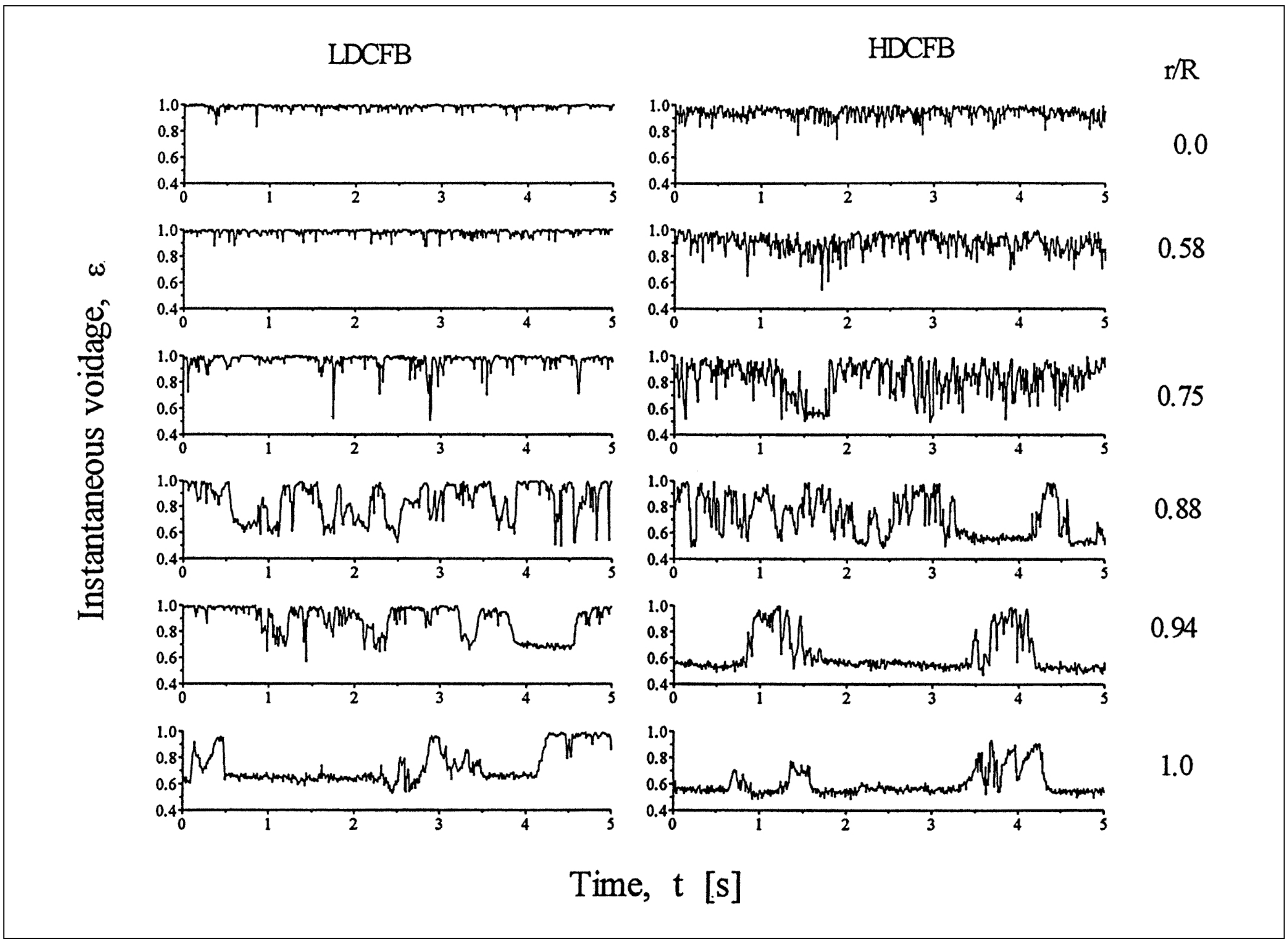

Figure 1. Instantaneous voidage traces for FCC particles 3.4 m above bottom of riser of diameter 76.2 mm for low-density (fast fluidization) operation ($U = 8.0$ m/s, $G_s = 93$ kg/m$^2 \cdot$s) and for high-density CFB conditions ($U = 7.7$ m/s, $G_s = 389$ kg/m$^2 \cdot$s).

Corresponding time-mean voidages: 0.945 and 0.806, respectively.

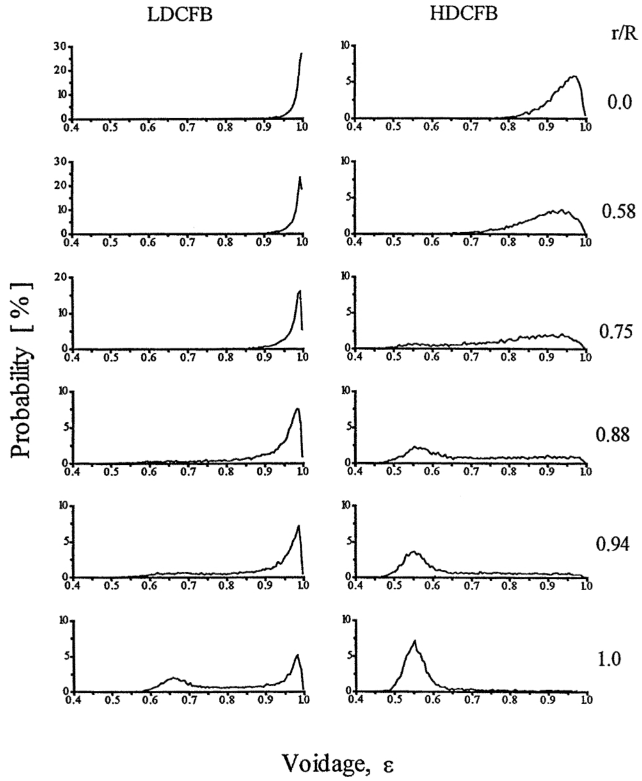

Figure 2. Probability distributions of local instantaneous voidage for the same positions and conditions as in Figure 1.

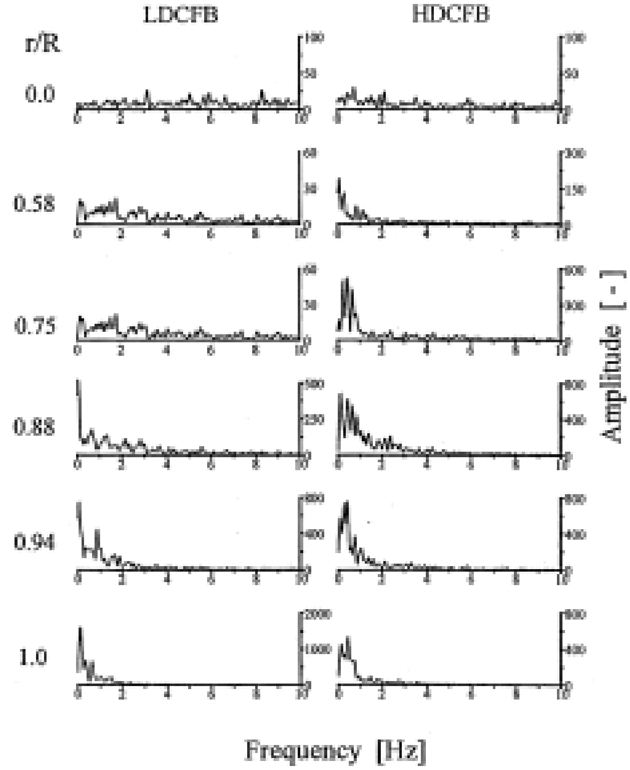

Figure 3. Power spectral densities of local voidage for the same positions and conditions as in Figures 1 and 2.

Radial distributions of time−mean voidage of local voidage are plotted in Figure 4, based on the same experimental study, for conditions corresponding to bubbling fluidization (labeled B), turbulent fluidization (T), dilute pneumatic conveying (P), as well as fast fluidization (marked as LDCFB) and a high-density circulating bed (HDCFB). The profile in the bubbling regime is nearly flat, clearly quite different from that in the other four cases, each of which shows a decrease in time−mean voidage with increasing radial coordinate, albeit at quite different levels. The corresponding radial profiles of standard deviations of local voidage appear in Figure 5. Here each profile has a somewhat unique character. The bubbling bed shows the greatest fluctuations at the axis of the column, whereas the pneumatic conveying and low-density (fast fluidization) profiles reach maxima at the outer wall. Both the turbulent bed and the HDCFB profiles reach maxima at intermediate radial positions and are nearly identical at the wall, though the standard deviations differ greatly in the interior of the column.

These results and those plotted elsewhere (Issangya, 1998; Bai et al., 1999) show that the high-density (HDCFB) operation does not fall neatly into any of the commonly studied flow regimes. It is possible, on the basis of the data and observations at the wall of the vessel, to affirm that their behavior differs significantly from the bubbling, slugging, and turbulent fluidization flow regimes. Hence these flow regimes are not considered further in this article. This leaves the following possibilities to be explored.

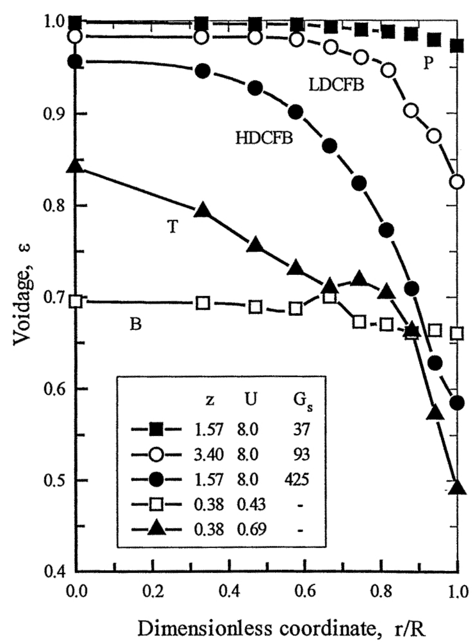

Figure 4. Radial profiles of time-mean voidage for HDCFB conditions with those for bubbling (B), turbulent fluidization (T), fast fluidization (LDCFB), and pneumatic conveying (P).

All results are for FCC particles in 76.2-mm-diameter riser with ambient air as the gas.

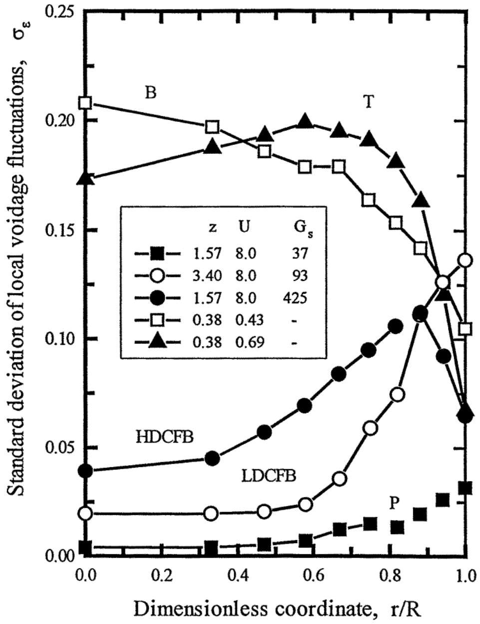

Figure 5. Radial profiles of standard deviation of local voidage fluctuations for the same data as in Figure 4.

1. *Pneumatic Conveying.* Most pneumatically conveyed suspensions contain no more than a few percent of particles by volume. Moreover, except when very dilute indeed, they are usually characterized by solids downflow at the wall. With solids concentrations of about 0.1 to 0.25 by volume and no net downflow at the wall, the high-density CFB suspensions considered in this article differ substantially from pneumatically conveyed suspensions. Observable features and key trends (such as in axial dispersion) differ widely. A new definition would be needed if the high-density systems were to be included as part of an expanded pneumatic transport flow regime. Moreover, as shown in Figure 4, fast fluidization (initiated by type A (accumulative) choking) intervenes between pneumatic conveying and the HDCFB conditions, and it would be unprecedented to return to a flow regime having left it when a variable is changed continuously.

2. *Fast Fluidization.* Most of those who have worked on high-density systems of the type discussed here have assumed that they are encompassed within the fast fluidization (FF) flow regime. Fast fluidized beds were first described under the term "highly expanded fluid beds" by Reh (1971), with the term "fast fluidization" emerging several years later (Grace, 1974). Fast fluidization has many definitions, key elements being as follows:

- Yerushalmi et al. (1978): relatively high solids concentration; aggregation of solids in clusters and strands; extensive backmixing of solids; slip velocities $\gg v_T$.
- Li and Kwauk (1980): a dilute zone at the top and a dense zone below, with a point of inflection usually separating these two regions.
- Leung (1980): clusters; some solids downflow; slip velocity $> v_T$; pressure gradient decreases when the gas velocity is increased at constant G_s.
- Kwauk et al. (1986): solids clusters in a dilute continuum of sparsely dispersed discrete particles.
- Rhodes and Geldart (1986): operation above [classic] choking point; insensitive to changes in U and G_s.
- Takeuchi et al. (1986): steady cocurrent upflow of gas–solids mixture; denser bed below, coexisting with relatively dilute suspension above.
- Horio and Morishita (1988): dilute phase continuous, dense phase dispersed.
- Karri and Knowlton (1991): between [classic] choking boundary and gas velocity at which pressure gradient reaches a minimum for a given G_s.
- Bai et al. (1993): aggregation tendency; S-shaped or exponential solids density profile; core-annulus segregation.

Except for the Rhodes and Geldart definition, the high-density riser does not satisfy any of these definitions. There is no net downflow at the wall, limited tendency to form clusters, relatively flat axial density profiles with no dilute zone at the top, and no point of inflection or S-shaped profile. Clearly, if the high-density systems are assigned to the fast fluidization regime, then a new definition would be required, and this definition would have to be broad enough to encompass both high-density systems, where solids travel upward at the wall, and lower density fast fluidized suspensions, which exhibit core-annulus behavior, with downward solids motion at the wall.

3. *Dense-Phase Conveying.* As indicated earlier, there is considerable confusion with respect to the terms "dense phase conveying" and "dense phase flow," the latter used in a recent article from our group (Bi and Grace, 1999). Setting aside use of these terms to cover slugging, turbulent, and bubbling beds, we examine several definitions:

- Kunii and Levenspiel (1969): $G_s/(\rho U)$ is required to be greater than about 80 or solids volumetric concentration, C, greater than about 0.1.
- Yerushalmi and Avidan (1985): These authors use the term "riser transport reactor" to mean what is called dense-phase conveying here. No definition is given, but they speculate that the difference between this flow regime and fast fluidization probably lies in the disposition of the solids.
- Drahos et al. (1988): Their "dense flow" is characterized by "clouds of particles and voids free from particles," and is held to be separate from bubbling beds.
- Bai et al. (1993): These authors used the term dense pneumatic conveying to denote a transitional flow regime where there is a relatively high solids concentration and a more uniform flow structure than in conventional fast fluidized beds.
- Gupta and Berruti (1998): These authors considered this regime to be an extension of turbulent fluidization and specified $G_s > 100$ kg/m^2·s, $C > 0.15$, and net solids flowing upward at every radial position as the defining features.

In contrast to the definitions of fast fluidization which the HDCFB does not meet, flow in the high-density riser is consistent with the last two of the preceding definitions, as well as with the second part of the first definition. However, the fourth definition defines one flow regime in terms of another, while the last one adopts arbitrary limits which, as shown below, do not accurately distinguish between the two flow regimes. To avoid the confusion in terminology with classically choked dense-phase systems that have much higher solids holdups, we propose a new label—*Dense-Suspension Upflow*—and provide what we believe to be a better definition to cover the unique features of this flow regime: "We define dense-suspension upflow (DSU) to be the flow regime where there is net upflow of solids across the entire riser, strong interactions between particles, gas velocity/solids flux conditions beyond any of the types of choking, and overall volumetric solids concentration of order 0.1 to 0.25, with little axial variation." It is clear that what we have heretofore called high-density circulating fluidized bed (HDCFB) operation corresponds to the DSU regime as defined here. Defined in this manner, DSU also meets the criteria established earlier for flow regimes:

1. It has *distinctive features:* DSU is distinguished from fast fluidization by upflow of particles at the wall, as well as by higher solids concentrations and lower axial gradients; from pneumatic conveying by much greater densities; and from turbulent, slugging and bubbling beds by the lack of voids or other large-scale structures.

2. Key characteristics, for example, axial density profiles and axial dispersion of both gas and solids, show *different trends* than for the neighboring flow regimes.

3. Axial density profiles (such as Weinstein et al., 1984; Isangya et al., 1997a) indicate that *fully developed* conditions can be achieved under DSU conditions, providing that the column is tall enough. *Statistically steady* conditions are also possible.

Likely Explanation for the Initiation of Dense Phase Upflow

Consider a transparent riser in which gas is forced upward at a constant superficial velocity of 5 to 10 m/s, with group A particles such as FCC injected near the bottom, at first very slowly and then at a gradually increasing rate. At first the particles are carried upward individually in *dilute pneumatic conveying*, faster in the core than near the wall due to the no-slip condition imposed by the wall on the gas. There is substantial lateral transport of particles to and from the wall due to gas turbulence. With increasing G_s, lift forces induced by shear in the wall layer, coupled with the low gas velocity in the boundary layer, combine to allow particles to congregate at the wall (Senior and Grace, 1999), forming streamers, causing type A (accumulative) choking and initiating the flow regime of *fast fluidization* (Bi et al., 1993).

As G_s is increased further, particles travel downward in a wall layer of increasing thickness (Bi et al., 1996). The core of the riser is occupied by a relatively dilute, turbulent, upward-traveling suspension flow, which exerts some shear on the downward-moving outer wall layer, sometimes dislodging and reversing the direction of wall clusters. As G_s increases, more and more of the core is occupied by particles for two reasons: (1) less of the riser cross section is available to the core as the wall layer grows thicker; and (2) more particles must be transported through the core to provide for the increased solids flux. This increased particle concentration in the core tends to damp out turbulence there, while also increasing the effective viscosity of the rising suspension, thereby imposing more shear on the descending wall layer. Because the wall layer itself is thicker, the influence of the gas no-slip condition at the wall itself is more remote. Given these factors, a point is reached where the shear imposed by the inner flow is sufficient to reverse (on average) the direction of the particle motion near the wall, initiating DSU. As G_s increases even further, a relatively dense suspension travels upward across the entire riser. Eventually, the carrying capacity of the gas is exceeded and the suspension must collapse, due presumably to type C (or B) choking, and it is then likely that *turbulent fluidization* will ensue.

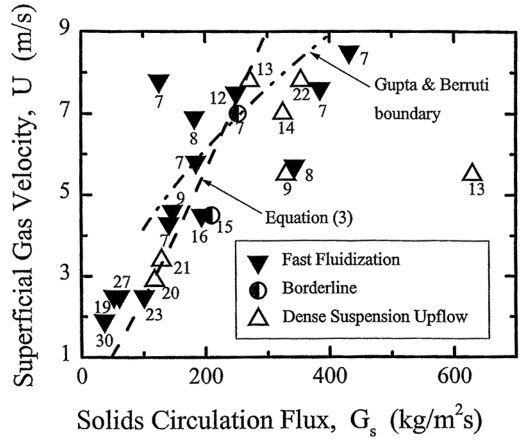

Figure 6. Data for catalyst particles with $C \geq 0.07$ where one can say whether the riser was operating in the fast-fluidization or dense-suspension upflow regime.

For sources of data, see Table 2. Closed symbols: fast fluidization; half-closed symbols: borderline; open symbols: dense suspension upflow. Numbers represent values of C, expressed as percentages.

Boundaries of the Dense-Suspension Upflow Regime

It remains to propose the lower boundary of the DSU flow regime. Operation in this flow regime clearly requires both high superficial gas velocities (at least several meters per second) and high solids fluxes (at least 200 kg/m$^2 \cdot$s). The regime has been primarily explored for Geldart type-A powders, though the recent data of Karri and Knowlton (1998) with sand indicate that it can also be achieved with group B particles.

Existing data for FCC and other group A catalysts showing operation in the DSU regime (open, upward-pointing triangles) and on the boundary (half-open circles) are shown in Figure 6, together with data for neighboring fast fluidization (FF) conditions (closed, downward-pointing triangles). The criteria used to distinguish the regimes are:

- For Bader et al. (1988), Wei et al. (1997), and Issangya (1998): flux sampling at the wall, with net downward fluxes there showing FF and upward fluxes DSU
- For van Swaaij et al. (1970): shear stress at wall, downward (FF), or upward (DSU)
- For van Zoonen (1962): particle velocity direction at wall indicated by Prandtl tube
- For Weinstein et al. (1984) and Contractor et al. (1994): flattening of axial voidage profiles at values of about 0.8 indicates the DSU regime.

Table 2. Properties of Particles and Column Dimensions for Experimental Data Employed in Figures 6 and 7

Author(s)	Particles	Diameter (μm)	Density (kg/m^3)	Riser Diam. (mm)	Riser ht. (m)
Bader et al. (1988)	FCC	76 (mean)	1,714	305	12.2
Contractor et al. (1994)	VPO Catalyst	70 (mean)	1,570	152	27.4
Issangya (1998)	FCC	70 (mean)	1,600	76	6.1
Karri and Knowlton (1998)	Sand	175 (mean)	2,643	203	14
van Swaaij et al. (1970)	FCC	Not given	Not given	180	Not given
van Zoonen (1962)	FCC	20–150 (range)	Not given	51	10
Wei et al. (1997)	FCC	54 (mean)	1,398	186	8.5
Weinstein et al. (1984)	HPZ20 Catalyst	49 (mean)	1,450	152	8.5

Note: Air under ambient conditions was used in each of these studies.

Some of the scatter in the figure arises because the properties of the catalyst particles were not identical, as shown in Table 2. The column diameters for the various studies also differed widely, again shown in Table 2; column diameter is known to be a major factor affecting the transition to turbulent fluidization (Chehbouni et al., 1995), and may also influence the lower DSU boundary. The measurements were also taken at differing levels in risers of various overall heights, and it is not known how or whether this factor can affect the results.

The numbers next to the points in Figure 6 indicate the approximate cross-sectional mean solids concentration by volume, C, expressed as percentages. Data have been excluded when C was less than 0.07 to avoid the possibility of including data for the lower concentration conveying condition (see Karri and Knowlton, 1998) where no downflow occurs at the wall. This restriction means that we are unable to use some flux data (such as Azzi et al., 1991; Herb et al., 1992), where the suspension density was too low. Except for three points from van Swaaij et al. (1970), there is a relatively clear boundary between the FF and DSU regimes. The corresponding boundary between FF and dense-phase conveying proposed by Gupta and Berruti (1998), shown in the figure, is very similar to the saturation carrying-capacity correlation of Bai and Kato (1996); neither does a very good job of separating the FF and DSU regimes.

Clearly both the gas velocity and the solids flux are important in establishing the lower DSU boundary. Moreover, it would appear from the C values that the solids volumetric concentration corresponding to the transition decreases as both U and G_s increase along the proposed boundary. It is not clear why the three van Swaaij et al. (1970) points lie beyond the proposed boundary. Shear-stress measurements must have been difficult under such high-flux conditions, and their accuracy is unknown. Moreover, it is not clear whether the shear stress changes direction at the same point as the local solids flux at the wall, just as we have shown (Bi et al., 1996) that particle velocity and solids flux do not reach zero at the same position near the wall.

If variables like column diameter, particle-size distribution, interparticle forces, and height are assumed, as a first approximation, to be of secondary importance, the superficial gas velocity corresponding to the onset of DSU should be a function of five variables, that is,

$$U = f\left(G_s, \bar{d}_p, \rho, g\Delta\rho, \mu\right), \tag{1}$$

where $\Delta\rho = \rho_p - \rho$ (density difference). Since there are three dimensions (length, mass, and time), this equation can be rewritten in dimensionless form as,

$$U^* = F\left(d_p^*, G_s^*\right), \tag{2}$$

where $U^* = U[\rho^2/\mu g\Delta\rho]^{1/3}$ is a dimensionless gas velocity; $d_p^* = Ar^{1/3} = \bar{d}_p[\rho\Delta\rho g/\mu^2]^{1/3}$ is a dimensionless particle diameter; and $G_s^* = G_s/(\rho U)$, often called the solids loading ratio, is the dimensionless solids circulation flux. Figure 7 plots U^* against G_s^*, with corresponding d_p^* values given as numbers adjacent to the points. The transition to DSU is seen to occur at loading ratios of 30 to 40. The van Swaaij et al. (1970) data are omitted for the reasons discussed earlier and also because their particle diameter and density were not provided. The sand (group B) data of Karri and Knowlton (1998), based on flux sampling at the wall showing whether solids travel upward or downward there, are included in Figure 7. There is no clear evidence that the sand data (for which $d_p^* = 8.0$) differ when plotted in this manner from those for the group A catalyst particles (where $d_p^* = 1.8$ to 3.0). Hence a single boundary is drawn on Figure 7, independent of d_p^*, to separate FF from the DSU regime, leading to the dimensionally consistent equation for the onset of DSU:

$$U_{\mathrm{DSU}} = 0.0113\, G_s^{1.192}\rho^{-1.064}(\mu g\Delta\rho)^{-0.064}. \tag{3}$$

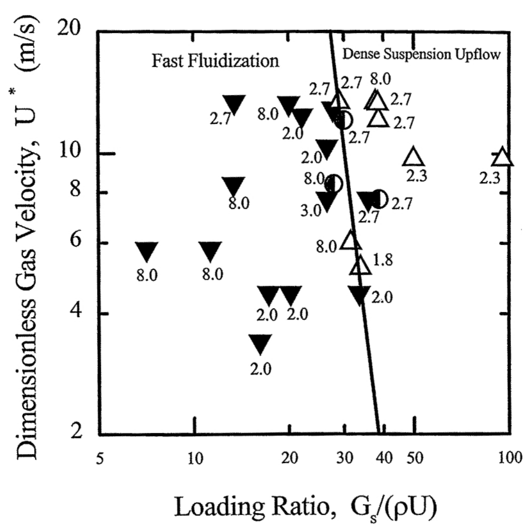

Figure 7. Dimensionless gas velocity vs. loading ratio for catalyst and sand particles.

Numbers show d_p^* values. Sources of data appear in Table 2.

Table 3. Key Features of Gas–Solid Fluidization and Vertical Flow

Flow Regime	Minimum Superficial Velocity	Upper Bed Surface	Direction of Particle Motion at Wall	Overall Voidage	Key Macroscales	Axial Dispersion (1 = lowest; 6 = highest)
Bubbling	Minimum bubbling	Distinct, wavy	Mostly down	~ 0.4 to ~ 0.65	Bubble diameter	6
Slug flow	Minimum slugging	Distinct, periodic up–down motion	Mostly down	~ 0.6 to ~ 0.7	Column diameter, slug length	5
Turbulent fluidization	U_c: minimum pressure fluctuations	Diffuse	Mostly down	~ 0.65 to ~ 0.8	Scale of turbulence, void dimension	3
Fast fluidization	See Bi et al. (1995)	None	Mostly down	~ 0.8 to ~ 0.97	Wall layer thickness, strand length	4
Dense suspension upflow	U_{DSU}, Eq. 3	None	Mostly upward	~ 0.75 to ~ 0.9	Scale of turbulence	2
Dilute pneumatic conveying	Transport velocity	None	Mostly upward	~ 0.97 to 1.0	Scale of turbulence	1

This line is also plotted on Figure 6 where it again successfully separates points belonging to the FF and DSU regimes. This boundary is also consistent with the condition where axial gas dispersion reaches a maximum with increasing superficial gas velocity (Liu et al., 1999). Equation 3 should only be used with great caution outside the range for which it was obtained, that is, air at ambient conditions as the gas; $1.8 \leq d_p^* \leq 8.0$; $C \geq 0.07$; $7 \leq G_s/(\rho U) \leq 100$; 51 mm $\leq D_r \leq$ 305 mm; 6.1 m $\leq H \leq$ 27.4 m. In particular, the dependences on gas density, viscosity, temperature, and pressure need to be confirmed experimentally, since these have not been varied at all at this point. In addition, more data are needed to characterize the dependence on d_p^* and d_p, which do not appear in the preceding equation, and to extend the range of column diameters investigated.

This article has also only considered fully developed flow conditions. Much of the earlier work on gas–solids flow regimes (such as Bai and Kato, 1995) and much of the practical interest are concerned with regions where flow is in the process of developing. As already noted, complex flow patterns can occur under developing conditions. As far as DSU is concerned, it is clear that being able to achieve the flow regime depends, among other factors, on being able to provide sufficient pressure drop and acceleration of particles in the entry region. However, once the flow regime has been achieved in a tall column, the flow regime appears to be robust and self-sustaining over a considerable range of superficial gas velocities and solids fluxes.

Summary of Regime Key Features

Key features of the DSU regime are summarized in Table 3 together with those of the other key flow regimes for upward flow in gas–solids systems. More work is clearly required to delineate fully and quantitatively such properties as gas and particle axial and radial mixing and the upper limit of the DSU regime.

Conclusions

High-density circulating fluidized-bed (HDCFB) systems offer significant advantages for reactions involving gas and particles, in particular low backmixing of gas and solids coupled with high solids loading and excellent gas–particle contacting. Local voidage measurements indicate that there are significant differences in hydrodynamics for systems operated under HDCFB conditions compared with those corresponding to previously recognized gas–solid flow regimes, including fast fluidization. Comparison of the behavior with definitions in the literature suggests that HDCFB hydrodynamics do not belong to the FF flow regime. Instead, they correspond to what some have termed "dense phase conveying." However, this term has been used in such different manners in the literature that a new less ambiguous name, "dense suspension upflow," is suggested as a new label for the flow regime, together with a distinctive definition. An explanation and lower boundary for this flow regime are suggested, based on available data for group A and B particles. More data are needed to confirm this boundary and to extend it to other types of particles, gases other than air, elevated temperatures and pressures, and a broader range of column diameters.

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Notation

Ar = Archimedes number
D_r = riser diameter, m
d_p = mean particle diameter, m
g = acceleration of gravity, m/s^2
H = overall riser height, m
U_c = superficial gas velocity at which amplitude of pressure fluctuations reaches a maximum, m/s
v_T = terminal settling velocity of particles of mean size, m/s
z = height above bottom of riser, m

Greek letters

ϵ = voidage
μ = gas viscosity, Pa·s
ρ = gas density, kg/m^3
ρ_p = particle density, kg/m^3
σ_ϵ = standard deviation of local voidage

Literature Cited

Azzi, M., P. Turlier, J. F. Large, and J. R. Bernard, "Use of a Momentum Probe and Gamma Densitometry to Study Local Properties of Fast Fluidized Beds," *Circulating Fluidized Bed Technology III*, P. Basu, M. Horio, and M. Hasatani, eds., Pergamon, Oxford, p. 189 (1991).

Bader, R., J. Findlay, and T. M. Knowlton, "Gas/Solid Flow Patterns in a 30.5-cm Diameter Circulating Fluidized Bed," *Circulating Fluidized Bed Technology II*, P. Basu and J. F. Large, eds., Pergamon, Oxford, p. 123 (1988).

Bai, D., A. S. Issangya, and J. R. Grace, "Characteristics of Gas Fluidized Beds in Different Flow Regimes," *Ind. Eng. Chem. Res.*, **38**, 803 (1999).
Bai, D., Y. Jin, and Z. Yu, "Flow Regimes in Circulating Fluidized Beds," *Chem. Eng. Technol.*, **16**, 307 (1993).
Bai, D., and K. Kato, "Saturation Carrying Capacity of Gas and Flow Regimes in CFB," *J. Chem. Eng. Jpn.*, **28**, 179 (1995).
Bai, D., E. Shibuya, N. Nakagawa, and K. Kato, "Characterization of Gas Fluidization Regimes Using Pressure Fluctuations," *Powder Technol.*, **87**, 105 (1996).
Bi, H. T., and J. R. Grace, "Introduction to Circulating Fluidized Beds," *Circulating Fluidized Beds*, Chap. 1, J. R. Grace, A. A. Avidan, and T. M. Knowlton, eds., Chapman & Hall, London (1997).
Bi, H. T., and J. R. Grace, "Flow Patterns in High-Velocity Fluidized Beds and Pneumatic Conveying," *Can. J. Chem. Eng.*, **77**, 223 (1999).
Bi, H. T., J. R. Grace, and J. Zhu, "On Types of Choking in Pneumatic Systems," *Int. J. Multiphase Flow*, **19**, 1077 (1993).
Bi, H. T., J. R. Grace, and J. Zhu, "Regime Transitions Affecting Gas-Solids Suspension and Fluidized Beds," *Trans. Inst. Chem. Eng.*, **73**, 154 (1995).
Bi, H. T., J. Zhou, S. Z. Qin, and J. R. Grace, "Annular Wall Layer Thickness in Circulating Fluidized Bed Risers," *Can. J. Chem. Eng.*, **74**, 811 (1996).
Bi, H. T., and J. Zhou, "Static Instability Analysis of Circulating Fluidized Beds and Concept of High-Density Risers," *AIChE J.*, **39**, 1272 (1993).
Chehbouni, A., J. Chaouki, C. Guy, and D. Klvana, "Effets de différents paramètres sur les vitesses de transition de la fluidisation en régime turbulent," *Can. J. Chem. Eng.*, **73**, 41 (1995).
Contractor, R. M., G. S. Patience, D. I. Garnett, H. S. Horowitz, G. M. Sisler, and H. E. Bergna, "A New Process for n-Butane Oxidation to Maleic Anhydride Using a Circulating Fluidized Bed Reactor," *Circulating Fluidized Bed Technology IV*, A. A. Avidan, ed., AIChE, New York, p. 387 (1994).
Derouin, C., D. Nevicato, M. Forissier, G. Wild, and J. R. Bernard, "Hydrodynamics of Riser Units and Their Impact on FCC Operation," *Ind. Eng. Chem. Res.*, **36**, 4504 (1997).
Drahos, J., J. Cermak, R. Guardani, and K. Schugerl, "Characterization of Flow Regime Transitions in a Circulating Fluidized Bed," *Powder Technol.*, **56**, 41 (1988).
Dry, R. J., and C. C. White, "Gas Residence-Time Characteristics in a High-Velocity Circulating Fluidised Bed of FCC Catalyst," *Powder Technol.*, **58**, 17 (1989).
Grace, J. R., "Fluidization and Its Application to Coal Treatment and Allied Processes," *AIChE Symp. Ser.*, **70**(141), 21 (1974).
Gupta, S. K., and F. Berruti, "Modeling Considerations for Large Scale High Density Risers," *Fluidization IX*, L. S. Fan and T. M. Knowlton, eds., Engineering Foundation, New York, p. 205 (1998).
Herb, B., S. Dou, K. Tuzla, and J. C. Chen, "Solid Mass Fluxes in Circulating Fluidized Beds," *Powder Technol.*, **70**, 197 (1992).
Hirama, T., H. Takeuchi, and T. Chiba, "Regime Classification of Macroscopic Gas-Solid Flow in a Circulating Fluidized Bed Riser," *Powder Technol.*, **70**, 215 (1992).
Horio, M., and K. Morishita, "Flow Regimes of High Velocity Fluidization," *Jpn. J. Multiphase Flow*, **2**, 117 (1988).
Issangya, A. S., "Flow Dynamics in High Density Circulating Fluidized Beds," PhD Diss., Univ. of British Columbia, Vancouver, B.C., Canada (1998).
Issangya, A. S., D. Bai, H. T. Bi, K. S. Lim, J. Zhu, and J. R. Grace, "Axial Solids Holdup Profiles in a High-Density Circulating Fluidized Bed Riser," *Circulating Fluidized Bed Technology V*, M. Kwauk and J. Li, eds., Science Press, Beijing, p. 60 (1997a).
Issangya, A. S., D. Bai, J. R. Grace, K. S. Lim, and J. Zhu, "Flow Behavior in the Riser of a High-Density Circulating Fluidized Bed," *AIChE Symp. Ser.*, **93**(317), 25 (1997b).
Issangya, A. S., D. Bai, J. R. Grace, and J. Zhu, "Solids Flux Profiles in a High Density Circulating Fluidized Bed Riser," *Fluidization IX*, L. S. Fan and T. M. Knowlton, eds., Engineering Foundation, New York, p. 197 (1998).
Karri, S. B. R., and T. M. Knowlton, "A Practical Definition of the Fast Fluidization Regime," *Circulating Fluidized Bed Technology III*, P. Basu, M. Horio, and M. Hasatani, eds., Pergamon, Oxford, p. 67 (1991).
Karri, S. B. R., and T. M. Knowlton, "Flow Direction and Size Segregation of Annulus Solids in a Riser," *Fluidization IX*, L. S. Fan and T. M. Knowlton, eds., Engineering Foundation, New York, p. 189 (1998).
Konrad, K., "Dense-Phase Pneumatic Conveying: A Review," *Powder Technol.*, **49**, 1 (1986).
Kunii, D., and O. Levenspiel, *Fluidization Engineering*, Wiley, New York (1969).
Kwauk, M., N. Wang, Y. Li, B. Chen, and Z. Shen, "Fast Fluidization at ICM," *Circulating Fluidized Bed Technology*, P. Basu, ed., Pergamon, Toronto, p. 33 (1986).
Leung, L. S., "The Ups and Downs of Gas-Solid Flow—A Review," *Fluidization*, J. R. Grace and J. M. Matsen, eds., Plenum, New York, p. 25 (1980).
Leung, L. S., and R. J. Wiles, "A Quantitative Design Procedure for Vertical Pneumatic Conveying Systems," *Ind. Eng. Chem. Proc. Des. Dev.*, **15**, 552 (1976).
Li, Y., and M. Kwauk, "The Dynamics of Fast Fluidization," *Fluidization*, J. R. Grace and J. M. Matsen, eds., Plenum, New York, p. 537 (1980).
Liu, J., J. R. Grace, H. T. Bi, and J. Zhu, "Gas Mixing in a High Density Circulating Fluidized Bed Riser," *Circulating Fluidized Bed Technology VI*, J. Werther, ed., pp. 399-404 (1999).
Reh, L., "Fluidized Bed Processing," *Chem. Eng. Progr.*, **67**(2), 58 (1971).
Rhodes, M., and D. Geldart, "The Hydrodynamics of Re-Circulating Fluidized Beds," *Circulating Fluidized Bed Technology*, P. Basu, ed., Pergamon, Toronto, p. 193 (1986).
Saxton, A. L., and A. C. Worley, "Modern Catalytic-Cracking Design," *Oil & Gas J.*, **68**(20), 82 (1970).
Schuurmans, H. J. A., "Measurements in a Commercial Catalytic Cracking Unit," *Ind. Eng. Chem. Proc. Des. Dev.*, **19**, 267 (1980).
Senior, R. C., and J. R. Grace, "Formation of Particle Streamers in the Wall Region of Circulating Fluidized Bed Risers," *Fluidization IX*, L.-S. Fan and T. M. Knowlton, eds., Engineering Foundation, New York, p. 165 (1998).
Takeuchi, H., T. Hirama, T. Chiba, J. Biswas, and L. S. Leung, "A Quantitative Definition and Flow Diagram for Fast Fluidization," *Powder Technol.*, **47**, 195 (1986).
van Swaaij, W. P. M., C. Buurman, and J. W. van Breugel, "Shear Stresses on the Wall of a Dense Phase Riser," *Chem. Eng. Sci.*, **25**, 1818 (1970).
van Zoonen, D., "Measurements of Diffusional Phenomena and Velocity Profiles in a Vertical Riser," Symp. Int. Fluids and Particles, Inst. Chem. Eng. London, p. 65 (1962).
Wei, F., F. Lu, Y. Jin, and Z. Yu, "Mass Flux Profiles in a High Density Circulating Fluidized Bed," *Powder Technol.*, **91**, 189 (1997).
Weinstein, H., R. A. Graff, M. Meller, and M. J. Shao, "The Influence of the Imposed Pressure Drop across a Fast Fluidized Bed," *Fluidization IV*, D. Kunii and R. Toei, eds., Engineering Foundation, New York, p. 299 (1984).
Yang, W. C., "A Model for the Dynamics of a Circulating Fluidized Bed Loop," *Circulating Fluidized Bed Technology II*, P. Basu and J. F. Large, eds., Pergamon, Oxford, p. 181 (1988).
Yerushalmi, J., and A. Avidan, "High-Velocity Fluidization," *Fluidization*, 2nd ed., Chap. 7, J. F. Davidson, R. Clift, and D. Harrison, eds., Academic Press, London, p. 225 (1985).
Yerushalmi, J., N. T. Cankurt, D. Geldart, and B. Liss, "Flow Regimes in Vertical Gas-Solid Contact Systems," *AIChE Symp. Ser.*, **74**(176), 1 (1978).
Zhu, J., and H. T. Bi, "Distinctions between Low Density and High Density Circulating Fluidized Beds," *Can. J. Chem. Eng.*, **29**, 644 (1995).
Zijerveld, R. C., F. Johnsson, A. Marzocchella, J. C. Schouten, and C. M. van den Bleek, "Fluidization Regimes and Transitions from Fixed Bed to Dilute Transport Flow," *Powder Technol.*, **95**, 185 (1998).

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