Static Instability Analysis of Circulating Fluidized Beds and Concept of High-Density Risers

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An instability analysis has been carried out to elucidate the unsteady flow conditions encountered in the typical circulating fluidized bed units in light of the conveyor-solids feeder interaction. The results successfully predict the critical velocity and the maximum solids circulation rates reported in the literature and explain the origin of such unstable conditions. Furthermore, the simulation, for the first time, reveals the importance of unit structure in improving the performance of circulating fluidized bed systems. Finally, the concept of a high-density circulating fluidized bed is proposed.

Introduction

In a gas-solid vertical co-current upflow system, particles are carried up the riser as a dispersed suspension at high gas velocity and low solids flux, called dilute-phase transport. As the gas velocity is reduced at a fixed solids flux, solids concentration increases. Eventually a point is reached where the suspension collapses, and particles are then transported up the riser in the form of slugging/bubbling flow. This point is generally defined as the "choking" point by Zenz and Othmer (1960), characterized by the formation of slugs and severe instability in the system. It has also been observed that choking is dependent on the gas-solid-tube system (Zenz, 1949; Yousfi and Gau, 1974; Yang, 1975; Smith, 1978). In some systems such as circulating fluidized beds, however, other types of unstable conditions may occur prior to choking.

Let us explain the formation of the nonchoking type unstable conditions. Zenz and Othmer (1960) noted that an unstable condition can result when the blower cannot offer sufficient pressure head to support the whole bed. With blowers characterized by reducing volumetric delivery with increasing delivery pressure, Doig (1963) and Leung et al. (1971) analyzed such an instability process, as shown in Figure 1. The solid lines represent pressure drop vs. gas flow rate in the riser, while the dashed line gives the characteristics of the blower. For a fixed solids flow rate, the figure shows two possible operating points A and B. From a sensitivity analysis, it can be shown that operating point B is inherently unstable. A small reduction in the gas flow rate at B would result in an increase in the pressure drop, leading to a further decrease in the gas flow rate and eventual blockage of the riser. For Group B and D particles, the analysis of Bandrowski et al. (1981) and Matsumoto et al. (1982) shows that a similar instability also exists for a given blower-conveyor system at which the blower characteristic curve intercepts the conveying system characteristic curve tangentially. Furthermore, the gas velocity at this critical point is generally higher than the choking velocity and can be reduced toward the choking velocity by improving the blower characteristics.

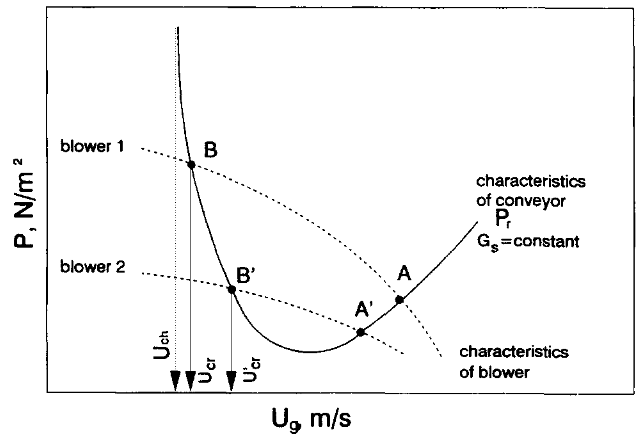

Figure 1. Operational instability due to insufficient blower pressure head for solids conveyor (adapted from Doig, 1963).

In the circulating fluidized beds where the risers are always coupled with the downcomers, an unstable state is also attained

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at which steady operation at a given solids flux becomes impossible although no choking phenomenon is observed around this point (Knowlton and Bachovchin, 1976; Takeuchi et al., 1986; Bai et al., 1987; Bader et al., 1988; Hirama et al., 1992). This unstable point also depends on solids inventory in the standpipe for the unit with a standpipe solids feeder, with lower critical velocity for higher solids inventory (Hirama et al., 1992). Furthermore, in the same unit as used by Takeuchi et al. (1986), Hirama et al. (1992) showed that such an unstable condition can be circumvented by utilizing a screw feeder instead of a standpipe as the solids feeding system so that the interaction between the riser and the downcomer is avoided. Such an unstable state results from the buildup of an inappropriate pressure balance between the riser side and the downcomer side. Therefore, this kind of instability is different from the choking condition and from the unstable condition which results from the interaction between the blower and conveyor.

To elucidate the origin of this flow instability, an analysis is carried out based on the concept of pressure balance in the whole circulation loop. The simulation results are compared with experimental data.

Once this instability due to the inappropriate pressure balance between the riser and the downcomer is avoided with the guidance of this analysis, a circulating fluidized bed should be able to operate at much higher solids flux and with much higher solids concentration in the riser. The concept of high density riser is thus introduced after the analysis, which is of significant importance to both the process industry and academic research.

Instability Analysis

A typical circulating fluidized bed can be treated in a simple manner as composed of a riser, a downcomer, a solids control valve and a gas-solids separator such as a cyclone (see Figure 2). In such a system, a pressure loop forms when the particles in the downcomer are fluidized (Kwauk et al., 1986; Yang, 1988; Rhodes and Geldart, 1987). If we assume that the static bed height in the downcomer is L_0 before fluidization (that is, when all particles are stored in the downcomer) and take the pressure at the outlet of the gas-solids separator/cyclone as zero, the pressure head at the bottom of the riser, P_r, and at the bottom of the downcomer, P_d, can be calculated by:

$$P_r = \rho_s(1-\epsilon)gH + \Delta P_c + \Delta P_{fs} + \Delta P_{ac} \tag{1}$$

$$P_d = \rho_s(1-\epsilon_{mf})gL_0 - [\rho_s(1-\epsilon)gH + \rho_s(1-\epsilon_e)gL_e]\left(\frac{D_t}{D_d}\right)^2 + \Delta P'_{ac} - \Delta P'_{fs} \tag{2}$$

where H is the riser height and L_e is the equivalent length of the exit (between riser and cyclone) section. ϵ and ϵ_e are their corresponding voidages. In Eq. 1, the first term comes from the pressure drop in the riser, and the rest are from the pressure drops across the cyclone (ΔP_c), due to solids to wall friction (ΔP_{fs}) and solids acceleration (ΔP_{ac}). In Eq. 2, the pressure head at the bottom of the downcomer, P_d, is obtained by first deducting the solids holdups in the riser and the exit pipe section from the total solids inventory (first term), and then considering the pressure drops due to solids to wall friction ($\Delta P'_{fs}$) and solids acceleration ($\Delta P'_{ac}$).

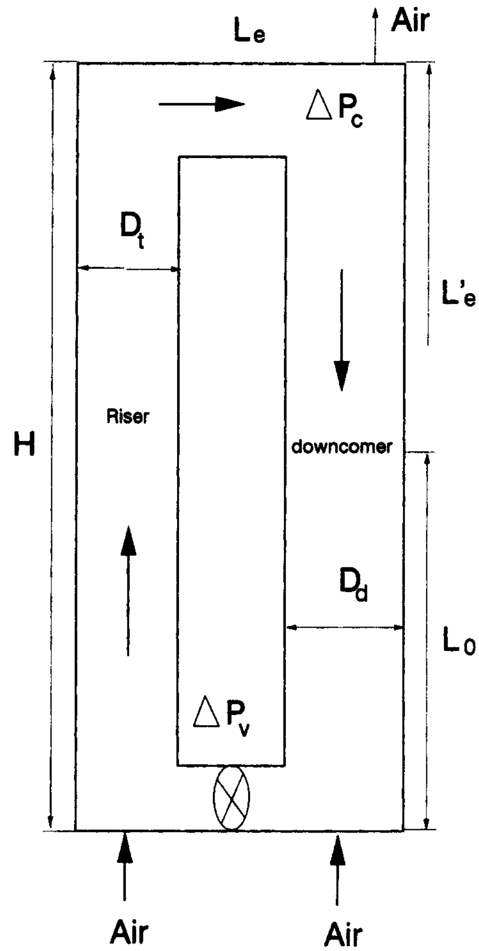

Figure 2. Typical circulating fluidized bed system.

In the riser, the pressure drops due to solids acceleration and friction are not significant at low solids circulation rates but need to be considered at high solids circulation rates. Particles can be considered to accelerate from zero velocity at the riser bottom to the fully developed velocity at the top. As a first approximation, ΔP_{ac} can be estimated by:

$$\Delta P_{ac} = 0.5\,\frac{G_s^2}{\rho_s} \tag{3}$$

Several correlations are available for particle-wall friction (see Leung, 1980). In the present analysis, the Kono-Saito equation (Kono and Saito, 1969) is adopted:

$$\Delta P_{fs} = 0.057\,g(H+L_e)\,\frac{G_s}{\sqrt{gD_t}} \tag{4}$$

The pressure drops due to solids acceleration and friction in the downcomer, $\Delta P'_{ac}$ and $\Delta P'_{fs}$, can be estimated from the similar equations.

The pressure drop across the cyclone can be approximated by:

$$\Delta P_c = \frac{1}{2}\,\zeta\rho_g U_g^2 \tag{5}$$

where ζ is the friction coefficient and is chosen as 50 (Rhodes and Geldart, 1987).

Assuming that the slip velocity is equal to the particle terminal velocity in the exit section, the voidage above the riser exit can be estimated by:

$$1-\epsilon_e=\frac{G_s}{\rho_s(U_g-U_t)} \tag{6}$$

This equation has been justified in laboratory-scale circulating fluidized bed (CFB) risers with solids circulation rate up to 600 kg/m^2•s (Kunii and Levenspiel, 1990). In commercial risers, it has been reported that the slip velocity could be higher than the terminal velocity of single particles (Matsen, 1976). Appropriate estimation of the slip velocity, however, has not been provided. Since U_g is much higher than U_t, the small difference between U_t and the actual slip velocity will not cause significant error in the calculation.

The difference between P_r and P_d is the pressure drop available for the solids flow control valve, that is:

$$\Delta P_0=P_d-P_r \tag{7}$$

For slide valves or other similar valves where the solids flow is controlled by the change of opening area, the relationship between pressure drop across the control valve and solids flux has been approximated (Jones and Davidson, 1965; Rudolph et al., 1991) as:

$$\Delta P_v=\frac{1}{2C_d^2\rho_s(1-\epsilon_{mf})}\left(\frac{G_s}{D_v/D_s}\right)^2 \tag{8}$$

where C_d is a constant ranging from 0.69 to 0.8. D_v is the equivalent diameter of the open area of the control valve. The maximum opening, D_{v0}, can be as much as the inside diameter of the particle discharge pipe, D_s.

Under steady-state operation, ΔP_v is adjusted to be equal to ΔP_0 by varying the opening area of the control valve (or the aeration air flow of nonmechanical valves) to maintain a fixed solids feed rate under a given superficial gas velocity. However, when the control valve has been completely opened (that is, D_v reaches D_{v0}), a further increase in P_r with reducing superficial gas velocity in the riser makes ΔP_0 smaller than ΔP_{v0}. The system then cannot remain at steady state at the prescribed solids circulation rate. Such a process is illustrated in Figure 3. In the figure, the solid line represents the characteristic curve of P_r, the dashed line represents the maximum available pressure head from the downcomer minus the minimum pressure drop across the valve, that is, $P_d-\Delta P_{v0}$. At gas velocities higher than U_{cr}, the pressure drop across the control valve, ΔP_v, is adjusted to meet the requirement for pressure balance in the whole loop, that is, $\Delta P_v=P_d-P_r$ to maintain a pre-set solids flux. When the gas velocity is decreased in the riser, the pressure drop across the control valve can be reduced by opening the valve to balance the system pressure. However, beyond a certain point the pressure drop across the control valve cannot be further reduced because the control valve has been completely opened. It is then conceivable that either the gas velocity needs to be raised so that the solids circulation rate can remain the same or the solids circulation rate will sharply decrease in order to maintain a steady operation at a given gas velocity. The latter case appears to correspond to the instability phenomenon observed by Takeuchi et al. (1986) and Bai et al. (1987). The purpose of this instability study is to establish a relationship between the observed instability and the operating and geometry conditions.

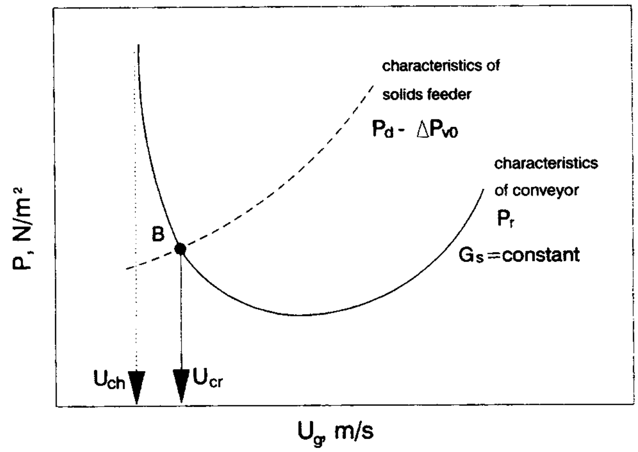

Figure 3. Operational instability due to the unbalance of pressures at the riser bottom and the downcomer.

To quantify the current instability analysis, the mean bed voidage under the critical condition is obtained by combining Eqs. 1, 2, and 7, and the criterion of $\Delta P_0=\Delta P_{v0}$ at the critical point to give:

$$1-\epsilon=\frac{(1-\epsilon_{mf})L_0-(1-\epsilon_e)L_e\left(\dfrac{D_t}{D_d}\right)^2-\dfrac{(\Delta P_c+\Delta P_{v0}+\Delta P_{fs}+\Delta P_{ac}-\Delta P'_{ac}+\Delta P'_{fs})}{\rho_s g}}{H\left\{1+\left(\dfrac{D_t}{D_d}\right)^2\right\}} \tag{9}$$

After substituting $(1-\epsilon_e)$, ΔP_c, ΔP_{v0}, ΔP_{fs}, ΔP_{ac}, $\Delta P'_{ac}$, and $\Delta P'_{fs}$ with Eqs. 3, 4, 5, 6 and 8, there are three unknowns left, that is, ϵ, G_s, and U_{cr}. Another relationship between the three unknowns is necessary for one to determine the critical superficial gas velocity, U_{cr}, for a given solids circulation rate, G_s, or to determine the critical solids circulation rate under a fixed superficial gas velocity.

Several approaches have been made to provide the relationship among bed average voidage, superficial gas velocity and solids circulation rate. Experimentally, Kato et al. (1989) and Zhang et al. (1990) gave correlations based on their own experimental data. Such correlations, however, tend to be unit dependent. A one-dimensional diffusion model was proposed by Kwauk et al. (1986). Core-annular flow structure models were proposed by Bai et al. (1988) and Senior and Brereton (1992). However, to use these models several model parameters need to be fitted experimentally due to the lack of data.

Another kind of model (Rhodes and Geldart, 1987; Yang, 1988; Bolton and Davidson, 1988; Kunii and Levenspiel, 1990) extends particle entrainment models for conventional fluidized beds to circulating fluidized beds. The only parameter needed with this approach is the decay constant, a. Since it is much simpler, the particle entrainment approach is adopted here.

The one-dimensional entrainment model of Kunii and Levenspiel (1990) is adopted in this calculation. The bed average voidage is expressed as:

$$1-\epsilon=\frac{\epsilon_e-\epsilon_d}{aH}+1-\epsilon_d-\left(1-\frac{H_d}{H}\right)(\epsilon^*-\epsilon_d) \tag{10}$$

where, H_d, the height of the bottom dense region, is given by:

$$H_d=H-\frac{1}{a}\ln\left(\frac{\epsilon^*-\epsilon_d}{\epsilon^*-\epsilon_e}\right) \tag{11}$$

Substituting Eq. 11 into Eq. 10, we have:

$$1-\epsilon=(1-\epsilon_d)+\frac{1}{aH}\left[(\epsilon_e-\epsilon_d)-(\epsilon^*-\epsilon_d)\ln\left(\frac{\epsilon^*-\epsilon_d}{\epsilon^*-\epsilon_e}\right)\right] \tag{12}$$

where ϵ^*, the saturated voidage far away from the exit of the column, can be approximated by:

$$\epsilon^*=1-\frac{G_s^*}{\rho_s(U_g-U_t)} \tag{13}$$

The saturation carrying capacity, G_s^*, is estimated using the recent correlation of Bi and Fan (1991), which is evaluated from the experimental data of Takeuchi et al. (1986), Chen et al. (1980), Drahos et al. (1988), and Bi et al. (1991), giving:

$$\frac{U_g}{\sqrt{gd_p}}=21.6\left(\frac{G_s^*}{\rho_g U_g}\right)^{0.542} Ar^{0.105} \tag{14}$$

The voidage in the bottom dense region, ϵ_d, ranges from about 0.75 to 0.85. For fine particle systems, the equation of King (1989) is used for estimation:

$$\epsilon_d=\frac{U_g+1}{U_g+2} \tag{15}$$

The decay constant, a, is an important parameter. In bubbling fluidized beds, Wen and Chen (1982) suggested a value of 4, although a range of 2 to 6 has generally been reported in the literature. Kunii and Levenspiel (1990) collected a wide range of literature data for high velocity fluidized beds and found that a varies from 0.3 to 2.5. The value tends to be a function of particle properties, but is relatively insensitive to column diameter and superficial gas velocity at high gas velocities. Rhodes and Geldart (1987) found that $a=0.5$ best fitted their experimental data obtained in a riser of 8 m height and 0.15 m diameter using fine Alumina particles. A similar value, $a=0.47$, was suggested by Bolton and Davidson (1988) for a column of 0.15 m in diameter using FCC (fluid catalytic cracking) particles. For fine particle systems, a value of $a=0.5$ is reasonable, and this is used in the present model simulation.

Once the relationship among bed average voidage, superficial gas velocity and solids circulation rate has been established, this critical point at which stable operation of the circulating fluidized bed system becomes impossible can be predicted with the following iteration procedures. For a given solids circulation rate, a critical gas velocity is first assumed. Bed average voidage is then calculated from Eq. 9 and Eq. 12, separately. If the two bed voidage values calculated from the two equations are different, a new gas velocity is assumed to repeat the calculation until the two bed voidage values calculated from both equations falls within a pre-set error range (for example, 0.001). This gas velocity is then taken as the critical velocity corresponding to a given solids circulation rate. For a given superficial gas velocity, the critical solids circulation rate can be estimated in a similar way. When both gas velocity and solids circulation rate are determined experimentally at the critical point of operation, the bed average voidage can be predicted directly from Eq. 9, together with Eqs. 3–6 and 8.

Model Verification and Predictions

Our instability analysis results are first verified against the data obtained by Hirama et al. (1992) under the so-called critical conditions. In their tests, the lowest possible superficial gas velocities (critical velocities) for the proper operation of their system were determined experimentally for several solids circulation rates. The corresponding mean solids holdups were also measured at these critical conditions. To compare, the mean solids holdup at this critical point is calculated using our study results. Column geometry parameters and operating conditions of their tests are tabulated in Table 1. The maximum open area of the valve is assumed to be the same as the size of the discharge pipe, and the inside diameter of the control valve reaches its maximum at the critical point, $D_v=D_{v0}$. The mean solids holdups at these critical points can then be readily calculated using Eqs. 1 through 9.

Table 1 shows the calculated results. Excellent agreement between the predicted and the measured bed mean voidages is obvious for the range of solids circulation rate tested, that is, 10 to 80 kg/m²·s. This further suggests that the so-called crit-

Table 1. Comparison of the Experimental Data of Hirama et al. (1992) with the Calculated Results

$D_t=100$ mm $D_d=200$ mm $D_s=50$ mm					
$H=5.5$ m $L_e=1.0$ m $C_d=0.75$					
Particles	L_0 m	G_s kg/m²·s	U_g m/s	ϵ_{exp}	ϵ_{cal}
HA54	1.1	10	1.17	0.917	0.922
$d_p=54$ μm	2.7	10	0.75	0.783	0.806
$\rho_s=750$ kg/m³					
FCC69	1.1	80	2.94	0.934	0.931
$d_p=69$ μm	2.7	80	2.19	0.835	0.813
$\rho_s=930$ kg/m³					

Table 2. Bed Parameters and Particle Properties Used by Gao et al. (1991)

$D_t = 90$ mm $D_d = 200$ mm $D_e = 50$ mm				
$H = 84$ m $L_e = 1.0$ m $C_d = 0.75$				
Particles	ρ_s kg/m³	ρ_b kg/m³	d_p μm	L_0 m
FCC	1,020	529	62	1.93,3.10,4.04
Catalyst	1,780	1,049	82	1.53,2.23,2.64
Silicagel	760	465	205	1.80,2.66,3.59

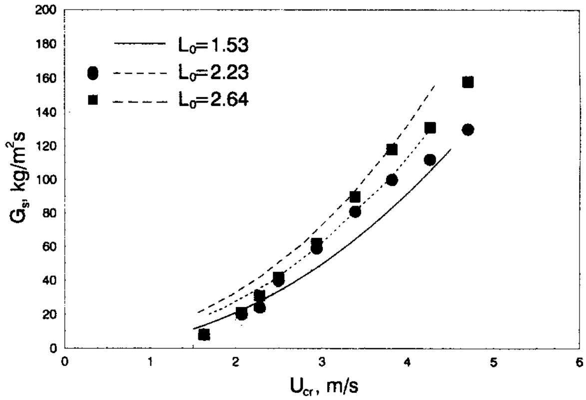

Figure 5. Comparison of model simulation with experimental data of Gao et al. (1990) for catalyst particles.

ical conditions determined experimentally by Hirama et al. (1992) result from the instability of the whole unit due to pressure imbalance. This agreement also clearly shows that our instability analysis approach predicts the occurrence of system instability very well.

This analysis is further validated through comparison with the data of Gao et al. (1991), where the static bed height in the downcomer is specified. Using a different experimental approach from that of Hirama et al. (1992), Gao et al. (1991) determined the critical velocities following the procedure of Schnitzlein and Weinstein (1988) in which a maximum solids circulation is reached at a given gas velocity when the solids control valve is completely opened. Bed structure parameters and particle properties are listed in Table 2. The mean bed solids holdups in the riser, however, were not reported. Hence, Eqs. 10 to 15 are used to estimate ϵ. Figures 4, 5, and 6 compare model prediction and experimental data. Again, good agreement is obtained for the two kinds of Group A particles used, that is, FCC and catalyst particles, and one kind of Group B particles, that is, silicagel particles.

From the foregoing analysis, it is clear that the maximum solids circulation rate corresponding to the critical operating condition in a CFB unit is strongly affected by the unit geometry. Total solids inventory in the system also affect the critical operating conditions. In light of the pressure balance between the riser and the downcomer, more particles will stay in the riser when total solids inventory is increased, as reflected in Eq. 9. To maintain the riser operating at higher solids holdup for a given gas velocity, the solids circulation rate would then be increased, as implicated in Eqs. 11 and 12. From Eq. 9, it shows that the standpipe-to-riser diameter ratio, the ratio of the static bed height in the standpipe to the riser height and the opening of the solids control valve all play very important roles. These influencing factors will be discussed based on our model simulation below.

For a typical laboratory-scale circulating fluidized bed using typical FCC particles, Figures 7, 8, 9 and 10 give a set of simulation results showing the influence of unit geometry and total solids inventory to the maximum solids circulation rate before instability occurs. For comparison, the saturation solids circulation rate from Eq. 14 and the choking velocity from the Yousfi and Gau (1974) equation, which has been found to be the most accurate correlation for Group A particles (Teo and Leung, 1984), are also plotted as dashed lines. As expected, the maximum solids circulation rate achieves higher values with increasing solids inventory, with a larger standpipe diameter or with a reduced flow resistance through the solids control valve. Figures 7, 8 and 9 also show that the solids circulation rate first increases rapidly with increasing gas velocity due to the significant gain in the saturation carrying capacity of the gas phase. However, beyond a certain gas velocity, the increase of solids circulation rate slows down, an indication that solids circulation starts to be restricted by the solids feeding system, because solids feed rate cannot catch up with the rapid increase in solids entrainment rate. In addition, pressure loss due to

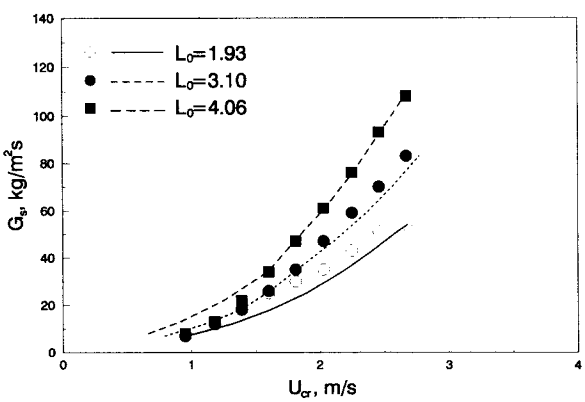

Figure 4. Comparison of model simulation with experimental data of Gao et al. (1990) for FCC particles.

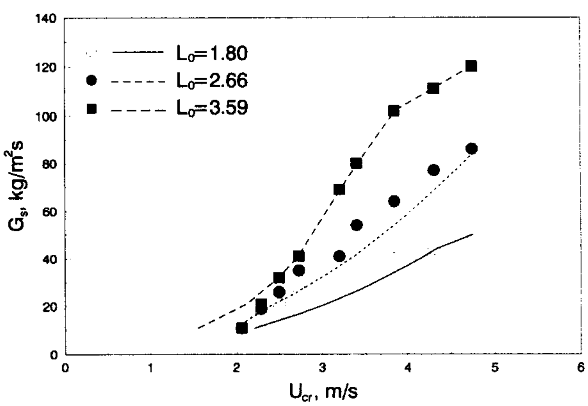

Figure 6. Comparison of model simulation with experimental data of Gao et al. (1990) for Silicagel particles.

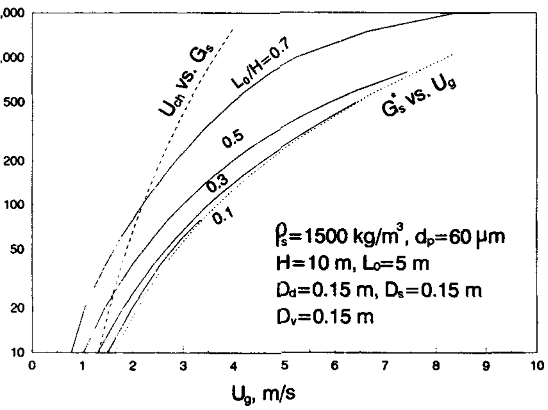

Figure 7. Effect of solids inventory on the solids circulation rate.

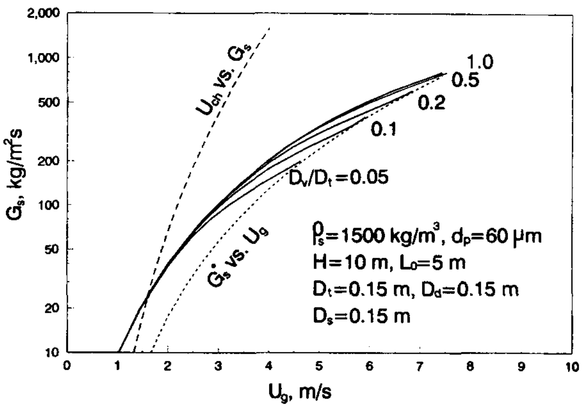

Figure 9. Effect of solids control valve on the solids circulation rate.

particle-wall friction also reduces the available pressure head for particle transportation. Eventually, the solids circulation rate curve is seen to merge with the saturation velocity curve. The solids feeding system is thus more important when the unit is operated in a high gas velocity range. In this range, increasing the gas velocity can only slightly increase solids circulation rate because solids circulation rate is no longer sensitive to the variation of gas velocity. To achieve higher solids circulation rate, a better measure is to increase the pressure available for solids feeding by adding more particles to the downcomer and/or reducing the pressure loss through the solids control valve and the gas-solids separator.

For the effect of solids control valve, Figure 9 shows that solids circulation rate is less affected by the control device at low solids circulation rate. At a high solids circulation rate, however, the control valve provides an important regulation function. Solids circulation rate is less influenced when the opening area of the valve is adjusted from 50% to 100% due to low pressure losses in this range. Therefore, from the proper regulation point of view, the open ratio of the valve needs to be kept under 50% to have a significant fraction of pressure drop across it. The maximum available solids circulation rate is correspondingly lower than expected with the control valve fully opened.

Figure 10 demonstrates the effect of pressure on the critical velocity in which the variation of decay constant and dense phase voidage with pressure is neglected. It is seen that the critical velocity decreases with increasing pressure, in agreement with the experimental results of Knowlton and Bachovchin (1976). It can be seen that such a decrease is mainly due to the increase of saturated entrainment rate with the increase of pressure as observed by Chan and Knowlton (1983). A more accurate evaluation of such effect still needs the information of the effects of pressure on the decay constant and dense phase voidage.

It should be noted that the accuracy of our prediction depends on the accuracy of the equations adopted in the calculation, that is, Eqs. 3-6 and Eqs. 10-15, some of which may not hold in larger scale commercial systems. More accurate correlations can be incorporated into the calculation when they become available in the future.

Concept of the High-Density Riser

An important application of the present analysis is to offer guidance for achieving high capacity of circulating fluidized beds by increasing the solids holdup and solids throughput in the riser. While circulating fluidized bed systems have been applied to many industrial processes including fluid catalytic cracking and combustion, some of the key operating conditions for FCC riser reactors and CFB combustors are dramatically different. As shown in Table 3, FCC risers operate at much

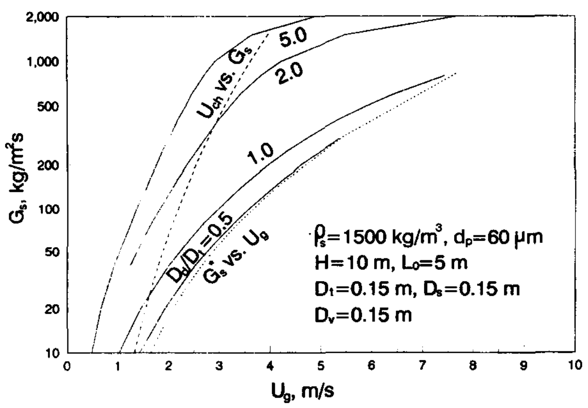

Figure 8. Effect of standpipe size on the solids circulation rate.

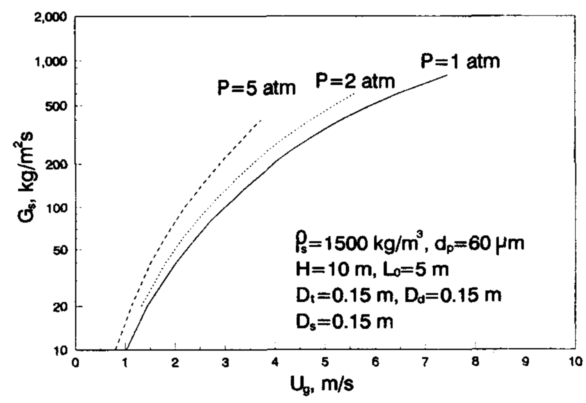

Figure 10. Effect of pressure on the solids circulation rate.

Table 3. Key Differences in Operating Conditions Between FCC Riser Reactor and CFB Combustor

Operating Conditions	FCC Riser Reactor	CFB Combustor
Particle Density	1,100–1,700 kg/m^3	1,800–2,600 kg/m^3
Superficial Gas Velocity	8–30 m/s (increasing with height)	5–9 m/s
Net Solids Flux	400–900 kg/m^2•s	10–100 kg/m^2•s
Apparent Suspension Density in the Developed Region	50–120 kg/m^3	10–40 kg/m^3

higher solids circulation rate and with a higher solids holdup (apparent density) in the riser. However, a recent survey conducted by the authors (Zhu and Bi, 1993) has shown that almost all reported circulating fluidized bed studies have been conducted with solids circulation rate below 200 kg/m^2•s, compared with 400–900 kg/m^2•s in FCC riser reactors. Hydrodynamic studies in risers with higher solids throughput is clearly lagging behind the industry need.

To make researchers aware of this situation, it is very useful to distinguish the concept of low-density riser or low-density circulating fluidized bed (LDCFB) corresponding to the operating conditions in a CFB combustor with the concept of high-density riser or high-density circulating fluidized bed (HDCFB) corresponding to the operating conditions in FCC riser reactors. A systematic study on hydrodynamics of a high-density riser will lead to a much better understanding of the existing FCC systems and to improvement in future design. In addition, the development of the high-density riser will also allow much higher solids holdup to be achieved in the riser (for example, more than 10%). Such high solids holdup is required for some catalytic processes in the petroleum industry where a catalytic/oil ratio of the order of 20 is needed compared to 5–9 in FCC processes. To derive optimal utilization of limited sources of crude oil, it is of great interest to develop the high-density risers in which a voidage of around 0.85 to 0.95 and a solids throughput as high as 1,000 kg/m^2•s are possible. A sound fundamental and theoretical study would provide a very valuable source of information for further application of the high-density risers in the petroleum and related industries. Such a high-density riser may also be useful in other processes to increase productivity.

As discussed above, almost all studies reported have been on low-density risers, mainly due to the restriction on the solids feeding system. From the riser-blower interaction point of view, a stable dense-phase transport operation mode is attainable for the fine Geldart Group A particles, typically FCC particles, if a sufficient pressure head can be provided to support the conveying line (Zenz, 1949). In light of the present analysis, it is further required that sufficient pressure head be provided to the downcomer side for pushing particles through the solids feeding system into the riser. From a pressure balance point of view (Figure 11), high pressure head can be provided by storing more particles in a large standpipe, utilizing large standpipe-riser diameter ratio and reducing the resistance of the solids control valve as much as possible. An alternative method is to superimpose pressure on the standpipe. This concept was utilized in several studies such as those of Yousfi and Gau (1974) and Lu (1988), in which fluidized bed was used as the solids feeder and the pressure in the fluidized bed was adjusted by means of an exhaust valve. Another approach is to break up the undesirable pressure balance in the whole unit. Using screw feeder, Drahos et al. (1988), Mori et al. (1991) and Hirama et al. (1992) showed that the solids circulation rate can be independently adjusted to achieve higher solids holdup in the small-scale laboratory units because the pressure balance between the riser and standpipe is circumvented. For industrial applications, however, a screw type solids feeder could not supply sufficient solids to meet the requirement of a large-scale commercial unit. Also, the fluidized bed type feeder or so-called internal circulating bed (Fusey et al., 1986) is only applicable to small units. The standpipe is, therefore, the only plausible feeder of industrial interest and worth extensive investigation.

From the above analysis, it is clear that the design of a high-density riser/circulating fluidized bed (HDCFB) needs to consider three important factors, that is, standpipe structure, the solids discharge pipe and control device, and total solids inventory in the system. Other hydrodynamic factors can also influence the maximum available solids flux and bed density. To achieve extremely high solids flux, a practical design concept for laboratory scale HDCFB is proposed here as depicted in Figure 12. The pressure loop in the CFB is terminated after the solids separation device, and separated particles are carried up to the standpipe through a second riser with a diameter much larger than the first riser diameter to ensure it operates in the dilute transport mode. In this way, a standpipe much higher than the (first) riser can be built to create enough back pressure for the operation of a high density riser. Since the second riser is much larger and will be operated at more dilute conditions, the pressure loss through this section is expected to be small. The solids feeding system and the gas-solids sep-

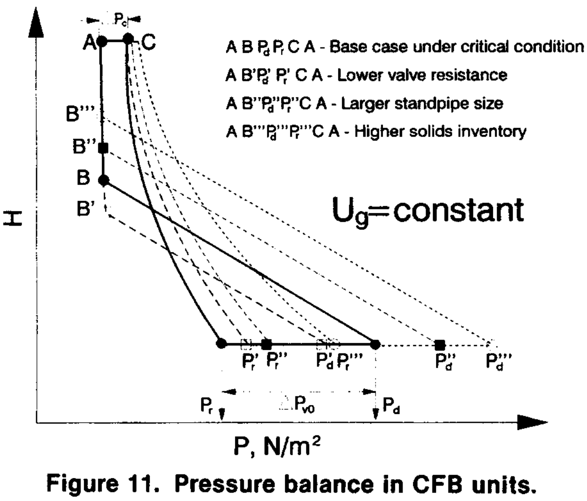

Figure 11. Pressure balance in CFB units.

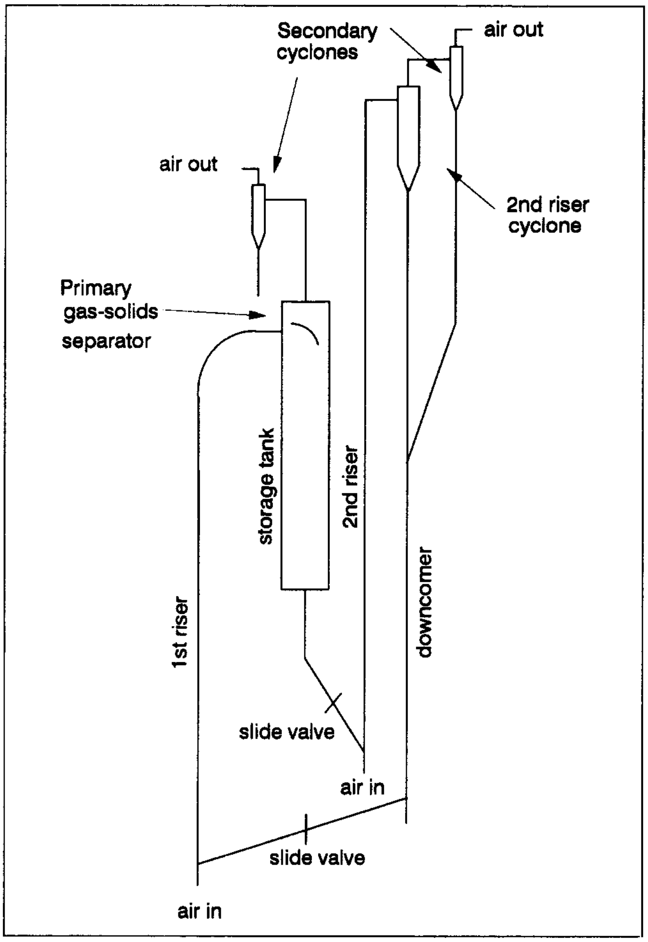

Figure 12. Concept of high-density circulating fluidized beds/risers.

aration system should also be designed in such a way that the pressure losses be minimized. Such a prototype high-density circulating fluidized bed is now under construction in the University of British Columbia for initial tests.

Conclusion

An instability analysis of the circulating fluidized bed system based on the pressure balance shows that the unstable operations at the critical superficial gas velocity and beyond the maximum solids circulation rates reported in some circulating fluidized bed systems are actually the same phenomenon, resulting from pressure imbalance in the whole unit. The consequence of this instability is then that the prescribed solids circulation rate cannot be maintained. Such an instability study successfully predicts the critical conditions reported by Hirama et al. (1992) and the maximum solids flux points cited by Gao et al. (1991). Model simulation further shows that the operating instability and the maximum solids circulation rate attainable are strongly influenced by the total solids inventory and the unit geometry such as standpipe size and solids feeding device. To achieve high-density and high solids flux operation, sufficient back pressure needs to be provided at the bottom of the riser, by increasing the pressure buildup in the standpipe and/or by lowering the resistance through the solids control valve. Based on the above analysis, a concept of high-density riser/circulating fluidized bed (HDCFB) is proposed and a schematic setup for the proposed HDCFB is given.

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Notation

a = decay constant, 1/m
Ar = Archimedes number
C_d = constant as defined in Eq. 8
d_p = mean particle size, m
D_d = standpipe diameter, m
D_s = diameter of solids discharge pipe, m
D_t = riser diameter, m
D_v = equivalent diameter of the open area of the control valve, m
D_{v0} = maximum equivalent diameter of the control valve, m
G_s = solids circulation rate, $kg/m^2 \cdot s$
G_s^* = saturation carrying capacity, $kg/m^2 \cdot s$
H = riser height, m
H_d = height of the bottom dense region in the riser, m
L_e = equivalent length of horizontal pipe between the riser exit and the cyclone, m
L_0 = static bed height in the standpipe, m
P_d = pressure head at the bottom of the standpipe, N/m^2
P_r = pressure head at the bottom of the riser, N/m^2
ΔP_{ac} = pressure loss due to solids acceleration in the riser, N/m^2
$\Delta P'_{ac}$ = pressure loss due to solids acceleration in the downcomer, N/m^2
ΔP_c = pressure loss through the cyclone or other gas-solid separators, N/m^2
ΔP_{fs} = pressure loss due to solids to wall friction in the riser, N/m^2
$\Delta P'_{fs}$ = pressure loss due to solids to wall friction in the downcomer, N/m^2
ΔP_v = pressure loss through the solids control valve, N/m^2
ΔP_{v0} = pressure loss through the solids control valve when the valve is in fully open position, N/m^2
ΔP_0 = the driving force available to push particles from the downcomer to the riser, as defined in Eq. 7, N/m^2
Re_t = Reynolds number at particle terminal velocity, $\rho_g U_t d_p/\mu_g$
U_{cr} = critical superficial gas velocity, m/s
U_g = superficial gas velocity, m/s
U_t = terminal velocity of single particles, m/s

Greek letters

ϵ = average bed voidage
ϵ_d = voidage at the bottom dense region of the riser
ϵ_e = voidage at the exit of the riser
ϵ_{mf} = voidage at minimum fluidization
ϵ^* = voidage at saturation condition
μ_g = viscosity of the gas, N/m
ρ_g = density of the gas, kg/m^3
ρ_s = density of the particles, kg/m^3
ζ = friction coefficient of cyclone

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