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THE GAS–SOLID–SOLID PACKED CONTACTOR: HYDRODYNAMIC BEHAVIOUR OF COUNTER-CURRENT TRICKLE FLOW OF COARSE AND DENSE PARTICLES WITH A SUSPENSION OF FINE PARTICLES

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Abstract—The development of a new recovery system involving direct contact of a dilute suspension of fine particles with a rain of coarse and dense particles in a packed column is described. The experiments are complemented by an analysis of the hydrodynamic behaviour. The experimental results show that the behaviour of a dilute multiphase system may be described by two or more diphase systems.

Key Words: multiphase system, raining particles contactor, dilute suspension, packed column, pressure profile, solids hold-up

1. INTRODUCTION

The gas–solid–solid packed contactor (GSSPC) was developed for the thermal treatment of sand wastes from foundries. However, it also has applications in a number of processes involving chemical reactions, such as the regeneration of catalysts and the recovery of hydrocarbonaceous fluids from particulate solids (Benali 1989).

Early studies on the circulation of solid particles in a packed reactor were performed by Sutherland *et al.* (1963), Sutherland & Wong (1964), Park *et al.* (1968), Capes & McIlhney (1968, 1972) and Claus (1974). The latter authors examined the effect of fixed packing on the hydrodynamic properties of a gas fluidized bed and its prospects for industrial use. Sutherland *et al.* (1963) showed that the presence of packing in a fluidized bed reduces the pressure drop to a value which is linked to the ratio of the particle diameter to the packing diameter (d_p/d_{pc}). Moreover, in a hydrodynamic analysis of a packed fluidized bed (PFB), Claus (1974) demonstrated the existence of two flow regimes, *the dense regime* (longer residence time) and *the dilute regime* (shorter residence time). He also observed a remarkable contribution of the packing in retaining solids in the contact section. Fundamentally, the PFB has a similar behaviour to the gas–liquid systems studied by Richardson & Zaki (1954). In recent developments of the raining packed bed exchanger (RPBE) used as a direct contact system, Tiboutine (1981) observed two regimes, *the dispersed* and *segregated regimes*. This author has also observed that the overall pressure drops show a monotonic increase with the gas superficial velocity (V_f). So, all the plotted curves exhibit a small characteristic "hump" over the entire range of $V_f = 1.6$ to 3.4 m/s. More recently, Westerterp & Kuczynski (1987) carried out a gas–solid trickle flow study in a packed column at relatively low velocities and proposed a two-phase model based on the concept of "trickle void porosity". They assumed that Ergun's equation remained valid for the prediction of the pressure drop in the packed section. Whilst substantial progress has been achieved in gas–solid trickle flow, there remains the problem that this assumption cannot describe the intimate gas–solid contacting. Indeed, Molodstof (1985) and Benali (1989) have shown that Ergun's equation can be used only in a quiescent medium such as a plug moving flow, a homogeneous fluidized bed or a sedimentation, where the relative positions

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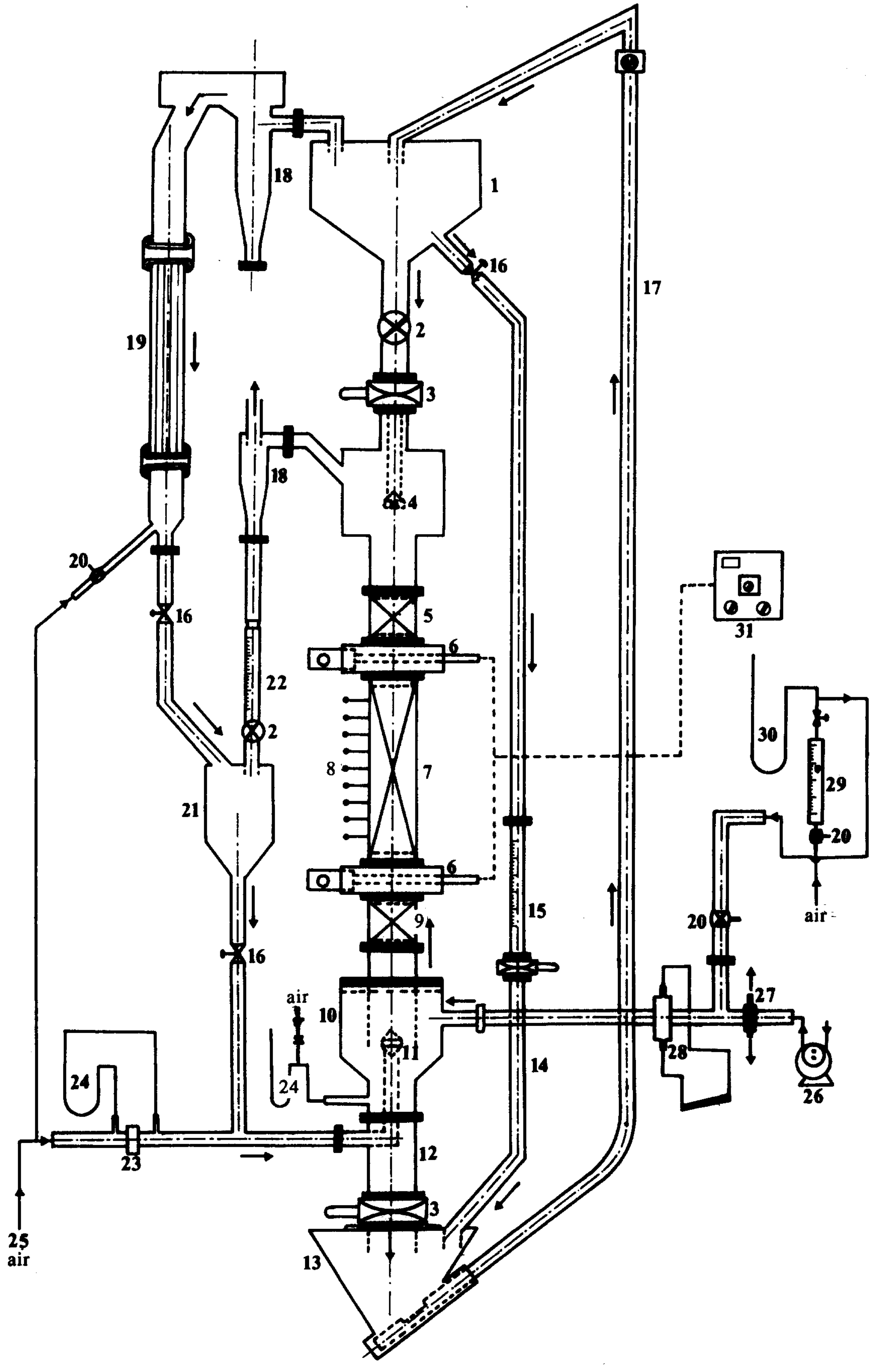

Figure 1—*caption opposite.*

Table 1. Properties of the packing materials

d_{pc} (mm)	25
δ_{pc} (mm)	0.62
h_{pc} (mm)	25
ϵ_{pc} (—)	0.848

Table 2. Physical properties of the solids

Type of solids	d (μm)	ρ (kg/m^3)	Class[a]
Sand (fine particles)	179	2646	A
Zirconia (coarse and dense particles)	1320	3774	B

[a]From Geldart's (1973) classification of powders.

Table 3. Range of operating conditions

Parameters	Air–sand	Air–zirconia	Air–sand–zirconia
V_f (m/s)	2.10–8.44	2.44–8.63	3.40–9.58
W_s (kg/s)	0.030–0.120	—	0.058–0.120
W_z (kg/s)	—	0.070–0.254	0.068–0.260
$10^4\langle\alpha_s\rangle$ (—)	10.70–64.46	—	25.78–178.08
$10^3\langle\alpha_z\rangle$ (—)	—	7.44–74.64	9.46–170.09

of the particles remain unchanged. In gas–solid suspension flows, such as those in pneumatic conveying and gas–solid trickle flow reactors, each phase is fundamentally random in nature. Otherwise, this system involves an entrainment of the particles at high gas superficial velocities ($V_f > 4$ m/s). Because of this limitation and the resurgence of interest in gas–solid trickle flow reactors operated at high gas velocities (4–10 m/s), another configuration of these contactors has to be developed.

The purpose of this paper is to present the concept of the GSSPC and to analyse its hydrodynamic behaviour by means of measurement of the pressure drops and the average dynamic hold-up of the solids in the packed section.

2. EXPERIMENTAL-SCALE UNIT AND PROCEDURES

A schematic of the experimental equipment is illustrated in figure 1. It consists of a plexiglass packed test section with an internal diameter and length equal to 114 and 700 mm, respectively. Cylindrical Pall rings (table 1) are used as packing and they are placed in a regular arrangement. A close-circuit system for coarse solids circulation is employed. The principle of operation is based on the vertical pneumatic transport of fine particles, a trickle flow of coarse and dense particles and on gas–solids separation. The fine particles are introduced into the packed section by means of a conical injector system. The coarse and dense particles are transported using a screw elevator. The flow of the feed solids can be adjusted by changing the aperture of the sleeve valve below the hopper. Two cyclones are used for gas–solids separation. The remaining fines are retained by a bag filter and are recycled in the circulation system of solids. A fan is employed as the air feeder whose flowrate is measured by means of an orifice and is regulated by a "Vari-Phi" pulley. In order to prevent short-circuiting of the gas through a return line of coarse and dense particles, a gas-lock is installed at the bottom of the standpipe. The average dynamic hold-up of solids by volume unit is determined using the weight of solids recovered following the closing of the slide valves. The distance between the two slide valves is 700 mm. The average dynamic hold-up of solids is given as follows:

$$\langle\alpha_i\rangle = \frac{4m_i}{\pi\rho_i D^2 L(\epsilon_{pc} - \langle\alpha_j\rangle)}, \quad i \neq j; \qquad [1]$$

(Figure 1 Opposite)

Figure 1. Schematic view of the GSSPC: **1,** feed hopper of coarse particles; **2,** pneumatic sleeve valve; **3,** orifice valve; **4,** conical distributor; **5,** packed section; **6,** pneumatic slide valves; **7,** packed test section; **8,** water manometers; **9,** packed section; **10,** gas injection chamber; **11,** fine particles injector; **12,** measurement system of coarse particles; **13,** collect hopper; **14,** return-line; **15,** graduated Pyrex column for measurement of the coarse particles flowrate; **16,** manual sleeve valve; **17,** elevator; **18,** cyclone; **19,** filter bag; **20,** ball valve; **21,** feed hopper of fine particles; **22,** graduated Pyrex column for measurement of the fine particles flowrate; **23,** diaphragm; **24,** U-water manometer; **25,** compressed air; **26,** fan; **27,** induction valve; **28,** diaphragm; **29,** rotameter; **30,** U-mercury manometer; **31,** electric control system.

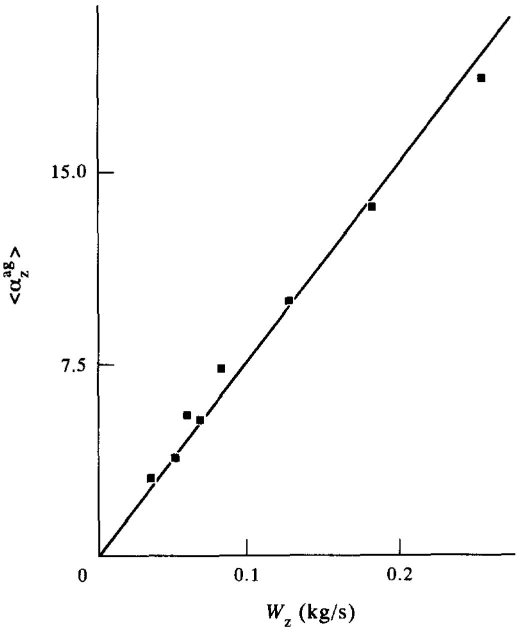

Figure 2. Average dynamic hold-up of zirconia particles (without gas) as a function of the mass flowrate of solids.

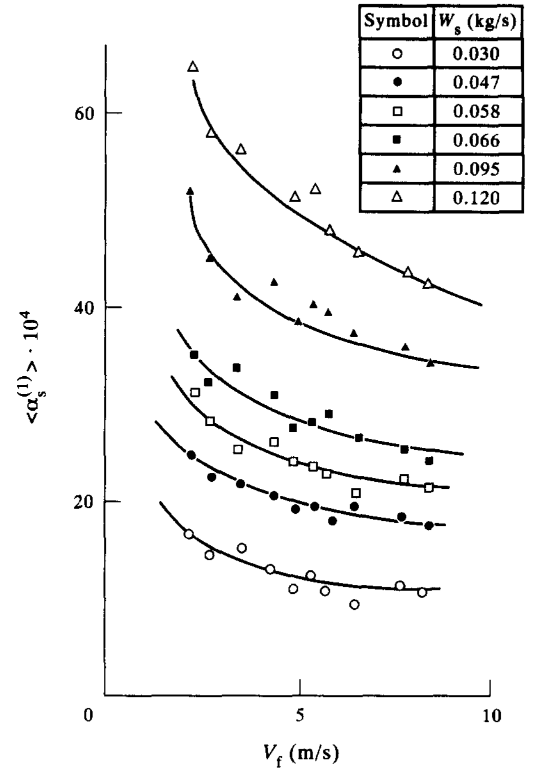

Figure 3. Average dynamic hold-up of sand particles as a function of the gas superficial velocity; *air–sand system.*

where i and j, represent different solid species (e.g. sand, zirconia etc.), such as $\langle\alpha_i\rangle + \langle\alpha_j\rangle + \epsilon_{pc} = 1$, m_i is the mass of solids retained between the two slide valves, ρ_i is the density of the different solid species, D is the internal diameter of the column, L is the distance between the two slide valves and ϵ_{pc} is the porosity of the packing.

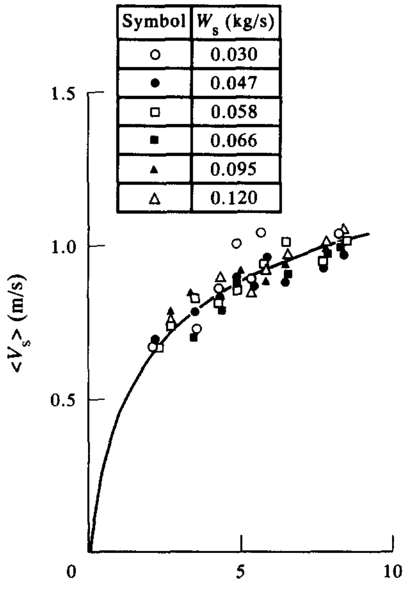

Figure 4. Mean velocity of sand particles as a function of the gas superficial velocity; *air–sand system.*

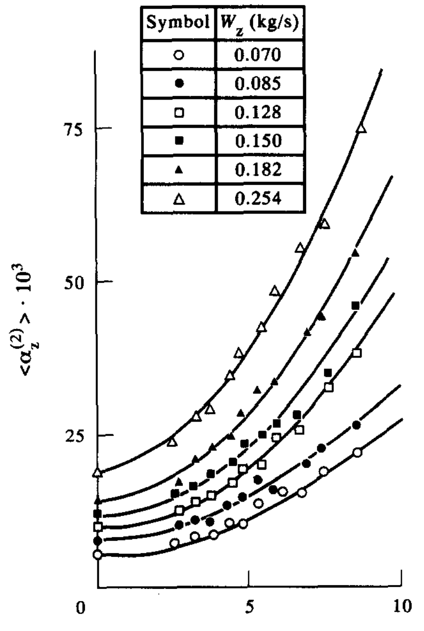

Figure 5. Average dynamic hold-up of zirconia particles as a function of the gas superficial velocity; *air–zirconia system.*

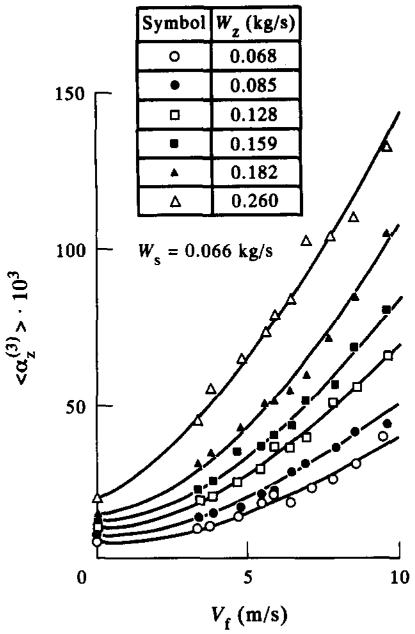

Figure 6. Average dynamic hold-up of zirconia particles as a function of the gas superficial velocity; *air–sand–zirconia system.*

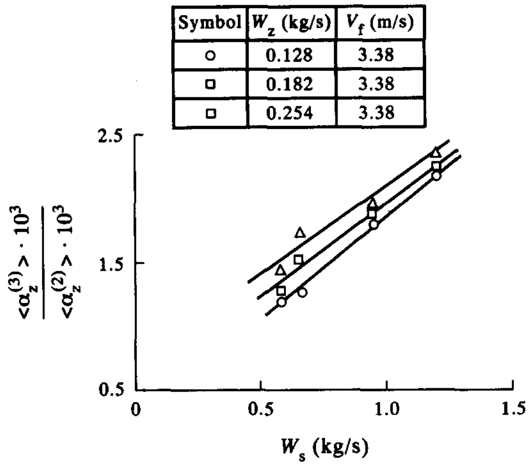

Figure 7. Contribution of sand particles to supporting the zirconia particles; *air–sand–zirconia system.*

In addition, the particle-size distribution of the average dynamic hold-up between the two slide valves is measured during the hydrodynamic experiments. The pressure drops are measured by inclined water manometers which are connected to the external walls of the column. The physical properties of the solids are specified in table 2. The range of the chosen operating conditions is given in table 3.

3. EXPERIMENTAL RESULTS AND DISCUSSION

The mean particle velocity (V_i, where i = s (sand) or z (zirconia)), the actual gas velocity ($V_{\mathrm{f,act}}$) and the relative velocity between the gas and each solid phase (U_{rel}) are deduced from the average values of the solids hold-up using the following equations:

$$\langle V_{\mathrm{f,act}} \rangle = \frac{V_{\mathrm{f}}}{\epsilon_{\mathrm{pc}} - \langle \alpha_{\mathrm{s}} \rangle - \langle \alpha_{\mathrm{z}} \rangle}, \qquad [2]$$

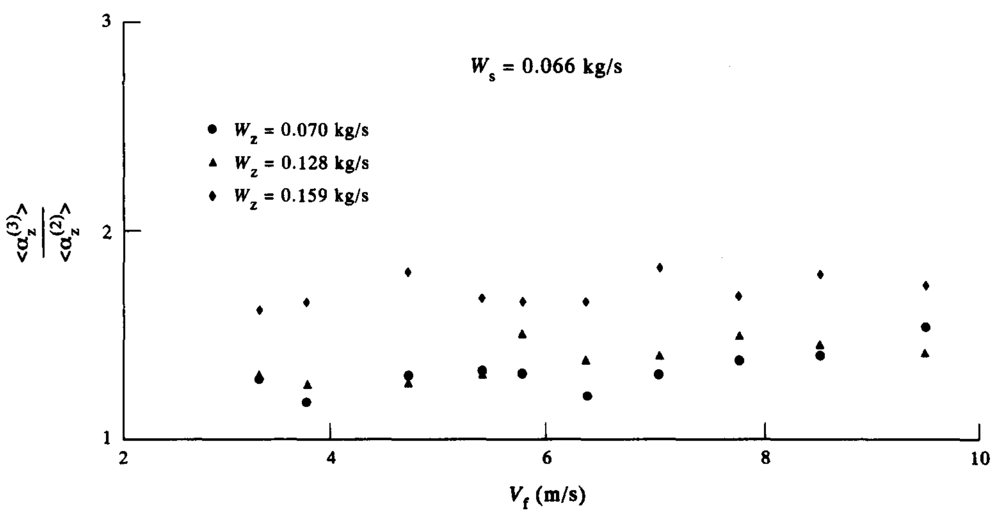

Figure 8. Contribution of sand particles to supporting the zirconia particles: effect of the gas superficial velocity; *air–sand–zirconia system.*

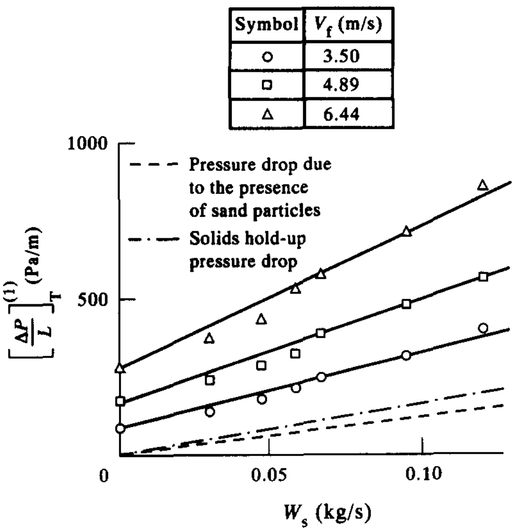

Figure 9. Overall pressure drop as a function of the mass flowrate of sand particles; *air–sand system.*

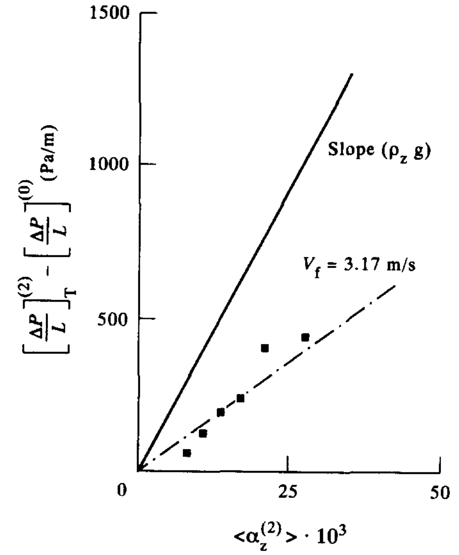

Figure 10. Pressure drop due to the presence of zirconia particles as a function of the average dynamic hold-up of zirconia particles; *air–zirconia system.*

$$\langle V_i \rangle = \frac{W_i}{\rho_i A_c \langle \alpha_i \rangle}, \quad i = \mathrm{s}, \mathrm{z}, \tag{3}$$

$$\langle U_{\mathrm{rel,z}} \rangle = \langle V_{\mathrm{f,act}} \rangle - \langle V_{\mathrm{z}} \rangle \tag{4}$$

and

$$\langle U_{\mathrm{rel,s}} \rangle = \langle V_{\mathrm{f,act}} \rangle - \langle V_{\mathrm{s}} \rangle. \tag{5}$$

3.1. Average Dynamic Hold-up of Solids

3.1.1. Trickle flow of zirconia particles in the absence of gas

Figure 2 shows the variation of the average dynamic hold-up of zirconia particles ($\langle \alpha_z^{ag} \rangle$) as a function of the mass flowrate of solids (W_z). This curve exhibits a linear relationship between the

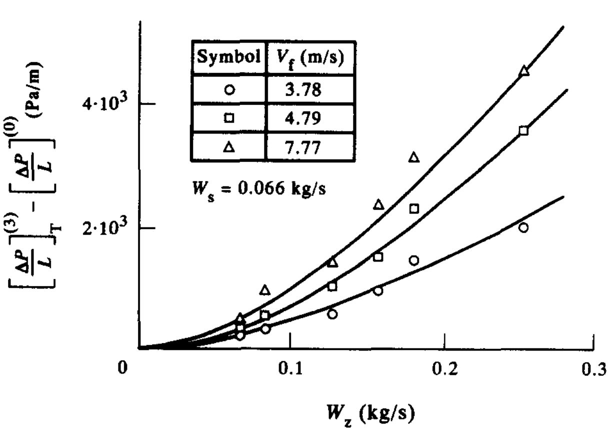

Figure 11. Pressure drop due to the presence of solids as a function of the mass flowrate of zirconia particles; *air–sand–zirconia system.*

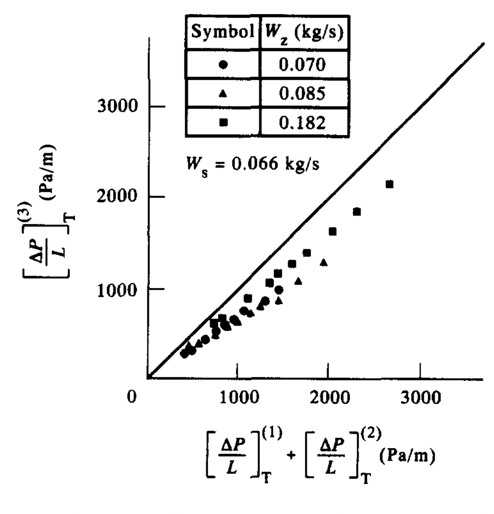

Figure 12. Comparison between the pressure drop measured in the air–sand–zirconia system and the sum of the pressure drops given by the air–sand and air–zirconia systems.

two variables, i.e. $\langle \alpha_z^{ag} \rangle$ is proportional to the mass flowrate of zirconia particles. Therefore, the mean velocity of the particles across the packed section is directly proportional to the ratio and, subsequently, remains constant.

3.1.2. Co-current upward flow of sand particles and a gas

Typical variations of the measured average dynamic hold-up of sand particles ($\langle \alpha_s \rangle$) and the deduced mean solids velocity (V_s) as a function of the gas superficial velocity (V_f) for different mass flowrates of sand (W_s) are summarized in figures 3 and 4, respectively. As shown in figure 3, the average dynamic hold-up decreases with increasing gas superficial velocity, and increases with increasing mass flowrate of sand particles. This is an expected result since the average dynamic hold-up, at given solid flowrate, is inversely proportional to the solids velocity, which should increase with increasing gas superficial velocity.

The same data are plotted as mean sand particles velocity as a function of gas superficial velocity in figure 4. These results show an increase in the mean sand particle velocity as the gas superficial velocity is increased. Also note that the average relative velocity between the gas and solids increases with increasing gas superficial velocity.

3.1.3. Counter-current trickle flow of zirconia particles and a gas

The curves presented in figure 5 show the increase in the average dynamic hold-up of zirconia particles with the gas superficial velocity and the mass flowrate of particles. At low gas superficial velocity ($0 \leqslant V_f \leqslant 2.5$ m/s), the variation in the average dynamic hold-up is negligible, viz. the contribution of the gas phase to breaking down the particle trickles is negligible in this range of the gas superficial velocity.

3.1.4. Counter-current trickle flow of zirconia particles and a suspension of fine particles

The trends in the variation of the measured average dynamic hold-up of zirconia particles are modified by the presence of the sand particles in the upward gas stream (figure 6). This means that the zirconia particles trapped in the packed section are not only supported by the gas and the packing but also by the sand particles. Accordingly, it is clear that the presence of the sand increases

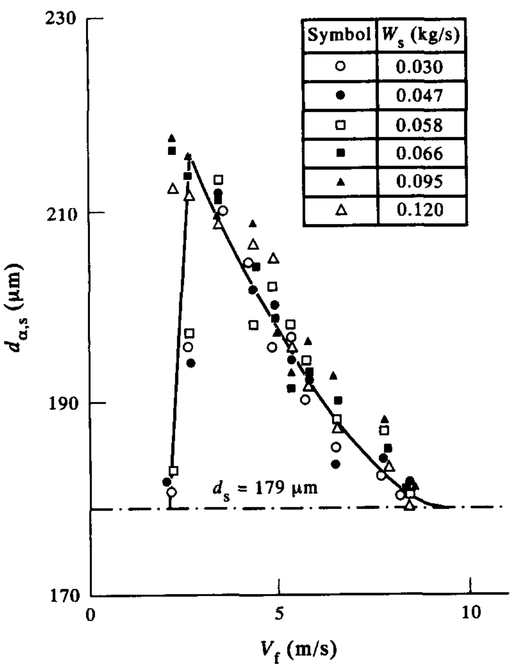

Figure 13. Mean diameter of the dynamic hold-up of sand particles as a function of the gas superficial velocity; *air–sand system.*

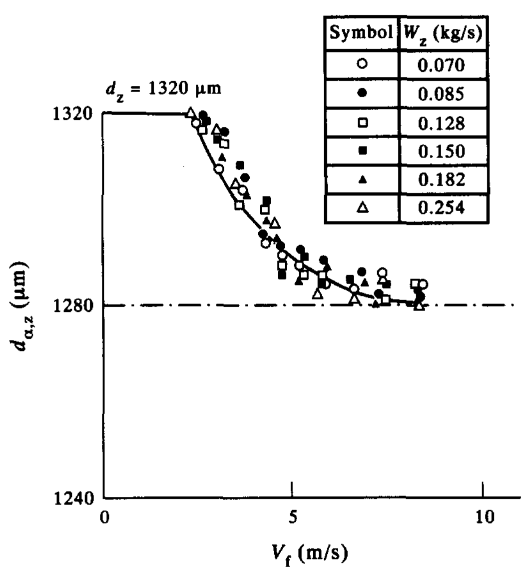

Figure 14. Mean diameter of the dynamic hold-up of zirconia particles as a function of the gas superficial velocity; *air–zirconia system.*

the pressure gradient in the gas, which can then support the zirconia particles (figure 7). This result is summarized by the following expression:

$$\frac{\langle \alpha \rangle_z^{(3)}}{\langle \alpha \rangle_z^{(2)}} = \langle \alpha \rangle_z^{(2)} + 17.45 W_z^{-0.16} W_s. \qquad [6]$$

Equation [6] shows that the average dynamic hold-up of solids particles in an air–sand–zirconia system [indicated by superscript (3)] can be predicted satisfactorily by means of measurements carried out in an air–zirconia system [indicated by superscript (2)]. However, figure 8 shows that the gas superficial velocity does not affect the values of the hold-up ratio.

3.2. Pressure Drops Along the Packed Section

3.2.1. Co-current upward flow of sand particles and a gas

The variations in the overall pressure drop per unit height of packing with the mass flowrate of sand particles are illustrated in figure 9. All profiles increase with increasing mass flowrate of sand and their trends can be described by a linear function following the form

$$\left[\frac{\Delta P}{L}\right]_T^{(1)} = aW_s + b, \qquad [7]$$

where a and b vary with the gas superficial velocity. The overall curves have a positive slope and their intercept corresponds well with the pressure drop of a gas flowing at the same superficial velocity without the solids.

A remarkable aspect of the data in figure 9 is the fact that at low gas velocities (2.10–2.70 m/s) the pressure drop due to the presence of sand particles is less than the solids hold-up pressure drop ($\rho_s \alpha_s g$), viz. the dynamic hold-up of sand particles is solely responsible for the difference between the suspension pressure drop and that due to the upward gas flow in the packed column without the solids. This observation demonstrates that the particles only lean against the packing and column walls and subsequently they slide down the walls. These results may be compared with those obtained in vertical conveying by Capes & Nakaruma (1973), who measured the shear stress between the solid particles and the internal walls of the transport line: these authors suggest that the particles are not only sliding down but are also recirculating at the pipe walls. This finding agrees with the experimental results obtained by Muzyka (1985) with sand particles ($d_p = 172$ and 242 μm) and those obtained by Yousfi & Gau (1974) with FCC particles ($d_p = 20$ and 50 μm). This effect at low gas velocities is also confirmed in figure 12, where the mean diameter of solids hold-up increases with the gas superficial velocity while it decreases beyond the relatively higher gas velocities ($V_f > 2.70$ m/s), viz. the mean velocity of the fraction of fine particles of the solids hold-up is higher than that of the fraction of coarse particles, which is lower in proportion to their presence in the sand particles feed.

3.2.2. Counter-current trickle flow of zirconia particles and a gas

The measured pressure profiles have similar characteristics to those obtained above. However, the pressure drop of zirconia particles is lower than the weight of the trapped solids (figure 10). This implies that part of the weight of zirconia particles is supported by the packing elements.

3.2.3. Counter-current trickle flow of zirconia particles and a suspension of fine particles

Figure 11 shows similar trends to those obtained in the previous sections. Therefore, the presence of sand particles in the upward gas induces an additional pressure drop due to their collisions with the packing element walls and zirconia particles.

The results in figure 12 add to the evidence presented above, suggesting that the laws governing the overall suspension are dependent on the particle–particle interaction: the total pressure drop is less than that given by adding the pressure drops of each of the two-phase systems (air–sand and air–zirconia).

3.3. Particle-size Distributions of the Solids Hold-up

Figure 13 shows that the mean diameter of the dynamic hold-up of sand particles decreases as the gas velocity increases, approaching the mean diameter ($d_s = 179$ μm) of the solids circulating in the system at high gas velocity. At low gas superficial velocity ($2.10 < V_f < 270$ m/s), a deviation from this behaviour is observed. This result is due to the fact that the largest fraction of sand particles is not conveyed upward by the gas stream.

Figure 14 shows a decrease in the mean diameter of the dynamic hold-up of zirconia particles with increasing gas superficial velocity. These results also demonstrate the existence of a dispersion of the particle velocity of each size distribution class around the mean particle velocity. Moreover, both figures 13 and 14 show that the mean diameter of the dynamic hold-up of solids is not affected by the mass flowrate of solids.

4. CONCLUSION

The concept of raining coarse and dense particles in counter-flow to entrained fine particles is developed as a means of contact heat exchange between solid phases. It is apparent that the presence of fine particles in the upward air stream plays an important role in this system.

This study presents important results regarding the hydrodynamic characterization of dilute suspensions. It is found that the behaviour of a dilute tri-phase system can be predicted satisfactorily by two-phase systems.

It is clear, from the above discussion, that experimental observations will play a central role in the analysis of the hydrodynamic characteristics of the direct contact mechanism.

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