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GAS HOLDUP AND PRESSURE DROP IN THREE-PHASE VERTICAL FLOWS OF GAS-LIQUID-FINE SOLID PARTICLES SYSTEM

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Key Words: Multiphase Flow, Gas Holdup, Pressure Drop, Vertical Upflow, Vertical Downflow, Slurry

To obtain information on the hydrodynamics of gas-liquid-fine solid particles flow systems, gas holdup and pressure drop in vertical upflow and downflow tubes were measured at comparatively high fluid velocities.

The following experimental results were obtained.

1) Within the range of experimental conditions, gas holdups in vertical upflow tubes are independent of tube diameter, average size and concentration of solid particles.

2) Frictional pressure drops in vertical upflow tubes are independent of the average size of solid particles, but increase with the concentration of solid particles.

3) Gas holdup in vertical downflow tubes, except at low gas and high slurry velocities, are independent of tube diameter, average size and concentration of solid particles.

4) Frictional pressure drops in vertical downflow tubes are independent of the average size of solid particles, but increase with the concentration of solid particles.

Introduction

Numerous studies have been conducted on the hydrodynamics of gas-liquid-solid flow, both in bubble columns with suspended solid particles and in three-phase fluidized beds. In these cases, comparatively low fluid vertical upflow velocities are examined. However, in preheater and transportation pipe lines for the coal liquefaction process comparatively high fluid velocities exist in a gas-liquid-solid system. Little information is available on such systems due to the complex properties of the flow and the difficulty in obtaining experimental results.[9] For the coal liquefaction process in particular a better understanding of the three-phase hydrodynamics could lead to improved predictions of the coal dissolution rate.[2]

In the present work, measurements of gas holdup and pressure drop in vertical upflow and downflow tubes were carried out at comparatively high fluid velocities to obtain information on the hydrodynamics of gas-liquid-fine solid particle flow systems over a wide range of operating conditions.

1. Experimental

Air, city water and fine glass spheres were used as the gas, liquid and solid, respectively. Three cuts of glass particles were used, as listed in **Table 1**. The finest cut (A) had a mean particle size of just under 30 μm, the medium cut (B) was 63 μm and the large cut (C) was just under 100 μm. A schematic diagram of the experimental apparatus is shown in **Fig. 1**. Air from a compressor flows through an air filter and an air/oil separator to eliminate impurities in

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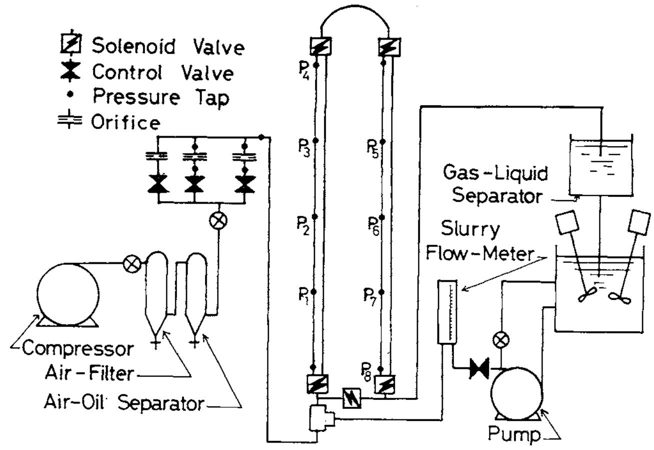

Fig. 1. Schematic diagram of experimental apparatus.

Table 1. Properties of glass spheres

Glass spheres	Density [g/cm³]	Average size [μm] dp_{32}	dp_{50}
A	2.52	29	28
B	2.52	63	63
C	2.52	98	94

the air before being mixed with slurry in the T-tube mixer. The mixed flow of air and slurry passes through vertical upflow and downflow columns to a gas-liquid separator tank and then on to a stirred-tank reservoir. Electric solenoid valves at both ends of the vertical upflow and downflow columns are used to measure the gas holdup in each column. Five holes, each 1 mm in diameter, are drilled in each side of the column at 60 cm spacing to enable pressure measurements to be taken with a manometer. A small sedimentation tank just after each tap on the column prevents gas and fine glass particles from flowing into the manometer line.

Measurement

At given flow velocities of air and slurry containing a given concentration of solid particles, pressure drop and gas holdup measurements were taken once stable operation had been attained for at least 5 minutes. Pressure drop was measured using the manometer tubes. Gas holdup was determined by instataneously closing the electric solenoid valves at both ends of the columns. The holdup was then measured directly as the gas volume fraction remaining in the columns once the gas and slurry had separated.

Two experiments for pressure drop and seven to 10 experiments for gas holdup were carried out at each operating condition. Their average values were then taken.

The concentration of solid particles in slurry was determined by drying samples of slurry in the gas-liquid separator and obtaining the average weight fraction of solid particles.

2. Results and Discussion

The operating conditions for the downward and upward tubes used in this work are listed in **Table 2**.

2.1 Gas holdup in upflow tubes

Relationships between gas holdup and gas velocity for various solid concentrations are shown in **Figs. 2** and **3**. Since almost the same results are obtained in Figs. 2 and 3 for different experimental conditions, we conclude that there is no effect of D_T, d_p and C_S on ε_G.

Observed values of ε_G in this gas-liquid-solid multiphase flow are compared to the equation of Nicklin *et al.* (Eq. (1))[6] proposed for the upflow slugging regime in gas-liquid two-phase flow in **Fig. 4**.

$$\varepsilon_G = \frac{U_G/U_T}{1.20 + 0.35/\sqrt{Fr_T}} \tag{1}$$

where

$$U_T = U_G + U_L \quad \text{and} \quad Fr_T = U_T^2/(gD_T)\,.$$

A comparison between this work's results and the two-phase correlations of Govier *et al.*[1] and Hughmark[3] was also made, and it was concluded that Eq. (1) proposed by Nicklin *et al.* shows the best agreement with the observed values.

2.2 Frictional pressure drop in upflow tubes

The relationship between frictional pressure drop per unit length and gas velocity for various d_p and C_S values are shown in **Fig. 5** for $U_L = 60$ cm/s, using a tube of $D_T = 1.55$ cm. The relationships between $\Delta P_f/L$ and U_G for Solid C at various C_S values are shown in **Fig. 6** at $U_L = 60$ cm/s using $D_T = 2.59$ cm. It is evident from these figures that a considerable effect of C_S on $\Delta P_f/L$ is observed although no effect of d_p on $\Delta P_f/L$ is detectable. The solid lines in the figures represent the results calculated from the Lockhart-Martinelli correlation,[5] which show good agreement with the pressure drop data taken by Toda *et al.*[9] at compara-

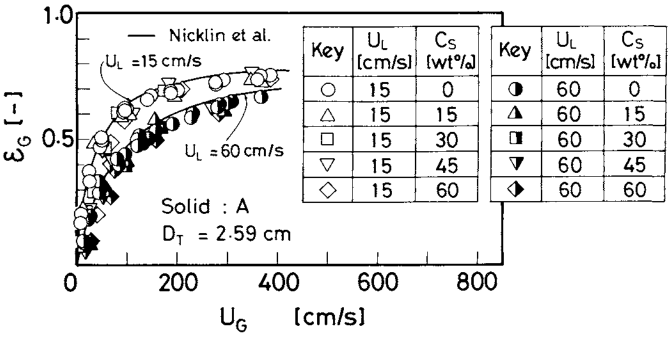

Fig. 2. Effect of solid concentration on gas holdup at D_T=2.59 cm using Solid A in vertical upflow.

Table 2. Experimental conditions

Flow	Upward	Downward
Tube diameter [mm]	26, 15	26, 15
Tube length [mm]	2785	2797
Air flow rate U_G [cm/sec]	0–800	0–800
Slurry flow rate U_L [cm/sec]	0– 80	0– 80
Solids concn. in slurry C_S [wt%]	0– 65	0– 65

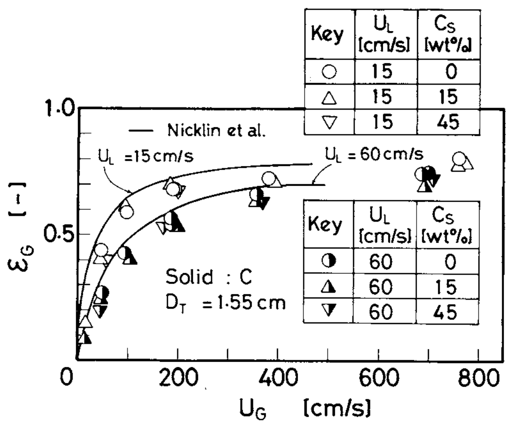

Fig. 3. Effect of solid concentration on gas holdup at D_T= 1.55 cm using Solid C in vertical upflow.

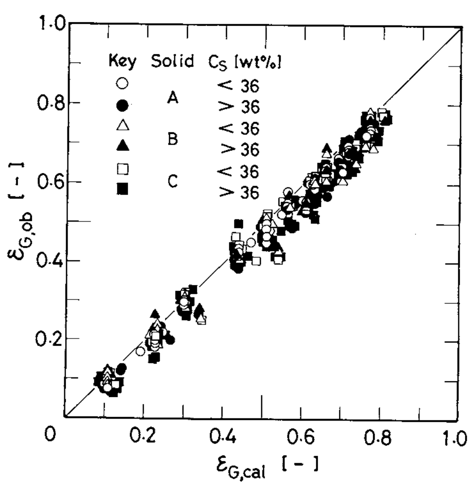

Fig. 4. Comparison of observed gas holdup in gas-liquid-solid flow with calculated value from Nicklin *et al.*[6] (vertical upflow).

tively low gas velocities. Thomas's following equation[8] was used for the estimation of slurry viscosity, which was adopted as viscosity in the calculation.

$$\mu_L=\mu_{\text{water}}[1.0+25\phi^2+0.062 \times\exp\{1.875\phi/(1-1.595\phi)\}] \quad \text{for} \quad 0\leqq\phi\leqq0.5 \qquad (2)$$

Also, slurry density was adopted as density in the calculation. It is apparent that agreement between the correlations and the observed data is diminished in the high-gas velocity region. Observed frictional pressure drop data, excluding negative pressure drop points, for gas-liquid-solid multiphase flow were compared to calculated values from the Lockhart-Martinelli correlation. The comparison showed that $\Delta P_f/L$ is underestimated by Lockhart-Martinelli in the region of comparatively high pressure drops obtained at high gas velocities. A simple single-flow model assuming a frictional slurry velocity $u_L=U_L/(1-\varepsilon_G)$, proposed by Hughmark[4] for the two-phase horizontal slugging flow regime, was also examined to describe the pressure drops at comparatively high gas velocities. Observed frictional pressure drop data for gas-liquid-solid multiphase flow are compared with values calculated from Hughmark's single-flow model in **Fig. 7**. The simple single-flow model shows better agreement with the observed frictional pressure drop than does the Lockhart-Martinelli correlation.

2.3 Gas holdup in downflow tubes

The relation between ε_G and U_G for U_L=15 cm/s is shown in **Fig. 8** under various conditions. It is apparent from the figure that the same gas holdups as in gas-liquid two-phase flow are obtained at low slurry velocities under the same experimental conditions of tube diameter, average size of solid particles and solid particle concentration. The solid lines in the figure represent results calculated from the following equation, i.e., a modification of the equation of Oshinowo *et al.*[7] proposed for the slugging flow regime in vertical downflow.

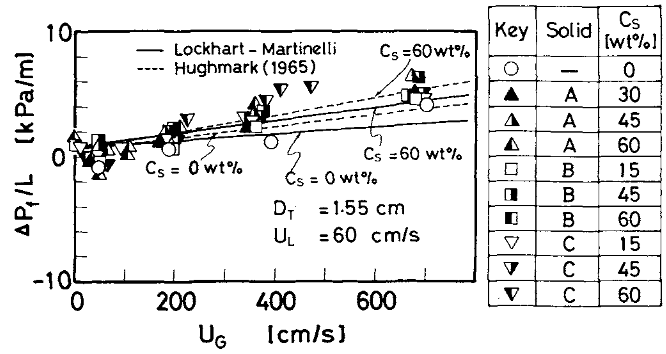

Fig. 5. Effects of average size and concentration of solid particles on frictional pressure drop at D_T= 1.55 cm in vertical upflow.

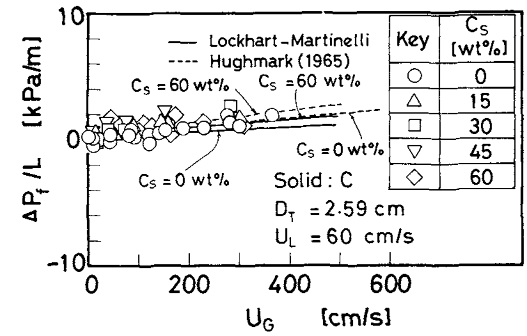

Fig. 6. Effect of solid concentration on frictional pressure drop at D_T=2.59 cm using Solid C in vertical upflow.

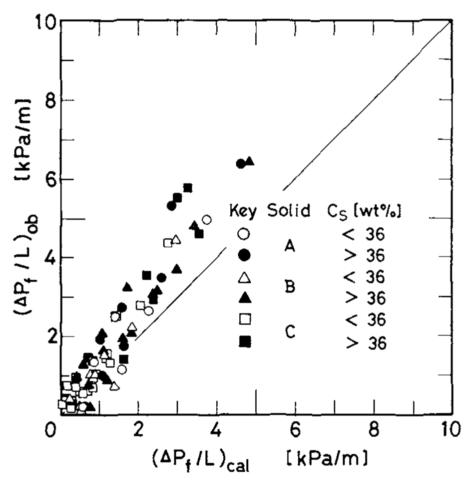

Fig. 7. Comparison of observed frictional pressure drop in gas-liquid-solid flow with calculated value from Hughmark[4] (vertical upflow).

$$\varepsilon_G = \frac{U_G/U_T}{1.10 - 0.33\sqrt{Fr_T}} \tag{3}$$

There is good agreement between the calculated and the observed values.

The relation between ε_G and U_G for D_T=1.55 and 2.59 cm at U_L=60 cm/s is shown in **Figs. 9** and **10** under various operating conditions. It is evident from these figures that when U_G>200 cm/s, the same gas holdup as in gas-liquid two-phase flow is obtained both at U_L=60 cm/s and at U_L=15 cm/s under the same experimental conditions for D_T, d_p and C_S. At lower gas velocities there is apparently a big discrepancy in ε_G between gas-liquid-solid three-phase flow and gas-liquid two-phase flow. That is, higher gas holdups are obtained at higher concentrations of solid particles. To investigate the effect of U_L on ε_G at low gas velocities, experimental results for ε_G at various U_L values are shown in **Fig. 11** for U_G=12.5–15 cm/s and D_T=2.59 cm. From the figure, much higher gas holdups than those in gas-liquid two-phase flow are observed with increasing U_L at high concentrations of solid particles. The observation of flow pattern with increasing U_L revealed that a transition from the film flow regime to the slug flow regime occurs at U_L=40 cm/s for the gas-liquid two-phase flow, whereas such a transition for the gas-liquid-solid multiphase flow is observed at higher slurry velocities for high solid-particle concentrations. This phenomenon causes the discrepancy in ε_G between gas-liquid flow and gas-liquid-solid flow at conditions of low U_G and high U_L. The same results as in Fig. 11 were obtained for D_T=1.55 cm.

Observed data for gas-liquid-solid multiphase flow are compared with values calculated from Eq. (3) in **Fig. 12**. The swarm of painted elliptical keys in the figure represent the data obtained with $U_G \leq 100$ cm/s, $U_L \geq 60$ cm/s and $C_S > 36$ wt%.

Except for these data points, the experimental results can be reasonably predicted by Eq. (3).

2.4 Frictional pressure drop in downflow tubes

The relationship observed between $\Delta P_f/L$ and U_G is shown in **Fig. 13** for D_T=2.59 cm, U_L=15 cm/s and C_S=60 wt% for three kinds of solid particles. In **Fig. 14**, the same kind of plot as in Fig. 13 is shown for various C_S values, except now D_T=1.55 cm and U_L= 60 cm/s. It is apparent from these figures that ΔP_f is independent of d_p, but increases with C_S. The simple single-flow model with $u_L = U_L/(1-\varepsilon_G)$ proposed by Hughmark,[4] which is proven to be applicable to the gas-liquid-solid vertical upflow in the foregoing section, was examined for pressure drop estimation. The

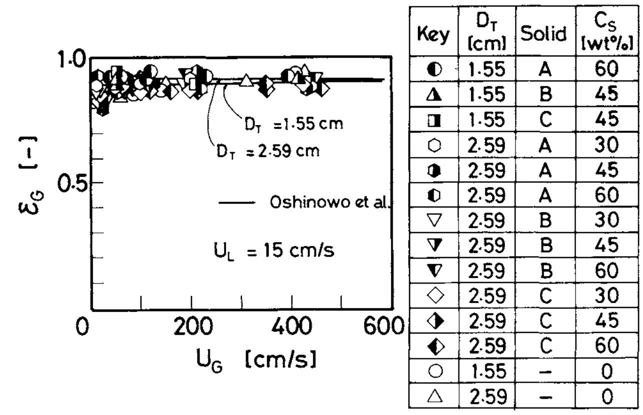

Fig. 8. Relation between gas holdup and gas velocity at $U_L = 15$ cm/s in vertical downflow.

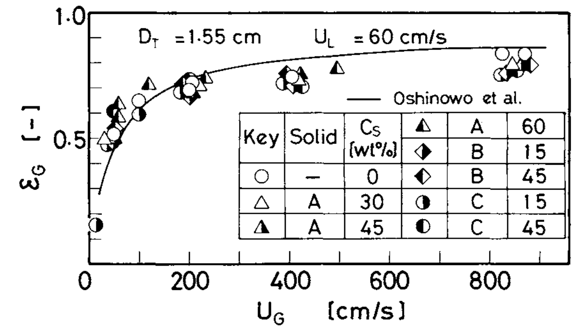

Fig. 9. Relation between gas holdup and gas velocity at $D_T = 1.55$ cm and $U_L = 60$ cm/s in vertical downflow.

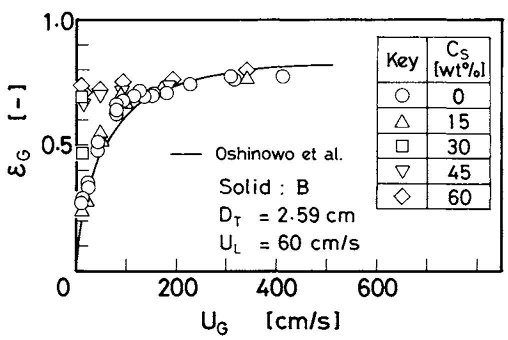

Fig. 10. Relation between gas holdup and gas velocity at $D_T = 2.59$ cm and $U_L = 60$ cm/s in vertical downflow.

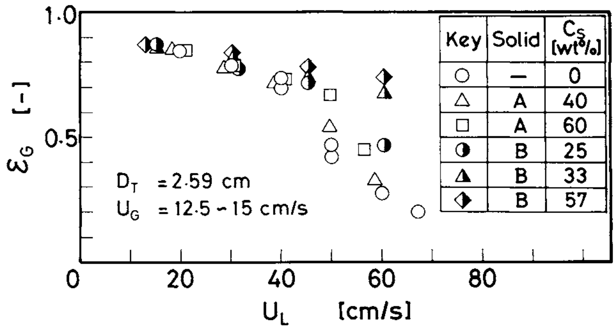

Fig. 11. Effect of slurry velocity on gas holdup at $D_T = 2.59$ cm and $U_G = 12.5$–15 cm/s in vertical downflow.

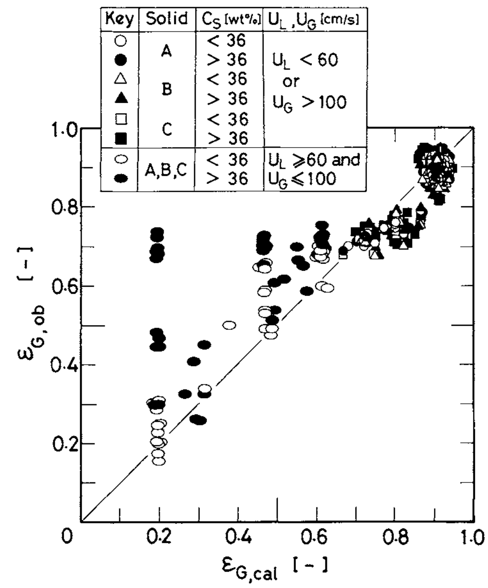

Fig. 12. Comparison of observed gas holdup in gas-liquid-solid flow with calculated value from Oshinowo *et al.*[7] (vertical downflow).

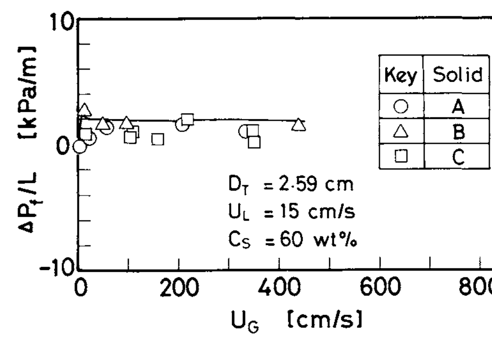

Fig. 13. Effect of solid particle size on frictional pressure drop at $D_T = 2.59$ cm, $U_L = 15$ cm/s and $C_S = 60$ wt% in vertical downflow.

solid lines in Figs. 13 and 14 are calculated from this model. Values calculated from the model are compared with the observed data in **Fig. 15**. Experiment and prediction show comparatively good agreement.

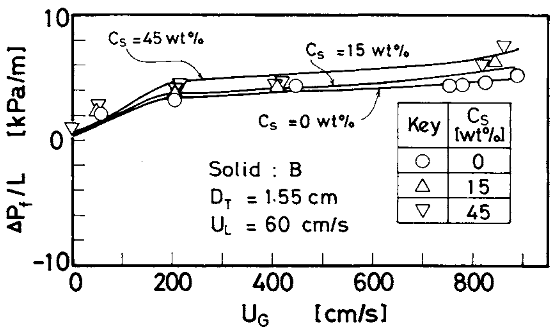

Fig. 14. Effect of solid concentration on frictional pressure drop at $D_T = 1.55$ cm and $U_L = 60$ cm/s using Solid B in vertical downflow.

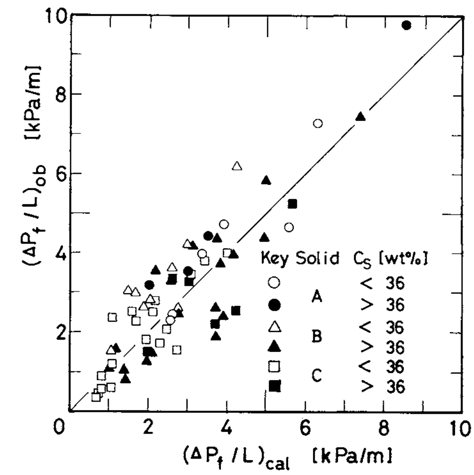

Fig. 15. Comparison of observed frictional pressure drop in gas-liquid-solid flow with calculated value from Hughmark[4] (vertical downflow).

Conclusion

Gas holdups and pressure drops in gas-liquid-solid multiphase flow were measured at comparatively high gas and slurry velocities. The following results were obtained.

1) Within the range of experimental conditions, gas holdups in vertical upflow tubes are independent of tube diameter, average size and concentration of solid particles. That is, gas holdup in gas-liquid-solid multiphase flow is the same as in gas-liquid two-phase flow at high gas and slurry velocities. An expression by Nicklin *et al.* (Equation (1)), developed for two-phase flow, can be used to estimate gas holdup in multiphase flow at the high gas and slurry velocities.

2) Frictional pressure drops in vertical upflow tubes are independent of the average size of solid particles, but are dependent on the concentration of solid particles. After a correction for slurry viscosity and density, the Lockhart-Martinelli equation can be used to estimate the pressure at low gas velocities. However, Hughmark's model assuming a single flow with $u_L = U_L/(1-\varepsilon_G)$ is better for pressure drop estimation at comparatively high gas velocities.

3) Gas holdups at low gas and high slurry velocities in vertical downflow tubes are much larger than those under the same fluid velocity conditions in gas-liquid two-phase flow. This phenomenon can be explained by the fact that the transition state from the film to slug flow regimes with increasing U_L moves to the flow region at higher slurry velocities by a solid effect on the flow regime. Except at the low gas and high slurry velocities, gas holdups in gas-liquid-solid three-phase flow are the same as those in gas-liquid two-phase flow, being estimated from a modified form of the equation of Oshinowo *et al.* (Eq. (3)).

4) Frictional pressure drops in vertical downflow tubes are independent of the average size of solid particles but are dependent on the concentration of solid particles. The frictional pressure drop can be estimated by using the simple single-flow model proposed by Hughmark.[4]

Nomenclature

C_S	=	solid particle concentration in slurry	[wt%]
D_T	=	tube diameter	[cm]
d_{p32}	=	Sauter average size	[μm]
d_{p50}	=	50% particle size	[μm]
Fr_T	=	$U_T^2/(gD_T)$, Froude number based on total fluid velocity	[—]
g	=	gravitational acceleration	[cm/s²]
L	=	tube length	[cm]
ΔP_f	=	frictional pressure drop	[Pa]
U_G	=	superficial gas velocity	[cm/s]
U_L	=	superficial liquid or slurry velocity	[cm/s]
U_T	=	$U_G + U_L$, superficial total fluid velocity	[cm/s]
u_L	=	$U_L/(1-\varepsilon_G)$, apparent linear velocity of slurry	[cm/s]
ε_G	=	gas holdup	[—]
μ_L	=	viscosity of liquid or slurry	[g·cm/s]
μ_{water}	=	viscosity of water	[g·cm/s]
ϕ	=	volume fraction of suspended solid particles	[—]

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