Drift Flux Distribution Parameter in Three-Phase Air-Lift Reactors

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Gas hold-up and liquid circulation velocity data in a three-phase system involving alginate beads in an internal-loop airlift reactor, reported by Lu *et al.* (1995), have been analyzed to evaluate the distribution parameter in drift flux model. The calculated distribution parameter values were significantly greater than 1.0 (the value used by Lu *et al.* in their modeling) and also affected by the solid volume fraction. An empirical correlation of this effect has been presented.

Key words:
Drift-flux model, gas holdup, liquid circulation velocity, multiphase reactors, design

Introduction

Multiphase reactors are commonly used in industry to carryout a variety of chemical and biological transformations. Modeling of hydrodynamics in these reactors remains a major task and several different approaches are utilized for this purpose.[1,2] Computational fluid dynamics (CFD) based modeling is increasingly becoming popular and although computational abilities are rapidly expanding,[3] the grid-size still remains quite coarse.[1] Models involving correlations with volume-average parameters remain popular due to ease of their use and ability to reasonably predict critical system variables (hold-ups of different fractions in various parts of reactors) for different operational parameters.[3–10] These models involve the mechanical energy balance equation for liquid flow which is combined with relationships between volume fractions and velocities of different phases to define the system.[5–7,11–13]

In order to relate the gas holdup with the velocities of various phases in the riser and downcomer sections of air-lift reactors, the drift-flux model of Zuber and Findley[14] is commonly used.[7,15,16] For the two-phase systems, these relationships for the riser section may be written as:

$$\frac{u_G}{\varepsilon_G} = C_0(u_G + u_L) + v_G \qquad (1)$$

A similar equation may also be written for the downcomer section with a negative sign before the

bubble slip velocity (v_G). According to Turner,[17] the bubble slip velocity (v_G) equals the terminal bubble rise velocity (v_∞). Although the drift-flux equation was originally proposed for two-phase gas-liquid system, many authors have suggested that it can be used also for the three-phase gas-liquid-solid systems as well, irrespective of whether solids are uniformly distributed in the liquid phase or not[15,18,19] with incorporation of solid-phase velocity in the respective equations as well. When the solids are uniformly distributed in liquid, the system behaves as a pseudo- two-phase system in which the properties of liquid may be replaced by those of solid-liquid suspension.[15,20]

C_0, the distribution parameter, represents radial distributions of gas phase in the reactor sections. If the gas bubbles are uniformly distributed, the value of C_0 equals 1.0. If bubbles prefer the center of column, $C_0 > 1$, and if bubbles prefer the space near walls, $C_0 < 1$.[9] Using calcium alginate gel beads (ε_S = 0.1, 0.2), Nakao *et al.*[19] found the values of C_0 between 2.32 and 2.57 for an internal loop airlift reactor. These values were explained on the basis of A_D/A_R ratio (= 1.78) and the flow patterns in the riser. Several other authors[6,21–24] have also reported the value of $C_0 > 1$ in three-phase systems; still values around or equal to 1.0 continue to be used by several researchers.[11,16]

Lu *et al.*[16] reported gas hold-up and liquid circulation velocities in a two and three phase internal loop airlift reactor. These authors used calcium alginate beads (solid volume fraction ε_S = 0, 0.05, 0.10, 0.20, and 0.30) as solid phase. These authors analyzed their experimental data assuming a value

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of the drift-flux distribution parameter $C_0 = 1$. Using the experimental data of Lu *et al.*,[16] we have calculated the value of C_0. Our analysis reveals that C_0 is significantly greater than 1 and it is also affected by the volume fraction of solids in the three-phase system. These results are presented here.

Experimental set-up of Lu *et al.* (1995): The air-lift reactor used by these authors was an internal circulation airlift reactor consisting of an outer tube of 0.18-m internal diameter and 2.5-m length, a draught tube (0.12-m internal diameter, 0.005-m wall thickness, and 1.1-m long) located 0.1-m above the base of the outer tube, and a sparger (0.09-m diameter ring made of 0.01-m diameter pipe with 0.0005-m holes located 0.01-m apart). Height of fluid (water) in the reactor was 1.4 m and the solid phase consisted of Ca-alginate beads (density 1030 kg·m^{-3}) of different average diameters (0.001 m, 0.0021 m, 0.0036 m). Several different solid volume fractions were utilized (0, 0.05, 0.1, 0.2, and 0.3) and the air flow rate was varied between 0 and 1.7 x 10^{-3} m^3·s^{-1}. This geometry meant that gas-liquid separation at the top of the column was not very efficient and considerable amount of gas was present in the downcomer.

Analysis

Lu *et al.*[16] reported linear liquid velocity in riser (v_{LR}), and gas volume fractions in the riser (ε_{GR}) and the downcomer (ε_{GD}) for different gas flow rates (Q_G) through the riser column in a suspension of alginate beads. From these data, the superficial gas velocity in riser (u_{GR}) was calculated as

$$u_{GR} = \frac{Q_G + u_{GD}A_D}{A_R} \qquad (2)$$

Here, u_{GD} is the superficial velocity of gas in the downcomer; it was estimated using the correlation proposed by Siegel *et al.*[25]

$$u_{GD} = 3.5\varepsilon_{GD}^2 + 0.22\varepsilon_{GD} + 1.1 \cdot 10^{-4} \qquad (3)$$

Superficial liquid velocity in the riser, u_{LR}, was calculated from measured linear liquid velocity (v_{LR}) as per suggestion made by Lu *et al.*[16]:

$$u_{LR} = v_{LR} \cdot (1 - \varepsilon_{GR}) \qquad (4)$$

These values of u_{GR}, u_{LR}, and ε_{GR} were plotted according to equation (1) and the parameter C_0 was estimated for the different solid valid fractions. In these cases, a constant value of bubble rise velocity ($v_\infty = 0.25$ m/s) could be used satisfactorily. This value of bubble rise velocity has been used by Lu *et*

al.[16] also and is in agreement with the measurements of Heijnen and van't Riet.[26] The estimates of C_0 for 2.1 mm diameter alginate beads at different solid volume fractions (ε_s from 0–30%) ranged from 1.77 to 2.2 and are presented in Table 1. If a value of v_∞ higher than 0.25 m/s was permitted in the estimations, the values of C_0 decreased slightly but in no case went lower than 1.6.

T a b l e 1 – *Effect of solid volume fraction on distribution parameter C0 (particle diameter 2.1 mm)*

ε_s	C_0^*	R^2
0	1.77	0.92
0.05	1.88	0.96
0.10	1.92	0.96
0.20	1.99	0.97
0.30	2.20	0.96

*The calculations were performed with v = 0.25 m/s as suggested by Lu et al. (1995).

Efforts to use a single value of C_0 for all the solid loadings resulted in large scatter (a value of $C_0 = 1.96$ with a correlation coefficient of 0.78). Clearly, the effect of solid volume fraction on the gas distribution parameter C_0 is significant. The estimates of C_0 suggest a tendency of the bubbles to prefer the center of the riser.

When the estimates of C_0 from Table 1 were correlated with the volume fraction of solids in the riser, the following relationship ($R^2 = 0.95$) was obtained:

$$C_0 = 1.78 + 1.31\varepsilon_s \qquad (5)$$

This trend of dependence of C_0 on solid volume fraction was observed even when bubble rise velocities greater than 0.25 m/s were permitted in the calculations. A plot of all the 2.1 mm diameter experimental data of Lu *et al.*,[16] according to equation (1) and incorporating equation (5) is shown in Fig. 1. The agreement between the experimental data and calculated values is good.

Discussion

The results of calculations presented in Table 1 and Fig. 1 show that the values of distribution parameter C_0 are significantly greater than 1.0, the value used by Lu *et al.*[16] in their model predictions. These authors combined the drift flux model (assuming $C_0 = 1.0$ and $v_\infty = 0.25$ m/s) with empirical correlations for gas holdups in the riser and downcomer and with the mechanical energy bal-
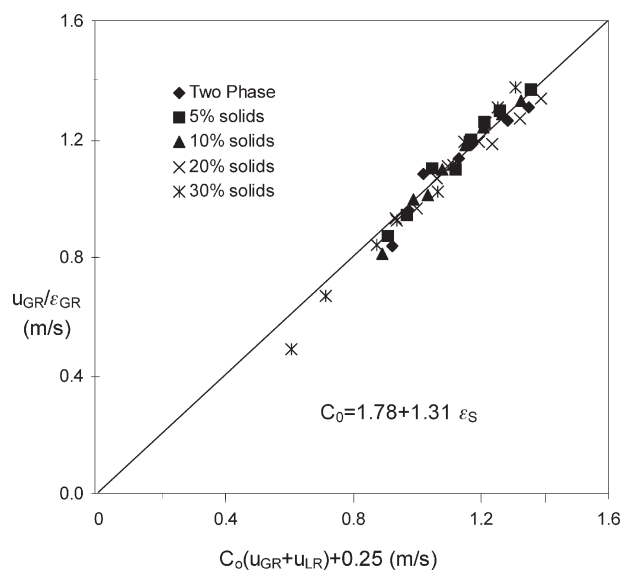
F i g . 1 – *A plot of two- and three-phase data of Lu et al. (1995) according to drift flux model (equation 1).*

ance equation to predict the riser liquid velocity (u_{LR}). u_{LR} was predicted satisfactorily for low solids loading, but was overestimated at higher loading of solids. It is not clear if the disparities between the experimental observations and the model predictions are due to inaccurate value of the distribution parameter, but it is clear that a constant value of 1.0 for C_0 cannot be justified by the experimental data.

Similar conclusions about the value of C_0 have been drawn by several other researchers as well. The observation that C_0 depends on the solid volume fraction has also been reported in literature.[6,24] A cursory survey of the values of C_0 reported in published literature and the effect of solid volume fractions on this parameter presents a very confusing picture with no easily discerned trends. Clearly C_0 depends on the prevalent hydrodynamic regime, physico-chemical characteristics of the system as well as on the operational parameters. At present, no correlation relating C_0 to the operational and physico-chemical parameters in air-lift reactors exists, making this to be a weak link in the use of drift-flux approach in modeling of flow behavior in multiphase systems. Since drift-flux is often utilized in the CFD models of the air-lift reactors, it is recommended that efforts be directed to developing such a correlation.

Conclusions

The results presented in this communication show that the gas distribution parameter C_0 can be significantly higher than the value of 1.0 expected for uniformly distributed gas bubbles. This value is affected also by the volume fraction of solids in the reactor.

Nomenclature

A_D – Area of cross section of downcomer, m^2

A_R – Area of cross section of riser, m^2

C_0 – Ratio of gas-holdup weighted-average and average velocity of multiphase system at specified cross section of multiphase flow, –

Q_G – Input Gas flow rate, m^3/s

u_{GD} – Superficial velocity of gas in the downcomer, m/s

u_{GR} – Superficial velocity of gas in the riser, m/s

u_{LR} – Superficial velocity of liquid in the riser, m/s

v_G – Gas-holdup weighted-average drift velocity of gas at specified cross section of multiphase flow, m/s

v_{LR} – Linear liquid velocity in the riser, m/s

v_∞ – Terminal bubble rise velocity of bubbles, m/s

ε_{GD} – Average gas holdup in the downcomer, –

ε_{GR} – Average gas holdup in the riser, –

ε_s – Volume fraction of solids in reactor, –

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