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Hydrodynamic model for three-phase internal- and external-loop airlift reactors

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Abstract

A mathematical model predicting the hydrodynamic behaviour of three-phase airlift reactors, working with low-density solids and with high solids loading, was developed. The model allows for the prediction of local gas holdup and liquid velocity in airlift bioreactors. Model was validated for an external-loop airlift reactor and an internal-loop airlift reactor with an enlarged degassing zone, being a good agreement obtained between calculated and experimental data.

Keywords: Three-phase reactor; Hydrodynamic model; Gas holdup; Liquid velocity

1. Introduction

Several models describing satisfactorily the hydrodynamics of two-phase airlift reactors have been developed (Chisti, Halard & Moo-Young, 1988; Garcia-Calvo & Letón, 1996; Kemblowski, Przywarski & Diab, 1993). However, models describing the hydrodynamics of three-phase airlift reactors are limited, especially with low-density solids as the solid phase, which is the case in biotechnology processes.

Chisti et al. (1988) developed a two-phase model that was extended to a three-phase system by Livingston and Zhang (1993), to predict the liquid circulation velocity. A pseudo-homogeneous phase density for liquid–solid phase, considering different values for the riser and the downcomer, was used.

Lu, Hwang and Chang (1995) also developed a mathematical model for the prediction of the liquid velocity and of the gas holdup for three-phase airlift reactors, assuming that particles are well dispersed in the reactor. The solid phase and the liquid phase were regarded as a "pseudo-homogeneous mixture phase" and the three-phase airlift reactor system was then reduced to a two-phase system, containing the solid–liquid mixture phase and the gas phase.

In this study, a model that allows for the estimation of gas holdup and liquid velocity in three-phase internal-loop (ILR) and external-loop (ELR) airlift reactors, working with high solids loading, was developed. The concept of "pseudo-homogeneous mixture phase" was also employed.

2. Theory

In the present work some assumptions were made in the development of the mathematical model: The airlift reactor consists in four sections — the riser, the downcomer, the top and the bottom sections; the solid and the liquid phase were considered as a "pseudo-homogeneous mixture phase", with a constant density for the entire reactor; for the internal-loop reactor, the values of solids holdup used were the experimental ones and, for the external-loop reactor, it was considered that the distribution of solids is almost uniform, being the solids holdup equal, in every section of the reactor, to the solids loading.

2.1. Riser gas holdup

Riser gas holdup estimation was done using the equation proposed by Bando, Nishimura, Sota, Hattori, Sakai and Kuraishi (1990) for a three-phase system, as a modification of the Zuber and Findlay model (Clark

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& Flemmer, 1985; Lu et al., 1995; Snape, Zahradník, Fialová & Thomas, 1995).

$$\varepsilon_{gr} = \frac{U_{gr}}{C(U_{gr} + U_{lr} + U_{sr}) + U_{bt}}. \quad (1)$$

The riser superficial velocity of the solid particles relative to the reactor walls (U_{sr}) is given by

$$U_{sr} = \frac{\varepsilon_{sr} U_{lr}}{(1 - \varepsilon_{gr} - \varepsilon_{sr})} - \varepsilon_{sr} U_{st}. \quad (2)$$

The solids settling velocity U_{st} was calculated using a correlation for spherical particles and Reynolds number between 1000 and 350 000 (Perry & Green, 1984):

$$U_{st} = 1.73\sqrt{g D_p(\rho_p - \rho)/\rho}. \quad (3)$$

Replacing U_{sr} (Eq. (2)) into Eq. (1), the riser gas holdup is given by

$$\varepsilon_{gr} = \frac{U_{gr}}{C[U_{gr} + U_{lr}(1 + \varepsilon_{sr}/(1 - \varepsilon_{gr} - \varepsilon_{sr})) - \varepsilon_{sr} U_{st}] + U_{bt}}. \quad (4)$$

In the present study, Eq. (4) was used to fit the experimental values of riser gas holdup, optimising the values of the distribution factor C and the terminal rise velocity of a single bubble U_{bt}.

2.2. Downcomer gas holdup

Several authors (Chisti, 1989) found linear relationships between riser and downcomer gas holdup. Therefore, with the value of riser gas holdup given by Eq. (4), downcomer gas holdup for the internal-loop reactor was calculated using the equation:

$$\varepsilon_{gd} = a\varepsilon_{gr} + b. \quad (5)$$

Calculation of parameters a and b was included in the optimisation procedure.

For the external-loop reactor, the downcomer gas holdup is negligible.

2.3. Riser superficial liquid velocity

An energy balance was used for the prediction of the riser superficial liquid velocity. The basis of the balance is to equate the head differential that causes liquid circulation between the riser and the downcomer (P_h) and the head losses due to friction ($-\Delta P_{\text{loss}}$).

P_h is given by

$$P_h = [(\rho_h \varepsilon_{hd} + \rho_g \varepsilon_{gd}) - (\rho_h \varepsilon_{hr} + \rho_g \varepsilon_{gr})] g\, H_{r,d}, \quad (6)$$

where ρ_h is the pseudo-homogeneous-phase density.

Since ρ_h is much larger than ρ_g and as the pseudo-homogeneous-phase density can be expressed by

$$\varepsilon_h = 1 - \varepsilon_g. \quad (7)$$

Eq. (6) becomes in the form:

$$P_h = (\varepsilon_{gr} - \varepsilon_{gd})\rho_h g H_{r,d}. \quad (8)$$

The total frictional loss in the airlift reactor is

$$-\Delta P_{\text{loss}} = \sum(-\Delta P_f)_i. \quad (9)$$

$(-\Delta P_f)_i$ is the frictional loss in each section i of the reactor and can be obtained by (Brodkey & Hershey, 1988):

$$(-\Delta P_f)_i = \frac{1}{2}\rho_h k_{fi} V_{li}^2, \quad (10)$$

where k_{fi} is the friction coefficient in section i of the reactor.

2.3.1. Friction coefficients

Standard "one-phase flow" equations were used to calculate the friction coefficients in specific parts of each airlift reactor:

- Internal-loop airlift reactor: In the reactor tubes (riser and downcomer) and bottom of the reactor; the friction coefficient in the top is negligible.
- External-loop airlift reactor: In the reactor tubes (riser, downcomer, top and bottom sections), in fittings and diameter changes.

Reactor tubes. The friction loss coefficient in lines of circular cross section was calculated according to (Brodkey & Hershey, 1988):

$$k_{fi} = 4f_i \frac{H_i}{D_i}, \quad (11)$$

where f_i is the friction factor of the pseudo-homogeneous mixture in section i.

The Blasius equation (Perry & Green, 1984), for one-phase and turbulent flow, was applied to a three-phase system, determining f for the pseudo-homogeneous mixture

$$f = 0.0791 Re_h^{-0.25}, \quad (12)$$

where Reynolds number of the pseudo-homogeneous phase for section i (Re_{hi}) is given by

$$Re_{hi} = \frac{\rho_{hi} V_{li} D_i}{\mu_{hi}}. \quad (13)$$

A correction factor α, proposed by Garcia-Calvo and Letón (1996) for systems where two-phase flow is present, was also introduced. So, considering that μ_h and ρ_h are constant for the entire reactor, Eq. (11) becomes

$$k_{fi} = \frac{4*\alpha*0.0791 H_i \mu_h^{0.25}}{\rho_h^{0.25} D_i^{1.25} V_{li}^{0.25}}. \tag{14}$$

As ρ_h, μ_h and α depend on the solids loading, a parameter β was considered

$$\beta = \alpha*4*0.0791*\mu_h^{0.25}\rho_h^{-0.25}. \tag{15}$$

Bottom and top friction coefficient for the internal-loop reactor. Since $k_{fb} \gg k_{ft}$ for internal loop airlift reactors, k_{ft} is negligible and k_{fb} was calculated by (Chisti et al., 1988):

$$k_{fb} = 11.4\left(\frac{A_d}{A_b}\right)^{0.79} \tag{16}$$

valid for an A_d/A_b range of 0.2–1.8. A_b is the free area below the draught tube.

Fittings. In the external-loop airlift reactor there are two elbows ("screwed long radius 90° ell") — connecting the top tube to the downcomer — $K_{ft,d}$ — and the downcomer to the bottom tube — $K_{fd,b}$, which depend on the nominal size, a "sharp edged entrance" from the riser into the top section ($K_{fr,t}$) and a "sharp edged exit" from the bottom section into the riser ($K_{fb,r}$). From Brodkey and Hershey (1988), $K_{ft,d} = 0.25$, $K_{fd,b} = 0.40$, $K_{fr,t} = 0.50$,$K_{fb,r} = 1.0$.

Diameter change. There are two types of diameter changes in the external-loop reactor, a "gradual contraction" in the top section (K_{fgc}) and a "sudden contraction" in the downcomer (K_{fsc}). From Brodkey and Hershey (1988):

$$K_{fgc} = 0.04,$$

$$k_{fsc} = 0.42\left(1 - \frac{d_2^2}{d_1^2}\right), \frac{d_2}{d_1} \leqslant 0.76. \tag{17}$$

At steady state (Lu et al., 1995):

$$P_h = -\Delta P_{\text{loss}}. \tag{18}$$

Combining this equation with Eqs. (8)–(10) and (14) and knowing that

$$V_{li} = \frac{U_{li}}{1 - \varepsilon_{gi} - \varepsilon_{si}}, \tag{19}$$

and that for a section i

$$A_i U_{li} = A_r U_{lr}, \tag{20}$$

the final equations for the riser superficial velocity become

Internal-loop Airlift Reactor:

$$2gH(\varepsilon_{gr} - \varepsilon_{gd}) = \left(\frac{A_r}{A_d}\right)^2 \frac{k_{fb}}{(1 - \varepsilon_{gd} - \varepsilon_{sd})^2} U_{lr}^2 + \left[\frac{D_r^{-1.25}}{(1 - \varepsilon_{gr} - \varepsilon_{sr})^{1.75}} + \left(\frac{A_r}{A_d}\right)^{1.75} \times \frac{(D_c - D_r)^{-1.25}}{(1 - \varepsilon_{gd} - \varepsilon_{sd})^{1.75}}\right] H\beta U_{lr}^{1.75}. \tag{21}$$

External-loop airlift reactor:

$$2gH\varepsilon_{gr} = \left[\frac{(H_r + H_t)D_r^{-1.25}}{(1 - \varepsilon_{gr} - \varepsilon_s)^{1.75}} + \left(\frac{A_r}{A_d}\right)^{1.75} \frac{(H_d + H_b)D_d^{-1.25}}{(1 - \varepsilon_s)^{1.75}}\right] \times \beta U_{lr}^{1.75} + k_{fr,t}\frac{U_{lr}^2}{(1 - \varepsilon_{gr} - \varepsilon_s)^2} + \left[\left(\frac{A_r}{A_{t,d}}\right)^2 (k_{ft,d} + k_{fgc}) + \left(\frac{A_r}{A_d}\right)^2 (k_{fsc} + k_{fd,b} + k_{fb,r})\right] \frac{U_{lr}^2}{(1 - \varepsilon_s)^2}. \tag{22}$$

3. Experimental

The internal-loop airlift reactor (ILR) used is of the concentric-tube type, with an enlarged degassing zone, as described in Freitas and Teixeira (1997). The diameters of the downcomer and the riser are 0.142 and 0.062 m, respectively. The height of the draught tube is 1.190 m and its bottom edge is 0.086 m above the bottom of the reactor.

The glass wall external-loop airlift reactor (ELR) used, similar to the reactor shown in Snape et al. (1995), has a downcomer and a riser diameter of 0.05 and 0.158 m, respectively, with 2.07 m height. The top section has a height and a diameter of 0.36 and 0.158 m, respectively, with a contraction that connects to a bend of 0.107 m of diameter. At the end of this, there is another contraction to reduce the diameter of the bend to the downcomer diameter. This has the same diameter as the bottom section and the bend that connects them. Both reactors have a working volume of 60 l.

Air was used as the gas phase and injected through perforated plates with 1 mm holes. The airflow rate was varied in a way that the riser superficial gas velocities studied were between 0.01 and 0.50 m/s, for the ILR, and between 0.03 and 0.17 m/s, for the ELR.

The liquid-phase used was water, in the case of the ELR, and an aqueous solution of 10 g ethanol/l, in the ILR.

Ca-alginate beads with a density of 1023 ± 1 kg/m^3 and a diameter of, approximately, 2 mm were applied as solid phase, for different solids loading (0, 10, 20 and 30% v/v).

Riser gas holdup was determined with a manometer, in the ELR, and with pressure transducers, in the ILR (Freitas et al., 1997).

The measurements of the liquid velocity were done, in the ELR, using a conductivity pulse technique (Snape et al., 1995) and, for the ILR, with the pH pulse technique (Freitas et al., 1997).

The optimisation of the parameters C, U_{bt}, a, b and β (Eqs. (4), (5), (21) and (22)), allowing for the prediction of values of riser and downcomer (for the ILR) gas holdup and the riser superficial liquid velocity, was done using a developed computer software.

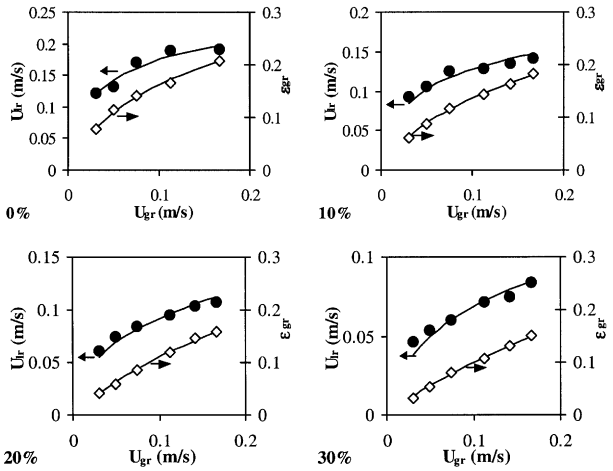

Fig. 2. Riser superficial liquid velocity (●) and riser gas holdup (◇) vs. riser superficial gas velocity (— predicted values), for different solids loading (ELR).

4. Results and discussion

The results of simulations of gas holdup and riser liquid velocity, for the internal- and the external-loop reactors, carried out for each solids loading, are presented in Figs. 1 and 2. As can be seen, there is, in general, a good agreement between the predicted and the measured values of gas holdup in the riser and in the downcomer. However, for the ILR, when working with 30% of solids and low airflow rates, values predicted for the riser gas holdup and riser liquid velocity are significantly different from the measured ones. For this reactor, the circulation becomes very difficult when working with high solids loading and low airflow rates. In these conditions, due to the existence of a zone at the bottom of the reactor where a high concentration of solids occurs, the assumption of the existence of two-phases is no longer valid. For the ELR, the deviation of the predicted values from the experimental ones is not significant. This is because the riser in the ELR is very large, what allows for a good circulation of the fluid, even if there is a high concentration of solids.

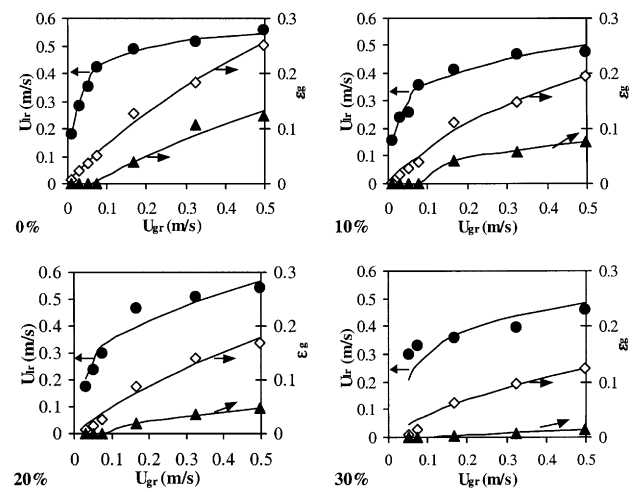

Fig. 1. Riser superficial liquid velocity (●), riser gas holdup (◇) and downcomer gas holdup (▲) vs. riser superficial gas velocity (— predicted values), for different solids loading (ILR).

The distribution factor C is an index of the flow pattern and it is equal to 1 when the flow distribution is radially uniform. However, the fact that C is near unity in magnitude should not be misinterpreted as indicating that plug flow prevails. The magnitude of the distribution parameter is due to the uniformity of the void distributions rather than the character of the phase velocity profile (Young, Carbonell & Ollis, 1991). According to Zuber and Findlay (Shah & Deckwer, 1983), if the holdup and velocity drop linearly from the centre of the tube to the wall, the value of the parameter C varies from 1.5 to 1, for fully established profiles. It can be seen from the values of C (Table 1) that, after an increase (very small for the ELR) till 10% of solids, it decreases with the increase of solids loading. With the introduction of solids (till 10%) the flux is perturbed but, when more solids are added, the flux becomes more uniform and values of C approach 1. Circulation restrictions in the ILR may explain the non-uniformity of the flux, for 30% of solids.

Values of the terminal velocity of a single bubble (U_{bt}) increase, in general, with the increase of solids loading, being higher for the ILR. Such results can be ascribed to the character of the bubble bed in the riser. In the "drift-flux model" (Young et al., 1991), U_{bt} is the terminal velocity of a single bubble, assuming that bubbles do not interact, that is, each bubble moves independently and is not affected by the presence of other bubbles. For the ILR, values of the parameter U_{bt} show that, even for the low solids loading, for which the fits are very good, values are high, comparing with values obtained by other authors (Bando et al., 1990; Lu et al., 1995). Even for 0%

Table 1
Values of the optimised parameters C, U_{bt}, β, a and b

Solids loading	Internal-loop reactor					External-loop reactor		
(%)	C	Ubt (m/s)	β	a	b	C	Ubt (m/s)	β
0	1.13	0.77	0.11	0.644	− 0.032	2.04	0.060	0.0044
10	1.74	0.68	0.11	0.354	0.0068	2.14	0.22	0.010
20	1.13	1.38	0.076	0.280	− 0.0023	1.53	0.58	0.016
30	2.21	1.22	0.069	0.184	− 0.0083	1.20	0.77	0.024

of solids, the U_{bt} value is outside the range of 0.2–0.4 m/s presented in literature for two-phase systems. This may be explained by the small riser cross-sectional area, causing the existence of a swarm of bubbles rising with very high velocities. For the ELR, although the U_{bt} increases, it only presents values higher than 0.4 m/s for the higher solids loading (20 and 30%). In these cases, the spaces between the solids are small what leads to an increase of the interaction between the bubbles, increasing coalescence with the consequent increase of bubbles diameter. The larger the size of the bubbles the higher the values of the riser velocity.

Also from Table 1, it can be seen that the parameter β responds in different ways to the increase of solids loading, for the two types of reactors. For the internal-loop airlift reactor, β decreases for solids loading higher than 10%. Since β is directly proportional to the pseudo-homogeneous phase viscosity and to the correction factor α and inversely proportional to density (Eq. (15)), the decrease of β indicates that, for these high solids concentration, the density of the "pseudo-homogeneous-phase" is the main factor responsible for the high losses. On the contrary, β increases with solids loading in the ELR what could be the result of an increase of the correction factor. As the geometries of the two reactors are very distinct, specially in what concerns the ratio between the riser and the downcomer diameters (for the ILR, $D_r/D_d < 1$ and for the ELR, $D_r/D_d > 1$), changes on viscosity, density and friction with the increase of solids loading will have different consequences on the hydrodynamics of both reactors. Furthermore, if, as an approximation, the viscosity and density of water (0.001 N s and 1000 kg/m^3) are considered to obtain the values of β, the correction factor α should have a value of about 10, for the ILR, and of about 1, for the ELR. Garcia-Calvo and Letón (1996) proposed for a two-phase flow system a value of 2. This means that the third phase has a higher influence on the hydrodynamics of the internal-loop than on the external-loop airlift reactor, deriving from the differences on their geometry.

Parameter a (Eq. (5)), for the ILR, decreases with the increase of solids loading since the larger bubbles formed rise faster and enter in the downcomer in lower amounts.

5. Conclusions

From the results presented, it can be concluded that the model predicts the experimental values found for both types of airlift reactors with high accuracy (with an error of $\pm$ 10%), despite the high number of simplifying assumptions introduced for calculations. Only for the internal-loop reactor some difficulties on estimation of the values for 30% of solids and low airflow rates were found.

For both reactors, for the estimated parameters, a similar effect of the solids loading on hydrodynamics was found. The distribution parameter presents some oscillations showing that, depending on the amount of solids, the solid-phase affects the flux in different ways. The terminal velocity of a single bubble increases with the increase of solids loading, as a consequence of the increase of coalescence deriving from the increase of the interaction between the bubbles. The parameter β exhibits different trends for the two reactors, resulting from their distinct geometries.

Notation

A	cross-sectional area, m^2
a, b	parameters
A_b	free area below the draught tube, m^2
C	distribution parameter
d_1,d_2	diameter of the tubes in a sudden contraction, m
D_p	solid particle diameter, m
D	diameter, m
f	friction factor of the pseudo-homogeneous mixture
g	gravitational acceleration, m/s^2
H	length, m
k_f	friction loss coefficient
P_h	hydrostatic pressure difference between riser and downcomer, Pa
Re	Reynolds number
U	superficial velocity, m/s
U_{bt}	terminal velocity of a single bubble, m/s

U_{st}	solids settling velocity, m/s
V	linear velocity, m/s

Greek letters

α	correction factor
β	parameter
$(-\Delta P_f)$	friction loss, Pa
$-\Delta P_{loss}$	total frictional loss in the reactor, Pa
ε	holdup
μ	viscosity, N s
ρ	density, kg/m^3

Subscripts

b	bottom section
d	downcomer
g	gas
gc	gradual contraction
h	homogeneous-phase
i	section
l	liquid
p	solid particle
r	riser
s	solid
sc	sudden contraction
t	top section

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