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Experimental study and modeling on liquid dispersion in external-loop airlift slurry reactors

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Abstract

Liquid dispersion in an external-loop airlift slurry reactor was experimentally studied. The effects of the superficial gas velocity, concentration of fine particle, flowing resistance, and liquid level in the gas–liquid separator on the liquid dispersion coefficient were investigated. A liquid dispersion model was proposed based on Taylor dispersion equation to predict the liquid dispersion coefficient. According to this model, the dispersion coefficient and the combination factor $u_L^2\varepsilon_L$ have a linear relationship. Validation of the liquid dispersion model together with the hydrodynamic model in our previous work was made by comparing the experimental and predicted results. The good agreement showed that the models could predict the liquid dispersion coefficient and the hydrodynamic behaviors of an airlift slurry reactor in a wide range of operating conditions with a satisfactory accuracy.

Keywords: Airlift slurry reactor; Liquid dispersion model; Dispersion coefficient

1. Introduction

External-loop airlift reactors are widely used in chemical, petrochemical, biochemical, and environmental processes. However, their design and scale-up are usually empirical due to the complexity of multiphase hydrodynamics, mixing and mass transfer behaviors. Mixing property of the airlift reactor is one of the most important design parameters (Fields & Slater, [1,25]). Good mixing is essential for some processes to enhance the chemical reaction rate, and decrease the side reaction caused by non-uniform profiles of concentration or temperature. The mixing behavior is also crucial for reactor scale-up and optimal operation.

In an airlift reactor, the gas phase exists as bubbles and its velocity is high so that the mixing of the gas phase is usually negligible and simplified as a plug flow [2]. Therefore, most studies on the mixing behavior focused on the liquid phase. The mixing of the liquid phase is mainly induced by turbulence, bubble rising and bubble wake. It was reported that the liquid concentration could be assumed uniform in the radial direction because the radial dispersion coefficients were much larger than the axial dispersion coefficients [3,4]. So this work focused on the axial dispersion neglecting the radial dispersion. To describe the liquid mixing, the dispersion coefficient, Peclet number, and mixing time were used in the literature [5]. The mixing time was used widely [6,7], but different definitions were used and caused difficulties of comparing the results by different investigators. The dispersion coefficient has a clear definition and was used to describe the liquid mixing behavior in this work. Some models based on the relation between the liquid dispersion coefficient and the operating conditions have been proposed in the literature, but most of them were limited in the gas–liquid system. Furthermore, the published models did not take into account the flowing resistance in an airlift reactor. In addition, several models on dispersion coefficient were proposed in the literature [8–10]. However, these modes were just presented in a general but complicated form, not for a specific system such as external-loop airlift slurry reactors.

In this work, the liquid dispersion in an external-loop airlift slurry reactor was experimentally studied. The effects of the superficial gas velocity, concentration of fine particles, flowing resistance, and liquid level in gas–liquid separator on the liquid dispersion coefficient were investigated. A liquid dispersion model was proposed based on Taylor dispersion equation to predict the liquid dispersion coefficient and validated by the

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Nomenclature

C	tracer concentration (mol/L)
D	axial dispersion coefficient (m^2/s)
U_L	average liquid velocity (m/s)
u_L	local liquid velocity (m/s)
u_0	liquid velocity in the center of reactor (m/s)
U_g	superficial gas velocity (m/s)
η	dimensionless radial distance ($= r/R$)
υ_t	turbulent viscosity coefficient (m^2/s)
g	gravity acceleration (m^2/s)
H	height (m)
P	pressure (Pa)
α	degree of angle between the valve and the vertical
d	column diameter (m)
s	cross-section of the riser/downer (m^2)
ε	phase holdup
n	exponent constant
ρ	density (kg/m^3)
κ	friction coefficient

Subscripts

b	bottom
t	top of the riser
d	downcomer
r	riser
g	gas
L	liquid
s	solid

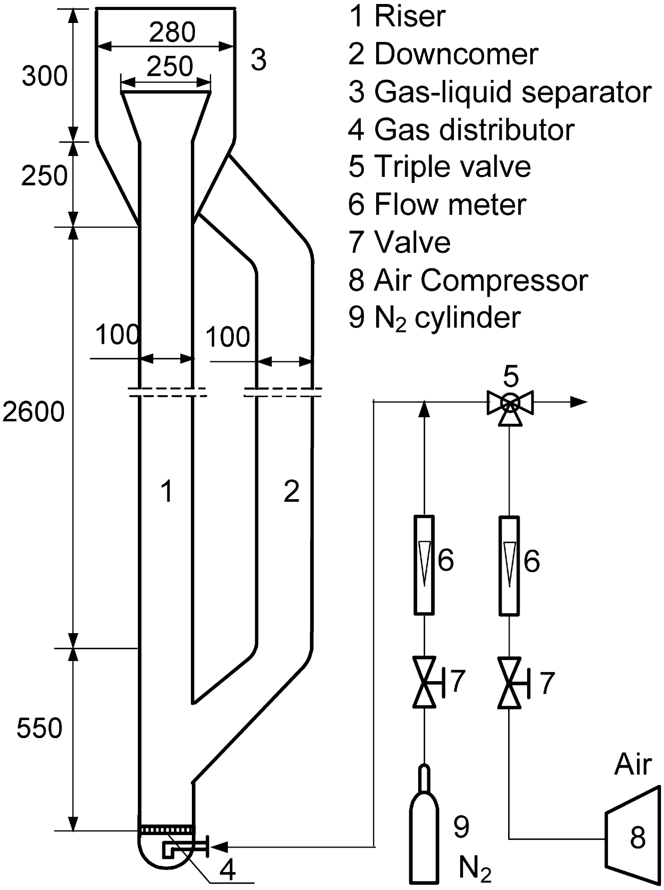

Fig. 1. The schematic diagram of the experimental apparatus.

experimental data. The liquid dispersion model and the hydrodynamic models proposed in our previous work [11] were validated and good agreements were achieved.

2. Experimental

The schematic diagram of the experimental setup is shown in Fig. 1. The external-loop airlift reactor was made of Plexiglas. It included three main parts: the riser with an inner diameter of 100 mm and a height of 3200 mm, the downcomer with an inner diameter of 100 mm, and a gas–liquid separator with an inner diameter of 280 mm.

The gas distributor at the bottom of the riser was a perforated plate with orifice diameter 1 mm and free area of 0.25%. Air, water, and FCC catalyst were used as the gas, liquid and solid phases, respectively. The average diameter and density of the solid particles are 60 μm and 2177 kg/m^3, respectively. The total volume of the liquid–solid slurry phase was kept 71 L at the beginning. The average solid concentration (ε_s) in this paper was defined as the volume fraction of the solid particle in the liquid–solid slurry. The average solid concentration (ε_s) can be calculated at the beginning directly. In order to study the effect of the flowing resistance on the flow behavior, a butterfly valve was installed in the downcomer. Different flowing resistance was realized by regulating the valve opening. The flowing resistance was expressed as the valve opening angle α ($\alpha = 0$ means full opening).

The gas holdup in the riser was measured with the manometric technique [12]. The tracer input-response technique was used to measure the liquid circulation velocity and the liquid axial dispersion coefficient. The KCl solution used as the tracer was injected into the reactor, and the response signal of the conductivity was sampled with an A/D card and stored in a computer. With the assumption that the tracer concentration C is uniform in the radial direction, the mass balance of the dispersed plug flow gives:

$$\frac{\partial C}{\partial t} + U_L \frac{\partial C}{\partial x} = D_L \frac{\partial^2 C}{\partial x^2} \qquad (1)$$

where D_L is the axial dispersion coefficient and U_L is the fluid axial velocity. Then the least-squares method was used to obtain the liquid circulation velocity and dispersion coefficient. The methods were described in detail in the previous paper [1].

3. Dispersion coefficient model

The Taylor dispersion model was proposed in the early studies [8,9]. In these studies, dispersion in a single-phase turbulent pipe flow was studied experimentally and theoretically. Vial et al. [13,14] used the mixing length model by analogy with the Taylor dispersion model to describe the liquid dispersion in an external-loop airlift reactor and obtained reasonable results. Therefore, the Taylor dispersion model was also employed as a basis in the present study. Taylor [8] proposed an axial dispersion model based on the turbulence theory as follows [10]:

$$D_{\mathrm{L}} = -2R^2 \int_0^1 \frac{(1-\varepsilon_{\mathrm{g}})}{\nu_{\mathrm{t}}}(u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta \times \int_1^\eta \frac{\mathrm{d}\eta}{\eta} \int_0^\eta (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta \tag{2}$$

where u_{L} is the local liquid velocity at different radial positions, $\eta = r/R$ the dimensionless radial distance, U_{L} the cross-sectional averaged axial liquid velocity, $1-\varepsilon_{\mathrm{g}} = \varepsilon_{\mathrm{Ls}}$ the solid–liquid holdup, and ν_{t} is the turbulent viscosity. For the gas–liquid–solid three-phase slurry system, several works showed that particles with typical particle sizes smaller than 100 μm are uniformly suspended in both the axial and radial directions in a slurry system under conditions of low solid concentrations [15,16]. The liquid and solid are treated as a pseudo homogeneous suspension here at low to medium solid concentrations, and liquid holdup (ε_{L}) takes the value of the solid–liquid holdup ($\varepsilon_{\mathrm{Ls}}$) in the following for the sake of simplicity.

The turbulent viscosity and liquid holdup in the radial direction can be considered as constant [17], because the radial profile of the gas holdup (ε_{g}, $\varepsilon_{\mathrm{L}} + \varepsilon_{\mathrm{g}} = 1$) is relatively uniform in an external-loop airlift reactor [18] and the gas holdup is much smaller than the liquids–solid holdup. Hence, the axial dispersion coefficient at radial position r is given by:

$$D_{\mathrm{L}} = 2R^2 \frac{\varepsilon_{\mathrm{L}}}{\nu_{\mathrm{t}}} \int_0^1 (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta \int_\eta^1 \frac{\mathrm{d}\eta}{\eta} \int_0^\eta (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta \tag{3}$$

Then Eq. (3) can be transformed into:

$$D_{\mathrm{L}} = 2R^2 \frac{\varepsilon_{\mathrm{L}}}{\nu_{\mathrm{t}}} \int_0^1 \frac{\mathrm{d}\eta}{\eta} \int_0^\eta (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta \int_0^\eta (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta \tag{4}$$

The radial profile of the liquid velocity was correlated as [19,20]:

$$u_{\mathrm{L}} = u_0(1 - \eta^n) \tag{5}$$

where u_0 is the liquid velocity in the center, n is empirical constant, dependent on both operating conditions and reactor structure parameters. Thus, the cross-sectional average liquid velocity is:

$$U_{\mathrm{L}} = u_0 2 \int_0^1 \eta(1 - \eta^n) \, \mathrm{d}\eta \tag{6}$$

The relationship between the center-line liquid velocity and average liquid velocity is:

$$u_0 = U_{\mathrm{L}} \frac{n+2}{n} \tag{7}$$

Hence,

$$u_{\mathrm{L}} - U_{\mathrm{L}} = (u_0 - U_{\mathrm{L}}) - u_0 \eta^n \tag{8}$$

The integration in Eq. (4) can be calculated as:

$$\int_0^\eta (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta = \int_0^\eta ((u_0 - U_{\mathrm{L}}) - u_0\eta^n)\eta \, \mathrm{d}\eta \tag{9}$$

$$\int_0^\eta (u_{\mathrm{L}} - U_{\mathrm{L}})\eta \, \mathrm{d}\eta = \frac{u_0 - U_{\mathrm{L}}}{2}\eta^2 - \frac{u_0}{n+2}\eta^{n+2} \tag{10}$$

Substituting Eq. (10) into Eq. (4) yields:

$$D_{\mathrm{L}} = R^2 \frac{\varepsilon_{\mathrm{L}}}{\nu_{\mathrm{t}}} \int_0^1 \left(\frac{(u_0 - U_{\mathrm{L}})^2}{4}\eta^3 - \frac{u_0(u_0 - U_{\mathrm{L}})}{n+2}\eta^{n+3} + \frac{(u_0)^2}{(n+2)^2}\eta^{2n+3} \right) \mathrm{d}\eta \tag{11}$$

Eq. (11) can be reduced as:

$$D_{\mathrm{L}} = R^2 \frac{\varepsilon_{\mathrm{L}}}{\nu_{\mathrm{t}}} \left(\frac{(u_0 - U_{\mathrm{L}})^2}{16} - \frac{u_0(u_0 - U_{\mathrm{L}})}{(n+2)(n+4)} + \frac{(u_0)^2}{2(n+2)^3} \right) \tag{12}$$

Substituting Eq. (7) into Eq. (12) leads to:

$$D_{\mathrm{L}} = R^2 U_{\mathrm{L}}^2 \frac{\varepsilon_{\mathrm{L}}}{\nu_{\mathrm{t}}} \left(\frac{1}{4n^2} - \frac{2}{n^2(n+4)} + \frac{1}{2n^2(n+2)} \right) \tag{13}$$

As a result, the axial dispersion coefficient can be expressed as:

$$D_{\mathrm{L}} = \frac{R^2}{4\nu_{\mathrm{t}}(n+4)(n+2)} U_{\mathrm{L}}^2 \varepsilon_{\mathrm{L}} \approx k U_{\mathrm{L}}^2 \varepsilon_{\mathrm{L}} \tag{14}$$

Generally, n decreased with an increase in the superficial gas velocity. In contrast, the turbulent viscosity coefficient ν_{t} increased with an increase in the superficial gas velocity. So $k = R^2/(4\nu_{\mathrm{t}}(n+4)(n+2))$ is approximately constant in a certain range of conditions. The dispersion coefficient and the combination factor $u_{\mathrm{L}}^2 \varepsilon_{\mathrm{L}}$ have a linear relationship which can be validated by experiments, the relationship can be used to estimate the dispersion coefficient conveniently if the radial profile of the liquid velocity is given.

The value of k can be estimated from literatures [13,20–22]. From Fig. 2 in the literature [20], n had values of 2.37 and 1.99 in a reactor of 0.089 m in diameter when the superficial gas velocity was 0.96 and 8.4 cm/s, respectively. In a reactor of 0.14 m in diameter, n had values of 1.85 and 1.59 at superficial gas velocity of 0.96 and 8.4 cm/s, respectively. In a reactor of 0.23 m, n had values of 2.02 and 1.91 at superficial gas velocity of 1.6 and 3.2 cm/s, respectively [21]. In a reactor of 0.15 m, n had values of 1.98 and 1.67 at superficial gas velocity of 2.3 and 11.7 cm/s, respectively [13]. It can be seen that n decreased slightly with the superficial velocity increasing. Unfortunately, the radial profiles of liquid velocity were not measured when the superficial gas velocity was from 5 to 35 cm/s and the reactor diameter is 0.1 m, but from the above discussions, the average value of n in the present study can be approximately assigned as 1.5. From Figs. 7 and 8 in the literature [22], it could be seen that the turbulent

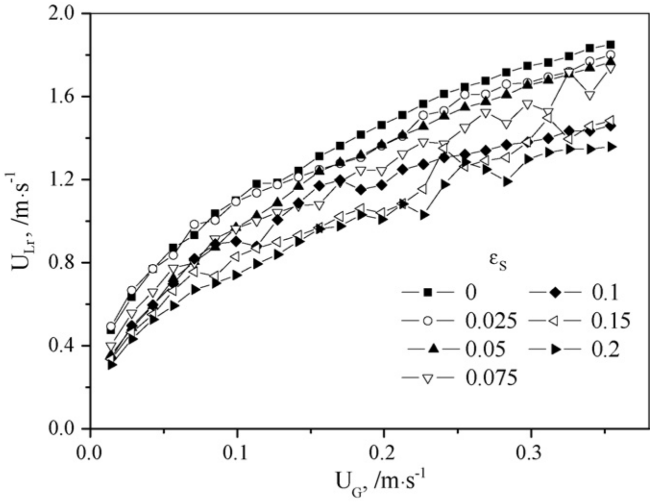

Fig. 2. Effect of the superficial gas velocity on the liquid circulation velocity at different solid holdups.

viscosity ν_t almost kept constant when the superficial gas velocity changed from 5 to 35 cm/s, but had great dependence upon the bubble column diameter *2R*. According to the results reported by Ueyama and Miyauchi [22], the turbulent viscosity ν_t in the present study can be assigned 0.0006 m^2/s (U_g = 5–35 cm/s, $2R = 0.1$ m). So $k = R^2/(4\nu_t(n+4)(n+2))$ is approximately constant in a certain range of conditions, and should has a value of about 0.055, with $R = 0.05$ m, $\nu_t = 0.0006$ m^2/s and $n = 1.5$ from the above discussion.

4. Results and discussion

To validate the model, the relationship between the liquid dispersion coefficient, circulation liquid velocity and the holdup of the liquid–solid homogenous phase at different operating conditions were studied. The effects of the solid holdup, valve opening and the liquid height on the liquid dispersion coefficient were experimentally measured.

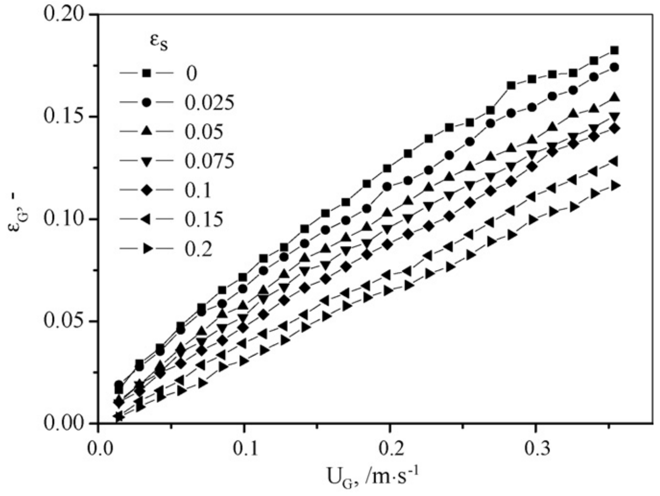

Fig. 3. Effect of the superficial gas velocity on the gas holdup at different solid holdups.

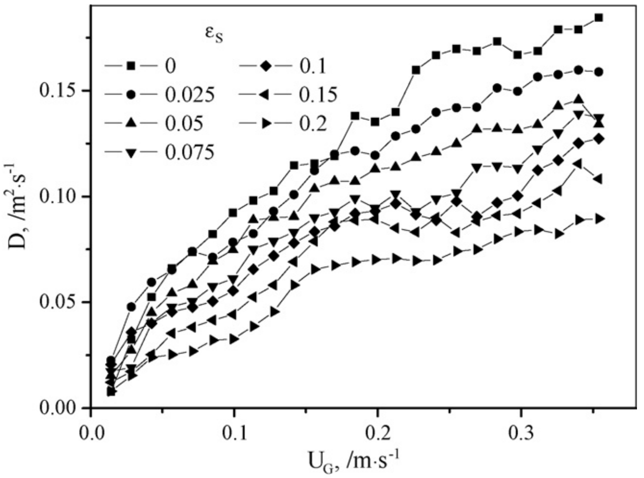

Fig. 4. Effect of the superficial gas velocity on the liquid dispersion coefficient at different solid holdups.

The effects of the superficial gas velocity and solid holdup on the liquid circulation velocity, gas holdup and liquid dispersion coefficient were shown in Figs. 2–4, respectively. The liquid velocity, gas holdup and liquid dispersion coefficient increased monotonically with an increase in the superficial gas velocity, but decreased with an increase in the solid holdup. Li et al. [23] also found that the liquid dispersion coefficient increased with increasing superficial gas and liquid velocities. However, they did not consider the effect of the solid holdup on the liquid dispersion. In our experiments, the decrease of the liquid dispersion coefficient with increasing solid holdup was caused by the decrease of liquid circulation velocity with increasing solid holdup, which in turn decreased the turbulent intensity and liquid dispersion.

The effects of the superficial gas velocity and flowing resistance on the liquid circulation velocity, gas holdup and liquid dispersion coefficient were shown in Figs. 5–7. The gas holdup increased monotonically with an increase in the flowing resistance, but the liquid circulation velocity and liquid dispersion

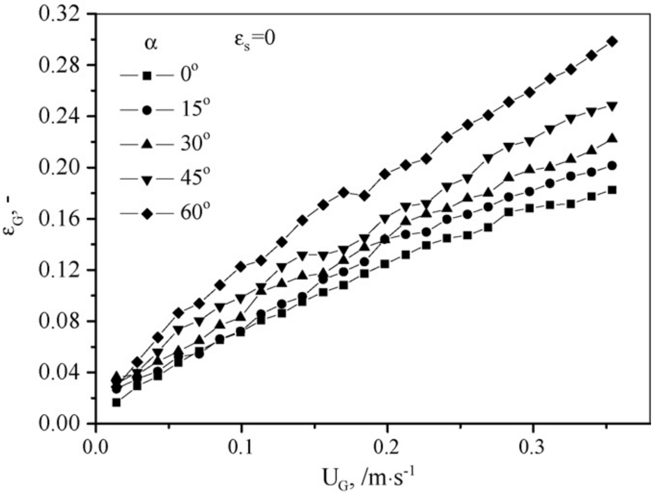

Fig. 5. Effect of the superficial gas velocity on the gas holdup with different flowing resistance.

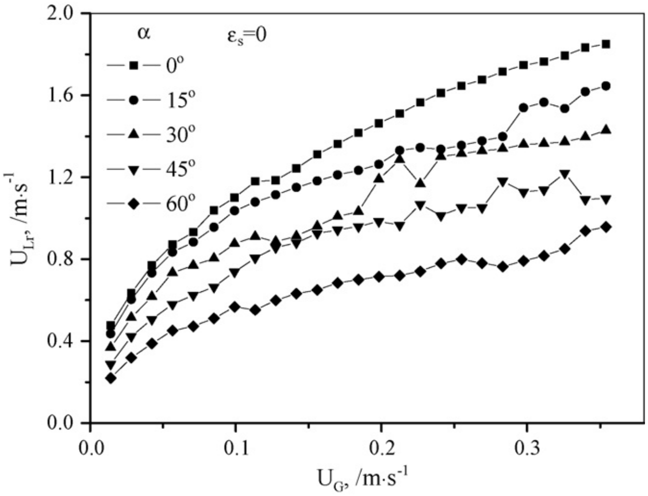

Fig. 6. Effect of the superficial gas velocity on the liquid circulation velocity with different flowing resistance.

coefficient decreased with increasing flowing resistance. Generally speaking, the liquid dispersion coefficient is dependent on the gas holdup and liquid circulation velocity. When the flowing resistance increased, the liquid circulation velocity decreased and the gas holdup increased, both tend to decrease the liquid dispersion coefficient.

The effects of the superficial gas velocity and liquid level in the gas–liquid separator on the liquid circulation velocity, gas holdup and liquid dispersion coefficient were shown in Figs. 8–10. It was found that the gas holdup decreased monotonically with the liquid level, but the liquid circulation velocity and liquid dispersion coefficient increased with the liquid level. With increasing liquid level, the flowing resistance at the top section decreased, which in turn increased the liquid circulation velocity and liquid dispersion coefficient. The similar results were found by Lu et al. [6] in internal-loop airlift reactors.

All the liquid dispersion coefficients at different operating conditions were shown in Figs. 11–13. The following relation-

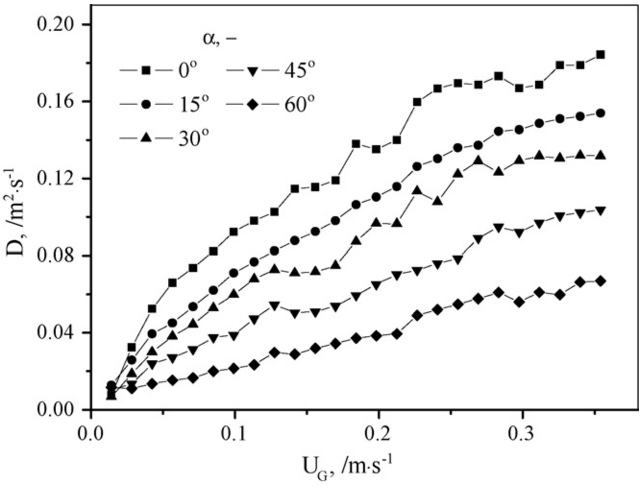

Fig. 7. Effect of the superficial gas velocity on the liquid dispersion coefficient with different flowing resistance.

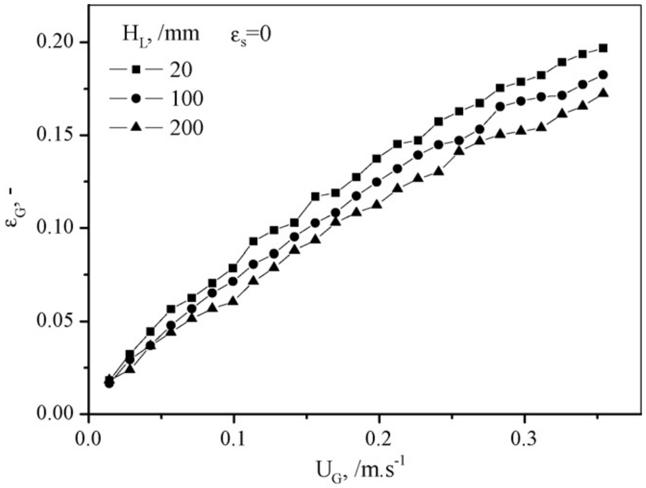

Fig. 8. Effect of the liquid level in the gas–liquid separator on the gas holdup.

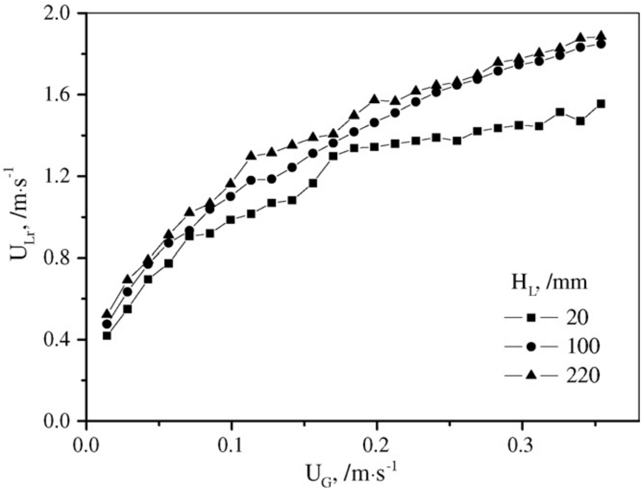

Fig. 9. Effect of the liquid level in the gas–liquid separator on the liquid circulation velocity.

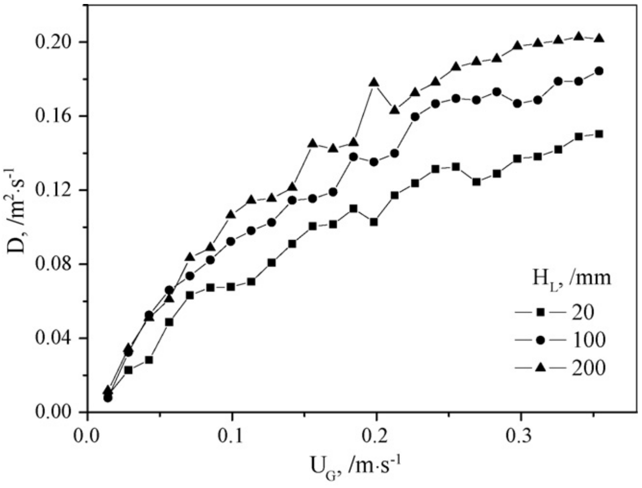

Fig. 10. Effect of the liquid level in the gas–liquid separator on the liquid dispersion coefficient.

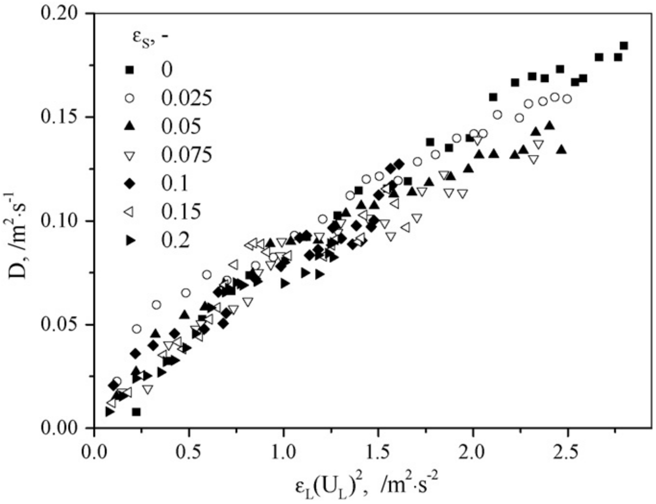

Fig. 11. Relationship between D and $\varepsilon_L(U_L)^2$ at different concentrations of fine particles.

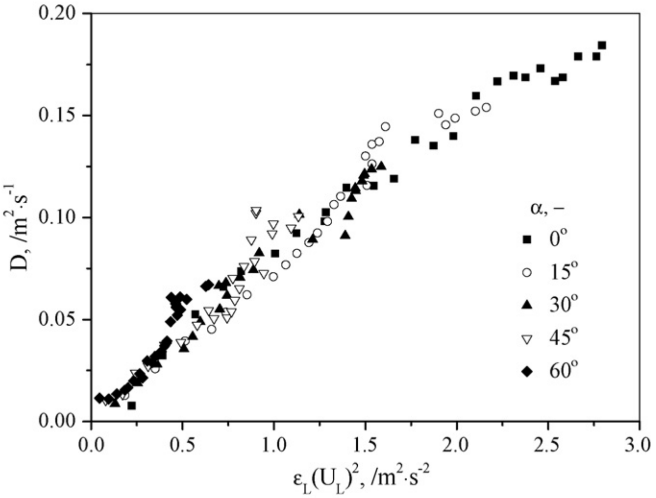

Fig. 12. Relationship between D and $\varepsilon_L(U_L)^2$ at different flowing resistance.

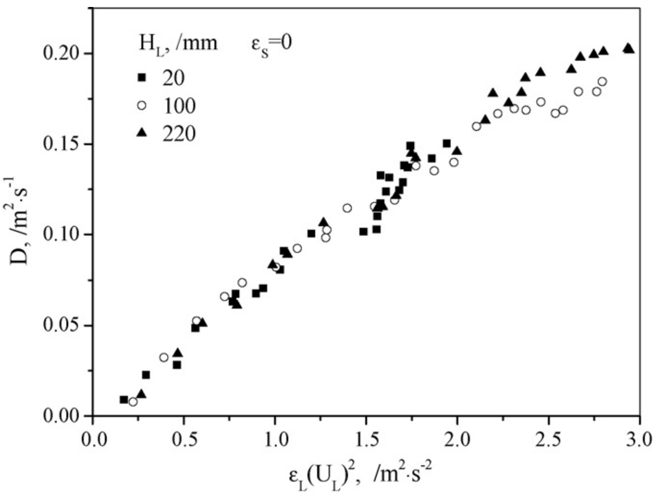

Fig. 13. Relationship between D and $\varepsilon_L(U_L)^2$ at different liquid levels in the gas–liquid separator.

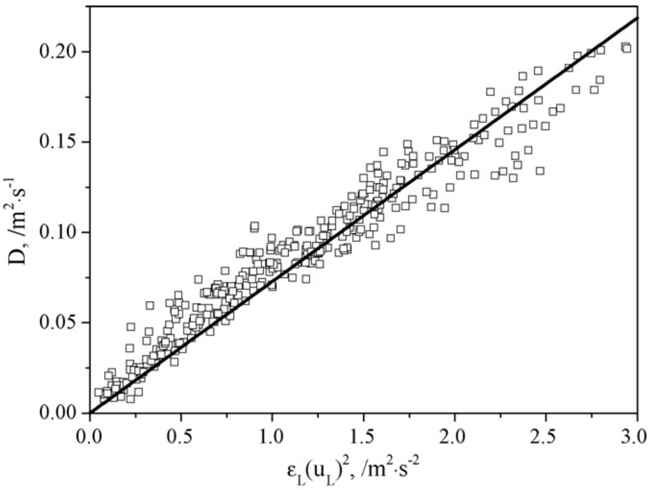

Fig. 14. The regressive results of the liquid dispersion coefficient.

ship was obtained with the linear least-square method:

$$D_L = 0.072 U_L^2 \varepsilon_L \tag{15}$$

The coefficient in Eq. (15) is close to the value estimated from literatures, showing that the proposed model can give a good prediction of the liquid dispersion coefficient. Fig. 14 shows that the correlation agrees with the experimental data satisfactorily.

The proposed model was validated by the other experimental results in the literature. By fitting the data given by Vial et al. [14], the correlation $D_L = 0.224\ u_L^2 \varepsilon_L$ was achieved, with the relative coefficient of 0.984, as shown in Fig. 15. The linear relationship of $u_L^2 \varepsilon_L$ and D_L was also found.

Other methods for calculation of the mixing time and recirculation liquid velocity from the tracer response were reported in the literature [24]. However, it is the first time to propose a model that shows a linear relationship between the liquid dispersion coefficient and the combination factor $u_L^2 \varepsilon_L$. This model shows

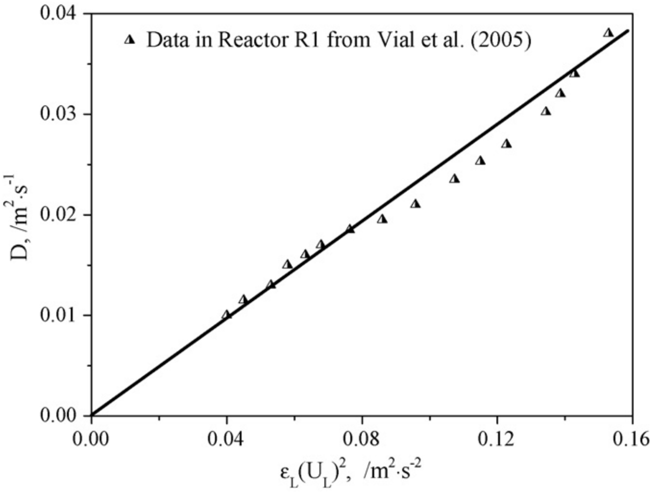

Fig. 15. The correlation between $U_L^2 \varepsilon_L$ and D_L (data from [14], $D_L = 0.224\ U_L^2 \varepsilon_L$, the relative coefficient is 0.984).

that the dispersion coefficient is linear with the liquid holdup, and quadratic with the liquid velocity. This means that the liquid dispersion is more dependent on the liquid velocity than the phase holdup, which was confused in the literature, especially when explaining why the liquid dispersion coefficient increased with the superficial gas velocity.

5. Model validation

In this section, the liquid dispersion model was validated together with the hydrodynamic model by comparing with the experiment results. The hydrodynamic models for prediction of the liquid circulation velocity and the gas holdup in an external-loop airlift reaction model were based on the momentum and energy balance [11], and the final form was:

$$P_b U_g \ln\left(1+\frac{\rho_h g H}{P_t}\right) = \frac{1}{2}k_1\frac{H}{s}\rho_h U_L^2(U_L+0.4(1+20\varepsilon_s)) \qquad (16)$$

$$\varepsilon_g = \frac{P_b U_g \ln(1+(\rho_h g H/P_t))}{\rho_h g H(U_L+0.4(1+20\varepsilon_s))} \qquad (17)$$

where P_b and P_t are the pressure at the bottom and the top, respectively, ε_s the solid holdup in the riser, and k_1 an frictional coefficient which can be determined with Fanning equation, U_g the superficial gas velocity, H the reactor height, s the ratio between the cross-section areas of the riser and downcomer, and g is the acceleration of gravity. The liquid circulation velocity calculated by Eq. (16) and the gas holdup by Eq. (17) are substituted into Eq. (15), and the predicted results are compared with the experimental data to validate the models.

The comparison of the measured and predicted liquid dispersion coefficients at different concentrations of fine particles is shown in Fig. 16. The satisfactory agreement shows that Eqs. (15)–(17) can predict the dispersion coefficient and the hydrodynamic behavior of an airlift slurry reactor operated at different concentration of fine particles satisfactorily.

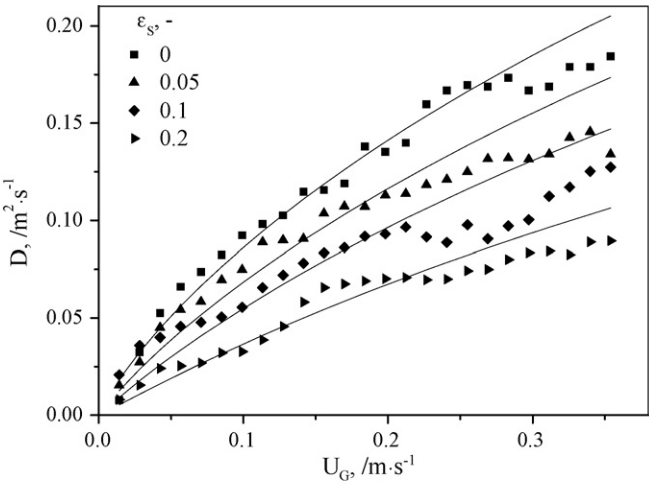

Fig. 16. Comparison of experimental and predicted values of dispersion coefficient at different concentrations of fine particles.

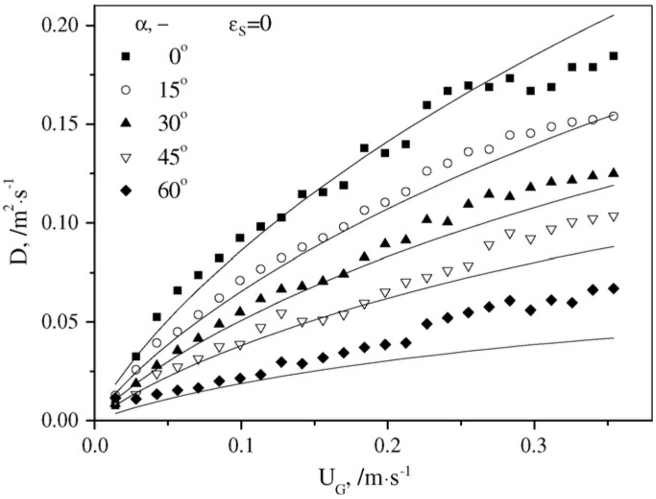

Fig. 17. Comparison of experimental and predicted values of dispersion coefficient with different flowing resistance.

At different flowing resistances, comparison of the measured and predicted liquid dispersion coefficients is shown in Fig. 17. The agreement is satisfactory at a large valve opening. However, the deviation is relatively large at a small vale opening. With the valve closed, the liquid circulation velocity will become zero. Such a case is out of the application range of the liquid dispersion model proposed in this work. Further study is needed to approach a unified description of the liquid dispersion both in a bubble column and an airlift reactor.

Comparison of measured and predicted liquid dispersion coefficients with different liquid levels is shown in Fig. 18. Good agreement was obtained, indicating that Eqs. (15)–(17) can give good predictions of the dispersion coefficient and the hydrodynamic behavior of an airlift slurry loop reactor operated at different liquid levels in the gas–liquid separator.

Comparison of the measured and predicted liquid dispersion coefficients in all operating conditions is shown in Fig. 19. The results show that Eqs. (15)–(17) can predict the liquid disper-

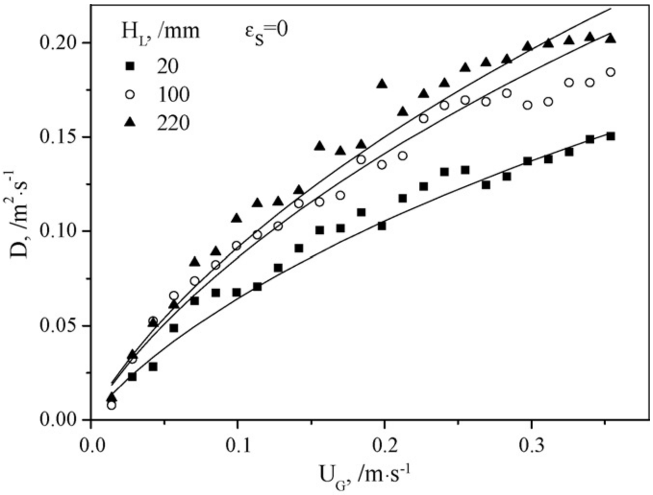

Fig. 18. Comparison of experimental and predicted values of dispersion coefficient at different liquid levels.

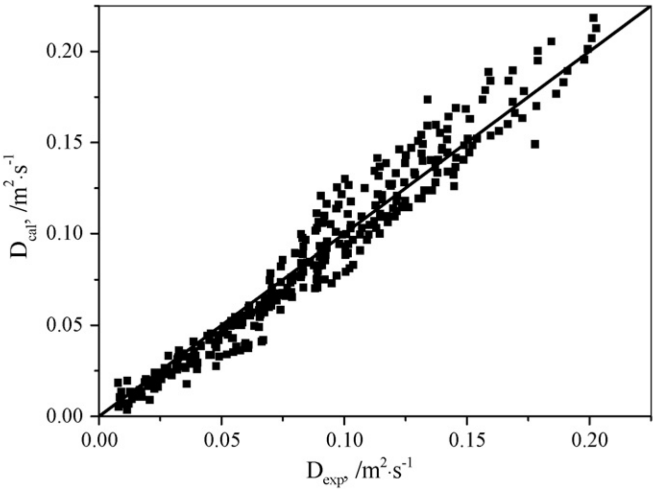

Fig. 19. Comparison of experimental and predicted values of dispersion coefficient in all operating conditions.

sion coefficients and the hydrodynamic behaviors of an airlift slurry loop reactor in a wide range of operating conditions with a satisfactory accuracy.

6. Conclusions

The effects of the superficial gas velocity, concentration of fine particles, flowing resistance, liquid level in the gas–liquid separator on the liquid dispersion coefficient in an external-loop slurry reactor were experimentally studied. A liquid dispersion coefficient model was proposed based on Taylor dispersion equation. The model was validated by comparing with the experimental results. The study leads to the following conclusions:

(1) The liquid dispersion coefficient model based on the Taylor dispersion equation for an external-loop airlift slurry reactor showed that the liquid dispersion coefficient D_{L} was linear with the combination factor $U_{\mathrm{L}}^2\varepsilon_{\mathrm{L}}$. The coefficient obtained by experiments is close to the value estimated from literatures.

(2) The liquid dispersion coefficient increase with an increase in the superficial gas velocity, liquid velocity, or liquid level in the gas–liquid separator, and decreased with an increase in the concentration of fine particles or flowing resistance.

(3) The proposed liquid dispersion model together with the hydrodynamic model was validated by comparing with the experimental results. The good agreement shows that the models can predict the liquid dispersion coefficient and the hydrodynamic behaviors of an airlift slurry loop reactor in a wide range of operating conditions with satisfactory accuracy.

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