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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)
v_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

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

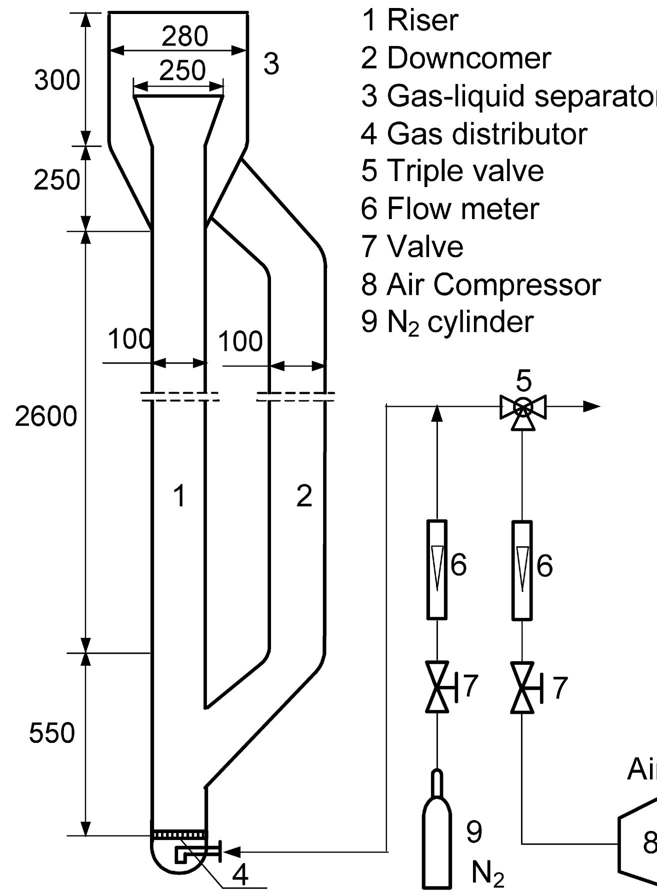

Fig. 1. The schematic diagram of the experimental apparatus.

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:

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

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