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Gas–liquid mass transfer in co-current three-phase fluidized beds with non-Newtonian fluids: Theoretical models based on the energy dissipation rate

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ABSTRACT

The gas–liquid mass transfer in the co-current three-phase fluidized bed with non-Newtonian liquids was studied. Gas hold-ups (ϕ_g) and volumetric gas–liquid mass transfer coefficients (k_La) in a bed of glass beads fluidized in non-Newtonian liquids with gas phase were measured. The gas hold-ups affecting gas–liquid mass transfer rate increased and decreased with increasing particle size and liquid velocity, respectively. The k_La coefficients were evaluated from the dynamic gassing-out method based on the tanks-in-series model describing non-ideal mixing in the three-phase fluidized bed. While an increase in the liquid velocity decreased the k_La coefficient, an increase in the particle size enhanced the k_La coefficient. The increase in purely viscous non-Newtonian flow behaviors reduced gas hold-up and gas–liquid mass transfer rate because of the bubble coalescing nature of the highly viscous non-Newtonian fluid. The theoretical models for ϕ_g and k_La in three-phase fluidized beds with non-Newtonian fluids were developed on the basis of the energy dissipation rate. They could reasonably fit the present experimental data.

1. Introduction

Three-phase fluidized beds, in which solid particles are fluidized by co-current, upward flow of liquid and gas, are considered to be one of the vital techniques of multiphase flow contacting operation and widely used in the chemical, petrochemical and biochemical industries [1]. During recent years, therefore, considerable progress has been made in exploring and understanding the hydrodynamics and heat and mass transfer in gas–liquid–solid three-phase fluidized beds (e.g. [1–4]). Although several fluidized beds operate with non-Newtonian liquids in food and polymer processing and biotechnology, most of the work on three-phase fluidized beds has been restricted to Newtonian liquid systems. Some attempts have been conducted to elucidate hydrodynamics and heat and mass transfer in three-phase fluidized beds with non-Newtonian liquids (e.g. [5–8]). The gas hold-up and gas–liquid mass transfer rate are most important design parameters in the modeling of three-phase fluidized bed reactors. In spite of several studies, gas hold-up and gas–liquid mass transfer in three-phase fluidized beds are not completely understood yet. Hence, it is necessary to study the influence of non-Newtonian flow behaviors on the gas hold-up and gas–liquid mass transfer in three-phase fluidized beds.

Patwari et al. [9] and Schumpe et al. [10] measured the gas hold-up (ϕ_g) and volumetric gas–liquid mass transfer coefficient (k_La) in a co-current fluidized bed of glass spheres (3–8 mm diameter) with pseudoplastic liquids and found significant decrease in the k_La coefficients with increasing effective viscosity. Schumpe et al. [10] found that increasing the liquid velocity reduced the solids hold-up and then increased the k_La coefficient. They proposed an empirical correlation for the k_La coefficient. Kang et al. [11] examined oxygen transfer in three-phase fluidized-bed bioreactors with floating bubble breakers and found that the k_La coefficient decreased with increasing effective viscosity of pseudoplastic liquids but its effect was rather smaller as compared with that observed by Schumpe et al. [10]. Lee et al. [12] measured gas hold-ups and k_La coefficients in a co-current fluidized bed with carboxymethyl cellulose aqueous solutions (CMC). They found that the gas hold-up increased with gas velocity but decreased with liquid velocity. They also found that the k_La coefficient increased with gas and liquid velocities and particle size. Tavakoli et al. [13] studied the k_La coefficient in a co-current three-phase fluidized bed with non-Newtonian liquids and found that solid particles enhanced the bubble coalescence and as a result caused the reduction of the k_La value. However, they did not discuss the effects of liquid velocity on k_La coefficient. In their empirical correlation, therefore, the liquid velocity is not included. Chen and Leu [14] measured gas hold-up and k_La coefficient in a co-current fluidized bed with CMC aqueous solutions. In their study, the superficial liquid flow rate was kept at 0.02 m s^{-1} and hence effects of liquid velocity on the

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Nomenclature

A	cross-sectional area of bed (m^2)
a	specific surface area (m^{-1})
C	dissolved oxygen concentration ($mg\,L^{-1}$)
C'	proportionality constant in Eq. (23)
D	diffusivity ($m^2\,s^{-1}$)
D_c	bed diameter (m)
d_p	particle diameter (m)
E_z	axial dispersion coefficient ($m^2\,s^{-1}$)
$f(\phi_s)$	function in apparent consistency index
g	gravitational acceleration ($m\,s^{-2}$)
H_e	Henry's constant
K	consistency index in power-law model ($Pa\,s^n$)
$\bar{K}$	apparent consistency index in power-law model of solid–liquid two-phase system ($Pa\,s^n$)
k	roundup of j/P
k_L	liquid-phase mass transfer coefficient ($m\,s^{-1}$)
L	bed height (m)
M_s	total solid mass in bed (kg)
N_G	number of hypothetical tanks for the gas phase
N_L	number of hypothetical tanks for the liquid phase
n	flow index in power-law model
P	ratio of the number of tanks for the gas phase to that for the liquid phase ($=N_G/N_L$)
Q_L	volumetric liquid flow rate ($m^3\,s^{-1}$)
t	time (s)
U_g	superficial gas velocity ($m\,s^{-1}$)
U_l	superficial liquid velocity ($m\,s^{-1}$)
U_{lmf}	minimum fluidization liquid velocity ($m\,s^{-1}$)
V_L	liquid volume (m^3)
Y	oxygen gas concentration ($mg\,L^{-1}$)

Greek letters

ε	energy dissipation rate per unit mass of liquid ($W\,kg^{-1}$)
$\bar{\varepsilon}$	apparent energy dissipation rate per unit mass of pseudo-homogeneous liquid ($W\,kg^{-1}$)
ϕ_g	gas hold-up
ϕ_s	solids hold-up in gas–liquid–solid three-phase system
ϕ'_s	solids hold-up in solid–liquid two-phase system
$\dot{\gamma}$	shear rate (s^{-1})
ρ_l	liquid density ($kg\,m^{-3}$)
$\bar{\rho}_l$	mean density of solid–liquid two-phase system ($kg\,m^{-3}$)
ρ_s	solid density ($kg\,m^{-3}$)
σ	surface tension ($N\,m^{-1}$)
τ	shear stress (Pa)

Subscripts

air	air
i	*i*-th tank
in	inlet
j	*j*-th tank
sat	saturated

k_La coefficient were not examined as well as the study of Tavakoli et al. [13]. The significant decreases in k_La with increasing effective viscosity were observed and the empirical correlations for ϕ_g and k_La were proposed. Recently, Sivasubramanian [15] measured the gas–liquid volumetric mass transfer coefficients with CMC aqueous liquids in a three-phase inverse fluidized-bed. Particles of two different densities, PP (polypropylene: $\rho_s = 830\,kg\,m^{-3}$) and LDPE (low-density polyethylene: $\rho_s = 940\,kg\,m^{-3}$) having diameter of 4, 6 and 8 mm were used. The k_La coefficients were found to increase with decreasing the liquid velocity. The empirical correlations for Newtonian and non-Newtonian systems were proposed, respectively. According to their empirical correlations the k_La coefficient increased with an increase in particle size. It is clear from the literature survey that rather limited work has been carried out on the gas–liquid mass transfer in three-phase fluidized beds with non-Newtonian fluids. Particularly, still the results for effects of liquid velocity and particle size are diverse in the literature. Furthermore, all proposed correlations for ϕ_g and k_La are purely empirical. In the correlations, non-Newtonian flow behaviors are taken into account by introducing effective viscosity concepts. Their physical meanings are very vague. Unfortunately, they cannot provide fundamental insight into the mechanism of gas–liquid mass transfer in three-phase fluidized beds.

The objective of this study is to examine effects of liquid velocity and particle size on gas–liquid mass transfer in co-current three-phase fluidized beds with non-Newtonian fluids. In this study, the dynamic response of dissolved oxygen concentration (DO) based on the tanks-in-series model has been applied instead of the steady-state method which has been widely used but requires troublesome measurements of axial dissolved oxygen concentration profiles. The proposed dynamic method requires DO concentration measurement only at one monitoring point in spite of the fact that it takes account of non-ideal mixing in a three-phase fluidized bed. Due to the complexity of the flow characteristics, modeling of gas–liquid mass transfer in a co-current three-phase fluidized bed remains difficult. On the basis of the energy dissipation rate concept, the theoretical correlations for gas hold-up and volumetric gas–liquid mass transfer coefficient in co-current gas–Newtonian or non-Newtonian liquid–solid three-phase systems were developed and their predictions were compared with the present experimental data.

2. Experimental

2.1. Experimental setup

A schematic diagram of the experimental setup for gas–liquid mass transfer in the co-current three-phase fluidized bed is shown in Fig. 1. The column was constructed of polyvinyl chloride resin with a length of 1.1 m and an inner diameter of 0.068 m. The gas and liquid streams were merged in the bottom section of the column and injected through a perforated plate gas–liquid distributor containing 89 holes of 1 mm diameter at the bed bottom. The ranges of superficial gas and liquid velocities used in this study were $U_g = 0.046$–$0.14\,m\,s^{-1}$ and $U_l = 0$–$0.065\,m\,s^{-1}$, respectively. Gas and liquid were disengaged at the column top. The liquid was overflowed into the annulus of the upper reservoir and returned to the liquid storage tank. The liquid was recirculated to the fluidized bed by the pump. The flow rates of gas and liquid were measured and controlled with calibrated rotameters. All particles were retained in the column by the screen attached to the top of the column.

2.2. Solid particles and liquid

The bed particles were 3, 5 and 7 mm diameter glass beads with a density of $2500\,kg\,m^{-3}$. Tap water was used as a Newtonian liquid. Aqueous solutions of carboxymethyl cellulose (CMC) and xanthan gum (XG) exhibited purely viscous non-Newtonian flow behaviors described by a power-law model:

$$\tau = K\dot{\gamma}^n \quad (1)$$