Refereed Proceedings

The 13th International Conference on Fluidization - New Paradigm in Fluidization Engineering

Engineering Conferences International   *Year* 2010

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EFFECTS OF LIQUID SWIRLING ON GAS-TO-LIQUID MASSTRANSFER IN THREE-PHASE FLUIDIZED BEDS

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Abstract

The swirling flow mode of liquid phase was adopted to promote the gas-to-liquid mass transfer in three-phase(gas-liquid-solid) fluidized beds. Effects of gas(0.01-0.09m/s) and liquid(0.035-0.172m/s) velocities, particle size(1.7-6.0mm) and swirling ratio of liquid phase(0-0.5) on the volumetric gas-to-liquid mass transfer coefficient in the bed were examined. The mass transfer coefficient increased up to 70% by adjusting the swirling flow of liquid phase, especially when the gas velocity is relatively low range. The value of gas-to-liquid mass transfer coefficient was well correlated in terms of dimensionless groups which were derived from the dimensional analysis on the mass transfer system.

Introduction

Three-phase fluidized-bed can be utilized in various kinds of processes and reaction systems since the process has numerous advantages, such as higher chemical reaction rate due to effective contacting among the reacting phase with higher heat and mass transfer rates, and ease of continuous operation compared to the other contacting mode. The principal application is for gas-liquid reaction in which catalyst is used to enhance the reactions. The Fisher-Tropshch process, coal liquefaction, the hydrogenation of liquid petroleum fractions, the hydrogenation of unsaturated fats are example of such process(1-4). It is understood that the information on oxygen transfer is one of the essential factors for the design, scale up and operation of the environmental and biochemical reactors or contactors employing the three-phase fluidized beds. In three-phase fluidized, the mass transfer coefficient can be affected by the gas holdup and residence time of bubbles(5-7). The formation of swirling flow of gas-liquid mixture in the bed can be one of the promise schemes to increase the residence time of bubbles by increasing the travel path in the bed. The swirling flow of gas-liquid mixture could be realized by injection of swirling liquid into the bed at the wall of the column tangentially(7-8). It could be understood that the increase of residence time of bubbles could lead to the increase of contacting between the bubbles and the continuous liquid medium. The swirling flow of gas-liquid mixture in the riser could increase the turbulence intensity in the bed(8-9). In addition, the swirling flow of continuous liquid medium could lead to the increase in

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the periodicity and stability of the hydrodynamic behavior in the system, which is a highly important aspect in view of control and design of the dynamic systems.
However, there has been little attention on the enhancement of gas-to-liquid mass transfer in three-phase fluidized beds. Therefore, this study was intended to enhance the gas-to-liquid mass transfer in the bed by means of swirling flow of continuous liquid medium.

Experimental

Experiments were carried out in a three-phase fluidized bed with a mode of liquid swirling, which can be called a swirling fluidized bed, as can be seen in Fig. 1(8-9). The diameter and height of the column were 0.102 m ID and 3.5 m, respectively. A glass bead whose diameter is either 1.7, 3.0, 4.0 or 6.0 mm (ρ_s = 2,500 kg/m^3) was used as a fluidized solid particle, compressed filtered air as a gas phase, and tap water was used as a continuous liquid medium, respectively. The swirling liquid, which is a secondary liquid stream, was injected into the bed tangentially through the injection port which was installed at the wall of the column at 0.2 m from the distributor. The ratio of swirling liquid to the primary liquid stream, which was injected from the bottom of the bed through a liquid calming section and a distributor, was determined by means of volumetric flow ratio of swirling liquid based on the total volumetric flow introduced into the bed. The swirling liquid ratio was in the range of 0–0.5 by volume flow. Although the swirling ratio of liquid phase was changed the total amount of liquid phase was same in a given superficial liquid flow rate.

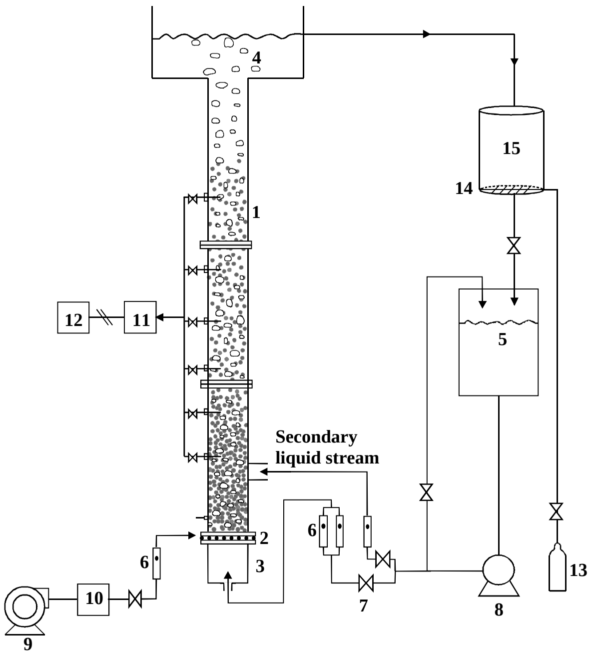

Fig. 1 Experimental apparatus:

1. Fluidized bed
2. Distributor
3. Liquid calming section
4. Liquid weir
5. Liquid reservoir
6. Flowmeter
7. Control valve
8. Liquid pump
9. Air compressor
10. Gas filter & regulator
11. Sample vessel
12. DO meter
13. N_2 Tank
14. N_2 injection distributor
15. Purge Tank

To obtain the volumetric gas-to-liquid mass transfer coefficient in the bed, the steady-state axial dispersion model was employed. Based on the model, the oxygen balance in the liquid phase around a differential volumetric section of the bed was yielded with appropriate boundary conditions. The values of volumetric gas-to-liquid mass transfer coefficient, k_La, were estimated by fitting the analytical solution of the model equation to the concentration profiles of dissolved oxygen in the axial direction of the column(10-14).

Results and Discussion

Typical concentration profiles of dissolved oxygen in the liquid phase in the three-phase swirling fluidized bed can be seen in Fig. 2, which were well fitted to the axial dispersion model(10-14). The value of volumetric mass transfer coefficient (k_La) was recovered from the concentration profile in the bed.
Effects of gas velocity on the mass transfer coefficient in the three-phase fluidized bed with liquid swirling can be seen in Fig. 3. In this figure, the values of mass transfer coefficient increased with increasing gas velocity. This is primarily due to the increase of contact area for gas-to-liquid transfer arising from the increase of gas holdup with increasing U_G(1-4, 15-17). In addition, the increase of gas holdup leads to the increase of turbulence in the beds, which can affect the gas-to-liquid mass transfer positively, since micro-eddies can be formed easily in splitting bubbles in the beds. It has been understood that the micro-eddies can be responsible for the increases in the surface renewal rate and turbulence at the gas–liquid contact area, which can be result in the consequent increase in the gas-to-liquid transfer coefficient[15]. In Fig. 3, the values of mass transfer coefficient of this study were comparable with those obtained in three phase beds without liquid swirling in the literature(17).

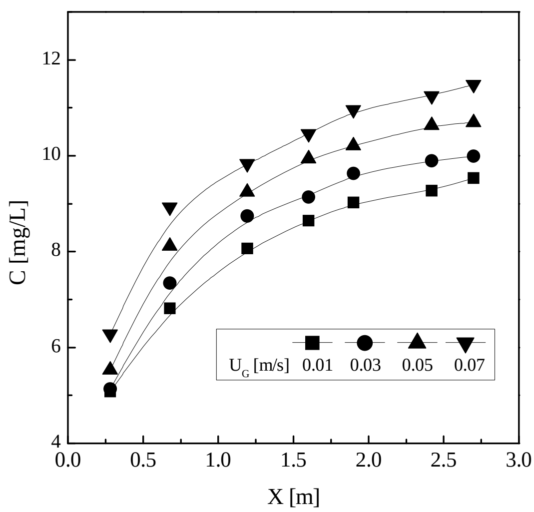

Fig. 2. Typical examples of dissolved oxygen concentration profile in the axial direction of the bed (U_L=0.103 m/s, d_p=3.0 mm, R_S=0.3)

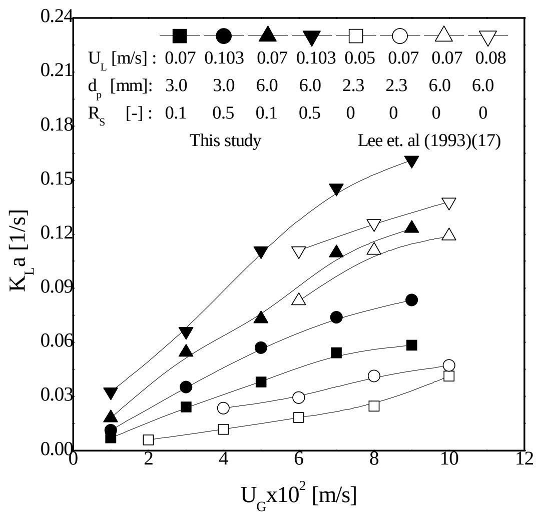

Fig. 3. Effects of gas velocity on the k_La in the three-phase swirling fluidized beds

Effects of liquid velocity on the gas-to-liquid mass transfer coefficient in the three-phase swirling fluidized bed can be seen in Fig. 4. In this figure, the mass transfer coefficient exhibits a local maximum with a variation of liquid velocity in the bed. In the relatively low range of liquid velocity, an increase in liquid velocity could promote the bed turbulence by means of vigorous solid suspension that may effectively break rising bubbles in the bed. Thereby, the gas/liquid interfacial area could increase due to a decrease in bubble size, and consequently results in an increase in gas-to-liquid mass transfer with increasing liquid velocity(4, 15). However, at the higher liquid velocity, the fluctuations of solid particles fluidized by liquid flow might decrease due to the considerable reduction of solid holdup in the bed, owing to the high bed expansion, with increasing liquid velocity. Moreover, the bubble rising velocity can increase with increasing liquid velocity, and bring about a reduction of gas holdup in the gas/liquid interfacial area. In consequence, at the higher liquid velocity the value of k_La tends to decrease with a further increase in liquid velocity. Therefore, the value of gas-to-liquid mass transfer coefficient shows a local maximum with a variation of liquid velocity. It can be noted in this figure that the decrease trend of mass transfer coefficient at the relatively high liquid velocity can be improved by means of the swirling flow of liquid phase. In other words, the efficiency of gas-liquid contacting for mass transfer could be improved by increasing the turbulence as well as residence time of gas bubbles, at the relatively high liquid velocity range.

Effects of particle size on the gas-to-liquid mass transfer coefficient in the three-phase swirling fluidized bed can be seen in Fig. 5. In this figure, the mass transfer coefficient increases with increasing the particle size in all the cases studied. The reason why the k_La value increases with the increase of particle size can be that the larger particle may have the greater inertial force to break down the bubbles and generate the turbulence sufficient for the gas-to-liquid mass transfer at the bubble–liquid interfaces in the three phase beds. It has been reported bubble holdup increases and its size decreases with an increase in solid particle size in three phase fluidized beds due to bubble breaking potential of fluidized solid particles. The value of gas-liquid mass transfer coefficient shows the higher value in the three phase beds of liquid swirling than that in the beds without liquid swirling, which was reported in the literature, in the similar operating conditions.

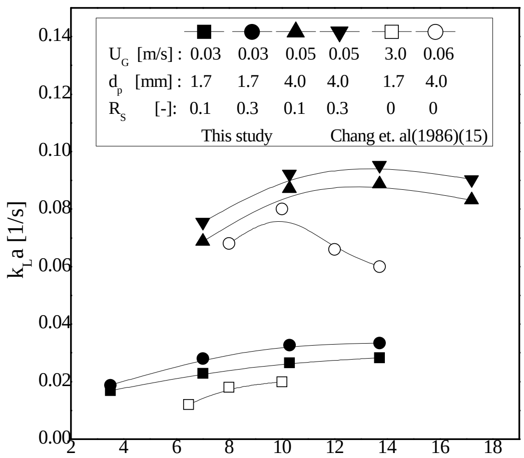

Fig. 4. Effects of liquid velocity on the k_La in the three-phase swirling fluidized beds

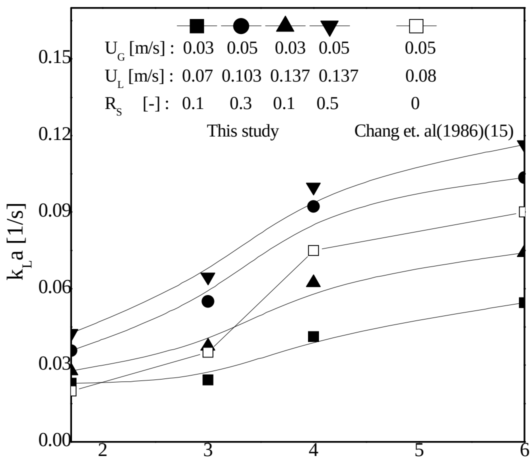

Fig. 5. Effects of particle size on the k_La in the three-phase swirling fluidized beds

Effects of swirling ratio of liquid phase on the mass transfer coefficient in the three-phase swirling fluidized bed can be seen in Fig. 6. In this figure, the mass transfer coefficient increases up to 70% with increasing the swirling ratio. This can be due to that the swirling liquid can generate the shear force which induces the upward flowing liquid to flow radially and tangentially in the column, because the swirling liquid has been injected by means of tangentially into the bubble column. In addition, the radial and tangential flow of liquid phase can generate turbulence by interrupting the upward flow of liquid-bubble mixture. The shear force and turbulence generated due to the swirling flow of liquid phase can break the bubble effectively in the column. Thus, the bubble size decreases while the gas holdup increases, with an increase in the ratio of swirling liquid. It seems that the effects of liquid swirling is noticeable when the gas velocity is relatively low, because, the turbulence for the gas-to-liquid mass transfer would not be sufficient in those conditions.

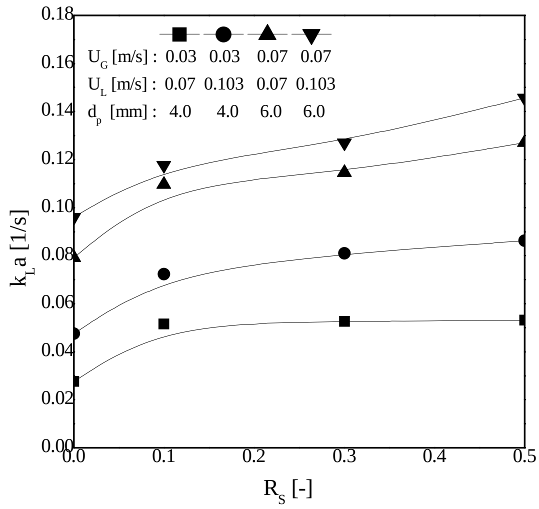

Fig. 6. Effects of swirling ratio on the $k_L a$ in the three-phase fluidized bed

The oxygen transfer coefficient value obtained from this study were correlated in terms of operation variables as Eq. (6) As can be seen in Fig. 7, Eq. (6) fit well with the experimentally obtained values with a correlation coefficient of 0.91. Since the contacting and flow behavior of gas, liquid and solid phases in the bed are highly stochastic and irregular, the values of gas-to-liquid mass transfer coefficient were correlated by means of dimensionless groups, as Eq. (7). As can be seen in Fig. 8, Eq. (7) also fit well with the experimentally obtained values with a correlation coefficient of 0.90.

$$k_L a = 114.4(U_G)^{0.812}(U_L)^{0.444}(d_p)^{0.729}(R_S+1)^{0.596} \tag{6}$$

$$St = 0.191\left(\frac{U_G}{U_G+U_L}\right)^{1.065}\left(\frac{d_p(U_G+U_L)\rho_L}{\mu_L}\right)^{0.581}(R_S+1)^{0.438} \tag{7}$$

For the application of mass transfer data in scaling-up of the reactor, the mass transfer coefficient could be expressed as Stanton number based on the column diameter and gas velocity(St_m) from the dimensional analysis. The values of St_m were correlated in terms of dimensionless groups such as Peclet(Pe), Schmidt(Sc), Weber(We), Froud(Fr) numbers, as Eq(8), which were derived from the dimensional analysis on this mass transfer system. As can be seen in Fig. 9, the values of St_m were well correlated as Eq.(8) with a correlation coefficient of 0.96.

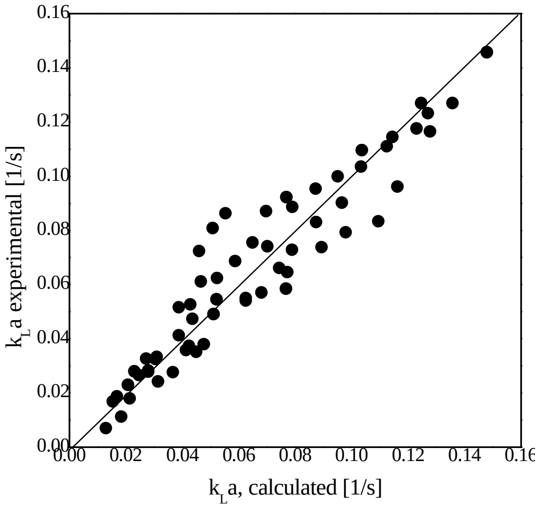

Fig. 7. Comparison between the experim- entally obtained and calculated values of k_La in a three-phase swirling fluidized beds.

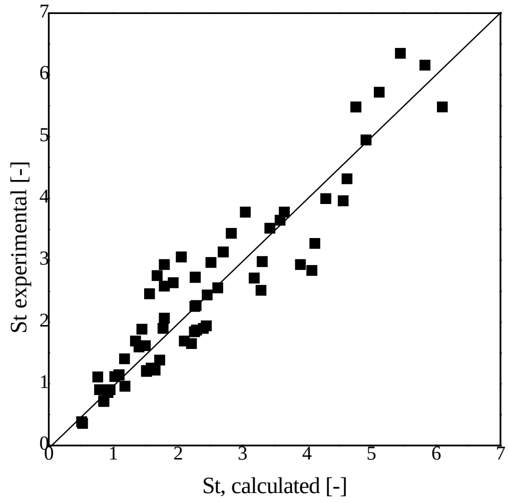

Fig. 8 Comparison between the experim-entally obtained and calculated values of Stanton number in a three-phase swirling fluidized beds.

$$St_m = 0.058\ Pe^{-0.67}\ Fr^{0.22}\ We^{0.02} \left(\frac{R_s}{R_s + 1} \right)^{0.99} \tag{8}$$

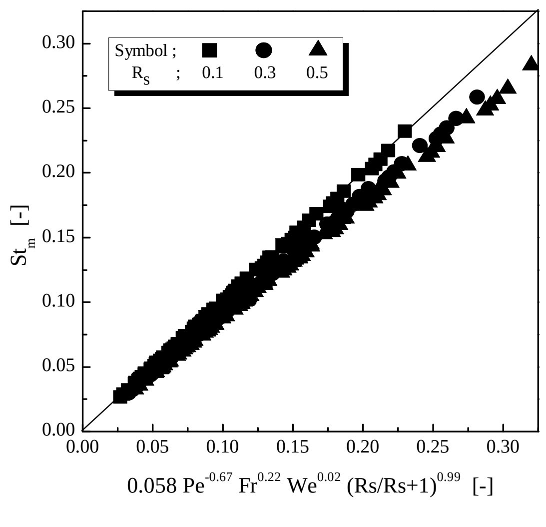

Fig. 9. Correlation of St_m in terms of dimensionless groups.

Conclusion

Effects of liquid swirling on the gas-to-liquid mass transfer in a three-phase (swirling) fluidized bed were well analyzed by considering the effects of operating variables. The gas-to-liquid mass transfer coefficients were successfully recovered `by adopting the axial dispersion model, since the concentration profiles of dissolved oxygen in the axial direction were well fitted to the model, as in three phase fluidized beds without liquid swirling. The values of the volumetric gas-to-liquid mass transfer coefficient were strongly dependent on the gas velocity in the beds. The gas-to-liquid mass transfer coefficient increased with increasing gas velocity, size of fluidized solid particles or liquid swirling ratio. However, the value of the mass transfer coefficient exhibits maximum value with a variation of liquid velocity in the bed. The gas-to-liquid mass transfer coefficients were well correlated with operation variables and dimensionless groups.

Nomenclature

C	concentration of dissolved oxygen [mol/L]
d_p	particle diameter [mm]
Fr	Froud number ($U_G^2/g\ D$) [-]
k_La	liquid side volumetric mass transfer coefficient [1/s]
L	bed height [m]
P	pressure [MPa]
Pe	Peclet number ($U_G \cdot D/D_L$) [-]
Re	Reynolds number ($d_P U_L \rho_L/\mu_L$) [-]
R_S	Liquid Swirling ratio [-]
Sc	Schmidt number ($\mu_L/((\rho_L-\rho_G) \cdot D_L)$) [-]
St	Stanton number ($k_La \cdot L/U_L$) [-]
St_m	Stanton number ($k_La \cdot D/U_G$) [-]
U_G	gas velocity [m/s]
U_L	liquid velocity [m/s]
We	Weber number ($D \cdot U_G^2 \cdot (\rho_L-\rho_G)\ /\sigma_L$) [-]
μ_L	liquid viscosity [Pa.s]
ρ_L	liquid density [kg/m^3]
ρ_S	solid density [kg/m^3]

<Subscript>

G	gas phase
L	liquid phase
S	solid phase

Acknowledgements

This work was supported by the Korea Energy Research Institute(A7-2802).

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