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SPECIFIC GAS-LIQUID INTERFACIAL AREA AND LIQUID-PHASE MASS TRANSFER COEFFICIENT IN A SLURRY BUBBLE COLUMN

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Key Words: Slurry Bubble Column, Bubble Column, Specific Interfacial Area, Bubble Size, Mass Transfer Coefficient

Introduction

Slurry bubble columns have been widely applied to such industrial processes as coal liquefaction, Fischer-Tropsch synthesis, biological waste water treatment and fermentation. In these processes, the rate of gas-liquid mass transfer, which is characterized by the volumetric mass transfer coefficient $k_L a$, may be controlling to the overall rate. To explain further the gas-liquid mass transfer phenomena in a slurry bubble column, it is very important to study the behavior of k_L , the liquid-phase mass transfer coefficient, and a , the specific gas-liquid interfacial area, independently. Little information, however, is available about a and k_L in a slurry bubble column.

From the experimental data on $k_L a$, gas holdup and volume-surface mean bubble diameter in our previous works^{3,15,16} the values of a and k_L in a slurry bubble column were separately evaluated and their behavior

was investigated and correlated in the present paper.

1. Specific Gas-Liquid Interfacial Area, a

The value of a can be evaluated from the mean (or cross-sectionally averaged) gas holdup, $\bar{\epsilon}_g$, and the volume-surface mean bubble diameter, d_{vs} , as follows.

$$a = 6\bar{\epsilon}_g/d_{vs} \quad (1)$$

Our previous study¹⁶ showed that the data on $\bar{\epsilon}_g$ were expressed by the correlation of Koide *et al.*⁷ for heterogeneous flow in the range where the mean solid holdup, $\bar{\epsilon}_s$, was less than 0.4. It was also shown that the data on d_{vs} was expressed well by the following equation.³

$$d_{vs} = 0.59(V_D/\bar{\epsilon}_g)^2/g \quad (2)$$

where g is the gravitational acceleration and V_D is the drift flux of gas defined as follows.¹⁴

$$V_D = U_g(1 - \bar{\epsilon}_g) - U_l\bar{\epsilon}_g(1 - \bar{\epsilon}_g)/\bar{\epsilon}_l \quad (3)$$

Figures 1 and 2 show the variation of a evaluated

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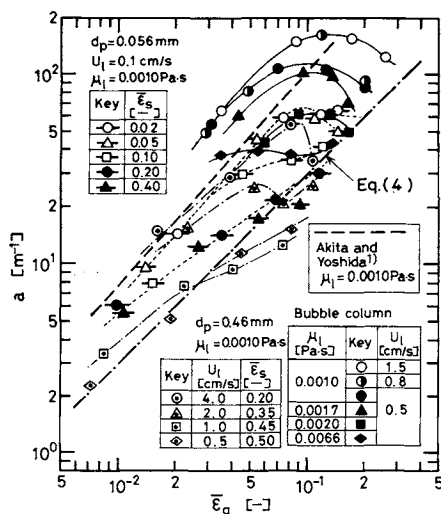


Fig. 1. Variation of a with $\bar{\epsilon}_g$ for a bubble column and a slurry bubble column.

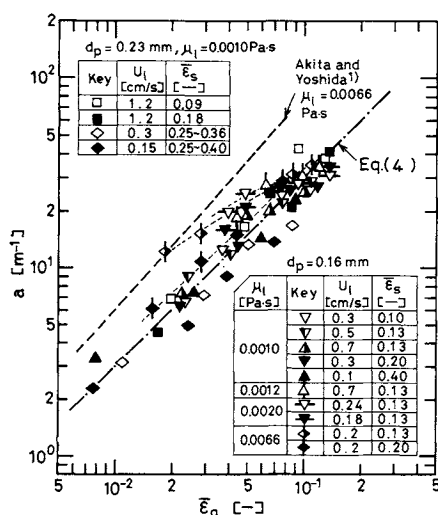


Fig. 2. Variation of a with $\bar{\epsilon}_g$ for a slurry bubble column.

from the original data on $\bar{\epsilon}_g$ and d_{vs} in a slurry bubble column and a bubble column. The values of a for bubble column calculated by the equation of Akita and Yoshida¹⁾ are also shown for comparison. It is seen that for the bubble column the value of a is twice that estimated from Akita's equation in the range of small $\bar{\epsilon}_g$, takes on a maximum and then decreases with increasing $\bar{\epsilon}_g$, while it decreases with increasing liquid viscosity, μ_l . The reduction of a with $\bar{\epsilon}_g$ is due to the change of flow pattern from homogeneous flow to heterogeneous flow, as noted in the previous paper.^{3,16)} For the slurry bubble column, the value of a increases with increasing $\bar{\epsilon}_g$ and decreases with increasing $\bar{\epsilon}_s$. The flow pattern also changes from homogeneous flow to heterogeneous flow, even at small $\bar{\epsilon}_g$ and $\bar{\epsilon}_s$ values. For the slurry bubble column with high solid content, the value of a may be expressed as

$$a = 300\bar{\epsilon}_g \quad (4)$$

where the unit of a is m^{-1} . The decrease of a with $\bar{\epsilon}_s$, however, is not significant at high values of μ_l . It is also noted that for the slurry bubble column with low solid content and for the bubble column the values of a also tend to approach the line expressed by Eq. (4) at high $\bar{\epsilon}_g$ values.

As shown in Fig. 3, the values of a obtained experimentally for the heterogeneous flow region agree fairly well with those estimated by using the estimated values of $\bar{\epsilon}_g$ (from Koide's equation⁷⁾) and d_{vs} (from Eq. (2)). The deviation between the values of $a_{\text{exp.}}$ and those of $a_{\text{est.}}$, however, is somewhat larger for the bubble column than for the slurry bubble column.

Figure 4 shows the variation of a with $\bar{\epsilon}_g$ for various bubble columns.^{1,2,4-6,8,11-13)} It is seen that the values of a for the columns containing CMC (carboxy methyl cellulose) solution or slurry having very fine particles decrease to the value of a in the slurry bubble column with high solid content (expressed by Eq. (4)) with increasing μ_l . For columns having a single- or multi-nozzle type gas distributor, the value of $\bar{\epsilon}_g$ can be estimated by Koide's equation within a deviation of about $\pm 30\%$. The values of a for the electrolyte system containing fine particles^{2,6,11)} are estimated fairly well from Eqs. (1)–(3) by using their original $\bar{\epsilon}_g$ data in the range of $U_g \leq 6 \text{ cm/s}$. However, the deviation of the estimated value from the observed one is large for the column containing CMC.^{4,5,12)} The values of a for non-electrolyte system^{8,13)} agree well with those estimated by the above method.

2. Liquid-Phase Mass Transfer Coefficient, k_l

Figure 5 shows the values of k_l evaluated from dividing $k_l a$ in the previous paper¹⁵⁾ by a . The relationships between k_l and d_{vs} for bubble columns in the literature^{1,9)} are also shown in the figure. It is seen that the value of k_l in the present work is proportional to the value of d_{vs} and decreases with increasing μ_l , while it is independent of d_p and $\bar{\epsilon}_s$. The degree of dependence of k_l on d_{vs} in the present work is larger than those observed by other workers.^{1,9)} However, the values of k_l for the bubble column in the present work agree well with those observed by Akita and Yoshida¹⁾ and Nakao *et al.*⁹⁾ The degree of dependence of k_l on μ_l for the slurry bubble column is also larger than that for bubble column observed by Akita and Yoshida.¹⁾

On the basis of dimensional analysis and the least-squares method, k_l values of the present work can be correlated by the following equation.

$$Sh = 4.5 \times 10^{-4} Sc^{0.5} Ga^{0.8} Bo^{-0.2} \quad (5)$$

where $Sh = k_l d_{vs} / D_l$, $Sc = \mu_l / (\rho_l D_l)$, $Ga = g d_{vs}^3 \rho_l^2 / \mu_l^2$

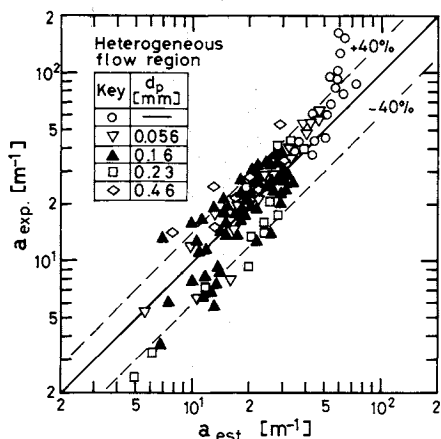


Fig. 3. Comparison of estimated and observed a values.

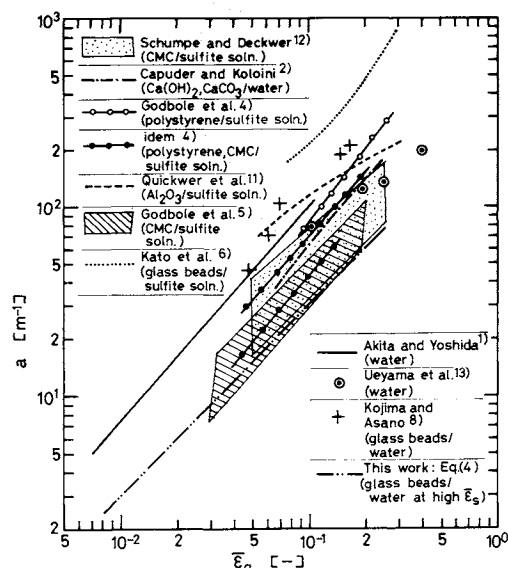


Fig. 4. Variation of a with $\bar{\epsilon}_g$ for various gas-liquid and gas-liquid-solid systems.

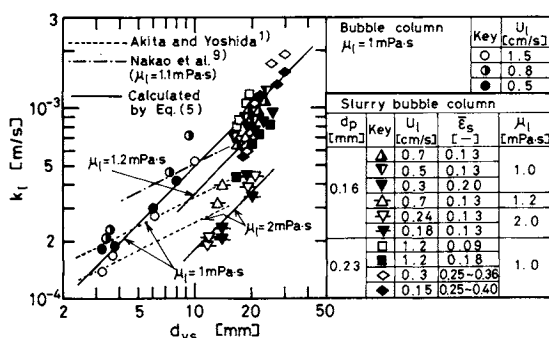


Fig. 5. Variation of k_l with d_{vs} for a bubble column and a slurry bubble column.

and $Bo = gd_{vs}^2 \rho_l / \gamma_l$ and the exponent of Sc is assumed to be 0.5 on the basis of the surface renewal theory. Eq. (5) is valid in the ranges of $480 \leq Sc \leq 1600$, $3.2 \times 10^5 \leq Ga \leq 2.6 \times 10^8$ and $1.4 \leq Bo \leq 120$. The coefficient of variation for Eq. (5) is 27%. The solid lines in Fig. 5 indicate the values of k_l estimated from Eq.

(5). It is seen that the calculated values agree well with the corresponding data.

By using Eqs. (1)–(3) and (5) in conjunction with individual phase holdups data, the value of $k_l a$ in the slurry bubble column can also be estimated from the relation, $k_l a = 6k_l \bar{\epsilon}_g / d_{vs}$. For instance, the data on $k_l a$ of the authors,¹⁵ Koide *et al.*⁷) and Nguyen-Tien¹⁰) are expressed within the deviation of about $\pm 40\%$ in the ranges of $U_g \leq 10 \text{ cm/s}$ and $\mu_l \leq 0.002 \text{ Pa}\cdot\text{s}$.

Nomenclature

a	= specific gas-liquid interfacial area based on unit column volume	[m ⁻¹]
Bo	= Bond number ($= gd_{vs}^2 \rho_l / \gamma_l$)	[—]
D_l	= molecular diffusivity of gaseous component in liquid phase	[m ² /s]
d_p	= particle diameter	[m]
d_{vs}	= volume-surface mean bubble diameter	[m]
Ga	= Galilei number ($= gd_{vs}^3 \rho_l^2 / \mu_l^2$)	[—]
g	= gravitational acceleration	[m/s ²]
k_l	= liquid-phase mass transfer coefficient	[m/s]
$k_l a$	= volumetric liquid-phase mass transfer coefficient based on unit column volume	[s ⁻¹]
Sc	= Schmidt number ($= \mu_l / (\rho_l D_l)$)	[—]
Sh	= Sherwood number ($= k_l d_{vs} / D_l$)	[—]
U_g	= superficial gas velocity	[m/s]
U_l	= superficial liquid velocity	[m/s]
V_D	= drift flux of gas	[m/s]
γ_l	= surface tension of liquid	[N/m]
$\bar{\epsilon}_g$	= mean (cross-sectionally averaged) gas holdup	[—]
$\bar{\epsilon}_l$	= mean liquid holdup	[—]
$\bar{\epsilon}_s$	= mean solid holdup	[—]
μ_l	= liquid viscosity	[Pa·s]
ρ_l	= liquid density	[kg/m ³]

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PARAMETER ESTIMATION IN A COOLING TOWER PROCESS

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Key Words: Parameter Estimation, Cooling Tower Process, Flowgraph Representation, Process Simulation, Computational Technique

In the previous paper,¹⁾ the solution to the packed tower process problem was proposed in the form of sequential cell flowgraph representation. There, the design problem of determining the tower height subject to the known input-output conditions data and the known parameter values (volumetric coefficients of heat and mass transfer) was treated. Alternatively, the parameter estimation subject to the known tower height and the known input-output conditions data is important for cases where a process is in operation.

In this report, a computational technique for parameter estimation in a packed tower process is proposed with the aim of eliminating the troublesome graphical routine, and an application to a cooling tower process is shown.

1. Proposed Method

The following algorithm for the parameter estimation in a packed tower process is proposed, where sequential cell flowgraph representation is used:

1) The height of tower Z is divided into N equal parts, and fundamental equations for the n -th part (called the n -th cell) from the tower bottom are arranged, using a model with gas phase, liquid phase, and interface.

2) Individual fundamental equation is converted to flowgraph unit and all units are combined to form an n -th cell flowgraph. A sequential cell flowgraph is constructed by arranging the n -th cell flowgraph simplified in boxed cell from $n=1$ to N .

3) Using the multi-variable searching technique in the sequential cell flowgraph, the value of the parameter is estimated as a limit when Δz goes to zero.

2. Application to a Cooling Tower Process Problem

A problem is proposed for estimating the values of the volumetric coefficients of heat and mass transfer in a cooling tower process, subject to the following data: temperature of air and water at tower top and bottom t_{GT} , t_{GB} , t_{LT} , t_{LB} , humidity of air at tower top and bottom H_T , H_B , dry air flow rate G , flow rate of water at tower top L_T , specific heat of water C_{PL} , and tower height Z .

As for steps 1 and 2, the model, the fundamental equations of the n -th cell and the n -th cell flowgraph are shown in the literature.¹⁾ Since the volumetric coefficient of heat transfer $h_{Gn}a$ in the Lewis equation depends on the cell number, its value should be determined as the average value between the tower top and bottom. A sequential cell flowgraph is shown in Fig. 1, where the following equations are used:

$$N = Z/\Delta Z \quad (1)$$

$$dv2 = dvH + dvL + dv t_L + dv t_G \quad (2)$$

Calculation will be commenced in the first cell with left-hand side variables t_{GB} , H_B , L_B , t_{LB} , Z and ΔZ , together with h_La and k_Ha . Cell-by-cell calculation must be continued to the N -th cell, where N is an integer determined by Eq. (1).

Since there is one-to-one correspondence between two parameters and $dv2$ in Eq. (2), we can determine h_La and k_Ha so as to minimize $dv2$ for a small value of ΔZ , using the multi-variable searching technique. This problem is denoted as follows:

$$h_La, k_Ha: dv2 = \min.$$

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