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GAS HOLDUPS AND FRICTION FACTORS OF GAS-LIQUID TWO-PHASE FLOW IN AN AIR-LIFT BUBBLE COLUMN

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***Key Words*:** Chemical Reactor, Bubble Column, Air-lift, Gas-Liquid Flow, Gas Holdup, Friction Factor

Gas holdups and friction factors were measured for gas-liquid parallel two-phase flow in an air-lift bubble column.

Velocities of gas and liquid, liquid viscosity, and surface tension were varied. The results were correlated with dimensionless equations.

As to the gas holdups (ε_G), Akita-Yoshida's equation[1] for batchwise processes was modified to the following form to suit continuous parallel gas-liquid flow operations:

$$\varepsilon_G(1-\varepsilon_G)^{-4}=0.20(gD^2\rho_L/\gamma)^{1/8}\ (gD^3/\nu_L^2)^{1/12}(U_{GL}/(gD)^{1/2})$$

As to the friction factors (f), the following correlation was proposed.

$$f=0.0468(U_L/(gD)^{1/2})^{-1.1}\varepsilon_G^{0.5}$$

These equations are useful for estimations of gas holdups, distribution of gas holdup along the column height, and liquid circulation velocities between a riser and a downcomer in an air-lift bubble column.

Introduction

Air-lift bubble columns are widely used as gas-liquid or gas-liquid-solid reactors in the chemical industry, water treatment apparatus in waste water engineering and in other services. The column consists of two cylindrical columns, i.e. a riser and a downcomer. They are connected in a loop with elbow joints or other devices at the bottom and a gas-liquid separator at the top. The riser is a column into which gas is sparged, while the downcomer is a pipe that returns the liquid from the top of the riser.

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In the riser, the gas-liquid mixture flows upward and the superficial liquid velocity in the riser rises as high as $1\ \mathrm{m \cdot s^{-1}}$. This high liquid velocity results in a high heat transfer rate through the column wall. If solid particles are suspended in the column, they are distributed uniformly throughout the column. Their uniform distribution is favorable for gas-liquid-solid three-phase reactions such as fermentation reactions.

The gas-liquid mixture at the top of the riser is separated into gas and liquid phases there. Only the liquid phase is brought into the top of the downcomer, flows downward through it, and is returned to the bottom of the riser. Thus a loop is completed for the liquid flow.

Numerous investigations[2,4-7,9,10,13] have been reported of gas-liquid two-phase flow and mass transfer in such columns, but no successful conclusions have yet been obtained. Therefore, we started our investigation in order to derive correlations for gas holdups and friction factors of gas-liquid flow.

For gas holdups, Akita-Yoshida's equation[1] was modified so as to be applicable to parallel gas-liquid flow as well as batchwise processes. The friction factors of gas-liquid two-phase flow in the riser were calculated with the data of pressure drops and were correlated with a dimensionless equation.

1. Experimentals

Figure 1 shows the experimental apparatus. The air-lift column, A, consisted of two columns, each 0.148 m in diameter. The length of the riser B was 8.020 m and that of the downcomer C was 7.795 m. The total length of the column was 8.600 m including elbow joints at the column bottom and a gas-liquid separator at the top. The volumes of the column components were 0.138 m^3 for the riser, 0.134 m^3 for the downcomer, 0.112 m^3 for the separator and 0.112 m^3 for the elbow joints.

At the bottom of the riser, a multiple-orifice nozzle sparger was installed. It was made of a perforated polyvinyl chloride plate mounted at one end of a nominal 15A pipe. The twenty holes were each 1.0 mm in diameter. It had been confirmed previously that the sparger gave the same gas holdup as was reported by Yoshida-Akita[14] with a single-orifice sparger.

To measure static pressure within the riser, two glass-pipe pressure taps were inserted into the riser at two heights through the column wall and their outside ends were connected to glass manometers F. The static pressure was measured at the heights of 0.695 m and 7.830 m. If the reading of the manometer connected to the pressure tap at the height z is h, static pressure P is calculated by the following equation.

$$P=\pi+\rho_L g(h-z) \tag{1}$$

The heights of installations and accessories were all measured from the column bottom. Liquid circulation velocity between riser and downcomer was measured with an orifice flowmeter installed midway in the downcomer at a height of 4.625 m. The diameter of the orifice was 11.0 cm. The liquid circulation velocity was regulated by the gate valve G, set at the height of 0.995 m in the downcomer. The average gas holdup ε_G in the riser was calculated by

$$\varepsilon_G=(V_F-V_L)/V_R \tag{2}$$

where V_L is the volume of clear liquid in the column, V_R the volume of the riser and V_F the fictitious volume of gas-liquid mixture in the column. The gas-liquid

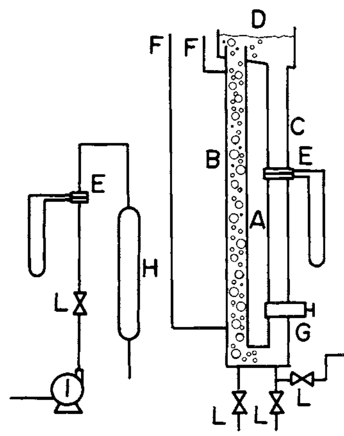

Fig. 1. Experimental apparatus

Table 1. Properties of liquids

Liquids	Temp. [K]	Density [$kg\cdot m^{-3}$]	Viscosity [$g\cdot m^{-1}\cdot s^{-1}$]	Surf. tension [$g\cdot s^{-2}$]
Water	283	999.7	1.308	74.20
5 $kg\cdot m^{-3}$ CMC	283	998.4	27.58	71.22
8 $kg\cdot m^{-3}$ CMC	283	1003	56.47	71.08
0.05% *n*-hepthanol	293	998.9	1.024	49.42
0.15 $kmol\cdot m^{-3}$ Na_2SO_4	283	1019	1.377	73.06

surface in the separator rippled violently and showed a complicated shape. Just above the riser top, the surface swelled steeply, to heights of up to 1 to 20 cm higher than the average horizontal surface of the remaining parts of the separator. The height of the latter surface was regarded as the height of the gas-liquid surface in the separator with which V_F was calculated. Preliminary experiments without liquid circulation showed that gas holdups measured by the method described above agreed well with those measured by manometer differences. This fact proved that the method above was adequate for the measurement of gas holdups in the riser because, without liquid circulation, gas holdups measured by the two methods must be equal to each other.

Operation of the column was carried out batchwise with respect to the liquid phase and continuously with respect to the gas phase.

Properties of liquids are listed in **Table 1**. Density was measured with an Ostwald pycnometer. Liquid viscosity, measured with a Canon-Fenske viscometer, varied from 1.02 to 56.5 $g\cdot m^{-1}\cdot s^{-1}$. Surface tension, measured with a stalagmometer, varied from 49.4 to 74.2 $g\cdot s^{-2}$.

2. Results and Discussion

2.1 Results of gas holdup measurement

The effect of liquid circulation velocities on gas

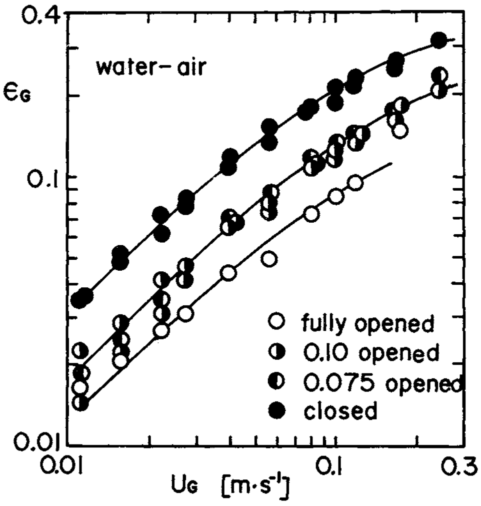

Fig. 2. Gas holdup for air–water system

holdups was investigated at various opening ratios of the gate valve.

Figure 2 shows results of gas holdup measurement for the air–water system. Parameters in the figure are four opening ratios of the gate valve: zero, 0.075, 0.1 and 1. The opening ratio of the valve was defined as the ratio of the cross-sectional area of the flow channel at the gate to the cross-sectional area of the riser. The figure shows that gas holdup decreases with increasing opening ratio. This phenomenon could be explained as the result of the reduction of relative gas velocity to liquid by the increase in opening ratio.

2.2 Modification of Akita-Yoshida's equation of gas holdup

Almost all the measurements of gas holdups were made at liquid velocities less than $0.1\ \mathrm{m\cdot s^{-1}}$, so that the effect of liquid velocities on gas holdups was neglected in the correlation. However, in the air-lift bubble column the effect is considerable, since liquid circulation velocity reaches as high as $1\ \mathrm{m\cdot s^{-1}}$.

To apply Akita-Yoshida's equation[1] of gas hold-up, which was originally derived from batchwise operation, we will here explain how to modify the equation with the drift-flux model.

Drift-flux j_{GL} was defined as[11]

$$j_{GL} \equiv U_G - \varepsilon_G(U_G + U_L) \tag{3}$$

Eq. (3) was rearranged to give

$$j_{GL} = (1-\varepsilon_G)U_{GL} \tag{4}$$

where U_{GL} is a superficial gas velocity relative to liquid phase and was defined as

$$U_{GL} \equiv U_G - U_L\varepsilon_G/(1-\varepsilon_G) \tag{5}$$

For the drift-flux j_{GL}, Wallis proposed the empirical equation[12] given as

$$j_{GL} = v_\infty \varepsilon_G(1-\varepsilon_G)^n \tag{6}$$

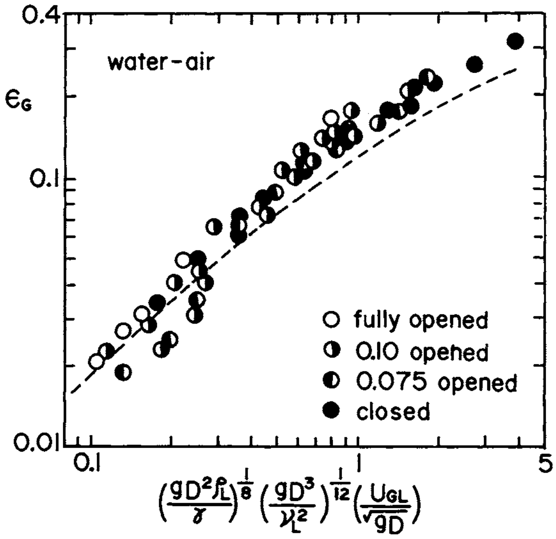

Fig. 3. General correlation of gas holdup

Combination of Eqs. (4) and (6) gives

$$\varepsilon_G(1-\varepsilon_G)^{n-1} = U_{GL}/v_\infty \tag{7}$$

If the operation is carried out without liquid flow, U_{GL} in Eq. (7) is replaced by U_G. The resultant equation was compared with Akita-Yoshida's equation[1] to give

$$n = -3 \tag{8}$$

$$v_\infty = 5(gD)^{1/2}(gD^2\rho_L/\gamma)^{-1/8}(gD^3/\nu_L^2)^{-1/12} \tag{9}$$

Substitution of Eqs. (8) and (9) into (7) gives

$$\varepsilon_G(1-\varepsilon_G)^{-4} = 0.2(gD^2\rho_L/\gamma)^{1/8} \times (gD^3/\nu_L^2)^{1/12}(U_{GL}/(gD)^{1/2}) \tag{10}$$

This is a modified form of Akita-Yoshida's equation[1] of gas holdup.

2.3 Gas holdup correlation by Eq. 10

In **Fig. 3**, the gas holdup of the air–water system is plotted against the product $(gD^2\rho_L/\gamma)^{1/8}\cdot(gD^3/\nu_L^2)^{1/12} \times (U_{GL}/(gD)^{1/2})$. The data in the figure are the same as those in Fig. 2.

In Fig. 2, the gas holdup at a fully opened valve was 60% less than that at a completely closed valve, if compared at the same U_G. However, Fig. 3 shows that the gas holdup is independent of the opening ratio of the valve if plotted against the dimensionless product appearing in the abscissa. The dotted line in the figure shows Eq. (10).

It appears that at high gas velocity, Eq. (10) predicts gas holdup 20% smaller than the observed value. The deviation could be ascribed to errors in measurement.

Otake *et al.*[9] measured gas holdup in parallel and countercurrent gas-liquid flow and correlated their results with an exponential function within $\pm 30\%$. Wachi *et al.*[10] also measured gas holdup in an air-lift column and compared their results with a semi-theoretical equation. With use of Eq. (10), we could correlate their data within $\pm 30\%$.

In conclusion, Eq. (10) could be applied to counter-current and parallel flow operations as well as batch-wise ones.

2.4 Relation between static pressure gradient $\Delta h/\Delta z$ and gas holdup

Static pressures were measured at the heights of $z_1=0.695$ m and $z_2=7.830$ m in the riser and the average static pressure gradient $\Delta h/\Delta z$ was calculated from

$$\Delta h/\Delta z \equiv (h_2-h_1)/(z_2-z_1)=(h_2-h_1)/7.135 \tag{11}$$

Here, the static pressure is expressed in terms of manometer heights h_1 and h_2.

Mori *et al.*[8] measured local pressure gradient dh/dz along a riser of an air-lift column and found that the local gradient remains invariable throughout the riser. Therefore, the average gradient $\Delta h/\Delta z$ can also be regarded as the local value dh/dz.

Since the gradient $\Delta h/\Delta z$ is considered to be the sum of gravity force exerted on the unit volume of gas-liquid mixture and frictional force of the two-phase flow, it must be a function of gas holdup and two-phase flow velocity.

Figure 4 shows the relation between the gradient $\Delta h/\Delta z$ and gas holdup ε_G. From the figure, linearity can be assumed between these quantities, i.e.

$$\Delta h/\Delta z=\lambda\varepsilon_G \tag{12}$$

where λ is a proportionality coefficient varying with the opening ratio of the gate valve.

The figure shows that the smaller the opening ratio, the larger becomes the constant λ, and for the completely closed valve λ is unity.

2.5 Calculation of the friction factors f for gas-liquid two-phase flow

Consider an infinitesimally short cylinder of gas-liquid mixture in the riser whose height is dz and diameter is that of the riser D.

In **Fig. 5**, all forces exerted on the cylinder in the riser are described schematically: two pressure forces acting on the lower and the upper surfaces $P(\pi/4)D^2$ and $-(P+dP)(\pi/4)D^2$; gravitational force $-g\rho_L\times(1-\varepsilon_G)(\pi/4)D^2dz$; and frictional force acting on the cylinder wall $-\tau_W\pi Ddz$.

Since the principle of force balance requires that the sum of these forces must be zero, the calculation gives

$$dP/dz+g\rho_L(1-\varepsilon_G)+(4/D)\tau_W=0\,. \tag{13}$$

By substituting P in Eq. (1) into Eq. (13) and rearranging, finally we obtain

$$\tau_W=(\rho_L gD/4)(\varepsilon_G-\Delta h/\Delta z) \tag{14}$$

Here, we define the friction factor f as

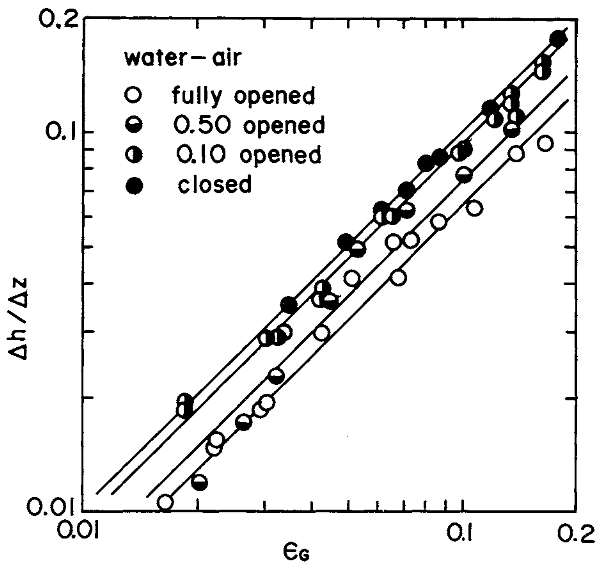

Fig. 4. Pressure gradient vs. gas holdup

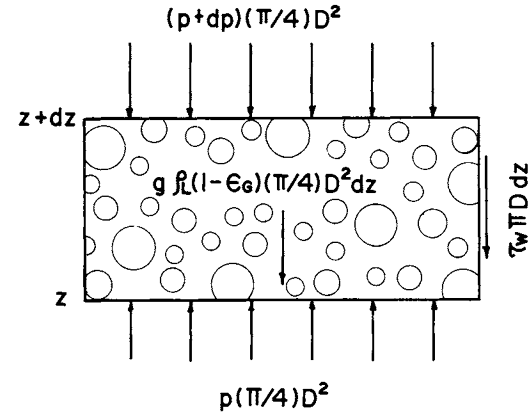

Fig. 5. Forces acting on a cylinder of gas-liquid mixture

$$\tau_W=(1/2)fG_M^2/\rho_M \tag{15}$$

where G_M is mass flow rate of the gas-liquid two-phase mixture and ρ_M is its density.

These quantities are defined and approximated as

$$G_M\equiv\rho_G U_G+\rho_L U_L \fallingdotseq \rho_L U_L \tag{16}$$

$$\rho_M\equiv\varepsilon_G\rho_G+(1-\varepsilon_G)\rho_L \fallingdotseq (1-\varepsilon_G)\rho_L \tag{17}$$

The contributions of gas phase to G_M and ρ_M are sufficiently small to be neglected.

Combining Eqs. (14) with (15) and substituting Eqs. (16) and (17) into the resultant equation, we get

$$f=(1/2)(U_L/(gD)^{1/2})^{-2}(\varepsilon_G-\Delta h/\Delta z)(1-\varepsilon_G)\,. \tag{18}$$

This equation was used to calculate the friction factor f.

2.6 Comparison of the friction factors of two-phase fluid flow with those of single-phase fluid flow

For a single-phase fluid flow, it is customary to plot the friction factor f against the Reynolds number Re.

The Reynolds number Re for the gas-liquid mixture

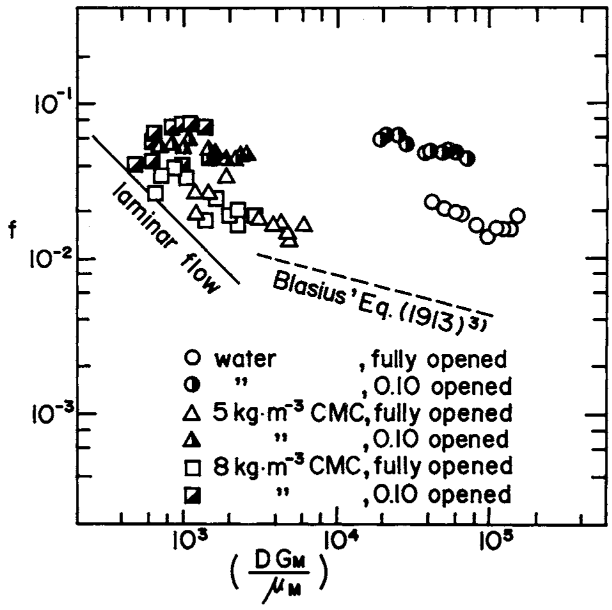

Fig. 6. Friction factor vs. Reynolds number

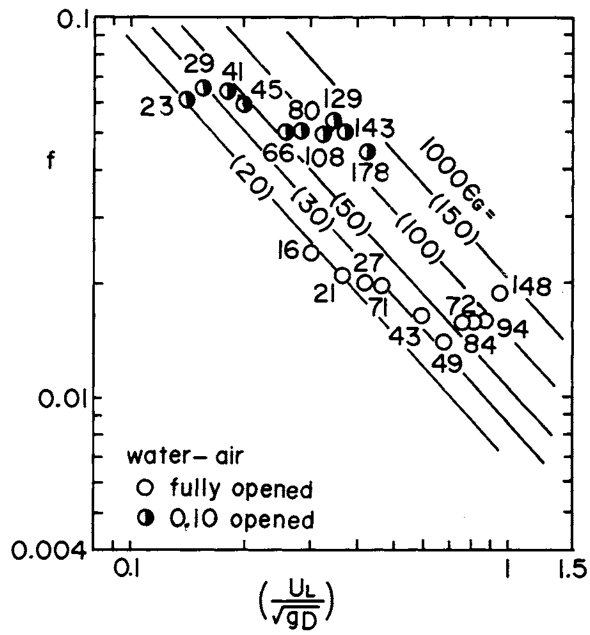

Fig. 7. Plot of f against $U_L/(gD)^{1/2}$

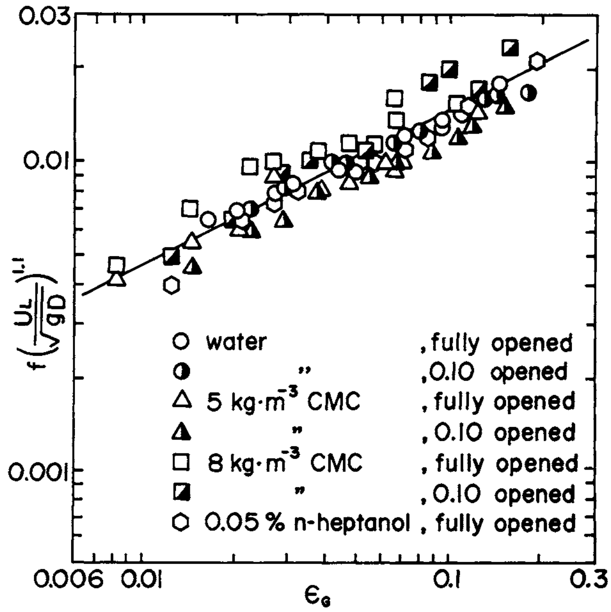

Fig. 8. Plot of $f\,(U_L/(gD)^{1/2})^{1.1}$ against ε_G

was defined as

$$Re \equiv DG_M/\mu_M \tag{19}$$

where μ_M is viscosity of the gas-liquid mixture and is defined and approximated as

$$\mu_M \equiv \mu_G\varepsilon_G + \mu_L(1-\varepsilon_G) \fallingdotseq \mu_L(1-\varepsilon_G)\,, \tag{20}$$

since $\mu_G\varepsilon_G$ is sufficiently small compared with $\mu_L(1-\varepsilon_G)$.

In **Fig. 6**, the friction factor f of the gas-liquid two-phase flow is plotted against the Reynolds number Re. The figure shows that f decreases as either Re or the opening ratio of the valve increases.

As already stated in **2.1**, the larger the opening ratio, the smaller becomes the gas holdup. Therefore, the friction factor f decreases as either Re increases or the gas holdup ε_G decreases.

In the figure, the friction factor of single-phase fluid flow is also shown. The friction factor for the two-phase fluids is seen to be larger than that for the single-phase fluids.

The fact that gas holdup increases friction factor can partly be ascribed to the fact that at a given U_L, shear stress τ_W increases with increasing gas holdup ε_G because of the increase in the linear liquid velocity $U_L/(1-\varepsilon_G)$.

2.7 Correlation of the friction factor f

Variables which affect the friction factor f are superficial liquid velocity U_L, liquid density ρ_L, liquid viscosity μ_L, surface tension γ and gas holdup ε_G. Dimensional analysis gives the following final relationship:

$$f = k(U_L/(gD)^{1/2})^p(gD^2\rho_L/\gamma)^q(gD^3/\nu_L^2)^r(\varepsilon_G)^s \tag{21}$$

where $U_L/(gD)^{1/2}$ is the Froude number, $gD^2\rho_L/\gamma$ the Bond number and gD^3/ν_L^2 the Galilei number. The proportional constant k and the indices p, q, r and s must be determined experimentally.

In **Fig. 7**, the friction factors for the water–air system are plotted against $U_L/(gD)^{1/2}$. The numeral attached to each datum point shows gas holdup multiplied by 1000. The lines in the figure which were drawn through the data are those of constant gas holdup. The slope of the lines or the exponent p in Eq. (21) is found to be -1.1. By substitution of -1.1 for p in Eq. (21) and rearrangement, the product $f(U_L/(gD)^{1/2})^{1.1}$ becomes a function of ε_G and liquid properties. In **Fig. 8**, the product $f(U_L/(gD)^{1/2})^{1.1}$ are plotted against ε_G for several gas-liquid systems and

Table 2. Values of $f(U_L/(gD)^{1/2})^{1.1}/\varepsilon_G^{-0.5}$

Liquids	$f(U_L/(gD)^{1/2})^{1.1}\varepsilon_G^{-0.5}$
Water	0.0457
$5\,kg\cdot m^{-3}$ CMC	0.0408
$8\,kg\cdot m^{-3}$ CMC	0.0545
0.05% *n*-heptanol	0.0438
$0.15\,kmol\cdot m^{-3}$ Na_2SO_4	0.0491
	AV. 0.0468

were found to be proportional to the square root of ε_G. Therefore, the index s in Eq. (21) was determined as 0.5. Thus the product $f(U_L/(gD)^{1/2})^{1.1}\varepsilon_G^{-0.5}$ for each liquid is listed in **Table 2**.

As seen from the table, the product varies from 0.0408 to 0.0545. The 30% scattering of the product can be ascribed to errors in measurement of ε_G and $\Delta h/\Delta z$. We can conclude that the value remains independent of the liquid properties. This means that both the indices q and r in Eq. (21) are zero.

By substitution of the values of the indices p, q, r and s into Eq. (21), the product $f(U_L/(gD)^{1/2})^{1.1}\varepsilon_G^{-0.5}$ became the proportional constant k. The most probable value of k was obtained by averaging the figures listed in Table 2 as

$$k=0.0468 \tag{22}$$

Thus, we obtained the following final expression.

$$f=0.0468(U_L/(gD)^{1/2})^{-1.1}\varepsilon_G^{0.5} \tag{23}$$

The range of variables in Eq. (23) is considered as

$$\varepsilon_G>0 \tag{24}$$

$$0<U_L/(gD)^{1/2}<1 \tag{25}$$

It is evident from Eq. (23) that f equals zero if ε_G is zero. This conclusion contradicts the behavior of the friction factor of single-phase fluid flow, where f has a finite value corresponding to a given Re. It also follows from Eq. (23) that f diverges when U_L becomes zero. This means that the equation cannot be defined at $U_L=0$.

Using Eq. (23), we can verify the statement mentioned in section 2.4 that dh/dz remains constant throughout the riser. The equation shows that the shear stress τ_W or $fU_L^2/2$ remains invariable throughout the riser insofar as gas holdup and superficial liquid velocity remain unchanged. Therefore, the constancy of dh/dz throughout the riser height is verified from Eq. (14).

The solid line drawn in Fig. 8 is an expression of Eq. (23). The figure shows that the scattering of our data from the line comes to ±20%.

Substitution of Eq. (23) into Eq. (18) gives

$$\Delta h/\Delta z=\varepsilon_G-0.0936(U_L/(gD)^{1/2})^{0.9}(\varepsilon_G^{0.5}/(1-\varepsilon_G)) \tag{26}$$

With Eq. (26), gas holdup ε_G was calculated from the data of $U_L/(gD)^{1/2}$ and $\Delta h/\Delta z$ by trial-and-error procedures. The calculated gas holdups differed from the measured ones by only ±10%.

Conclusions

Gas holdups and friction factors for the gas-liquid flow in an air-lift bubble column were investigated. These quantities were correlated by dimensionless equations.

The gas holdup equation of Akita-Yoshida[1] was modified to suit the parallel gas-liquid flow in a riser. The modified equation Eq. (10) can estimate the gas holdup within 20% deviation.

Friction factors of the gas-liquid two-phase flow was correlated by Eq. (23). The equation is valid over the range prescribed by Eqs. (24) and (25). If this equation is used for estimation of gas holdups, it can estimate the gas holdups within ±10% deviation.

These equations can be used to estimate the gas holdups, its distribution along a riser and the circulating velocities in an air-lift column.

Acknowledgment

Financial assistance by the Asahi Glass Foundation for Industrial Technology is gratefully acknowledged.

Nomenclature

D	= column diameter	[m]
f	= friction factor defined by Eq. (15)	[—]
G_M	= mass flow rate of gas-liquid mixture	$[kg\cdot m^{-2}\cdot s^{-1}]$
g	= gravitational constant, 9.800	$[m\cdot s^{-2}]$
h, h_i	= manometer heights	[m]
j_{GL}	= drift-flux defined by Eq. (3)	$[m\cdot s^{-1}]$
k	= proportional constant in Eq. (21)	[—]
n	= exponent in Eq. (6)	[—]
P	= static pressure	[Pa]
p, q	= exponents in Eq. (21)	[—]
Re	= Reynolds number defined by Eq. (19)	[—]
r, s	= exponents in Eq. (21)	[—]
U_G	= superficial gas velocity	$[m\cdot s^{-1}]$
U_L	= superficial liquid velocity	$[m\cdot s^{-1}]$
U_{GL}	= superficial gas velocity relative to liquid	$[m\cdot s^{-1}]$
V_F	= fictitious volume of gas-liquid mixture in a column	$[m^3]$
V_L	= volume of liquid in a column	$[m^3]$
V_R	= volume of a riser	$[m^3]$
v_∞	= proportional constant in Eq. (6)	[—]
z, z_i	= heights	[m]
γ	= surface tension	$[kg\cdot s^{-2}]$
ε_G	= gas holdup	[—]
λ	= proportionality coefficient in Eq. (12)	[—]
μ_G, μ_L	= viscosities of gas and liquid respectively	$[kg\cdot m^{-1}\cdot s^{-1}]$
μ_M	= viscosity of gas-liquid mixture defined by Eq. (20)	$[kg\cdot m^{-1}\cdot s^{-1}]$
ν_L	= kinematic viscosity of liquid	$[m^2\cdot s^{-1}]$
π	= atmospheric pressure	[Pa]

ρ_G, ρ_L = densities of gas and liquid respectively [$kg \cdot m^{-3}$]
τ_W = shear stress exerted by a column wall [$kg \cdot m^{-1} \cdot s^{-1}$]

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A RULE-BASED SIMULATION SYSTEM FOR DISCRETE EVENT SYSTEMS

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The increasing variety of variants in products and/or raw materials in a multiproduct batch process causes frequent alterations of the system configuration and the operational policies. Therefore, the simulation system for batch processes must be flexible enough to deal easily with the alterations. This paper presents a rule-based simulation system for discrete event systems, e.g. batch processes, which has the capability to examining the validity of a simulation model. The model we employ is composed of system state and events. The basic unit for representing the system state is the frame and the behavioral description of the event is provided by the if-then rule. To prevent the model from being incomplete, the simulation system clarifies the causal relationship between rules and frames and detects the rules which are never executed or which have possibilities of conflict. The relationship is also utilized for executing efficient simulation. The model employed deals easily with rapid alterations of the simulated system because it takes advantage of the modularity of a rule.

Introduction

The increasing variety of variants in products and/or raw materials in a multiproduct batch process causes frequent alterations of the system configuration and operational policies. Therefore, the simulation model describing the behavior of the batch process must be flexible enough to easily accommodate itself to the alterations.

State transitions of batch processes are generally regarded as instantaneous changes at specific points in time. These systems are often modeled as discrete event systems. Many special-purpose languages such as GPSS, SIMSCRIPT and SIMULA have been designed to build simulation models of discrete event systems.[2,4,7] Although these languages are powerful modeling tools for certain classes of systems, they suffer from a number of drawbacks, such as lack of flexibility. For example, GPSS obliges the user to build the simulation model as a flowchart composed of some predetermined blocks to which he/she is restricted. In SIMSCRIPT and SIMULA, any changes to the model require extensive program modifications because the structure of the model is embedded in the program. Therefore, these languages are unsuitable for investigation of operational policies.

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