



Interfacial area transport of vertical upward bubbly two-phase flow in an annulus

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Received 30 August 2002

Abstract

In relation to the development of the interfacial area transport equation in a subcooled boiling flow, the one-dimensional interfacial area transport equation was evaluated by the data taken in the hydrodynamic separate effect tests without phase change, or an adiabatic air–water bubbly flow in a vertical annulus. The annulus channel consisted of an inner rod with a diameter of 19.1 mm and an outer round tube with an inner diameter of 38.1 mm, and the hydraulic equivalent diameter was 19.1 mm. Twenty data sets consisting of five void fractions, about 0.050, 0.10, 0.15, 0.20, and 0.25, and four superficial liquid velocities, 0.272, 0.516, 1.03, and 2.08 m/s were used for the evaluation of the one-dimensional interfacial area transport equation. The one-dimensional interfacial area transport equation agreed with the data with an average relative deviation of $\pm 8.96\%$. Sensitivity analysis was also performed to investigate the effect of the initial bubble size on the interfacial area transport. It was shown that the dominant mechanism of the interfacial area transport was strongly dependent on the initial bubble size.

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Keywords: Interfacial area transport; Two-fluid model; Interfacial area concentration; Double-sensor conductivity probe; Gas–liquid bubbly flow; Multiphase flow

1. Introduction

In relation to the modeling of the interfacial transfer terms in the two-fluid model, the concept of the interfacial area transport equation has recently been proposed to develop the constitutive relation on the interfacial area concentration [1]. The dynamic changes in the two-phase flow structure can be predicted mechanistically by introducing the interfacial area transport equation. Such a capability does not exist in the current state-of-the-art nuclear thermal–hydraulic system analysis codes like RELAP5, TRAC and CATHARE. Thus, a successful development of the interfacial area trans-

port equation can make a quantum improvement in the two-fluid model formulation and the prediction accuracy of the system codes.

The present status of the development of the interfacial area transport equation was extensively reviewed in the previous paper [2]. In the first stage of the development of the interfacial area transport equation, adiabatic flow was the focus, and the interfacial area transport equation for the adiabatic flow was developed successfully by modeling sink and source terms of the interfacial area concentration due to bubble coalescence and breakup [3–5]. In the next stage, subcooled boiling flow would be the focus, and a preliminary local measurement for interfacial area concentration was initiated for subcooled boiling water flow in an internally heated annulus [6]. To develop the interfacial area transport equation for boiling flows in the internally heated annulus, sink and source terms due to phase change should be modeled based on rigorous and extensive boiling flow

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Nomenclature

A_i	interfacial area	We_{cr}	critical Weber number
a_i	interfacial area concentration	\bar{x}	spatial position
C	adjustable valuable	z	axial coordinate
C_{RC}	adjustable valuable	<i>Greek symbols</i>	
C_{WE}	adjustable valuable	α	void fraction
C_{TI}	adjustable valuable	α_{max}	maximum void fraction
D	diameter of outer round tube	ρ_g	gas density
D_0	diameter of inner rod	σ	interfacial tension
D_{bc}	critical bubble size	ϕ_{EXP}	change rate of interfacial area concentration due to bubble expansion
D_e	volume equivalent diameter	ϕ_j	change rate of interfacial area concentration due to bubble coalescence or breakup
D_H	hydraulic equivalent diameter	ϕ_{PH}	change rate of interfacial area concentration due to phase change
D_{Sm}	Sauter mean diameter	ϕ_{RC}	change rate of interfacial area concentration due to bubble coalescence caused by bubble random collision
$d\bar{x}$	spatial range	ϕ_{TI}	change rate of interfacial area concentration due to bubble breakup caused by turbulent impact
dV	volume range	ϕ_W	change rate of interfacial area concentration due to wall nucleation
f	particle density distribution function	ϕ_{WE}	change rate of interfacial area concentration due to bubble coalescence caused by wake entrainment
j_g	superficial gas velocity	ψ	factor depending on bubble shape ($1/(36\pi)$ for spherical bubble)
$j_{g,N}$	superficial gas velocity reduced at normal condition (atmospheric pressure and 20 °C)	<i>Subscripts</i>	
j_l	superficial liquid velocity	calc.	calculated value
i_{fg}	latent heat	meas.	measured value
n	exponent	Probe	quantity measured by double-sensor conductivity probe
R	radius of outer round tube	Rotameter	quantity measured by rotameter
R_0	radius of inner rod	γ -densitometer	quantity measured by γ -densitometer
R_j	change rate of particle number density due to particle coalescence or breakup	<i>Mathematical symbols</i>	
r	radial coordinate	$\langle \rangle$	area-averaged quantity
S_j	change rate of particle density distribution due to particle coalescence or breakup	$\langle \langle \rangle \rangle$	void fraction weighted cross-sectional area-averaged quantity
S_{PH}	change rate of particle density distribution due to phase change	$\langle \langle \rangle \rangle_a$	interfacial area concentration weighted cross-sectional area-averaged quantity
T_l	liquid temperature		
T_{sat}	saturation temperature		
t	time		
u_t	turbulent velocity		
V	particle volume		
V_{max}	maximum particle volume		
V_{min}	minimum particle volume		
\vec{v}_p	particle velocity		
\vec{v}_g	gas velocity		
v_g	gas velocity		
v_{gz}	gas velocity in z-direction		
We	Weber number		

data to be taken in the annular channel, and sink and source terms due to bubble coalescence and breakup modeled previously [3] should be evaluated separately based on adiabatic data to be taken in the same channel.

From this point of view, this study aims at evaluating the one-dimensional interfacial area transport equation with axial development data of local flow parameters (void fraction, interfacial area concentration, and interfacial velocity) of vertical upward air–water bubbly

flows in an annulus. The annulus test loop is scaled to a prototypic BWR based on scaling criteria for geometric, hydrodynamic, and thermal similarities [6]. The data obtained from the double-sensor conductivity probe give near complete information on the time-averaged local hydrodynamic parameters of bubbly flow to evaluate the sink and source terms of the interfacial area concentration. In this study, the one-dimensional interfacial area transport equation is evaluated by the data set taken in

the annulus loop. The detailed discussion is also given for the mechanism of the axial development of local flow parameters.

2. Interfacial area transport equation

For the purpose of modeling interfacial area transport, Kocamustafaogullari and Ishii [1] obtained the interfacial area transport equation based on statistical mechanics. The fluid particle number density distribution changes with the fluid particle contraction and expansion, entering and leaving, coalescence and disintegration, evaporation and condensation, nucleation and collapse. Simply accounting for these effects in a control volume yields the fluid particle transport equation:

$$\frac{\partial f}{\partial t} + \nabla \cdot (f \vec{v}_p) + \frac{\partial}{\partial V} \left(f \frac{dV}{dt} \right) = \sum_j S_j + S_{PH}, \quad (1)$$

where $f(\vec{x}, V, t)$ is the particle density distribution function, which is assumed to be continuous and specifies the probable number density of fluid particles at a given time t , in the spatial range $d\vec{x}$ about a position \vec{x} , with particle volumes between V and $V + dV$. $\vec{v}_p(\vec{x}, V, t)$ denotes the particle velocity, which is a function of the position, \vec{x} , particle volume, V , and time, t . For small rigid bubbles, the particle velocity, \vec{v}_p , is identical to the gas velocity, \vec{v}_g . The interaction term, $\sum_j S_j$, represents the net rate of change in the particle density distribution due to the particle coalescence and breakup processes. The second term of the right-hand side, S_{PH} is the fluid particle source or sink rate due to the phase change. For example, for a one-component bubbly flow, S_{PH} represents the bulk liquid bubble nucleation rate due to homogeneous and heterogeneous nucleation, and the collapse rate due to condensation for the subcooled boiling flow. The wall nucleation rate which is not included in S_{PH} must be specified as a boundary condition. The third term of the left-hand side in Eq. (1) represents the rate of change in the particle density distribution due to the pressure change and/or phase changes appearing on existing interfaces.

The interfacial area concentration transport equation of fluid particles can be obtained by multiplying the particle number density transport equation by the average interfacial area, $A_i(V)$, which is independent of the spatial coordinate system. This yields the following equation:

$$\frac{\partial f A_i(V)}{\partial t} + \nabla \cdot (f \vec{v}_g A_i(V)) + A_i(V) \frac{\partial}{\partial V} \left(f \frac{dV}{dt} \right) = \sum_j S_j A_i(V) + S_{PH} A_i(V). \quad (2)$$

For practical purposes, the fluid particle interfacial area transport equation is too detailed. Hence, it would be

much more useful to average an interfacial area transport equation over particle size groups that are determined according to particle mobilities. As a general approach, two-group interfacial area transport equations have recently been proposed by treating the bubbles in two groups such as the spherical/distorted bubble group (group one) and the cap/slug bubble group (group two) [5]. If only one group of bubbles is considered, the interfacial area transport equation can easily be obtained by integrating Eq. (2) from V_{\min} to V_{\max} and applying the Leibnitz rule. Then, we have the three-dimensional interfacial area concentration transport equation:

$$\begin{aligned} \frac{\partial a_i}{\partial t} + \nabla \cdot (a_i \vec{v}_g) &= \frac{2}{3} \frac{a_i}{\alpha} \left(\frac{\partial \alpha}{\partial t} + \nabla \cdot \alpha \vec{v}_g \right) + \frac{1}{3\psi} \left(\frac{\alpha}{a_i} \right)^2 \sum_j R_j + \pi D_{bc}^2 \\ &\times \left(1 - \frac{2}{3} \left(\frac{D_{bc}}{D_{Sm}} \right) \right) R_{RH}, \end{aligned} \quad (3)$$

where R_j is the rate of change of particle number due to coalescence or breakup, ψ is the shape factor defined by

$$\psi = \frac{1}{36\pi} \left(\frac{D_{Sm}}{D_e} \right)^3, \quad \left(\psi = \frac{1}{36\pi} \text{ for spherical bubbles} \right), \quad (4)$$

where D_e is the volume equivalent diameter, and D_{bc} is the critical bubble size beyond which it is possible for bubbles to grow due to evaporation, or for clusters of molecules to serve as nuclei for bubbles, as

$$D_{bc} = \frac{4\sigma T_{sat}}{\rho_g i_{fg} (T_f - T_{sat})}, \quad (5)$$

where σ , T_{sat} , ρ_g , i_{fg} and T_f are the surface tension, the saturation temperature, the gas density, the latent heat, and the liquid temperature, respectively.

The simplest form of the one-group interfacial area transport equation is the one-dimensional formulation obtained by applying cross-sectional area-averaging over Eq. (3). That is

$$\begin{aligned} \frac{\partial \langle a_i \rangle}{\partial t} + \frac{\partial}{\partial z} (\langle a_i \rangle \langle v_{gz} \rangle) &= \langle \phi_{EXP} \rangle + \sum_j \langle \phi_j \rangle + \langle \phi_{PH} \rangle + \langle \phi_W \rangle, \end{aligned} \quad (6)$$

where $\langle \phi_{EXP} \rangle$, $\langle \phi_j \rangle$, and $\langle \phi_{PH} \rangle$ are the change rate of the interfacial area concentration due to bubble expansion, bubble coalescence or breakup, and phase change defined as Eqs. (7)–(9), respectively.

$$\langle \phi_{EXP} \rangle \equiv \left(\frac{2 \langle a_i \rangle}{3 \langle \alpha \rangle} \right) \left\{ \frac{\partial \langle \alpha \rangle}{\partial t} + \frac{d \langle \alpha \rangle \langle v_{gz} \rangle}{dz} \right\}, \quad (7)$$

$$\langle \phi_j \rangle \equiv \frac{1}{3\psi} \left(\frac{\langle \alpha \rangle}{\langle a_i \rangle} \right)^2 \langle R_j \rangle, \quad (8)$$

$$\langle \phi_{PH} \rangle \equiv \pi D_{bc}^2 \left\{ 1 - \frac{2}{3} \left(\frac{D_{bc}}{\langle D_{Sm} \rangle} \right) \right\}. \quad (9)$$

$\langle \phi_W \rangle$ is the wall nucleation source, which is the most important term for subcooled boiling flow.

For gas-dispersed flows with no phase change, three major mechanisms are responsible for bubble coalescence and breakup that result in the interfacial area transport [3]. They are (1) bubble coalescence due to random collision driven by liquid turbulence, $\langle \phi_{RC} \rangle$, (2) bubble coalescence due to wake-entrainment caused by the relative motion between the bubbles in the wake region and the leading bubble, $\langle \phi_{WE} \rangle$ and (3) bubble breakup upon the impact of turbulent eddies, $\langle \phi_{TI} \rangle$. The source and sink terms of the interfacial area in Eq. (6) can be expressed as follows [3]:

$$\langle \phi_{RC} \rangle = 12\psi C_{RC} \left[\frac{\langle u_t \rangle \langle a_i \rangle^2}{\langle \alpha \rangle^{1/3} (\langle \alpha \rangle_{max}^{1/3} - \langle \alpha \rangle^{1/3})} \right] \times \left[1 - \exp \left(- \frac{C \langle \alpha \rangle_{max}^{1/3} \langle \alpha \rangle^{1/3}}{\langle \alpha \rangle_{max}^{1/3} - \langle \alpha \rangle^{1/3}} \right) \right], \quad (10)$$

$$\langle \phi_{WE} \rangle = 12\psi C_{WE} (\langle u_r \rangle \langle a_i \rangle^2), \quad (11)$$

$$\langle \phi_{TI} \rangle = \frac{1}{18} C_{TI} \left[\frac{\langle u_t \rangle \langle a_i \rangle^2}{\langle \alpha \rangle} \right] \left[1 - \frac{We_{cr}}{We} \right]^{1/2} \exp \left(- \frac{We_{cr}}{We} \right), \quad (12)$$

$We > We_{cr}.$

where $C_{RC} (= 0.0041)$, $C (= 3.0)$, $C_{WE} (= 0.0020)$, $C_{TI} (= 0.035)$ are adjustable values determined based on extensive data set taken in various adiabatic bubbly flows. u_t , α_{max} , u_r , We , and We_{cr} are the turbulent velocity, the maximum allowable void fraction ($= 0.80$), the relative velocity between gas and liquid phases, Weber number and critical Weber number ($= 2.0$), respectively.

3. Experimental

An experimental facility is designed to measure the relevant two-phase parameters necessary for developing constitutive models for the two-fluid model in subcooled boiling. It is scaled to a prototypic BWR based on scaling criteria for geometric, hydrodynamic, and thermal similarities [6]. The experimental facility, instrumentation, and data acquisition system are briefly described in this section. The detailed explanation can be found in the previous paper [6].

The two-phase flow experiment was performed by using a flow loop constructed at Thermal-hydraulics and Reactor Safety Laboratory in Purdue University.

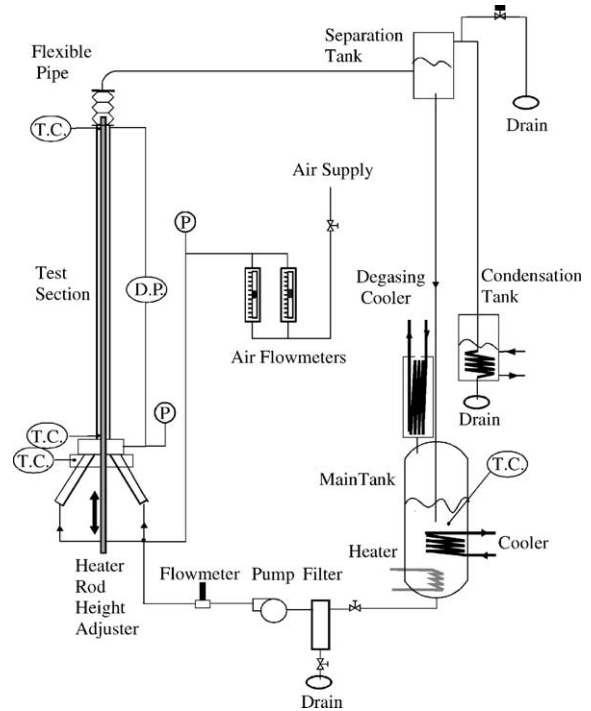


Fig. 1. Schematic diagram of experimental loop.

Fig. 1 shows the experimental facility layout. The water supply is held in the main tank. The tank is open to the atmosphere through a heat exchanger mounted to the top to prevent explosion or collapse and to degas from the water. There is a cartridge heater inside the tank to heat the water and maintain the inlet water temperature. A cooling line runs inside the tank to provide control of the inlet water temperature and post-experimental cooling of the tank. Water is pumped with a positive displacement, eccentric screw pump, capable of providing a constant head with minimum pressure oscillation. For the adiabatic air–water flow experiment, porous spargers with the pore size of 10 μm are used as air injectors. The water, which flows through a magnetic flowmeter, is divided into four separate flows and can then be mixed with air before it is injected into the test section to study adiabatic air–water bubbly flow.

The test section is an annular geometry that is formed by a clear polycarbonate tube on the outside and a cartridge heater on the inside. The test section is 38.1 mm inner diameter, D , and has a 3.18 mm wall thickness. The overall length of the heater is 2670 mm and has a 19.1 mm outer diameter, D_0 . The hydraulic equivalent diameter, D_H is 19.1 mm. The heated section of the heater rod is 1730 mm long. The maximum power of the heater is 20 kW and has a maximum surface heat flux of 0.193 MW/m². The heater rod has one thermocouple that is connected to the process controller to

provide feedback control. The heater rod can be traversed vertically to allow many axial locations to be studied with four instrument ports attached to the test section. At each port there is an electrical conductivity probe to measure axial development of local flow parameters. Pressure taps and thermocouples are placed at the inlet and exit of the test section. A differential pressure cell is connected between the inlet and outlet pressure taps. The two-phase mixture flows out of the test section to a separator tank and the gas phase is piped away and the water is returned to the main tank. Since the primary purpose of this study was to acquire the data in the hydrodynamic separate effect tests without phase change, the loop was operated with an adiabatic air–water flow in this experiment. The loop will be operated with a diabatic steam–water flow in a future study.

The flow rates of the air and water were measured with a rotameter and a magnetic flowmeter, respectively. The loop temperature was kept at a constant temperature (20 °C) within the deviation of ±0.2 °C by a heat exchanger installed in a water reservoir. The local flow measurements using the double-sensor conductivity probe [7] were performed at four axial locations of $z/D_H = 40.3, 61.7, 77.7,$ and $99.0,$ and 10 radial locations from $r/(R - R_0) = 0.05$ to $0.9.$ Here, R and R_0 are the inner radius of the outer tube and the outer radius of the heater, respectively. $r/(R - R_0) = 0$ and 1 indicate the walls of the heater and the outer tube, respectively. A γ -densitometer was installed at $z/D_H = 51.1$ in the loop to measure the area-averaged void fraction. The flow conditions in this experiment are tabulated in Table 1. The area-averaged superficial gas velocities in this experiment were roughly determined so as to provide the same area-averaged void fractions among different conditions of superficial liquid velocity, namely $\langle \alpha \rangle = 0.050, 0.10, 0.15, 0.20,$ and $0.25.$ The flow conditions covers typical bubbly flow conditions, since the primary purpose of this study is to acquire the data in the hydrodynamic separate effect tests without phase change to be the reference for future subcooled bubbly flow tests. It should be noted here that void fraction would increase along the axial direction on the order of 10% between $z/D_H = 40.3$ and 99.0 in the present ex-

perimental conditions due to the pressure reduction. This leads to a continuous developing flow along the flow direction.

Generally, a void distribution depends on an initial condition (bubble size, generation method and mixing condition), a flow condition (flow rates and physical properties), and a test section condition (geometry and wall surface) [8]. Although sophisticated experiments controlling the initial condition were performed [8], the initial condition was not controlled in this experiment, resulting in the change of the initial bubble size with the flow condition.

Fig. 2 shows the dependence of the Sauter mean diameter, $\langle D_{sm} \rangle,$ measured at the first measuring station of $z/D_H = 40.3$ on the void fraction, $\langle \alpha \rangle,$ or the superficial liquid velocity, $\langle j_l \rangle.$ The bubble size measured at the first measuring station increased with increasing the gas flow rate, but the effect of the gas flow rate on the bubble size came to be smaller for higher liquid flow rate. On the other hand, the bubble size increased with increasing the liquid flow rate for $\langle j_l \rangle < 0.5$ m/s and $\langle \alpha \rangle \leq 0.10,$ but the bubble size decreased with increasing the liquid flow rate for $\langle j_l \rangle \geq 1.0$ m/s. The increase in the gas flow rate or the increase in the liquid flow rate for $\langle j_l \rangle < 0.5$ m/s and $\langle \alpha \rangle \leq 0.10$ would enhance the bubble coalescence due to the collision of bubbles, resulting in the increase of the bubble size. On the other hand, the increase in the liquid flow rate for $\langle j_l \rangle \geq 1$ m/s would enhance the bubble breakup due to the liquid turbulence, resulting in the decrease of the bubble size. However, the effect of the liquid flow rate on the bubble size was not pronounced for $\langle j_l \rangle \leq 1$ m/s and $\langle \alpha \rangle > 0.10,$ where the bubble coalescence and breakup seem to be in the equilibrium state. In this experiment, bubbles with the diameters of about 3 and 2 mm were generated for $\langle j_l \rangle \leq 1$ m/s and $\langle j_l \rangle = 2$ m/s, respectively. The similar result was obtained for vertical air–water bubbly flow in a round tube with an inner diameter of 50.8 mm [9].

In order to verify the accuracy of local measurements, the area-averaged quantities obtained by integrating the local flow parameters over the flow channel were compared with those measured by other cross-calibration methods such as a γ -densitometer for void fraction, a photographic method for interfacial area

Table 1
Flow conditions in this experiment

Symbols	●	▲	■	▼	◆
Lines	—	-----	- · - · -	- · · · ·
$\langle j_l \rangle$ [m/s]	$\langle j_{g,N} \rangle$ [m/s]	$\langle j_{g,N} \rangle$ [m/s]	$\langle j_{g,N} \rangle$ [m/s]	$\langle j_{g,N} \rangle$ [m/s]	$\langle j_{g,N} \rangle$ [m/s]
0.272	0.0313	0.0506	0.0690	0.0888	0.105
0.516	0.0406	0.0687	0.103	0.135	0.176
1.03	0.0683	0.130	0.201	0.400	0.489
2.08	0.108	0.215	0.505	0.651	0.910

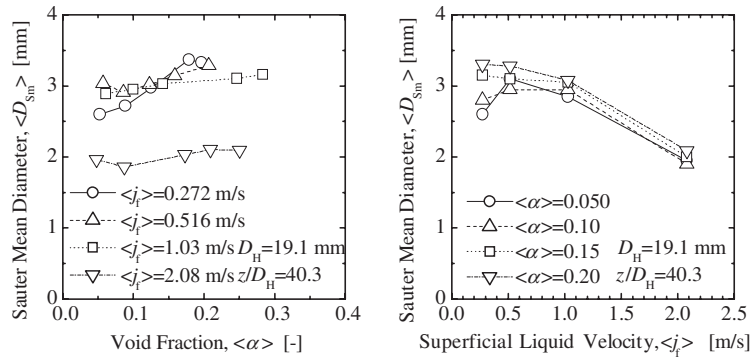


Fig. 2. Dependence of bubble size on void fraction and superficial liquid velocity.

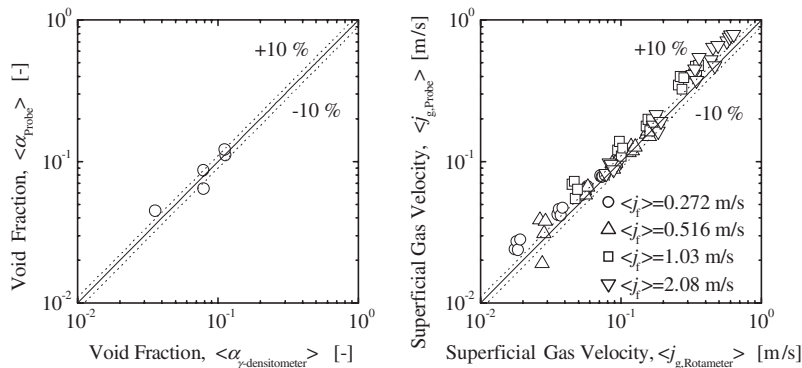


Fig. 3. Verification of the double sensor probe with other calibration methods.

concentration, and a rotameter for superficial gas velocity. As shown in Fig. 3, good agreements were obtained between the area-averaged void fraction and superficial gas velocity obtained from the local measurements and those measured by the cross-calibration methods with averaged relative deviations of $\pm 12.8\%$ and $\pm 14.9\%$, respectively. The separate experiment to evaluate the interfacial area concentration measured by the double sensor probe with the photographic method also showed a good agreement between the area-averaged interfacial area concentrations obtained from the double sensor probe method and those measured by the photographic method with an averaged relative deviation of $\pm 6.95\%$ [10].

4. Results and discussion

4.1. Axial development of local flow parameters

Some discussions on the flow characteristics of local flow parameters for gas and liquid phases in an annulus can be found in the previous papers [11,12]. Here, the

axial development of the local flow parameters will be briefly described as follows.

4.1.1. Void fraction

Fig. 4 shows the behavior of void fraction profiles measured at $z/D_H = 40.3$ (upper figures) and 99.0 (lower figures) in this experiment. The meanings of the symbols in Fig. 4 are found in Table 1. As can be seen from the figure, various phase distribution patterns similar to those in round tubes are observed in the present experiment, and void fraction profiles are found to be almost symmetrical with respect to the channel center, $r/(R - R_0) = 0.5$.

For $\langle j_l \rangle = 0.272$ m/s, broad core peak with plateau around the channel center and intermediate peak, which is characterized as broad peak in void fraction near the channel wall and plateau with medium void fraction around the channel center, are found for low (\bullet , \blacktriangle) and high (\blacksquare , \blacktriangledown , \blacklozenge) void fraction regions, respectively, at the first measuring station of $z/D_H = 40.3$. As the flow develops, the plateau observed for low void fraction region (\bullet , \blacktriangle) tends to be narrower. On the other hand, as the flow develops, two peaks observed for high void fraction

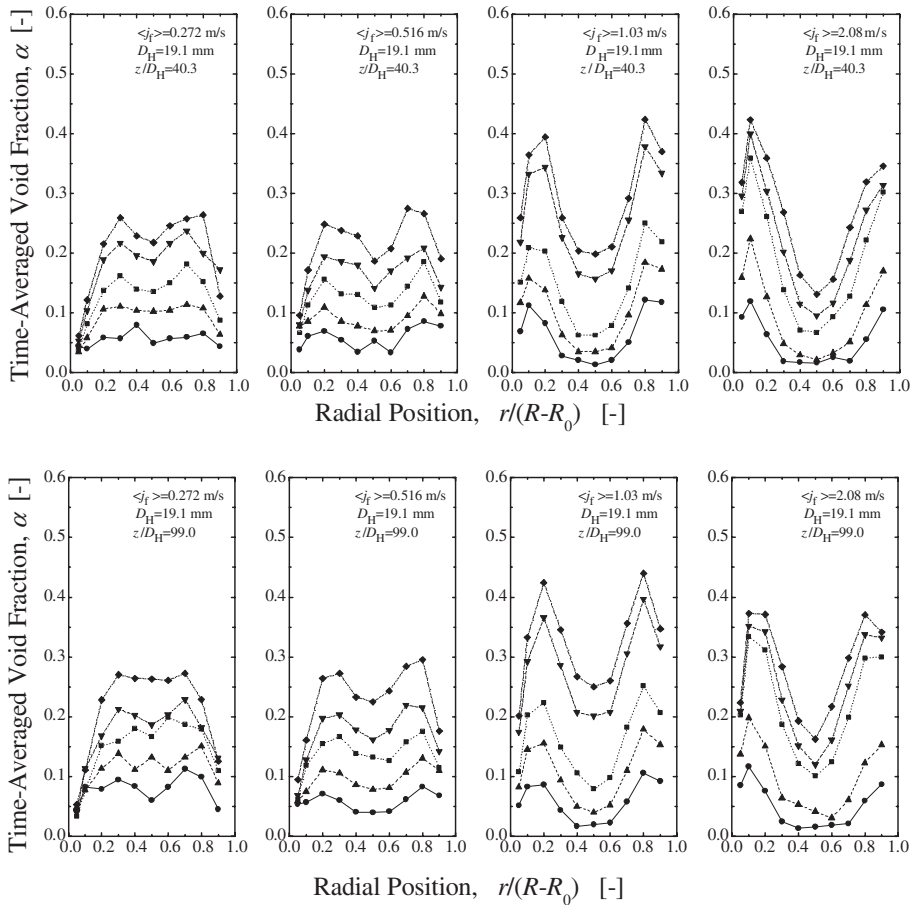


Fig. 4. Local void fraction profiles at $z/D_H = 40.3$ and 99.9 .

region (■, ▼, ◆) tend to move towards the channel center and to be merged into one core peak. For $\langle j_r \rangle = 0.516$ m/s, intermediate peak is observed at the first measuring station of $z/D_H = 40.3$. As the flow develops, the void fraction profiles are not changed for low void fraction region (●, ▲), but the trough of the void fraction profiles observed around the channel center comes to be shallower for high void fraction region (■, ▼, ◆). The similar tendency is observed for $\langle j_r \rangle = 1.03$ m/s. For $\langle j_r \rangle = 2.08$ m/s, wall peak is observed at the first measuring station of $z/D_H = 40.3$. As the flow develops, the void fraction profiles are not changed.

For $\langle j_r \rangle = 0.272, 0.516,$ and 1.03 m/s, the bubble diameter is about 3 mm, which is close to a critical bubble size of 3.6 mm pointed out by Zun [13], which gives the boundary between the wall and intermediate peaks. The bubble size is likely to determine the direction of the bubble migration. Thus, for tested low liquid flow conditions, bubbles tend to move towards the channel center gradually. For $\langle j_r \rangle = 2.08$ m/s, the bubble diameter is about 2 mm, and the bubble can stay near

the channel wall, resulting in insignificant axial change of the void fraction distribution.

4.1.2. Sauter mean diameter

Fig. 5 shows the behavior of Sauter mean diameter profiles, corresponding to that of void fraction profiles in Fig. 2. The meanings of the symbols in Fig. 5 are found in Table 1. The Sauter mean diameter profiles are uniform along the channel radius with some decrease in size near the wall, $r/(R - R_0) \leq 0.1$ and $0.9 \leq r/(R - R_0)$. Only a part of a bubble can pass the region close to the channel wall, resulting in apparent smaller Sauter mean diameter. The profiles are not changed significantly as the flow develops, although the bubble size increases up to 10–20% along the flow direction.

4.1.3. Interfacial area concentration

Fig. 6 shows the behavior of interfacial area concentration profiles, corresponding to that of void fraction profiles in Fig. 4. The meanings of the symbols in Fig. 6 are found in Table 1. As can be expected for

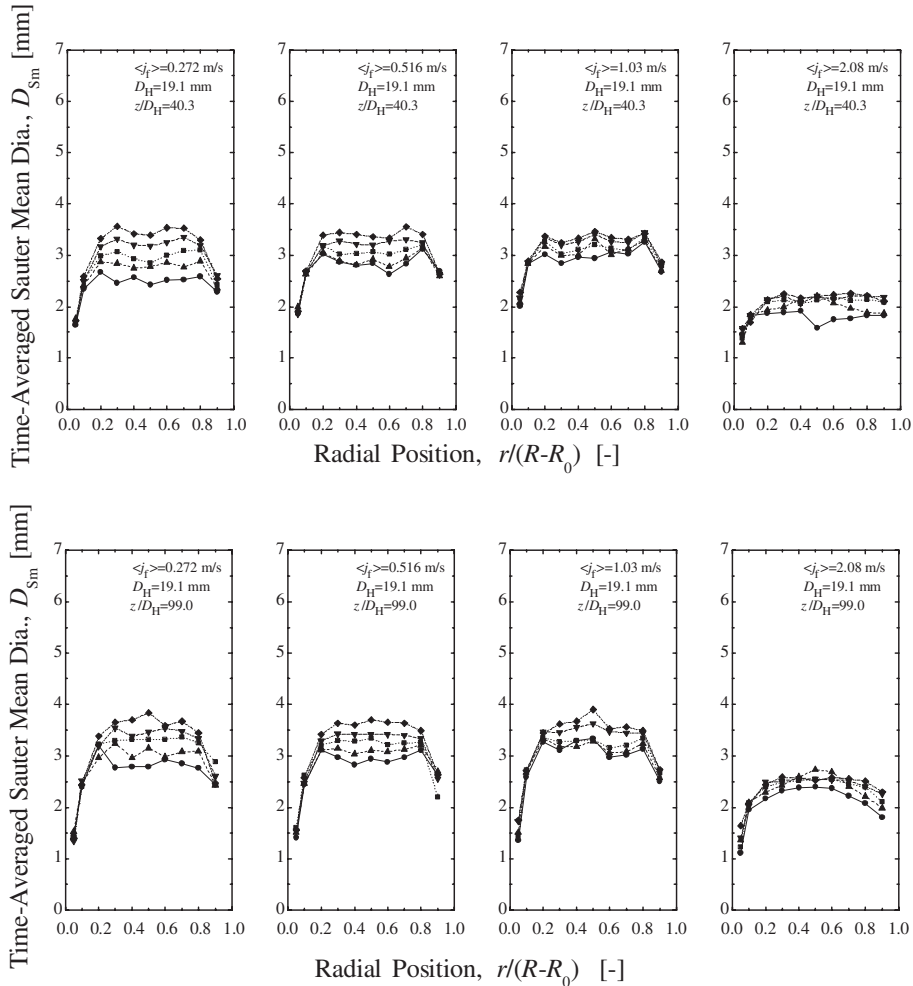


Fig. 5. Local Sauter mean diameter profiles at $z/D_H = 40.3$ and 99.9 .

bubbly flow, the interfacial area concentration profiles are similar to the void fraction profiles. The interfacial area concentration is proportional to the void fraction and inversely proportional to the Sauter mean diameter. Thus, since the Sauter mean diameter is almost uniform along the channel radius, the interfacial area concentration profiles display the same behavior as their respective void fraction profiles.

4.1.4. Interfacial velocity

Fig. 7 shows the behavior of interfacial velocity profiles, corresponding to that of void fraction profiles in Fig. 4. The meanings of the symbols in Fig. 7 are found in Table 1. As can be expected, the interfacial velocity has a power-law profile. As shown in the figure, measured interfacial velocities can be fitted by the following function reasonably well except for $\langle j_l \rangle = 2.08$ m/s and higher void fraction.

$$v_g = \frac{n+1}{n} \langle v_g \rangle \left\{ 1 - \left| \frac{2r - (R - R_0)}{R - R_0} \right| \right\}^{1/n}. \quad (13)$$

As the area-averaged void fraction increases, the exponent increases gradually, resulting in flatter interfacial velocity profile. As the superficial liquid velocity increases, the exponent decreases gradually and approaches an asymptotic value. Since the interfacial velocity profile would basically be similar to the respective liquid velocity profile [9], the interfacial velocity profile might be attributed to the balance of the bubble-induced turbulence and shear-induced turbulence. It was observed in a round tube that for low liquid superficial velocities ($\langle j_l \rangle \leq 1$ m/s) the introduction of bubbles into the liquid flow flattened the liquid velocity profile and the liquid velocity profile approached to that of developed single-phase flow with the increase of void fraction [9]. It was also reported that the effect of the bubble

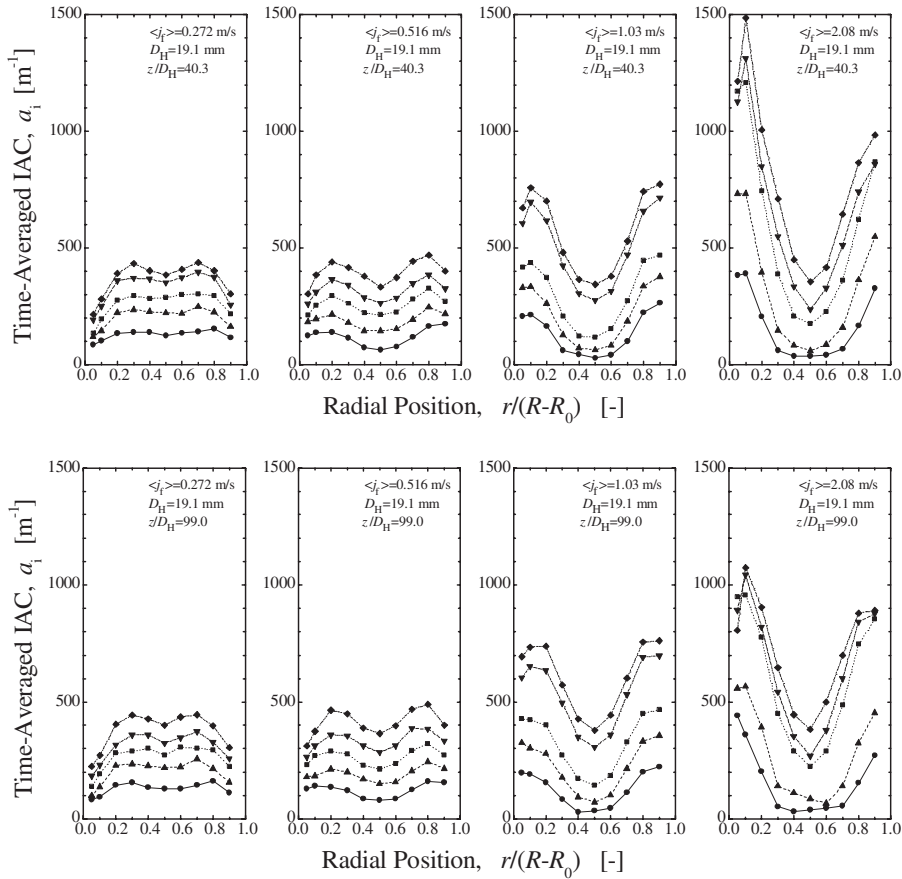


Fig. 6. Local interfacial area concentration profiles at $z/D_H = 40.3$ and 99.9 .

introduction into the liquid on the liquid velocity profile was diminishing with increasing gas and liquid velocities, and for high liquid velocities ($\langle j_l \rangle \geq 1$ m/s) the liquid velocity profile came to be the power law profile as the flow developed. Thus, for low or high liquid velocity, the bubble-induced or shear-induced turbulence, respectively, would play an important role in determining the liquid velocity profile as well as the interfacial velocity profile.

4.2. One-dimensional interfacial area transport

4.2.1. Model evaluation

Fig. 8 shows the axial developments of the area-averaged interfacial area concentrations measured in the annulus loop. The meanings of the symbols and the lines in Fig. 8 are found in Table 1. As can be seen from Fig. 8, significant axial changes of the interfacial area concentrations are not observed in most of the tested flow conditions. This may mainly be attributed to (i) axial measuring locations far from the test section inlet, (ii) relatively short distance between $z/D_H = 40.3$ and 99.0

corresponding to 1.12 m, and (iii) not extremely high void fraction and liquid velocity to promote the bubble interaction.

Millies and Mewes [14] classified the bubble coalescence and breakup processes in a bubble column into four basic processes along the flow direction such as (a) primary bubbles, (b) secondary bubbles, (c) equilibrium state of bubble coalescence and breakup, and (d) coalescence with the free surface, eventually formation of foam. The size of primary bubbles would strongly be influenced by the bubble distributor design. The primary bubbles would coalesce and breakup violently in the vicinity of the bubble distributor, and reach to the secondary bubbles. In fact, significant interfacial area transport was observed near a test section inlet even at relatively low void fraction and liquid velocity [10]. The size of the secondary bubbles would be a weak function of the bubble distributor design. This is particularly true for high liquid velocity condition [2]. The axial length for primary bubble region may be a function of liquid velocity, void fraction and initial bubble size distribution. Judging from the present test data, the bubble

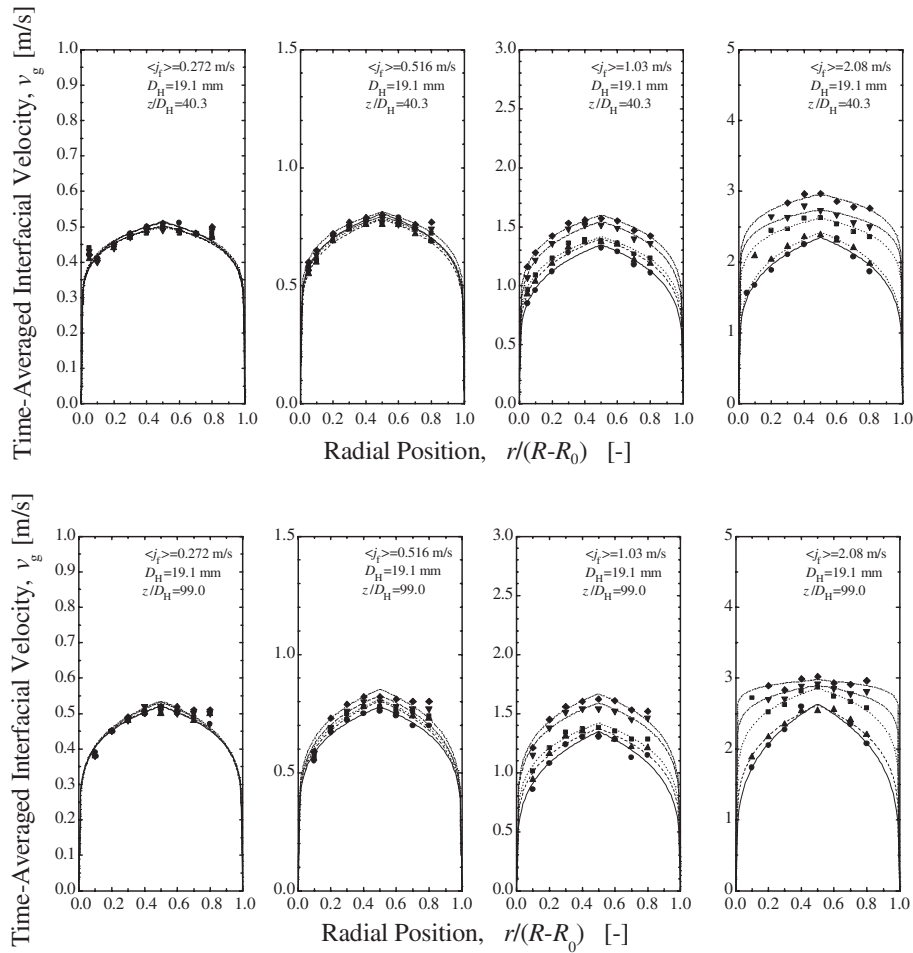


Fig. 7. Local interfacial velocity profiles at $z/D_H = 40.3$ and 99.9 .

coalescence and breakup processes seem to reach the secondary bubble region or equilibrium state region even at the first measuring station of z/D_H since the axial interfacial area transport at the downstream of the first measuring station is insignificant.

Unfortunately, the first two reasons, (i) and (ii), are inherent in the present flow loop, which was basically designed for subcooled boiling tests. In view of these, the data sets obtained in this study may not be ideal to evaluate the one-dimensional interfacial area transport equation. However, these test data may still be useful as the hydrodynamic separate effect test data without phase change to be the reference data for future subcooled bubbly flow tests and as the test data to check the applicability of the developed one-dimensional one-group interfacial area transport equation to the annulus loop. In what follows, possible discussions on the interfacial area transport mechanism will be described.

The top priority to improve the current nuclear thermal-hydraulic system analysis codes is to imple-

ment the one-dimensional interfacial area transport equation into the codes. Therefore, the applicability of the one-dimensional one-group interfacial area transport equation developed in the previous study [3] to the present annulus loop will be tested by the data obtained in this study as the first step. In Fig. 8, lines indicated the interfacial area concentrations calculated by the one-dimensional one-group interfacial area transport equation. The interfacial area concentrations measured at $z/D_H = 40.3$ are utilized as the initial values. It is recognized that the one-dimensional one-group interfacial area transport equation may reproduce proper trends of the interfacial area transport depending on flow parameters. As shown in Fig. 9, the one-dimensional one-group interfacial area transport equation gives excellent predictions of the interfacial area concentrations ranging over one order with an average relative deviation of $\pm 8.96\%$, provided the accurate initial values are given. Thus, in the interfacial area transport calculation, the importance of the initial value

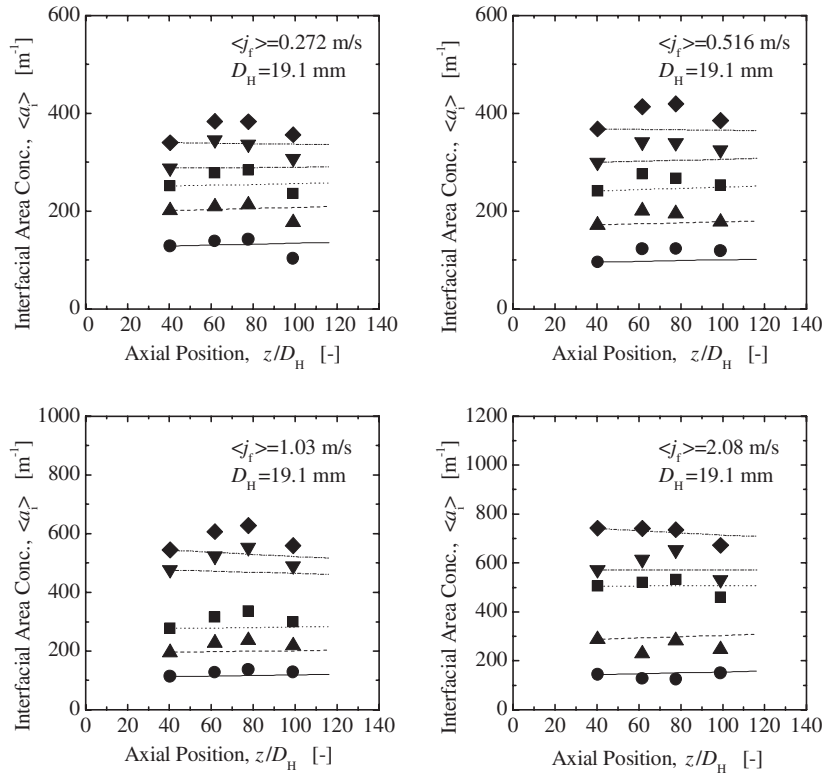


Fig. 8. Evaluation of one-dimensional interfacial area transport equation with data.

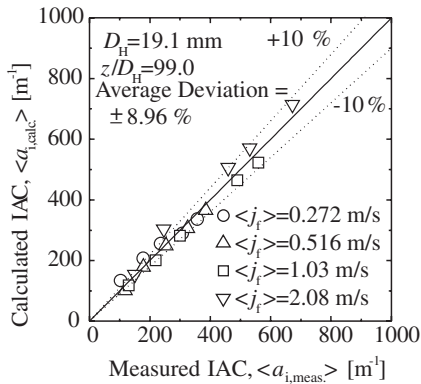


Fig. 9. Comparison of predicted and measured interfacial area concentrations.

should be emphasized. In subcooled boiling flow, the mechanistic modeling of the active nucleation site density, the bubble departure size, and the bubble departure frequency should be modeled accurately to provide accurate initial bubble size [15].

In order to evaluate the contribution of each source or sink term to interfacial area transport, the interfacial

area concentration change due to each mechanism along the axial direction is calculated. Figs. 10–12 show the contributions of bubble random collision, wake entrainment and bubble expansion to the interfacial area transport, respectively. The meanings of the lines in Figs. 10–12 are found in Table 1. Since the values of Weber number in the present experiment are smaller than the critical Weber number, the values of $\langle \phi_{TI} \rangle$ are calculated to be zero. As a general trend, the bubble expansion term, $\langle \phi_{EXP} \rangle$, governs the interfacial area transport at relatively low liquid velocity and void fraction, where bubble–bubble and bubble–eddy interactions are weak. The bubble coalescence terms, $\langle \phi_{RC} \rangle$ and $\langle \phi_{WE} \rangle$, are enhanced in the interfacial area transport equation at high void fraction, where the distance between bubbles is short enough to cause the bubble coalescence. It can be seen from these figures, the interfacial area concentration decrease rates due to the bubble coalescence terms, $\langle \phi_{RC} \rangle$ and $\langle \phi_{WE} \rangle$ are almost comparable to the interfacial area concentration increase rates due to the bubble expansion term, $\langle \phi_{EXP} \rangle$ for most of the present experimental conditions. These consequently lead to insignificant interfacial area transport along the flow direction for most of the present experimental conditions as shown in Fig. 8.

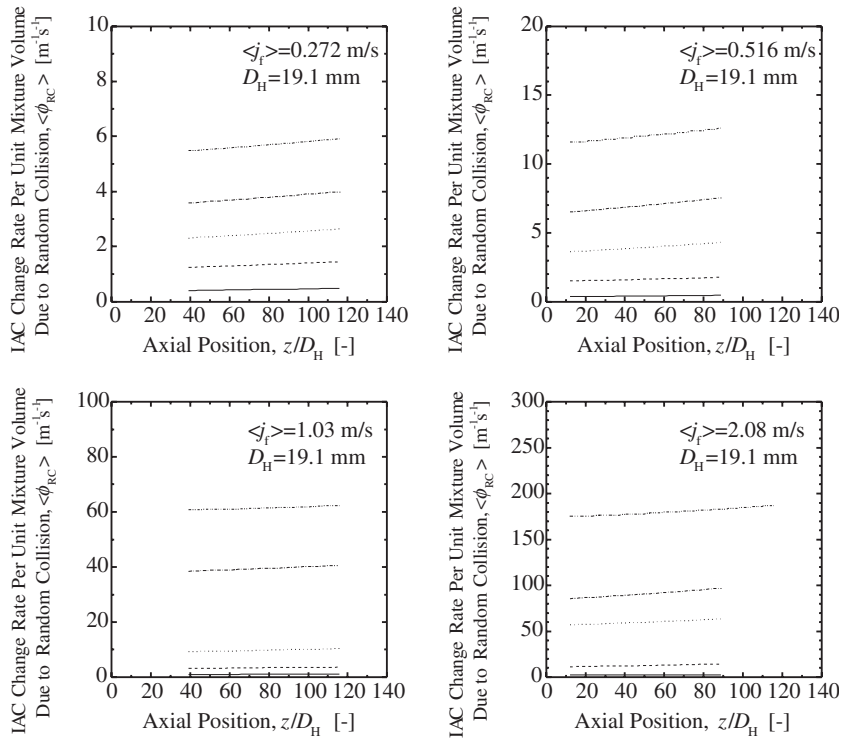


Fig. 10. Contribution of bubble coalescence due to bubble random collision to interfacial area transport.

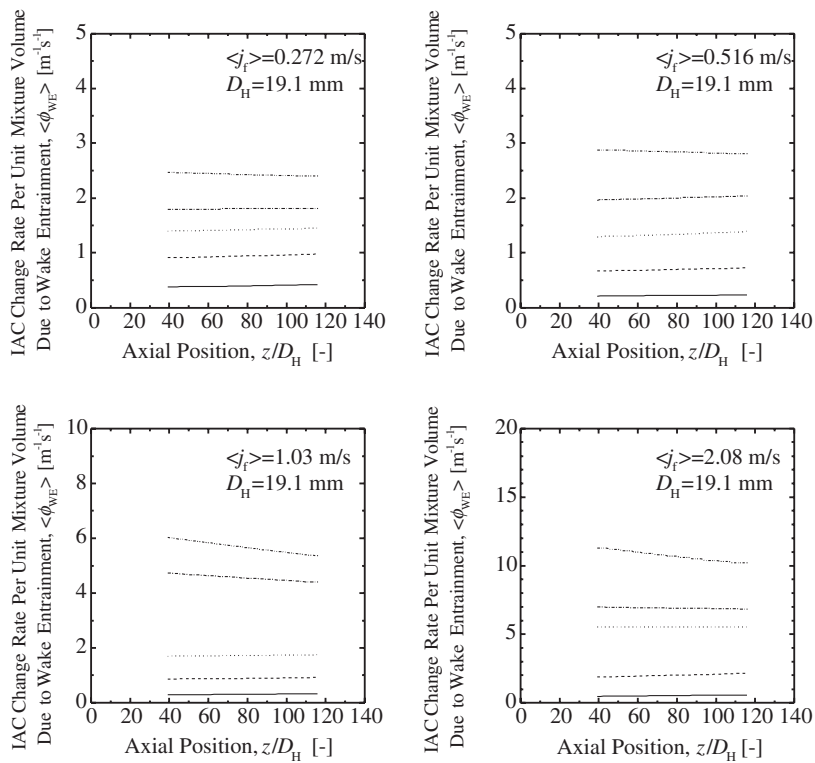


Fig. 11. Contribution of bubble coalescence due to wake entrainment to interfacial area transport.

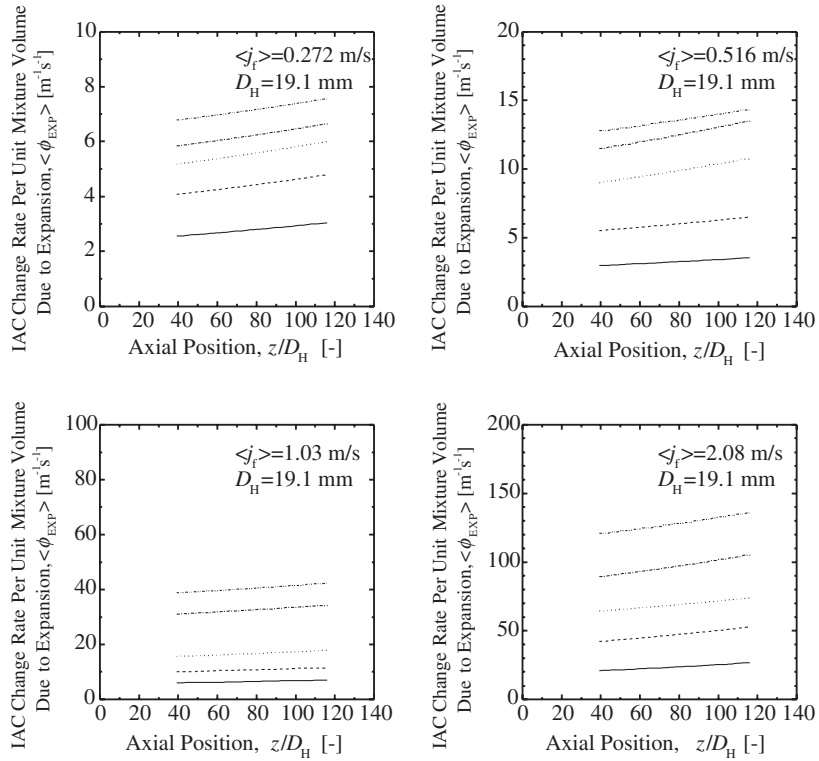


Fig. 12. Contribution of bubble expansion to interfacial area transport.

4.2.2. Sensitivity analysis on initial bubble size

The sensitivity analysis of the initial bubble size to the interfacial area concentration is performed to investigate the effect of the initial bubble size on the interfacial area transport. Fig. 13 depicts an example of the sensitivity analysis for $\langle j_{g,N} \rangle = 0.910$ m/s and $\langle j_f \rangle = 2.08$ m/s. As can be seen from Figs. 10–12, experimentally observed dominant mechanism on the interfacial area transport is bubble coalescence for $\langle D_{Sm,meas.} \rangle_{z/D_H=40.3} = 2.10$ mm, see the solid line. If smaller bubbles such as $\langle D_{Sm,meas.} \rangle_{z/D_H=40.3} = 1.0$ mm are generated at the inlet, the bubble coalescence is enhanced significantly, see the broken line. On the other hand, if larger bubbles such as $\langle D_{Sm,meas.} \rangle_{z/D_H=40.3} = 3.0$ mm are generated at the inlet, the dominant mechanism on the interfacial area transport becomes bubble breakup instead of bubble coalescence, see the chain line. Thus, the dominant mechanism of the interfacial area transport is strongly dependent on the initial bubble size, and the interfacial area transport equation can reproduce the dependence of the interfacial area transport on the initial bubble size reasonably.

In a future study, the experiment should be performed in the flow conditions where significant bubble interactions are observed to evaluate the interfacial area transport equation extensively. The local data should also be utilized to evaluate two-dimensional interfacial

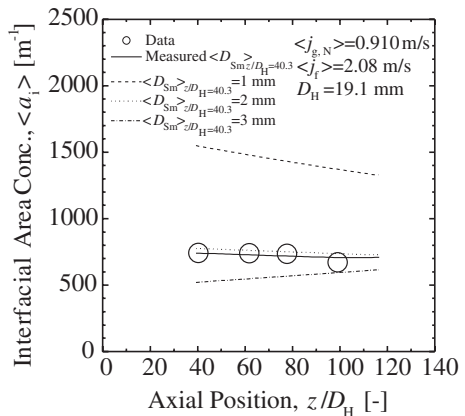


Fig. 13. Sensitivity analysis on initial bubble to interfacial area transport.

area transport with the help of various component models to calculate the flow field as the next step.

5. Conclusions

As a first step of the development of the interfacial area transport equation in a subcooled boiling flow, the

one-dimensional interfacial area transport equation was evaluated by the data taken in the hydrodynamic separate tests without phase change for an adiabatic air–water bubbly flow in a vertical annulus. The annulus channel consisted of an inner rod with a diameter of 19.1 mm and an outer round tube with an inner diameter of 38.1 mm, and the hydraulic equivalent diameter was 19.1 mm. Twenty data sets consisting of five void fractions, about 0.050, 0.10, 0.15, 0.20, and 0.25, and four superficial liquid velocities, 0.272, 0.516, 1.03, and 2.08 m/s were used for the evaluation of the one-dimensional interfacial area transport equation. The one-dimensional interfacial area transport equation agreed with the data with an average relative deviation of $\pm 8.96\%$. Sensitivity analysis was also performed to investigate the effect of the initial bubble size on the interfacial area transport. It was shown that the dominant mechanism of the interfacial area transport was strongly dependent on the initial bubble size.

Acknowledgements

The research project was supported by the Tokyo Electric Power Company (TEPCO). The authors would like to express their sincere appreciation for the support and guidance from Dr. Mori of the TEPCO.

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