INTERFACIAL AREA TRANSPORT EQAUTION EVALUATION METHODOLOGY IN LARGE DIAMETER PIPES

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Abstract

Two-phase flows in large diameter pipes are significant throughout nuclear power systems for accident analysis and, in BWRs, steady-state performance analysis. Currently predictive computer codes use static, flow-regime-dependent models to predict the interfacial area concentration, which describes the interfacial geometry of a two-phase flow. To improve the ability of these codes to predict the behavior and development of two-phase flows, it has been proposed that the Interfacial Area Transport Equation (IATE) be used. The IATE is well-developed for small diameter pipes, or those in which stable slug flows can exist, however for large diameter pipes where stable slug bubbles cannot form very little data is available. Therefore an experiment has been undertaken to measure the local profiles of various two-phase flow parameters at several axial positions in large diameter pipes of varying diameters. The results are presented here along with a method for evaluating the performance of the IATE and the results of the IATE performance evaluation for the experimental data.

Keywords: Interfacial Area, Large Diameter Pipes, Two-Fluid Model, Interfacial Area Transport Equation

1. Introduction

Two-phase flows occur in many common industrial applications, and many of these make use of large diameter pipes. This is especially true of the chemical and petroleum industries, where bubble column reactors and large pipe pumping systems are common. In the nuclear industry two phase flows often occur in large channels. For this reason, fundamental knowledge in this area is essential for nuclear reactor safety. In next-generation BWR systems, for example, flow through the reactor core is driven by natural circulation. To establish natural circulation a long chimney section, which behaves as a large diameter pipe, is required above the core. This region is extremely sensitive to variations in the flow, especially during reactor startup.

Flows in large pipes have significant differences from flows in small pipes. Once the channel diameter is larger than the maximum cap bubble size defined by Kataoka and Ishii[1] a variety of fundamental changes to the hydrodynamics of the flow begin to occur. Slug bubbles bridging the entire pipe diameter can no longer be sustained due to instability on the upper surface, which causes the bubble to collapse and break apart into smaller cap-shaped bubbles. This results in significant three-dimensional recirculatory behavior and causes significant changes to the void fraction and velocity profiles. For reactor safety it is vital that the ability to model and predict two-phase flows in such systems be developed. These models will be integrated into existing thermal-hydraulic analysis codes to be used in the prediction of reactor system behavior. The most accurate way of predicted system behavior is through full-scale testing, however in the nuclear industry full-scale tests are expensive and impractical. In place of these tests, the local phenomena in reactor systems are studied separately resulting in the development of mathematical models for predicting flow behavior under a variety of conditions. These models are then solved numerically using computers. For this approach to work reliable models with appropriate constitutive relations are essential.

Most analysis codes make use of the two-fluid model, which is currently the most practical model for two-phase flows. It is more accurate than simple models like the drift-flux model while being less computationally intensive than DNS or LES. The two-fluid model treats each phase separately, with balance equations for the mass, momentum and energy of each phase. For this reason it is also sometimes called the six-equation model. The equations are then time-averaged. This introduces terms representing the transfer of mass, momentum and energy between the phases. The detailed formulation of the two-fluid model equations can be found in Ishii and Hibiki[2]. The most significant challenge in modeling systems using the two-fluid model is developing relations for predicting interfacial transfer of mass, momentum and energy. These interfacial transfer terms can be thought of as being composed of two components: the first component is the amount of interface available for transfer, or interfacial area concentration, while the second is the driving potential for the transfer. In order to close the two-fluid model, accurate constitutive relations must be developed for the driving potential and the interfacial area concentration.

Currently the interfacial area concentration has been specified using static, flow-regime-dependent correlations. This approach has some shortcomings, as it is limited by the accuracy of the flow regime transition criteria and the experimental range for which they have been validated. This static nature also limits the ability of the models to predict dynamic features of flows during transient events and in developing flows, which is particularly important for large diameter pipes due to the relatively short lengths of components. This method can also give rise to numerical instabilities and bifurcations that can degrade or prevent convergence. Further, most of these models have been developed for small pipes rather than for large-diameter channels.

For these reasons Ishii and Kocamustafaogullari[3] proposed a more dynamic approach to prediction of interfacial area concentration by developing a transport equation for the interfacial area concentration, which is modified by Ishii and Hibiki[2] as

$$\frac{\partial a_i}{\partial t}+\nabla\cdot(a_i\vec{v}_i)=\frac{2}{3}\left(\frac{a_i}{\alpha}\right)\left\{\frac{\partial\alpha}{\partial t}+\nabla\cdot\left(\alpha\vec{v}_g\right)-\eta_{ph}\right\}+\frac{1}{3\psi}\left(\frac{\alpha}{a_i}\right)^2\sum_j R_j+\pi D_{bc}^2 R_{ph} \tag{1}$$

where $a_i(\vec{x},t)$ is the average interfacial area per unit volume of fluid and $\vec{v}_i(\vec{x},t)$ is the interfacial velocity. Additionally, ψ is a bubble shape factor, η_{ph} is the change in void fraction due to phase change, α is the void fraction, D_{bc} is the critical bubble size for condensation, and R is the source or sink reaction rate. Later further refinements to the IATE accounted for the expansion or compression of the gas phase due to pressure changes and developed accurate source and sink terms for one-group bubble interactions. This was followed by the development of the two-group IATE, which treats small spherical bubbles and large cap-shaped bubbles separately due to the differences in their hydrodynamics. This allowed improved accuracy in the slug and churn-turbulent flow regimes for small pipes. In order to complete development of the transport equation these source and sink processes must be modeled mechanistically, with the effect of the hydrodynamic differences between small and large diameter channels kept in mind. To validate and benchmark these models a significant database of local flow data covering a wide range of flow conditions is required.

One of the first experiments which focused on interfacial area in large diameter pipes was performed by Smith[4]. In this experiment, he used a four-sensor electrical conductivity probe to perform measurements in test sections with 0.102 m and 0.152 m diameters. He performed 31 experiments in flow conditions ranging from bubbly flow to churn-turbulent flow and measured the local void fraction, interfacial area concentration, bubble diameter and interface velocity for two groups of bubbles.

In 2003, Sun et al[5] reported additional data including void fraction and interfacial area concentration profiles as well as bubble number frequency data. The detailed structure of the two-phase flow was investigated to determine the development of the interfacial structure along the flow direction and provide a limited database for the future development of the two-group IATE. Void fractions for this data were as high as 0.45 and included several conditions in the cap bubbly flow regime as well as the bubbly flow regime.

Later Shoukri et al[6] studied the structure of two-phase flow in a test section with 0.2 m diameter. The radial distributions of void fraction, bubble velocity, bubble size and interfacial area concentration were measured using dual-sensor optical probes. Experimental conditions were limited to void fractions smaller than 0.04 as the experiment was intended to measure the hydrodynamics at low void fractions. The authors noted that wall-peak void profiles occur only when the void fraction was very small and that the bubble sizes for flows in large pipes were smaller than the bubble sizes reported in the literature for flow in small pipes. This results in a somewhat higher interfacial area concentration for flows in large pipes under similar flow conditions.

Shen et al[7] also studied two-phase flows in a 0.2 m diameter facility using dual-sensor optical probes. Void fractions for this study were as high as 0.4, however interfacial area concentration data was only reported for some flow conditions as dual-sensor probes cannot accurately measure interfacial area when cap-shaped Taylor bubbles are present, at void fractions higher than about 0.25.

Additionally Prasser[8] performed experiments using air and water in a 0.195 m diameter test facility. A double wire-mesh sensor was used to calculate the void fraction and bubble velocities. The use of the wire-mesh sensor allowed direct reconstruction of the three-dimensional structure of the two-phase flow so that the total interfacial area concentration could be calculated directly from the measured bubble shapes. This allows rapid calculation of the two-phase flow parameters, however the resolution of the wire-mesh sensors is only about 3 mm.

Constitutive models for interfacial area concentration sources and sinks were also developed by Smith[4], who mechanistically modeled the sources and sinks of interfacial area concentration due to the various bubble interaction methods shown in Fig. 1. The models predicted the data used to develop the models quite well, however additional data is needed for a wider variety of pipe diameters and flow conditions in order to fully validate the models. To this end and experiment has been undertaken to measure the interfacial area concentration, void fraction and bubble velocity in a round pipe of diameter 0.152 m.

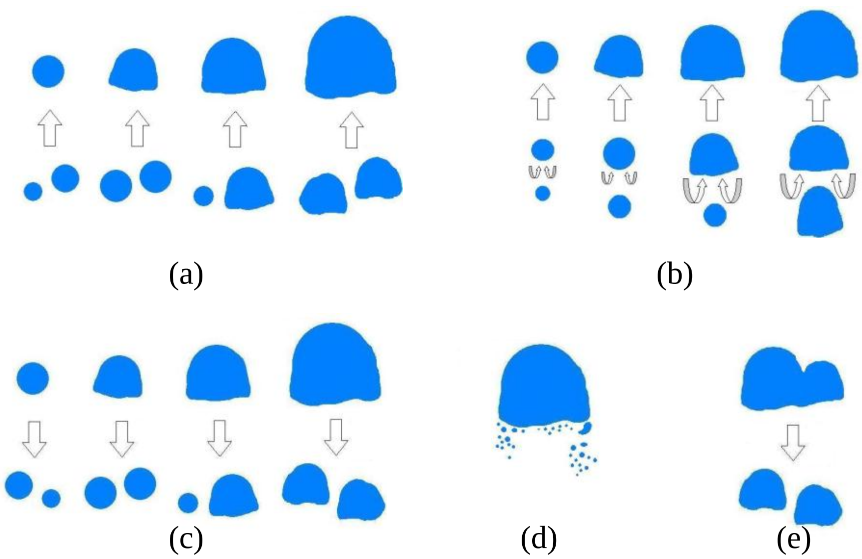

Figure 1 Source and Sink Term Mechanisms in the Two-Group IATE: (a) Random Collision, (b) Wake Entrainment, (c) Turbulent Impact, (d) Shearing-Off, (e) Surface Instability.

2. Experimental Facility

The experimental facility, shown in Fig. 2, consists of a 0.152 m diameter clear acrylic tests section. The maximum liquid flow rate is over 2 m/s, while the maximum gas flow rate is approximately 5 m/s. Liquid flow rate is measured using an electromagnetic flow meter accurate to within ±1%, while the gas flow rate is measured using Venturi gas flow meters accurate to within ±2%. Air and water enter the test section through an injector unit which produces uniform bubble injection. There are three measurement locations along the test section at z/D = 11.7, 17.7 and 33.9. Each measurement location includes pressure measurements and local void fraction and interfacial area concentration measurements using conductivity probes.

The test conditions for the experiments are shown in Fig. 3. The experiments cover flows from the bubbly flow region to the churn flow region so that the transitions can be investigated.

Superficial gas velocities are as high as 3.5 m/s, with void fractions up to 0.7. This is much higher than most of the data available in the literature for adiabatic flows.

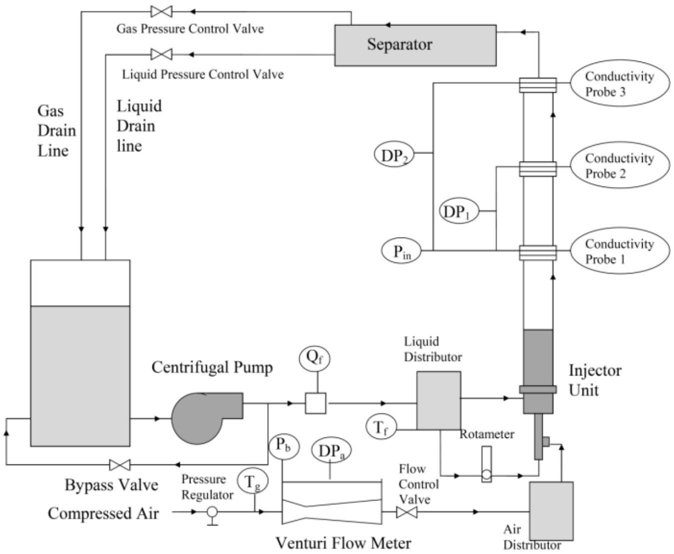

Figure 2 Schematic of Test Facility[9]

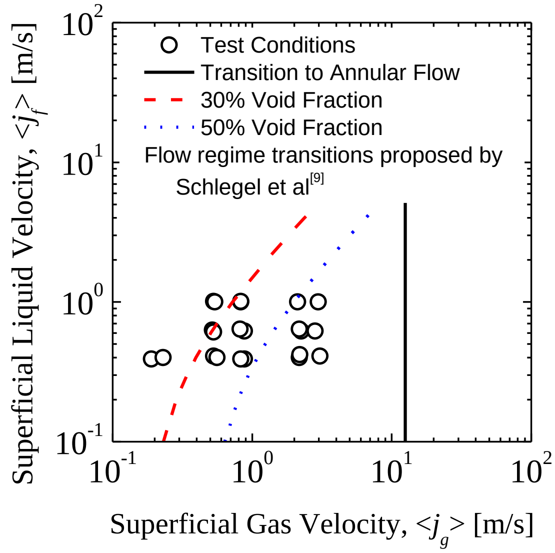

Figure 3 Test Conditions

The local void fraction, interfacial area concentration and interface velocity are measured using electrical conductivity probes. Electrical conductivity probes operate based on the fact that water conducts electricity well, while air does not. Therefore it is possible to tell whether an individual probe tip is surrounded by water or air by measuring the current passing from the probe to a common ground. In practice, this current is measured by converting to a non-dimensional voltage signal. By including multiple sensors spaced very closely, as in Fig. 4, one can calculate the velocity of the bubble interface by measuring the time taken for the surface to move from one sensor to another. Kataoka et al[10] developed a mathematical basis for calculating the interfacial area concentration based on the three-dimensional velocity that can be determined using the four-sensor conductivity probe. This approach was used by Kim et al[11] to develop an advanced signal processing scheme to calculate the interfacial area concentration from the raw voltage signal.

The accuracy of a conductivity probe measurement system depends on the accuracy of the measurements for sensor locations as well as the rate at which data is collected and the velocity of the bubble interface. This is because the sensor does not measure the exact time at which a bubble contacts the sensor, only whether the bubble contacted the sensor between two measurements. For the current experiment, the distances between sensors are measured using a microscope and a micrometer with an accuracy of 1%, and the data acquisition rate is 20 kHz. Using these settings, the maximum error expected is 12% under the highest flow rate conditions, with this value decreasing as the gas velocity decreases[11].

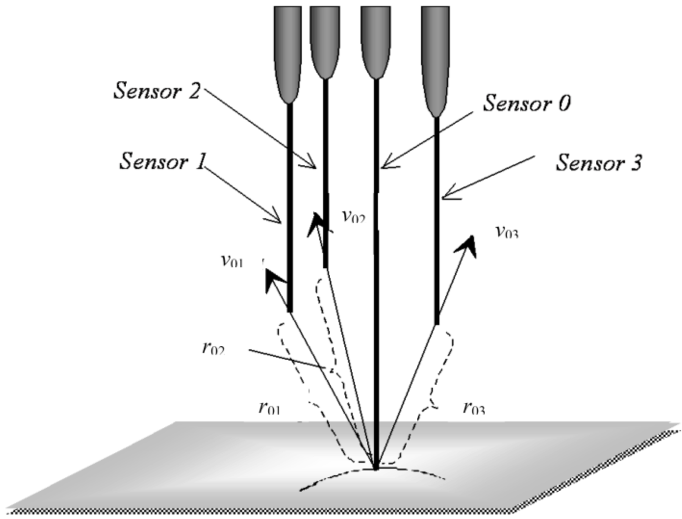

Figure 4 Four-Sensor Conductivity Probe Sensor Configuration[11]

The evaluation of the IATE begins with the local profile data. This data is then converted into area-averaged values using an area-weighted sum, resulting in area-averaged quantities at each of the three measurement locations. It is then necessary to apply the IATE numerically. A one-dimensional mesh is used, meaning that the IATE results are updated at finite locations along the test section. Then the two-group interfacial area and void transport equations are decoupled from the momentum equation in the two-fluid model by using a curve-fit of the experimentally measured void-weighted area-averaged gas velocities. This eliminates the need to solve the two-fluid model in addition to the IATE and simplifies the process. The void fraction and interfacial area concentration for each group at the lowest measurement location are then used as the inlet conditions. The IATE is then solved progressively along the test section using an iterative scheme, with the source and sink terms calculated at each location and applied through the finite distance between the current and next location. This process is then repeated for each axial mesh position until the calculations have been performed for the entire test section. Comparison with the measured values at the two remaining measurement locations allows evaluation of the development of void fraction and interfacial area concentration for each group.

3. Experimental Results

As mentioned previously, the evaluation of the IATE begins with the local profile data. For this reason experiments were performed for a variety of conditions to measure the local void fraction, interfacial area concentration, and other quantities of interest. A selected sample of the local profile data, for flow conditions with liquid velocity of 1 m/s, is shown in Figs. 5 through 8. Each plot shows a different measured parameter. In Fig. 5, the group 1 void fraction profiles show that for the lowest gas velocity, the group 1 void fraction peaks in the middle ranges of *r/R* rather than a real center or wall peak. This phenomenon disappears at higher gas velocities as the profile becomes largely flat except very near the pipe wall, which is as expected. In addition, the group 2 void fraction profile shows that the shape of the profile does not change much as the gas velocity increases, showing a strong center peak. This is true until the maximum void fraction reaches values over 0.95 locally, at which point the profile begins to flatten.

The interfacial area concentration profiles in Fig. 6 show the expected behavior. For group 1, the interfacial area concentration generally follows the void fraction profile. For group 2, there is a much stronger peak in interfacial area concentration than in the void fraction. This indicates that the measured bubble size is somewhat larger in the mid ranges of *r/R* for higher gas velocity conditions, namely churn-turbulent flow.

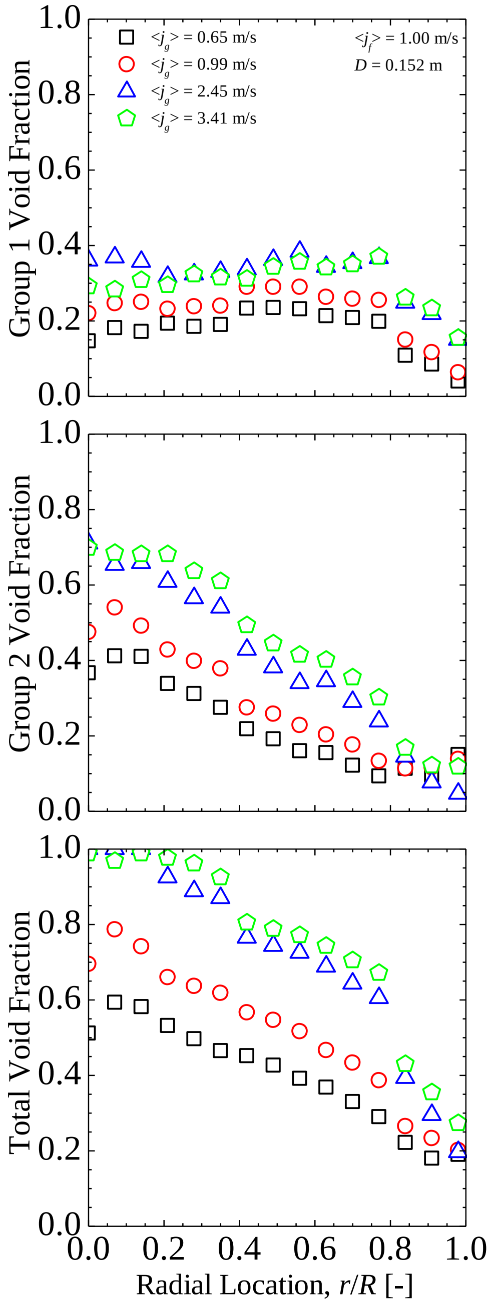

Figure 5 Some Void Fraction Profiles

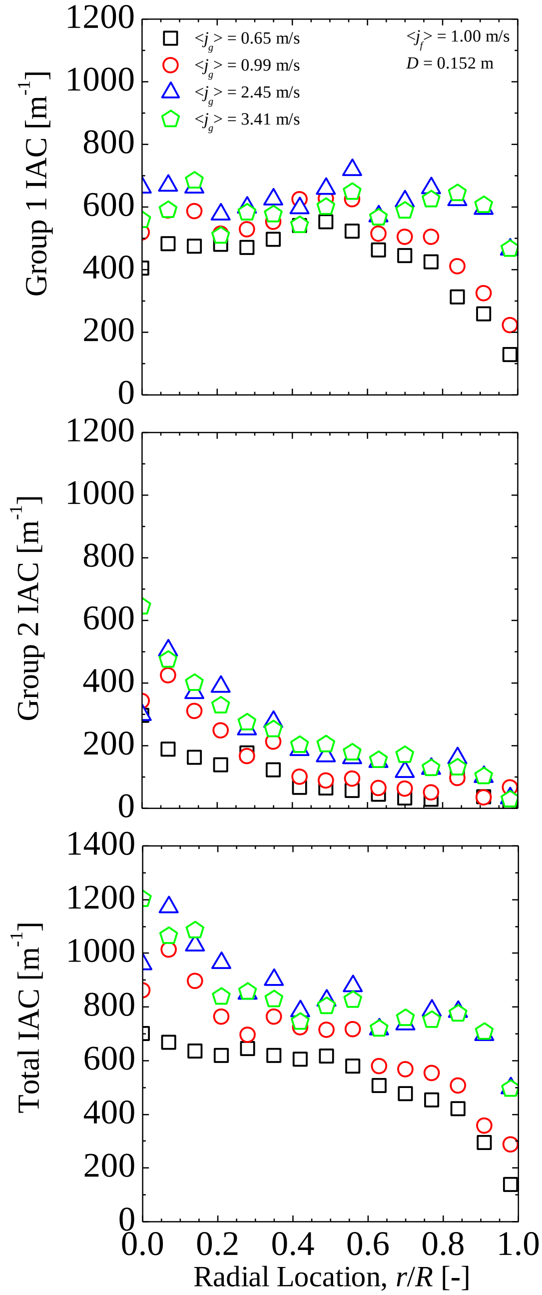

Figure 6 Some IAC Profiles

The gas velocity profiles in Fig. 7 also show some interesting patterns. For group 1, the gas velocity remains nearly constant across the entire radius of the pipe. This indicates, as described in previous studies, that enhanced turbulent mixing occurs[4]. The introduction of group 2 bubbles generates additional turbulence, especially near the walls, which results in a flatter velocity profile for the liquid. As group 1 bubbles, due to their small size, generally follow the liquid velocity the group 1 velocity profile is as shown. For group 2 however the velocity profile shows much more variation. There are several possible reasons for this. First is that it is expected that larger bubbles will migrate to the pipe center. These bubbles have more buoyancy and therefore a faster rise velocity. In addition, for the highest gas velocity values, churn-turbulent bubbles show wide variation in velocity simply due to the hydrodynamics of the flow. Group 2 bubbles also show much more distortion in bubble surfaces than group 1 bubbles, which may indicate that the measured velocity is actually the interface velocity rather than the gas velocity.

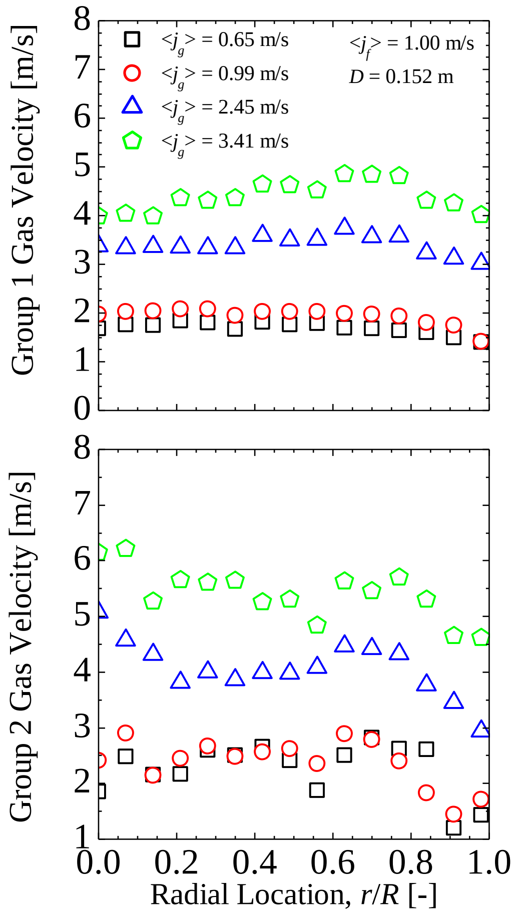

Figure 7 Some Velocity Profiles

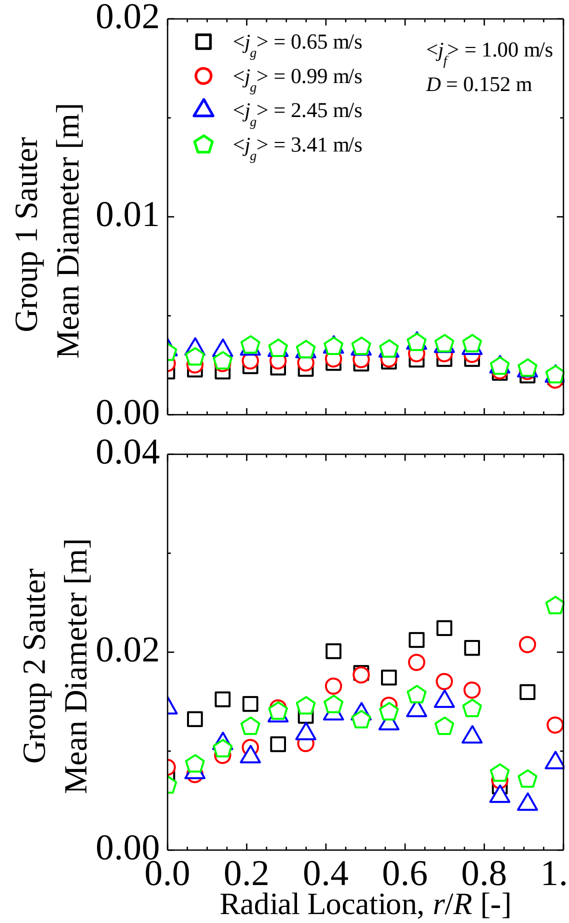

Figure 8 Some Bubble Size Profiles

Previous observations are largely confirmed by the measurements of the Sauter mean diameters for group 1 and group 2 bubbles shown in Fig. 8. For group 1 the measured diameter is nearly constant not only radially, but also for the various conditions. This shows that the hydrodynamics of the group 1 bubbles do not change much as the flow condition changes, so that the models developed for the one-group IATE for small bubbles should work even for flows with large bubbles and in a wide variety of geometries. The group 2 bubble size measurements show that the group 2 bubble sizes actually decrease as the gas velocity is increased. This likely indicates that increased turbulence generated by the presence of larger bubbles causes additional breakup. Also the maximum measured bubble sizes tend to occur at values of *r/R* near 0.6. This indicates that higher velocity bubbles near the pipe center draw liquid from the lower-velocity regions near the pipe wall, causing distortions in the bubble surface and enhancing recirculation effects. This can be noted in the data collected by Prasser[8], who used a wire mesh sensor to develop an image of the flow. That data shows that larger bubbles in the pipe center are very irregular in shape, but very large. This irregularity of shape may result in smaller bubble size measurements.

The measured data has been validated in several ways. First, the area-averaged value of the void fraction is compared to previous data collected using impedance void meters and to the predictions based on Kataoka and Ishii's drift-flux model[1]. It was shown that the agreement with the experimental data for both methods was within ±10% indicating that the void fraction measurement is reasonably accurate. It then remains to validate the velocity measurements. To do this, the gas volumetric flux, or $<\alpha \cdot v_g>$ as measured by the conductivity probes, is compared to the measured gas volumetric flux as measured by the Venturi gas flow meters in setting the experimental condition. It was found that the agreement between the two sets of data was within ±15%. This indicates that, when the uncertainty in the void fraction measurement is taken into account, that the uncertainty in the velocity measurement is about ±11.5%. This agrees well with the theoretical ±12% which was predicted.

The data from the local probes is then converted into area-averaged data and used in a numerical calculation as described above, resulting in plots such as those in Figs. 9 and 10. These plots show the interfacial area concentration and void fraction for both groups and the relative strength of the interfacial area transport mechanisms as predicted by Smith (2002). Wake entrainment is the dominant mechanism for group 2 bubbles, while the most significant source for group 1 bubbles is shearing off from group 2 and the most significant sink for group 1 is random collision between group 1 bubbles due to turbulence.

The evaluation of the IATE in this manner showed that most of the data was predicted well, to within ±10%, however some data showed as much as ±30% error. To gain greater insight into the performance of the model, the data was plotted in a flow condition map along with the flow regime transitions proposed for large diameter pipes by Schlegel et al[9]. The resulting performance map in Fig. 11 shows that for bubbly and cap-bubbly flows the data is predicted very well, however the increased distortion is shown for churn-turbulent flows. This indicates that the model does not perform as well for churn-turbulent flows. This is not unexpected, as the original data set used for development and benchmarking of the model was limited to liquid and gas superficial velocities of less than 1.0 m/s in pipes of this size[9]. Unfortunately this study did not include sufficient data in this region to allow development and validation of a new model. Additional data is needed for gas velocities higher than 1.0 m/s, or void fractions from 0.5 through the transition to annular flow, in order to complete the development of interfacial area transport models for large pipes.

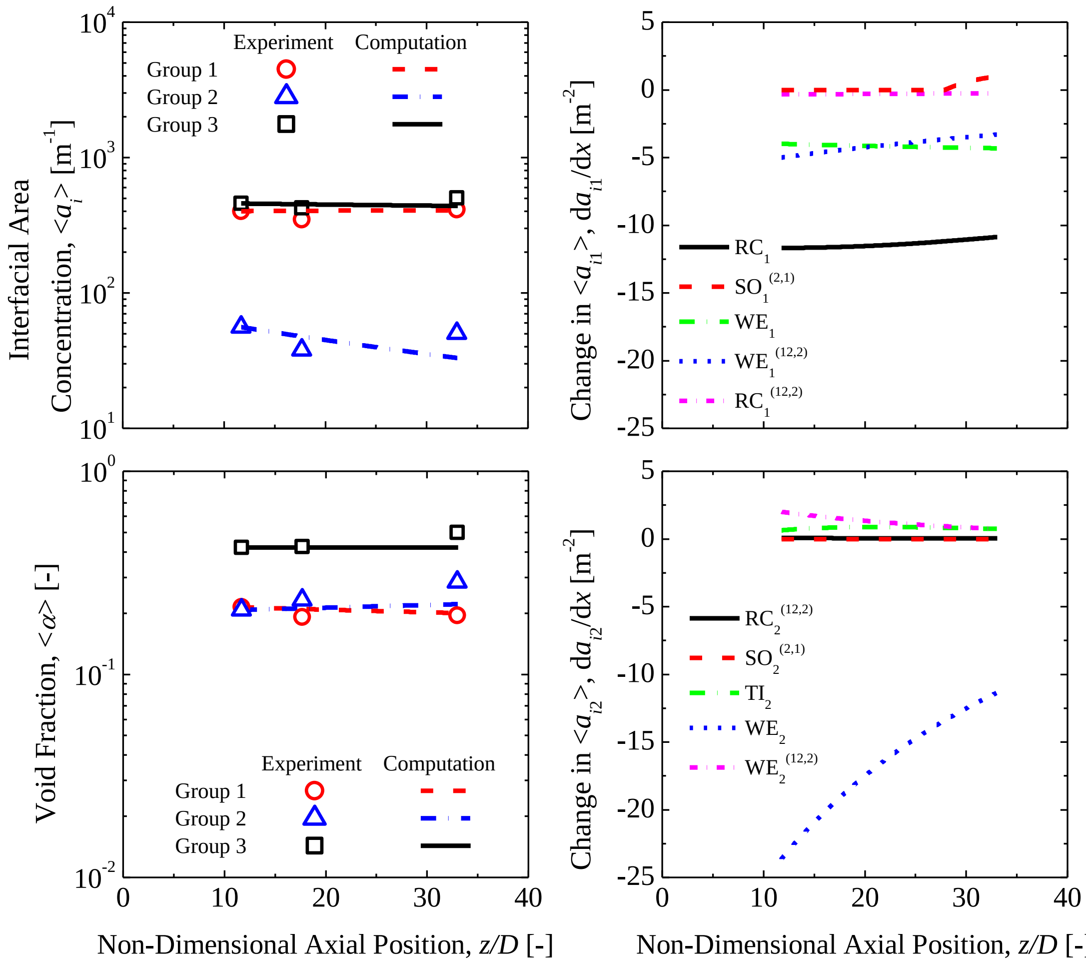

Figure 9 IATE Predictions for $\langle j_f \rangle$ = 1.00 m/s, $\langle j_g \rangle$ = 0.53 m/s

4. Conclusions

Two-phase flows in large diameter pipes are present in a wide variety of industrial applications, and understanding and modeling these flows accurately is essential for the safety analysis of future nuclear reactor systems. This modeling is done using the two-fluid model. Interfacial transfer terms in the two fluid model are modeled using the interfacial area and a driving potential. The interfacial area, which defines the geometry of the two phase flow, is best predicted using an interfacial area transport equation. Previous efforts have modeled the interfacial area concentration sources and sinks due to bubble interactions, however the present database is insufficient to validate these models over the full range of flow conditions which can be seen in reactor systems. To fill this void in the database, experiments were performed to collect local void fraction and interfacial area concentration data in a pipe of 0.152 m diameter for void fractions up to 0.7. The local profiles of the data were then used to calculate the area-averaged values and the resulting data was used as input to a computer simulation of the IATE. The computer simulation compared the IATE calculations to the collected data. It was found that the current models worked well for flow conditions within the range of experimental data used in

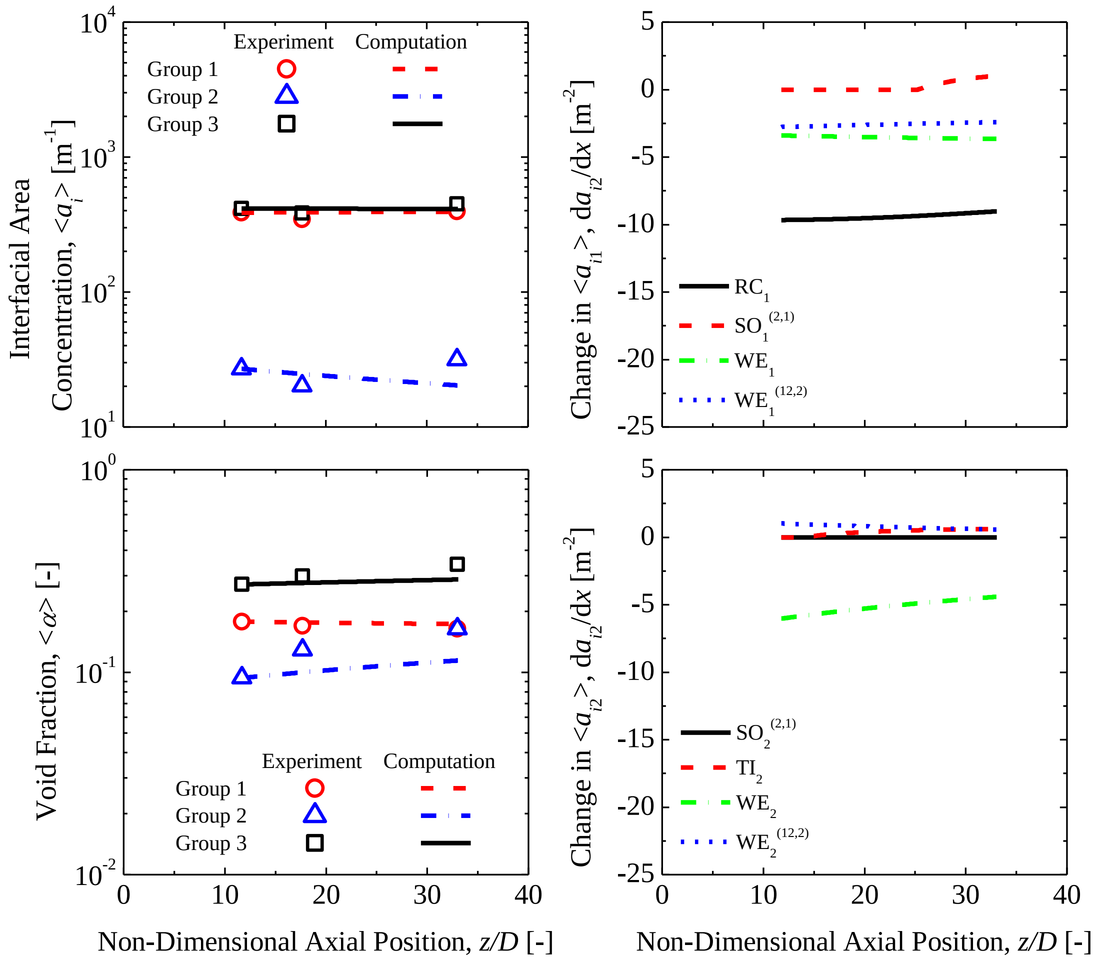

Figure 10 IATE Predictions for $\langle j_f \rangle$ = 0.40 m/s, $\langle j_g \rangle$ = 0.84 m/s

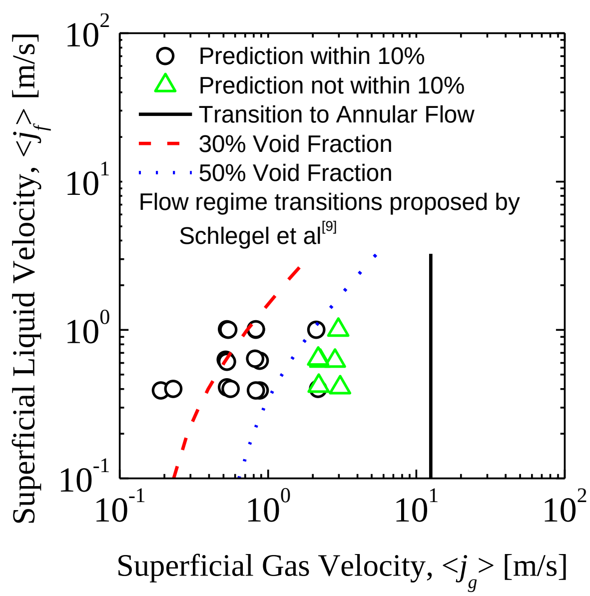

Figure 11 Final Performance Map for Existing IATE Models

the original benchmarking of the models, but that additional data is required for void fractions from 0.5 through the transition to annular flow in order to correct the models for use in churn-turbulent flows.

5. Acknowledgement

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