Characterization of Horizontal Air-Water Two-Phase Flow in a Round Pipe

Part II: Measurement of Local Two-Phase Parameters in Bubbly Flow

By:

Justin D. Talley, Ted Worosz, and Seungjin Kim*

230 Reber Building

Department of Mechanical and Nuclear Engineering

The Pennsylvania State University

University Park, PA 16802

Tel. (814)-867-1783

Fax. (814)-863-4848

Email: skim@psu.edu

* Corresponding author

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ABSTRACT

The current work seeks to develop an additional database in air-water horizontal bubbly flow through a 3.81 cm inner diameter test section with a total development length of approximately 250 diameters. The experimental conditions are chosen to cover a wide range of the bubbly flow regime based upon flow visualization using a high-speed video camera. A database of local time-averaged void fraction, bubble velocity, interfacial area concentration, and bubble Sauter mean diameter are acquired throughout the entire pipe cross-section using a four-sensor conductivity probe for nine flow conditions. To investigate the evolution of the flow, measurements are made at axial locations of 44, 116, and 244 diameters downstream of the inlet. Using this database, the effects of gas flow rate, liquid flow rate, and development length on the local and area-averaged two-phase flow parameters are presented. From the local profiles, it is found that highly turbulent liquid flow conditions cause the void fraction profile to become more uniform with increasing development length through a turbulent mixing process which counters the effect of buoyancy. Bubble interactions are also observed to play a significant role in the evolution of the flow such that the area-averaged void fraction and interfacial area concentration can have different axial trends.

KEYWORDS: horizontal two-phase flow, horizontal bubbly flow, four-sensor conductivity probe, interfacial area concentration, interfacial area transport equation

1. INTRODUCTION

In two-phase flows, the transfers of mass, momentum, and energy between the phases strongly depend on the configuration of the interfacial structure. Thus, in order to provide closure to the two-fluid model (Ishii and Hibiki, 2006), an accurate description of the interfacial structure is required. Therefore, in view of providing dynamic closure relations to the two-fluid model, the development of the interfacial area transport equation (IATE) has been a critical area of research in recent years (Ishii and Kim, 2004; Ishii et al., 2009; Kataoka, 2010). Dynamic prediction of the interfacial area concentration (a_i) is achieved through mechanistic models for bubble breakup and coalescence processes. However, experimental databases of local a_i measurements are required to develop and validate these models. While numerous air-water two-phase flow experiments have been performed in the vertical orientation (Ishii et al., 2009), relatively few studies have been made for horizontal flows. Among the available studies for horizontal two-phase flow in a straight pipe, listed in Part I of the current study (Talley et al., XXXX), most cannot be employed to develop an IATE, as they lack local a_i information due to the limitations of the measurement techniques. Only a few earlier studies that employed the double-sensor conductivity probe (Kocamustafaogullari and Wang, 1991; Huang, 1993; Kocamustafaogullari and Huang, 1994; Kocamustafaogullari et al., 1994) or double-sensor optical probe (Bottin et al., 2014) can provide such information for bubbly flow.

The most detailed study was performed by Huang (1993), where two sets of experiments were performed to investigate a_i in horizontal bubbly flow in a 5.03 cm inner diameter (ID) pipe over 15 meters in length. The first experiment investigated the profiles of local two-phase flow parameters obtained at a distance of 253 diameters downstream of the two-phase injector. The second experiment concentrated on the axial development of the profiles measured at locations

of $L/D = 25, 148, \text{ and } 253$. Both of these studies could be useful in determining the interfacial forces that dictate the phase distribution, such as lift, drag, wall lubrication, turbulent dispersion, etc., which are of particular importance for multiphase computational fluid dynamics (CFD) approaches. However, for developing an IATE applicable to horizontal bubbly flow, only the second study is useful because the measurements at multiple axial locations provide information on the evolution of the interfacial structures. This reflects the various bubble interaction mechanisms in the flow that require models in the IATE.

In the second experiment, local data was acquired throughout the cross-section using a double-sensor conductivity probe by traversing up to 23 radial positions along lines at every 22.5° for eight flow conditions (Huang, 1993). It is noted that this type of data is extremely useful in developing models and benchmarking multi-dimensional code calculations, as the entire distribution is measured. The data was acquired at 20 kHz for one second at each measurement point. Although valuable, this database has several limitations that should be considered before employing it to develop an IATE, or to benchmark code calculations. These include:

- (1) Improvements that have been made in the construction of the multi-sensor conductivity probe related to reducing intrusiveness, selecting sensor spacing for reliable data, and increasing the number of sensors to remove assumptions in calculating a_i (Kim et al., 2000);
- (2) Improvements that have been made in processing the conductivity probe signal (Kim et al., 2000);
- (3) Huang (1993) notes that the data acquisition frequency of 20 kHz may be too low to adequately capture bubbles moving at velocities up to 8 m/s;

- (4) The data acquisition time of one second may be too short to obtain statistically reliable velocity information, which is directly used to estimate a_i (Wu and Ishii, 1999).

More recently, several studies have investigated the effect of elbow restrictions on horizontal bubbly flow (Kim et al., 2007; Talley et al., 2009; Kim et al., 2010; Yadav et al., 2010). In these studies, a double-sensor conductivity probe was employed to acquire local data in a 5.03 cm ID test section with a total development length of over 20 meters. Although a large number of axial positions were employed to investigate the effect of the elbow on the distribution of the two-phase flow, the cross-sectional measurement mesh was significantly less dense than that utilized by Huang (1993). For measurement locations immediately downstream of the 90-degree or 45-degree elbow, the conductivity probe was traversed along the vertical and horizontal diameters of the pipe cross-section. In the remaining measurement locations, data along only the vertical diameter was acquired. Therefore, this data may be too limited to fully investigate the effects of the elbow (Yadav et al., 2011). Although a preliminary IATE was developed for this geometry (Talley and Kim, 2010), which can predict the experimental data within $\pm 20\%$, it is desirable to develop an IATE applicable to straight horizontal pipe flows. Then, modifications to account for the effect of the elbow restrictions can be included. Hence, more fundamental studies on horizontal bubbly flow in the absence of restrictions are necessary.

Therefore, the current work seeks to establish a horizontal two-phase flow test facility and to generate an additional database in the bubbly flow regime. This database can serve to benchmark the previous database (Huang, 1993), develop an IATE model, and benchmark both multiphase CFD and one-dimensional systems code calculations.

2. EXPERIMENTAL FACILITY

The details of the experimental facility can be found in Part I of the current study (Talley et al., XXXX). Only details related to the local measurements are presented in the current paper. Along the length of the test section, multiple instrumentation ports, shown in Fig. 1, are installed at $L/D = 44$, 116, and 244 to perform measurements of the flow. These instrumentation ports allow for additional flow visualization locations, as well as measurement of the pressure using a differential pressure transducer with an accuracy of $\pm 0.1\%$. Moreover, a four-sensor conductivity probe (Kim et al., 2000) can be installed to measure local two-phase flow parameters with an accuracy of $\pm 7\%$ (Wu and Ishii, 1999). The conductivity probe is capable of measuring local time-averaged void fraction (α), bubble frequency (f_b), bubble velocity (v_g), interfacial area concentration (a_i), and Sauter mean diameter (D_{sm}).

Since horizontal two-phase flows are asymmetric in the vertical direction because of buoyancy, the instrumentation port is designed such that the probe can be traversed along the diameter of the pipe at every 22.5° azimuthally to cover the entire cross-section during the experiment without stopping the flow. The probe is traversed throughout the pipe cross-section using up to 8 angles with up to 17 radial positions per line. In total, up to 81 local measurements are made in the cross-section if the flow is symmetric about the vertical diameter, and for an asymmetric condition, up to 129 points are measured. In general, a developed horizontal flow condition is symmetric about the vertical axis. However, the possibility of an asymmetric condition exists near the two-phase injector due to entrance effects; a procedure for determining the symmetry of the condition is discussed in Section 4.2. The measurement mesh in the current study is selected based on reducing a dense mesh while maintaining accurate area-averaged values. The measurement mesh for a symmetric condition is shown in Fig. 2. For each radial

position, data is acquired at 48 kHz to ensure signal quality when measuring bubbles moving at velocities up to 7 m/s. To obtain an accuracy of $\pm 7\%$ (Wu and Ishii, 1999), the data acquisition time is either 30 seconds or 60 seconds, depending on the local bubble frequency, to acquire at least 2,000 bubbles for all positions except those at the lowest positions in the pipe. For all measurements, a criterion of at least 500 bubbles is imposed for a measurement to be included in the database.

3. TEST CONDITIONS

As an initial study, in view of developing a database for a one-group interfacial area transport equation (Ishii and Kim, 2004) applicable to horizontal flows, local measurements are made only in the bubbly flow regime. Several considerations are made in selecting the test matrix for the local measurements. An additional flow visualization study is performed to verify the transition boundary between the bubbly and intermittent flow regimes. It is found that the theoretical boundary given by Taitel and Dukler (1976) shows good agreement with the experimental data. In addition to selecting flow conditions that cover a wide range of the bubbly flow regime, the test conditions are chosen to highlight the interaction mechanisms in bubbly flow. From the flow visualization study (Talley et al., XXXX), it is observed that turbulence-impact disintegration and random-collision coalescence are active in horizontal bubbly flow. However, depending on the flow condition, one mechanism can be dominant over the other. Therefore, additional flow visualization studies are performed to identify regions where a given mechanism is dominant. By doing so, the test matrix of nine conditions shown in Fig. 3 and Table I are obtained. Here, the superficial gas velocity is expressed under the atmospheric pressure condition.

As shown in Fig. 3, there are three categories of conditions: (1) turbulence-impact dominant, (2) random-collision dominant, and (3) both mechanisms active but neither dominant. It is found that for high superficial liquid velocity conditions, with lower void fraction, the turbulence-impact mechanism is dominant. For low liquid superficial velocities near the flow regime transition boundary, the bubbles pack more densely, and individual bubble interactions cannot be easily distinguished. However, bubbles near the bottom of the bubble layer are observed to fluctuate vertically in the pipe, due to the transport by turbulent eddies, and can enter the bubble layer. Here, the bubble layer is defined as the region where the bubbles are most densely packed. Since the bubbles beneath approach the bubble layer with the turbulence fluctuation velocity and do not emerge, it is assumed that random-collision is the dominant mechanism in these conditions. It should be noted that these conditions may also contain the effect of packing-induced coalescence due to the high local void fractions as discussed in Talley et al. (XXXX). However as the flow conditions are bubbly, it is not a dominant mechanism. Additional conditions are selected where both the turbulence-impact and random-collision mechanisms are active such that the net effect depends on the interplay of the coalescence and breakup processes.

4. RESULTS

The results of the experimental study are categorized into: (1) pressure drop, (2) evolution of local two-phase flow parameter profiles, and (3) one-dimensional transport of area-averaged two-phase flow parameters.

4.1 Pressure Drop

For the current experiment, a four-sensor conductivity probe is employed at three axial locations, namely, 44, 116, and 244 diameters downstream of the two-phase injector. These locations are chosen to minimize the inlet effect while maintaining a total development length of at least 200 diameters. Pressure measurements are made at these locations to determine the pressure drop along the length of the test section. The pressure transducer is first employed to measure the local gauge pressure at $L/D = 44$ for each of the nine conditions. Subsequently, measurements are made to obtain the differential pressure between 44 and 116 diameters, and 116 and 244 diameters downstream of the inlet, respectively. For each measurement, care is taken to ensure that air is not present in the pressure line that would distort the measured value. The change in gauge pressure along the length of the test section is shown in Fig.4.

From the figure, it can be observed that the pressure drop is similar for conditions with the same j_f . This is expected, as the major contributor to the pressure change in horizontal bubbly flow is the single-phase liquid frictional pressure drop. The pressure drop is found to increase for higher values of j_f as indicated by a steeper slope in the pressure. While the effect of $j_{g,atm}$ is to increase the system pressure, little effect is observed on the pressure drop, as the slope does not change significantly. Furthermore, no entrance effect on pressure is observed, as the pressure drop between 44 to 116 diameters and 116 to 244 diameters downstream of the inlet generally does not differ more than the accuracy of the pressure transducer.

To predict the frictional pressure drop in the test section, the Lockhart and Martinelli (1949) approach is employed, as given by:

$$\phi_f^2 = 1 + \frac{G}{X} + \frac{1}{X^2} \quad (1)$$

with the two-phase multiplier, ϕ_f , and the Martinelli parameter, X , defined as:

$$\phi_f^2 = \left(\frac{dp}{dz}\right)_F^{2\phi} / \left(\frac{dp}{dz}\right)_F^f, X^2 = \left(\frac{dp}{dz}\right)_F^f / \left(\frac{dp}{dz}\right)_F^g \quad (2)$$

where dp/dz denotes the pressure drop per unit length. Here, the subscript F denotes frictional loss, and the superscripts 2ϕ , f , and g denote the two-phase mixture, liquid phase, and gas phase, respectively. The frictional loss for each phase is obtained by:

$$\left(\frac{dp}{dz}\right)_F^k = \frac{2f}{D} \rho_k j_k^2 \quad (3)$$

where f , D , ρ , and j are the friction factor, pipe diameter, density, and superficial velocity, respectively. The script k denotes the phase index of f or g . Hence, the two-phase frictional multiplier is estimated by finding an appropriate value for the parameter C in Eq. (1). This value is found graphically from the experimental data as shown on a log-log plot in Fig. 5.

From the figure, it is observed that the best fit to the data is obtained using $C = 25$, with a disagreement of $\pm 1.04\%$. However, the conventional value of $C = 20$, for a turbulent-turbulent flow (Chisholm, 1967) also performs well with a disagreement of $\pm 1.37\%$. Here, the turbulent-turbulent value is selected in horizontal bubbly flow, since the transport of the dispersed bubbles is essentially determined by the liquid phase, which is highly turbulent. It can also be noted that the pressure drop is mainly dominated by the liquid phase frictional loss, as the multiplier is generally less than 1.05. The two-phase pressure drop is obtained from the product of the two-phase multiplier and the liquid frictional pressure drop calculated by Eq. (3).

4.2 Local Two-Phase Flow Parameters

In presenting the evolution of the local two-phase flow parameters, the influences of the following factors are discussed: (1) superficial gas velocity, (2) superficial liquid velocity, and

(3) development length. The effects of these on the void fraction, interfacial area concentration, Sauter mean diameter, and mean bubble velocity are considered.

Before analyzing the data obtained with the conductivity probe, benchmarking of the data is presented. The benchmark compares two measures of the volumetric gas flux through the pipe cross-section at a given axial location. The first measure of the local superficial gas velocity, $\langle j_g \rangle_{L/D}$, is obtained by area-averaging the product of the local void fraction and bubble velocity measurements obtained with the conductivity probe. This is given by:

$$\langle j_g \rangle_{L/D} = \langle \alpha v_g \rangle_{L/D} \quad (4)$$

Here, the subscripts g and L/D denote the gas phase and local axial position, respectively. The second measure of $\langle j_g \rangle_{L/D}$ is obtained from the measurements of the inlet gas flow rate and pressure at the axial location of interest. This is given by:

$$\langle j_g \rangle_{L/D} = \frac{Q_{g,L/D}}{A} = \frac{Q_{g,atm}}{A} \left(\frac{P_{atm}}{P_{atm} + P_{L/D}} \right) \quad (5)$$

where Q , A , and P are the volumetric flow rate, pipe cross-sectional area, and pressure, respectively. The subscript atm denotes the atmospheric pressure condition. Since $Q_{g,atm}$ is measured by the rotameters (Dwyer RMB Series, accuracy $\pm 3\%$ of scale), and $P_{L/D}$ is measured by the pressure transducer, Eqs. (4) and (5) compare local and global measurement techniques for the same parameter, respectively. Based on the accuracies of the four-sensor conductivity probe, pressure transducer, and flow meters, measurements that benchmark within approximately $\pm 10\%$ are considered reliable.

This procedure is also applied to verify the symmetry of the flow by testing the benchmark of the data after measuring half of the cross-section. If the data is found to benchmark well, and contain no significant asymmetry along the horizontal axis, the flow is

considered symmetric and the measurement is complete. However, if the data benchmarks poorly or strong asymmetry is observed along the horizontal axis, the remaining half of the cross-section is measured. The benchmark comparison is a more thorough test of the flow symmetry than a visual inspection alone, as it compares area-averaged local data against a globally measured parameter. If the area-average of the local data does not match the global measurement, information is missing in the local measurement.

Using this procedure, the complete conductivity probe benchmark for all nine flow conditions at the three axial measurement locations is shown in Fig. 6. From the figure, it can be observed that all the conditions generally benchmark within $\pm 10\%$, with an average disagreement of approximately $\pm 5\%$. Therefore, the data are reliable for analysis and model development.

4.2.1 *Effect of Superficial Gas Velocity*

The effect of increasing $j_{g,atm}$ for a fixed j_f on the measured contour plots of local void fraction and interfacial area concentration are shown in Fig. 7. In the figure, r/R denotes the radial location, with +1 and -1 being the upper (or left) and lower (or right) pipe wall in the direction of flow, respectively. The subscripts V and H denote the vertical axis (i.e., 90°) and horizontal axis (i.e., 0°), respectively. As shown in the figure, both α and a_i exhibit peaks near the upper wall consistent with the flow visualization studies (Talley et al., XXXX). This is expected, as buoyancy promotes the bubbles to move upwards. From the contour plots, it is observed that α and a_i are not uniform in the lateral (i.e., horizontal) direction, but rather, exhibit the effect of the pipe wall. This effect causes the curvature observed in the contour plots, as the local values of α and a_i approach zero at the wall. Hence, the curvature increases near the upper pipe wall. Therefore, it is necessary to measure the data along multiple angles to capture this

effect and to obtain correct area-averaged values. Specifically, area-averaged values may be overestimated by assuming a uniform profile in the lateral direction.

The profiles of α and a_i are found to be similar, since the current flow conditions are in the bubbly flow regime. However, the local values of both α and a_i exceed those generally observed in vertical bubbly flows. Here, the local values of void fraction are larger in the upper half of the pipe cross-section because of the buoyancy effect due to the flow orientation. As noted in Part I of the current study, the absence of the wake-entrainment mechanism in horizontal bubbly flow diminishes coalescence (Talley et al., XXXX). Therefore, the high values of a_i reflect the packing of numerous small bubbles, on the order of one to two millimeters in diameter, in the upper half of the pipe cross-section.

It can be observed that as the gas flow rate increases, the peak values increase. Moreover, the profiles become fuller, as the global values of α and a_i increase. This effect is more clearly observed in Fig. 8 where data only along the vertical axis are shown for the same liquid flow rate. From the figure, it can be observed that the peak position remains fairly constant between $r/R \sim 0.85 - 0.90$ regardless of the superficial gas velocity. This position corresponds to approximately one bubble diameter away from the upper pipe wall. Below this position the void profiles exhibit a linear trend, followed by an exponential tail region. It should be noted that in Fig. 8, all markers represent measured values, regardless of magnitude. For locations where a measurement cannot be made due to the absence of bubbles, no marker is presented.

While increasing $j_{g,atm}$ has a large effect on the α and a_i profiles, smaller effects are observed on the Sauter mean diameter and bubble velocity. Here, the Sauter mean diameter

represents a length scale of the bubble through the volume to surface area ratio, which can be expressed as:

$$D_{sm} = \frac{6\alpha}{a_i} \quad (6)$$

In the current study, where the bubbles are generally spherical or near spherical in shape, the Sauter mean diameter describes the average bubble diameter. Similarly, the velocity measured by the conductivity probe is the interfacial velocity, v_i , which for spherical and near spherical bubbles is approximately equal to the bubble velocity (Kim et al., 2000). For completeness, the local D_{sm} and v_g values for the conditions presented in Figs. 7 and 8 are shown in Fig. 9. It is found that the D_{sm} and v_g results in the bubble region are much less skewed than the α and a_i values; hence only data along the vertical axis are presented.

From Fig. 9 (a), it is observed that the D_{sm} profiles tend to exhibit a power law behavior with the largest bubbles residing near the center of the pipe. The bubble size is found to increase slightly with an increase in $j_{g,atm}$, which indicates that coalescence is occurring. This is supported by the results for the $j_{g,atm} = 0.50$ m/s condition, which has the largest bubble size, yet the highest system pressure for the $j_f = 5.00$ m/s conditions, as shown in Fig. 4.

It is found that the bubble velocities also tend to exhibit power law behavior, shown in Fig. 9 (b), indicating that the bubbles may closely follow the liquid. However, due to the high local void fractions, the direct measurement of the liquid velocity by optical or intrusive methods is problematic. Therefore, to investigate the ratio of the bubble velocity with respect to the liquid velocity, or slip ratio, the liquid velocity profile is estimated using the same approach employed by Kocamustafaogullari and Huang (1994) and Hibiki et al. (1998), where:

$$v_f(r) = v_{f,max} \left(1 - \frac{r}{R}\right)^{1/n} \quad (7)$$

with:

$$v_{f,max} = \frac{(n+1)(2n+1)}{2n^2} \frac{j_f}{1-\langle\alpha\rangle} \quad (8)$$

where v_f , $v_{f,max}$, n are the estimated local liquid velocity, maximum liquid velocity, and profile power, respectively. It should be noted that these expressions only consider a symmetric and nearly uniform void fraction profile in the pipe; therefore, the distribution effects in horizontal flow are not included. However, for high j_f conditions with low local void fractions, these assumptions become more valid.

It can be observed that the bubble velocity profile is not exactly symmetric, but higher in the lower void fraction section. Using Eqns. (7) and (8) with $n = 7$ (Kocamustafaogullari and Huang, 1994; Hibiki et al., 1998), it is found that the estimated local slip ratios are on the order of unity. For Fig. 9 (b), the values along the vertical axis vary from 0.73 to 0.99. The smaller values correspond to the high local α region and the larger values correspond to the low local α region. By comparing the measured v_g profiles to the estimated v_f profiles it can be inferred that the dense bubble layer affects the local liquid velocity, assuming that the shape of the v_g profile corresponds to the true v_f profile. This may make the use of Eqs. (7) and (8) questionable. Furthermore, it is found that the bubbles move faster lower in the pipe and it is speculated that the resistance of the dense bubble layer promotes the liquid to travel the path of less resistance found lower in the pipe cross-section. This leads to an asymmetric liquid velocity profile and is consistent with the findings of Iskandrani and Kojasoy (2001), where local liquid velocity measurements were made in horizontal bubbly flow using hot-film anemometry. Therefore, the symmetric liquid velocity profile assumed in Eq. (7) may not be valid in two-phase flow conditions with high local void fractions. Hence, the calculated local slip ratios may not be reliable. Considering the likelihood of a non-symmetric liquid velocity profile, there is more

confidence in the average liquid velocity than the local estimated profile. Therefore, a slip ratio based on void-weighted phase velocities may provide a more accurate description of how closely the bubbles follow the liquid. For Fig. 9 (b) this ratio is approximately 0.84, indicating that appreciable slip can occur in low void fraction bubbly flows, which is consistent with the findings of Andreussi et al. (1991).

Similar to the D_{sm} profiles, $j_{g,atm}$ has little effect on v_g . Moreover, a consistent trend in v_g with increasing $j_{g,atm}$ is not found throughout the cross-section. For vertical upward flows the slip ratio is generally greater than unity (i.e., bubbles move faster than the liquid) due to the effect of buoyancy. Since buoyancy acts in the direction of flow for vertical conditions, increasing $j_{g,atm}$ can lead to an increase in v_g . However, in horizontal flow where buoyancy is perpendicular to the flow direction, increasing $j_{g,atm}$ does not significantly influence the axial bubble velocity.

4.2.2 Effect of Superficial Liquid Velocity

The effect of increasing j_f on the measured contour plots of local void fraction and interfacial area concentration is shown in Fig. 10. An increase in the liquid superficial velocity is found to decrease the degree of peaking in the gas distribution and promote more uniform profiles. This occurs because as the turbulence level increases, turbulent mixing counters the effect of buoyancy. Again, the profiles of α and a_i are similar, as can be clearly seen in the data along the vertical axis shown in Fig. 11, since all conditions reside in the bubbly flow regime.

Unlike $j_{g,atm}$, which has little effect on D_{sm} and v_g , j_f is found to have a larger impact on these parameters as shown in Fig. 12. It is observed that D_{sm} decreases as j_f increases due to the increased turbulence in the flow, which promotes a smaller stable bubble size (Kocamustafaogullari et al., 1994). Moreover, it is also found that the bubbles become more

mixed throughout the cross-section with increasing j_f , as indicated by the lack of data in the lower half of the pipe for the lowest j_f condition.

Considering the gas velocity, it is observed that j_f plays a larger role than $j_{g,atm}$, as it is observed that increasing j_f causes an analogous increase in v_g . However, for all conditions it is found that the bubbles have a smaller velocity than the estimated liquid velocities. It is noted that the probe benchmark in Fig. 6, as well as bubble velocity measurements made using multiple images from the flow visualization study, support the probe measured bubble velocities (Talley et al., XXXX). As discussed previously, the estimated local v_f profiles may not be valid for conditions with large local void fractions which makes the determination of local slip ratios questionable. Hence, the slip ratio based on the void-weighted phase velocities is considered instead. For Fig. 12, the slip ratios based on void-weighted phase velocities vary in the range of 0.73 to 0.86. Moreover, it is found that the slip ratio approaches unity with increasing j_f . This trend is observed in all values of $j_{g,atm}$ for which j_f is varied. Hence, the void-weighted bubble velocity approaches the void-weighted liquid velocity as the bubbles become more distributed in the pipe with increasing j_f . This occurs because fewer bubbles reside in the low velocity region near the upper pipe wall with increasing j_f .

4.2.3 Effect of Development Length

To investigate the axial transport of the local interfacial structures, the local measurements obtained at $L/D = 44, 116$ and 244 are employed. Of particular importance in considering the local interfacial structures is the identification of entrance effects to be considered in the analysis of the one-dimensional transport. Only the conditions with the minimum and maximum entrance effect are presented. The development of the void fraction profiles for a fixed $j_{g,atm}$ with varied j_f are shown in Figs. 13 and 14. For the $j_f = 4.00$ m/s

conditions, shown in Fig. 13, it is found that the entrance effect is minimal, as the flow is symmetric about the vertical axis and the α profile continuously increases. Since the a_i profile is always similar to the α profile, only the α profiles are presented. For the $j_f = 6.00$ m/s conditions, shown in Fig. 14, the entrance effect becomes more pronounced. The entrance effect is found in both the symmetry of the flow about the vertical axis, as well as a higher peak in α and a_i at $L/D = 44$. Downstream of the first measurement port, the symmetry of the flow is restored, and the peak value decreases, presumably due to the dissipation of the entrance effect. Between $L/D = 116$ and 244 the void fraction profile always increases due to pressure drop. While the asymmetry in the α profile is the greatest in the $j_f = 5.00$ m/s conditions, the increased void peaking becomes more pronounced with increasing j_f . Therefore, it is believed that the entrance condition is related to the injection of the main liquid flow, j_{fl} . As the liquid flow rates increases, the bubbles are pushed more toward the upper pipe wall by the increasing inertia of the liquid phase.

The change in the local D_{sm} and v_g along the test section are shown in Figs. 15 and 16, respectively. Again, only the conditions with the minimum and maximum entrance effect are presented. In Fig. 15, a consistently increasing or decreasing trend in D_{sm} or v_g is not observed. However, it is found that the largest and fastest moving bubbles present at $L/D = 44$ transition to become smaller and slower by $L/D = 116$. Subsequently, between $L/D = 116$ and 244 the bubbles again grow and accelerate, but do not exceed the values observed in the entrance region. For D_{sm} , the larger bubbles in the entrance region initially disintegrate and then expand due to the pressure drop. The velocity trends are directly related to the difference in the drag, which depends on the bubble size.

In Fig. 16 it is observed that there is a consistent trend in D_{sm} and v_g , as both values are found to decrease with additional development. Therefore, the turbulence level in the $j_f = 6.00$ m/s conditions is sufficient to cause continuous disintegration of the bubbles. Moreover, it is found that the turbulence level in Fig. 16 is sufficient to cause significant mixing, as the bubble layer grows with increasing L/D . This contrasts the development shown in Fig. 15 where the bubble layer depth remains nearly constant.

4.3 Area-Averaged Two-Phase Flow Parameters

While local data are critical for computational tools and models which capture local phenomena, several prevalent nuclear reactor system codes, such as TRACE (2007) and RELAP (2009), employ one-dimensional approaches for pipe flows. In view of a one-dimensional approach, the area-averaged flow parameters are also investigated. These include: $\langle \alpha \rangle$, $\langle a_i \rangle$, $\langle \langle v_g \rangle \rangle$, and $\langle \langle D_{sm} \rangle \rangle$. Here, the $\langle \rangle$ and $\langle \langle \rangle \rangle$ brackets denote the area and void-weighted averages, respectively, given by:

$$\langle \xi \rangle = \frac{1}{A} \int \xi dA \quad (9)$$

$$\langle \langle \xi \rangle \rangle = \frac{\frac{1}{A} \int \alpha \xi dA}{\frac{1}{A} \int \alpha dA} = \frac{\langle \alpha \xi \rangle}{\langle \alpha \rangle} \quad (10)$$

where ξ is the parameter of interest. Here, the $\langle \langle D_{sm} \rangle \rangle$ is employed rather than the $\langle D_{sm} \rangle$ due to the horizontal orientation of the flow. This choice is made since below the bubble layer the local value of D_{sm} is undefined. If a value of zero is provided at these locations, then the $\langle D_{sm} \rangle$ does not represent the bubble size. By void-weighting instead, equivalent to frequency weighting, the value of $\langle \langle D_{sm} \rangle \rangle$ represents the bubble size, as only the bubble layer contributes to this value.

Since advection along the test section affects the $\langle \alpha \rangle$ and $\langle a_i \rangle$ transport, the $\langle \langle v_g \rangle \rangle$ is considered first. In Section 4.2.2, it is discussed that the bubble velocity is more strongly affected by the liquid velocity in horizontal flows, which is relatively constant due to the incompressible nature of the fluid. Hence, it is expected that the $\langle \langle v_g \rangle \rangle$ should not change significantly along the flow. This is generally true, as shown in Fig.17. However, in several conditions, the effects of the inlet condition and turbulent mixing are observed, as $\langle \langle v_g \rangle \rangle$ is also influenced by the phase distribution

From Section 4.2.3 it is observed that the inlet promotes increased bubble packing near the upper pipe wall, a lower velocity region, at $L/D = 44$. Therefore, as the turbulence mixes the flow, the bubble velocity generally increases between $L/D = 44$ and 116 (i.e., turbulent mixing dominates buoyancy). This is highlighted in Runs 6 through 9 ($j_f = 6.00$ m/s), where the inlet condition and turbulent mixing are the greatest. For these conditions, bubbles are present throughout the entire cross-section by $L/D = 116$. However, additional acceleration is observed between $L/D = 116$ and 244 for $j_f = 5.00$ m/s, where the bubble layer grows for the entire development length. In contrast, a decreasing trend in bubble velocity is observed between $L/D = 44$ and 116 for Runs 1 and 2 ($j_f = 4.00$ m/s). Here, these conditions have high local void fractions at the inlet and continue to increase with additional development length. Hence, $\langle \langle v_g \rangle \rangle$ decreases as the bubble packing in the low velocity region increases (i.e., buoyancy dominates turbulent mixing). These conditions clearly indicate the influence of the void distribution on the one-dimensional bubble velocity.

The transport of $\langle \alpha \rangle$ and $\langle a_i \rangle$ are shown in Figs. 18 through 20. The $\langle \alpha \rangle$ is found to increase along the length of the test section due to the expansion of bubbles with decreasing pressure. It should be noted that the trends of $\langle \alpha \rangle$ and $\langle a_i \rangle$ are similar in bubbly flow if

interactions are not occurring, or if coalescence and breakup are balanced. In general, it is found that $\langle \alpha \rangle$ and $\langle a_i \rangle$ trends are similar, with two notable exceptions. The first exception is Run 2 ($j_f = 4.00$ m/s and $j_{g,atm} = 0.25$ m/s), shown in Fig. 18, where it is observed that $\langle \alpha \rangle$ increases between $L/D = 116$ and 244 while the $\langle a_i \rangle$ remains constant. This indicates that coalescence is dominant in this condition, which is near the bubbly to plug flow regime transition (Taitel and Dukler, 1976). The second exception is Run 5 ($j_f = 5.00$ m/s and $j_{g,atm} = 0.50$ m/s), shown in Fig. 19, where it is observed that $\langle \alpha \rangle$ remains constant between $L/D = 44$ and 116 while the $\langle a_i \rangle$ increases. This indicates that breakup and pressure are dominant over coalescence. Here, the inlet condition forms elongated bubbles, which breakup downstream as turbulence mixes the flow. The $\langle \alpha \rangle$ remains constant between $L/D = 44$ and 116 because the void fraction associated with the elongated bubbles is disintegrated and accelerated, which counters the pressure expansion effect.

For $j_f = 5.00$ m/s, shown in Fig. 21, the $\langle \langle D_{sm} \rangle \rangle$ is found to decrease between $L/D = 44$ and 116 followed by an increase between $L/D = 116$ and 244 . The decrease is due to both the bubble velocity and breakup, while the increase results from the bubble expansion by pressure drop. However, for $j_f = 6.00$ m/s, shown in Fig. 22, $\langle \langle D_{sm} \rangle \rangle$ decreases with L/D . Hence, the turbulence level is sufficient to continuously break up the bubbles.

5. SUMMARY AND CONCLUSIONS

Despite the prevalence of horizontal pipes in nuclear reactor systems, limited studies are available which directly measure the interfacial characteristics of horizontal two-phase flows. Therefore, an experimental facility capable of simulating all separated and dispersed horizontal two-phase flow regimes is established. The test section is constructed from 3.81 cm acrylic

pipes with a total development length of approximately 250 diameters. Along the length of the test section, three local instrumentation ports are installed to investigate the axial evolution of two-phase flow parameters using a four-sensor conductivity probe. Moreover, due to the asymmetry of horizontal two-phase flow, the ports are designed such that they can be rotated to take local measurements throughout the entire cross-section without stopping the experiment.

A reliable database of local two-phase flow parameters is measured using a four-sensor conductivity probe at axial locations of $L/D = 44, 116, \text{ and } 244$ for nine flow conditions in the bubbly flow regime. Measurements of void fraction, bubble velocity, interfacial area concentration and bubble Sauter mean diameter are acquired throughout the pipe cross-section by traversing up to 17 radial locations along 8 angles. In total, this database consists of approximately 2,000 time-averaged local measurements. For all conditions, the data is found to benchmark within $\pm 10\%$ of the pressure and flow rate measurements, confirming the reliability of the data.

Using the local data acquired throughout the cross-section, contour plots of the measured two-phase flow parameters are generated. It is observed that while the bubble frequency is higher near the upper pipe wall, the lateral distribution is found to show the effect of the wall. Without the measurements made in the current study at different angles in the cross-section, inaccurate area-averaged values would result. Thus, with the current detailed measurement mesh, there is high confidence in the area-averaged parameters obtained.

From the data, it is found that increasing the gas flow rate increases the maximum value of the local void fraction, while increasing the liquid flow rate promotes a flatter profile. The profiles of α and a_i are found to be similar as the bubbles are spherical and one to two millimeters in diameter due to the highly turbulent flow conditions investigated. The bubble size

and velocity are found to be strongly influenced by the liquid superficial velocity, but not the gas superficial velocity due to the lack of buoyancy and hydrostatic effects in the direction of flow. The bubble size is found to decrease with increasing liquid flow rate due to increased turbulent fluctuations. Moreover, the bubble velocity increases with increasing liquid flow rate, as the bubbles closely follow the liquid.

Through investigating the one-dimensional transport, notable effects of the interactions are found for both coalescence and breakup processes, such that $\langle \alpha \rangle$ and $\langle a_i \rangle$ can have different axial trends. Due to the inlet condition, it is found that the bubbles tend to accelerate between $L/D = 44$ and 116 , followed by a fairly constant bubble velocity downstream. In the $j_f = 6.0$ m/s conditions, it is found that the bubble size continuously decreases along the test section due to turbulence-impact disintegration. In these conditions it is also found that the void fraction profiles become more uniform with increasing development through a turbulent mixing process which counters the effect of buoyancy.

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Table I: Test matrix for horizontal bubbly flow experiment.

j_f [m/s]	$j_{g,atm}$ [m/s] (Run #)			
6.00	0.15 (6)	0.25 (7)	0.50 (8)	1.00 (9)
5.00	0.15 (3)	0.25 (4)	0.50 (5)	
4.00	0.15 (1)	0.25 (2)

Fig. 1. Exploded view of the instrumentation port assembly.

Fig. 2. Four-sensor conductivity probe measurement mesh for a symmetric flow condition.

Fig. 3. Test matrix for horizontal bubbly flow experiment. Dashed line given by Taitel and Dukler (1976).

Fig. 4. Measured pressure drop in the bubbly flow experiment.

Fig. 5. Predicted two-phase multiplier using the Lockhart and Martinelli (1949) approach. Error bars shown: $\pm 1\%$.

Fig. 6. Benchmark of $\langle j_g \rangle_{L/D}$ calculated from four-sensor conductivity probe measurement versus $\langle j_g \rangle_{L/D}$ calculated from the flow rate and pressure measurement.

(a) $j_{g,atm} = 0.15$ m/s, $\langle \alpha \rangle = 0.026$, $\langle a_i \rangle = 100.98$ [1/m]

(b) $j_{g,atm} = 0.25$ m/s, $\langle \alpha \rangle = 0.042$, $\langle a_i \rangle = 156.03$ [1/m]

(c) $j_{g,atm} = 0.50$ m/s, $\langle \alpha \rangle = 0.078$, $\langle a_i \rangle = 268.56$ [1/m]

Fig. 7. Effect of $j_{g,atm}$ on measured α and a_i contour plots for $j_f = 5.00$ m/s at $L/D = 244$.
Left: Void fraction (α), Right: Interfacial area concentration (a_i).

Fig. 8. Effect of $j_{g,atm}$ on the local (a) α and (b) a_i for $j_f = 5.00$ m/s at $L/D = 244$. Vertical axis data (90°), arrows denote trend with increasing $j_{g,atm}$.

(a)

(b)

Fig. 9. Effect of $j_{g,atm}$ on the local (a) D_{sm} and (b) v_g for $j_f = 5.00$ m/s at $L/D = 244$. Vertical axis data (90°), arrow denotes trend with increasing $j_{g,atm}$.

(a) $j_f = 4.00$ m/s, $\langle \alpha \rangle = 0.038$, $\langle a_i \rangle = 111.55$ [1/m]

(b) $j_f = 5.00$ m/s, $\langle \alpha \rangle = 0.026$, $\langle a_i \rangle = 100.98$ [1/m]

(c) $j_f = 6.00$ m/s, $\langle \alpha \rangle = 0.017$, $\langle a_i \rangle = 78.18$ [1/m]

Fig. 10. Effect of j_f on measured α and a_i contour plots for $j_{g,atm} = 0.15$ m/s at $L/D = 244$.
Left: Void fraction (α), Right: Interfacial area concentration (a_i).

Fig. 11. Effect of j_f on the local (a) α and (b) a_i for $j_{g,atm} = 0.15$ m/s at $L/D = 244$. Vertical axis data (90°), arrows denote trend with increasing j_f .

(a)

(b)

Fig. 12. Effect of j_f on the local (a) D_{sm} and (b) v_g for $j_{g,atm} = 0.15$ m/s at $L/D = 244$. Vertical axis data (90°), arrows denote trend with increasing j_f .

(a) $L/D = 44$ $\langle \alpha \rangle = 0.032$

(b) $L/D = 116$, $\langle \alpha \rangle = 0.034$

(c) $L/D = 244$, $\langle \alpha \rangle = 0.038$

Fig. 13. Effect of L/D on measured α contour plots for $j_f = 4.00$ m/s and $j_{g,atm} = 0.15$ m/s.

(a) $L/D = 44$ $\langle \alpha \rangle = 0.014$

(b) $L/D = 116$, $\langle \alpha \rangle = 0.015$

(c) $L/D = 244$, $\langle \alpha \rangle = 0.017$

Fig. 14. Effect of L/D on measured α contour plots for $j_f = 6.00$ m/s and $j_{g,atm} = 0.15$ m/s.

(a)

(b)

Fig. 15. Effect of L/D on the local (a) D_{sm} and (b) v_g for $j_f = 4.00 \text{ m/s}$ and $j_{g,atm} = 0.15 \text{ m/s}$. Vertical axis data (90°).

(a)

(b)

Fig. 16. Effect of L/D on the local (a) D_{sm} and (b) v_g for $j_f = 6.00 \text{ m/s}$ and $j_{g,atm} = 0.15 \text{ m/s}$. Vertical axis data (90°), arrows denote trend with increasing L/D .

Fig. 17. Transport of void-weighted bubble velocity, $\langle v_g \rangle$.

Fig. 18. Transport of (a) area-averaged void fraction, $\langle \alpha \rangle$, and (b) area-averaged interfacial area concentration, $\langle a_i \rangle$, for $j_f = 4.00$ m/s.

Fig. 19. Transport of (a) area-averaged void fraction, $\langle \alpha \rangle$, and (b) area-averaged interfacial area concentration, $\langle a_i \rangle$, for $j_f = 5.00$ m/s.

Fig. 20. Transport of (a) area-averaged void fraction, $\langle \alpha \rangle$, and (b) area-averaged interfacial area concentration, $\langle a_i \rangle$, for $j_f = 6.00$ m/s.

Fig. 21. Transport of void-weighted Sauter mean diameter, $\langle\langle D_{sm} \rangle\rangle$, for $j_f = 5.00$ m/s.

Fig. 22. Transport of void-weighted Sauter mean diameter, $\langle\langle D_{sm} \rangle\rangle$, for $j_f = 6.00$ m/s.