

Chapter 2

Flow Characteristics and Void Fraction Prediction in Large Diameter Pipes

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Abstract Two phase flows in large diameter pipes have immense importance in a wide variety of industrial applications. As a first approximation for the prediction of a two-phase flow and as a beginning for the development of more complex models, the drift-flux model is often used to characterize and predict flows for many geometries and flow conditions. In this chapter, the flow characteristics in flows in large diameter pipes are illustrated based on the experimental data. The flow regimes and their transition criteria are discussed. The existing drift-flux models are summarized, their strengths and weaknesses are noted and the data that can be used to evaluate these models are presented. Based on the flow regime transitions in large diameter pipes, all of the available drift-flux models are evaluated systematically in both low (bubbly) and high (cap and churn-turbulent) void fraction flows. The drift-flux type correlations of Hibiki and Ishii [14] and Kataoka and Ishii [24] are found to be able to give the best predictions for the existing low and high void fraction databases respectively and are recommended for void fraction predictions in flows in large diameter pipes.

Keywords Void fraction • Flow regime • Distribution parameter • Drift velocity • Drift-flux model • Large diameter pipe

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Nomenclature

A	Area (m^2) or A parameter in Shen et al. [45] drift-flux model (-)
B	A parameter of Shen et al. [45] drift-flux model (-)
B_1	A parameter of Chexal et al. [4] drift-flux model (-)
C_0	Distribution parameter (-)
C_2	A parameter of Chexal et al. [4] drift-flux model (-)
C_3	A parameter of Chexal et al. [4] drift-flux model (-)
C_4	A parameter of Chexal et al. [4] drift-flux model (-)
C_∞	The asymptotic value of C_0 (-)
$C_{D\infty}$	Drag coefficient of a single particle in an infinite medium (-)
D	Diameter (m)
D_b	Bubble diameter (m)
D_H	Hydraulic diameter (m)
F	Placeholder value (-)
g	Gravitational acceleration (m/s^2)
j	Superficial velocity, namely, Volumetric flux (m/s)
K_0	A parameter of Chexal et al. [4] drift-flux model (-)
L	Chexal-Lellouche fluid parameter (-)
$N_{Re\infty}$	Particle Reynolds number in an infinite medium (-)
N_μ	Non-dimensional viscosity number for continuous and disperse phases (-)
$N_{\mu f}$	Non-dimensional viscosity number for liquid and gas phases (-)
r	A parameter of Chexal et al. [4] drift-flux model (-)
r_d	Radius of a particle (m)
Re	Reynolds number (-)
Re_f	Reynolds number for liquid phase only (-)
Re_g	Reynolds number for the gas phase only (-)
$\bar{V}_{gj}$	Mean transport drift velocity (m/s)
V_{gj}^0	A drift velocity parameter of Chexal et al. [4] drift-flux model (m/s)
v	Velocity (m/s)
v_{gj}	Drift velocity (m/s)
z	Height (m)

Greek Letters

α	Void fraction (-)
μ	Dynamic viscosity (Pa s)
ρ	Density (kg/m^3)
$\Delta\rho$	Density difference between the gas and liquid phases (kg/m^3)
ν	Kinematic viscosity (m^2/s)
σ	Surface tension (N-m)

Subscripts/Superscripts

*	Non-dimensional value
+	Non-dimensional value
∞	In an infinite medium

B	Value for bubbly flow
c	Critical value or continuous phase
cap	Cap bubble
f	Liquid phase
g	Gas phase
P	Value for cap-bubbly flow
r	Relative velocity between two phases
S	Value for slug flow
$sphere$	Spherical bubble

Operators

$\langle \rangle$	Area-averaged value
$\langle \langle \rangle \rangle$	Void-fraction-weighted mean value

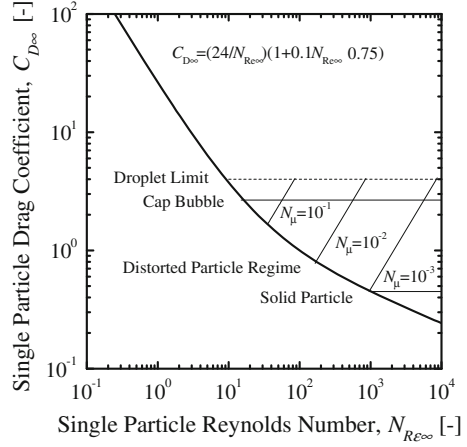
2.1 Introduction

Two-phase flows are essential in industries ranging from energy generation and petrochemical processing to pharmaceuticals. Two-phase flow systems have even been proposed for atmosphere scrubbing and waste disposal as part of the space program. The reason for this is that two-phase flows are extremely efficient for transferring mass or energy from one location to another and provide a simple method for generating large surface areas to allow chemical reactions between gases and water-soluble compounds.

Two-phase flows are also an essential component in nuclear energy systems. In Pressurized Water Reactors (PWR), the steam generator is a heat exchanger that creates the steam needed to run the turbine. The flow in this region is extremely complex and must be accurately predicted in order to properly predict the characteristics of the reactor during normal operation as well as transient scenarios. Likewise for Boiling Water Reactors (BWR), where steam is generated in the core itself. While this provides a negative reactor feedback, improving safety, it also means that two-phase flows are present in the reactor core. In BWRs which utilize natural circulation to establish liquid flow through the core, two-phase flow also exists in the large-diameter chimney section installed above the core to provide the hydrostatic head necessary for driving the natural circulation. The ability to predict two-phase flows in these rod bundle and large diameter channel regions is essential for predicting the safety performance and power characteristics of these types of reactors.

Two-phase flows are characterized by the existence of one or several interfaces and discontinuities in fluid properties at the interface. The interaction between the two phases causes the interfaces to change continuously, which may result in enhancement or reduction in the efficiency of transfer of energy or mass in the two-phase flows. The interactions between phases are closely associated with the drag forces.

Fig. 2.1 Single-particle drag coefficients [21, 22]



For a single particle moving in an infinite medium the drag coefficient, $C_{D\infty}$, depends on the particle Reynolds number and the viscosity number [21, 22] as shown in Fig. 2.1.

$$N_{Re\infty} = \frac{2r_d\rho_c|v_{r\infty}|}{\mu_c} \quad (2.1)$$

and

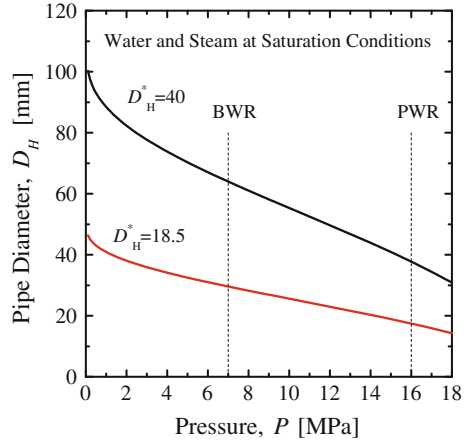
$$N_{\mu} = \frac{\mu_c}{\left(\rho_c\sigma\sqrt{\frac{\sigma}{g\Delta\rho}}\right)^{1/2}}. \quad (2.2)$$

where r_d , $v_{r\infty}$, μ_c , ρ_c , $\Delta\rho$, σ and g are the particle radius, the relative velocity of a single particle in an infinite medium, continuous phase viscosity, continuous phase density, density difference between the continuous and dispersed phases, surface tension and gravitational acceleration, respectively. Figure 2.1 shows that $C_{D\infty}$ greatly depends on the bubble shape (such as spherical, distorted or cap) and size.

Generally, bubbles are categorized into two groups based on these drag characteristics. Group 1 bubbles are defined as small spherical and ellipsoid (or distorted) bubbles. Group 2 bubbles are larger cap-shaped or irregularly shaped bubbles. These groups are determined based on the drag behavior of each bubble group. Thus this bubble classification provides a clear distinction between two types of bubbles with varying behavior, while bubbles within either group behave in relatively similar fashion to each other.

Materials are transported in industrial systems by tubes and pipes of varying sizes ranging from microchannels to large diameter pipes. The diameters of the tubes and pipes may limit the shape of the bubbles and restrict their ability to move within the channel. Thus the behavior of the bubbles is closely linked with the

Fig. 2.2 Pipe diameter change with pressure at water and steam saturation states



diameter of the tubes and pipes. Tubes and pipes with non-dimensional hydraulic diameter, defined as

$$D_H^* = \frac{D_H}{\sqrt{\sigma/g\Delta\rho}} \quad (2.3)$$

smaller than about 18.5 are considered ‘small diameter’ due to the presence of stable Taylor bubbles forming long gas slugs which occupy the entire cross section of the pipe. Pipes with non-dimensional diameter larger than 40 are considered large diameter pipes because Taylor bubbles occupying the entire pipe cross section can no longer be sustained. The intervening range, from non-dimensional diameter of 18.5–40, is a transition region between the two behaviors [37]. Pipe diameters corresponding to $D_H^* = 18.5$ and $D_H^* = 40$ of water and steam at saturation states are shown in Fig. 2.2 for different working pressures, with the operating pressures of Boiling Water Reactors (BWR) and Pressurized Water Reactors (PWR) highlighted.

Prediction of flows in large diameter pipes presents several unique challenges. Surface instability results in the breakup of large Taylor cap bubbles into smaller bubbles, and even these bubbles have highly distorted surfaces. This prevents the formation of stable slug bubbles [24]. The effects of this bubble breakup and surface instability on the flow are varied and complex, affecting all facets of the hydrodynamics of the flow. This makes the prediction of interfacial drag, interfacial structure, generation of turbulence, and other important phenomena difficult at best.

Many modeling efforts in two-phase flows begin with the drift-flux model [58]. Despite being less rigorous than the more detailed two-fluid model, the drift-flux model is extremely important for predicting two-phase flows because of its intrinsic simplicity and because the drift velocity and distribution parameter are directly related to the physical structure of the flow [22]. For this reason, the drift-flux model is a reasonable starting point for understanding and modeling the behavior of two-phase flows in large diameter pipes.

Due to the importance of large diameter pipes in two-phase systems and the challenge that predicting these flows poses, significant effort has already been made to develop the tools necessary to do so. Many researchers have performed hundreds of measurements and developed many drift-flux type models. Many of these models are empirical in nature and, while they are of limited use in predicting the wide range of two-phase flows in large diameter systems, provide valuable insight into the most important factors affecting the flow. Data has been collected in pipes with non-dimensional diameters as high as 220 and in adiabatic air-water and steam-water flows as well as boiling flows.

This chapter will focus on summarizing the most important characteristics of flows in large diameter pipes as well as the drift-flux models available. The characteristics of two-phase flows in large diameter pipes and their effects on the behavior and prediction of the flow will be discussed. Then the available drift flux models will be presented and the major properties, strengths and weaknesses of each will be discussed. Finally the data available for evaluating and analyzing the performance of the drift-flux models will be summarized and analyzed.

2.2 Characteristics of Large Diameter Pipes

In comparison with two-phase flows in small diameter pipes, flows in large diameter pipes are quite different. The differences are summarized as follows.

2.2.1 Surface Instability and Shearing off in Large Pipes

In small diameter channels, the channel walls restrict the growth of large bubbles, resulting in the formation of gas slugs. Because the diameter of these gas slugs is relatively small, the upper bubble surface remains stable and distortions in the bubble surface, which may be caused by turbulent fluctuations or by Kelvin–Helmholtz instability, are quickly damped. In large diameter pipes however, the diameter of the gas slugs can become very large. In this case, the shape of the upper bubble surface as not so restrained. Because of this when distortions to the bubble surface occur Taylor instabilities result in growth of the disturbance. This causes the bubble to break up, as shown in Fig. 2.3. Generally when discussing bubble breakup mechanisms, this mechanism is described as ‘surface instability’. The limited ability of bubbles to grow caused by this surface instability, resulting in the absence of large gas slugs, is the defining feature of large diameter channels.

The increased prevalence of bubble breakup due to surface instability results in some significant differences between flows in large and small diameter channels. First, in place of large gas slugs flows in large diameter channels are composed of either Taylor cap bubbles (at moderate void fractions) or unstable churn-turbulent bubbles (at high void fractions). This change affects the flow in a variety of ways.

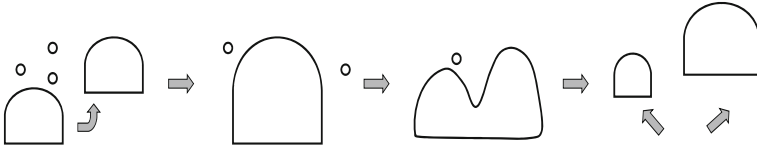
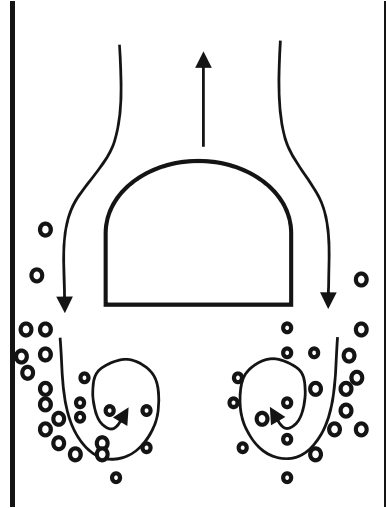


Fig. 2.3 Formation and breakup of cap bubbles larger than the maximum stable bubble size

Fig. 2.4 Enhanced turbulence and shearing-off of small bubbles at base of cap bubble



The increased number of Group 2 bubbles results in additional interfacial surface area in the flow. This can result in dramatically increased bubble-induced turbulence. Additionally, the presence of an increased number of cap bubbles results in many additional locations where small bubbles can be broken off from the base of larger bubbles, as shown in Fig. 2.4. This breakup mechanism, which is termed ‘shearing-off,’ may become the most significant source of small, dispersed bubbles in large diameter channels and result in a higher proportion of the void being composed of smaller Group 1 bubbles [9].

In spite of these differences, many of the other ways in which bubbles interact are very similar. In general, bubbles may coalesce by randomly bumping into each other within the flow. This type of coalescence, termed ‘random collision,’ occurs mainly with small bubbles and depends on the bubble concentration and size of small bubbles, as well as the strength of turbulence in the flow. Both the bubble concentration and size of Group 1 bubbles are not strongly dependent on the pipe size. Generally the turbulence characteristics of flows large diameter channels are quite different from those in small diameter channels, but this is generally accounted for using constitutive models. The second major type of bubble coalescence occurs when bubbles are caught up in the wake of a preceding bubble and run into each other. This is called ‘wake entrainment’ and is determined by the relative velocity of

the bubbles, their geometry, and the concentration of the bubbles. Again, this process is very similar in both large and small diameter channels, and the differences that do exist can be accounted for in the models for predicting relative velocity and concentration. The most significant cause of bubble breakup is the interaction of bubbles with turbulent eddies in the liquid. This mechanism is called ‘turbulent impact.’ Generally this mechanism is not very strong for Group 2 bubbles, but for Group 1 bubbles can be very important. This mechanism is dependent on the Group 1 bubble properties, which do not change much as the channel size increases, as well as the turbulence properties. Turbulence characteristics can be very different in large and small diameter channels, however this does not change the behavior of the mechanism and can be accounted for in the models for predicting turbulence. Thus for all of these types of bubble interactions, the actual physics is very similar in the various pipe sizes and the differences can be accounted for by using the correct models for predicting relative velocity, turbulence, and other important parameters [22].

2.2.2 Flow Regimes

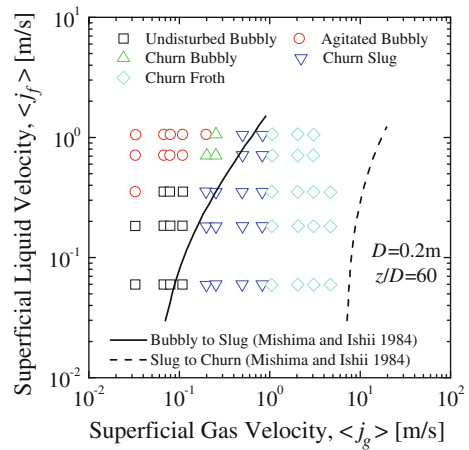
In small diameter pipes, the flow regimes typically noted include bubbly, slug, churn and annular flows [28]. As discussed in the previous section however, stable slug bubbles cannot exist in large diameter channels. This means that flow regime maps developed by various researchers for small diameter pipes cannot be applied to large diameter channels. This has been confirmed by observations from many sources [31, 38, 44, 59].

The first investigation which defined the flow regimes present in large diameter pipes in detail was that of [31], where flow regime characteristics were investigated in a pipe with diameter of 0.2 m. The flow regimes were classified as follows:

- Undisturbed bubbly: composed of very small numbers of small, spherical bubbles which do not show significant lateral movement.
- Agitated bubbly: composed of small, spherical bubbles which show some mixing but limited coalescence or breakup.
- Churn bubbly: composed of many small, spherical bubbles and characterized by very significant mixing and prevalent bubble breakup and coalescence.
- Churn slug: composed of many small, spherical bubbles as well as large but relatively stable cap-shaped bubbles.
- Churn-froth: composed of many small, spherical bubbles as well as unstable large bubbles that undergo frequent coalescence and breakup.

The results of their flow regime identification efforts can be seen in Fig. 2.5. These flow regimes classifications were determined by the visual observation of the researchers. Unfortunately, the flow conditions in this study were restricted to gas velocities no higher than 4.7 m/s. This means that the transition to annular flow was not able to be evaluated. Annular flow is characterized by a solid gas core surrounded by a liquid film. This gas core may also contain liquid droplets.

Fig. 2.5 Flow regimes observed by Ohnuki and Akimoto [31]



In view of the difficulties in visualizing the internal flow structure in a large diameter pipe, Smith [49] and Schlegel et al. [38] observed the flow regimes by means of electrical impedance void meters in pipes of various diameters. In these studies, the flow regimes were re-evaluated based on the transport properties of the bubbles. Bubbly flow includes the undisturbed bubbly, agitated bubbly and churn bubbly regimes defined by Ohnuki and Akimoto [31]. Cap-bubbly or cap-turbulent flow is identical to the churn-slug flow regime, while churn-turbulent flow is identical to the churn-froth flow regime. It is recommended that these flow regimes be used when discussing two-phase flows and in developing models, as these flow regimes were determined based on the transport properties of bubbles, namely bubble drag and relative velocity, and these transport properties determine the behavior of the bubbles and therefore the properties of two-phase flows. Shen et al. [43, 44] reported bubbly flow, churn flow and slug flow in a 0.2 m diameter pipe. In their case, slug flow is defined as intermittent upward-flowing large cap bubbles and is identical to cap-turbulent flow.

In small diameter pipes, the flow regime transition occur rather abruptly as the mechanisms which cause the change in flow regime occur under very specific conditions. For large diameter pipes however, all of these researchers have noted that the transitions between flow regimes are very gradual [45]. This is because in large diameter channels, Group 2 bubbles behave in a much more random fashion than for flows in small diameter channels, and the mechanisms which cause the transitions between flow regimes reflect this behavior.

2.2.3 Effect of Inlet Conditions on Flow

Another major characteristic of two-phase flows in large diameter pipes is the effect of the inlet condition on the flow pattern [13]. For small diameter channels the gas injection method has little effect on the flow, but in large diameter pipes the

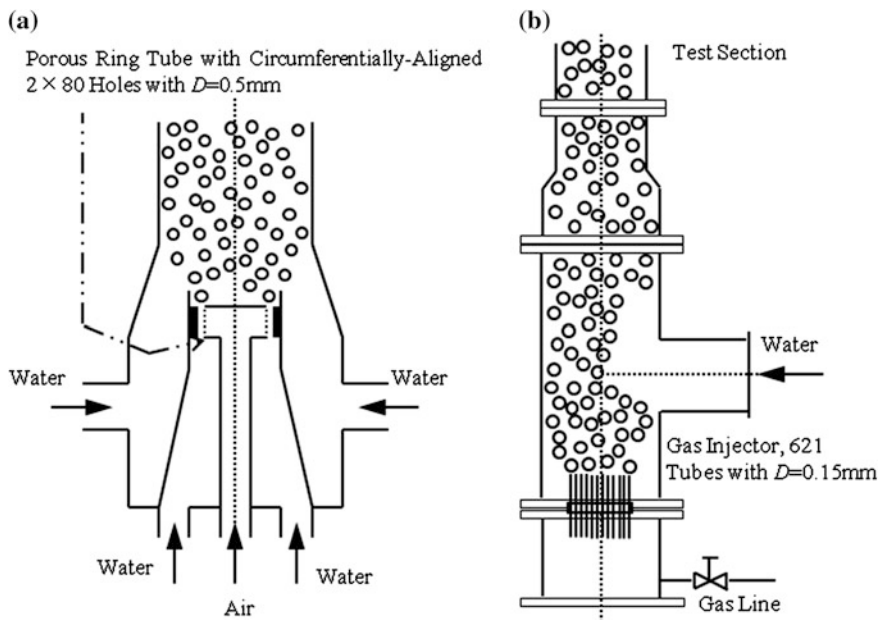


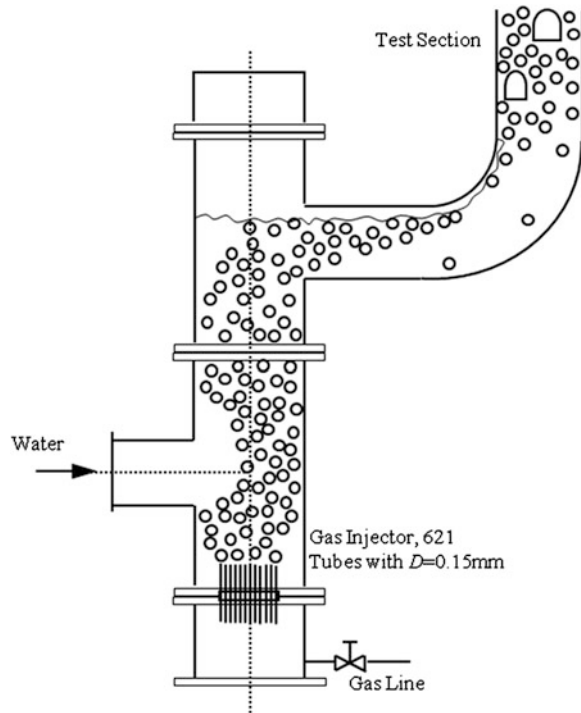
Fig. 2.6 Uniformly distributed bubble injection mixer. **a** Shear-cutting liquid and gas mixer. **b** Needle-injecting liquid and gas mixer

effect is pronounced for relatively low flow rates. At low void fraction conditions, almost uniformly distributed small, spherical bubbles are obtained by using the typical liquid and gas inlet mixers such as the shear-cutting liquid and gas mixer of Shen et al. [44] and the needle-injecting liquid and gas mixer of Hibiki and Ishii [13] shown in Fig. 2.6a, b, respectively. At identical gas and liquid injection rates, cap bubbles at the inlet are observed for bubble injection using a horizontal section in Fig. 2.7 [13] even though the condition is typically considered bubbly flow. Bubbles injected as large cap bubbles do not break up easily and are typically much less affected by turbulence because they are significantly larger than the turbulent eddy sizes which contain most of the turbulent kinetic energy. This means that cap bubbles can be found even at very low void fractions (<0.1) when the gas is introduced by using the bubble injection mixer with a horizontal section. This can significantly affect the flow under these conditions, with very different flow behavior at low void fractions under different injection conditions.

2.2.4 Relative Velocity in Large Diameter Pipes

Large-volume cap bubbles can be found in both small and large diameter channels. In the small diameter case, these appear as slug bubbles which occupy the entire pipe diameter. Generally, the slug bubble length is less significant when determining how

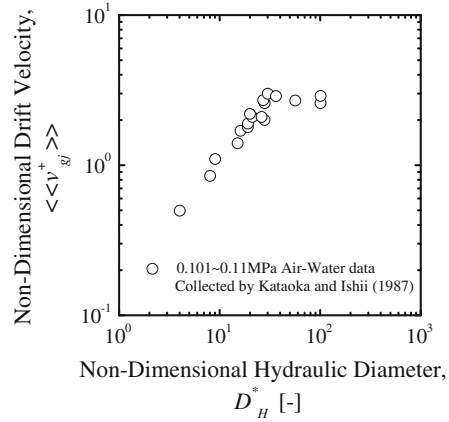
Fig. 2.7 Bubble injection mixer with elbow horizontal section



the bubble moves in a two-phase mixture and the dynamics of the flow around the nose and tail of the bubble almost entirely govern their motion. Thus the small diameter of the bubble nose and the effect of wall drag on the bubble limit the rise velocity of slug bubbles in the liquid. This effect can be seen for smaller pipe diameters in Fig. 2.8, which shows the drift velocity collected by Kataoka and Ishii [24] for several pipe diameters. As the pipe size increases the shape of the bubble nose changes and the effect of wall drag is reduced, leading to higher bubble relative velocity.

In large diameter pipes however, the figure shows that the drift velocity reaches a maximum and then remains constant. In larger channels the channel size is larger than the maximum stable bubble. When this occurs, the mechanics governing the relative velocity are no longer dependent on the channel size as the channel no longer restricts bubble motion. Additionally, the additional drag on large bubbles caused by the wall is no longer a factor because the Group 2 bubbles are generally no longer in contact with the wall. Because of this, the relative velocity is determined entirely by the bubble hydrodynamics without the effect of the wall and becomes independent of the pipe diameter [12, 43].

Fig. 2.8 Drift velocity change with pipe diameter



2.2.5 Profiles of Local Flow Parameters

Turbulence plays a vital role in determining the local void fraction and velocity profiles. In general, turbulence is generated in regions where a velocity gradient exists. For single-phase flows, this is in the region near the wall of the flow channel. In small diameter channels, this is very similar for two-phase flows as shown by Liu and Bankoff [26]. For large diameter channels, single-phase flows show very similar turbulence characteristics to flows in small diameter channels. On the other hand, while the existing data regarding turbulence characteristics in large diameter channels is admittedly limited, it indicates that two-phase flows are very different.

Based on the work of Serizawa and Kataoka [41], two-phase flows introduce bubble-induced turbulence as well as turbulence absorption at the interface. Absorption of turbulence at the interface is caused by distortions in the bubble surface. These distortions absorb energy because of surface tension, which results in a small reduction in turbulence. Bubble-induced turbulence is dependent on the relative velocity between the phases, drag force, and interfacial area concentration. For bubbly flows where only Group 1 bubbles are present this bubble-induced turbulence is similar in magnitude to the absorption of turbulence at the interface and can, under some conditions, actually result in reduction turbulence as compared to single-phase flows. Group 2 bubbles however have very different characteristics than Group 1 bubbles. While the relative velocity is similar, the drag force is much larger for Group 2 bubbles and the interfacial area concentration is much smaller. The net result of these two effects is that turbulent kinetic energy production from Group 2 bubbles is about an order of magnitude higher than that from Group 1 bubbles. Thus the bubble-induced turbulence created by Group 2 bubbles dominates in cap-bubbly and churn-turbulent flows for all types of flow channels.

This leads to the differences in turbulence characteristics for flows in small and large diameter channels. Generally, the Group 1 bubbles are similar in size and behavior regardless of channel size. However, based on the work of Ohnuki and Akimoto [32], the length scale at which turbulence is produced is very important. Turbulent eddies significantly larger than the bubble size tend to carry more energy and carry bubbles along, while those significantly smaller do not appreciably affect the bubbles. In large diameter channels, turbulence is produced at much larger scales than in smaller channels. This means that the turbulent kinetic energy and turbulent fluctuations may be higher in large channels even though the velocity gradient near the pipe wall may be smaller, resulting in significant changes to the bubble distribution and small changes in bubble size and behavior. As discussed in Sect. 2.2.1, large diameter pipes cannot sustain stable slug bubbles. In the cap-bubbly and churn-turbulent flow regimes, this leads to the existence of many smaller cap bubbles within the flow rather than one large gas slug. This drastically increases the interfacial area concentration of Group 2 bubbles, leading to a corresponding increase in the production of turbulent kinetic energy from these bubbles. Because of these two effects it is expected that flows in large diameter channels will have significantly stronger turbulent fluctuations than flows in small diameter pipes across nearly all flow conditions, but especially at relatively low liquid flow rates. This is illustrated in the data of Ohnuki and Akimoto [31].

These changes in the turbulence characteristics for large diameter channels result in very different flow behavior. In small diameter pipes, the lift force causes small spherical bubbles within a certain range of bubble diameter to concentrate near the pipe wall at low void fractions. This phenomenon is known as wall-peaking. The increased turbulence as well as larger scale of turbulence in large diameter channels results enhanced turbulent mixing of small, spherical bubbles. Because of this enhanced mixing as well as a reduction in lift force due to a smaller near-wall velocity gradient, the bubbles are distributed more evenly throughout the flow channel, resulting in elimination or reduction of the wall-peaking effect. This effect is shown in Fig. 2.9 using void profile data from experiments in 50.8 mm [15] and 200 mm [47] diameter channels. The figure shows the void fraction profile for very similar flow conditions and axial locations, with the only significant difference being the size of the channel. Even so, the void fraction profiles are quite different. Similar trends are shown in experiments performed by Smith et al. [51] in 101 and 152 mm diameter channels and by [39] in channels with diameter of 152 and 203 mm.

Large cap bubbles also have many other effects on the velocity profile in large channels. At lower flow rates, large bubbles tend to collect in higher-velocity regions within the pipe—namely the center of the pipe. Because of this, the velocity in this region may be enhanced at higher void fractions. This effect can be noted in the data of Ohnuki and Akimoto [30, 31] as well as the data of Smith et al. [51] Sun et al. [54], Shen et al. [59] and Schlegel et al. [39] in the gas velocity profiles at various flow conditions which show a center peak in the gas velocity profile which becomes more prominent as the void fraction increases.

Fig. 2.9 Wall-peaking effect comparison between small and large diameter pipes

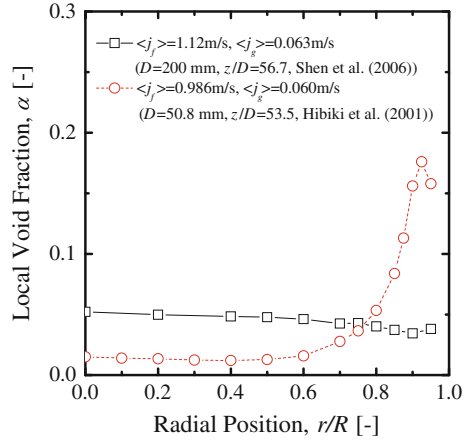
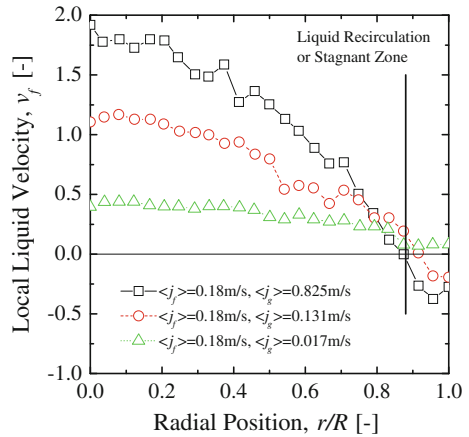


Fig. 2.10 Radial local liquid velocity in a large diameter pipe from Ohnuki et al. [33]



When the liquid velocity is very small, this effect can have some interesting consequences. As noted by Hills [16], when the liquid velocity is very small compared to the bubble rise velocity, the liquid is pulled along by the large bubbles in the center of the channel more quickly than the liquid is flowing near the channel wall. In this case, to maintain the total liquid volumetric flow rate, the liquid tends to move out of the pipe center toward the channel wall where it can be stagnant or even move against the total flow direction. This is illustrated by the data of Ohnuki et al. [33] in Fig. 2.10. The figure shows that at low gas velocities when cap bubbles do not exist, the liquid velocity is relatively flat. At higher gas velocities where larger bubbles begin to dominate, the velocity profile develops a strong center peak with negative local velocities near the channel wall. Based on the data from this and other studies, this effect tends to decrease as the liquid flow rate increases and for average liquid velocities higher than about 0.5 m/s this recirculation region near the channel wall is largely absent.

2.3 Flow Regime Transition Criteria

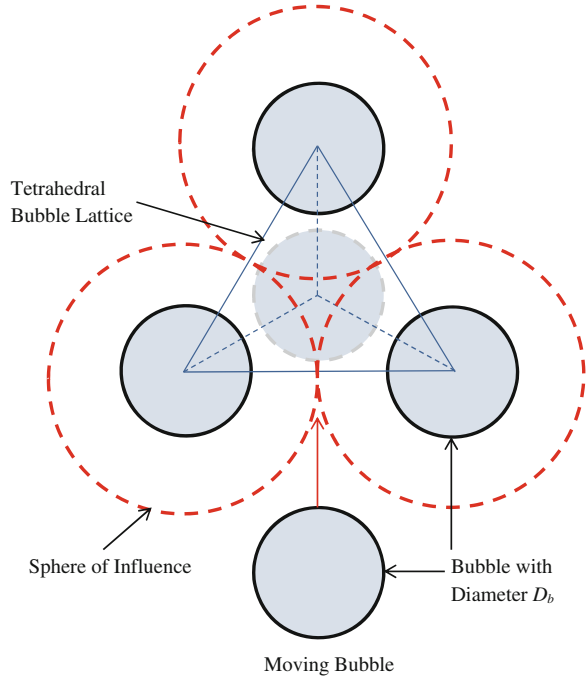
As many existing correlations are applicable only for certain flow patterns, accurate flow pattern maps which show the conditions for which certain flow patterns are present are key. For flows in small pipes, theoretically-based flow regime maps have been developed by Taitel et al. [55] and Mishima and Ishii [28], but the applicability of these models to large pipes has not been thoroughly investigated. Schlegel et al. [38] have developed a flow regime map for large diameter channels which has been partially validated using a large database of flow regime identification data.

Before defining flow regime transitions, it is first necessary to define the flow regimes. Based on experiment and as described in Sect. 2.2.2, there are four flow regimes with unique transport characteristics present in large diameter channels. The first is bubbly flow, which consists largely of small, spherical bubbles and occurs at lower void fraction. The second flow regime is cap-turbulent flow, which is characterized by large Taylor bubbles moving through the liquid phase. This results in significant agitation of the flow and causes the liquid and smaller bubbles to move in very chaotic fashion, although the cap bubbles are still relatively stable. The third flow regime is churn-turbulent flow, which is characterized by highly unstable large bubbles which coalesce and break up with extremely high frequency due to the proximity of large bubbles to each other and surface instability. This results in a very chaotic flow pattern with irregular bubble shapes. These three flow regimes approximately correspond to the agitated bubbly, churn slug, and churn froth flow regimes described by Ohnuki and Akimoto [31]. The final flow regime is annular/mist flow, which is characterized by a gas core containing liquid droplets and surrounded by a liquid film [38].

Both theoretical flow regime maps for small diameter channels [55, 28] considered packing of bubbles and a sudden increase in the coalescence rate as the mechanism for the transition from bubbly flow to slug flow. Taitel and Bornea [55] postulated a cubic bubble lattice while Mishima and Ishii [28] postulated a tetrahedral bubble lattice, shown in Fig. 2.11, resulting in predictions that the transition would occur at void fractions of 0.25 and 0.30, respectively. This mechanism is very likely the mechanism which drives the transition from bubbly to cap-turbulent flow in large diameter channels, and so it is expected that the transition boundary will be similar. In fact, experimental studies have shown that the transition region tends to lie between these two values of void fraction based on changes in the drift-flux parameters and on flow regime identification studies. Thus it is recommended that $\langle \alpha \rangle = 0.3$ be used as the flow regime transition criteria from bubbly to cap-turbulent flow in large diameter channels, but one must keep in mind that the transition begins at lower void fractions, about 0.25, and be cautious when applying models in this transition region [38].

In large diameter channels, Taylor bubbles occupying the entire diameter of the channel simply cannot exist for an extended length of time due to instability in the upper surface of such large bubbles. Therefore, stable slug bubbles cannot exist

Fig. 2.11 Maximum packing of spherical bubbles for transition to cap/slug flow



and the transition criteria from slug flow to churn-turbulent flow in small-diameter studies cannot be applied for the transition from cap-turbulent to churn-turbulent flow in large diameter channels. Instead, Schlegel et al. [38] proposed that the mechanism for the transition is the onset of rapid formation and disintegration of Taylor bubbles larger than the bubble size limit. This occurs when Taylor bubbles achieve the maximum packing criteria in similar fashion to bubbly flows. This process is illustrated for Taylor bubbles in Fig. 2.12. The ratio of Taylor bubble volume to the volume of the tetrahedron is approximately 0.3. Thus this should be the cap-bubble fraction at the transition to churn-turbulent flow. If the spherical bubble fraction of the remaining liquid volume is assumed to be 0.30 as in the bubbly flow analysis, then the total void fraction at the transition to churn-turbulent flow is [38]

$$\langle \alpha \rangle = \langle \alpha_{Cap} \rangle + \langle \alpha_{Sphere} \rangle (1 - \langle \alpha_{Cap} \rangle) = 0.51 \quad (2.4)$$

The transition to annular flow which is applicable to large diameter pipes is the entrainment condition postulated by Mishima and Ishii [28]. Interestingly, this is the same mechanism hypothesized by Taitel and Bornea [55], and the resulting transition boundaries for the two studies have almost identical formulations. In fact, the only difference is that in Mishima and Ishii's formulation, the constant 3.1 in Taitel and Bornea's model is replaced by the viscosity number, which scales the effect of the liquid viscosity on the flow and is equal to approximately 3.38 for air-

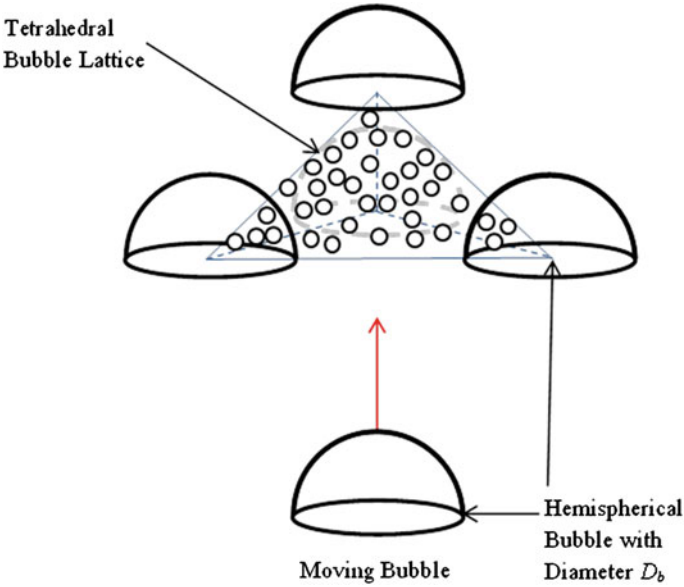


Fig. 2.12 Maximum packing of cap bubbles for transition to churn-turbulent flow

Table 2.1 Flow regime transition criteria for large diameter pipes

Flow regime transition	Criteria
From bubble to cap-turbulent flow	$\langle \alpha \rangle = 0.30$
From cap-turbulent to churn-turbulent flow	$\langle \alpha \rangle = 0.51$
From churn-turbulent flow to annular/mist flow	$\langle j_g \rangle = \left(\frac{\sigma g \Delta \rho}{\rho_g^2} \right)^{1/4} N_{\mu f}^{-0.2}$

water flows at atmospheric conditions. For this reason, the transition boundary for droplet entrainment proposed by Mishima and Ishii [28] is suggested for use in large diameter pipes to ensure generality [38].

The flow regime transition criteria proposed by Schlegel et al. [38] are summarized in Table 2.1. Based on these proposed transition criteria, the flow regime map is shown in Fig. 2.13 including the flow regime identification results of Schlegel et al. [38] using impedance void meter measurements and a self-organized neural network to determine the flow regime of each condition. As the figure shows, the proposed transition from bubbly to cap-turbulent flow agrees almost exactly with the data as does the proposed transition from cap-turbulent to churn-turbulent flow. Additionally, Figs. 2.14 and 2.15 show the comparison of these flow regime transition criteria with the observations of Ohnuki and Akimoto [32] and Smith [49]. Ohnuki and Akimoto [32] used visual observation and categorization to determine the flow regime. In Fig. 2.14, the undisturbed bubbly and agitated bubbly flow conditions have been combined as ‘Bubbly’ and the observations at the test

Fig. 2.13 Flow regime mapping and identification of Schlegel et al. [38]

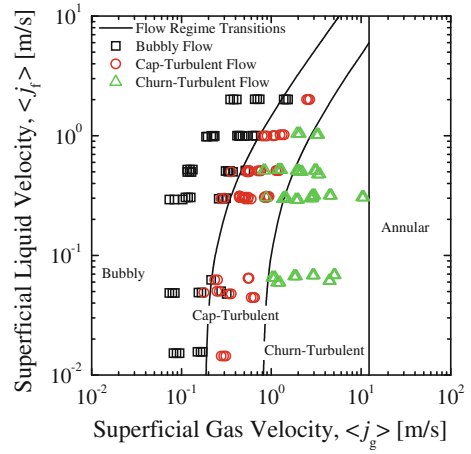


Fig. 2.14 Flow regime identification results of Ohnuki and Akimoto [31]

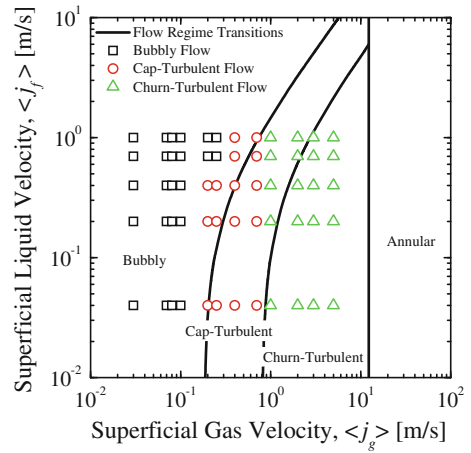
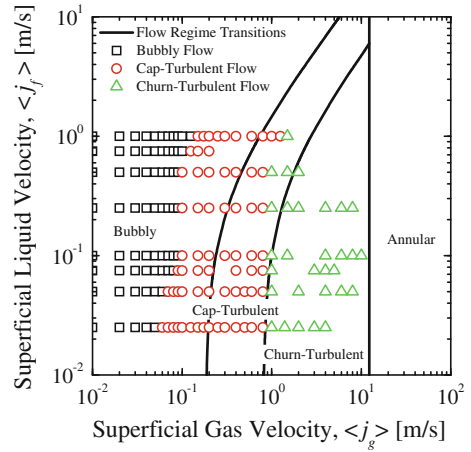


Fig. 2.15 Flow regime identification results of Smith [49]



section exit are used. Smith [49] used a neural network to categorize the flow regime based on data collected by impedance void meters to generate the results shown in Fig. 2.15. It should be noted that this result shows that cap-turbulent flow begins at significantly lower void fraction than the other studies. This may be caused by the injection method, which likely resulted in cap-bubbly injection when the gas flow rate reached values higher than 0.1 m/s. Otherwise, both experimental studies confirm the applicability of the flow regime transitions developed by Schlegel et al. [38] for large diameter channels.

2.4 Drift-Flux Models

2.4.1 Formulation of the Drift-Flux Model

The drift-flux model is one of the most practical and accurate models for two-phase-flow analysis. The model takes into account the effects of non-uniform velocity and void fraction profiles as well as the effect of the local relative velocity between phases. It has been utilized to solve many engineering problems involving two-phase-flow dynamics. In particular, its application to forced convection systems has been quite successful. The one-dimensional drift-flux model was derived by averaging the local drift velocity over the channel cross-section [58]. The model is given as

$$\frac{\langle j_g \rangle}{\langle \alpha \rangle} = \langle \langle v_g \rangle \rangle = C_0 \langle j \rangle + \langle \langle v_{gj} \rangle \rangle \quad (2.5)$$

where $\langle \rangle$ denotes the area average of a quantity, F , over the cross-sectional area (A) of the flow path, which is defined mathematically as

$$\langle F \rangle = \frac{1}{A} \int_A F dA \quad (2.6)$$

and α and j are the void fraction and the mixture volumetric flux, respectively. $\langle \langle v_g \rangle \rangle$ is the void-fraction-weighted mean gas velocity. The distribution parameter, C_0 , and the void-fraction-weighted mean drift velocity, $\langle \langle v_{gj} \rangle \rangle$, are respectively defined by the following equations,

$$C_0 = \frac{\langle \alpha j \rangle}{\langle \alpha \rangle \langle j \rangle} \quad (2.7)$$

$$\langle \langle v_{gj} \rangle \rangle = \frac{\langle v_{gj} \alpha \rangle}{\langle \alpha \rangle} \quad (2.8)$$

where v_{gj} is the local drift velocity of gas phase defined as

$$v_{gj} = v_g - j = (1 - \alpha)(v_g - v_f) \quad (2.9)$$

The v_g and v_f in the above equation are the gas and liquid velocities, respectively.

The void-fraction-weighted mean gas velocity, $\langle\langle v_g \rangle\rangle$, and the cross-sectional mean mixture volumetric flux, $\langle j \rangle$, are easily obtainable parameters in experiments. Therefore, Eq. (2.5) suggests a plot of $\langle\langle v_g \rangle\rangle$ versus $\langle j \rangle$. An important characteristic of such a plot is that, for two-phase flow regimes with fully developed void and velocity profiles, the data points cluster around a straight line. The value of the distribution parameter, C_o , can be obtained indirectly from the slope of the line, whereas the intercept of this line with the void-fraction-weighted mean gas velocity axis can be interpreted as the void-fraction-weighted mean drift velocity, $\langle\langle v_{gj} \rangle\rangle$. As the recent development of local sensor techniques [25–27, 40, 46, 61, 62] enables the measurement of the local flow parameters in bubbly flow, including void fraction as well as gas and liquid velocities, the values of C_o and $\langle\langle v_{gj} \rangle\rangle$ in bubbly flow can be determined directly by Eqs. (2.7) and (2.8) from experimental measurement of the local flow parameters.

It should be mentioned here that sometimes the following non-dimensional parameters are introduced to nondimensionalize the drift-flux model:

$$\langle j_g^+ \rangle = \frac{\langle j_g \rangle}{\left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4}} \quad (2.10)$$

$$\langle j_f^+ \rangle = \frac{\langle j_f \rangle}{\left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4}} \quad (2.11)$$

$$\langle j^+ \rangle = \frac{\langle j \rangle}{\left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4}} \quad (2.12)$$

$$\langle\langle v_g^+ \rangle\rangle = \frac{\langle\langle v_g \rangle\rangle}{\left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4}} \quad (2.13)$$

$$\langle\langle v_{gj}^+ \rangle\rangle = \frac{\langle\langle v_{gj} \rangle\rangle}{\left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4}} \quad (2.14)$$

The non-dimensionalized drift-flux model is expressed as

$$\frac{\langle j_g^+ \rangle}{\langle \alpha \rangle} = \langle \langle v_g^+ \rangle \rangle = C_0 \langle j^+ \rangle + \langle \langle v_{gj}^+ \rangle \rangle. \quad (2.15)$$

2.4.2 Summary of Drift-Flux Correlations for Large Diameter Pipes

Several important studies regarding the development of drift-flux correlations for large diameter pipes are summarized in this section. The details for each model can be found in Table 2.2.

Hills [16] performed tests with a large pipe with an inner diameter of 0.150 m and height of 10.5 m for gas superficial velocities of 0.070–3.5 m/s and liquid superficial velocities of 0–2.7 m/s. Hills developed a drift flux type correlations based on his own experimental data. However, the effect of physical properties on the distribution parameter and the drift velocity are not included in this correlation. Therefore, Hills's correlation may not be applicable to other fluid systems such as high pressure steam-water flow conditions.

Ishii's model [20] was developed as a comprehensive model for small-diameter channels and represents the starting point for many later models. An expression for the distribution parameter was developed semi-empirically, and models for the drift velocity in each flow regime were developed that account for dependence on void fraction, fluid properties and pipe diameter. Unfortunately, as the presence of slug bubbles was assumed and the effect of inlet conditions were not accounted for in the models for bubbly and slug flow, this model may not be completely applicable to large diameter pipe flows.

Shipley [48] conducted experiments with a large pipe of an inner diameter of 0.457 m and height of 5.64 m. A drift flux correlation was developed based on this data. It is noted that the second term in the right hand side of his correlation, corresponding to the drift velocity, can become infinitely large for a very large diameter pipe. This seems not to be physically realizable.

Clark and Flemmer [5] performed tests with a large pipe of an inner diameter of 0.10 m. Mixture volumetric fluxes are ranged from 0.7 to 2.7 m/s. A drift flux correlation was developed based on their own test data. However, their correlation did not take into account the effect of physical properties on the distribution parameter. Consequently, Clark and Flemmer's correlation may not be applicable to other fluid systems such as high pressure steam-water flow conditions. Clark and Flemmer also developed another modified drift-flux type correlation in 1986. It should be noted here that Clark and Flemmer [6] did not consider the effect of physical properties on the distribution parameter and the drift velocity in their

Table 2.2 Drift-flux models for large diameter channels

Researchers	Distribution parameter, C_0 (-)	Drift velocity, $\langle\langle v_{gj} \rangle\rangle$ (m/s)	Applicable range
Ishii [20]	$1.2 - 0.2\sqrt{\rho_g/\rho_f}$	$\sqrt{2}\left(\frac{\sigma g \Delta \rho}{\rho_f^2}\right)^{1/4} (1 - \langle\alpha\rangle)^{1.75}$	Bubbly
	$1.2 - 0.2\sqrt{\rho_g/\rho_f}$	$0.35\sqrt{g D_m \Delta \rho / \rho_f}$	Slug
	$1.2 - 0.2\sqrt{\rho_g/\rho_f}$	$\sqrt{2}\left(\frac{\sigma g \Delta \rho}{\rho_f^2}\right)^{1/4}$	Churn
Hills [16]	$\bar{V}_{gj} \equiv (C_0 - 1)\langle j \rangle + \langle\langle v_{gj} \rangle\rangle = \frac{1 - \langle\alpha\rangle}{\langle\alpha\rangle + 4\sqrt{\rho_g/\rho_f}} \left[\langle j \rangle + \sqrt{\frac{\Delta \rho D(1 - \langle\alpha\rangle)}{0.015 \rho_f}} \right]$		Annular
	$1.35/\langle j \rangle^{0.07}$	0.24	
	1.0	$(1 - \langle\alpha\rangle)(0.24 + 4.0\langle\alpha\rangle^{1.72})$	$\langle j_f \rangle > 0.3$
Shipley [48]	1.2	$\left\{ 0.24 + 0.35 \left(\frac{j_k}{j} \right)^2 \sqrt{g D \langle \alpha \rangle} \right\}$	$\langle j_f \rangle < 0.3$
Clark and Flemmer [5]	$0.934(1 + 1.42\langle\alpha\rangle)$	$1.53 \left(\frac{\sigma g}{\rho_f} \right)^{1/4}$	
Clark and Flemmer [6]	$(1.95\langle j_g \rangle + 0.93\langle j_f \rangle) / \langle j \rangle$	0.25	
Murase et al. [29]	$\left\{ 1.2 - 0.2\sqrt{\rho_g/\rho_f} \right\} \{ 1 - \exp(-18\langle\alpha\rangle) \}$	$\langle\alpha_c\rangle = 0.588 - 1.817\sqrt{\frac{\rho_k}{\rho_f}} + 2\left(\sqrt{\frac{\rho_k}{\rho_f}}\right) - 3.34\left(\sqrt{\frac{\rho_k}{\rho_f}}\right)^{3/2}$	$\langle\alpha\rangle < \langle\alpha_c\rangle$
	$1.0 + 0.2\left(1 - \sqrt{\rho_g/\rho_f}\right) \left\{ 1 - \left(\frac{\langle\alpha\rangle - \langle\alpha_c\rangle}{1 - \langle\alpha_c\rangle}\right)^2 \right\}$		$\langle\alpha\rangle > \langle\alpha_c\rangle$
		$7.1(D_H^*)^{1/8} \left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4} \left(\frac{1}{\langle\alpha\rangle} - C_0 \right) \langle\alpha\rangle^{1.6}$	$\langle\alpha\rangle < 0.33$
		$16.4(D_H^*)^{1/8} \left(\frac{\sigma g \Delta \rho}{\rho_f^2} \right)^{1/4} \left(\frac{1}{\langle\alpha\rangle} - C_0 \right) \langle\alpha\rangle \left\{ -\ln\left(1 - \langle\alpha\rangle^{1.45}\right) \right\}$	$\langle\alpha\rangle > 0.33$

(continued)

Table 2.2 (continued)

Researchers	Distribution parameter, C_0 (-)	Drift velocity, $\langle\langle v_{gj} \rangle\rangle$ (m/s)	Applicable range
Ishii and Kocamustafaoğullari [23]	Not given	$0.54 \sqrt{\frac{g D_H \Delta \rho}{\rho_f}}$	$D_H^* \leq 30$
	Not given	$3.0 \left(\frac{e g \Delta \rho}{\rho_f^2} \right)^{1/4}$	$D_H^* \geq 30$
Hirao et al. [17]	$1.2 - 0.2 \sqrt{\rho_g / \rho_f}$	$0.52 \sqrt{\frac{g D_H \Delta \rho}{\rho_f}}$	
Kataoka and Ishii [24]	$1.2 - 0.2 \sqrt{\rho_g / \rho_f}$	$V_{gj}^+ = 0.0019 (D_H^*)^{0.809} \left(\frac{\rho_g}{\rho_f} \right)^{-0.157} N_{Hf}^{-0.562}$	$N_{Hf} \leq 2.25 \times 10^{-3}$
			$D_H^* \leq 30$
			$N_{Hf} \leq 2.25 \times 10^{-3}$
			$D_H^* \geq 30$
			$N_{Hf} > 2.25 \times 10^{-3}$
Hibiki and Ishii [14]	$1.2 - 0.2 \sqrt{\rho_g / \rho_f}$	$V_{gj}^+ = 0.030 \left(\frac{\rho_g}{\rho_f} \right)^{-0.157} N_{Hf}^{-0.562}$	
		$V_{gj}^+ = 0.92 \left(\frac{\rho_g}{\rho_f} \right)^{-0.157}$	
	$\exp \left\{ 0.475 \left(\frac{j_g^+}{j^+} \right)^{1.69} \right\} \left(1 - \sqrt{\frac{\rho_g}{\rho_f}} \right) + \sqrt{\frac{\rho_g}{\rho_f}}$	$V_{gj}^+ = V_{gi,B}^+ \exp \left(-1.39 \langle j_g^+ \rangle \right)$	Bubbly flow
		$+ V_{gi,P}^+ \left\{ 1 - \exp \left(-1.39 \langle j_g^+ \rangle \right) \right\}$	$\langle \alpha \rangle < 0.3$
			$0 \leq \frac{\langle j_g^+ \rangle}{j^+} \leq 0.9$
	$\left\{ -2.88 \left(\frac{j_g^+}{j^+} \right) + 4.08 \right\} \left(1 - \sqrt{\frac{\rho_g}{\rho_f}} \right) + \sqrt{\frac{\rho_g}{\rho_f}}$	$V_{gj}^+ = V_{gi,B}^+ \exp \left(-1.39 \langle j_g^+ \rangle \right)$	Bubbly flow
		$+ V_{gi,P}^+ \left\{ 1 - \exp \left(-1.39 \langle j_g^+ \rangle \right) \right\}$	$\langle \alpha \rangle < 0.3$
	$1.2 \exp \left\{ 0.110 (j^+)^{2.22} \right\} \left(1 - \sqrt{\frac{\rho_g}{\rho_f}} \right) + \sqrt{\frac{\rho_g}{\rho_f}}$	$V_{gj}^+ = V_{gi,P}^+$	Cap bubbly flow
			$\langle \alpha \rangle < 0.3$
	$[0.6 \exp \{ -1.2 (\langle j^+ \rangle - 1.8) \} + 1.2]$		$0 \leq \langle j^+ \rangle \leq 1.8$
	$\left(1 - \sqrt{\frac{\rho_g}{\rho_f}} \right) + \sqrt{\frac{\rho_g}{\rho_f}}$		Cap bubbly flow
			$\langle \alpha \rangle < 0.3$
			$\langle j^+ \rangle \geq 1.8$

(continued)

Table 2.2 (continued)

Researchers	Distribution parameter, C_0 (-)	Drift velocity, $\langle\langle v_{gj} \rangle\rangle$ (m/s)	Applicable range
Cunningham and Yeh [7]	$\langle\alpha\rangle = 0.925 \left(\frac{\rho_k}{\rho_f}\right)^{0.239} \left[\frac{\langle j_s \rangle}{1.53(g\sigma/\rho_f)^{1/4}} \right]^a \left(\frac{\langle j_t \rangle}{\langle j \rangle}\right)^{0.6}$ where $a = 0.67$ if $\frac{\langle j_s \rangle}{1.53(g\sigma/\rho_f)^{1/4}} < 1$ $a = 0.47$ if $\frac{\langle j_s \rangle}{1.53(g\sigma/\rho_f)^{1/4}} \geq 1$ $L/(K_0 + (1 - K_0)\langle\alpha\rangle')$ where $K_0 = B_1 + (1 - B_1)\left(\frac{\rho_k}{\rho_f}\right)^{1/4}$ $L = \min\left[1.15\langle\alpha\rangle^{0.45}, 1.0\right]$ for air-water $r = \frac{1.0 + 1.57 \rho_k / \rho_f}{1 - B_1}$ B_1 is a variable depending on Re. $Re = \begin{cases} Re_g & Re_g > Re_f \text{ or } Re_g < 0 \\ Re_f & Re_g \leq Re_f \end{cases}$	$V_{gf}^0 C_g$ where $C_g = (1 - \langle\alpha\rangle)^{B_1}$ $V_{gf}^0 = 1.41 \left(\frac{\sigma_g \Delta p}{\rho_f^2}\right)^{1/4} C_2 C_3 C_4$ C_2, C_3, C_4 are variables depending on ρ_f, $\rho_g, D_H, Re_f = \frac{\langle j_t \rangle D_H \rho_f}{\mu_f}, Re_g = \frac{\langle j_t \rangle D_H \rho_g}{\mu_g}.$ $\frac{1}{2} \left(\langle\langle v_{g \rightarrow s} \rangle\rangle + \langle\langle v_{g \rightarrow B} \rangle\rangle \right) + \frac{1}{\pi} \left(\langle\langle v_{g \rightarrow s} \rangle\rangle - \langle\langle v_{g \rightarrow B} \rangle\rangle \right) \times \arctg \left\{ A \times \left(\frac{\langle j_s \rangle}{\langle j_t \rangle + \langle j_s \rangle} - B \right) \right\}$ where $\langle\langle v_{g \rightarrow s} \rangle\rangle = 0.52 \sqrt{\frac{D_g(\rho_f - \rho_g)}{\rho_f}}$ $\langle\langle v_{g \rightarrow B} \rangle\rangle = \sqrt{2} \left(\frac{g\sigma(\rho_f - \rho_g)}{\rho_f} \right)^{1/4}$ $A = 6.72 \langle j_f^+ \rangle^2 - 34.00 \langle j_f^+ \rangle + 69.94$ $B = 0.0851 \langle j_f^+ \rangle^2 - 0.414 \langle j_f^+ \rangle + 0.955$	
Shen et al. [45]	1		

correlation. Thus, the applicability of Clark and Flemmer's correlations to other fluid systems such as high pressure steam-water flow is still questionable.

Ishii and Kocamustafaogullari [23] developed a theoretical correlation for the drift velocity for cap bubble flow inside a large diameter channel. Their correlation suggests that the drift velocity increases with the channel diameter and reaches a constant value depending on physical properties at $D_H^* = 30$, corresponding to $D_H = 0.09$ m for air-water systems at atmospheric pressure. The correlation was derived under the assumption that the surface of the cap bubble was smooth. In real two-phase flows, large bubbles can be highly deformed due to natural turbulences in two-phase flow. They did not give any correlation for distribution parameter, however Ishii's correlation for the distribution parameter [20] is usually used with this correlation for the drift velocity.

Hirao et al. [17] conducted experiments with a large pipe of 0.102 m diameter. Their experimental flow conditions included steam-water two-phase flows with liquid and gas volumetric flux below 1 and 4 m/s, respectively. They also proposed a correlation for the drift velocity. It is noted that drift velocity obtained by their correlation can become infinitely large for a very large diameter pipe. This seems not to be physically realizable as well.

Murase et al. [29] performed steam-water two-phase experiments. The test section has four circular channels with an inner diameter 0.0405 and 1.055 m height. A single heater rod is installed inside each flow channel for generating steam flow. The correlations for distribution parameter and drift velocity were developed based on their experimental data.

Kataoka and Ishii [24] developed a drift velocity correlation for pool void fraction based on various experimental conditions. Their correlation shows that the drift velocity is dependent upon vessel diameter, system pressure, gas flux and fluid properties.

Chexal et al. [4] developed a generalized drift-flux correlation for air-water, steam-water, and refrigerant two-phase flows. It depends on pressure, temperature, hydraulic diameter and flow conditions of the two-phase flow but it is independent of flow regime. The correlation covers vertical (up and down), horizontal and inclined cocurrent flows and vertical countercurrent flows. The empirical correlation is usually called EPRI drift-flux model in RELAP 5. The drift-flux correlations for vertical air-water concurrent flows are shown in Table 2.2. The model is entirely empirical, and thus may have some inaccuracies when trying to predict flows in two-phase systems not included in the original benchmarking data set.

Hibiki and Ishii [14] found that the flow regime at the test section inlet affects the liquid circulation pattern resulting in inlet-flow-regime dependent distribution parameter and drift velocity. Two types of drift flux correlations based on different inlet flow regimes, namely bubbly and cap bubbly flows, were proposed. This drift-flux model includes the physical properties of the two phases and relates the distribution parameter and drift velocity to the local liquid and gas flow rates. It should be mentioned that the drift flux correlations developed by Kataoka and Ishii [24] were adopted for cap bubbly flow with void fraction greater than 0.3 in their

drift flux model. In Table 2.2, $V_{gj,P}$ is the pool-boiling drift velocity developed by Kataoka and Ishii [24], while $V_{gj,B}$ is the bubbly flow drift velocity given by Ishii [20].

Shen et al. [45] found an N -shaped void fraction development (increase-decrease-increase) of the void fraction in the flow direction as the two-phase flow develops from bubbly flow from to slug (large cap) flow at low liquid flow rates and developed a drift flux model for the two-phase flow ranging from bubbly flow from to slug flow in a vertical large diameter pipe to predict the N -shaped changing void fraction.

Other than the various drift-flux models, Cunningham and Yeh [7] developed a void fraction prediction model for rod bundle systems under pool conditions based on data from a simulated small-break LOCA (Loss-Of-Coolant Accident) experiment in a PWR fuel bundle assembly (480 heated rods). The model is a practical empirical correlation, but without the general form of the drift-flux model using a distribution parameter and a drift velocity.

2.5 Void Fraction Data in Large Diameter Pipes

Void fraction prediction in large diameter channels is essential for many applications. Because of this, many studies have been performed to measure the void fraction in large diameter pipes. These studies have been listed in Table 2.3 and are summarized here.

Among the first of these tests were those performed by Bailly et al. [1] who made void fraction measurements in pipes ranging from 0.152 to 0.610 m in diameter using an air-water system. This experiment was able to provide valuable data on the bubbly to cap-turbulent transition region, but was limited to void fractions smaller than 0.4 and to pool conditions. This means that the data does not provide any information on the behavior of churn-turbulent flows or flows with higher liquid fluxes.

Later several high-pressure steam-water flow experiments were performed to study the behavior of nuclear reactor systems during accident scenarios [3, 53, 56]. These studies provide important data for higher void fraction flows, but again are limited to pool conditions.

Several further experiments were performed by Hills [16] in a 0.152 m diameter pipe, but void fractions were limited to 0.55. Important data regarding low void fraction behavior at varied liquid velocities was obtained, but data at high void fractions that would allow evaluation of the churn-turbulent flow regime was not. Hall et al. [10] performed steam-water experiments in a 0.171 m diameter pipe for low liquid velocities and void fractions of up to 0.9. This data is valuable for evaluating the transition to annular flow, however very few data points were collected in this region. Hirao et al. [17] measured the void fraction in steam-water two-phase flow using a large scale apparatus with 0.1023 m in diameter. The data were obtained in a large range of flow conditions, but the number of data points is

Table 2.3 Void fraction data sets for large diameter channels

Researchers	Pipe diameter, D (m)	Superficial liquid velocity, $\langle j_f \rangle$ (m/s)	Superficial gas velocity, $\langle j_g \rangle$ (m/s)	Void fraction, $\langle \alpha \rangle$ (-)	Pressure, P (MPa)	Fluid system	Number of data points
Baily et al. [1]	0.152, 0.304, 0.610	0	up to 0.53	up to 0.4	~0.1	Air-Water	47
Wilson et al. [56]	0.101, 0.480	0	up to 1.9	up to 0.85	1.0–4.1	Steam-Water	115
Carrier [3]	0.456	0	up to 1.2	up to 0.82	4.1–13.8	Steam-Water	53
Styrkovich and Kutateladze [53]	0.238	0	up to 1.6	up to 0.7	0.61–9.3	Steam-Water	57
Hills [16]	0.150	up to 2.6	up to 3.2	up to 0.55	~0.1	Air-Water	394
Hall et al. [10]	0.171	up to 0.2	up to 0.75	up to 0.9	4.4	Steam-Water	90
Hashemi et al. [11]	0.305	up to 0.06	up to 1.2	up to 0.94	~0.1	Air-Water	30
Ohnuki and Akimoto [30]	0.480	up to 0.2	up to 0.9	up to 0.5	~0.1	Air-Water	75
Ohnuki and Akimoto [31]	0.200	up to 1	up to 4.7	up to 0.55	~0.1	Air-Water	58
Hirao et al. [17]	0.1023	up to 1.0	up to 4.0	up to 0.7	~1.5	Steam-Water	25
Hibiki and Ishii [13]	0.101	up to 0.4	up to 0.3	up to 0.3	~0.1	Nitrogen-Water	467
Yoneda et al. [57]	0.152	up to 0.6	up to 0.25	up to 0.35	~0.5	Steam-Water	12
Inoue [19]	0.200	up to 0.3	up to 0.35	up to 0.47	~0.1	Air-Water	84
Smith [50]	0.101, 0.152	up to 2	up to 8	up to 0.71	~0.1	Air-Water	106
Sawant et al. [36]	0.152	up to 2	up to 5.2	up to 0.83	~0.1	Air-Water	248
Shawkat et al. [42]	0.200	up to 0.7	up to 0.18	up to 0.15	~0.1	Air-Water	34
Schlegel et al. [37]	0.152, 0.203	up to 1	up to 8	up to 0.8	0.180–0.280	Air-Water	655
Prasser [60]	0.194	1 m/s	up to 0.53	up to 0.3	6.5	Steam-Water	7
Omebere-Iyari et al. [34]	0.194	up to 0.65	up to 0.95	up to 0.58	~4.64	Steam-Water	10
Shen et al. [45]	0.200	up to 0.311	up to 0.510	up to 0.39	~0.1	Air-Water	128

small. Hashemi et al. [11] performed experiments in a 0.305 m diameter facility with void fractions as high as 0.94, but again only very few data points were collected at such high void fractions.

A series of experiments investigating flow development in large diameter pipe systems were performed by Ohnuki and Akimoto [30, 31]. These experiments provided valuable insight into flow structure, but were again limited to void fractions less than 0.5. Yoneda et al. [57] performed some experiments using steam-water flows in pipes with diameters of up to 0.152 m, however few tests were reported. Also Inoue [19] performed air-water tests in a facility with 0.2 m diameter and for void fractions up to 0.47.

In relation to nuclear reactor accident and safety studies, Hibiki and Ishii [13] performed experiments focusing on bubbly flow behavior in a 0.101 m diameter pipe to understand the two-phase natural circulation and flow termination during a small break loss of coolant accident in light water reactors. One of the important aspects in their experiments was that inlet flow conditions were varied by inserting or removing a horizontal header section between the gas injector and the hot leg in order to investigate the effect of the gas phase inlet conditions on the flow regimes and flow interruption.

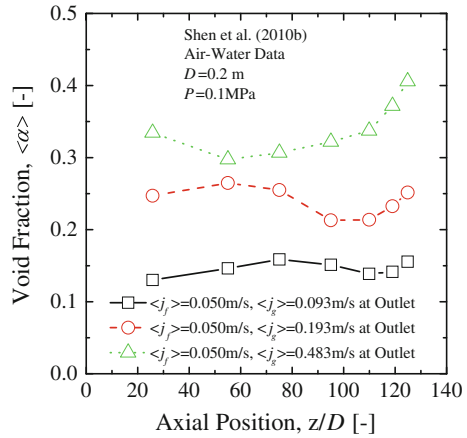
Sawant et al. [36] performed experiments in a 0.152 m diameter pipe with void fractions of up to 0.83, providing valuable data on flow pattern and flow behavior in the cap-turbulent and churn-turbulent regions. Shawkat et al. [42] performed very low void fraction experiments in a 0.2 m diameter pipe to evaluate the void profile, gas and liquid velocity profiles. Schlegel et al. [37] performed extensive experiments in pipes with 0.152 and 0.203 m diameter with void fractions as high as 0.8 to provide valuable data on the transition from cap-turbulent to churn-turbulent flows.

Omebere-Iyari et al. [34] carried out measurements of steam/water two-phase flow by using a wire mesh sensor in the 0.194 m diameter vertical pipe at the pressure of 4.64 MPa. Their data were taken in churn turbulent flow with void fraction as high as 0.58.

Shen et al. [45] performed experiments in a 0.2 m diameter pipe with a total length of 26 m at low liquid flow rates. They found that the void fraction tends to increase in the inlet region due to expansion of small spherical bubbles. This is followed by a decrease in the void fraction along the flow channel caused by an increase in the relative velocity as small bubbles begin to coalesce into larger Taylor bubbles. Eventually the void fraction begins to increase again as the coalescence rate decreases and expansion of Taylor bubbles begins to become important. This axial void fraction profile is shown in Fig. 2.16, which illustrates the evolution of the flow from dispersed, spherical bubbles to cap-bubbly flow in large diameter pipes.

In addition, early work in large diameter pipes was performed by Filiminov et al. [8], Styrikovich et al. [52] and Petukhov and Kolokoltsen [35] for steam-water flow in larger diameter channels. These studies have been published in Russian and the results were utilized in the development of Kataoka and Ishii's [24] correlation.

Fig. 2.16 *N*-shaped void fraction change along the flow development at low flow rate



These databases widely cover extensive experimental conditions such as flow channel diameters (0.102–0.480 m), pressures (0.1–13.8 MPa), superficial liquid velocity (0.0–2.6 m/s), superficial gas velocity (0.0–8.0 m/s), void fraction (0–0.94), bubble injection methods (bubbly or cap-bubbly injection), and fluid systems (air-water, nitrogen-water, and steam-water). However, the available data for void fraction in large diameter pipes still has some significant limitations. Especially for forced convection flows, data for the transition to annular flow region is insufficient to validate the models or flow regime transitions.

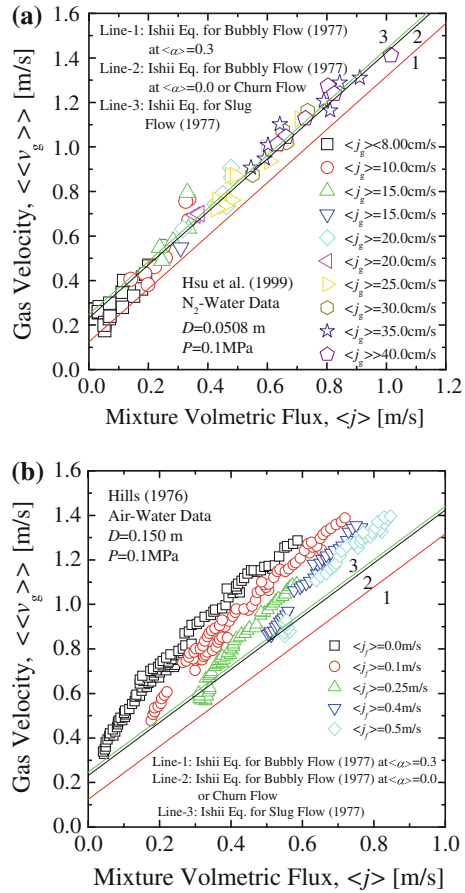
2.6 Comparisons and Analysis

In this section, the various models described in the previous section are evaluated based on the existing database of two-phase flow measurements in large diameter pipes.

2.6.1 Drift Velocity in Large Diameter Pipes

The drift velocity in the drift-flux model is the result of the relative velocity between the gas and liquid phases. Thus the increased relative velocity of bubbles in large diameter channels is expressed in the drift-flux model using the drift velocity. For relatively low void fractions such as those encountered in bubbly flow the drift velocity is determined by the rise velocity of small, spherical bubbles. Thus under these conditions the drift velocity in large diameter channels is similar to that in small diameter channels. When the flow condition becomes cap-turbulent however, the larger relative velocity of the cap bubbles results in a significantly higher drift

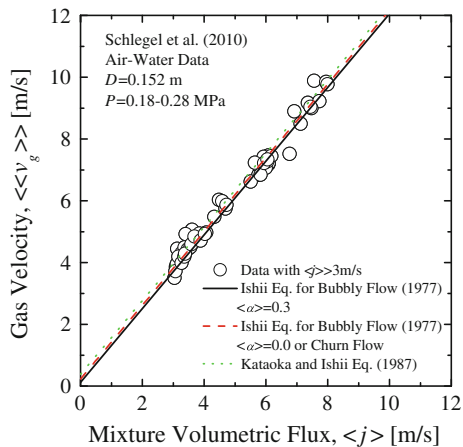
Fig. 2.17 Comparison of small diameter pipe data ($D = 50.8$ mm) with large diameter pipe data ($D = 150$ mm) at $P = 0.1$ MPa. **a** Hsu et al. [18] data for a small diameter pipe with $D = 50.8$ mm. **b** Hills [16] data for a large diameter pipe with $D = 150$ mm



velocity for drift-flux models in large diameter channels [14]. To illustrate this effect, data from a pipe with non-dimensional diameter of 18.5 ($D = 50.8$ mm, [18]) and 55 ($D = 152$ mm, [16]) are compared with the drift-flux correlations developed by Ishii [20] for small pipes in Fig. 2.17.

As shown in Fig. 2.17a, the data for bubbly flow in a vertical pipe with a non-dimensional diameter of 18.5 collapse to a single line. The drift flux correlations of Ishii [20] with the drift velocity for bubbly or churn or slug flow can predict the void fraction in the small diameter pipe over all flow conditions tested in the experiment of Hsu et al. [18]. On the other hand, Fig. 2.17b reveals the effect of the pipe diameter on the ability of drift flux correlations to predict the void fraction. The data for bubbly flow in a vertical pipe with an inner diameter of 150 mm do not fall on a single line. Generally the drift velocity gradually increases from that for bubbly flow to that for cap-bubbly flow over a transition region between bubbly flow and cap-bubbly flow as large bubbles begin to occur and eventually become dominant.

Fig. 2.18 Comparison of void fraction measured in a large pipe ($D = 0.152$ m) at high mixture volumetric flux with various drift flux correlations



2.6.2 Distribution Parameter in Large Diameter Pipes

Within the flow channel, there exists a profile of both velocity and void fraction. The result of these combined effects is that the gas phase tends to concentrate in the higher-velocity region in the center of the pipe [2]. This means that the gas is concentrated in a region of high velocity, while the liquid is concentrated in a region of low velocity. This profile effect is accounted for in the drift-flux model through the distribution parameter.

At low total volumetric fluxes, the change in drift velocity alone is not sufficient to explain all of the variation shown in Fig. 2.17. In addition, the discussion in Sect. 2.2.5 shows that the void and velocity profiles can vary significantly from those typically seen in small diameter pipes. This leads to significant changes in the distribution parameter under these conditions [14].

As the total mixture velocity is increased, the importance of the relative velocity and therefore the drift velocity decreases, while the effect of the void profile, expressed using the distribution parameter, becomes dominant. At very high mixture volumetric flux, it is not unreasonable to neglect the relative velocity and consider only the profile effect.

To illustrate this, the void fraction data collected by Schlegel et al. [37] has been compared to various drift-flux correlations in Fig. 2.18. The predicted drift velocities from Ishii's bubbly and churn flow equations and Kataoka and Ishii equation are 0.124, 0.231 and 0.381 m/s respectively for the experimental conditions shown in the figure. The difference in these drift velocity predictions does not significantly affect the prediction accuracy of the drift-flux model for these high mixture volumetric flux conditions. Figure 2.18 indicates that the distribution parameter given by Ishii [20] can be applicable to upward two-phase flow in a large diameter pipe at high mixture volumetric flux conditions.

2.6.3 Drift Flux Type Correlation Evaluations

In view of the great difference in flow characteristics for various flow regimes, the existing drift flux models are evaluated in two steps: for low and high void fraction flows respectively. The low void fraction flows correspond to the bubbly and bubbly-slug transition regions and the high void fraction flows correspond to the cap turbulent and churn turbulent flow regimes. The evaluation of these data sets with regards to the various models described previously is shown in the following figures. The first several figures, Figs. 2.19, 2.20, 2.21, 2.22, 2.23, 2.24, 2.25, 2.26, 2.27, 2.28, 2.29 and 2.30, show the comparison of bubbly flow data with the various models for void fractions smaller than 0.3. The remaining figures, Figs. 2.31, 2.32, 2.33, 2.34, 2.35, 2.36, 2.37, 2.38, 2.39 and 2.40, show the comparison of higher void fraction data, for void fractions larger than 0.3, with various models.

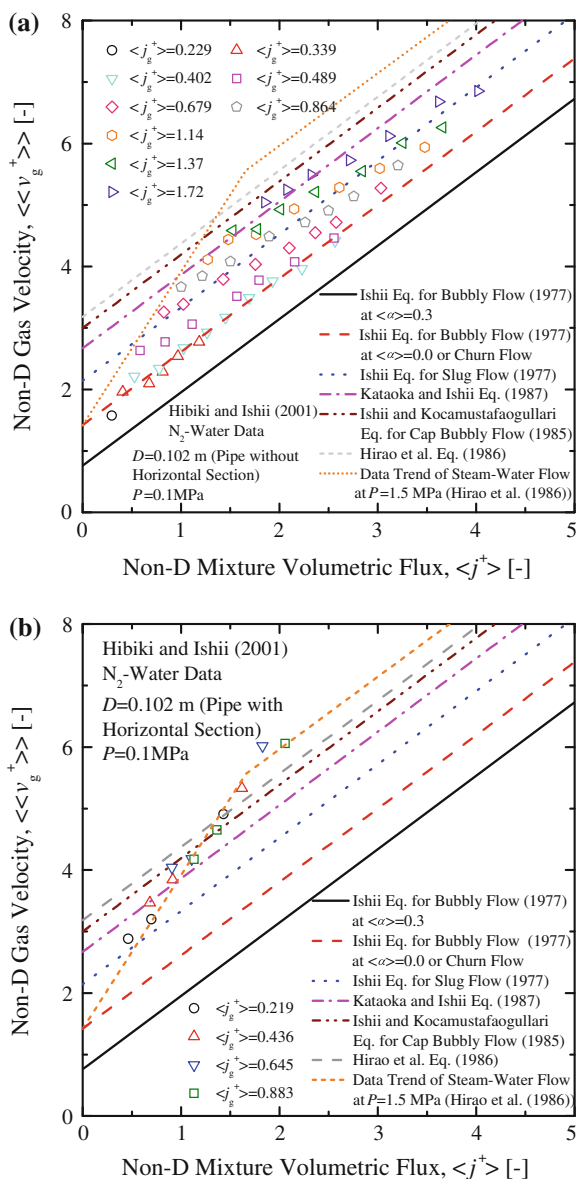
2.6.3.1 Evaluations with Low Flow Rate and Low Void Fraction Data

Figure 2.19 shows the comparison of data collected in a 0.102 m diameter pipe with various drift-flux correlations as a parameter of superficial gas velocity. Figure 2.19a shows the behavior of bubbly flows at low void fraction when the inlet flow regime is uniformly distributed bubbly flow. The bubbly data do not fall on a single line at low mixture volumetric flux. As the figure shows none of the correlations shown can accurately predict all of the data for bubbly flows in large diameter systems. Figure 2.19b shows the behavior of the flow when it is injected as cap-bubbly flow for void fractions smaller than 0.3, showing the strong effect that the inlet condition has on the flow. In this case, the flow behavior is mainly determined by the behavior of cap bubbles, which is why the data is predicted well by only a single line.

Figure 2.20 shows the comparison of Hills [16] correlation with the data of Hibiki and Ishii [13]. The overall agreement is fair; however the correlation cannot reproduce the behavior at very low void fraction and may not perform very well at higher liquid fluxes. Figures 2.21 and 2.22 show the comparison of Hibiki and Ishii's [13] data with the predictions using correlations of Shipley [48] and Clark and Flemmer [5] respectively. Neither model shows a reasonable agreement with the data. Figure 2.23 compares the correlation of Clark and Flemmer [6] with the data of Hibiki and Ishii [13]. The overall agreement between them is fairly good. However, for $\langle j_g^+ \rangle < 0.4$ the correlation overestimates the gas velocity, whereas for $0.5 < \langle j_g^+ \rangle < 1.2$ the correlation underestimates the gas velocity. It should be pointed out that these correlations were developed based on adiabatic air-water flow data. Since physical properties would affect the drift velocity significantly [24], the applicability of these correlations to high pressure or other fluid systems has not been validated.

Fig. 2.19 Comparison of various drift-flux correlations with air-water data in 0.102 m diameter pipe [13].

a The inlet flow regime is bubbly flow. **b** The inlet flow regime is cap-bubbly flow



The correlation of Murase et al. [29] is compared with the data of Hibiki and Ishii [13] in Fig. 2.24. The correlation does not predict the gas velocity well. The comparison with the correlation of Cunningham and Yeh [7] is shown in Fig. 2.25. Although the correlation was developed from data collecting during core uncovering experiments (namely pooling conditions), it greatly overestimates the low void fraction and low liquid flow rate data of Hibiki and Ishii [13].

Fig. 2.20 Comparison of Hills correlation [16] with air-water data in 0.102 m diameter pipe [13]

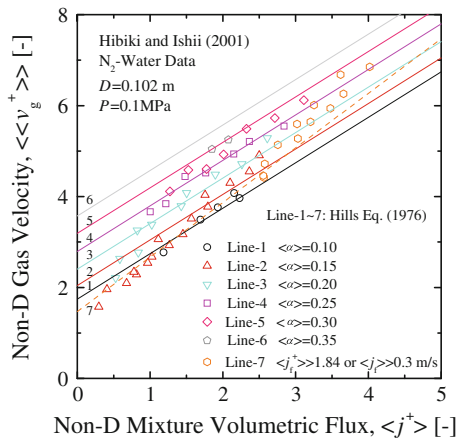


Fig. 2.21 Comparison of Shipley correlation [48] with air-water data in 0.102 m diameter pipe [13]

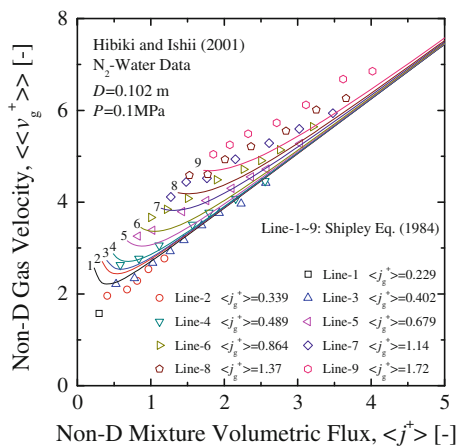


Fig. 2.22 Comparison of Clark and Flemmer correlation [5] with air-water data in 0.102 m diameter pipe [13]

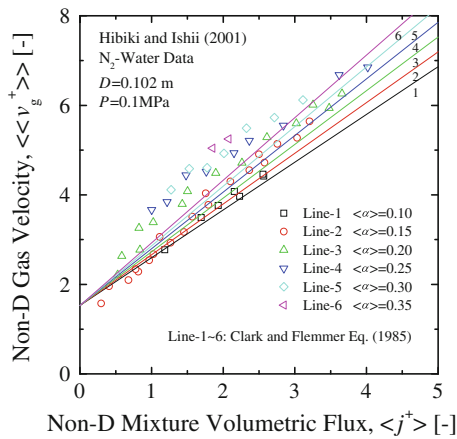


Fig. 2.23 Comparison of Clark and Flemmer correlation [6] with air-water data in 0.102 m diameter pipe [13]

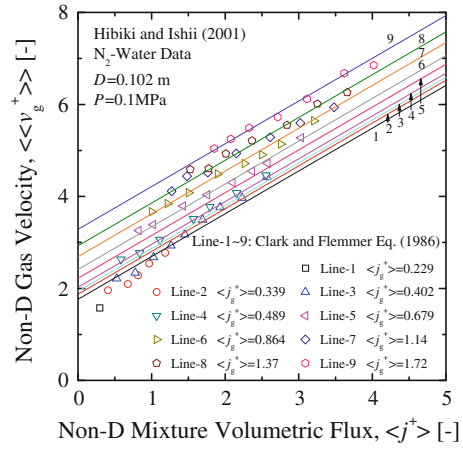


Fig. 2.24 Comparison of Murase et al. correlation [29] with air-water data in 0.102 m diameter pipe for bubbly flow injection [13]

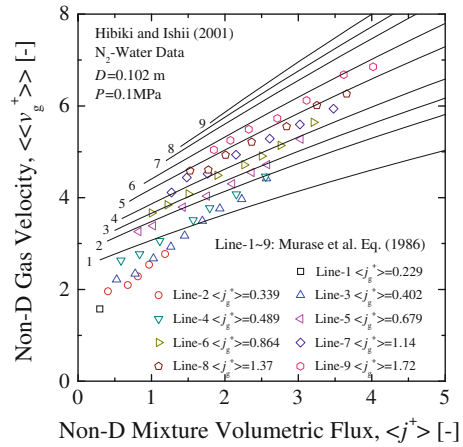


Fig. 2.25 Comparison of Cunningham and Yeh correlation [7] with air-water data in 0.102 m diameter pipe for bubbly flow injection [13]

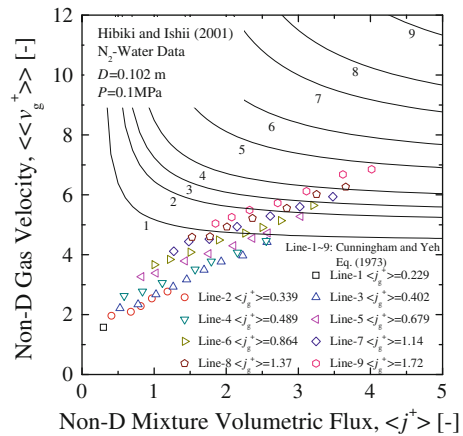


Fig. 2.26 Comparison of Chexal et al. correlation [4] with air-water data in 0.102 m diameter pipe for bubbly flow injection [13]

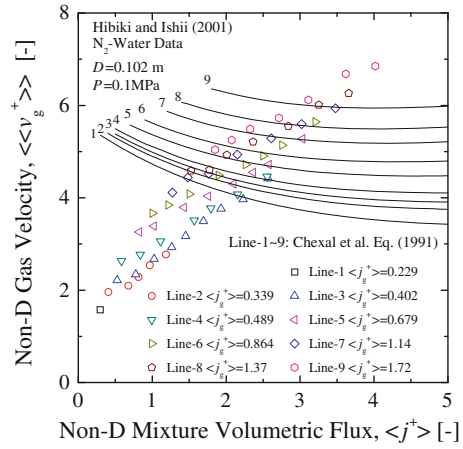


Fig. 2.27 Comparison of Shen et al. correlation [45] with air-water data in 0.102 m diameter pipe for bubbly flow injection [13]

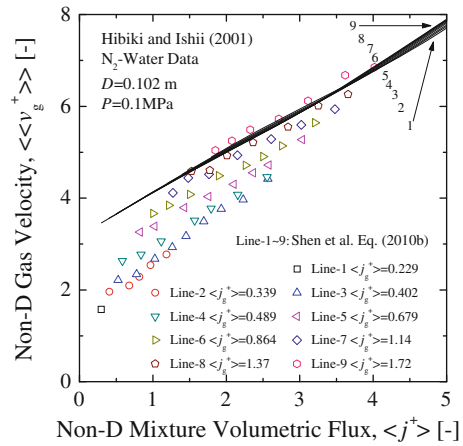


Fig. 2.28 Comparison of Hibiki and Ishii correlation [14] with air-water data in 0.102 m diameter pipe for bubbly flow injection [13]

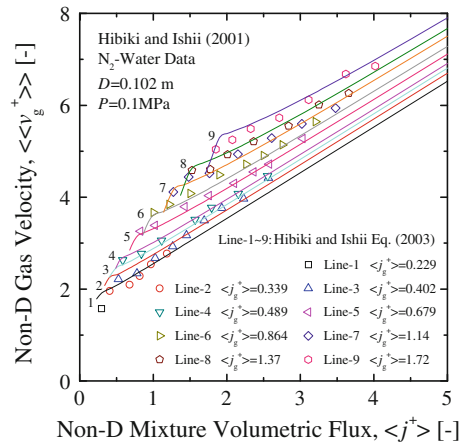


Fig. 2.29 Comparison of Hibiki and Ishii correlation [14] with air-water data taken by Hills [16] in 0.150 m diameter pipe

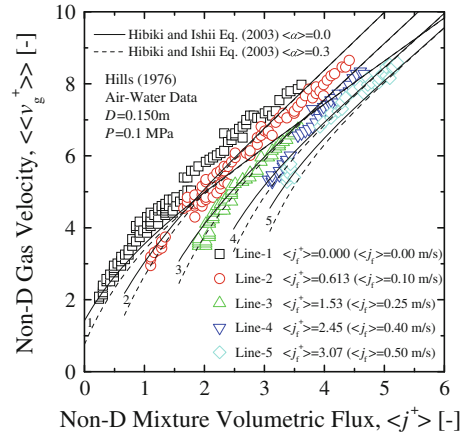


Fig. 2.30 Comparison of Hibiki and Ishii correlation [14] with air-water data taken by Hashemi et al. [11] taken in 0.305 m diameter pipe

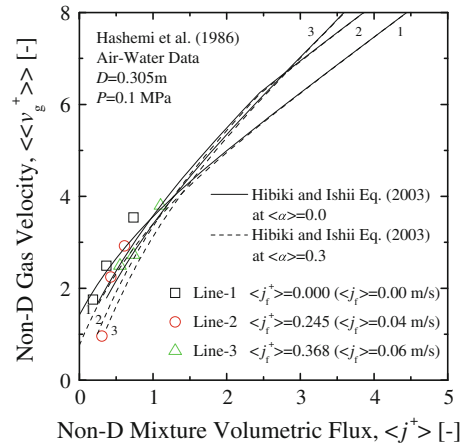


Fig. 2.31 Comparison of the correlations of Shipley [48] and Clark and Flemmer [5] with the data of Schlegel et al. [37]

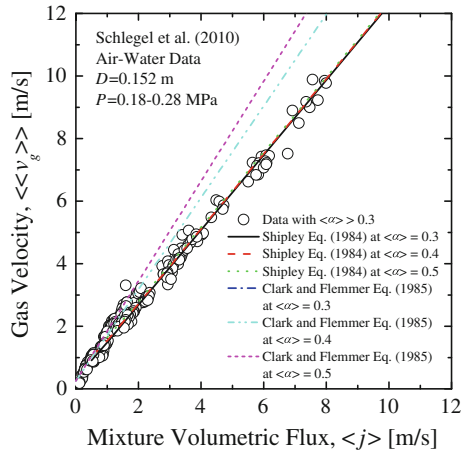


Fig. 2.32 Comparison of the correlations of Ishii [20], Hirao et al. [17], and Clark and Flemmer [6] with the data of Schlegel et al. [37]

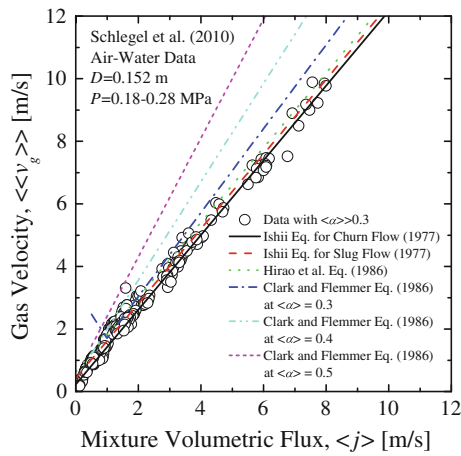


Fig. 2.33 Comparison of the correlations of Kataoka and Ishii [24] and Hills [16] with the data of Schlegel et al. [37]

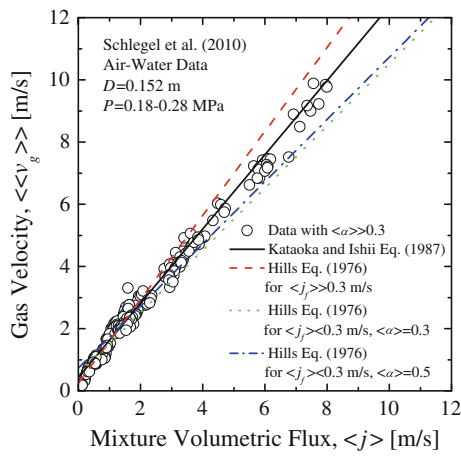


Fig. 2.34 Comparison of the correlations of Cunningham and Yeh [7] and Murase et al. [29] with the data of Schlegel et al. [37]

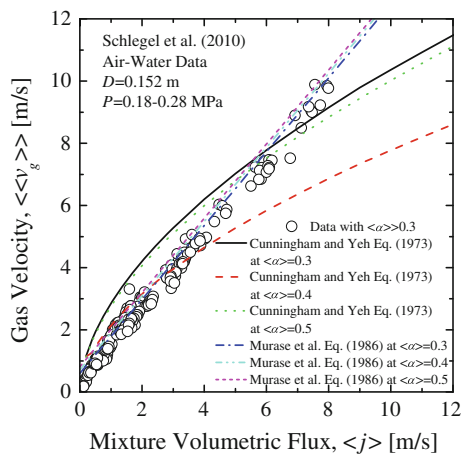


Fig. 2.35 Comparison of the correlations of Hibiki and Ishii [14] and Chexal et al. [4] with the data of Schlegel et al. [37]

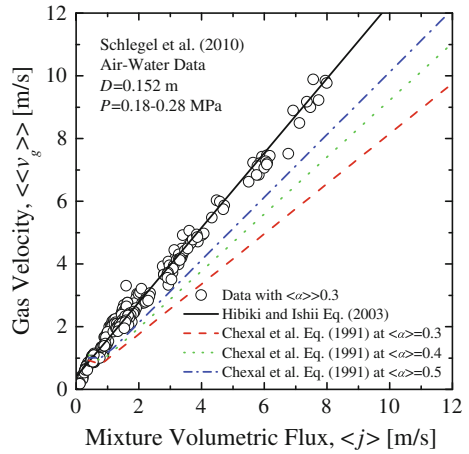


Fig. 2.36 Comparison of the correlations of Ishii and Kocamustafaogullari [23] and Shen et al. [45] with the data of Schlegel et al. [37]

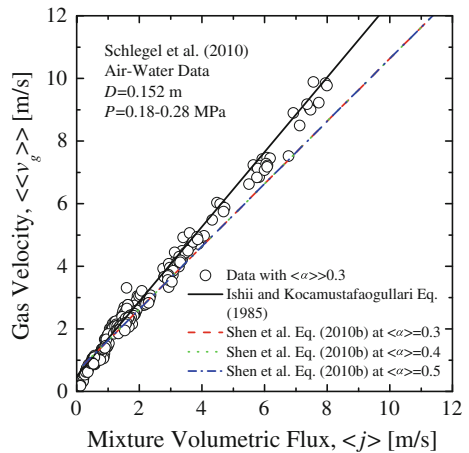


Fig. 2.37 Comparison of Hills' data [16] with the drift-flux model of Kataoka and Ishii [24] at high void fraction

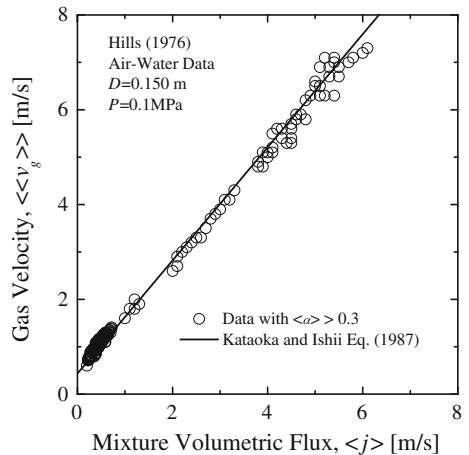


Fig. 2.38 Comparison of Sawant et al. [36] data with the drift-flux model of Kataoka and Ishii [24] at high void fraction

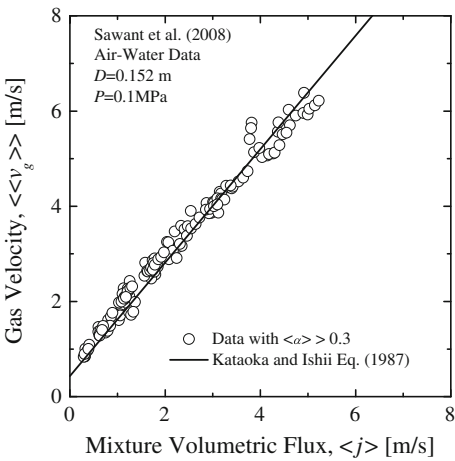


Fig. 2.39 Comparison of Schlegel et al. [37] data with the drift-flux model of Kataoka and Ishii [24] at high void fraction

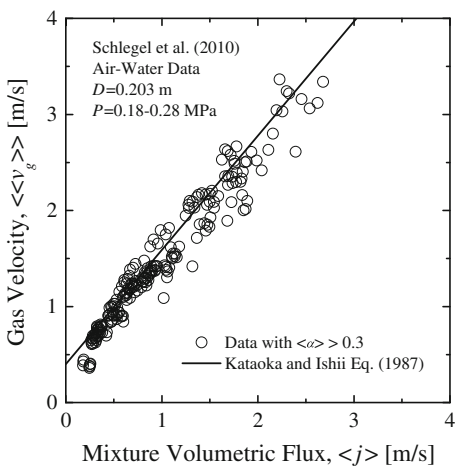
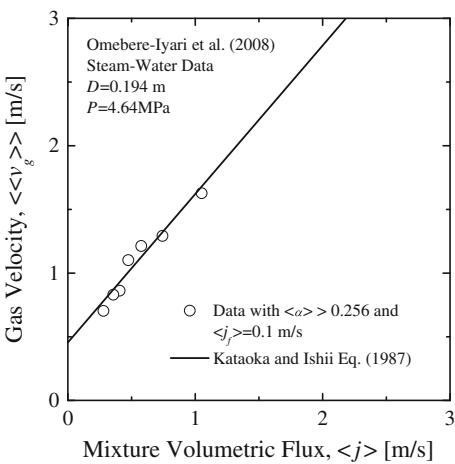


Fig. 2.40 Comparison of Omebere-Iyari et al. data [34] with the drift-flux model of Kataoka and Ishii [24] at high void fraction



In Fig. 2.26, the correlation developed by Chexal et al. [4] is presented in comparison with the data of Hibiki and Ishii [13]. The correlation of Chexal et al. [4] was developed based on a wide range data sets and has been shown to effectively predict the void fraction of the data set on which it is based. Figure 2.26 shows that its prediction for the experimental conditions of Hibiki and Ishii [13] does not capture the change in gas velocity as the mixture volumetric flux changes and the overall accuracy is not very good.

The correlation of Shen et al. [45] is compared with the data of Hibiki and Ishii [13] in Fig. 2.27. Although the correlation can predict the data in the flow transition from bubbly to cap turbulent flows, its agreement with the data of Hibiki and Ishii [13] is not acceptable and its accuracy at higher total flux conditions is questionable.

The comparison of the correlation developed by Hibiki and Ishii [14] with the data collected earlier by the same authors [13] is shown in Fig. 2.28. The solid lines indicate the predictions of the model for each superficial gas velocity. This drift-flux model, though complex, shows that it can accurately reproduce the experimentally measured gas velocity trends and values very well in Fig. 2.28.

The comparisons of the experimental data taken by Hibiki and Ishii [13] with the drift-flux type correlations in Figs. 2.19, 2.20, 2.21, 2.22, 2.23, 2.24, 2.25, 2.26, 2.27 and 2.28 show that the correlations developed by Hibiki and Ishii [14] gives the most reasonable predictions of the experimental data in large diameter pipes at low liquid flow rate and low void fraction conditions.

The correlation developed by Hibiki and Ishii [14] has also been compared to the data of other researchers. Figures 2.29 and 2.30 shows the comparisons of the correlations with the experimental data taken by Hills [16] and Hashemi et al. [11] at low liquid flow rate and low void fraction conditions. Solid and broken lines in the Figs. 2.29 and 2.30 indicate calculated values by the correlation of Hibiki and Ishii [14] at $\langle \alpha \rangle = 0$ and 0.3, respectively. Since Hills [16] and Hashemi et al. [11] carried out the experiments by changing the superficial gas velocity, keeping the superficial liquid velocity constant, the data in Figs. 2.29 and 2.30 are presented as a parameter of the superficial liquid velocity. The drift-flux correlation of Hibiki and Ishii [14] can predict the proper trend and the value of the experimental data very well. The correlation is found to have an average relative prediction error of $\pm 6.7\%$ for the 373 data sets of Hibiki and Ishii [13], Hills [16] and Hashemi et al. [11] shown in Figs. 2.28, 2.29, 2.30.

2.6.3.2 Evaluations with High Void Fraction Data

When the void fraction is higher than 0.3, the large cap bubbles are formed and play dominant roles in determining the flow behavior in large diameter pipes. The corresponding flow regimes fall into the regions of cap turbulent flow and churn turbulent flow. The flow characteristics in the high void fraction regions are expected to be different from those in the low void fraction regions. The data taken by Schlegel et al. [37] in pipes with 0.152 m diameter covered an extensive range

of void fractions. Among these data, the data with high void fraction ranging from 0.3 to 0.8 are valuable and typical. They are selected to evaluate all of the available drift-flux correlations in the flow regime transition from cap turbulent to churn turbulent flows.

Figure 2.31 shows the comparison of the high void fraction data collected in a 0.152 m diameter pipe with the drift-flux correlations of Shipley [48] and Clark and Flemmer [5]. Both of the correlations can predict the data at low mixture volumetric flux. Since Clark and Flemmer [5] did not give a physically accurate prediction for the distribution parameter the model agreement with high mixture volumetric flux data is not good. The correlations of Shipley [48] can predict the whole data well.

Comparison of the correlations of Ishii [20], Hirao et al. [17] and Clark and Flemmer [6] with the data of Schlegel et al. [37] is presented in Fig. 2.32. The churn and slug flow correlations of Ishii [20] were originally developed for high void fraction flow in a small diameter pipe. Although these models can acceptably predict the data for high mixture volumetric flux in large diameter pipes, this is due to a hydrodynamically accurate model for the distribution parameter. At lower mixture volumetric fluxes these models underestimate the data of Schlegel et al. [37] due to the differences in the drift velocity. The correlation of Clark and Flemmer [6] overestimated the data. The correlation of Hirao et al. [17] gave a reasonable prediction for the data.

The correlations of Kataoka and Ishii [24] and Hills [16] are compared with the data of Schlegel et al. [37] in Fig. 2.33. The correlation of Hills [16] predicts the high void fraction data well in the mixture volumetric flux ranging from 0.2 to 3 m/s. For low liquid flow rates however, the correlation underestimates the gas velocity and at high gas flow rates or pool conditions the correlation overestimates the gas velocity. The correlation of Kataoka and Ishii [24] was developed for the cap turbulent flow in pool conditions for large diameter pipes. Because of this, it reasonably predicts the high void fraction data of Schlegel et al. [37].

Figure 2.34 depicts the comparison of the correlations of Cunningham and Yeh [7] and Murase et al. [29] with the data of Schlegel et al. [37]. The correlation of Cunningham and Yeh [7] did not reasonably predict the data and its void fraction estimation is questionable in cap turbulent flow in a large diameter pipe. The correlation of Murase et al. [29] acceptably predicted the data of Schlegel et al. [37].

Comparison of the correlations of Hibiki and Ishii [14] and Chexal et al. [4] with the data of Schlegel et al. [37] is illustrated in Fig. 2.35. The generalized drift flux model of Chexal et al. [4] did not perform well. In most regions, it greatly underestimates the gas velocities in the large diameter pipe. Since the correlation of Hibiki and Ishii [14] is identical to the correlation of Kataoka and Ishii [24] in this region, it satisfactorily predicted the high void fraction data of Schlegel et al. [37].

Figure 2.36 shows the comparison of the correlations of Ishii and Kocamustafaogullari [23] and Shen et al. [45] with the data of Schlegel et al. [37]. Since the distribution parameter of Shen et al. [45] is set to be 1, its prediction for gas velocities in the high mixture volumetric flux region is not good. However, the

drift velocity correlation of Shen et al. [45] is similar to the correlation of Hirao et al. [17] for these experimental conditions and can predict the void fraction at low mixture volumetric flux. The prediction of Ishii and Kocamustafaogullari [23] shows reasonable agreement with the experimental data of Schlegel et al. [37].

From the above-mentioned discussions on the comparisons of the available drift flux models with the high void fraction data of Schlegel et al. [37] in Figs. 2.31, 2.32, 2.33, 2.34, 2.35 and 2.36, we know that the correlations of Shipley [48], Hirao et al. [17], Ishii and Kocamustafaogullari [23], Kataoka and Ishii [24] and Hibiki and Ishii [14] can satisfactorily predict the data in a large diameter pipe. Since physical properties are not included in the correlation of Shipley [48], it may be questionable to apply the correlation to high pressure or other fluid systems. The dependence of Hirao et al. [17] on diameter may not be physically realistic at very high diameters and therefore this correlation should not be used. Hibiki and Ishii [14] adopt the correlation of Kataoka and Ishii [24] when void fraction is greater than 0.3. So the drift-flux correlation of Kataoka and Ishii [24] is recommended for the high void fraction calculation in a large diameter pipe.

Many data sets for large diameter pipes other than that of Schlegel et al. [37] are available for relatively high void fraction conditions. Data was collected in a 0.152 m diameter pipe at 0.1 MPa pressure using an air-water mixture by Hills [16] and Sawant et al. [36]. Additional data was collected by Schlegel et al. [37] in a pipe with diameter of 0.203 m. Omebere-Iyari et al. [34] collected data using a steam-water system at pressure of 4.64 MPa in a pipe with diameter of 0.194 m. This data has been compared to the correlation of Kataoka and Ishii [24] in Figs. 2.37, 2.38, 2.39 and 2.40.

The comparison of the drift-flux correlation of Kataoka and Ishii [24] with these high void fraction data ranging from 0.1 to 4.64 MPa pressure and from air-water to steam-water shows good agreement. This serves to further illustrate the effectiveness of the drift-flux correlation of Kataoka and Ishii [24] for high void fraction flows. The impacts of different fluids and different pressures are included in the drift-flux correlation of Kataoka and Ishii [24] through the physical properties.

The comparisons in Figs. 2.33, 2.37, 2.38, 2.39 and 2.40 also illustrated that the pipe diameter ranging from 0.15 to 0.203 m has little effect on the behavior of the cap turbulent flow and churn turbulent flow and the performance of the drift-flux models in large diameter pipes.

In view of the evaluations of the 13 drift-flux type correlations from low and high void fraction data available now, the drift-flux correlations of Hibiki and Ishii [14] and Kataoka and Ishii [24] perform satisfactorily for low and high void fraction conditions respectively. They are recommended for void fraction prediction in the two-phase flow in large diameter pipes. Since the correlations for the distribution parameter and the drift velocity in a large diameter pipe have not been validated separately by detailed local flow data, they should not be used individually. In a future study, detailed local measurements of flow parameters for gas and liquid phases in a large diameter pipe are recommended to develop a detailed and more rigorous drift-flux correlation taking account of the detailed flow structure.

Because of the importance of predicting two-phase flows in large diameter pipes, many studies have been performed to measure the void fraction and other two-phase flow characteristics in such systems. These experiments have provided valuable insight into the behavior and structure of the flow, however there are still some limitations to the existing database. Very high void fraction data in the churn-turbulent and annular/mist flow regimes is still very limited, which inhibits the ability of researchers to develop models to predict the flow in this region. Further, data at very high liquid fluxes is limited to only a few data sets. These limitations must be remedied before a complete understanding of flows in large pipes can be realized.

2.6.4 Boiling Two-Phase Flow in Large Diameter Pipe

For a flow with generation of void at the wall due to nucleate boiling, the drift-flux models developed by Hibiki and Ishii [14] and Kataoka and Ishii [24] have not been validated by experimental data. For such a flow condition, the distribution parameter should have a near-zero value at the beginning of the boiling two-phase-flow region. With the increase in the cross-sectional mean void fraction, the peak of the local void fraction moves from the near-wall region to the central region. This will lead to the increase in the value of the distribution parameter as the void profile develops.

In view of these basic characteristics described above, Ishii [20] proposed the following simple correlation for the distribution parameter of boiling flow.

$$C_0 = \left\{ C_\infty - (C_\infty - 1) \sqrt{\frac{\rho_g}{\rho_f}} \right\} \{1 - \exp(k\langle\alpha\rangle)\} \quad (2.16)$$

where k is a coefficient. Ishii determined C_∞ (the asymptotic value of C_0) to be 1.2 and the coefficient, k , to be -18 with experimental data taken in relatively small channels (1977). For a wall nucleate boiling in a large diameter pipe, the above model should also be sound. Equation (2.16) suggests that the void fraction would be the dominant factor for determining the distribution parameter in boiling flow. The impacts of different fluids and different pressures are included through the physical properties. The changes in the distribution parameter with void fraction and pressure are shown in Fig. 2.41 for water and steam boiling flow at saturation states, with the operating pressures of BWR and PWR highlighted.

As mentioned above, as the cross-sectional mean void fraction increases, the peak of the local radial void fraction profile moves from the near-wall region to the central region. When the void fraction is high enough, the boiling two-phase flow would have flow characteristics similar to non-boiling two-phase flow. The comparisons in Sect. 2.6.3.2 showed that the high void fraction steam-water data without boiling were well predicted by the drift-flux correlation of Kataoka and

Fig. 2.41 Boiling flow distribution parameter changes with void fraction and pressure at saturation states

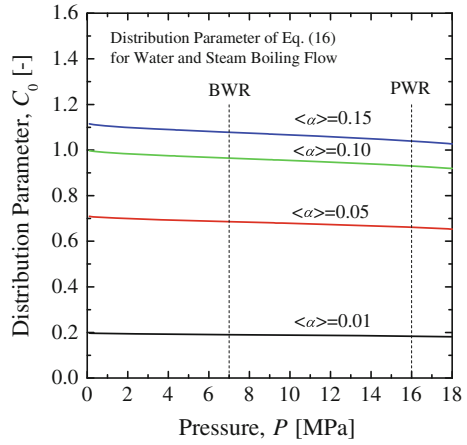
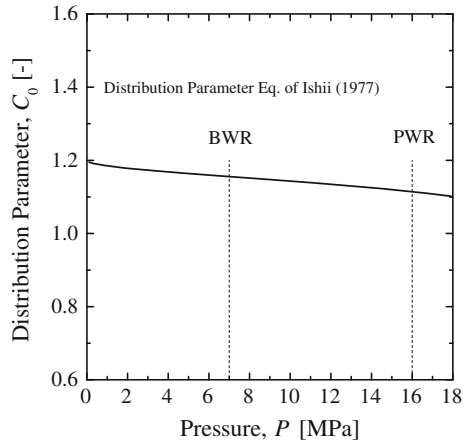
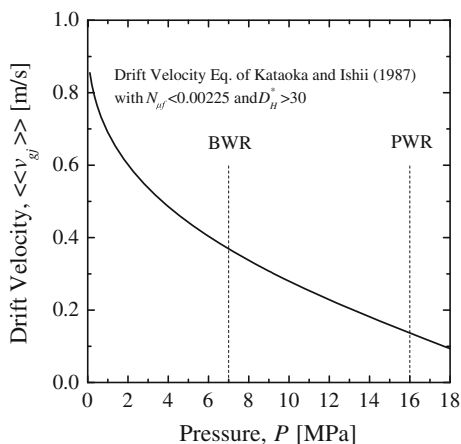


Fig. 2.42 Distribution parameter of Ishii [20] changes with pressure at saturation states



Ishii [24]. The model of Kataoka and Ishii consists of the original distribution parameter correlation of Ishii [20] and the drift velocity correlation of Kataoka and Ishii [24]. So the two correlations would be expected to be able to predict boiling two-phase flows with high void fraction. The distribution parameter and drift velocity correlations do not depend on the void fraction but change with the different fluids and different pressures through the physical properties. The changes in the distribution parameter and drift velocity with pressure are shown in Figs. 2.42 and 2.43 respectively for water and steam boiling flow at saturation states, with the operating pressures of BWR and PWR highlighted.

Fig. 2.43 Drift velocity of Kataoka and Ishii [24] changes with pressure at saturation states



2.7 Summary

Two-phase flows are essential in a wide variety of industrial applications, and in many of those applications large diameter channels form an integral component. To predict the performance of these systems, the ability to predict the void fraction in such large diameter channels is essential. This task is complicated, however, by the many differences between small and large diameter channels. The instability of cap bubbles leading to the lack of typical gas slugs in larger pipes leads to changes in the turbulence characteristics, flow pattern and bubble behavior, and relative velocity between the phases as the pipe diameter increases. The overall effect of these changes is that models for small diameter channels, which have been well-developed, cannot be used to predict flows in large diameter channels. To predict two-phase flows, the drift-flux model is commonly used due to its simplicity and ease of use. Many correlations have been developed using the framework provided by the drift-flux model in order to predict the behavior of two-phase flows in large diameter channels. Many of these correlations are empirically based on only a few data sets and so have limited applicability and accuracy.

In this chapter, the latest progress in two-phase flows in large diameter pipes is discussed. The flow characteristics in large diameter pipes are analyzed and summarized by using experimental data. The flow regimes and their transition criteria are discussed. The existing drift-flux models are collected and their strengths and weaknesses are noted. The up-to-date void fraction data that can be used to evaluate these models are presented. In view of the different flow regimes in the large diameter pipe, the existing drift-flux models are evaluated systematically in both low and high void fraction flows. Based on the existing database, the drift flux models of Hibiki and Ishii [14] and Kataoka and Ishii [24] seem to perform best for bubbly, cap-turbulent and churn-turbulent flow. They are recommended for use in predicting the void fraction in bubbly flow [14] and in cap-turbulent and churn-turbulent flows [24].

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