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Evolution of flow patterns, gas fraction profiles and bubble size distributions in gas-liquid flows in vertical tubes

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Abstract

Air-water flow at ambient conditions in a vertical pipe with an inner diameter of 51.2 mm is investigated. An electrode wire-mesh sensor enables the measurement of the phase distribution with a very high resolution in space and in time. Local bubble size distributions are calculated from the data. The measurements were done in different distances from the gas injection device. As a result, the development of bubble size distributions as well as the development of the radial gas fraction profiles can be studied. It was found, that the bubble size distribution as well as local effects determine the transition from bubble flow to slug flow. The data are used for the development of a model, which predicts the development of the bubble size distribution and the transition from bubble flow to slug flow in the case of stationary flow in a vertical pipe.

Keywords: Two-phase flow; Flow pattern transition; Bubble Flow; Slug flow

1 Introduction

Great efforts are made world-wide to develop CFD codes for two-phase flows in complicated three-dimensional geometries. In the case of bubbly flow, and especially for the correct prediction of the flow pattern transition from bubbly to slug flow, the codes must be equipped with constitutive laws describing the interaction between the gaseous and the liquid phases in a more detailed way than this is done by the wide-spread assumption of mono-disperse bubble flow.

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Recently attempts were made to solve this problem by the introduction of additional equations for the bubble density or similar parameters like bubble diameter, bubble volume or interfacial area. Rates for bubble coalescence and frequencies for bubble break-up, which form the source terms in these equations, are determined as local quantities. That means, they depend on local parameters of turbulence as well as on the local bubble size distribution. In order to reflect the fact that bubbles of different sizes develop different spatial distributions, it is necessary to introduce multi-bubble-size models. The main challenge is to find model equations quantifying the interaction between these bubble-size classes.

Gas-liquid flow in vertical pipes is a very good object for studying the corresponding phenomena. Here, the bubbles move under clear boundary conditions, resulting in a shear field of nearly constant structure where the bubbles rise for a comparatively long time. This allows to study the lateral motion of the bubbles in a shear flow by comparing distributions measured at different heights. It was shown, that the radial distribution of bubbles strongly depends on their diameter. In the case of a vertical upflow, smaller bubbles tend to move towards the wall, while large bubbles are preferably found in the centre. This was initially observed for single bubbles [1]. In an air-water system at ambient conditions the change of the direction of the lift force in the shear field occurred at a bubble diameter of about 5-6 mm. We could confirm this also for multi-disperse flow [2] (Fig. 2). This is very important for the evolution of the flow because the local bubble size distributions may differ significantly from those averaged over the cross section.

For this reason detailed investigations of gas-liquid flows in vertical tubes including the transition from bubble flow to slug flow along the flow path were made. The change of the bubble size distribution along the pipe as well as the changing radial profiles of the gas fraction represented by bubbles of different size were measured by fast wire-mesh sensors developed at our institute.

2 Experimental setup and instrumentation

Evolution of the bubble size distribution was studied in a vertical tube of 51.2 mm inner diameter supplied with air-water mixture at 30^0C. The distance between sensor and air injection was varied from 0.03 m to 3.03 m (inlet lengths 0.6-60 L/D). Gas and liquid superficial velocities were varied in a wide range. Stationary flow rates of air and water were used. About 150 combinations of the superficial velocities were considered. These include stable bubble flow, finely dispersed bubble flow, slug flow and annular flow at the upper end of the pipe. Transitions between the flow regimes were observed within the pipe.

Data were obtained by an electrode wire-mesh sensor (Fig. 1) performing measurement of the instantaneous conductivity distribution [3]. Two electrode

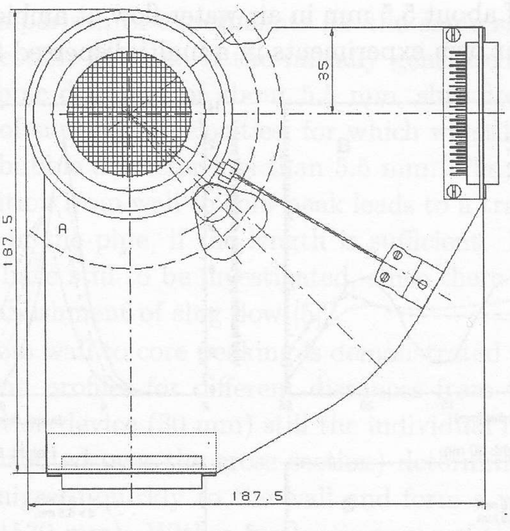

Figure 1. Scheme of the wire mesh sensor.

grids with 24 electrode wires each (diameter 120 μm) are placed at an axial distance of 1.5 mm behind each other. During signal acquisition, the electrodes of the first grid (transmitter wires) are supplied with short voltage pulses in a successive order. The currents arriving at the second grid (receiver wires) are digitalized by ADCs and stored in a data acquisition computer. Two sensors were put at a distance of 36 mm behind each other to measure velocities, too. For this sensor assembly, the time resolution of 2500 frames per second was achieved. The spatial resolution is given by the pitch of the electrodes and equals 2 mm.

The sensor delivers a sequence of two-dimensional distributions of the local instantaneous conductivity, measured in each mesh formed by two crossing wires i and j. Local instantaneous gas fractions are calculated assuming a linear dependence between gas fraction and conductivity. The result is a three-dimensional data array i, j, k, where k is the number of the instantaneous gas fraction distribution in the time sequence. A special procedure, described in [4] allows the identification of single bubbles and the determination of their volume and the equivalent bubble diameter. Using this procedure bubble size distributions as well as gas fraction profiles for bubbles within a predefined interval of bubble sizes can be calculated, the latter by using the method described in [2].

3 Experimental results

Tomiyama found a correlation for the bubble lift force by investigations of the behaviour of single bubbles within a well defined shear field [1]. It changes sign at a bubble diameter of about 5.5 mm in air-water flow at ambient conditions. This was confirmed by our own experiments in a multi-dispersed flow (Figs. 2c-2f).

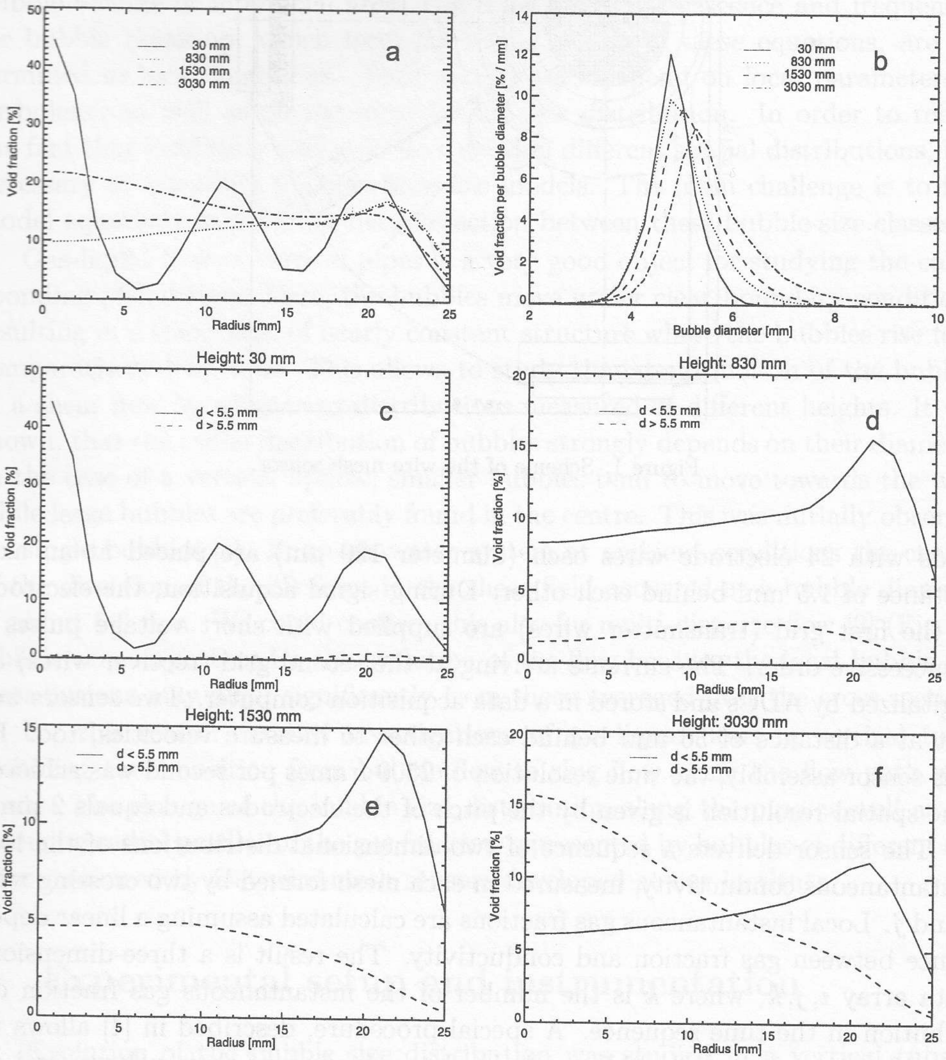

Figure 2. Radial gas fraction profiles and bubble size distributions for different distances from the gas inlet, superficial velocities: $j_l = 1$ m/s, $j_g = 0.14$ m/s.

Stable bubble flow was only observed for high water superficial velocities and low air superficial velocities. In this case a stable sharp wall peak of the gas fraction was found. In all cases with a core peak, the development of the bubble size distributions along the pipe indicates, that the coalescence rate exceeds the break-up rate. Although at $L/D \cong 60$ the transition to slug flow is not yet completed in many cases, it can be expected at larger distances. This is also evident from the comparison of experiments for the same superficial velocities, but different gas injection devices. If the initially generated bubbles are larger than the critical bubble diameter of about 5.5 mm, slug flow is established for many combinations of superficial velocities, for which we still found bubble flow in case of an initial bubble diameter less than 5.5 mm. The experiments clearly show, that the transition from wall to core peak leads to a transition to slug flow somewhere upwards in the pipe, if the length is sufficient. In pipes of a larger diameter the effects have still to be investigated, since there is a maximum pipe diameter for the establishment of slug flow [5].

The transition from wall to core peaking is demonstrated at Fig. 2. Figure 2a shows the gas fraction profiles for different distances from the gas inlet. Very close to the gas injection device (30 mm) still the individual injection nozzles (19 nozzles, equally distributed over the cross section) determine the radial profile. The small bubbles migrate quickly to the wall and form a wall peak of the gas fraction (830 mm – 1530 mm). With a further increase of the distance from the gas inlet a transition to core peaking is observed (3030 mm). Figure 2b shows the corresponding bubble size distributions. More and more larger bubbles are generated by coalescence.

The figures 2c to 2f show again gas fraction profiles for different distances from the inlet, but here they are subdivided according to the bubble diameter. Only few bubbles exceeding 5.5 mm are generated by the air injector. They appear due to coalescence at higher positions and migrate to the core of the pipe. These bubbles form a clear core peak at 3030 mm.

For other combinations of superficial velocities the transition from bubble to slug flow was observed. The process of transition is very fast, if a sufficient fraction of bubbles with a diameter between 10 mm and 15 mm is generated by coalescence. Figure 3 shows a typical evolution of the bubble size distribution in case of the transition. Bubbles larger than 10 mm in diameter coalesce with a high rate and form slugs. Consequently, a transition to a bi-modal bubble size distribution is observed.

4 The transition from bubble to slug flow

As the experimental data suggest, the transition from bubble to slug flow along the flow path in case of an co-current flow within a pipe is influenced by local effects depending on the bubble size. Bubble coalescence and bubble break-up, which cause the transition, depend on the local bubble densities n as well as on the dissipation rate of the turbulent kinetic energy ε:

coalescence rate: $\Gamma_{i,j} = f(d_i, d_j, \varepsilon)^* n_i^* n_j$

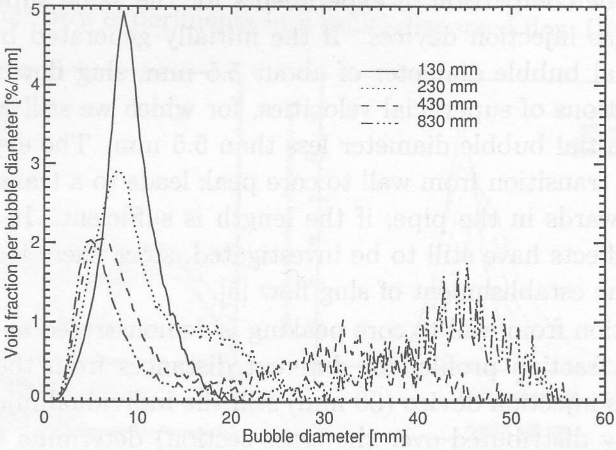

Figure 3. Development of the bubble size distribution during transition from bubble to slug flow, $j_l = 0.4$ m/s, $j_g = 0.34$ m/s.

$$\text{break-up rate: } \Omega_i = f(d_i, d_j, \varepsilon)^* n_i$$

i and j indicate the bubble class. The lift force found by Tomiyama [1] causes, that small bubbles (diameter < ca. 5.5 mm in case of air-water flow) can be found preferably in the wall region, while larger bubbles are accumulated in the core region.

Another important fact is, that the dissipation rate of turbulent energy is larger in the near wall region than in the core flow in most cases. The consequences for the transition to slug flow can be explained by help of Fig. 4. An upward air-water flow is considered. In both considered cases small bubbles (diameter < 5.5 mm) are injected. In the left side of the figure a low superficial gas velocity was assumed. The small bubbles tend to move towards the wall. The local gas fraction in the wall region is larger than the averaged gas fraction, but it is still low. In this case bubble coalescence and break-up are in equilibrium and an stable bubble flow is established.

If the gas superficial velocity is increased (Fig. 4, right side), the equilibrium between bubble coalescence and break-up is shifted towards a larger bubble diameter, because the coalescence rate increases with the square of the bubble density, while the break-up rate is only proportional to the bubble density. The bubble break-up rate strongly increases with the bubble diameter.

By a further increase of the gas superficial velocity, more and more large bubbles (diameter > 5.5 mm) are generated. They start to migrate towards the pipe centre. If enough large bubbles are generated by coalescence in the wall region, some of them can reach the core region without break-up. Because of the lower dissipation rate of turbulent energy they can then growth up by further coalescence at much lower break-up rates, typical for the low shear in the centre. This mechanism is the key for the transition from bubble to slug flow. That means, for an appropriate modelling of the transition a number of bubble classes as well as radial gas fraction profiles for each bubble class have to be considered.

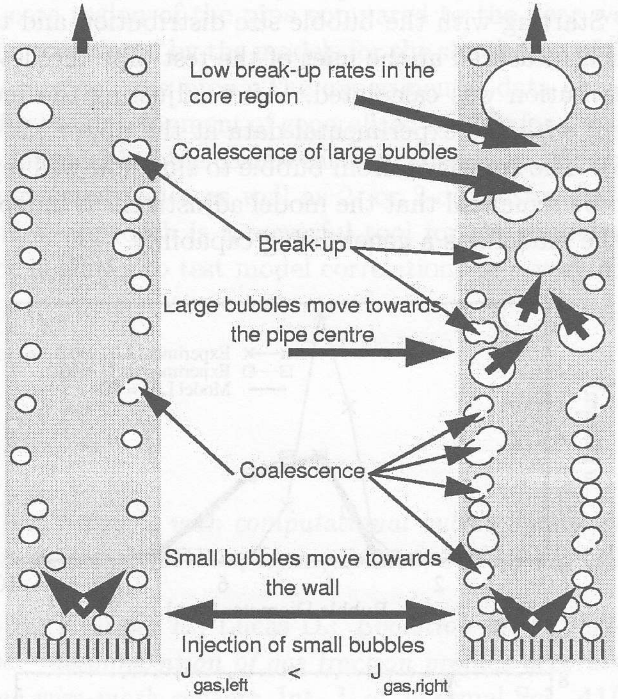

Figure 4. Stable bubble flow (left) and transition to slug flow (right).

5 Modelling

According to the importance of the bubble size distributions and the dependence of the parameter on the radial position, a model, which considers a large number of bubble classes was developed [6]. The radial profiles of the bubble density are the result of the non drag forces, acting perpendicularly to the flow direction. They are caused by the liquid shear flow (lift force), turbulence (turbulent dispersion force) and no-slip boundary at the tube wall (lubrication force). The model calculates the radial profiles of the bubble density for each bubble class as well as the radial profile for the energy dissipation rate per unit mass on a basis of a given overall bubble size distribution [7]. The model assumes an equilibrium of all forces acting on a bubble perpendicularly to the flow direction. In the result local bubble size distributions for all radial positions are available. Models for bubble coalescence and break-up use these distributions as an input to calculate local rates of bubble sources and sinks for each bubble class and each radial position individually.

First calculations were made using simplified assumptions for coalescence and break-up rates. Starting with the bubble size distribution and the radial profile of the gas fraction measured at the inlet of the test pipe the development of the bubble size distribution was calculated. After adjusting the model parameters a good agreement with the experimental data at the upper end of the pipe was achieved (Fig. 5). The transition from bubble to slug flow was reproduced by the model. It has to be remarked that the model adjustment is independent from the flow rates, i.e. the model has a generalising capability.

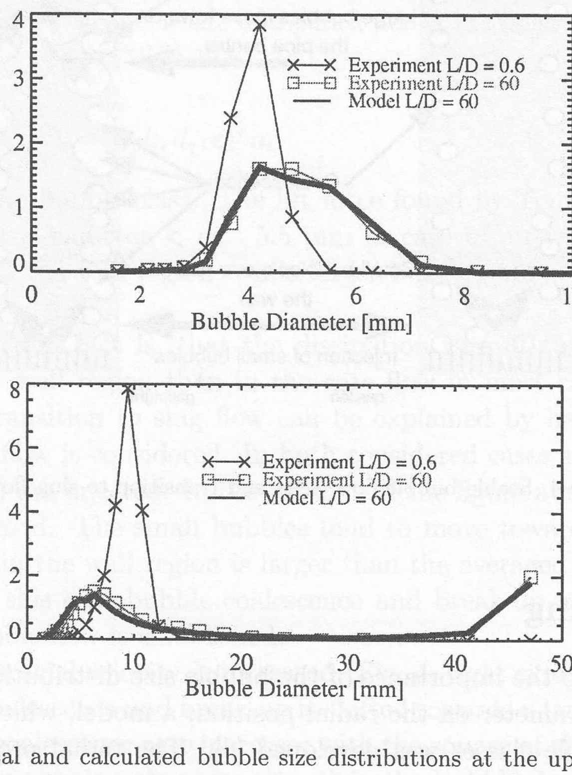

Figure 5. Experimental and calculated bubble size distributions at the upper end of the test pipe (3030 mm from the gas inlet). $j_l = 1$ m/s, a) $j_g = 0.09$ m/s, b) $j_g = 0.5$ m/s.

Using the new experimental data the models for bubble coalescence and break-up will be improved. The models for the non-drag bubble forces and the rates for coalescence and break-up can be applied in one-dimensional codes as well as in CFD-codes.

6 Conclusions

The experimental data clearly show, that the transition from bubble to slug flow is determined by the effects depending on the bubble sizes as well as on local effects. For an adequate modelling a large number of bubble classes (e.g. 25) has to be considered. There are very different conditions for bubble coalescence and break-up in the core region of the pipe compared to the near wall region. This fact has also to be considered by the models for the simulation of the development of the flow pattern along the pipe. The experimental data obtained in vertical pipes are useful for the development of generalised models for the non-drag bubble forces as well as for local bubble coalescence and break-up. These models can be used by one-dimensional codes as well as 2- or 3-dimensional CFD-codes. The presented simplified approach is a powerful tool for the model development for CFD codes, since it allows to test model correlations in a very efficient manner.

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