4. Experimental Characterisation of Gas-Liquid Flow: Major Hydrodynamics Parameters

4.1 Introduction

The current and the following chapter report and discuss experimental data obtained from the gas-liquid flow in the 2D column. These chapters are subdivided in sections in which all measured variables are separately discussed. The experimental data is presented in a logical order according to its complexity and links with other measured parameters. This chapter focus on a generic description of the two-phase flow and the parameters that globally characterise flow of gas-liquid mixtures, such as flow regimes, pressure drop, gas volume fraction and interfacial area. The next chapter reports the two-phase flow variables that are specifically coupled with a given phase, i.e., bubble properties and liquid field characteristics. The experimental techniques and procedures used to measure these quantities were described in Chapter 3.

The experimental data to be presented first is a comparison of two-phase flow patterns occurring in unpacked and packed columns with special focus given on the bubble-slug regime transition. The flow regimes were determined by visual inspection of the tracks obtained by video camera. Then, the pressure drop measured in single- and two-phase flow are presented both for the unpacked and packed column. Finally, the gas volume fraction and the specific interfacial area are reported and their spatial distribution and lateral
profiles are discussed. A correlation between these two quantities is shown and the impact of the packing and surfactant addition on the phase distribution is presented.

4.2 Flow Regimes in Gas-Liquid Flow

Knowing the flow regime in two-phase flow is as important as knowing whether the flow is laminar or turbulent in single-phase flow. Interfacial momentum, heat and mass transfer rates depend strongly on the flow regimes that finally determine the overall efficiency of chemical reactors or heat exchangers. Flow regimes are mainly determined by the ratio of gas and liquid velocities and, in a less extent, by the physical properties of the two phases and the column geometry. Flow patterns commonly encountered in larger pipes and channels, such as, bubbly, slug, churn and annular flows, occur also in narrow flat channels. Some differences do however exist in the gas-liquid interface configurations between the flows in larger channels and narrow flat channels. In this section, the definition of flow patterns is given first, followed by the presentation of the flow regime maps obtained in the 2D flat column used in this work. Finally, the comparison of the experimental results with available theoretical models and the flow regime maps of other researchers is performed.

4.2.1 Flow Regime Definitions

Two-phase regimes are classified according to the gas–liquid distributions in the channel. Since in the present case, the aspect ratio of the duct is large and the gap is very narrow, large bubbles in narrow passages are flattened to the point where the radius of curvature is almost independent of the transversal position. Therefore, bubbles can be considered two-dimensional and the flow regimes may be determined based upon the shape of the bubbles observed through the wider wall. The various flow regimes encountered in the 2D flat column are illustrated in Figure 4.1a for the unpacked column and in Figure 4.1b for the packed column. Since the flow regimes in the unpacked column have similar characteristics as in the packed column, the following descriptions can be applied in both situations.

- Bubbly flow is characterised by circular or oval bubbles, with a relatively narrow bubble size distribution, and with almost no bubble coalescence. If the bubble dimension is larger than the channel gap thickness, bubbles become flattened and
appear as flat objects of pancake like shape. It occurs at low to moderate gas flow rates. At high liquid flow rates, increased turbulent intensity in the liquid phase breaks up the bubbles into smaller ones, and this bubble regime has been referred in the literature as dispersed bubbly flow.

Figure 4.1 Examples of different flow patterns observed in: a) unpacked and; b) packed flow.
• **Slug flow** can be encountered at moderate liquid and gas rates. In the slug regime, large wobbling bubbles of cap shape are observed together with small circular bubbles. Therefore, its bubble size distribution is usually characterised by two separated peaks. Due to different bubble rise velocities, bubble coalescence plays here an important role and taking place especially right above the gas distributor. If the liquid rate is increased, turbulent fluctuations cause the break up of slug bubbles and the transition to dispersed bubbly regime occurs.

• **Churn regime** is characterised by a broad bubble size distribution and irregular bubble shapes, with the smaller bubbles flowing in the wake of the large bubbles. It is a highly turbulent and dynamic situation of two-phase flow, where bubble-bubble interactions are very strong. Both bubble coalescence and bubble break-up are very intensive processes in this regime. The visual discrimination between slug and churn regime is rather subjective and the transition has been usually based upon the fact that large bubbles become noticeable irregular, deformed and unstable. Many small vortices with entrapped small bubbles can be also observed.

• **Annular flow** occurs at high gas rates when the gas occupies the majority of the flowing area and most of the liquid moves up in the form of a wavy liquid film. In the 2D unpacked column, this flow regime is hardly distinguished visually from the highly turbulent churn regime. In flow with packing, annular flow occurs when the gas occupies almost totally the space inside the channels between particles. Liquid moves up only in the form of thin bridges intermittently connecting the neighbouring packing particles.

### 4.2.2 Flow Regimes Identification

Various methods for flow regime identification have been used for narrow channels, ranging from visual observations to examination of the statistical properties of gas volume fraction or pressure fluctuations. Most investigators have relied on visual observations by still photography as for example Lowry and Kawaji (1988), Troniewski and Ulbrich, (1984), Mishima et al. (1993), or the use of high-speed video cameras (Wilmarth and Ishii, 1994). Non-visual techniques are considered less subjective, but the criteria used for each flow description need still be properly established. Non-visual methods have been usually based on statistical characterisation of the gas volume fraction fluctuations by the X-ray
Attenuation technique (Jones and Zuber, 1979) or by electrical conductivity (Ali et al., 1993). Attempts to visualise two-phase flow in metallic ducts by neutron radiography have been done by Mishima and Hibiki (1996). A promising method to determine flow patterns appears to be based on the pressure signal PDF — *probability density function* (Tutu, 1982; Tutu, 1984; Matsui, 1984 and Wambsganss et al., 1991). The characteristics of pressure PDF’s are different for each flow regime. A single peak occurs at low gas volume fraction in bubbly flow; a double peak was observed for slug and churn regimes, and single broad peak at high volume fraction for annular flow. These simple statistical criteria can be easily employed in industrial applications for the automatic detection of two-phase flow regimes. Despite of their indisputable advantages, the visual determination of the flow regimes remains a fundamental and probably the most accurate way for the flow regime detection.

In the present study, visual analysis of video tracks was performed for each flow condition and for two column configurations (packed/unpacked). The transitions between the flow regimes are usually represented in the form of flow pattern maps where the boundaries between flow regimes are plotted as a function of gas and liquid superficial velocities. In the measured interval of gas-liquid superficial velocities, the detection and distinction of the four above mentioned flow regimes was carried out and the flow pattern maps were constructed for the unpacked and the packed column, as shown in Figure 4.2a and Figure 4.2b, respectively.

![Flow regime maps](image)

**Figure 4.2** Flow regime maps with the indication of experimental points for: a) unpacked and; b) packed column.
Bubbly flow regime in the packed column occurs in a range of $U_{GS}$ values that is more than twice the range observed in the unpacked column. The shift in the bubbly-slug transition for the packed column is caused mainly by the inhibition of bubble coalescence due to the presence of packing. The main mechanism of bubble coalescence observed in the unpacked column was mostly due to the entrainment of trailing bubbles into the wake of leading bubbles and a consequent acceleration of the trailing bubbles towards the leading one. In the presence of packing, the bubble wake effect is strongly dissipated and bubbles interfere almost exclusively due to a periodic acceleration and deceleration when rising through the system of staggered packing particles. The observed bubbly-slug transition in the unpacked column depends on $U_{LS}$, since the bubble residence time decreases with the increased liquid velocity and bubbles do not have sufficient time to coalesce in the relatively short column. Slug regime in the packed column extends over a relatively narrow range of gas flow rates, compared with the unpacked column. As $U_{GS}$ rises, the packing breaks down the large slugs into small bubbles and the flow regime changes to churn regime. Annular regime occurs almost at the same gas velocities in both cases, unpacked and packed. In the packed column, the annular regime extents over higher $U_{LS}$ than that one in unpacked column.

However, at this point it has to be mentioned that the annular regime observed in the unpacked column is not the typical one, known from tubular columns. Due to the narrow gap and unstable wavy liquid film, the continuous liquid film on the facing wider walls is quite difficult to establish and maintain. The liquid films on the facing walls often interact resulting in unstable liquid structures or intermittent bridges extending over whole column area.

From the flow regime maps in Figure 4.2a and Figure 4.2b, is clear that the main impact of packing is in the substantial increase on the range of operation of the bubbly flow regime for the packed bed column. The packing acts distinctly as a flow organiser in the column.

### 4.2.3 Comparison with Existing Data

The flow regime maps obtained in the unpacked column were compared to existing data for air-water flow in narrow gaps. Figure 4.3 shows the transition zones obtained for the vertical upflow in 2 mm test section, compared with data reported by other authors. The transitions zones differ significantly from each other due to different definitions of flow
regimes and partially, to the different aspect ratio used in the experiments. Wilmarth and Ishii (1994) used a relatively narrow channel width of $1.5 \times 10^{-2}$ m, while Lowry and Kawaji (1988) used a broader channel of width $8.0 \times 10^{-2}$ m. Another important factor is the position for the observation of the flow regimes. In all cases, the column length was not sufficiently long to establish the flow structure that would be steady further downstream. In the present case, it was observed that especially at low $U_{LS}$, the flow structure was in the developing state and it was strongly forming further downstream. Therefore, the flow conditions and flow patterns present near the distributor might be completely different from those near the free surface. Here, the centre of the column was chosen as the observation point, representing a compromise between the flow regimes at the bottom and at the top of the column.

The bubbly-slug transition in quiescent liquid was observed approximately at $U_{GS} = 4.0 \times 10^{-2}$ m/s, a value lower than those observed by Mishima et al. (1993), and Wilmarth and Ishii (1994), which is $U_{GS} \approx 0.07$ m/s. On the other hand, Tzeng (1997) reported bubbly-slug transition at the even lower value of $U_{GS} = 1.0 \times 10^{-2}$ m/s. Further, the present observations show a stronger dependence of the bubbly-slug transition on $U_{GS}$ when compared with the published data. The slug-churn transition gives a good match with the experimental data for quiescent liquid but it differs significantly as the liquid rate increases. In the present case, slug regime was not detected for $U_{LS} > 0.1$ m/s which was not the case in the other data. The churn-annular transition in this work is located between the transitions observed by Lowry and Kawaji (1988) and Mishima et al. (1993).

The data for the packed column were compared with the work of Rong et al. (1993), the only work found dealing with similar kind of packing and flow conditions. Unfortunately, their flow pattern map is restricted to the zone of medium and high fluid velocities for $U_{GS} > 3.0 \times 10^{-1}$ m/s and $U_{LS} > 5.0 \times 10^{-2}$ m/s, and smaller packing particles ($D_p = 8.2 \times 10^{-3}$ m) were used. Despite these differences, relatively good match for the bubbly-slug transition was obtained in limited interval of fluid velocities. The churn-annular transition observed by Rong et al. (1993) is located at higher $U_{GS}$ values compared to the present transition if extrapolated to the range of measured velocities. This discrepancy can be again attributed to the uncertainty arising from visual observation and different definition of churn-annular transition.
4.2.4 Comparison with Transition Correlations

To predict flow regime transitions, various theories have been proposed for circular pipes without packing with diameter greater than $10^{-2}$ m. Widely used theories were proposed by Taitel et al. (1980), Mishima and Ishii (1984) for vertical flow and by Barnea (1983) for a wide range of pipe inclinations. For flat narrow channels, Lowry and Kawaji (1988)
found that the churn-annular transition differs substantially from that proposed by Taitel et al. (1980) and suggested new transition criteria based on the balance between frictional and interfacial forces. The summary of relevant transition criteria for vertical upward flow is given in Table 4.1.

As can be seen from Figure 4.5, the bubbly-slug transition is in good agreement with transition criteria established by Taitel (1980) and Mishima and Ishii (1984) although it exhibits a different slope. Good agreement was found in the case of the slug-churn transition with the transition proposed by Taitel (1980) due to similar definition of slug flow. For the purpose of comparison, the relative entrance length $l_e/D_H$ (see Table 4.1) used in Taitel’s model was set to 100 that approximately corresponds to the centre of the 2D column used in this work.

The churn-annular transition in this work was found for gas superficial velocities between 1.5 and 2 m/s that is a smaller value than the values given by the criteria proposed by Mishima and Ishii (1984), and Taitel (1980). On the other hand, the transition line of Lowry and Kawaji (1988) predicts this transition at smaller $U_G$ values since their model is based on different assumptions and it was specifically developed for the flow in narrow slots.

**Table 4.1  Flow regime transitions criteria used in vertical upflow.**

<table>
<thead>
<tr>
<th></th>
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<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Bubble-Slug</td>
<td>$U_{ls} = 3.0U_G - 1.15 \left( \frac{g \Delta \rho \sigma_{G-L}}{\rho_L} \right)^{0.25}$</td>
<td>$U_{ls} = \frac{3.33}{C_0} - 1 \frac{U_G}{C_0} - 0.076 \left( \frac{g \Delta \rho \sigma_{G-L}}{\rho_L} \right)^{0.25}$</td>
<td>—</td>
</tr>
<tr>
<td>Slug-Churn</td>
<td>$l_e/D_H = 40.6 \left( \frac{U_{MS}}{gD_H} + 0.22 \right) \alpha^{0.26}$</td>
<td>$1 - \alpha^{0.13} = \left[ \frac{(C_0 - 1)U_{ms} + 0.35 \sqrt{\Delta \rho g D_H} / \rho_L}{U_{ms} + 0.75 \sqrt{\Delta \rho g D_H} / \rho_L} \right]^{0.26}$</td>
<td>—</td>
</tr>
<tr>
<td>Churn-Annular</td>
<td>$U_{gs} \rho_G^{0.55} \rho_L^{0.25} = 3.1$</td>
<td>$U_{gs} = \left( \frac{g \Delta \rho \sigma_{G-L}}{\rho_L} \right) \left( \alpha - 0.1 \right)^{0.5}$</td>
<td>$\frac{(1 - \alpha) \mu}{\alpha^2} = \frac{2 - 2\alpha \mu}{\sigma_{G-L}} U_{ls}$</td>
</tr>
</tbody>
</table>

$C_0 = 1.35 - 0.35 \rho_G / \rho_L$; $U_{MS} = U_{ls} + U_G$; $D_H = 4 A / P_c$; $\Delta \rho = \rho_L - \rho_G$
Figure 4.5  Flow regime map comparison with: theoretical predictions of Mishima and Ishii (1984), Taitel et al. (1980), Lowry and Kawaji (1988).

4.3 Pressure Drop

The following subsections report the pressure drop data from single and two-phase flow measured by the pressure transducer technique. Further, the different sources of pressure drop are discussed and the indirect calculations of the gas volume fraction from the pressure drop are presented.

4.3.1 Single-Phase Flow in the Unpacked Column

Pressure drop of single-phase flow in tubes and rectangular channels is a relatively well studied subject and in the case of simple column geometry it can be directly predicted from the solution of the Navier-Stokes equations. It is presented here to provide a complete review of the experimental data and for further correlation with multiphase flow pressure drop. Generally, the total pressure drop can be divided into three main components: gravitational, accelerational and frictional, as follows

$$dp^T = dp^G + dp^A + dp^F$$  \hspace{1cm} (4.1)

In steady, adiabatic flows of constant cross-sectional area, the accelerational pressure drop approaches zero, if the flow field is well developed. The gravitational component is important for vertical and inclined columns and it can be easily expressed as the hydrostatic pressure gradient.
\[ \left( \frac{dp}{dx} \right)_L = \rho_L g \sin \theta \]  \hspace{1cm} (4.2)

where \( \theta \) is column inclination from the horizontal.

The usual way for prediction and comparison of the frictional pressure gradient in channels of any geometry is to correlate it in form of the Darcy’s formula

\[ \left( \frac{dp}{dx} \right)_L = f \frac{\rho U_{ls}^2}{2D_H} \]  \hspace{1cm} (4.3)

where and \( D_H \) is hydraulic diameter defined in standard way and \( f \) is friction factor. The latter one is in circular tubes given in terms of the Reynolds number as

\[
f = \begin{cases} 
64/Re & \text{for } Re \leq 2100 \\
0.3164/Re^{0.25} & \text{for } 2100 > Re > 100000 
\end{cases} \hspace{1cm} (4.4)
\]

In the rectangular channels, the friction factor can be approximated from the equation of laminar flow between two infinite parallel plates – Couette flow – but taking into account the finite width of the channel through the hydraulic diameter, \( D_H = 4A/P_w \), as

\[
f = \frac{24\mu_L D_H}{\rho_L \delta^2 U_{ls}} \hspace{1cm} (4.5)
\]

where \( \delta \) is channel thickness. However, this last relationship neglects the edge effect on pressure drop that can result to deviations between the predicted and measured pressure drop, mainly at small column geometric ratios, \( k_w = \delta/W \).

According to Troniewski and Ulbrich (1984), the single-phase friction factor in rectangular channels can be predicted by

\[
f = \begin{cases} 
64/Re_{DH}^* & \text{for } Re_{DH}^* \leq 2100 \\
0.3164 \left( Re_{DH}^* \right)^{0.25} & \text{for } Re_{DH}^* > 2100 
\end{cases} \hspace{1cm} (4.6)
\]

where

\[
Re_{DH}^* = \frac{U_{ls}^* D_H \rho_L}{\mu_L} \hspace{1cm} (4.7)
\]
and

\[ U_{LS}^* = U_{LS} \left( \frac{2}{k_w} \right)^{0.16} \]  \hspace{1cm} (4.8)

Figure 4.6 compares the experimental friction factor with the models introduced above. The experimental data are located between the lines corresponding to Couette flow (Equation 4.5) and the Troniewski and Ulbrich (1984) correlation, (Equation 4.6). The latter model was derived especially for rectangular columns with smaller aspect ratios, \( k_w < 20 \), where the edge effect plays an important role resulting in a higher pressure gradient. The Couette flow model does not account for the edge effect, therefore it slightly underestimates the pressure gradient. All experimental data in single-phase flow were collected for laminar flow with Reynolds number not exceeding 1000, due to restricted operating range of the liquid pump. Correlating the experimental data in this range of Reynolds numbers gives the following empirical formula for the friction factor

\[ f = 92R_{DH}^{-1} \hspace{0.5cm} (10 < R_{DH} < 1200) \]  \hspace{1cm} (4.9)

where the empirical parameter \( C_L = 92 \) is close to the value 94 obtained by Ali et al. (1993) for a gap thickness of \( 1.47 \times 10^{-3} \) m.
4.3.2 Single-Phase Flow in the Packed Column

The pressure drop of single-phase flow through packed bed has been extensively studied and detailed accounts may be found in literature (Ergun, 1952; Bemer and Kalis, 1978; Rong and Kawaji, 1993; Takatsu and Masuoka, 1998). Among commonly used correlation for the pressure drop in packed bed, belong the Chilton-Colburn and Ergun’s correlations (see Bird et al., 1976). These correlations were derived for 3D infinitely large columns filled with spherical packing. Ergun’s equation was derived for transient flow regime \(10 < Re_p < 1000\); where \(Re_p\) represents the Reynolds number based on particle diameter, \(D_p\), and it is usually given in form

\[
\frac{dp}{dx} = \frac{150\mu_L}{D_p^2} \left(1 - \varepsilon_p\right)^2 \frac{D_p}{\varepsilon_p} U_{LS} + \frac{1.75 \rho_L}{D_p} \left(1 - \varepsilon_p\right) \varepsilon_p U_{LS}^2
\]  

(4.10)

where \(\varepsilon_p\) is the packing porosity. In the present case, the packing particles are of cylindrical shape, thus the particle diameter; \(D_p\); in the Ergun’s equation should be replaced by the characteristic scale of the void space in the bed, which is the mean hydraulic diameter of the bed, defined as

\[
D_H = \frac{4A_c}{P_w}.
\]  

(4.11)

where \(A_c\) is inside the packed structure and \(P_w\) is the mean wetted perimeter. Since the packing structure is composed from the unit cells, as the one shown in Figure 4.7, the mean cross-sectional area can be calculated as the ratio of the void space volume of the unit cell and the height of the unit cell as

\[
A_c = \frac{\varepsilon_p V_{cell}}{D_p}
\]  

(4.12)

where \(V_{cell}\) is the unit cell volume. Similarly, the mean wetted perimeter can be obtained as the ratio of the wetted area and height of the unit cell as

\[
P_w = \frac{2\varepsilon_p V_{cell}/\delta + \pi D_p \delta}{D_p}
\]  

(4.13)
Inserting Equations 4.12 and Equation 4.13 into Equation 4.11, the following relationship for the mean hydraulic diameter can be obtained:

\[ D_H = D_p \left[ \frac{D_p}{2 \delta} + \frac{1 - \varepsilon_p}{\varepsilon_p} \right]^{-1} \]  (4.14)

The modified Erguns’ equation can be rewritten in the form of the friction factor defined by Equation 4.3, as

\[ f_p = \left( \frac{dp}{dx} \right)^2 \frac{2D_H^2}{\rho U_L^2} \frac{300}{Re_{DH}} \left( \frac{1 - \varepsilon_p}{\varepsilon_p} \right)^2 + 3.5 \frac{1 - \varepsilon_p}{\varepsilon_p} \]  (4.15)

where \( Re_{DH} = \rho_L U_L \varepsilon_p D_H / \mu_L \).

The experimental data was further compared with the empirical correlation of Knudsen and Katz (1958) for flow across staggered tube banks, which for laminar flow takes the form

\[ f_p = \frac{2400 \varepsilon_p \left( \frac{1 - \varepsilon_p}{\pi} \right)^{0.3}}{\pi} \frac{1}{Re_{DH}} \]  (4.16)

and for turbulent flow is given by
Single-phase friction factor values measured at two downstream positions as a function of $Re_{DH}$ is presented in Figure 4.8, and compared with the modified Erguns’ equation, and Equations 4.16 and 4.17. Erguns’ equation overestimates the friction factor and this may be attributed to the enhancement of turbulence by complicated 3D structures of the solid matrix in flow through packed beds.

Equations 4.16 and 4.17 for staggered tube banks provide a good match with the experimental data, although some discrepancies are noticeable. For example, the measured liquid friction factor in the column bottom section is slightly smaller compared to the one predicted by Knudsen and Katz (1958). This might be explained by the flow development near the column entrance. The experimental data indicates the transition to turbulent regime at, $Re_{DH} = 150$ that nearly agree with the cross point of empirical predictions of for laminar regime and turbulent regime suggested by Knudsen and Katz (1958).

![Graph showing experimental single-phase friction factor compared with the modified Erguns’ equation and equations for staggered bank of tubes (Knudsen and Katz, 1958).](image-url)

**Figure 4.8** Experimental single-phase friction factor compared with the modified Erguns’ equation and equations for staggered bank of tubes (Knudsen and Katz, 1958).
Rong and Kawaji (1993) correlated experimental data on pressure drop obtained for various particle size and porosities in the turbulent regime \((Re_{DH} > 400)\) and finned narrow channels by the empirical relationship

\[
f_p = C_f \text{Re}_D^{1-m}
\]  

(4.18)

where \(C_f\) and \(m\) are correlation parameters.

The present data was correlated in a similar manner and the results both for the liquid and for the gas phase are illustrated in Figure 4.9. The comparison of the parameters \(C_f\) and \(m\) obtained by Rong and Kawaji (1993) and those obtained for the present data is shown in Table 4.2. The value of the correlation slope \(m\) is in good accordance with the one obtained Rong and Kawaji (1993), however, the parameter \(C_f\) is significantly larger than the ones reported by Rong and Kawaji (1993). This discrepancy may be due to the smaller gap thickness of the present column or by a different definition of \(Re_{DH}\), since Rong and Kawaji (1993) did not clearly specify whether the liquid superficial or the liquid interstitial velocity was used in their correlation. The small differences between correlation parameters obtained at the bottom and the top column section can be attributed to the flow development near the column entrance.

\[\text{Correlations}
\]

\[
\begin{align*}
\text{Liquid (bottom): } f_p &= 6.92 \text{Re}_{DH}^{-0.314} \\
\text{Liquid (top): } f_p &= 6.27 \text{Re}_{DH}^{-0.274} \\
\text{Gas (top): } f_p &= 6.09 \text{Re}_{DH}^{-0.219}
\end{align*}
\]

Figure 4.9  Single-phase friction factor in turbulent regime correlated with Reynolds number as suggested by Rong and Kawaji (1993).
Table 4.2  Comparison of coefficients used to correlate friction factor in packed column according Equation 4.18.

<table>
<thead>
<tr>
<th>Source</th>
<th>(D_r \times 10^3) [m]</th>
<th>(\delta \times 10^3) [m]</th>
<th>(\varepsilon_r)</th>
<th>(C_f)</th>
<th>(m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Rong and Kawaji (1993)</td>
<td>8.2</td>
<td>2.6</td>
<td>0.47</td>
<td>2.432</td>
<td>0.249</td>
</tr>
<tr>
<td></td>
<td>6.1</td>
<td>2.6</td>
<td>0.49</td>
<td>4.487</td>
<td>0.253</td>
</tr>
<tr>
<td></td>
<td>3.2</td>
<td>2.3</td>
<td>0.67</td>
<td>4.805</td>
<td>0.320</td>
</tr>
<tr>
<td>Present Work</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Liquid phase (bottom)</td>
<td>10</td>
<td>2.0</td>
<td>0.60</td>
<td>6.93</td>
<td>0.314</td>
</tr>
<tr>
<td>Liquid phase (top)</td>
<td>10</td>
<td>2.0</td>
<td>0.60</td>
<td>6.27</td>
<td>0.274</td>
</tr>
<tr>
<td>Gas phase</td>
<td>10</td>
<td>2.0</td>
<td>0.60</td>
<td>6.09</td>
<td>0.219</td>
</tr>
</tbody>
</table>

4.3.3 Two-Phase Pressure Drop

It is known that the pressure drop in two-phase flow is larger than that in a single-phase flow. While the introduction of a second phase results in a reduction of the flow area of both phases, this alone cannot account for the dramatic increase of pressure drop. The difference in phase densities creates an additional body force that results in different phase velocities, non-homogeneous phase distribution and an increased shear rate in the denser phase. Consequently, a large loss of energy occurs, especially in the continuous phase and the contribution of the accelerational pressure drop on the total pressure rises proportionally to the level of flow heterogeneity.

As in the case of single-phase flow, the total pressure gradient can be split into frictional, accelerational and gravitational contributions. However, the expression for these contributions become more complicated due to the presence of the second phase that causes additional sources of the accelerational pressure drop if the gas volume fraction changes downstream or if the flow is not well-developed. If the average gas volume fraction is known, the gravitational pressure gradient can be calculated as follows

\[
\left(\frac{dp}{dx}\right)_g = \left[(1-\alpha)\rho_L + \alpha\rho_g\right]g \sin \theta
\] (4.19)

The measured total pressure drop as a function of gas superficial velocity is presented in Figure 4.10a and Figure 4.10b for the packed and unpacked columns, respectively, with \(U_{LS}\) as a parameter. In the case of the packed column, the frictional component of the
pressure gradient is a square function of the liquid superficial velocity, and it strongly contributes to the total pressure gradient. Therefore, a change of the curve trend from descending to ascending is observed as $U_{LS}$ increases.

Figure 4.10 Two-phase flow total pressure gradient vs. $U_{GS}$ at different liquid velocities for: a) unpacked column; b) packed column.
The two-phase friction pressure drop is given by the difference between the total pressure drop and the sum of gravitational and accelerational components, and is commonly correlated by Lockhard-Martinelli correlation

\[
\left( \frac{dp}{dx} \right)_T = \Phi_L \left( \frac{dp}{dx} \right)_L = \Phi_G \left( \frac{dp}{dx} \right)_G
\]  

(4.20)

where \( \Phi_L \) and \( \Phi_G \) are the two-phase friction multipliers that are correlated by means of the Lockhard-Martinelli parameter, \( \chi \), which is defined as the ratio of the frictional pressure drop of liquid to that of gas, when each phase is assumed to flow alone in the channel

\[
\chi = \frac{\left( \frac{dp}{dx} \right)_L}{\left( \frac{dp}{dx} \right)_G}
\]  

(4.21)

An useful correlation to predict the friction multiplier \( \Phi_L \) is that proposed by Chisholm and Laird (1958) based upon the assumption of separated flow model (see Wallis, 1969)

\[
\Phi_L^2 = \left[ 1 + \frac{1}{\chi^2} \right]^{\frac{2-m}{2}}
\]  

(4.22)

where \( m \) is the same parameter as in Equation 4.18. For laminar flow, \( m = 1 \), one can obtain

\[
\Phi_L^2 = 1 + \left( \frac{2}{\chi^2} \right) + \left( \frac{1}{\chi^2} \right)
\]  

(4.23)

Since the assumption of the separated flow is non-adequate in most cases, Equation 4.23 is for correlation purposes usually replaced by

\[
\Phi_L^2 = 1 + \left( C/\chi \right) + \left( 1/\chi^2 \right)
\]  

(4.24)

where \( C \) is the correlation parameter obtained by the fitting of experimental data

To correlate pressure drop data in this way, information about the two-phase frictional drop is needed. This can be calculated from the measured total pressure drop by subtracting the gravitational component, providing that accelerational one is negligible. This assumption can be justified only in the case of the unpacked column, since the cross-sectional area is constant and bubbles follow a nearly rectilinear path, especially at low gas
volume fraction. In the case of unpacked flow, gas volume fraction data obtained from imaging was used to calculate the gravitational pressure drop. The Lockhart-Martinelli correlation for unpacked column is plotted in Figure 4.11a. The parameter $C$ depends on the hydraulic diameter and fluids velocity (Mishima et al., 1993). As can be seen, all data points fall within the interval marked by Equation 4.23 with the coefficient $C$ between 15 to 40. Despite of the large interval of $C$, this result is in good agreement with the data of Ali et al. (1993) for a channel gap of $1.47 \times 10^{-3}$ m, who correlated their data by the parameter $C$ ranging from 10 to 20. Sadatomi et al. (1982) reported $C = 21$ for different noncircular columns. The relatively large scatter exhibited by the present data can be caused by covering of a relatively broad spectrum of flow regimes and by possible errors caused by the measurement of gas volume fraction using the imaging technique.

In flow with packing, the accelerational component of pressure drop cannot be neglected and it is rather difficult to make any assumptions about its amplitude. Despite this fact, the data from the packed column was correlated in a similar manner as in the previous case to enable comparison with the results of Rong and Kawaji (1993). For the measured interval of gas and liquid superficial velocities, all data points shown in Figure 4.11b are bounded by the curves given by Equation 4.23 with $C = 5$ and $C = 30$. This is in good agreement with Rong and Kawaji (1993) who fitted their data by $5 < C < 50$. However, it is obvious that the experimental data does not follow the same trend as Equation 4.23 and some adjustment to the Chisholm and Laird (1958) correlation is needed for better representation. As mentioned before, Equation 4.23 is based upon the simplification of separated flow in a pipe and negligible interactions between the two phases. This assumption is rather difficult to apply in the case of the flow with packing since the phases are well mixed and both the wall interactions and the interfacial interactions are of increased level, compared to the flow in a smooth channel. The fitting curve that more closely correlates the present data trend was found to be

\[
\phi_L^2 = \left[1 + \left(\frac{C}{\chi}\right)^{1/16}\right]^{(4-m)/2}
\]  

(4.25)

where $m = 0.29$ was taken from the single-phase flow correlation for the liquid phase. The parameter $C$ was tuned to value 1.5. However, it has to be noted that the above correlation is merely empirical and without any physical background.
Figure 4.11  Lockhard-Martinelli correlation of pressure drop in: a) unpacked and; b) packed column.
4.4 Gas Volume Fraction

The gas volume fraction, $\alpha$, or gas hold-up is a fundamental quantity in the description and analysis of multiphase flow as it affects the flow regime, pressure drop, fluid residence time and heat and mass transfer characteristics. Generally, its magnitude and its distribution are influenced mainly by the gas and the liquid flow rates, but also by the fluid properties and the column geometry. Several methods have been used to measure either the local or the channel-averaged volume fraction. One of the simplest techniques to determine $\alpha$ is the quick-closing valve method (Nickin and Davidson, 1961). Lowry and Kawaji (1988) and Ali et al. (1993) measured the electrical conductivity of a two-phase mixture between pairs of electrodes in the channel walls to obtain both chord-averaged and channel averaged data. An impedance probe used by Costigan and Whalley (1996) was characterised by high sampling rate that allowed determination of flow regimes by PDF of gas volume fraction. A glass fibre probe was used by Groen et al. (1995) to measure local values of $\alpha$ and to characterise liquid circulation structures in a bubble column. Mishima et al. (1993) used both image processing and neutron radiography to obtain 2D void distribution and channel averaged volume fraction data. Hibiki et al. (1995) used neutron radiography to measure $\alpha$ in opaque metallic rectangular ducts.

The local instantaneous gas volume fraction, $\alpha$, is defined as the fraction of the volume element that is occupied at any instant of time with the gas phase. Then, according to Wallis (1969), space-averaged gas volume fraction is defined as

$$\bar{\alpha} = \frac{\iiint \alpha(r,t) \, dr \, dt}{\iiint \, dr \, dt}$$

The simplest approach to estimate gas volume fraction is to assume that the flow is homogeneous, that is, both phases flow at the same velocity. Under these circumstances, the space-averaged gas volume fraction equals the volumetric gas quality

$$\bar{\alpha} = \frac{Q_G}{Q_M}$$

where $Q_G$ and $Q_M$ are the gas and the mixture flow rates, respectively.
However, the space-averaged gas volume fraction is generally smaller than the volumetric quality due to higher velocity of the gas relative to the liquid phases. Several approaches have been suggested to predict the space-averaged gas volume fraction with accounting for the different phase velocities, including the drift-flux model, incorporation of a slip ratio or direct correlations of $\alpha$ (see Wallis, 1969). The drift-flux model is widely used to correlate $\bar{\alpha}$ for bubbly and slug regimes (Zuber and Findlay, 1965; Wallis, 1969). In this model the average gas velocity is given by

$$\bar{U}_G = \frac{U_{GS}}{\bar{\alpha}} = C_0 U_{MS} + U_{G, drift}$$ (4.28)

where $U_{MS}$ is the mixture volumetric flux equal to $U_{GS} + U_{LS}$, $C_0$ is the distribution parameter, and $U_{G, drift}$ is the mean drift velocity. The distribution parameter accounts for the differences in velocity and the gas volume fraction profiles in along the cross sectional area of flow. For circular columns, its value is about 1.2 (Nicklin et al., 1962) but in narrow rectangular channels, it is usually given by the Ishii (1977) correlation in the form

$$C_0 = 1.35 - 0.35 \sqrt{\rho_G / \rho_L}$$ (4.29)

The mean drift velocity $U_{G, drift}$ represents the difference between the gas velocity and the mixture mean velocity and it is a function of terminal bubble rise velocity in stagnant liquid.

The drift-flux model was also used here to correlate the experimental gas volume fraction obtained by the image processing technique. By averaging of gas volume fraction over the whole column width at given axial positions, the average gas velocity, $\bar{U}_G$, can be calculated as in Equation 4.28. The calculated gas velocity is then plotted versus known mixture volumetric flux, as shown in Figure 4.12 for the unpacked column and in Figure 4.13 for the packed column. As can be seen, the data from the slug, churn and annular regimes can be well correlated by drift flux model. In the bubbly regime, no linear dependency of the gas velocity on the mixture velocity was observed. This was confirmed for the column both with and without packing where the non-linear dependency was observed even in the slug regime. The main reason of these non-linearities in the bubble regime is explained by the presence of relatively small bubbles whose drift velocity depends strongly on their size and on the presence of other bubbles. This does not happen...
in the slug regime where larger bubbles occur and their drift velocity is mainly dependent on the column size. Figure 4.12 reveals another interesting fact; the slope of correlation lines, $C_0$, increases with the downstream position. As mentioned above, the magnitude of distribution parameter, $C_0$, indicates the level of the flow uniformity in the lateral direction and its rise downstream indicates an increased heterogeneity of the flow in the upper column section. This was also confirmed visually, where an increased number of large bubbles, concentrated near the column centreline, was observed in the upper section of the column.

**Figure 4.12**  Gas velocity vs. mixture velocity at different downstream positions in the unpacked column. Drift-flux model correlation for the slug and churn regime.

**Figure 4.13**  Gas velocity vs. mixture velocity in the packed column. Drift-flux model correlation for the churn regime.
The distribution parameter in the packed column, does not vary with the axial position, thus, only one correlation line is present in Figure 4.13. Its invariability indirectly confirms the higher uniformity of the phase distribution in the packed column.

Table 4.3 gives a comparison of current distribution parameters with existing data (Iida and Takahashi, 1976; Jones and Zuber, 1979; Sadatomi et al., 1982; Mishima et al., 1988; Moriyama and Inoue, 1991) measured in rectangular ducts of various aspect ratios. The mean value of $C_0$ from the unpacked column is in good agreement with data of Mishima et al. (1988) and Iida and Takahashi (1976). From Table 4.3 is obvious that despite the use of different columns, the discrepancies in the distribution parameters are of the order 20% that confirms the good practical applicability of the drift-flux model to correlate gas volume fraction data.

### Table 4.3  Comparison of distribution parameters obtained by various authors.

<table>
<thead>
<tr>
<th>Source</th>
<th>$\delta \times 10^3$ [m]</th>
<th>$W \times 10^3$ [m]</th>
<th>$C_0$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Moriyama and Inoue(1991)</td>
<td>0.5</td>
<td>30</td>
<td>1.4</td>
</tr>
<tr>
<td>Iida and Takahashi (1976)</td>
<td>0.7</td>
<td>40</td>
<td>1.35</td>
</tr>
<tr>
<td>Jones and Zuber (1979)</td>
<td>5</td>
<td>63.5</td>
<td>1.2</td>
</tr>
<tr>
<td>Mishima et al. (1988)</td>
<td>1</td>
<td>40</td>
<td>1.35</td>
</tr>
<tr>
<td>Sadatomi et al. (1982)</td>
<td>7</td>
<td>20.6</td>
<td>1.23</td>
</tr>
<tr>
<td>Sadatomi et al. (1982)</td>
<td>7</td>
<td>50</td>
<td>1.16</td>
</tr>
<tr>
<td>Present work (without packing)</td>
<td>2</td>
<td>200</td>
<td>1.34</td>
</tr>
<tr>
<td>Present work (with packing)</td>
<td>2</td>
<td>200</td>
<td>1.22</td>
</tr>
</tbody>
</table>

### 4.4.1  Comparison of Imaging and Pressure Drop Measurements

The averaged gas volume fraction for the unpacked column is plotted in Figure 4.14 as a function of the gas superficial velocity for different values of $U_{LS}$. As expected, the averaged gas volume fraction is an increasing function of $U_{GS}$ and it decreases as the liquid flux rises. Results from two different techniques, pressure transducers and imaging, are compared. It can be noticed that the results from both techniques are in a good agreement only at low gas velocities in the bubbly and slug regimes. At higher values of
that correspond to the churn regime (approximately at $U_{LS} > 0.5 \text{ m/s}$), the discrepancies between the data measured by pressure transducers and by imaging increase, mainly due to the more pronounced effect of the accelerational pressure drop that was assumed being negligible. Therefore, it can be concluded that the gas volume fraction determined by pressure measurements provides reliable results only in bubbly and slug regimes.

Similar indirect determination of the gas volume fraction was performed for the flow in the packed column. However, large discrepancies between the calculated and directly measured gas volume fraction were found. These can be attributed mainly to the increased accelerational pressure drop component and to the non-linear pressure drop dependency on $U_{LS}$. Therefore, this method cannot be used for practical purposes.

![Comparison of space-averaged gas volume fraction obtained by imaging, IP, (void symbols) and pressure drop measurements, PD, (filled symbols) in unpacked column.](image)

### 4.4.2 Lateral Distribution of Gas Volume Fraction

Gas volume fraction profiles obtained by the standard imaging and the BIV technique are illustrated and compared in Figure 4.15a and Figure 4.15b for the unpacked and packed case, respectively. The lateral profiles at three different downstream positions ($X = 0.5; 3.0; 5.5$) are compared for the same gas rate, $U_{GS} = 0.01 \text{ m/s}$, and stagnant liquid conditions, $U_{GS} = 0 \text{ m/s}$. From these plots, it is quite clear that the gas phase has a tendency to concentrate in the column centre, but this process is reduced in the packed column. Due
to lower bubble rise velocities in the packed column, which will be shown later, the magnitude of $\alpha$ is higher in the packed column at the same flow conditions. Calculated uncertainty intervals for the BIV technique are smaller than in the case of imaging due to the higher number of frames used for the image analysis and due to a larger image size used at BIV. Further, the uncertainty intervals are significantly larger in the column centre that can be explained by the presence of larger bubbles resulting in higher flow heterogeneity and thus higher scatter of gas volume fraction data. Both techniques provide relatively similar results, where all discrepancies are within the estimated uncertainty intervals.

Figure 4.15  Comparison of gas volume fraction profiles for: a) unpacked and; b) packed column. Flow conditions: $U_{LS} = 0.0$ m/s; $U_{GS} = 0.011$ m/s.
The above mentioned trends were consistent over the entire range of gas fluxes used in the experiments that is documented by Figure 4.16 obtained at $U_{gs} = 0.054 \text{ m/s}$ in stagnant liquid corresponding to fully developed slug regime.

Figure 4.17 represents the lateral $\alpha$ profiles measured in the flowing liquid conditions and at the same $U_{gs}$ as in Figure 4.15. In contrary to Figure 4.15, the data presented in Figure 4.17 correspond to the bubble regime (see Figure 4.2). It was observed that bubbles migration towards the column centre is reduced and bubbles rise along almost rectilinear paths. In these flow conditions, the lateral bubble motion is partially controlled by the liquid lateral shear that pushes bubble towards the lateral walls.

**Figure 4.16** Comparison of gas volume fraction for: a) unpacked and; b) packed column. Flow conditions: $U_{ls} = 0.0 \text{ m/s}$; $U_{gs} = 0.054 \text{ m/s}$. 
Consequently, the effect of the inward lift forces, *Bernoulli like force*, *wall-lubrication force*, *aerodynamic-like force*, is reduced. Detailed explanation of these forces will be given in Chapter 6. One can see that the gas volume fraction profiles are relatively non-uniform and scattered due to the above-mentioned non-uniformity of the gas distribution and due to decreased lateral bubble mixing at high $U_{LS}$. Hence, at low gas rate, the $\alpha$ profiles depend strongly on the location of the active nozzles and their effect is evident even at the column top, as illustrated in Figure 4.17.

*Figure 4.17*  Comparison of gas volume fraction for a) unpacked and; b) packed column. Flow conditions: $U_{LS} = 0.193$ m/s; $U_{GS} = 0.054$ m/s.
4.4.3 Effect of Surfactant on Gas Volume Fraction

As already mentioned in the Chapter 1, the character of gas-liquid dispersion is considerably influenced by the fluid physical properties, where surface tension is of high importance. An even more important parameter than the absolute value of surface tension is the value of its derivative $\frac{d\sigma_{g-l}}{da}$, where $a$ is the surfactant activity. This was confirmed experimentally by Marrucci and Nicodemo (1967) and Prince and Blanch (1990a) who studied different aqueous solutions of inorganic salts. The measured surface tension of these solutions was almost unaffected by the presence of salt or it was slightly higher than that for pure water. Nevertheless, the bubble size distribution was significantly altered and strong suppression of bubble coalescence was observed above certain salt concentration. Moreover, the ion valency influences the minimal salt concentration at which complete coalescence suppression was observed. According to Marrucci (1969), the coalescence efficiency depends on the draining rate of liquid bridge between two sufficiently close bubbles. Since, the surface-active agents have the tendency to concentrate on the gas-liquid interface, electrically active layers can be created on the liquid bridge sides with a repulsive effect on the approaching bubbles and thus slowing down the coalescence process.

During the testing of the BIV-LIF technique, a fluorescent dye – Rhodamine B – was added to the water in the concentration of 2 ppm. The addition of this dye resulted in significant change of the two-phase flow hydrodynamic parameters, such as bubble size, bubble velocity, and gas volume fraction in spite of the fact that the interfacial tension was maintained constant. This remarkable difference in the gas volume fraction profiles can be seen from Figure 4.18, where the comparison between the systems with and without Rhodamine B was performed. In stagnant liquid conditions, the system with surfactant exhibits remarkably higher and more uniformly distributed gas volume fraction than the system without surfactant. However, these differences diminish as the liquid flux increases.

In packed columns and flowing liquid conditions (Figure 4.19b), an interesting fact was observed. The measured gas volume fraction was smaller than that one for pure water system. This can be only explained by an enhanced turbulence intensity that increases the bubble break-up rate in the system with pure water and produces a large number of small
bubbles of about $1 \times 10^{-3}$ m diameter. On the other hand, as it will be shown in the next chapter, in the packed column with the addition of surfactant, the bubble size distribution is little affected by the liquid superficial velocity. Therefore, there is an indirect indication that the addition of surface-active component suppresses both the coalescence and the break-up rates.

Figure 4.18  Effect of Rhodamine B addition to water on $\alpha$ lateral profiles in: a) unpacked and; b) packed column. Flow conditions: $U_{LS} = 0.0$ m/s; $U_{GS} = 0.054$ m/s.
Figure 4.19  Effect of Rhodamine B addition to water on $\alpha$ in: a) unpacked and; b) packed column. Flow conditions: $U_{LS} = 0.063 \text{ m/s}; U_{GS} = 0.054 \text{ m/s}$.

4.5 Interfacial Area Concentration

The specific interfacial area in two-phase flow is one of the most important parameters of bubble columns. Except for the case of extremely slow reactions, its value determines the rate of the overall mass and energy transfer in two-phase flow. The interfacial area concentration is defined as the local interfacial area, $A_i$ per unit of the volume occupied by the gas-liquid mixture, $V_M$, that is
Since the interfacial area in 3D columns cannot be measured directly, it is indirectly estimated from the mean gas volume fraction and the mean bubble diameter. If only spherical bubbles with diameter \( d_B \), are assumed to be present in column, the interfacial area can be estimated from the relation:

\[
a_i = \frac{6\alpha}{d_B}
\]  

At present, several other methods are available to measure interfacial area: video analysis (Wilmarth and Ishii, 1997), light attenuation (Mishima et al., 1988), ultrasonic, photoelectric capillary and chemical adsorption (Desejus and Kawaji, 1990). Chemical adsorption of the gas active component into the liquid phase seems to provide most reliable results and it is applicable for all flow regimes. On the other hand, it is necessary always strictly to distinguish between the coalescence-suppressing and coalescence-promoting systems. When the column is transparent, the most popular technique is to use photographic or video method. In the present case, video analysis was used based on the assumption that bubbles are thin 2D disks defined by their perimeter and column gap thickness, as was explained in Chapter 3 (Section 3.5.2).

Figure 4.20 shows the measured space-averaged interfacial area concentration, \( a_i \), over the upper-half part of the column (3.0 < \( X \) < 5.0) plotted against the averaged gas volume fraction values, \( \alpha \), for all flow conditions and flow configurations. As can be seen, the presence of packing results in higher \( a_i \) for all flow regimes. The observed dependence on the liquid rate for the packed case may be explained by the increase in the bubble break-up rate caused by higher turbulence levels generated by the liquid motion. There is clearly a remarkably simple correlation between \( a_i \) and \( \alpha \) that is only dependent on the flow configuration, unpacked/packed, and only in the packed case is dependent on the liquid flow rate. Figure 4.21 shows the same correlation when 2 ppm of Rhodamine B were added to the liquid phase, e.g. tap water. One can see that the situation changes radically, and about 30% increase of the interfacial area concentration is observed at the same gas volume fractions as in previous case. This can be attributed to the smaller coalescence rate resulting in smaller bubble size and larger interfacial area. For the unpacked column,
Mishima et al. (1988) correlated the interfacial area concentration with gas volume fraction as

\[ a_i = A \alpha^B \]  

(4.32)

where \( A \) and \( B \) are empirical constants.

**Figure 4.20** Interfacial area concentration data versus gas volume fraction data for packed and unpacked column with tap water-air system.

**Figure 4.21** Interfacial area concentration data versus gas volume fraction data for packed and unpacked column with addition of 2 ppm Rhodamine B.
The comparison of the empirical constants $A$ and $B$ with the results obtained by Mishima et al. (1988) is given in Table 4.4 for each flow regime and flow configuration. One can see that linearity of the correlation decays in the order bubbly, slug, churn regime. Further, in the system with Rhodamine B, the parameter $A$ increased approximately between 20% and 30% compared to the system without surfactant.

<table>
<thead>
<tr>
<th></th>
<th>Bubbly</th>
<th>Slug</th>
<th>Churn</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$A$</td>
<td>$B$</td>
<td>$A$</td>
</tr>
<tr>
<td>Mishima et al. (1988)</td>
<td>-</td>
<td>-</td>
<td>9.52</td>
</tr>
<tr>
<td>Present Data - without Packing</td>
<td>14.2</td>
<td>0.98</td>
<td>11.9</td>
</tr>
<tr>
<td>Present Data - with Packing, Stagnant Liquid</td>
<td>14.8</td>
<td>0.90</td>
<td>14.9</td>
</tr>
<tr>
<td>Present Data - with Packing, Moving Liquid</td>
<td>16.8</td>
<td>0.92</td>
<td>17.8</td>
</tr>
<tr>
<td>Present Data - without Packing, 2 ppm of Rhodamine B</td>
<td> </td>
<td> </td>
<td> </td>
</tr>
<tr>
<td>Present Data - with Packing, 2 ppm of Rhodamine B</td>
<td> </td>
<td> </td>
<td> </td>
</tr>
</tbody>
</table>

For all regimes: $A = 14.3$; $B = 0.83$

For all regimes: $A = 22.8$; $B = 0.94$

4.6 Conclusions

In this chapter, experimental data concerning the flow regimes transitions, pressure drop, gas volume fraction and interfacial from the 2D column were reported. The flow regimes maps for the unpacked and packed column were constructed and compared. The flow regime map for unpacked column was compared with published experimental flow regime transitions and with theoretical predictions. Relatively large discrepancies were found not only between the present data and other experimental studies but also between published data themselves. This was attributed to the differences in flow regime definitions and relatively subjective evaluation of the flow regime transitions. It
was found that the presence of packing increases the range of gas flow rates at which bubbly flow occurs.

The pressure gradient in single- and two-phase flow was reported both for unpacked and packed column. The data obtained from single-phase flow showed good agreement with available theoretical predictions. Two-phase pressure drop was correlated by the Lockhard-Martinelli correlation and it was compared with published data. In the case of the unpacked column, it was shown that the Chisholm-Laird correlation was capable to correlate pressure gradient data but for the packed column, it did not follow accurately the data and some modifications are needed.

The gas volume fraction was correlated by the drift flux model for the data from slug, churn and annular regimes. From the measured dynamics of pressure fluctuations, the indirectly measured gas volume fraction showed good agreement with the directly measured data by imaging technique only in the bubbly and slug regimes. Plotting the gas volume fraction profiles at different downstream locations revealed a strong migration of the gas phase towards the column centre. This tendency was smaller in the packed column where also a higher magnitude of gas volume fraction was observed at the same flow conditions. The addition of the Rhodamine B to the liquid phase resulted in an increased uniformity of phase distribution. Moreover, a higher magnitude of gas volume fraction was observed except in the case of the packed reactor with flowing liquid conditions.

The relationship between gas volume fraction and interfacial area has been investigated for various flow regimes. It has been shown that packing substantially increases the interfacial area and this relationship is dependent on liquid flow rates. Further, the addition of surfactant resulted in the increased interfacial area in both unpacked and packed column.