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**Department of Chemical and Environmental
Engineering**

**Gas-High Viscosity Oil Flow in Vertical
Large Diameter Pipes**

by

Shara Kamal Mohammed

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Abstract

Gas flow in columns of high viscosity liquids is found in heavy oil and bitumen production, polymer manufacturing and in Volcanology as silicate magmas in the volcanic conduit. Predicting the characteristics of the hydrodynamics of gas flow, under such conditions is essential in both design and safety assessments. In oil and gas industry, it is very important in the design of the equipment and the very long pipelines. Further, it is important in the design and safety of the industrial equipment and the pipelines in the polymer manufacturing. Finally, the ability to predict natural phenomenon and develop the knowledge about volcanoes activity and the nature of eruptions in volcanoes is important for the assessment of environmental risks.

The majority of the works, which have studied gas-liquid flow in pipes, have been carried out mainly by using water or liquids of low viscosities (<1 Pa.s). Knowledge, regarding the hydrodynamics of gas flow in high viscosity liquids and large diameter columns, is still limited though despite the importance of this subject. In this work, the characteristics of gas-high viscosity oils in large diameter columns were studied over a wide range of gas flow rates. Two column geometries were used: one of 240 mm and the other of 290 mm internal diameter. The columns were initially filled with stagnant Silicone oil of viscosities 360 and 330 Pa.s respectively. Electrical Capacitance Tomography (ECT) technique was employed for the data measurements besides a high-resolution camera.

In general, 4 flow patterns are found over the selected range of gas flow rates. These are seen to differ from the ones in lower viscosity liquids. First, bubbly flow consists of single spherical bubbles that rise at a constant velocity in the centre of the column at low gas flow rates. Second, slug flow

Abstract

which consists of long bubbles (Taylor bubble) with rounded top and end, with a diameter almost equal to the column diameter and separated by liquid slugs. The third flow pattern is the transition to churn flow which occurs due to further increase of gas flow rate. The falling film, around the very long bubbles, accumulates to create regions of high frequency activity that consists of liquid bridges. These regions are named "churn" regions. The length of this region increases gradually with increasing the gas flow rate. At very high gas flow rates, the gas flows through an open wavy non-symmetrical core in the column of the viscous oil (churn regions). The length of the churn regions increases significantly at this flow regime which can be named churn flow regime. Small bubbles of millimetres to centimetres diameter are seen to accumulate in the column due to the high viscosity and low velocity of the liquid motion. These bubbles generate due to the bubble eruption at the top section, bubble coalescence along the column and at the gas injection points at the bottom of the column.

Dedications

I dedicate this thesis to

The Peshmarga and the martyrs of Kurdistan.

My mother, my father

My sister, my brother

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Notations

Symbol	Description	Unit
S	Bubble diameter basing on surface tension	m
C_{d*}	Average drag coefficient	
C_o	Ratio of the centreline of average velocity	
D	Pipe diameter	
d_B	Bubble diameter	m
d_h	Gas nozzle diameter	m
E_o	Eötvös number	
f	Frequency	s^{-1}
F_D	Factor relating the amount of the liquid falling film	
Fr	Froude number	
g	Gravitational acceleration	m/s^2
H_{liquid}	Initial height of the liquid	m
H_{mix}	Height of the gas-liquid mixture	m
L	Bubble diameter basing on viscous drag force	m
L_S	Liquid slug length	M
L_{TB}	Taylor bubble length	m
L_u	Unit slug length	m
R	Buoyancy Reynolds number	
R_p	Pipe radius	m
T	Bubble diameter basing on liquid inertia force	m
U_{TB}	Rise velocity of Taylor bubble	m/s
U_{sb}	Rise velocity of small bubbles	m/s
U_g^*	Dimensionless gas superficial velocity	
U_{gs}	Gas superficial velocity	m/s

U_l	Liquid superficial velocity	m/s
U_{st}	Structure valocity	m/s
U_T	Terminal velocity of bubbles in stagnant liquid	m/s
V^*	volumetric gas flow rate	m ³ /s
V_B	Bubble volume	m ³
V_g	Volume fraction of gas	
V_l	Volume fraction of liquid	

Greek Symbols

α	Dimensionless inertial parameter	
α_s	Void fraction of liquid slug (from PDF)	
α_{TB}	Void fraction of Taylor bubble (from PDF)	
δ	Film thickness	m
ϵ_g	Void fraction (gas holdup)	
$\epsilon_{g(avg)}$	Overall void fraction	
ϵ_l	Liquid holdup	
μ	Dynamic viscosity	Pa.s
μ_e	Effective suspension viscosity	Pa.s
ν	Liquid kinematic viscosity	m ² /s
P_{ave}	effective average suspension density	
ρ_g	Gas density	Kg/m ³
ρ_l	Liquid density	Kg/m ³
σ	Surface tension of the liquid	N/m

Abbreviations

ECT	Electrical Capacitance Tomography
PDF	Probability Density Function

Chapter 1

Introduction

1.1 Introduction

When gas and liquid flow in vertical columns the distribution of the phases take up different forms. These structures of the flow depend on the flow rates as well as the physical properties of both fluids, the methods of mixing of the fluids, the pipe geometry, and the degree of inclination. The hydrodynamics of gas flowing in high viscosity liquids and large diameter columns is very complex. It is associated with variation of the structure velocity and pressure distributions along the column. The structure becomes more complicated with increasing the gas flow rate when the transition to churn flow occurs. The flow patterns of gas-high viscosity liquids vary significantly compared with the two-phase flow structure with water and other low viscosity liquids. Bubbly flow, for instance, in water or low viscosity liquids, consists of non-uniform bubbles which vary in size, shape, and concentration, and rise with spiral or zigzag trajectory; while it consists of one large spherical bubble rise at the centre of the column in high viscosity liquids. The characteristics of slug flow also change with increasing liquid viscosity. Slug flow occurs when coalescence begins between the bubbles, with increasing the gas flow rate, to create larger bubbles. The diameter of these large bubbles is approximately equal to the pipe diameter. They rise in a bullet shape with a flat tail and turbulent bubbly wake and these are called Taylor bubble. The area beyond each Taylor bubble occupied by the turbulent small bubbles is called liquid slug. With increasing the liquid viscosity, the tail of the Taylor bubbles becomes rounded shape while the top of the bubble

remains rounded. The variation of the structure of Taylor bubble in different liquid viscosities was studied for the first time by Viana et al. (2003). They employed liquids of viscosities ranged between 0.001-3.9 Pa.s. The liquid slug beyond Taylor bubbles is also affected by changing the liquid viscosity. The small bubbles in the liquids' slug become less turbulent with increasing the liquid viscosity. In liquids of very high viscosity, they rise in straight lines and constant velocity within the liquid slug and move downward within the falling film around Taylor bubbles. On the other hand, churn flow in low viscosity liquids is recognised as churning or oscillatory motion of the gas-liquid mixture with a net upward flow. Unstable flow structure occurs at high gas flow rates due to the breaking down of the Taylor bubbles. This flow pattern has been studied by many researchers employing water and low viscosity liquids such as Kaya et al. (2000), de Carvalho and Ferreira (2000), Furukawa and Fukano (2001), Sawai and Kaji (2001) and Azzopardi and Wren (2004). However, no studies have been done to investigate churn flow regime in a very high viscosity liquids. As the characteristic of churn flow in very viscous liquids has not been studied previously, it is examined for the first time in the present work.

High viscosity liquids can be found in many industrial applications. First, heavy oil and bitumen productions have very high viscosities and low surface tension compared with water. For example, Orinoco belt crude oils and Athabasca bitumen have viscosities of 3 to >1000 Pa.s and 2000 Pa.s respectively and surface tension of 0.01-0.03 N/m (Terebey et al. (1984)). Second, in the polymer manufacturing and processing, the viscosity ranges between 100-1000 Pa.s and surface tension approximately equal to the heavy oil. Third, in Volcanology, the silicate magmas in volcanic conduit consist of gas, silica-rich and crystals. The viscosity range of the common type of magma is 100-1000 Pa.s and about 0.4 N/m surface tension (Mysen

et al. (1982), McBirney and Murase (1984), and Gonnermann and Manga (2013)). However, the value of the viscosity depends on the chemical composition, which is in turn influenced by processes such as mineral crystallisation and gas exclusion. The recent studies, which have been involved in the volcanology, look at the behaviour of the Taylor bubbles are Pioli et al. (2012) Del Bello et al. (2015), and Capponi et al. (2016). In addition, the surface tensions of fluids used in chemical processing industries are lower than water at atmospheric temperature and pressure. Figure 1.1 displays some examples of liquids with a high viscosity which have been discussed earlier.



Figure 1.1: High viscosity liquids examples. Heavy oil and bitumen, polymer manufacture, magmas in volcanic conduit.

Predicting the characteristics of the hydrodynamics of gas flow in liquids of high viscosities is essential in the design and safety aspects. Besides, predicting the design parameters such as pressure gradient, void fraction or gas holdup, structure velocity and the bubbles dimensions has a significant role in the gas-liquid systems. In oil and gas industry, it is very important in the design of the equipment and the very long pipe. In addition, the design and safety of the industrial equipment and the pipelines in the polymer manufacturing. Finally, to predict some natural phenomenon and develop the knowledge about “strombolian activity” and the nature of the eruption in volcanos. Strombolian activity is a type of volcanic eruptions that creates

frequent and moderate eruptions of fluid lava. This explosion is created from the eruption of the large bubbles which rise through the magma columns the volcanic conduits (Blackburn et al. (1976), Jaupart and Vergnolle (1988), and Parfitt and Wilson (1995)).

Some progress has been made in recent years towards understanding the characteristics of gas flow in viscous liquids. for example, Pioli et al. (2012) investigated the effect gas flow rate, liquid viscosity, and the pipe geometry have on the characteristics of the gas-liquid flow. Liquids with a range of viscosities of 3-300 Pa.s and surface tension of 0.07-0.4 N/m were employed to study the dynamics of gas-liquid flow of basaltic magma in vertical conduit. Another subsequent work by Azzopardi et al. (2014b) studied the structure of air flowing in very viscous liquid and 240 mm diameter column. Glucose syrup of 300 Pa.s viscosity and 0.08 N/m surface tension were employed in their investigations.

Knowledge about the hydrodynamics of gas flow in high viscosity liquids and large diameter columns is still limited in spite of its importance. Therefore, the study of gas flow in viscous liquids has become an important aspect in the two-phase flow systems. In the present work, new information about very high viscosity and low surface tension liquids is presented. The characteristics of the hydrodynamics of gas flow in high viscosity liquids are studied in large diameter columns over a wide range of gas flow rates. The flow structures in bubbly and slug flow regimes are shown in very viscous liquids in addition to churn flow, which is introduced for the first time.

1.2 Aim and Objectives

The aim of this work, which is carried out at the University of Nottingham, is to study the characteristics of the hydrodynamics of gas passage in very high

viscosity oils and large diameter columns. Stagnant Silicone oils of 330 and 360 Pa.s viscosity have been employed in this investigation. Silicone oil has an excellent dielectric property (permittivity) that could be used with air in the Electrical Capacitance Tomography (ECT). Wide ranges of gas superficial velocities have been applied to create different flow patterns in 2 columns of 240 and 290 mm diameter.

The objectives of this work are:

1. To study the behaviour of gas flow in high viscosity liquids in different flow regimes.
2. To introduce the characteristics of churn flow in stagnant 330 Pa.s and 290 mm diameter column.
3. To investigate the effect of the pipe diameter on the flow structure in gas-high viscosity liquids.
4. To study the bubble creation using single gas injection point.
5. To investigate the effect of the methods of gas injection (the number and the position of the gas nozzle) on the flow structure in gas-high viscosity liquids.
6. To study the hydrodynamics of the bubble eruption at the top section of the column as well as the liquid oscillation there.
7. To study the hydrodynamics of the small bubbles (millimetres to centimetres diameter) generation in high viscosity liquids. In addition to the removal of small bubbles after concentration has ceased.

1.3 Thesis Outline

This project studies the hydrodynamics of gas flow in high viscosity oils and large diameter columns due to the lack of information about such systems. This thesis, which consists of 9 chapters, provides new information about the

bubbly and slug flow and introduces churn flow in high viscosity liquids and large diameter columns. It also contains information about the very small bubbles which occupy the viscous liquid. This thesis is divided into:

[Chapter 1](#) provides a brief introduction to the present work and also introduces the aims and objectives of this work and shows a scope of the thesis.

[Chapter 2](#) reviews the previous work in literature and provides a background for the gas-liquid flow in vertical pipes. It also identifies the gaps in the literature.

[Chapter 3](#) describes the experimental arrangements and the design of the rigs which are employed in this work. It also shows the flow facilities, fluids properties and the measurement techniques. Finally, it explains the methods of the data analysis which have been applied for the data processing.

[Chapter 4](#) contains an overview of the whole range of the gas flow rates used in the experiments of the effect of gas flow rate on the flow structure. It compares the parameters, such as pressure gradient, void fraction, structure velocity, film thickness, bubble dimensions and frequency, at different flow patterns. It also shows the effect of the pipe diameter on the flow structure over a wide range of gas flow rates.

[Chapter 5](#) includes the results of bubbly and slug flow patterns in high viscosity liquids. It also reports the mechanisms of the bubble generation in high viscosity liquids using single nozzle. The results obtained from the Electrical Capacitance Tomography (ECT) have been compared which the results obtained using the high-resolution camera and with models from literature. The bubble coalescence along the column is also presented in this chapter. Finally, the top surface oscillation and the bubble eruption at the top

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surface of the liquid are also examined in this chapter.

[**Chapter 6**](#) presents the effects of the method of gas injection on the flow structure in viscous liquids. This includes changing the number and the positions of the gas injection points. The effect of the gas injection time is also shown in this chapter.

[**Chapter 7**](#) introduces churn flow in high viscosity liquids and large diameter column. The transition to churn flow in 2 different diameter columns is also introduced. Parameters such as void fraction, PDFs, structure velocity, film thickness and velocity are presented.

[**Chapter 8**](#) contains the results of the small bubbles (millimetres to centimetres diameter) experiments which occupy the liquid slugs and the falling film. The generation, concentration, dispersing, and rising of the small bubbles are included in this chapter.

[**Chapter 9**](#) provides a conclusion for the main findings of this work and recommendations for further work.

Chapter 2

Literature Review

2.1 Introduction

This chapter explores gas-liquid flow in vertical pipes. It focuses on the work that has been carried out using viscous liquids. However, information is still very limited on the subject of gas-high viscosity liquids flow despite the importance of this subject. Predicting the hydrodynamic characteristics of gas flow in viscous liquids is essential in the heavy oil and bitumen (3-2000 Pa.s), polymer manufacturing and processing (100-1000 Pa.s) and for the Silicic magmas in volcanic conduits (100-10⁹ Pa.s). Only a few studies in literature employed high viscosity liquids in the investigations of gas-liquid flow in vertical large diameter pipes. In this chapter, the mechanism of the bubble generation at a single gas nozzle in the liquid is explained. The mechanisms of generating very small bubbles, which accumulate in the liquid in the pipe, are also described. Slug flow in vertical pipes and different liquid viscosities are shown in addition to the methods of determining the dimensions of the unit slugs. The equations, which applied for the comparison with the experimental data in the present work, are also displayed in this chapter. Moreover, the analytical methods for the determination of the void fraction, Probability Density Function (PDF) and the structure velocity are also included. The transition to churn flow is explained. Besides, a survey about the works carried out on this flow regime is shown. Finally, a background about the permittivity measurement technique which is applied in this work is presented at the end of this chapter.

2.2 Bubble Formation and the Gas Injection Method

Bubble formation in high viscosity liquids using single air injection was studied by several researchers to determine the bubble size and rise velocity. The prediction of the bubble diameter is essential in the design of the heat and mass transfer equipment. Bubbles play an important role in many industrial processes. They are employed to enhance the heat and mass transfer in the polymer, biochemical and other industries by circulating the gas phase to the viscous liquid phase.

The geometry of a bubble created at a single cylindrical nozzle in a liquid, neglecting the effect of the walls, at constant gas flow rate is controlled by several forces. The bubble is created in two stages. First, the expansion stage, when the volume of the bubble increases while it is still attached to the nozzle. Second, the detachment stage, the bubble moves away from the nozzle due to the effect of the forces which correspond to the increase in the bubble size. Gaddis and Vogelpohl (1986) developed an implicit equation to calculate the bubble diameter similar to the model produced by Ramakrishnan et al. (1969). This equation has been examined and validated experimentally with 4 different liquid viscosities (0.001-1 Pa.s) and nozzle diameters (0.2-6 mm). They used a curve fitting experiments to produce an explicit algorithm to determine the diameter of the bubbles. The equation to determine the bubble diameter, d_B which was driven by Gaddis and Vogelpohl (1986) considered 3 main cases. First, is the surface tension, S which controls the diameter of the detached bubble and considering the volumetric gas flow rate, V^* is equal to 0:

$$S = \left[\frac{6 d_h \sigma}{\rho_l g} \right]^{1/3} . \quad 2.1$$

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Where: d_h, σ, ρ_l, g and v are the diameter of the nozzle, the surface tension of the liquid, liquid density, gravitational acceleration and kinematic viscosity respectively.

The second case, which controls the bubble diameter, is the viscous drag, L in high viscosity liquids and intermediate gas flow rates:

$$L = \left[\frac{81 v V^*}{\pi g} \right]^{1/4} . \quad 2.2$$

And finally, the liquid inertia force, T for low viscosity liquids at high gas flow rates:

$$T = \left[\frac{135 V^{*2}}{4\pi^2 g} \right]^{1/5} . \quad 2.3$$

The summation of the 3 cases to determine the bubble detached diameter, d_B produces Equation 2.4:

$$d_B = \left[S^{4/3} + L + T^{4/5} \right]^{1/4} . \quad 2.4$$

Substituting Equations 2.1, 2.2, and 2.3 in Equation 2.4 gives:

$$d_B = \left[\left(\frac{6 d_h \sigma}{\rho_l g} \right)^{4/3} + \left(\frac{81 v V^*}{\pi g} \right) + \left(\frac{135 V^{*2}}{4\pi^2 g} \right)^{4/5} \right]^{1/4} . \quad 2.5$$

Another theory was proposed by Davidson and Schüler (1960) to predict the bubble volume at constant pressure and gas flow rate in high viscosity liquids. A wide range of gas flow rates (0-50 ml/s) and viscosity (0.5-1.4 Pa.s) were employed to obtain a good agreement with their theoretical approach. The bubble volume, V_B was calculated from:

$$V_B = \left(\frac{4\pi}{3} \right)^{1/4} \left(\frac{15 v V^*}{2g} \right)^{1/4} . \quad 2.6$$

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Equation 2.6 might be applicable for liquids with a limited range of surface tension because it does not take the surface tension of the liquid in consideration. Similarly, Snabre and Magnifotcham (1998) produced an implicit equation independently from Gaddis and Vogelpohl (1986) to calculate the bubble diameter:

$$\frac{\pi}{3} d_B^3 \rho_l g = \left(\frac{81 C_d^*}{16} + 9 \alpha \right) \frac{\rho V^{*2}}{\pi d_B^2} + \pi d_h \sigma, \quad 2.7$$

where, C_d^* , α and d_h are average drag coefficient, dimensionless inertial parameter and the gas nozzle inner diameter respectively.

In a study on the shape and the volume of the bubbles in viscous liquids of (0.001-1.1 Pa.s), Terasaka and Tsuge (1990) showed that the bubble volume is directly proportional to the gas chamber volume, orifice diameter and liquid viscosity. Furthermore, the time of the bubble formation in the higher viscosity liquids is higher, as the bubble volume is larger.

A number of studies looking at the effect of viscosity and the nozzle diameter on the bubble size and formation have been reported. Davidson and Schuler (1997), Davidson and Harrison (1963), and Ramakrishnan et al. (1969) showed that the bubble diameter increases with increasing liquid viscosity. However, Datta et al. (1950) found different results, noting that the bubble size decreases with increasing viscosity. On the other hand, in studies by Jamialahmadi et al. (2001) on the effect of the liquid physical properties, orifice diameter, and liquid height on the bubble size, concluded that the bubble size is dependent mainly on the surface tension of the liquid. In addition to the liquid viscosity, which is shown to have a considerable effect at high gas flow rates, these studies also showed that the bubble size is considerably proportional to the diameter of the nozzle and it is not affected by the liquid height in the column. A number of studies on the effect of the

gas injection nozzle diameter showed a similar result for the strong effect of the nozzle diameter (Kumar et al. (1976)). However, Akita and Yoshida (1974) reported that the bubble size is independent of the nozzle diameter.

Another study by Ohnuki and Akimoto (1996), which was carried out to investigate the effect of the injection methods on the structure of the air-water flow in a large vertical pipe, employed two types of injections in their experiments; sinter and nozzle air injections. They found that the flow structure is totally different along the test section. The lower part of the test section near the air injection showed a specific structure affected by the injection method. While the upper part did not show a considerable effect by the air injection according to the phase and the differential pressure distribution in the pipe.

2.3 Slug Flow

Slug flow in vertical pipes is a common flow type that can be found in many industrial applications. It occurs in two-phase gas and oil pipelines such as risers (subsea platforms). It can also be found in some process equipment such as vapour-liquid contractor or absorbers and vapour generators. In low viscosity liquids, slug flow starts when coalescence begins between the bubbles to create larger bubbles in the two-phase gas-liquid flow. The large bubbles diameter is approximately equal to the pipe diameter. They rise in a bullet shape and are known as Taylor bubbles. Taylor bubbles usually flow upward and cause the surrounded thin liquid film to fall down as a falling film. Small bubbles occupy the liquid slug beyond the tail of each Taylor bubble. In vertical pipes, Taylor bubbles seem to be symmetric in shape. However, in the case of inclined pipes, the nose of the bubbles could shift toward the pipe wall (Azzopardi (2006)). According to previous studies, this flow regime does not appear for gas-water and low viscosity liquids flow in

large diameter pipes (150 and 200 mm). Thus, the bubble flow is directly transit to churn (Chishti et al. (1998), Ohnuki and Akimoto (2000), and Hewakandamby et al. (2014)). However, slug flow seems to appear in large diameter pipes ≥ 127 mm in viscous liquids (Azzopardi et al. (2014b)), this will be the focus of present study.

The liquid viscosity significantly affects the structure of Taylor bubbles. In low viscosity liquids, it has a flat tail with a turbulent bubbly wake, which either moves down or coalesces with the Taylor bubble. While, by increasing the liquid viscosity, the shape of the Taylor bubble's tail becomes rounded, for a Buoyancy Reynolds number $R < 15$, and the turbulence in the wake disappears (White and Beardmore (1962), Viana et al. (2003), and Del Bello et al. (2012)). Reynolds number Re is $Re = UL/\nu$, while Buoyancy Reynolds number $R = ReH^2/L^2$. Where U, ν, L and H are velocity, kinematic viscosity, horizontal and vertical length scales respectively, Bartello and Tobias (2013).

In viscous liquids, the Taylor bubble is divided into four parts: nose, body (upper), body (lower), tail and wake. The first part is bullet shaped or hemispherical shaped nose. The nose is followed by the body, which is the second region, surrounded by falling liquid film. The third part is the tail of the bubble. This part has a different structure; it could be flat, concave or semi-hemispherical. The last part is the wake, which could be close or open, laminar or turbulent. The second region (the body) can be subdivided, as it is shown in Figure 2.1, into two parts: the upper part, where the developed film is accelerating and thinning and the lower part, where equal forces are acting on the film around the bubble and creating constant film thickness (Bugg et al. (1998) and Feng (2008)).

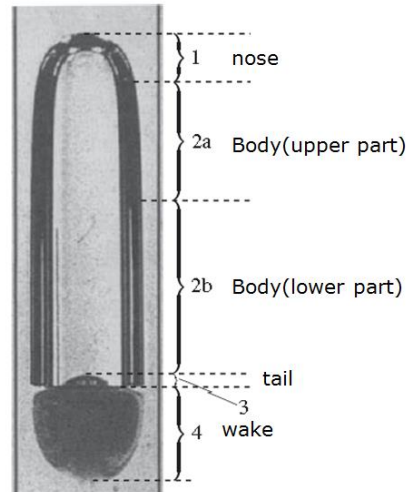


Figure 2.1: Taylor bubble structure in viscous liquids (Llewellyn et al. (2012)).

In a part of a comprehensive study by Viana et al. (2003) that focused on the rise velocity of the Taylor bubble in vertical pipes, liquids of viscosities ranged between 0.001-3.9 Pa.s and high-speed camera were employed in their work. Figure 2.2 compares Taylor bubbles rising in liquids of 4 different viscosities and a 76.2 mm diameter column. The photos were taken by the high-speed camera NAC HWS1000, 5000 frames per second. 3 different structures for the bubble wake appeared in the photos: close rounded, close air cup liquid hold up and open turbulent with "bubbles detachment". The variation in the structures of the wakes' shapes is corresponded to change in viscosities from low to high. As the Taylor bubble in (a) is rising in water, the wake consists of unsteady small bubbles with a high turbulence section. The bubbles in the wake contain more air which is dragged by the fast falling film around the Taylor bubble. The second bubble in (b) is rising in 4.8 Pa.s machinery oil. The nose of the bubble is similar to the bubble in (a). The bottom of the bubble is steady, flat, and identified by an air cup that contains liquid. Small bubbles may be entrained into this section, or sometimes it could break up into a small bubble. Bubbles (c and d) are rising in Silicone oils with viscosities of 1.3 and 3.9 Pa.s. The shape of Taylor bubbles in both cases is similar. They have the same rounded top and convex bottom. The

bottoms are surrounded by a hemispherical “air membrane” that is trapped in liquid. No bubbles are left behind the Taylor bubble (in the liquid slug). The only difference between the two bubbles is the amount of the liquid, which is trapped in the bottom of the bubble. It decreases with increasing the viscosity of the liquid.

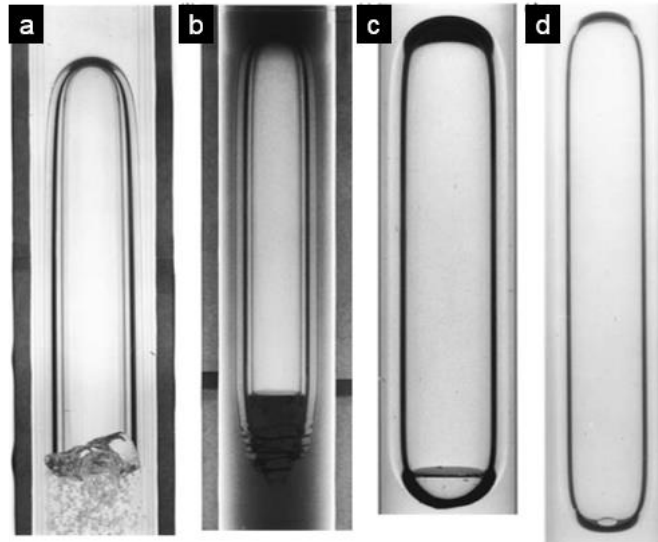


Figure 2.2: Photos of Taylor bubbles rising through 76.2 mm diameter column in liquids of different viscosities (Viana et al. (2003)).

When a large bubble rises in a column of liquid, it tends to expand to become as the same pressure as the fluid static pressure in the column. This leads to generate bubbles in meter-size, in the case of neglecting the surface tension force (Seyfried and Freundt (2000)). Therefore, no energetic explosion occurs as the pressure inside the large bubbles is close to the atmospheric pressure. However, in high viscosity systems, the viscous and the inertial forces hinder the bubble expansion, Prousevitch et al. (1993) and James et al. (2008). James et al. (2004) studied the pressure change associated with the rise of the large bubbles in magmas conduits and the bubble eruption at the top surface. They employed liquids with viscosities of 0.001, 0.09 and 0.9 Pa.s in vertical and inclined pipes. They stated that the inclination of the column promotes the coalescence of the bubbles then increasing the bubbles size and rise velocity. The pressure gradients are

significantly affected by the rise of the large bubbles along the conduit. Different frequencies of bubbles and slugs were recorded at different locations along the column in the work of James et al. (2004). On the other hand, at the top surface of the column, the length of the large bubbles controlled the pressure gradients. The overpressure was calculated numerically by Vergniolle (1998) and James et al. (2008) by applying the large bubbles rise models which are not considered the effect of the top surface. The overpressure, which is produced by the expansion of large bubbles rising in the column against the viscous and inertial hindrance, was also studied by James et al. (2009). They presented a model to determine the rise velocity of the large bubble that is in equilibrium with the local static pressure. They assumed that the values of the measurements at the top surface are a function of overpressure associated with the large bubble rise at the column top surface. In another study on the nature of the strombolian explosion, Del Bello et al. (2015) found that the existence of the viscous liquid slug at the top of the volcanic conduit controls the hydrodynamics of the bubble eruption. The viscous liquid slug increases the energy of the bubble explosion by increasing the bubble overpressure.

The importance of slug flow in industrial and engineering fields motivates most of the earliest works and studies of the behaviour of Taylor bubbles. Theoretical studies were published by Dumitrescu (1943), Davies and Taylor (1950), Goldsmith and Mason (1962a), and Harland (1965), whilst by (1950), Goldsmith and Mason (1962a) White and Beardmore (1962), Allan and Mason (1962), Campos and Carvalho (1988), Viana et al. (2003), Brauner and Ullmann (2004) and Nogueira et al. (2006) published experimental studies. Also numerical studies (Taha and Cui (2006), Zheng et al. (2008) and Feng (2008)). The main aim of the above works was to study the rise velocity characteristic of the Taylor bubble. A number of

studies focused on the behaviour of falling film around Taylor bubble such as Karapantsios et al. (1989). Bugg and Saad (2002), Van Hout et al. (2002), Lel et al. (2005), and Llewellyn et al. (2012).

Most of the previous studies focused on air-water mixtures. However, the surface tensions of fluids used in chemical processing industries are lower than water at atmospheric temperature and pressure. Slug flow in high viscosity liquids can be found in many industrial applications. For example, heavy oil and bitumen production have a very high viscosities and low surface tension compared to water. Terebey et al. (1984) gave a viscosity value of 3 to >1000 Pa.s and 2000 Pa.s for Orinoco belt crude oils and Athabasca bitumen respectively and surface tension of about 0.4 N/m. Another example of high viscosity liquids is in the polymer manufacturing and processing (100-1000 Pa.s) viscosity and surface tension approximately equal to the heavy oils. In addition, the silicate magmas in volcanic conduit, which consist of gas, silica-rich and crystals. The viscosity range of the least viscous type of magma (basalt magma) is 100-1000 Pa.s. While the highest viscosities magma is between $100-10^9$ Pa.s and surface tension of about 0.4 N/m. However, the value of the viscosity depends on the chemical composition, which in turn is influenced by processes such as mineral crystallisation and gas exclusion (Mysen et al., 1982, McBirney and Murase, 1984, Gonnermann and Manga, 2013). The previous studies, which are involved in the Volcanology, looked at the behaviour of the Taylor bubbles. As it is believed, these bubbles cause the "strombolian activity". Blackburn et al. (1976), Seyfried and Freundt (2000), James et al. (2009), Pioli et al. (2012), Del Bello et al. (2015), and Capponi et al. (2016) pointed out that the nature of eruptions in the volcanos is affected by the liquid slug rise. The falling magma also causes the overpressure developing during the rise of the slug which is considered as a key role for the determination of the "eruption

explosivity" (Del Bello et al. (2012)).

2.3.1 Taylor Bubble and Liquid Slug Length

Predicting the dimension of the slug unit in viscous liquids is very important in designing the very long pipelines in oil and gas industry. In addition, it provides more information about the characteristics of the magmas in the volcanic conduits. The majority of the works carried out to study the dimensions of the Taylor bubble and liquid slug employed water or low viscosity liquids. Therefore, information about the dimensions of the slug unit in the very viscous liquid is limited.

Khatib and Richardson (1984) proposed an equation to calculate the length of a Taylor bubble numerically. They used air in water and slurry mixtures in 38.8 mm diameter vertical pipe and electrical conductivity technique. This method, which is based on the conductance that is different between the two fluids, is to measure the cross-sectional phase distribution in the pipe. In their equation, they used the values of the Probability PDF and the averaged void fraction to calculate the Taylor bubble, L_{TB} and liquid slug, L_s lengths. As it is explained later in this chapter, slug flow is pronounced by two peaks of PDF. The one with the higher void fraction represents the Taylor bubble α_{TB} , while the one with the lower correspond to the liquid slug, α_s . The length of the slug unit, L_u (see Figure 2.3) can be written as:

$$L_u = L_s + L_{TB} . \quad 2.8$$

It can be also calculated from the structure velocity, U_{st} and the frequency, f :

$$L_u = \frac{U_{st}}{f} , \quad 2.9$$

$$\frac{L_s}{L_u} = \frac{\varepsilon_{g(avg)} - \alpha_{TB}}{\alpha_s - \alpha_{TB}}, \quad 2.10$$

where $\varepsilon_{g(avg)}$ is the overall void fraction.

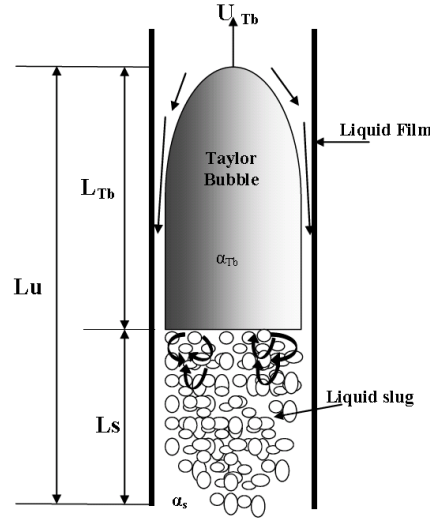


Figure 2.3 Taylor bubble and liquid slug in gas-water mixture

The length of the slug unit was also studied by Abdulkadir et al. (2014) who employed air-Silicone oil of 5.3 Pa.s. They reported that Taylor bubble and liquid slug lengths increase with an increasing gas flow rate. Also, the length of the liquid slug is seen to change because of the coalescence of the bubbles existed at the tail of Taylor bubbles. This result shows an agreement with Akagawa and Sakaguchi (1966), Fernandes (1981) and Van Hout et al. (2002). Using stagnant glucose syrup of 300 Pa.s in a 240 mm diameter column, Azzopardi et al. (2014b) determined the slug unit length over a wide range of gas flow rates using Electrical Capacitance Tomography (ECT). The lengths of Taylor bubbles increased with increasing the gas flow rate while that of the liquid slugs decreased over the same range of gas flow rate. They also reported agreement between their results and the approach by Khatib and Richardson (1984), which was also employed in the present work. The work by Azzopardi et al. (2014b) was followed a work carried out by Pioli et al. (2012) to study the dynamics of gas-liquid flow of basaltic magma in vertical conduits. In this work, a wide range of liquid viscosities was

employed over a wide range of gas flow rates in different diameter columns. Similar results were reported for the lengths of the slug units.

2.3.2 Falling Film around Taylor Bubble

The thickness of the falling liquid film around the large bubbles is dependent on the amount of the liquid flowing down on the pipe wall. Llewellyn et al. (2012) studied the effect of the physical parameters on film thickness around Taylor bubble rising in a vertical cylindrical pipe. They found that the relationship between the film thickness and the flow rate of falling film is related to the flow type (laminar or turbulent). In their experiment, they had a pipe with a closed top in which all displaced liquid had to fall past the Taylor bubble. While, if the top of the pipe is open, the top surface area can rise and hence not all the liquid will fall down past the Taylor bubble. Goldsmith and Mason (1962b), Brown (1965), and Padmanaban (2006) applied the Navier-Stokes equation in cylindrical coordinates for thicker films in cylindrical pipes. This results in an equation based on the film flow rate and the bubble rising velocity, U_B ;

$$U_B \pi (R_p - \delta)^2 F_D = \left(\frac{\pi \rho_l g}{8l} \right) \left[-4(R_p - \delta)^2 R_p^2 + R_p^4 - 4(R_p - \delta)^4 \ln \left(\frac{R_p - \delta}{R_p} \right) + 3(R_p - \delta)^4 \right]. \quad 2.11$$

R_p Is the radius of the pipe, δ is the film thickness around the large bubbles, and F_D is a factor relating the amount of the liquid film falling down.

In vertical slug flow, the annular film around Taylor bubble is considered to behave as a falling film without interfacial shear. The thickness of the falling film around Taylor bubble depends on its flow rate (Azzopardi et al. (2014a)) The falling film, which is created at the top surface, is assumed to be identical with the film at the bottom in term of the relationship between the thickness

and the flow rate (Fernandes et al. (1983)). Most of the previous studies of the liquid film flow were concentrated on the film in annular flow and low viscosity liquids. Dukler and Bergelin (1952), Fernandes et al. (1983), Goldsmith and Mason (1962b), Batchelor (2000) and Llewellyn et al. (2012) studied the behaviour of the falling film around Taylor bubble both theoretically and experimentally.

Slug flow in viscous oil is characterised by smaller bubbles distributed in the whole liquid (falling film and liquid slug), Azzopardi et al. (2014b), therefore they have to be considered in the determination of the void fraction of Taylor bubbles, α_{TB} . They can be measured by:

$$\alpha_{TB} = \frac{\frac{\pi}{4}[(D - 2\delta)^2 + \{D^2 - (D - 2\delta)^2\}\alpha_s]}{\frac{\pi D^2}{4}}, \quad 2.12$$

where D is the inner diameter of the pipe, and δ is the film thickness around Taylor bubble.

The falling film thickness to the pipe radius, R_p can be determined from the void fraction of Taylor bubble, α_{TB} and the void fraction of small bubbles, which are distributed in liquid slug and the falling film, α_s (Azzopardi et al. (2014b)):

$$\frac{\delta}{R_p} = 1 - \sqrt{1 - \frac{1 - \alpha_{TB}}{1 - \alpha_s}}. \quad 2.13$$

2.4 Churn Flow

Churn flow, which is an intermediate flow that appears between the slug and annular flow regimes, is an important flow pattern in the gas-liquid two-phase flow. The pressure drop, mass and heat transfer are design parameters which have a significant role in the safety and the stability of gas-liquid systems. In churn flow, the gas-liquid travels upward and downward in a churning or oscillating movement with a resultant of upward flow. This oscillating motion is a result of two forces, gravity and shear forces, acting in opposing directions on the falling film around Taylor bubbles. Churn flow was first defined by Zuber and Findlay (1965) as "churn-turbulent" bubbly flow. Then, it was considered as an intermediate region that occurs between the slug and churn flow regimes by Hewitt and Hall-Taylor (1970). And finally, it was defined as churn flow regime by Taitel et al. (1980). Figure 2.4 illustrates the flow map in vertical pipes for low viscosity liquids by Taitel et al. (1980). Further, Hewitt et al. (1985) reported similarities between these two flow regimes by employing a visual technique. In addition, a modified form of slug flow model was applied by De Cachard and Delhay (1996) for churn flow. However, pressure drop in pipes at churn flow regime is seen to be better predicted by annular flow model, Holt et al. (1999). Churn flow was correlated to be a developing form of slug flow regime. On the other hand, Barbosa Jr et al. (2001) defined this flow regime as it appears after the breakdown of slug flow when the velocity increases. Since churn flow is similar to annular flow regime in features, a number of studies considered it as annular flow. For instance, Nicklin (1962) used the name "semi-annular" flow to recognise the annular properties. Churn flow is distinguished from the annular flow regime based on 3 criteria.

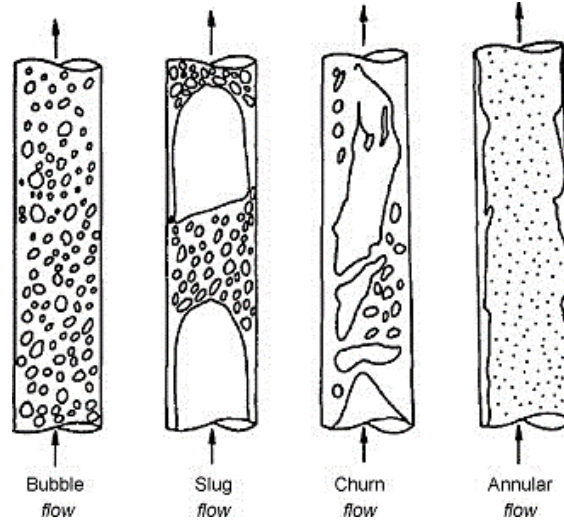


Figure 2.4: Two-phase flow map in vertical pipes for water (Taitel et al. (1980)).

First, the change of the pressure gradient inside the pipe with an increasing gas flow rate. The friction and the intensive wave activity which result from the sudden breakdown of slug flow increase the pressure gradient significantly inside the pipe. Owen (1986) presented the change of the pressure gradient data against the dimensionless gas velocity for air-water flow in a 32 mm diameter vertical pipe. According to his results in Figure 2.5, the pressure gradient values showed a sudden drop at slug flow regime then increased at the transition to churn flow due to the high frequency wave activity and friction. However, in Owen's work, the values of the pressure gradients decreased with an increasing gas flow rate at a constant liquid flow rate. The decrease in the amount of the interaction between the gas and the liquid is a reason for the decrease of the pressure gradient at the selected gas flow rate. Then by increasing the gas flow rate, the values of the pressure gradients increased again. A comparison between Owen (1986) work and the present work is elucidated in Chapter 4 and Chapter 7. The dimensionless gas superficial velocity, U_g^* in Figure 2.5 is calculated from:

$$U_g^* = U_{gs} \left[\frac{\rho_g}{g D (\rho_l - \rho_g)} \right]^{1/2}, \quad 2.14$$

where U_{gs} is the gas superficial velocity. g, D, ρ_l and ρ_g are the gravitational acceleration, pipe diameter, liquid and gas density respectively.

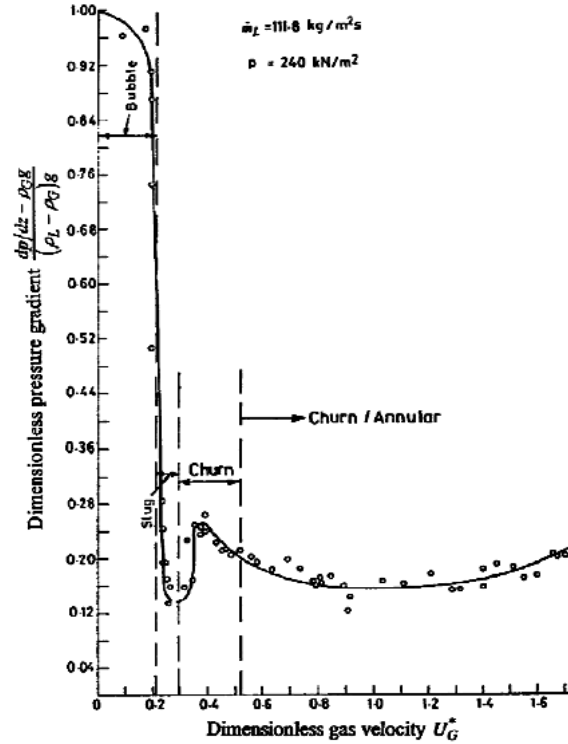


Figure 2.5: Dimensionless pressure gradient distribution at different gas flow rates in vertical pipes by Owen (1986)

Second, the mechanism of the two-phase flow. Churn flow establishes when the falling liquid film accumulates and starts to flow upward in a form of large waves. This flow mechanism, which was proposed by Hewitt et al. (1985), is similar to the flooding phenomenon due to the flow of large waves. Figure 2.6 illustrates the flow mechanism suggested by Hewitt et al. (1985). It shows the upward direction of the large waves on the pipe wall. The draining film and the flooding phenomenon in vertical pipes are described in more detail later in this chapter.

The third criteria, for starting churn flow, is the entrainment of the droplets from the large liquid fractions which travel as large flooding waves along the column. Wallis (1962) found that increasing the gas velocity decreases the entrained fraction to a minimum value (churn flow). Then a further increase

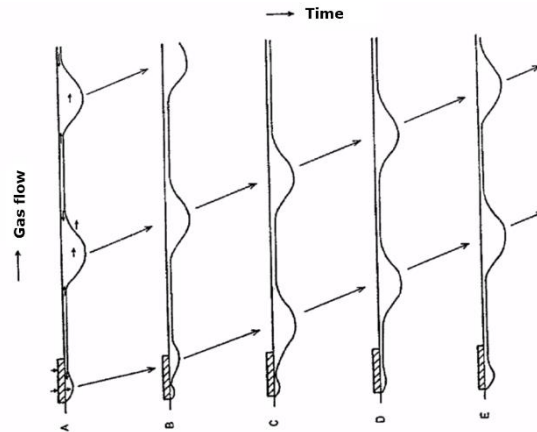


Figure 2.6: Mechanism of churn flow by Hewitt et al. (1985)

in the upward gas flow rate increases the liquid entrained fraction in the annular flow. Churn flow in low viscosity liquids was also studied by Brauner and Barnea (1986), Sekoguchi and Mori (1997), Kaya et al. (2000), de Carvalho and Ferreira (2000), Furukawa and Fukano (2001), Sawai and Kaji (2001) and Azzopardi and Wren (2004).

2.4.1 Transition to Churn Flow

Transition to churn flow occurs at critical gas flow rates when the length of Taylor bubbles in slug regime decreases and the liquid slugs become more aerated (Chen and Brill (1997)). Different mechanisms for the transition to churn flow or "slug to churn" regime were suggested for gas-low viscosity liquids. First, Dukler and Taitel (1976) proposed that the transition to churn flow occurs when the lengths of the liquid slugs between Taylor bubbles decrease significantly. At this point, the thin slug bridges collapse due to the effect of Taylor bubbles. The second suggestion was by Taitel et al. (1980) who considered churn flow as a developed form of slug flow, as it is shown earlier. Mishima and Ishii (1984) suggested a mechanism similar to the first one by Dukler and Taitel (1976). They assumed that the transition to churn flow occurs when the mean void fraction of Taylor bubbles becomes less than the mean void fraction of the liquid slugs. The reason is that the liquid slugs

become more aerated, so the overall void fraction exceeds the void fraction of Taylor bubbles. On the other hand, both Brauner and Barnea (1986) and Barnea (1987) assumed that when the void fraction of the liquid slugs becomes equal to the maximum value to the "bubble cubic volumetric packing (0.5)", then the transition to churn flow starts. This mechanism proposes that the transition from slug to churn flow starts when the void fraction of liquid slugs become equal to the maximum volumetric package along with a line of constant structure velocity. Brauner and Barnea (1986) reported a value of 0.5 for the maximum volumetric packing in their work. They also postulated that the transition occurs when the regions of the high gas fraction in the liquid slugs destroy the structured slug flow rapidly. The transition to churn flow was also related to the flooding of the liquid which drains from the Taylor bubbles. The last theory was introduced by McQuillan and Whalley (1985), Nicklin (1962), Wallis (1969) and Govan et al. (1991). The next section explains Flooding thoroughly. Finally, Chen and Brill (1997) introduced an equation for the transition from slug to churn flow which has better agreement with the experimental data comparing with the previous models. They based their analysis of the mechanisms of transition to churn on the effect of the wake of Taylor bubble.

The effect of the liquid viscosity on the characteristics of transition to churn flow was studied by Pioli et al. (2012), using a range of liquid viscosities and column diameters. They suggested that the transition to churn flow occurs at lower gas flow rates with increasing the viscosity of the liquid and decreasing the pipe diameter.

2.4.2 Flooding Churn Flow

Flooding vertical pipes occur when the critical gas velocity, which flows upward in a vertical pipe, reaches a point when the created waves from the falling liquid film can grow very large. The flow pattern changes when the very large waves, which have the same interfacial velocity as the liquid film, start to bridge the column. Above the flooding point, the liquid flows upwards as very large waves or bridges, while just below this point the liquid film flows downwards. Predicting the flooding point is essential for the designing and the safety of the industrial equipment and the pipelines. Flooding affects the performance of the chemical plants such as falling film chemical reactor, packed towers and the vertical tube condenser. Flooding has been studied widely by many researchers employing water and low viscosity liquids to predict its onset point. In some studies, the upward liquid flow was considered as onset point because of the interfacial waves. Whereas, in others, the physical mechanisms were considered in the flooding phenomenon such as the flow reversal within the liquid film. The principles of the interfacial wave models are that the waves become large to a point which bridge or close the pipe or, the flooding starts when the gas pushes the waves in addition to the accumulated falling waves from the original one. None of the models, which were proposed by Shearer and Nedderman (1965), Cetinbudaklar and Jameson (1969), Imura et al. (1977), Ueda and Suzuki (1978) and Govan et al. (1991) ,in general agreed with the experimental studies employing wide range of experimental conditions (Karimi and Kawaji (2000)).

According to a work carried out by Govan et al. (1991), flooding occurs by the creation of single waves near the liquid injection point which, then, travel upward by the gas above the liquid inlet point. After this point, a fraction of

the liquid flows downward and another fraction travels upward the liquid inlet point in the column. Govan et al. (1991) did not report any significant difference when comparing the measured holdup and the pressure gradient for the region above the liquid inlet in the present and the absent of the falling film below the injector. On the other hand, Karimi and Kawaji (2000) studied the hydrodynamics of the falling liquid in a 50.8 mm diameter column using deodorised kerosene and a high-speed camera. They pointed out that the interfacial waves do not flow downward and do not block the column at the onset of flooding.

2.5 Void Fraction

Void fraction, or gas fraction, is an important parameter which could be determined in the gas-liquid flow. It is a dimensionless value that represents the cross-sectional area occupied by the gas in gas-liquid flows. The values of void fraction vary between 0-1 or 0-100% when it is based on the percentage (Bertola (2003)). Void fraction is controlled by a number of parameters such as the dimensions of the pipe, the physical properties of the fluids, pressure, flow rates and the inclination angle of the pipe. Changing the temperature and the pressure of the fluids changes the values of void fraction continuously along the pipe (Saleh (2002) and Bertola (2003)). Therefore, predicting the value of the void fraction is crucial in the transition between the flow pattern and in heat transfer and control equipment, for example, control valves, two-phase pumps, and separators. The void fraction is also an important key parameter to predict the pressure drop in any industrial vertical pipe due to the acceleration and the gravitation of the fluids.

A review of the recent literature yielded a limited number of studies of the void fraction in viscous liquids. Spisak and Idzik (1994) employed air-

lubricating oils of 4.8 Pa.s viscosity to measure the void fraction in 25 mm column in slug flow regime. A new empirical correlation was proposed by Kabir and Hasan (1990) to measure the void fraction using commercial oils to study the pressure drop and the flow patterns in the oil wells. Glycerol oil/water of 0.007 Pa.s viscosity was employed with air to study the void fraction in annular flow in 26 mm diameter column by Fukano and Furukawa (1998). The same mixture, but with viscosities up to 0.55 Pa.s, was used by McNeil and Stuart (2003) for the prediction of the void fraction in annular flow. Schmidt et al. (2008) employed a range of liquid viscosities (7-9 Pa.s) in a column of 54.5 mm diameter to study the void fraction distributions using γ -ray densitometry. Moreover, the most recent work was the one carried out by Pioli et al. (2012) so as to study the characteristics of the two-phase flow of the basaltic magmas, using a wide range of liquid viscosities (3-300 Pa.s). Following the Pioli's work, Azzopardi et al. (2014b) employed Glucose syrup of 300 Pa.s to study the properties of the gas flow in viscous liquids. The void fraction was measured using the ECT and the level swell in a 240 mm diameter column.

Using the values of void fraction, the structure velocity (Taylor bubble and liquid slug velocities) and the phase distribution inside the pipe (bubble length and liquid film thickness) can be determined. Void fraction increases with an increasing gas flow rate under constant liquid flow rate in vertical and horizontal pipes. It can be determined from the liquid hold up, which is the volume fraction of cross section occupied by the liquid; this can be determined by using optical or electrical probes (Hewitt (1978)). In this work, ECT was used to determine the ratio of the liquid hold up in the column.

In gas-liquid flow, the summation of the volume fractions is equal to 1.

$$\varepsilon_g + \varepsilon_l = 1 \quad , \quad 2.15$$

$$\varepsilon_g = \frac{V_g}{V_g + V_l} \quad , \quad 2.16$$

And

$$\varepsilon_l = \frac{V_l}{V_l + V_g} \quad . \quad 2.17$$

Then,

$$\varepsilon_g = 1 - \varepsilon_l \quad , \quad 2.18$$

where ε_g is void fraction, ε_l is liquid holdup, V_l and V_g are the volume fraction of gas and liquid phase respectively.

The overall average void fraction, $\varepsilon_{g(avg)}$ can also be measured from the level swell by the determination of the initial height of the liquid, H_{liquid} before gas injection and the height of the gas-liquid mixture, H_{mix} after gas injection. This can be calculated based on the assumption that the volume of the liquid in the column remains constant (Al-Oufi et al. (2010));

$$\varepsilon_{g(avg)} = \frac{H_{mix} - H_{liquid}}{H_{mix}} \quad . \quad 2.19$$

2.6 Probability Density Function

The Probability density function (PDF) is a variation of the amplitude frequencies of void fraction, which is used for the identifications of flow patterns. Typical results for the PDF are shown in Figure 2.7 applied by Jones Jr and Zuber (1975) who employed air-water in vertical flow applying rectangular test section and x-ray technique.

According to the figure, bubbly flow in (a) consists of one single peak at low

void fraction corresponding to the spherical bubbles flow. Slug flow (b) contains two peaks of PDFs. One at high void represents the large bubble (Taylor bubble) rising in the column. Second, small peak represents the small bubbles in the liquid slug, which have low void fraction compared with Taylor bubbles. Finally, annular flow in (c), a single peak with high void fraction appears to be corresponding with the gas core in the middle of the column.

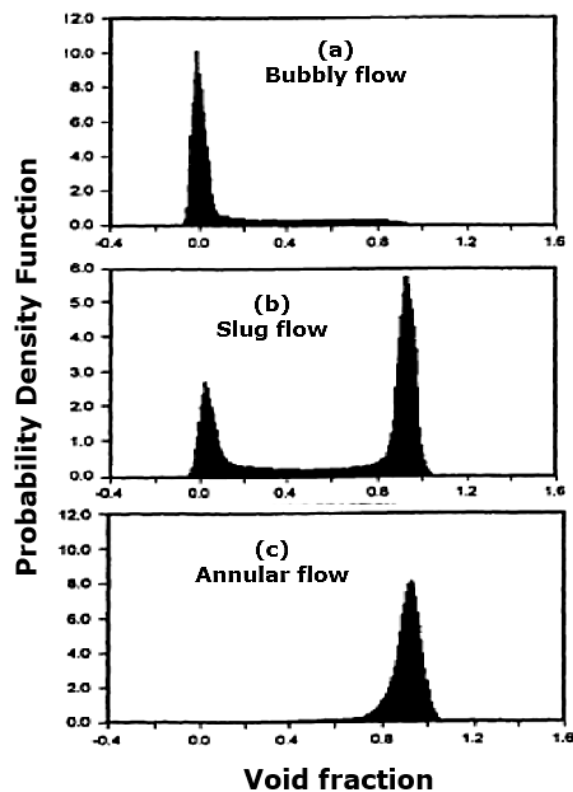


Figure 2.7. Probability density function as a function of void fraction (Jones Jr and Zuber (1975)).

Similar results were reported by Costigan and Whalley (1997) using PDF for the identification of the flow pattern. They showed that the bubbly flow occurs when the void fraction is between 0.2-0.3 at all times. Stable slug appears when the two PDF peaks are around 0.4 and 0.8, (0.5-0.6) for churn flow, and finally, 0.8-0.9 for annular flow. See Figure 2.8. Other investigations followed these studies applying PDF for flow pattern identification by using different pipe geometries, Vince and Lahey Jr (1982), Kelessidis and Dukler (1989), and Das and Pattanayak (1993).

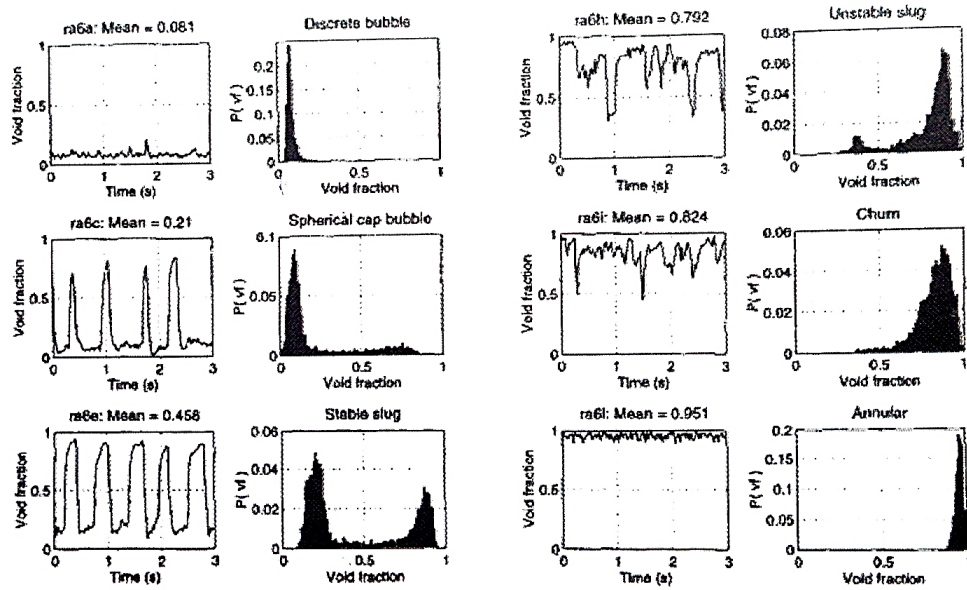


Figure 2.8. Time series and PDF profiles by Costigan and Whalley (1997) for air-water flow in 32 mm diameter pipe.

2.7 Structure Velocity

Studying the dynamics of Taylor bubbles in high viscosity liquids is essential in the many industrial applications such as in the gas and oil wells drilling and the riser tube. Pressure drop, heat and mass transfer coefficients and the hold-up are mainly dependent on the structure velocity. The rise velocity of Taylor bubble U_{TB} , which bridges the pipe cross section area in stagnant liquid, was determined both theoretically and experimentally by Davies and Taylor (1950) and Nicklin (1962).

$$U_{TB} = F_r \sqrt{g D_B}, \quad 2.20$$

where: F_r is Froude number, a dimensionless velocity represents the ratio of the gravitational and inertial forces which are equal to 0.351 and 0.328. These values are constants proposed by Dumitrescu (1943) and Davies and Taylor (1950) to represent F_r in equation 2.20. Dumitrescu (1943), proposed F_r equal to 0.351 for the first time analytically. Then, Davies and Taylor

(1950) presented a value of 0.328 analytically and experimentally.

Viana et al. (2003) presented an equation to determine the value of F_r basing on the dimensionless inverse velocity Buoyancy Reynolds number, R . ρ_l and ρ_g are the density of the liquid and the gas respectively.

$$R = \frac{\sqrt{D_B^3 g (\rho_l - \rho_g) \rho_l}}{\mu} , \quad 2.21$$

and Eötvös number, E_o which is the ratio of the interfacial tension and viscous forces;

$$E_o = \frac{g \rho_l D_B^2}{\sigma} . \quad 2.22$$

They found that for $R > 200$;

$$F_r = \frac{0.34}{(1 + 3805/E_o^{3.06})^{0.58}} , \quad 2.23$$

while for $R < 10$

$$F_r = \frac{9.494 \times 10^{-3}}{(1 + 6197/E_o^{2.561})^{0.5793}} . \quad 2.24$$

They also presented a universal correlation for the flat region of high Buoyancy Reynolds number and inclined region of low Buoyancy Reynolds number. These two regions are separated by a transition region ($10 < R < 200$), which was described by fitting the data to a "logistic dose curve", the equation has 13 empirical constants. The accurate values of F_r were shown by Viana et al. (2003) and Azzopardi et al. (2014b) for liquids up to 300 Pa.s viscosity.

The rise velocity of the isolated Taylor bubble, U_{TB} in a flowing turbulent liquid can be determined by the following relationship, Nicklin (1962);

$$U_{TB} = C_o(U_{gs} + U_{ls}) + K F_r \sqrt{g D_B} , \quad 2.25$$

U_{gs} and U_{ls} are the gas and liquid superficial velocities respectively,

$$K = \frac{0.905}{(1 - \varepsilon_{gs})^{3.95}} . \quad 2.26$$

The coefficient C_o represents the ratio of the maximum centreline velocity of the bubble to the mean upward liquid velocity. Nicklin (1962) suggested a value of 1.2 for the coefficient C_o . Collins et al. (1978) suggested a slight modification of Nicklin (1962) relationship, basing on a strong theoretical support. He suggested a value of 1.29 for C_o in Equation 2.25 the difference between the two C_o values is due to the difference of pipe diameters. However, according to Nicklin (1962) the higher values of C_o are more suitable when the flow rates decrease. Another study by Dukler and Fabre (1994) and Guet et al. (2004) proposed more complicated equation for C_o value. It can be written as:

$$C_o = \frac{C_{Bc}}{\left[1 + \left(\frac{R_{em}}{R_{ec}}\right)^2\right]} + \frac{C_{0,Re=\infty}}{\left[1 + \left(\frac{R_{ec}}{R_{em}}\right)^2\right]} , \quad 2.27$$

where: $C_{Bc} = 5$ or 2.27, $R_{ec} = 4000$, $C_{0,Re=\infty} = 1.2$, and

$$R_{em} = \frac{D_B \rho_l (U_{gs} + U_{ls})}{\mu} . \quad 2.28$$

The rise velocity of Taylor bubble is affected by the dynamic viscosity of the liquid, the density, the interfacial tension between the gas and the liquid, and the gravitational acceleration and finally the pipe diameter (White and Beardmore (1962) and Wallis (1969)). However, in case when the value of E_o is more than 40, the behaviour of Taylor bubble is independent of the interfacial tension which is negligible. Thus, the viscous and inertial forces dominate the Taylor bubble rise velocity. These conditions are applied in all

volcanic slugs and some industrial fields (Seyfried and Freundt (2000)). There are further studies carried out to determine the velocity of Taylor bubbles in viscous liquids by Bhaga and Weber (1981), Philip et al. (1990), Das et al. (1998), Van Hout et al. (2001), Bugg and Saad (2002), and Talaia (2007).

2.8 Small Bubbles in High Viscosity Liquids

The small bubbles, which are created due to the interaction between the gas and the liquid, accumulate in the liquid due to the slow motion of the liquid. Small bubbles holdup in viscous liquids has a numerous importance in chemical and industrial processing. Most of the studies on the small bubbles bed were conducted by using water or low viscosity liquids Shah et al. (1982) and Lubbert et al. (1988). However, a number of studies used viscous liquids with viscosities about 1 Pa.s, Schumpe and Deckwer (1987), Philip et al. (1990), Deckwer and Field (1992), Kawalec-Pietrenko (1992), and Kastánek et al. (1993b).

The small bubbles are created from the eruption of the large bubble at the top surface of the liquid. The curved film which retracts from the ruptured bubbles, traps air forming unstable toroidal geometry. Therefore, a ring of small bubbles is created at the top section of the liquid. The mechanism of the small bubble generation was studied by Bird et al. (2010) using 2 synchronised high-speed cameras. One at the side and another below the bubbles. Figure 2.9 displays the two stages of the small bubbles creation from the eruption of larger bubbles in 0.31 Pa.s viscosity glycerol-water. First is the collapse of the liquid film above the large bubble and the trapping of the air after falling down. The second stage is the forming of the small bubbles after the breakup of each torus of trapped air. The creation of the small bubbles in viscous liquids (1.5 Pa.s castor oil) was studied earlier by

Pandit et al. (1987). They produced two mechanisms for generating small bubbles of 0.5-2 mm diameter. First, from the eruption of large bubbles of 10-50 mm diameter. And second, from the gas injection points.

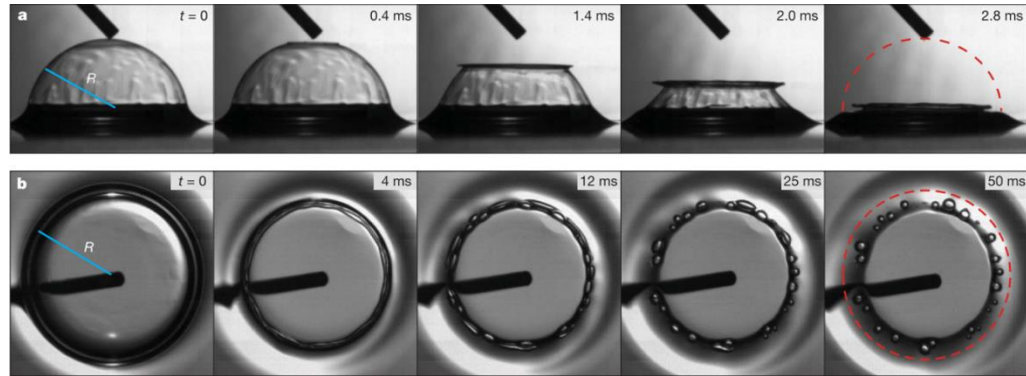


Figure 2.9: High-speed images showing the mechanism of the small bubbles creation at the top section of the glycerol-water mixture with a viscosity of 0.31 Pa.s is from the side and b is from below. Bird et al. (2010).

The first method, which is from the eruption of the large bubble, is the same as that which is produced later by Bird et al. (2010). The second source of the small bubbles is from the coalescence and the break-up of the large bubbles at the gas nozzles. According to Otake et al. (1977) and Pandit et al. (1987), the small bubbles are generated from the tail which is created during the coalescence of the bubbles at the gas inlet point. This tail breaks up to smaller bubbles after it gets elongated. Then the small bubbles are carried up by the liquid motion. A similar mechanism for the small bubbles generation was shown by Pioli et al. (2012) and Azzopardi et al. (2014b). Bird et al. (2010) reported that the amount of the trapped gas which is produced from the eruption of the large bubble is controlled by several factors. First, the capability of the collapsed liquid film to bend inward at the top surface of the liquid. Second, the stability of the area surrounding the large bubble at the top surface. And finally, the amount of the energy which is required for the large bubble to rupture. Therefore, increasing the pressure inside the large bubbles increases the curvature of the film and produces more small bubbles.

The small bubbles, which are generated in high viscosity liquids, accumulate in the system due to their low rising velocity. At low gas flow rates, the liquid slugs rise at low velocity and carry the small bubbles. The rise velocity of the small bubbles increases with increasing the gas flow rate due to increasing the rise velocity of the large bubbles which in turn push the liquid slug upward. This circulation motion of the liquid enhances the mixing and the releasing of the small bubbles. Philip et al. (1990) studied the gas holdup, bubbles rise velocity and the liquid circulation in 0.15 m diameter column using liquids viscosities range of 1-3 Pa.s. They noticed small bubbles of <1 mm diameters, which may contribute about 50% of the gas holdup. These small bubbles are seen to accumulate in the pipe due to their low rising velocity. Less than 1% of the injected gas to the system appeared to convert to small bubbles in high viscosity liquids. These small bubbles generate due to the coalescence of the large bubbles at the gas injection points and due to the bubbles eruption at the top surface of the column. The void fraction, due to small bubbles in slug flow, increases with increasing the liquid viscosity which in turn decreases the velocity of the slug units. This low velocity of the liquid produces less circulation for the liquid in the column due to low rise velocity of the liquid slugs. The void fraction (the small bubble concentration) increases also with increasing the time of the gas injection and the gas flow rate. If considering the big difference in the liquids viscosities between the present work and the work by Philip et al. (1990), similar behaviour is seen to occur in the present study. The concentration of the small bubbles increases with increasing the time of gas injection as well as increasing the gas flow rate. However, a different conclusion was reported for the void fraction of the small bubbles in gas-viscous liquids flow by Kuncová and Zahradník (1995). They stated that the small bubbles holdup increases with an increasing liquid viscosity, while it decreases with an

increasing gas flow rate. Similar studies in literature focused on the behaviour of the small bubbles in gas-liquid flow by Azzopardi and Zaidi (2000), Schäfer et al. (2002), Mena et al. (2005), Majumder et al. (2006), and Bröder and Sommerfeld (2007).

The size distribution of the small bubbles was studied by using digital image methods by a number of researchers. Lau et al. (2013) and Aoyama et al. (2016) studied the small bubbles size distribution, gas holdup and the shapes of the single bubbles rising in stagnant liquids of different viscosities. In both studies, a method by Otsu (1979) was used for the small bubbles image processing. The method is described in detail in chapter 3. The small bubbles size distributions were also studied by (Lage and Esposito (1999). Wongsuchoto et al. (2003), Mandal et al. (2005), Bordel et al. (2006), and Montante et al. (2008)).

The rising velocity of a single bubble in the stagnant liquid is controlled mainly by the buoyant and drag forces. These forces are dependent on the bubble diameter, fluid properties, and gravity. In high viscosity liquids, total drag force of sphere is a result of the viscous force at very small Reynolds number $Re \leq 0.1$. On the other hand, the buoyant force, which is mainly a factor of the fluids densities and bubble diameter, is very small comparing with the viscous force. Therefore, in the very high viscosity liquids, the terminal velocity of the individual small bubble is mainly controlled by the viscosity force. (Gaddis and Vogelpohl (1986), Tomiyama (2002), and Talaia (2007))

In two-phase systems, the motion of a particle at high population is affected by the presence of the surrounding particles. Stokes' law can be applied considering the small bubbles as particles to determine their suspension density in the liquid (Rhodes (2008)). The effective suspension viscosity, μ_e

and the effective average suspension density, ρ_{ave} can be calculated from:

$$\mu_e = \frac{\mu}{\varepsilon_l} , \quad 2.29$$

$$\rho_{ave} = \varepsilon_l \rho_l + (1 - \varepsilon_l) \rho_g , \quad 2.30$$

where: μ is the dynamic viscosity of the liquid, ε_l is the volume fraction occupied by the liquid. ρ_l and ρ_g are the density of the liquid and the gas respectively. The terminal velocity, U_T of an isolated small bubble rising in stagnant the liquid was determined by Stokes (1880);

$$U_T = \frac{1}{18} \frac{g d_b^2}{\mu} (\rho_l - \rho_g) , \quad 2.31$$

d_b is the diameter of the spherical small bubbles in stagnant high viscosity liquid. g is the gravitational acceleration. The rising velocity of the small bubbles in stagnant liquids was studied by (Hartunian and Sears (1957), White and Beardmore (1962), Bhaga and Weber (1981), Tomiyama et al. (2002), Talaia (2007) and Acuña and Finch (2010)).

2.9 Electrical Capacitance Tomography

Capacitance method is one of the common techniques used for void fraction measurement in the gas-liquid flow. The idea is to use a pair of electrodes around the pipe wall and determine the electrical capacitance in between. The electrodes can be placed inside or outside the test section according to the sensitivity and response time requirements. They use a "dielectric constant difference" of each liquid and gas. The advantage of this technique can be summed up in: it is a non-intrusive method, it is able to capture fast transient phenomena, and its cost efficient. Using the relationship between the phase distribution and the electrical capacitance, the average cross-section void fraction can be determined. The very first capacitance sensor

was used by (Gregory and Mattar (1973)). They deduced a pair of helical capacitor planes after testing a number of designs. Basing on the pipe diameter, they recommended using a specific dimension and inclinations for the capacitor electrodes. Another design developed by Kouba (1986) consisted of three electrode rings located equally on the pipe wall. His results seemed to be similar to the helical sensor results. However, the overall length of the sensor was shorter than the previous particularly in the case of large diameter pipes.

In slug flow, the parallel ring sensor appeared to be better than the helical sensor because the instantaneous values of void fraction were not averaged along the pipe section. However, the helical wrap could be more flexible than the parallel rings. Its length can be increased by adding more turns and increasing the length of the electrodes. For this reason, it would be more applicable in the case of the measurements that demand longer test section. The helical sensor is described in more details in a study by Siemens (1992). A comparison between 2 rings, 3 rings, vertical split, and horizontal split models was carried out by Butler et al. (1995). They showed, experimentally, that in the case of disconnecting one of the outer rings, a small change occurred in the sensor output. Then, Elkow and Rezkallah (1996) pointed out that a new capacitance sensor is required to be designed due to the implications of the helical sensor. These include the poor sensitivity, poor shielding as well as the nonlinear response of the sensor.

The Electrical Capacitance Tomography (ECT) is a non-intrusive technique, which is used for reconstructing images and flow velocity measurement in the two-phase gas-liquid flow. It has the ability to visualise the cross-sectional permittivity distribution of the two-phase based on the capacitance inside the dielectric pipes. The advantages of this technique can be

summarised as following: it is a non-intrusive sensor that can be located outside the test section (in the case of the non-metallic pipes); low manufacturing cost and flexibility for different size pipes; finally, high data acquisition frequency. (Huang et al. (1989), Xie et al. (1989), Dyakowski (1996), and Isaksen (1996)).

ECT can be used as gas-liquid such it is the case in the present work (Huang et al. (1989), Xie et al. (1989), Jeanmeure et al. (2002) and Gamio et al. (2005)). It is also used for gas-solid investigations (Jaworski and Dyakowski (2001)) and Azzopardi B.J. (2008)), liquid-liquid such as water-oil, (Bolton et al., 1999), kerosene-water (Hasan and Azzopardi (2007)), And gas-liquid-liquid such as gas-oil-tap water (Johansen et al. (1996) and Mwambela and Johansen (2001)).

2.10 Conclusions

From the survey of the relevant published materials presented in this chapter, it can be concluded that it is dominated by employing water or liquids with low viscosity and high surface tension. Only a few studies carried out using high viscosity liquids such as Azzopardi et al. (2014b) who used glucose syrup of 300 Pa.s viscosity, 0.08 N/m surface tension and the ECT sensor. This work followed an earlier paper by Pioli et al. (2012) who studied the dynamics of the two-phase flow of basaltic magmas. Different liquids of viscosity ranging between 3 to 300 Pa.s were used over a wide range of gas flow rates in different diameter columns using pressure transmitters. On the other hand, Viana et al. (2003) presented a universal correlation to determine the rise velocity of Taylor bubbles. They employed liquids of 0.001-3.9 Pa.s viscosity and 76.2 mm diameter to study the characteristics of Taylor bubble rising in different viscosity liquids. These three studies above used almost the highest viscosity liquid among the other works of gas-liquid

Chapter Two: Literature Review

flow in literature so far. Table 2.1 shows a number of the previous works on two-phase flow carried out using different liquid viscosities and pipe diameters.

Table 2.1: Summary of previous work carried out employing liquid of different viscosity, Reynolds number, and column diameter.

Reference	Liquid viscosity Pa.s	Reynold number	Pipe diameter mm
Khatib and Richardson (1984)	0.001	388-291000	38.8
Pandit et al. (1987)	1.5	1.4-11.4	152
Philip et al. (1990)	1-3	130-200	150
Spisak and Idzik (1994)	4.8	0.4-0.9	25
Fukano and Furukawa (1998).	0.007	40721-203608	26
Karimi and Kawaji (2000)	0.00164	1391-6584	50.8
Van Hout et al. (2002)	0.001	4350	25
Bugg and Saad (2002)	0.084	27	19
Viana et al. (2003)	0.001-3.9	2.69-21445	76.2
James et al. (2004)	0.001-0.9	9-10 ⁴	38
Nogueira et al. (2006)	0.001-1.5	0.5-1403	32
Schmidt et al. (2008)	7-9	0-1.2x10 ⁵	0.0545
Bird et al. (2010)	0.31	10 ⁻⁵ -10 ⁴	_____
Del Bello et al. (2012)	10-10 ⁴	40-4000	3000,6000
Pioli et al. (2012)	3-300	1-10 ⁶	240
Abdulkadir et al. (2014)	0.0053	4778-15359	67
Azzopardi et al. (2014a)	0.001	2680-36850	67
Azzopardi et al. (2014b)	300	0.058-0.812	240

There is a distinctive lack of information regarding the flow characteristics in systems involving gas and high viscosity liquids despite the importance of such systems in the oil and polymers industries besides the knowledge of volcanology. Besides, no works have so far introduced churn flow in liquids about 360 Pa.s. In addition, no works have been reported on the bubble formation using single and different configuration of the gas injection using

such viscous oil. Finally, there are no studies on the generation, dispersion and rise of small bubbles in high viscosity (330-360 Pa.s) liquids. The present work aims to provide new information regarding the gas flow in very high viscosity liquids and large diameter columns in addition to introducing churn flow in such high viscosity liquids.

Chapter 3

Experimental Arrangements and Methodologies

3.1 Introduction

This chapter includes the procedure of the experiments that was carried out to study the characteristics of gas-high viscosity liquids flow. It also contains the equipment used in the high viscosity oils experiments such as the sensors and flow indicators. The design of both rigs, fluid properties, description of experimental facilities and the methods of the experiments and analysis are also presented. The calibration of the equipment that is employed in the present work is shown in Appendix A.

3.2 High Viscosity Oil Rigs

In the present work, two rigs of similar design are used to study the characteristics of gas-high viscosity liquids flow in large diameter columns. Each rig consists of a large vertical column open to the atmosphere. The columns contain Silicone oils with viscosity of 360 and 330 Pa.s. The ECT sensor is placed at about 2.56 m from the gas inlet section in the 240 mm diameter column in order to allow full development of the flow structure. In the 290 mm diameter column the ECT sensor is placed at 2.4 m from the gas inlet section in the bottom of the column for the same reason. The sensor is used for the measurement of the time of cross-sectional phase distribution in the column. The ECT sensor is connected to a TFLR 5000 electronic box and a PC for data acquisition and processing. The advanced ECT system Tomoflow R5000 is developed for more accurate measurements including

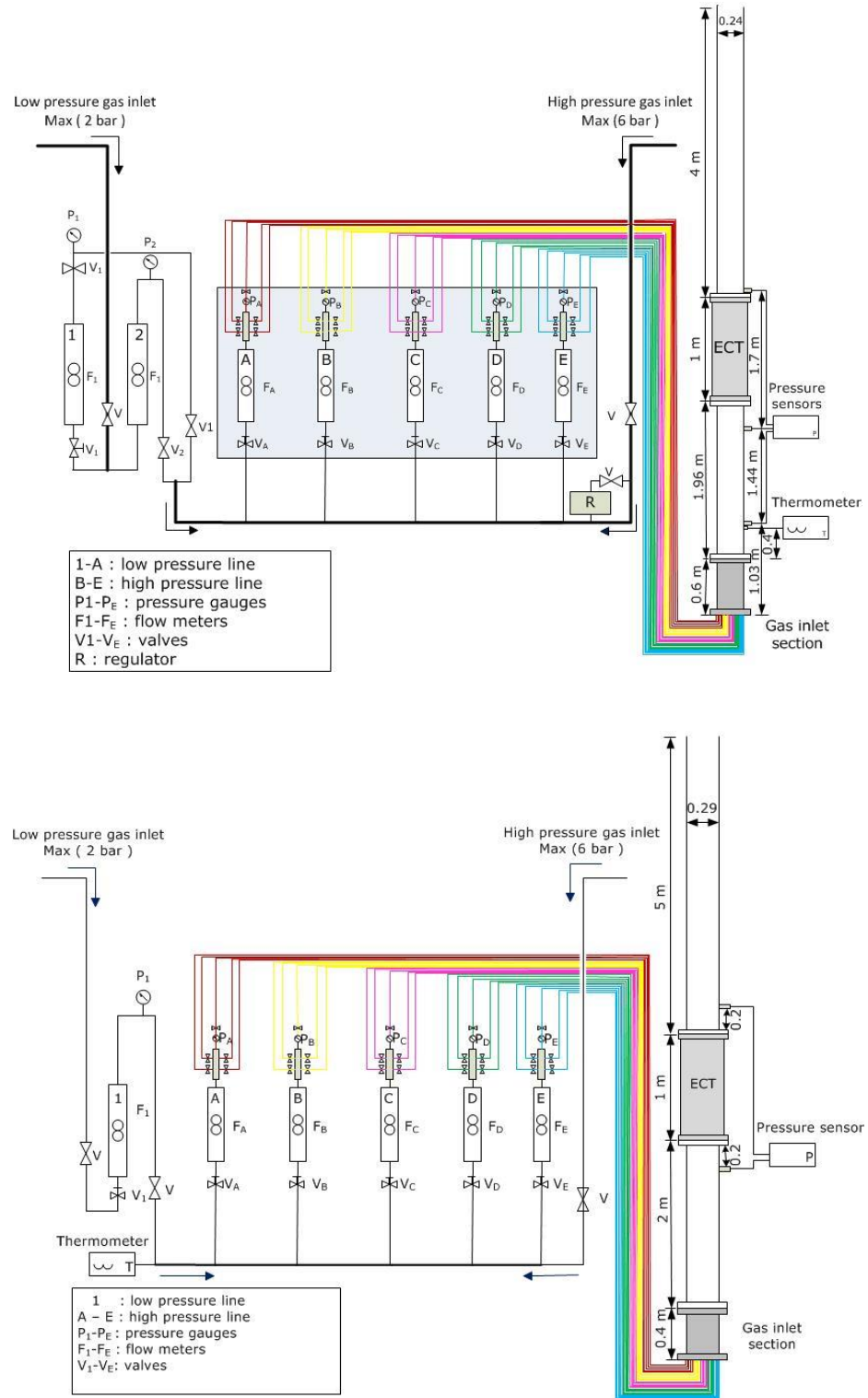


Figure 3.1: The two setups employed to study the gas-high viscosity liquid flow in large diameter pipes. The coloured gas lines correspond to the colours of the gas injection points in Figure 3.2.

concentration, velocity and phase distribution of the mixture in the pipe cross section. It has an excitation frequency of 10 MHz and maximum capture rate

of 5000 frames per second. These measurement values within the TFLR 5000 can be constructed and set in the control PC using a software. Figure 3.1 displays both rigs which are used in the present work showing the gas injection arrangements, instrumentation, and the dimensions. The rig is also provided with a temperature measurement (thermocouple) to indicate the temperature of the air at the gas lines besides the air-oil mixture in the column. Both columns contain stagnant Silicone oil with initial heights of about 3.30 m. The properties of the fluids can be found in the table in the next section in this chapter. The gas injection system consists of two main gas lines to obtain a wide range of gas flow rates, high pressure of 6 bar and low-pressure of 2 bar. The high-pressure line is connected to 5 flow meters and pressure gauges. Then each line is divided into 5 more lines. Thus, the total number of the gas injection lines is 25 lines connected to the bottom section of the columns. These 25 gas injection points are distributed equally at the bottom of the column. Figure 3.2 shows the distribution of the gas injection points at the bottom of both columns.

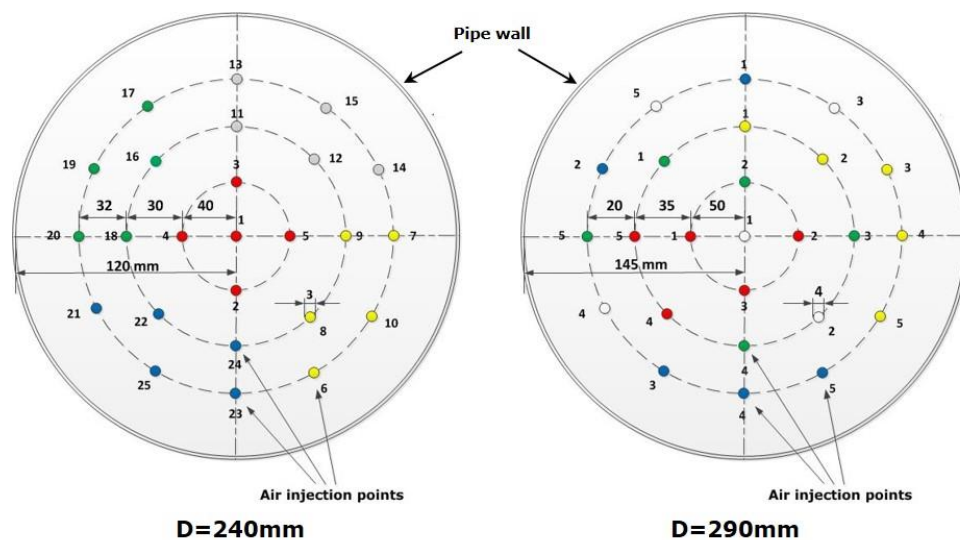


Figure 3.2: Configuration of the gas inlet section at the bottom of the columns. The colour of injection points corresponds to the colours of the tubes in both rigs. The numbers at the inlet section correspond to the numbering on valves on the coloured tubes in Figure 3.1.

The 5 colours of the gas injection points correspond to the 5 different colours

of the gas lines in both rigs. The gas injection system is designed in a way that allows a full control of the gas flow inside the column by changing the number and the position of the gas injection points. The low-pressure gas line has the same structure as the high-pressure line in the rig. Figure A.1 in Appendix A displays the photos of the 240 mm diameter high viscosity oil rig showing the flow meters and the ECT system. The figure also shows the top view (inner section) of the gas inlet section with the 25 gas nozzles. And finally, a side view photo of the gas inlet section, including the 25 gas inlet tubes, is also shown in the same figure.

3.2.1 Fluid Properties

Two types of fluids were used in this study, stagnant Silicone oil and air. Silicone oil was chosen because it is transparent, non-toxic, high flash point, low solidifying point, cost effective and excellent dielectric properties for the ECT. The fluids and their properties are displayed in Table 3.1.

Table 3.1: The properties of the fluids

Fluid (1)	Silicone oil					
Rig	Column diameter m	Viscosity Pa.s	Density Kg/m ³	Surface tension N/m	Eötvös number $G\rho D^2/\delta$	Buoyancy Reynold number
1	0.24	360	950	0.02	26812	0.97
2	0.29	330	950	0.02	39148	1.4
Fluid (2)	Air					
Rig	Column diameter m	Gas injections diameters mm	Density Kg/m ³	Low-pressure gas line bar	High-pressure gas line bar	Temperature °C
1	0.24	3	1.225	2	6	20-23
2	0.29	4	1.225	2	6	19-22.5

3.2.2 Flow Facilities and Measurements

Both rigs are supplied by 2 gas lines from the laboratory compressed air

main supply. The 2 main gas lines of 2 and 6 bars are used for low and high gas flow rates respectively. Both main lines in both rigs are connected to 5 Rotameters and pressure gauges, which indicate a wide range of flow measurements. Each line of the 5 is also divided into more 5 lines, which can be controlled by valves separately. This allows a full control of the number and the location of the gas injection points which can be used to inject gas into the column. Thus the total number of the gas lines in each rig is 25 lines. These lines are divided into groups of 5 similar colours in order to locate the injection points. The 25 valves are also numbered, corresponding to the numbers on the gas inlet section as it is illustrated in the figures in Section 3.2 and Figure A.2 in Appendix A. Moreover, 3 additional Rotameters and pressure gauges are installed on both rigs (2 on the first rig and 1 on the second rig). These flow meters and pressure gauges are used with the low-pressure line for low gas flow rates. They are located between the 2 bar main line and the 25 gas injection lines. The purpose of installing these indicators is to obtain a wide range of gas flow rates including the very low gas flow rates. The gas flow rate was adjusted and measured using the following flow meters; Fisher 2000 (10-50 L/min), KDN MOBREY 2000 (50-100 L/min), Solartron Mobry (10-100L/min) and MPB industries LTD 1750 (100-1500 L/min) Rotameters. A correction equation, from the manufacturer, was applied to correct the flow rates using the values of the pressure and the temperatures obtained for each flow rate. Thermocouple REX-D100, TC LTD and Fisher gauges were used for this purpose. Figure A.2 in Appendix A displays the Rotameters, pressure gauges and the 25 gas lines described above.

3.2.3 Electrical Capacitance Tomography

The Electrical Capacitance Tomography System (ECT), which is used for the time series of void fraction measurements, consists of: an ECT (the sensor), an electronic box used for data acquisition, and finally, a computer used for data processing and image reconstruction. This non-intrusive sensor measures the cross section distribution of capacitance in the column, which can be correlated to the cross section phase distribution for the non-conductive liquids. In this work, the TFLR 5000, ect 32V2 was used for data acquisition in the viscous oil column. Figure A.3 in Appendix A shows the ECT sensor placed on the rig connected to the electronic box. The ECT sensor, in Figure 3.3, consists of two planes; each plane contains eight measurement electrodes radially located around the pipe.

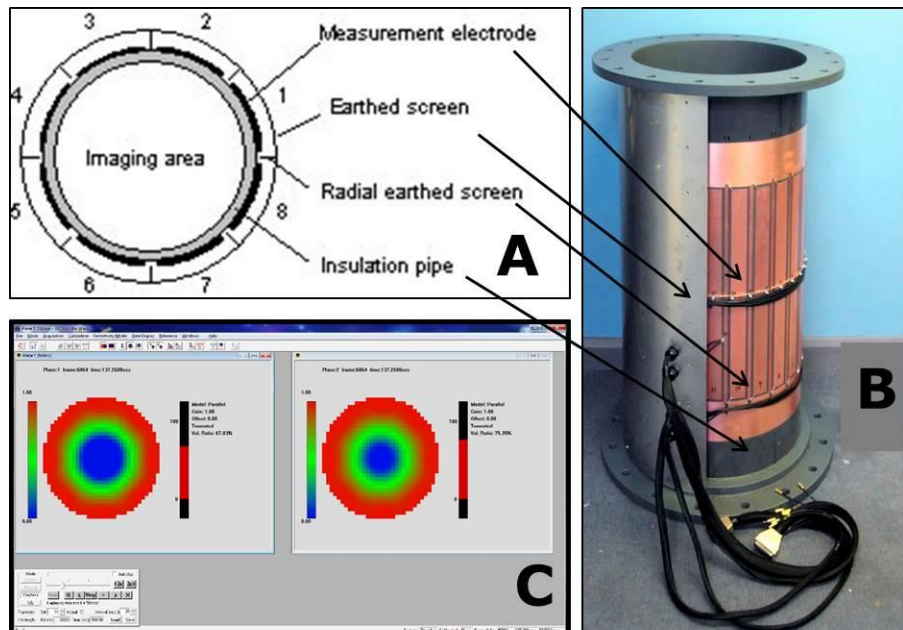


Figure 3.3: A is the cross section of the ECT electrodes showing the 8 electrodes (one plane). B is ECT sensor showing the electrodes mounted around the pipe after removing the outer screen. C is the ECT software showing the cross section distribution of the air-Silicone oil mixture in the column. The red colour represents the Silicone oil, the blue colour is the air, and the green colour is the interface between both fluids.

In order to protect the electrical field of these electrodes, equal numbers of guarded electrodes are located on each side. This pair of planes is used as a

non-intrusive sensor to measure the structure velocities by cross correlating the output signal from plane to another. The 16 electrode sensor uses the two planes of 36 mm spacing to measure the structure velocity in the column. The TFLR 5000 electronic box was connected to the sensor for the data acquisition and control software PTL (ECT 32) for reconstructing the image and for the adjustments such as time and frequency. The sensor is capable of measuring up to 10 MHz. The measurement precision of the ECT depends on many factors such as the properties of the fluid and the calibration conditions. The appropriate method to measure the accuracy is to set up a bench test which is almost impossible due to the very high viscosity of the oil used in this work. The validation of the ECT was checked by comparing the results with the Wire Mesh Sensor (WMS) by Azzopardi et al. (2010). The WMS was tested against the gamma densitometer and approved by Sharaf et al. (2011). Detailed information about the ECT is described by Abdulkareem (2011).

The ECT was calibrated by obtaining measurements for the low permittivity fluid ones when it was empty (gas only) and next when it was filled with high permittivity fluid (Silicone oil only). In the second measurements (high permittivity), the Silicone oil was left for a certain time to allow the small trapped bubbles to rise. These calibration files were then used in the dynamic measurements since the ECT output images are obtained from permittivity difference between the two phases in the gas-liquid flow. Then, the calibration procedure should be undertaken properly due to the difference in permittivity between the oil and gas phases to get a better image. For example, during the full section test section calibration (full pipe) some of the small gas bubbles were trapped in the Silicone oil due to the high viscosity. In order to obtain an accurate measurement, calibration file reading was taken before injecting the gas, and the same file was uploaded

for each experiment. The same procedure was followed for the low permittivity file (gas only). The high viscosity of the Silicone oil makes it very difficult to empty the test section for each set of experiment for the calibration. However, using the saved calibration files could be efficient when taking in consideration that the change in temperature does not affect the permittivity, Belt (2007).

3.2.4 Viscosity Measurement

The viscosity of the Silicone oils which used in both columns was tested by using the Viscometer DV-II+Pro. This Viscometer measures the oil viscosity at certain shear rates (Viscosity is a measure of the resistance of the fluid to flow). In the bench test, the effect of temperature and the trapped small bubbles were also examined. The oils were tested in the temperature range of (12-33 °C) using a rotating spindle. The viscosity of the oil was measured in the absence of the small bubbles once, then another test was carried out in their presence. The photo of the bench test is displayed in Figure 3.4.

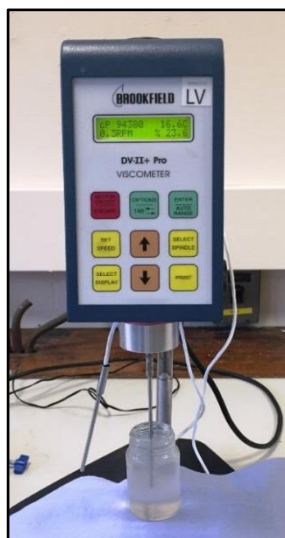


Figure 3.4: Photo of the bench test to measure the viscosity of the Silicone oils used in both columns using the Viscometer DV-II+Pro.

The photo displays the Viscometer that was used in the bench viscosity test with a sample of the oil employed in the experiments. The results of this test

are shown in the next chapter, Viscosity-temperature tests for the Silicone oils used in the experiments of the 240 and 290 mm diameter columns. The results showed a decrease in the viscosity of the oils with increasing temperature. It also showed a decrease in viscosity in the presence of the small bubbles.

3.2.5 High-Resolution Visualization Technique

Two types of cameras were employed in the present work to determine the size and the velocity of the large and the small bubbles in the 290 mm diameter column. First, the high-resolution camera, Canon EOS 600D, 18-megapixel resolution, which was used for large and small bubbles experiments. This camera was employed to determine the size of the large bubbles rising in the 330 Pa.s Silicone oil for different gas flow rates. The camera was also used to determine the velocity, size and concentration of the small bubbles in stagnant viscous Silicone oil. Photos of the high-resolution camera are shown with the experimental set up for the small bubbles rise experiments in the next section. The second camera is the Dantec Dynamics high-speed camera (Vision research, Phantom V12:20-30 VDC), EX SIGMA 105mm 1:2.8 DG MACRO. This camera was used to determine the velocity and the size of the large bubbles rising in the 290 mm diameter column for different gas flow rates. Figure A.8 in Appendix A shows the high-speed camera and the software used for data processing on the PC. A large spherical bubble rising at a flow rate of 10 L/min in the 330 Pa.s Silicone oil in 290 mm diameter column was captured by the camera. The collected photos from both cameras were analysed using ImageJ software. This software is able to determine the bubbles dimensions after setting up the calibration values for the images. The images were calibrated using measuring tape placed on the column.

3.3 Operating Procedure

Operating systems of high viscosity liquids have different requirements compared to the common low viscosity liquids which are used in the majority of gas-liquid flow studies. Running high viscosity oils experiments require more time for flow development at the beginning of the experiment as well as during changing the gas flow rates. Usually, the gas lines get blocked by the viscous oil due to the high back pressure of the oil in the column. Thus, more time is required to push the oil and clear the gas lines at the beginning of the experiments. Moreover, at the beginning of each experiment, the low-pressure gas line should be used to clear the 25 pipes. This will allow the creation of less amount of small bubbles, which increase the void fraction and also for safety reasons. The experiments also require longer data acquisition time because of the characteristics of the air flow in high viscosity liquids. The bubbles rise at low velocities compared to lower viscosity liquids. For this reason, a lower sampling rate and longer time acquisition are required in such viscous oils. For example, 50 Hz can be considered a reasonable frequency for the data acquisition for the oils which were used in this work. During the experiments, a longer time is also required when changing the gas flow rates. This allows a full development for the next flow rate, particularly when the transition between the flow patterns occurs. At very high gas flow rates, the air makes a very loud sound when flows through the passage of the oil column. As for the experiments of the methods of gas injection, the number and the position of the gas nozzles can be controlled by the 25 small valves which are located after the 5 Rotameters. All the valves are numbered and the gas lines are coloured to allow flexibility of the gas injection methods. The experimental procedure is concluded in:

1. Check for any crack or leak on the column.

2. Check the gas injection points at the bottom of the column.
3. Check the main valves of the high and low-pressure lines; they should be closed as they are designed to be used separately, but they both are connected to the same flow meters.
4. Start the gas injection with the lowest values of gas flow rates (2 bar line). This can be done first by opening the small valves then opening the main valve. Next, allow a time of about 10 min to clear the lines from the trapped Silicone oil.
5. Operate the ECT box and software by setting up the sample rate and the time (50 Hz and 600 Sec for most of the experiments in this work).
6. Set up the flow rate and start recording the data using the ECT software and save the files after each run.
7. For each run, the temperature and the pressure should be monitored to be used for the correction of the flow rates values. The height of the air-Silicone oil mixture might also be measured for the calculations of the top surface oscillation.
8. Change the flow rates and wait for about 5 minutes before the next data acquisition.
9. Close the low-pressure line before starting using the high-pressure gas line.
10. For better results, the values of the flow rates might be distributed equally at the bottom of the column by dividing the flow rate between the 5 Rotameters equally.
11. The value of the maximum gas flow rate can be selected just before the column starts to vibrate for safety.
12. At the end of the experiments, the main valves should be closed, so the small 25 valves will release the back pressure.

13. To obtain more accurate results, the rig might be left for about 5 days before running the next set of experiments. This duration is nearly enough to allow the small bubbles, which trapped in the column, to rise and clear the oil.

3.4 High Viscosity Oil Experiments

This section shows the experimental arrangement for all experiments of high viscosity oils in the present work. Gas flow conditions, data acquisition arrangement, and gas injection methods are displayed in the tables in this chapter. ECT was used to collect the measurements of the cross-sectional distribution of the gas-liquid besides the visualisation technique used for the measurements of the dimension of the bubbles. Values of temperature and pressure were collected for all experiments and used convert Rotameter readings to volumetric flow rates.

3.4.1 Experiments to Monitor the Effect of Gas Flow Rates

This section includes the experiments conducted to study the effect of changing gas flow rates on the structure of the flow in high viscosity oils. Two rigs with different diameter columns were employed, see Section 3.2. Table 3.2 illustrates the ranges of the gas flow rates which were applied in the experiments of the effect of the gas flow rate on both rigs. It shows the range of the gas flow rates for the different flow patterns. The number of the gas injection points and the data acquisition conditions are also displayed in the table.

Table 3.2: The experimental setup of the effect of gas flow rate on the structure of the flow in high viscosity liquids experiments in both rigs. The U_{gs} in the table is the gas superficial velocity.

	240 mm diameter column	290 mm diameter column
Flow rates (All)	10-1400 L/min U_{gs} :0.005-0.314 m/s	10-3200 L/min U_{gs} :0.003-0.566 m/s
Bubbly flow	10-20 L/min U_{gs} :0.005-0.011 m/s	10-50 L/min U_{gs} :0.003-0.016 m/s
Slug flow	30-600 L/min U_{gs} :0.016-0.114 m/s	70-800 L/min U_{gs} :0.022-0.088m/s
Transition to churn flow	700-1400 L/min U_{gs} :0.127-0.314 m/s	1000-1800 L/min U_{gs} :0.108-0.243 m/s
Churn flow	—————	2000-3200 L/min U_{gs} :0.276-0.566 m/s
No. runs	21x4	22x3
No.gas injection points	5-25	5-25
Frame rates	50 Hz	50, 300, 500 Hz
Sampling time	600 Sec	600, 300, 180 Sec

3.4.2 Size of the Large Bubbles

The size of the bubbles was determined by using ECT and it was compared with the sizes calculated from equations by Gaddis and Vogelpohl (1986), Snabre and Magnifotcham (1998), and Davidson and Schuler (1997) at the same range of gas flow rates. A single gas injection, at the centre of the gas injection section, was used to create the bubbles at the bottom of the column, see Table 3.3.

A range of the gas flow rates was selected to create bubbles with diameters almost equal to $2/3$ of the diameter of the columns. This value was selected experimentally by increasing the gas flow rate to the point when the shape of the bubble changed from spherical to ellipsoidal. In general, bubbles rise in liquids with a spherical shape due to the effect of the surface tension force. Then, when the size of the bubbles increases, at higher gas flow rates, the effect of the pipe wall starts to appear. The bubbles, in this case, become

elongated/ellipsoidal due to the shearing force. Therefore, selected range was chosen to eliminate the effect of the pipe wall on the bubbles to match the bubbles which were calculated theoretically, as they assumed bubbles in space. Photos were taken using the high-resolution camera (Canon EOS 600D) to compare the volumes with the bubbles in the previous methods in the 290 mm diameter column.

Table 3.3: The experimental setup for the bubble size experiments in both rigs. U_{gs} is the gas superficial velocity.

	240 mm diameter column	290 mm diameter column
Flow rates (All)	2-15 L/min U_{gs} :0.0009-0.005 m/s	10-3200 L/min U_{gs} :0.003-0.026 m/s
No. gas injection points	1	1
Gas injection position	Centre	Centre
No. runs	5	6
Frame rates	50 Hz	50 Hz
Sampling time	600 Sec	600 Sec

The experimental setup of the bubbles size measurements is displayed in another experiment to determine the diameter of the bubbles using 2 experimental methods, ECT and high-resolution camera, which was carried out on the 290 mm diameter column. The results from the high-resolution camera (Canon EOS 600D) were corrected using a method by Rana et al. (2015) and also compared to the ECT results. Information about the experimental setup is shown in Table 3.4. The results of the bubble size experiments are shown in Sections 5.3 and 5.4 in Chapter 5.

Table 3.4: The experimental setup for the bubble size experiments in both rigs using the visualisation techniques. U_{gs} is gas superficial velocity.

	ECT	High-resolution camera
Flow rates (All)	10-100 L/min U_{gs} :0.003-0.035 m/s	10-100 L/min U_{gs} :0.003-0.03 m/s
Column diameter	290 mm	290 mm
No.gas injection points	5	5
No. runs	6	9
Frame rates	50 Hz	_____
Sampling time	600 Sec	_____

3.4.3 Pressure Gradient Experiments

The aim of this experiment is to study the effect of gas flow rate on the variation of pressure and the characteristics of the flow in each flow pattern inside both columns. A wide range of gas flow rates was used to inject gas into the stagnant high viscosities Silicone oils in order to create different flow regimes. The produced voltage signals from the transmitters were converted to pressure in bar using “LabViwe” software. The LabViwe block diagram is displayed in Figure A.9 in Appendix A.

The sealed pressure transmitters were mounted along both columns to indicate a continuous pressure outputs. In the 290 mm diameter column, the 2 Keller, PR-25/8797.1-5 transmitters are installed at 2.2 and 3.6 m from the gas inlet section. The pressure measurements were collected at a sampling rate of 50 Hz for 300 Sec at flow rate range of 10-3000 L/min. Whereas, in the 240 mm diameter column, experiments were carried out by Dr Laura Pioli from Section des Sciences de la Terre et de l’Environnement, Université de Genève, Switzerland using 3 transmitters. Table 3.5 illustrates

information about the pressure gradient experiments.

Table 3.5: The experimental setup for the pressure gradient experiments in both rigs.

	240 mm diameter column	290 mm diameter column
Gas superficial velocity (All)	0.0008-0.43 m/s	0.003-0.517 m/s
Liquid viscosity	360 Pa.s	330 Pa.s
Transmitters type	Keller PD-25	Keller, PR-25
No. of transmitters	3	2
Locations from the gas inlet section	1.02, 2.47, 4.17 m	2.2, 3.6 m
Transmitters sensitivities	0.1,0.1,0.02 bar/volt	0.4, 0.4 bar/volt
No.gas injection points	5-25	5-25
Frame rates	100 Hz	50 Hz

3.4.4 Experiments on Gas Injection Methods

The aim of these experiments is to study the effect of changing the configuration of the gas injection on flow structure. The positions of a single gas injection point were changed at constant gas flow rates to examine the effect of the pipe wall. The number of gas injection nozzles were changed to study the effect of coalescence between the bubbles at the gas inlet section.

3.4.4.1 Effects of Air Injection Positions on the Flow Structure

The experiments were carried out using a single air injection point (one nozzle out of the 25). Seven different positions at the bottom of the column were used to inject the gas. Figure 3.5 illustrates the configuration of the gas injection points at the bottom of the 240 and 290 mm diameter columns. The 25 injection points (all circles) correspond to the 25 small pipes in the

rig. The large circles in bold correspond to the location of the gas injection points which was used in this experiment. Numbers and colours on the gas injection points correspond to the similar ones of the gas lines in the rigs.

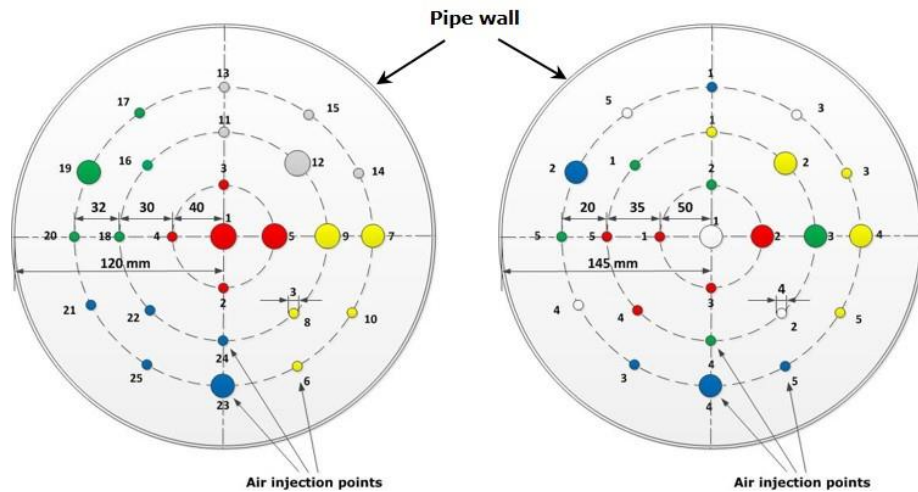


Figure 3.5: The distribution of the gas injections at the bottom section of the 240 and 290 mm diameter columns. The numbers and the colours correspond to the air pipes in the rig. The circles in bold correspond to the nozzles used in this experiment.

A constant gas flow rate was used for all positions in order to compare the effect of changing the positions of the gas inlet point. Tables 3.6 and 3.7 illustrate the experimental setup and the gas injection arrangements for the experiment respectively.

Table 3.6: The experimental setup of the effect of gas injection positions experiments in both rigs.

	240 mm diameter column	290 mm diameter column
Flow rates (All)	150 L/min U _{gs} :0.04 m/s	150 L/min U _{gs} :0.03 m/s
No. runs	7x3	7x3
No.gas injection points	1	1
Frame rates	50 Hz	50 Hz
Sampling time	600 Sec	600 Sec

Sampling rate of 50 Hz for 10 minutes was applied at the data acquisition

time for each position. The gas injection was initiated from the nozzle at the centre of the column then to the nozzles which are close to the pipe wall.

Table 3.7: The locations of the single gas injection points at the inlet sections at the bottom of both columns. R, W, Y, G, Y, and B represent the colours of the gas pipes (i.e. Red, White, ... etc). The numbers correspond to the gas injection points at the bottom of the column. See Figure 3.5, for example, R1 is the red circle at the centre of the 240 mm diameter column.

Runs	240 mm diameter column				290 mm diameter column			
	Distance between gas injectors and the centre of the column							
	0	40	70	102	0	50	85	105
1	R ₁				W ₁			
2		R ₅				R ₂		
3			Y ₉				G ₃	
4				Y ₇				Y ₄
5			W ₁₂				Y ₂	
6				B ₂₄				B ₄
7				G ₁₉				B ₂

Gas superficial velocities of 0.04 and 0.03 m/s were employed in the 240 and 290 mm diameter column respectively. These were the closest values which could be obtained experimentally because of the variation in dimensions of the two rigs. Moreover, these gas flow rates were chosen because they produce slug flow including less coalescence between the bubbles to show the effect of changing the position of gas injection points.

3.4.4.2 Number of Gas Injection Points

In this experiment, two types of arrangement for the gas injection were used to create the bubbles in both columns. Single injection point, which was located in the centre of the gas injection section, and 5 injection points distributed equally at the bottom of the columns. The aim is to examine the effect of the coalescence between the bubbles at the gas injection points on the flow structure when using more than one gas injection. Table 3.8 shows the experimental arrangements.

Table 3.8: The experimental setup of the effect of the number of the gas injection points experiments in both rigs.

	240 mm diameter column		290 mm diameter column	
No.gas injection points	1 nozzle	5 nozzles	1 nozzle	5 nozzles
Flow rates	10-300 L/M U _{gs} : 0.005-0.1 m/s	10-300 L/M U _{gs} :0.005-0.063 m/s	10-200 L/M U _{gs} :0.003-0.047 m/s	10-100 L/M U _{gs} :0.003-0.035 m/s
No. runs	8	10	7	7
Frame rates	50 Hz	50Hz	50Hz	50Hz
Sampling time	600 Sec	600 Sec	600 Sec	600 Sec

3.4.4.3 Distance Between Gas Injection Points

This experiment is to examine the effect of the distance between gas injection points on the flow structure in high viscosity oils. 3 different spacing between 2 gas injection points in parallel were applied. At the same time, the effect of the distance between gas injection points was studied at 3 different gas flow rates which created 3 different flow regimes. The flow regimes were bubbly, slug, and transition to churn in both columns. Figure 3.6 and Table 3.9 show the positions of the gas injection points which employed in this work for both columns.

Table 3.10 provides parameters for the experimental setup.

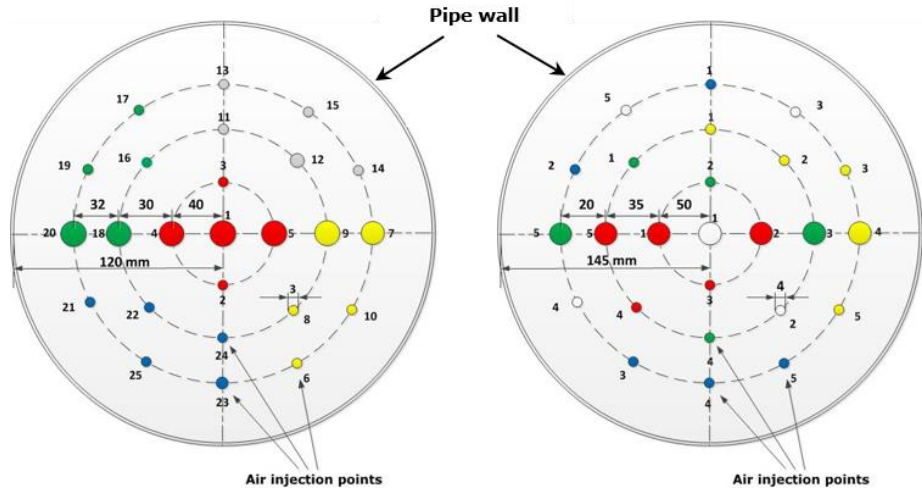


Figure 3.6: The distribution of the gas injections at the bottom section of the 240 and 290 mm diameter columns. The numbers and the colours correspond to the air pipes in the rig. The circles in bold correspond to the nozzles used in this experiment.

Table 3.9: The positions of gas injection points at the gas inlet sections at the bottom of both columns. R, W, Y, G, Y, and B represent the colours of the gas pipes (i.e. Red, White, ... etc). The numbers correspond to the gas injection points at the bottom of the columns. see Figure 3.6

Runs	240 mm diameter column			290 mm diameter column		
	Distance between gas injection points (mm)					
	80	140	204	100	170	210
1	R4&R5			R1&R2		
2		G18&Y9			R5&G3	
3			G20&Y7			G5&Y4

Table 3.10: The experimental setup of the effect of the distance between the gas injection points experiments in both rigs using 2 gas injection points in parallel.

	240 mm diameter column	290 mm diameter column
Flow rates (All)	10, 20, 300 L/min Ugs:0.005, 0.01, 0.1 m/s	20, 50, 450 L/min Ugs:0.006, 0.016, 0.1 m/s
No. runs	3x3	3x3
No.gas injection points	2	2
Frame rates	50 Hz	50 Hz
Sampling time	600 Sec	600 Sec

3.4.5 Effect of Gas Injection Time

This set of experiments was conducted so as to study the effect of the time of gas injection on the characteristics of the flow in high viscosity oils. The gas was injected at a constant flow rate for several hours and data was collected periodically using the ECT. Table 3.11 illustrates the experimental set up in both rigs.

Table 3.11: Experimental setup of the effect of gas injection time experiments in both rigs

	240 mm diameter column	290 mm diameter column
Flow rates (All)	150 L/min U _{gs} :0.026 m/s	50, 150, 500 L/min U _{gs} :0.015,0.047,0.178 m/s
No. runs	1 Data collected every 30 min	1x3 Data collected every 30 min
Experiment time	3.5 hours	6 hours
No.gas injection points	5	5
Frame rates	50 Hz	50 Hz
Sampling time	600 Sec	600 Sec

This experiment was repeated in the 290 mm diameter column, at different gas flow rates, to compare the behaviour at different flow regimes. The top surface oscillations of the air-Silicone oil mixture were measured and compared at different times and gas flow rates.

3.4.6 Small Bubbles Experiments

This set of experiments was carried out in order to study the characteristics of the small bubbles (centimetre to millimetre size) in high viscosity oils. The experiments were divided into two parts: generation and rise of the small bubbles. In the first one, the gas was injected at a gas superficial velocity of 0.006 m/s for 1 hour. Photos were taken at the top and bottom sections (about 60 mm from the bottom of the column) of the 290 mm diameter

column. Photos were captured every 10 minutes and the gas injection was stopped for every photo shooting. The size, velocity and rate of increasing concentration of the bubbles were determined using ImageJ software. The second experiment was the small bubbles rise in stagnant high viscosity liquid. Photos were taken at the same locations as the first experiment but, this time, the liquid was stagnant. Photos were taken for 5 days at the top section of the column. This is due to the very low rise velocity of small bubbles because of the very high viscosity of the Silicone oil. However, photos were taken for about 2 hours only at the bottom section of the column. This was the time required to clear the lower part of the column from the small bubbles. The size, velocity and rate of decreasing the population of the small bubbles were determined as well. High-resolution camera, Canon EOS 600D, 18-megapixel resolution was used with LED panel light PS0606-6W which was installed behind the column. Finally, horizontal and vertical resolutions of 72 dpi were obtained for the photos which were taken by the camera. Figure 3.7 displays the experimental setup.

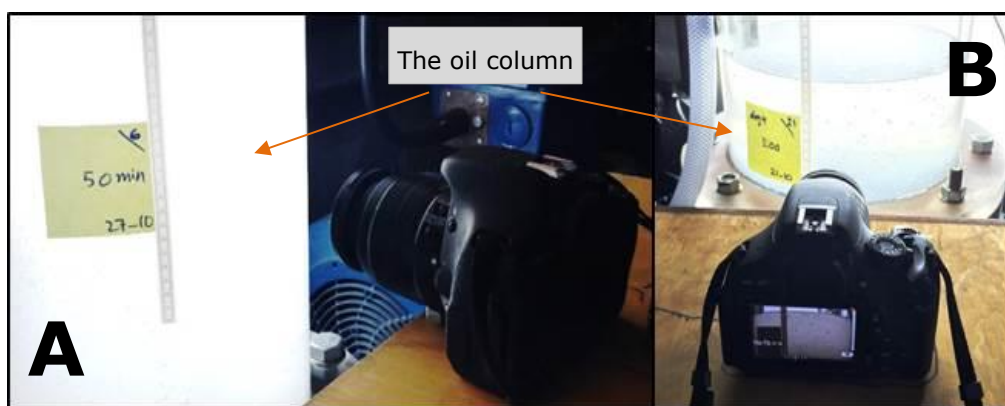


Figure 3.7: Experimental setup for the small bubbles rising in stagnant high viscosity oil experiments showing the high-resolution camera at the bottom (A) and the top (B) of the 290 mm diameter column.

3.5 Methods of Analysis

3.5.1 Large Bubble Experiments

In these experiments, the effect of gas flow rate on the characteristics of the flow in high viscosity oils was studied in two different large diameter columns. Data of time series of void fraction, obtained from the twin planes of the ECT, were used to study the structure of flow in the air-viscous Silicone oil. A wide range of gas flow rates was applied in 360 and 330 Pa.s Silicone oils in 240 and 290 mm diameter columns. PDF, mean void fraction, level swell, structure velocity, large bubble and liquid slug lengths, frequency, liquid film thickness, and time of churn were determined.

The PDF, which is the variation of the amplitude frequencies of void fraction, was determined by using the histogram method. The frequency of the void fractions of the large bubbles and the small bubbles in the liquid slugs was also determined. The PDF files were also determined for the bubbles and the churn areas individually. Data from the time series for the long bubbles was extracted from the time series of void fraction of the two-phase flow and separated from that in the churn regime. The PDFs values were calculated separately and, then, compared with the average values of the PDF.

The averaged void fraction for each gas flow rate was determined by calculating the values of the void fraction obtained from the time series. These values were also used to determine the falling liquid film thickness around the large bubbles. It was determined by applying the geometric relationship

$$\delta = \left(\frac{D}{2}\right)(1 - \sqrt{\varepsilon_g}) \quad (3.1)$$

Where δ is the film thickness, D is pipe diameter and ε_g is void fraction

obtained from the ECT.

The structure velocity was calculated from the cross-correlation of the void fraction signals which obtained from the two ECT planes. The time delay between the signals of the planes Δt was used in conjunction with the distance between the planes L_p to calculate the structure velocity, U_{st} .

$$U_{st} = L_p / \Delta t \quad (3.2)$$

Figure 3.8 shows 30 s of signals obtained from the ECT planes for the gas superficial velocity of 0.022 m/s in 330 Pa.s Silicone oil and 290 mm diameter column.

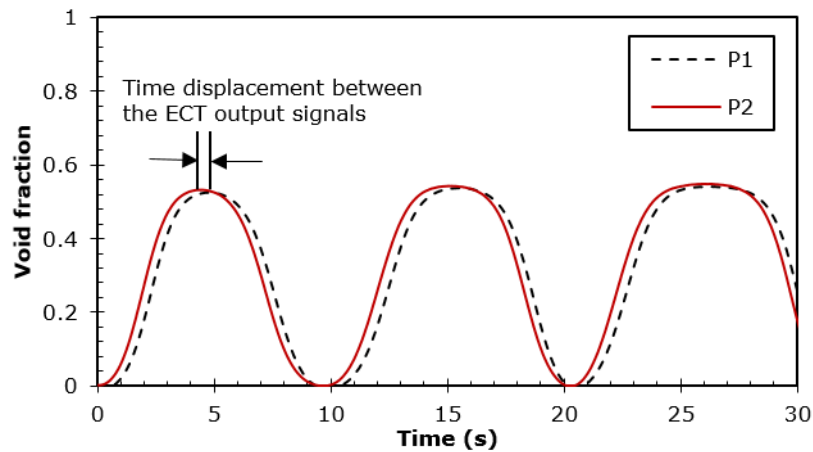


Figure 3.8: Time series of void fraction signals from both ECT planes for slug flow in 330 Pa.s Silicone oil and 290 mm diameter column. The gas superficial velocity is equal to 0.022 m/s.

The lengths of Taylor bubble and liquid slug were determined from the time series data of void fraction obtained from the ECT planes. They were determined by calculating the time of each bubble/liquid slug passing through the ECT planes for each bubble/liquid slug velocity.

$$L_B = U_B * t_B \quad (3.3)$$

$$L_S = U_S * t_S \quad (3.4)$$

Where L_B , U_B and t_B are Taylor bubble length, velocity and the time of the

bubble passing through the ECT planes. L_s , U_s and t_s are the liquid slug length, velocity and the time of the liquid slug passing through the ECT planes.

The Taylor bubbles and liquid slug lengths which obtained for 600 s were averaged for each gas flow rates and plotted with the gas superficial velocity of both columns. The experimental results were also compared with an approach by Khatib and Richardson (1984) which is explained in Chapter 2. The length of the individual large bubbles was also determined using data of void fraction time series obtained from the ECT. First, the time of each large bubble/liquid slug was extracted individually. Then the lengths were calculated by multiplying the time of each bubble by the structure velocity, which was determined from cross correlating the void fraction from the two planes of the ECT. The individual bubble lengths versus the time of each bubble for different gas flow rates is displayed in Chapter 5.

The frequency of the large bubbles was calculated from the time series data by counting the number of the bubbles per unit time for each gas flow rate. A method, which is developed by Hazuku et al. (2008), was applied to determine the large bubble frequency. This method consists of counting the number of the large liquid fractions in the vertical annular flow. This method was validated by Waltrich et al. (2013) by using visual observation. Figure 3.9 illustrates the method of counting the large bubbles in the present work using 2 threshold lines. The first line in red (lower) represents the mean values of void fraction and the second line in black is the mean void fraction of the values above of the first line. The areas in blue are the bubbles which pass through both lines. The advantage of this method is excluding liquid slugs and the churn (liquid bridges in churn flow) from the bubble frequency calculations, e.g., the churn area from 0 to 4000 frames at the time series

in the figure. However, a number of large bubbles could be excluded from the bubble counting, for example, the 3 bubbles at 1100 to 1200 frames in the figure. This is due to the low void fraction of these bubbles which less than the upper threshold line.

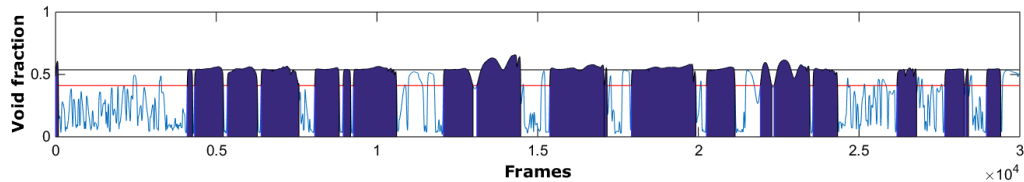


Figure 3.9: The time series of void fraction and the bubble counting using Hazuku et al. (2008) method. The bubbles are rising in 330 Pa.s Silicone oil and 290 mm diameter column at a gas superficial velocity of 0.139 m/s. Red is mean value of the void fraction, the black line is the mean void fraction of the values above of the red line, and areas in blue represent bubbles which pass through both lines.

Finally, the percentage of churn in Figures 4.11 and 7.7 was determined in both 240 and 290 mm diameter columns from the time series of void fraction. The time of the churns was extracted from the overall flow time (churn and bubbles). Then the overall churn time for each flow rate was divided by the total time of the flow to obtain the fraction of the churn.

3.5.2 Small Bubbles Experiments

The characteristics of the small bubbles in high viscosity liquids and large diameter columns have been studied in this work. The photos, collected at the top and the bottom of the column, were analysed according to the type of the experiments. First, in the bubble concentration analysis, photos of the column were obtained in the absence of the bubbles to obtain the background which will be used as a threshold. The RGB photos of the bubbles were inverted to greyscale images. Then they were converted to a binary image by using the threshold, applying the method by Otsu (1979). Using a Matlab code, the small bubbles were selected by specifying the number of objects. It labels each object (bubble) in the image with an integer value. Since the small bubbles have a spherical shape, a circle with the same area

of the small bubbles was fitted. This helped to obtain another set of useful data such as the area, major, and minor axes of the small bubbles. Then the area was converted to volume by multiplying by the thickness (depth) of the tested area. The thickness was selected from the average diameter of the small bubbles. The percentage of the determined volume of the small bubbles was calculated by dividing the volume of the bubbles by the total volume of the image. The calculated volume of the small bubbles is shown in Figures 8.3, 8.10 and 8.11.

Second, the diameters of the small bubbles were determined using the ImageJ software. This software enables to define the bubble boundary and calculate the area, circumference, minor axis, major axis, and the diameter of the selected objects. The images were calibrated by using a measuring tape placed on the column for scaling purpose. The diameters of about 95-57 bubbles were determined to calculate the average diameters of the small bubbles at each time. Figures 8.5, 8.14 and 8.15.

Finally, the rise velocity of the small bubbles was also determined by using the ImageJ software. Since ImageJ can locate the x and y-axis of the selected objects. Then it was possible to measure the rise velocity by tracing the single bubbles. Videos of the small bubbles rising in stagnant oil were recorded for 5 minutes during 5 days at the top section. Another video was recorded at the bottom of the column on the first day of the experiment. Then the videos were converted to photos of an equal time interval, using a Matlab code. Then, 8 bubbles were selected and each one was tracked for 6 positions. Then, the velocity was calculated by dividing the change in the position of each bubble by the time between both positions. Therefore, the calculated velocities were an average of 6 velocities of 8 different diameter bubbles, See Figures 8.16 and 8.19 in Chapter 8.

Chapter 4

Transitions Between the Flow Regimes

4.1 Introduction

This chapter presents the results for the whole range of experiments carried out to study the effect of the rate of air flow passing through a stagnant high viscosity oil. It also shows the change of the flow structures which occur in large vertical columns considering a wide range of gas superficial velocities. Further, it compares the behaviour of the flow in two different diameter columns (240 and 290mm). Photographs, used to illustrate the characteristics of the flow and images reconstructed from ECT measurements, are presented. In addition, the time series of the void fraction and the PDF files, which illustrate the structure of the flow in each gas flow rate and column diameter, are considered. The mean void fraction in the column from the ECT signals and the level swell are determined. Structure velocity, Taylor bubble and liquid slug lengths, and liquid film thickness are extracted and compared, for both columns, from the time series data. The pressure difference inside the column also is measured using two pressure sensors. This chapter provides a general overview showing all flow patterns seen for the range of the flow rates studied. The characteristic of each flow regime will be explained in detail in the next chapters.

4.2 The Characteristic of Air flow in Viscous Liquids

The flow of air in high viscosity liquids shows significantly different characteristic compared with water and lower viscosity liquids which are the

most commonly employed in gas-liquid studies. In these sets of experiments, a wide range of air flow rates was used to show the full picture of the flow and cover the whole flow regimes. Silicone oils with viscosities of 330 and 360 Pa.s were used in the two vertical columns of 290 and 240 mm diameters respectively. In the smaller column (240 mm), 21 runs with a range of 10-1400 L/min flow rate (this is equal to gas superficial velocities 0.005-0.314 m/s) were used. While in the larger column, a wider range of 10-3200 L/min corresponding to the gas superficial velocities of 0.003-0.566 m/s with 22 runs were applied. In both columns, 25 nozzles were used to inject the air into the bottom of the column. Details of the rigs and the gas inlet distributors are shown in Chapter 3. The aim of these experiments was to examine the effect of the changing gas flow rate on the flow structure inside the column.

4.2.1 Visual Observation and Image Reconstruction

Visual methods, which are the simplest way to observe the flow structure through the column, were the first step applied in collecting information. These approaches are more applicable for low fluid velocities. The photographic method becomes less efficient in case of high gas-liquid superficial velocities. Then the description of the flow becomes very difficult and is based on personal interpretations (Hewitt and Nicholls, 1969). However, Hewitt and Hall-Taylor (1970) pointed out that the visual observation method can be reliable if based on the different descriptions of the flow pattern in two-phase flow named by human observation, such as bubble, liquid slug, annular, mist, wave entertainment, and turbulent.

In general, in gas-viscous liquid flow, the gas flows mainly in large bubbles (Taylor bubble) which occupy almost the entire cross-sectional area for the majority of the gas flow rates. These large bubbles are accompanied by small

bubbles which range in size from very small bubbles of millimetres to large ones of centimetre size.

The large bubbles or Taylor bubbles, which can be seen in Figure 4.1, dominate the flow structure in such viscous oils. They rise in the centre of

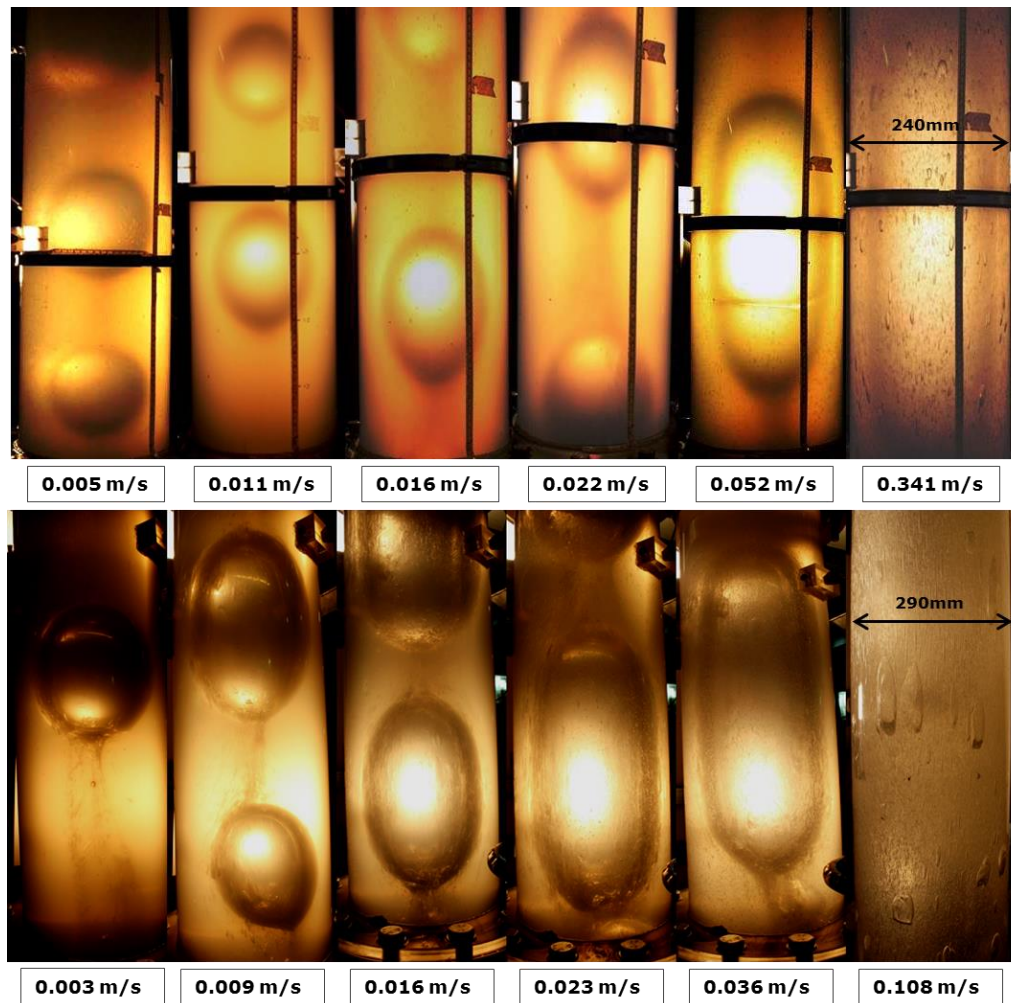


Figure 4.1: Photographic images show the effect of increasing gas flow rate along columns of 240 and 290 mm diameter stagnant 360 and 330 Pa.s Silicone oil and respectively.

the column because of wall effects. The shape of these bubbles is mainly dependent on the gas flow rate. In low gas superficial velocities (0.005 and 0.003 m/s), the bubble rises in constant velocity with spherical shape due to the surface tension of the liquid. The bubble, followed by a wake, contains a small number of small bubbles that rise in vertical lines and increase with increasing the gas flow rate. These small bubbles in the wake of the large

bubbles (liquid slug) were noted to be more plentiful in the larger columns for the same gas flow rate range. However, in the smaller column (240 mm), it seemed to be, almost, no small bubbles in the wake of the bubbles specifically in lower gas flow rates. Increasing the gas flow rate affects the structure of the liquid slug, which becomes more turbulent with small bubbles that either join the Taylor bubbles or remain as a liquid slug in the wake. The rate of increase of these small bubbles appeared to be not only dependent on the gas flow rate, but also on the time over which gas was injected into the column. The effect of time and flow rate on this type of bubbles in the liquid slug is illustrated in the constant velocity experiments results in Section 6.2 in Chapter 6. The shape of the large bubbles changes with increasing flow rate. First, they become ellipsoidal, due to the shear force, as 0.005-0.022 m/s in the 240 mm columns and 0.003-0.016 m/s in the larger column in the photos. Then, with gas superficial velocities above 0.022 m/s in the smaller column and more than 0.016 m/s in the larger column, the shape of the bubbles changes. They develop into long bubbles with rounded nose and end and no recirculation region in the wake. The diameter of the bubbles does not show any significant change, with the increasing gas flow rate, that can be seen clearly in the film thickness analysis in Section 4.2.7.2 later in this chapter. The falling film around Taylor bubble, which contains the small bubbles, show a different behaviour in high gas flow rates when the flow regime changes to churn flow. This will be explained in Chapter 7, Churn Flow.

The second type of bubbles in gas-viscous liquids flow is small bubbles. They are generated mainly from bubbles bursting at the top of the liquid in the column. They are also formed at the air injection nozzles at the bottom of the column, and also from the coalescence between the bubbles while rising in the column. The diameter of these bubbles varies between millimetres to

a few centimetres. They can be classified into two types of bubbles. The first one is a temporary small bubble of centimetres that can be seen mainly on the column wall. This type of bubble seems to be larger, more frequent, and less organised in shape in the larger diameter column, as it can be seen in Figure 4.1. They also increase in quantity with increasing gas flow rate and time. At the same time, the growth of these small bubbles occurs close to the gas inlet section in the column and lower part of the column rather than in the other sections of the column. They are created from the coalescence of the smaller bubbles and merge into Taylor bubble after a period of time. They have an ellipsoidal shape when they are small, but have a more complex shape when they increase in size. The second type of small bubbles in high viscosity liquids are millimetres in diameter and spherical in shape. This type of bubble is distributed all over the liquid in the column. They can be seen in the liquid slug between the Taylor bubbles as well in the falling film around Taylor bubbles. In a study carried out by Azzopardi et al. (2014b) to examine the characteristic of gas flow in high viscosity liquids, very small bubbles of ($\sim 100\ \mu\text{m}$) were reported. Glucose syrup of 300 Pa.s viscosity and 0.08 N/s surface tension was used in their experiment with a column diameter of 240mm. The behaviour of the flow structure they reported in their study generally agrees with that of the present work. However, in this work, the very small bubbles appeared to be quite large and visible to the naked eye. More detail about the character of these bubbles can be found in Chapter 8.

In Figure 4.1, 0.341 and 0.108 m/s gas superficial velocities show a different behaviour for the flow. As it is clear from the photo it is quite difficult to see the flow structure inside the column. The column in this stage is full of small bubbles which cover the entire wall and is not possible to recognise the large bubbles which become very long. In this case, the time series of the void

fraction and the PDF files can provide the information on the flow structure in the column. The structure of the flow changes significantly as the flow pattern goes from slug to churn. It starts from the flooding of the falling film around Taylor bubbles. The drained liquid in the flooded film moves in different directions during the transition. Then at the very high flow rate, where the churn flow starts, the air flows in a random shape conduit. Chapter 7 discusses churn flow in high viscosity liquids.

Figure 4.2 shows a schematic drawing for the flow patterns in high viscosity liquids and vertical large diameter pipes. It displays the structure of gas flow in 330 Pa.s Silicone oil and 290 mm diameter column. Descriptions of each flow regime in the figure are shown in details in Chapter 5 (bubbly and slug flow) and Chapter 7 (churn flow). Also, the original photos for the corresponding gas superficial velocities in this figure are displayed in Figure 4.1, and Chapter 7.

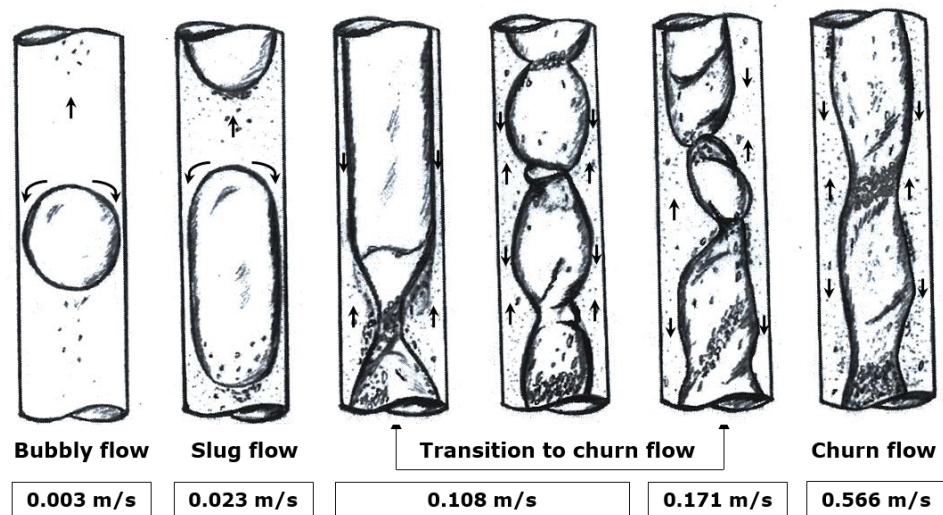


Figure 4.2: Schematic drawing showing the flow patterns in 330 Pa.s Silicone oil and 290 mm diameter vertical pipe. The arrows in the figure correspond to the direction of the liquid flow.

The internal structure of the air-Silicone oil mixture distribution inside the bubble column can be obtained by using the ECT image reconstruction technique. The ECT32 software allows the images to be viewed in both online

and offline mode. The Plot3D software is a stand-alone Matlab program captures the capacitance measurements by employing the PTL ECT32 software. The software displays sets of multiple two-dimensional images frames as 2 or 3 dimensions slices. The Plot3D software allows two-dimensional images to be reconstructed and displayed. The software uses a series of pixels to reconstruct the normalised permittivity distribution in the test section. It indicates the pixels' normalised permittivity by a range of a colour scale of red, green, and blue. The blue colour represents the lower normalised permittivity fluid, which is equal to zero in the calibration of the sensor. However, the red colour (pixel) corresponds to the fluid with the higher value of normalised permittivity that is equal to 1. The green colour represents the mixed area where there is both air and Silicone oil. The normalised permittivity distribution is proportional to the fractional concentration of the fluid with the higher permittivity. Figure 4.3 shows the single axial slices for 1000 frames of 30000 for different gas flow rates. The images are a reconstruction of air-Silicone oil 330 Pa.s viscosity in the 290 mm columns. These 3-D slices display 1000 frames of cross-sectional void fraction measurements for the gas-liquid flow in the column. These images cannot quantify the phase distribution, but they can produce a picture of the structure of the flow. As it is clear from the images, Taylor bubbles pass the measurement volume with a uniform shape. The shape of the bubbles is spherical or semi-spherical with a diameter close to the column diameter. This shows a similar result with photos 0.003 m/s in Figure 4.1. The blue represents the low permittivity fluid that is air, while the red is the Silicone oil. The structure of Taylor bubbles changes with increasing the gas superficial velocity. The bubbles get longer due to the coalescence between them. When the flow regime changes to churn flow, the nose of the bubbles keeps the rounded shape due to the shear force until a very high gas flow

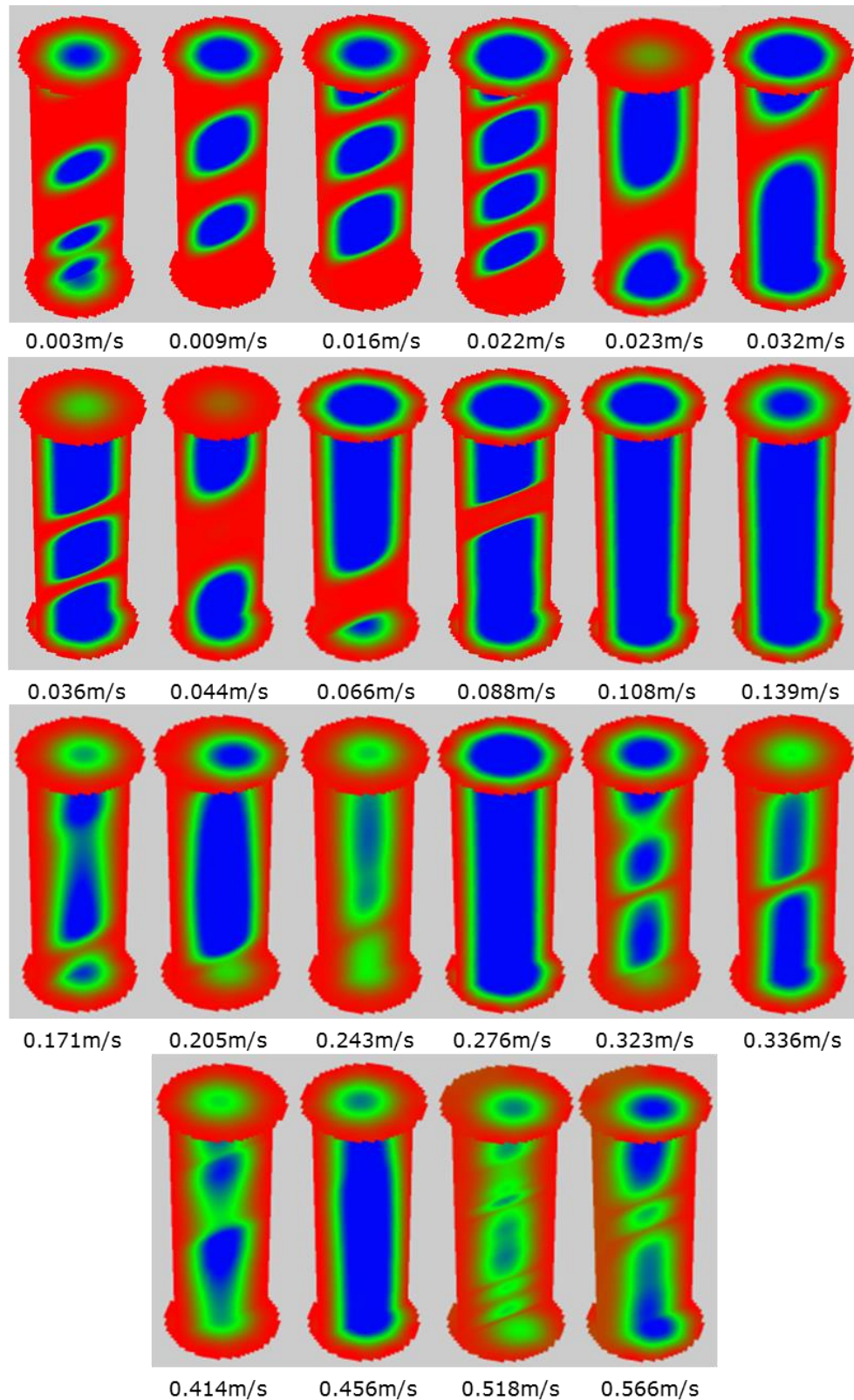


Figure 4.3: Reconstructed images of the air-viscous Silicone oil for different gas superficial velocities in 290mm diameter column. Blue, red and Green colours represent air, Silicone oil and the mixed area between both fluids respectively. The images are sets of multiple two-dimensional image frames as 3 D slices.

rate. Then, they seem to have almost a flat or slightly rounded shape. The

tail of the bubbles appears to be more affected by velocity. They turn flat with increasing the velocity and get a random structure due to the coalescence with the upstream Taylor bubbles. The wake is also affected by the change of the gas flow rate in terms of increasing the turbulence amount and the amount of the small bubbles in the falling film. According to the axial slices, the flow pattern changes first from bubbly to slug (0.003-0.088 m/s) then from transition to churn flow (0.108-0.566 m/s). The churn flow can be seen clearly in images for 0.414 and 0.518 m/s. The photos in Figure 4.1 show an agreement with the axial slices from the ECT in Figure 4.3 for the selected range of gas superficial velocities. The accurate results for the flow transitions can be found in the void fraction and PDF files in the next section.

4.2.2 Time Series of Cross-Sectional Averaged Void

Fraction

The time series of the cross-sectional averaged void fraction provides a useful way of examining the flow structure. The information about phase distribution was obtained by using the ECT, which provides time and cross-section resolved information. The time series illustrates the variation of the void fraction with time. In this work, an acquisition time of 600 seconds and a sampling rate of 50 frames per second were used for all runs, i.e., 2x30000 cross section per run. Figure 4.4 and 4.5 show the void fraction time series plots for each gas superficial velocity for all experiments on both rigs. The figures show a change in flow pattern from bubbly to transition to churn flow in the 240 mm diameter column in Figure 4.4 and from bubbly to churn flow in Figure 4.5 for the 290 mm diameter column. Time series of the void fraction for different flow patterns was presented by Costigan and Whalley (1997) for air-water flow in 32mm diameter pipe, see Figure 2.8.

In general, the maximum value of void fraction in both columns is

approximately 0.5 at most of the gas flow rates. This could be related to the diameter of the bubbles that do not exceed $2/3$ the diameter of the pipe. At low gas flow rates, the bubbles rise with a spherical shape, as no effect of the pipe wall due to the small size of the bubbles. Then, with increasing the gas flow rate, the bubbles elongate due to the shear force by the pipe wall. The effect of the pipe wall starts at that gas flow rate when the diameter of the bubble is approximately equal to $2/3$ the diameter of the pipe. Further increase in gas flow rates increases the length of the bubbles. At the same time, the diameter of the bubbles remains approximately constant after they become equal to $2/3$ of the pipe even at high gas flow rates. Therefore, the gas volume fraction of the gas does not exceed 0.5 in both columns.

Figure 4.4 illustrates the flow patterns in the 240 mm diameter column at 21 gas flow rates. It also shows how the flow patterns change with the gas superficial velocity. In the first two graphs 0.005 and 0.011 m/s, and at low gas superficial velocity, the bubbles are almost at the same size and the rise velocity is almost constant. They also appear to be as about the same length as the liquid slug. At this stage, the flow seems bubbly; it consists of one spherical bubble rise in the centre of the column at a constant velocity. By increasing the gas superficial velocity to 0.016 and 0.028 m/s, the bubbles become longer, but still have the uniform structure with less volume of liquid slug. In the next six plots 0.036-0.114 m/s, the length of Taylor bubbles increases dramatically due to the coalescence of these bubbles. The length of the slug decreases with increasing the gas flow rate, as it is clear in the figure, the time of the liquid slug decreases gradually. Therefore, the slug flow can be considered to occur in this range of flow rate. Transition to churn flow can be found in the rest of gas flow rates where a dramatic change in the structure of the flow was observed. This flow pattern starts with changing to unstable slug, as the Taylor bubbles are considerably longer than previous

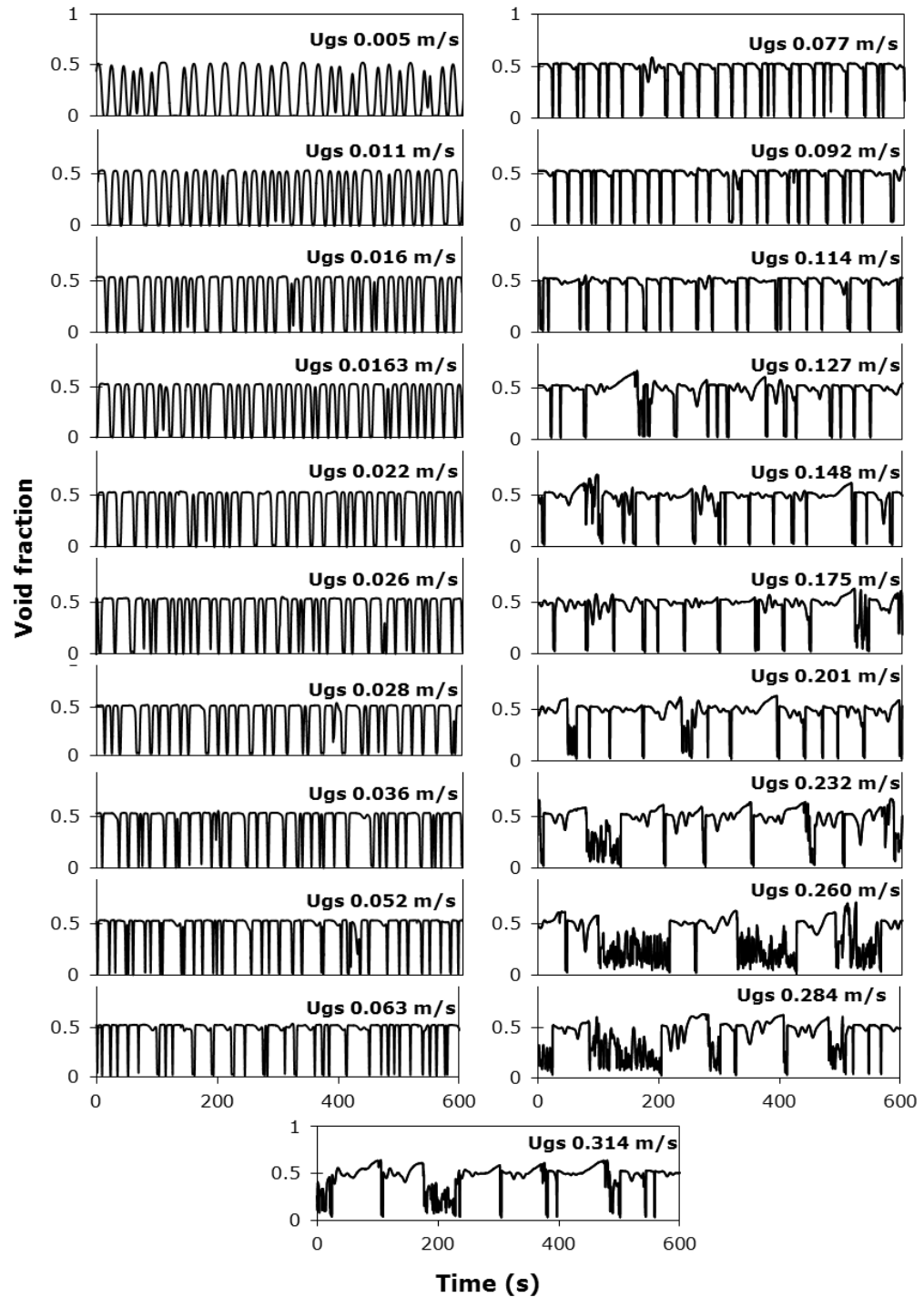


Figure 4.4: Time series of the cross-sectional averaged void fraction for different gas superficial velocities in stagnant 360Pa.s Silicone oil 240 mm diameter column.

ones and the liquid slug is much shorter. The bubbles lose their uniform structure due to the increase of the coalescence between the large bubbles. In the last plot 0.341 m/s, the variation of the void fraction is completely different in contrast to the previous ones. As the length of the bubbles

increases dramatically to occupy the majority of the column volume, the frequency of slugs decreases significantly. The slug length becomes much shorter than at other flow rates. In other words, the flow pattern consists of very long bubbles separated by churning parts of the lower void fraction that become more frequent as the flow rate increases.

Figure 4.5 shows all the time series obtained for the larger column 290 mm. Here, a wide range of gas flow rates was studied. The bubbly flow in this column appears between 0.003-0.016 m/s. It starts with small bubbles with low void fraction and low frequency. The liquid slug lengths in this stage are higher than the bubbles. The bubbles then become larger, the void fraction higher and the frequency increases. There is no widespread evidence of bubble coalescence. At gas superficial velocities 0.022-0.088 m/s, where slug flow starts, the flow becomes less structured especially in the last two graphs in this range. The bubbles become longer and the coalescence increases with increasing the gas flow rate. The structure of the flow totally changes in the next graphs, 0.108-0.243 m/s. It consists of very long bubbles with high void fraction separated by low void fraction churning regions. The churn parts start to appear at the beginning of the transition stage and become more frequent and longer with increasing flow rate. In this stage, the air flows through the oil in a form of conduit. The diameter of this conduit changes and it corresponds to the film thickness around this passage. This behaviour of the air dominates most of the flow structure. However, one very long bubble might appear in a form of an open column with a very long falling film, which accumulates in the bottom of the column to create another conduit. The length of this conduit increases until the whole falling film drains. Chapter 7 illustrates the churn flow in detail.

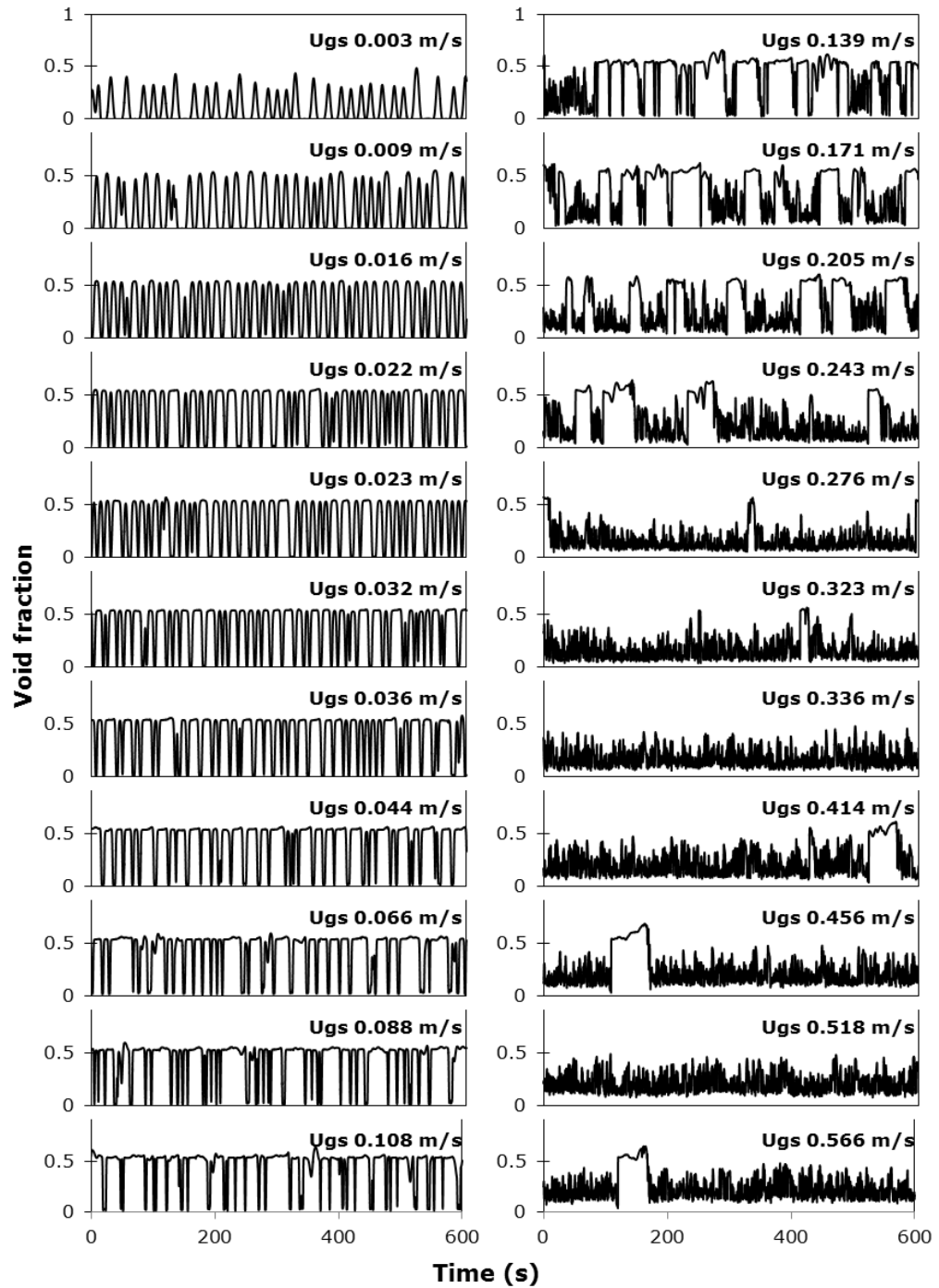


Figure 4.5: Time series of the cross-sectional averaged void fraction for different gas superficial velocities in stagnant 330Pa.s Silicone oil 290 mm diameter column.

Comparing the flow structures according to the column diameter, the flow showed a slight difference in the behaviour. In the smaller column, the bubbly flow occurred until 0.011 m/s, while it existed for a wider range, 0.016 m/s, in the larger column. The slug flow seemed to occur for a wider range in the smaller column, 0.016-0.114 m/s, while it only appeared in the

range 0.022-0.088 m/s in the larger one. The transition to churn flow was longer in the smaller column. However, in the larger column, a churn flow was recorded with shorter transition region.

4.2.3 Pressure and Pressure Gradient

Variation of pressure with changing gas flow rates was measured to study the characteristics of the flow in each flow pattern inside both columns. Three pressure transducers at 1.02, 2.47 and 4.17 m and two transducers at 2.2 and 3.6 m from the air inlet section were used to measure the pressure in the 240 and 290 mm columns respectively. A wide range of gas superficial velocities of 0.0008-0.43 m/s in the 240 mm and 0.003-0.517 m/s for the larger column were used to study the pressure drop.

Figure 4.6 and 4.7 illustrate the changing in pressure with time in air-viscous oil mixture in 240 and 290 mm columns respectively for different gas flow rates. The first figure is from an experiment carried out by Dr Laura Pioli from Section des Sciences de la Terre et de l'Environnement, Université de Genève, Switzerland on the smaller column. The second (Figure 4.7) represents results from this work on the larger column. At low gas superficial velocities of 0.0008-0.0096 m/s, Figure 4.6, the pressure drop is approximately constant due to the small size and low frequency of the bubbles (long liquid slugs). Oscillation in pressure decreases gradually with increasing gas flow rates; this might be due to the thick film around the air conduit in the column. This is quite similar to the results from the 290 mm column (see bubbly flow at 0.003 m/s). Increasing gas flow rate (slug flow) causes fluctuation in the pressure signals. This oscillation in the pressure increases with increasing the gas superficial velocity (see 0.0155, 0.028, and 0.061 m/s graphs in Figure 4.6).

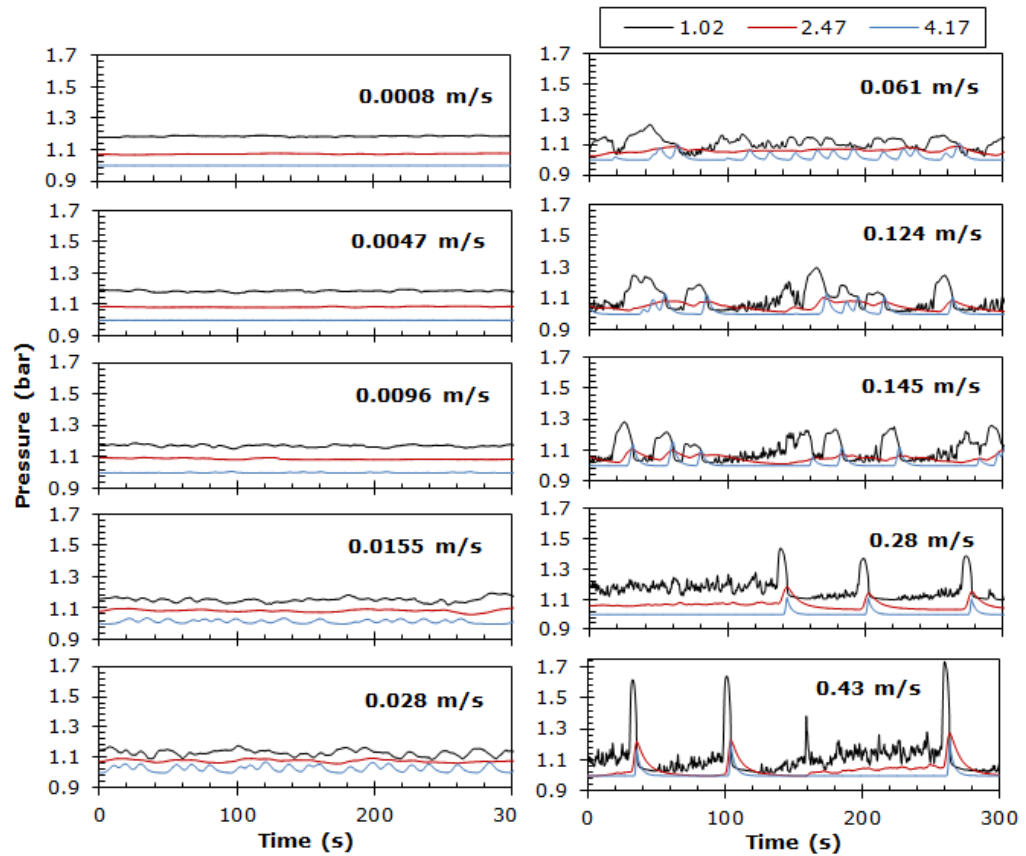


Figure 4.6: Effect of gas-viscous oil flow on the pressure inside a 240 mm column for different gas flow rates. Three pressure transducers of 1.02 (black line), 2.47 (red line), and 4.17m (blue line) spacing has used the distance between the first sensor and the air injection points is equal to 1.03 m.

Likewise, it starts to fluctuate steadily during slug flow at gas superficial velocities of 0.035 and 0.053 m/s in the larger column. This is due to the high frequency of the long Taylor bubbles which occupy the cross-section area of the column. On the other hand, this oscillation seems higher and more complicated than the ones in the larger column for the same range of flow rates. Unlike the 290 mm column, in high gas superficial velocities (0.124-0.43 m/s), the pressure signal shows a synchronous oscillation for all transducers particularly the lower one. The transducer in the lower part of the small diameter column seems to be more affected by churning. This different behaviour might be attributed to, firstly the different positions of

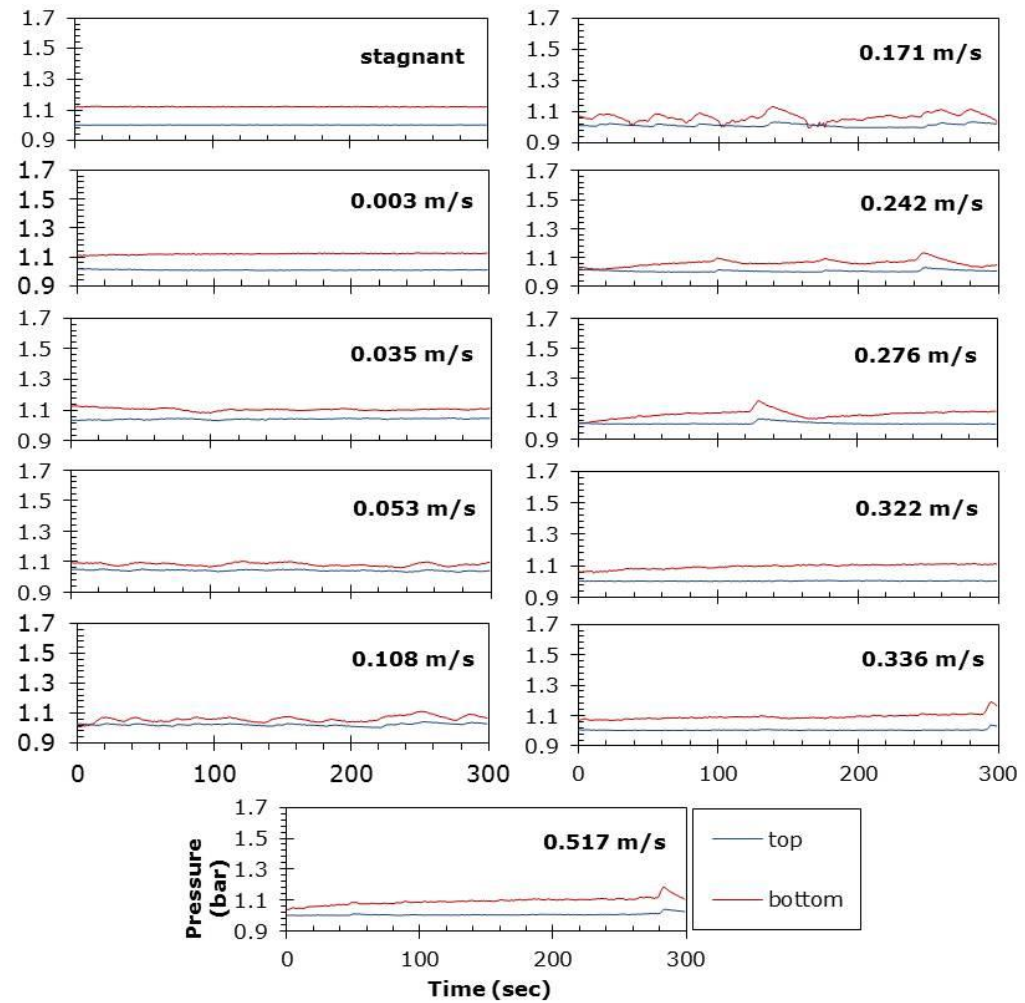


Figure 4.7: Effect of gas-viscous oil flow on the pressure inside the 290 mm diameter column for different gas flow rates. The distance between the two sensors is 1.4m, the bottom sensor is located at 2.2m from the gas inlet section.

the pressure transducers, except the 2.47 and 2.2 m on the 240 and 290 mm columns respectively which look similar. Secondly, to the effect of pipe diameter on the pressure drop which seems to be more affected in the smaller diameter column. In the larger diameter column, the pressure seems approximately linear, while it reaches maximum fluctuation in the smaller diameter column. Churn flow seems to have less turbulent and thicker film with increasing the pipe diameter. Figure 4.7 shows the comparison of these plots with the gas superficial velocities of 0.108-0.242 m/s. The fluctuation of the pressure outputs reaches the maximum due to increasing the collisions of the bubbles in the transition to churn flow. The pressure gradient in churn

flow can be seen in the last four flow rates. The oscillation in pressure decreases gradually with increasing the gas flow rates; this might be due to the thick film around the air conduit in the column.

Comparison between pressure gradient, which is equal to the pressure difference divided by the distance between the sensors, and void fraction for a wide range of gas flow rates in the column can be found in Figure 4.8. Increasing the gas flow rate increases the void fraction and decreases the pressure gradient which reaches its minimum at the start of the transition to churn flow (2.5 KPa/m). Then it starts to increase gradually to the end of the churn flow region where the void fraction seems steady. The pressure gradient in the column is basically dependent on the liquid portion of the gas-liquid mixture. Therefore, in bubbly flow, which consists of bubbles with a low void fraction (thicker liquid film), the pressure gradient is relatively higher.

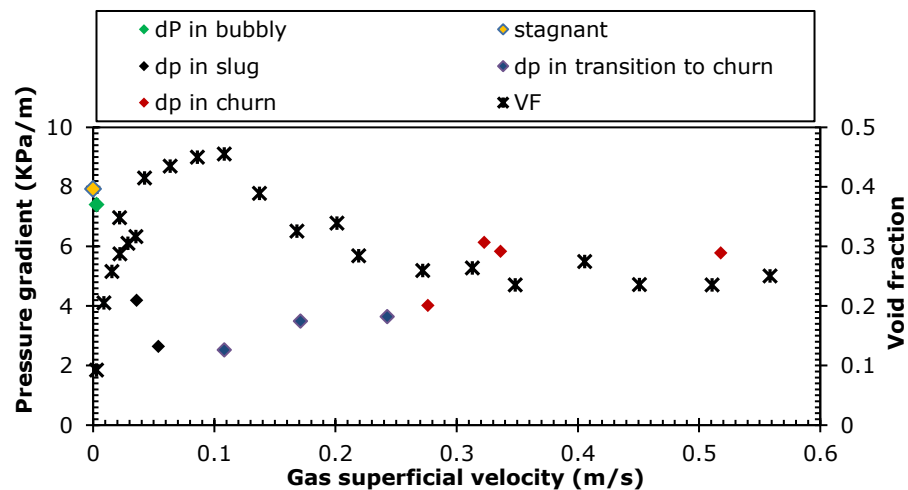


Figure 4.8: Mean pressure gradient and mean void fraction with changing the flow patterns of air flow in 330 Pa.s Silicone oil and 290 mm diameter column.

Increasing the gas flow rate increases the void fraction due to the large Taylor bubbles which almost occupy the pipe cross section area. Increasing the frequency of the bubbles and decreasing the liquid slug length are also considered factors that decrease the pressure gradient. With increasing flow

rate where churn starts to appear, pressure gradient starts to increase because of: a- increasing the liquid film thickness around the air conduit in the column, and b- increasing the interaction between the two fluids due to increasing the high frequency activity areas (increasing the velocity of the waves and the interaction between both fluids).

The behaviour of the pressure gradient with changing the gas flow rate does not agree with the results reported by Pioli et al. (2012). Stagnant glucose syrup of 300 Pa.s and 1450 Kg/m³ density was used in 240 mm column with a maximum gas superficial velocities 0.7 m/s. They pointed out that in lower gas flow rates the change in pressure decreases rapidly, then at churn flow (the higher gas flow rates), it becomes an “asymptotic trend” of declining, while this work shows a different trend. It shows a sharp drop in the pressure gradient in the bubbly and slug flows and a steady increase during the transition to churn and churn flow. It also reports the minimum pressure gradient at a gas superficial velocity of 0.108 m/s which is equal to 2.5 KPa/m, while Pioli et al. (2012) recorded 3 KPa/m at a gas superficial velocity of 0.7 m/s. The reason behind the difference in the results might be attributed to the difference in column diameters and the density of the liquids in both experiments.

Nevertheless, the results from the large column show agreement in trend with the 240 mm diameter column, which is presented in Figure 4.9, for a similar range of gas flow rate. The dimensionless pressure gradients decrease in the bubbly and slug flow regions, while it increases again toward churn flow. The trend of pressure gradient results in both columns agree with the results presented by Owen (1986), compare Figure 2.5 in Chapter 2.

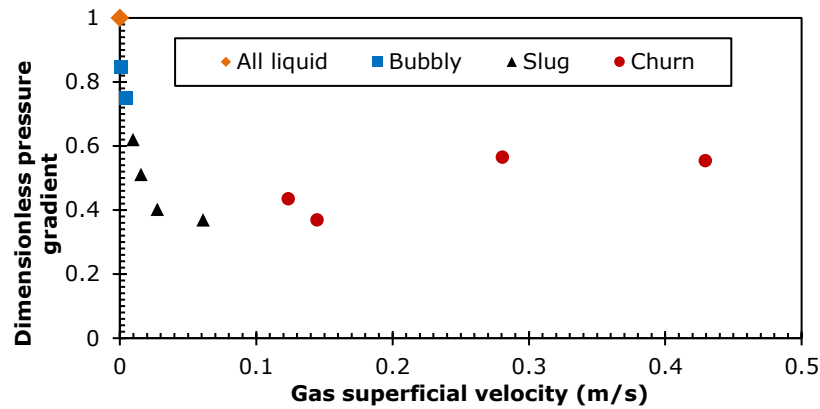


Figure 4.9: Dimensionless pressure gradients and the flow regimes of air flow in 360 Pa. Silicone oil and 240 mm diameter column.

The time-averaged pressure change with the gas superficial velocities in the column is shown in Figure 4.10.

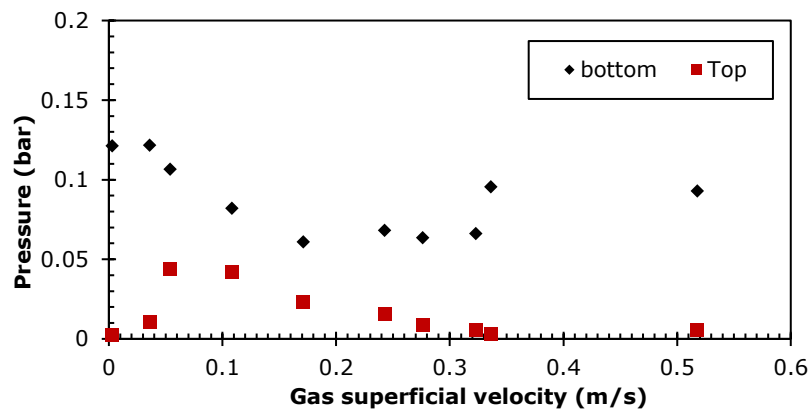


Figure 4.10: The effect of gas flow rate on the pressure drop inside 290 mm diameter column and 330 Pa.s Silicone oil using two pressure sensors separated by of 1.4 m.

The overall picture shows a significant difference in the pressure between the two sensors in the low and very high gas flow rates. This difference decreases during the transition to churn flow regime due to the long Taylor bubbles which approximately occupy the whole length of the column.

Figure 4.11 represents the time of churning in the 240 and 290 mm columns obtained from the ECT and the pressure transducers. In general, the time of churning increases in both columns with increasing the gas superficial velocity. However, it seems to be higher in the larger diameter for the same gas flow rates (see the range from 0.1 to 0.223 m/s). This figure can be

compared with the time series data (Figure 4.4 and Figure 4.5) and the pressure results to obtain the full picture of the characteristics of the flow.

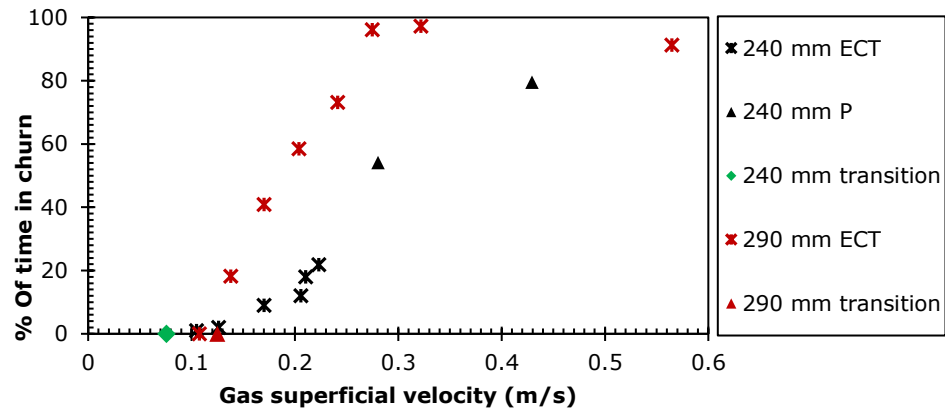


Figure 4.11: Comparing time of churn in both columns for different gas superficial velocities and two different diameter columns.

Finally, Figure 4.12 compares the pressure difference at different points in the two columns for similar gas superficial velocities and flow patterns. In general, the values of the pressure difference between 2.47-4.17 m and 2.2-3.6 m at 240 and 290 mm column respectively seem to be close for most of the flow patterns except churn flow. Starting with the bubbly flow at 0.0047 and 0.0003 m/s in the 240 and 290 mm columns respectively, the pressure difference seems to have a constant trend in both columns. It starts to fluctuate in the second plot at the gas superficial velocities of 0.061 and 0.053 m/s (slug flow) for the smaller diameter column while it is almost linear in the larger diameter column. Then the fluctuation of the pressure difference increases more and becomes complicated with increasing the gas superficial velocity to 0.145 in the 240 mm column.

However, the fluctuation increases slightly in the larger diameter column at the transition to churn region. Finally, churn flow shows a different trend in both columns. It shows a synchronised fluctuation with large peaks in the smaller diameter column at 1.2-2.47 m and 0.43 m/s, while it is slightly fluctuating at 2.47-4.17 m in the same column. The pressure difference in

the larger diameter column shows approximately a linear trend at 2.2-3.6 m from the column due to the thick liquid film around the gas conduit. The difference in the trends between both columns might occur due to different positions of the pressure measurements. It also might belong to the effect of the pipe diameter, which appears more in the smaller diameter pipes.

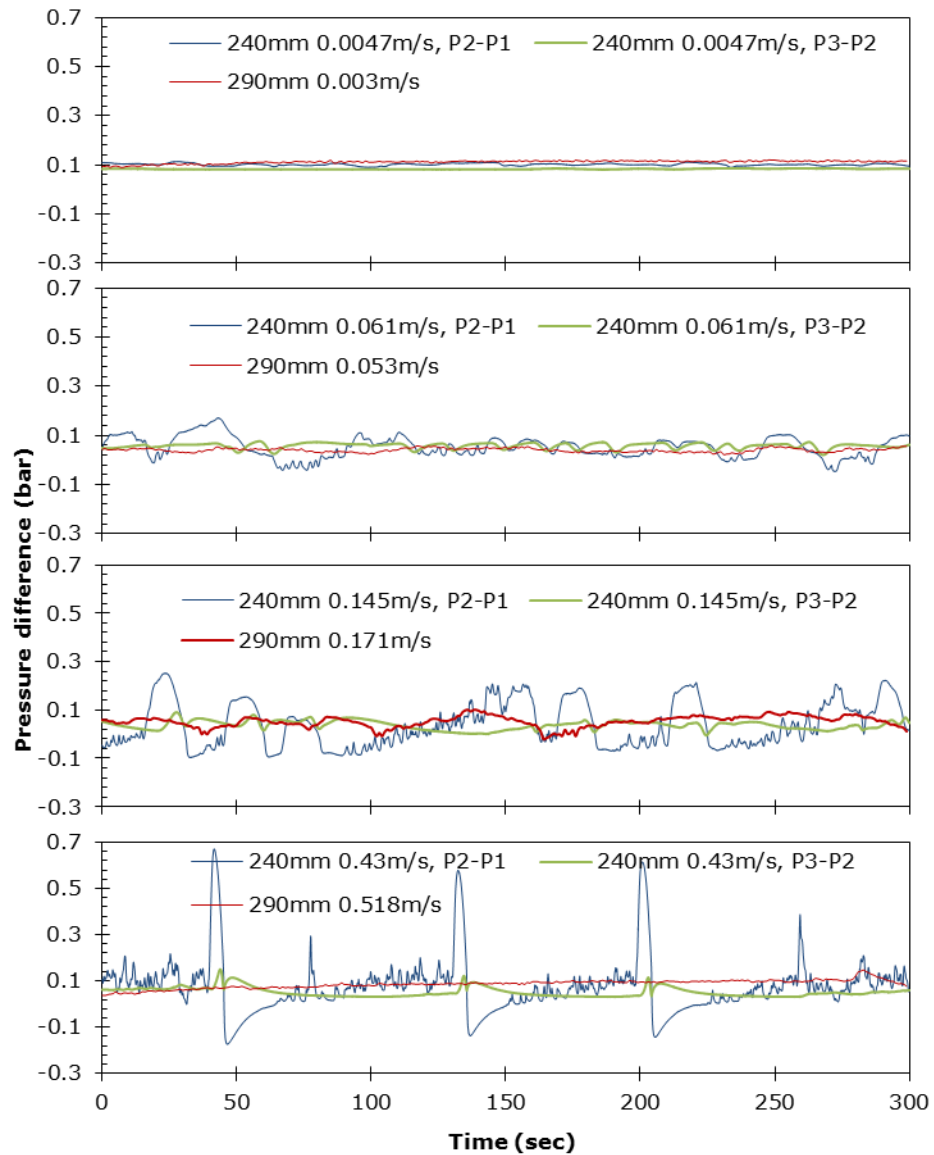


Figure 4.12: comparing the pressure difference at different positions in two different diameter columns. The blue line is a pressure difference between 1.2-2.47m, the green line is between 2.47-4.17m in the 240 mm diameter column. Red line is between 2.2-2.37m in the 290 mm diameter column.

4.2.4 Effect of Temperature

The effect of temperature on the structure of gas flow in 360 Pa.s Silicone oil was studied for the 240 mm diameter column. Oil temperatures of 20, 22, 23, and 30 °C, which are corresponding to 360, 340, 337, and 302 Pa.s respectively, were used in these experiments with gas superficial velocities of 0.005-0.3 m/s. The thermocouple, which was used to measure the temperature, was located at 1 m above the air injection points at the bottom of the column. A viscosity-temperature test was carried out by using a Viscometer (see Chapter 3) and samples of the Silicone oils with different temperatures to determine the viscosity of the oil used in both columns in different temperatures. The viscosity-temperature trend for the oils in both columns can be seen in Figure 4.13. The viscosity of the oil decreased with increasing the temperature as it was expected.

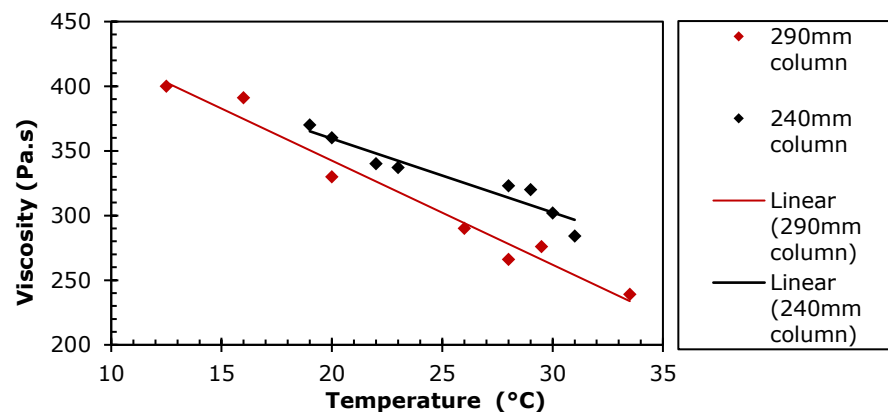


Figure 4.13: Viscosity-temperature tests for the Silicone oils used in the experiments of the 240 and 290 mm diameter columns.

In Figure 4.14, which illustrates the effect of changing the temperature on the mean void fraction of the mixture in a 240 mm diameter column, the mean void fraction seems to be independent of the temperature for the most of the temperatures. However, for the viscosity (360 Pa.s), the mean void fraction appears to have minimum values comparing with the other viscosities.

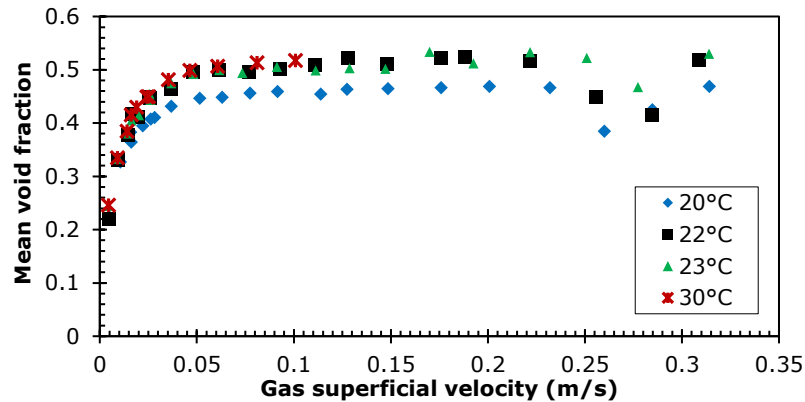


Figure 4.14: Effect of temperature on the mean void fraction in 240 mm diameter column.

Structure velocity shows approximately different trend compared to the mean void fraction. The values of structure velocity seem almost constant with changing temperature, while they become more scattered with increasing the gas superficial velocities above 0.07 m/s, see Figure 4.15.

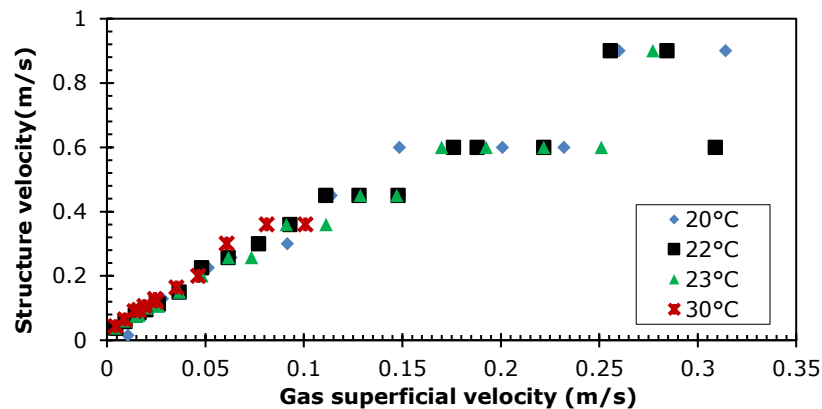


Figure 4.15: Effect of temperature on the structure velocity of air flow in 302-360 Pa.s Silicone oil in 240 mm diameter column.

Taylor bubbles in Figure 4.16 show a similar behaviour by changing the temperature. They appear more affected by changing the temperature at high gas flow rates. This behaviour might belong to the unstable structure of the flow at high gas flow rates besides the effect of temperature on the viscosity.

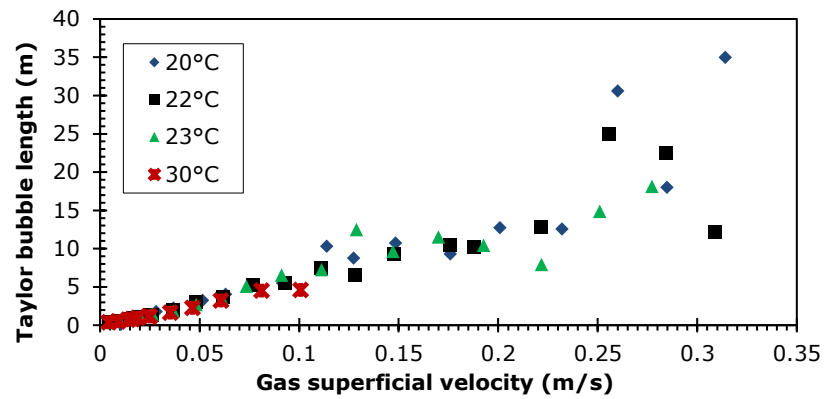


Figure 4.16: Taylor bubble lengths with changing temperature of the Silicone oil in 240 mm diameter column.

Unlike Taylor bubble lengths, the film thickness seems almost constant with changing the temperature except for the values recorded at 20°C (see Figure 4.17).

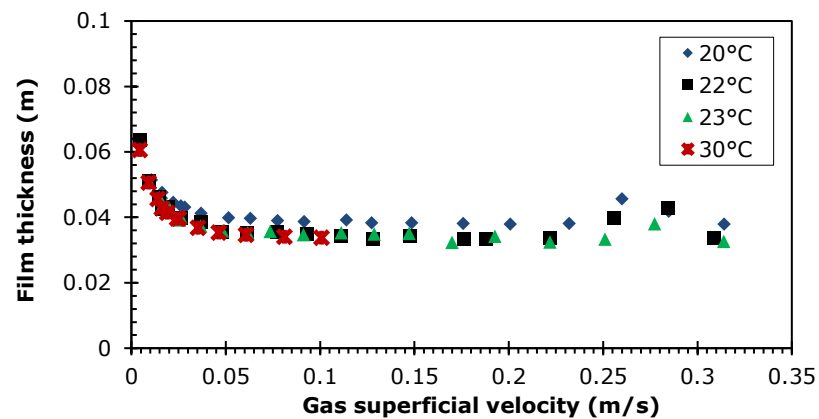


Figure 4.17: Liquid film thickness around air flow in stagnant 302-360 Pa.s Silicone oil at different temperatures.

4.2.5 Probability Density Function

The probability density function (PDF) is used to present the void fraction data measured by the ECT. From the work of Costigan and Whalley (1997), in liquids of low viscosities, the bubbly flow is recognised by a peak at low void fraction and spherical cap bubble flow by a peak and a tail towards high void fraction. Slug flow is characterised by the existence of two peaks. The small peak corresponds to the value of the liquid slug, whereas the large

peak to the Taylor bubbles. In unstable slug flow, the value of the lower peak is significantly lower than in regular slug flow. This low peak starts to lose its identity until it disappears. Then the churn flow appears with a single high peak accompanied by a tail towards lower void fraction.

The PDFs files of viscous liquids for both diameters, obtained in the present work, show trends different from those by Costigan and Whalley (1997), see Chapter 2. Starting with bubbly flow, the PDF appears to have the same characteristics as spherical cap bubbles of water. It consists of two peaks, one with high void fraction and low PDF that corresponds to the spherical Taylor bubble and another with zero void fraction and high PDF. This can be seen for the gas superficial velocity of 0.005 m/s in Figure 4.18 and 0.003 m/s and 0.009 m/s in Figure 4.19. Increasing gas superficial velocities to 0.011 m/s in the 240 mm diameter column and 0.015 m/s in the larger column increases the PDF of the large spherical bubbles. Likewise, the PDFs of slug flow in the viscous oil appear to be more similar to the unstable slug than slug flow in water. The PDF shows two peaks of void fraction with a considerable difference. The lower peaks are at the low values and the high peaks are at higher values of void fraction, Taylor bubbles, 0.016-0.114 m/s and 0.022-0.088 m/s in the 240 and 290 mm diameter columns respectively. The slug flow in both columns records the higher PDF value for the Taylor bubbles. The value of PDF decreases with increasing the gas superficial velocity. This indicates the decrease of the number of Taylor bubbles due to the bubble coalescence to create longer bubbles. However, in the transition to churn flow, the values of the PDF of the higher void fraction decreased gradually and showed two different aspects. Firstly, in the smaller diameter column, at gas superficial velocities of 0.127-0.314 m/s, the single high void peak decreased dramatically pointing to the growth of the bubbles and developed to more than one peak. The second peak of low void and PDF

remained the same.

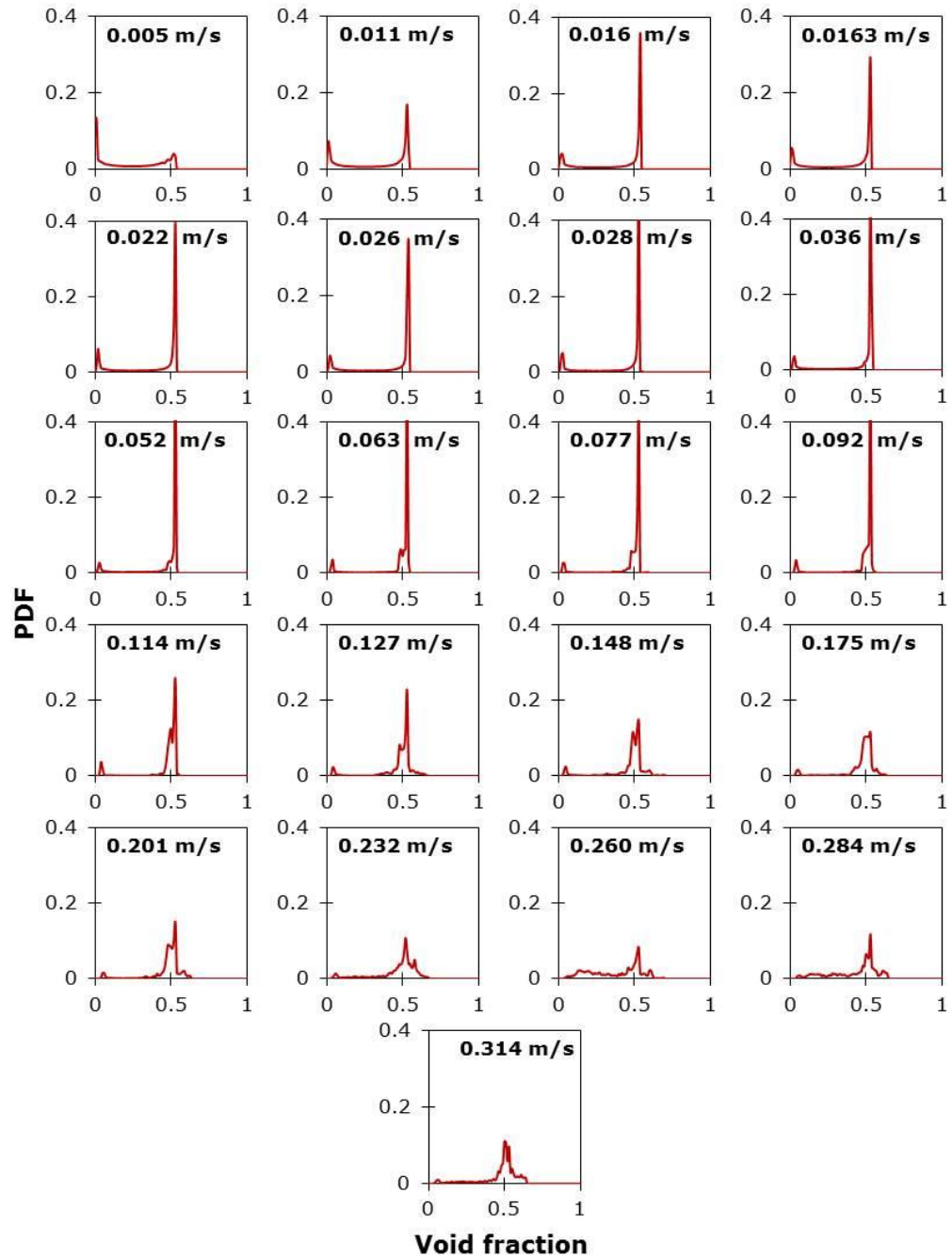


Figure 4.18: Probability density function profiles for air flow in stagnant 360Pa.s Silicone oil in 240 mm diameter column.

Secondly, in the 290 mm column, 0.108-0.243 m/s the higher void fraction peak decreased dramatically to become smaller than the lower void fraction peak. In other words, the PDF of the Taylor bubbles became less than the small bubbles in the liquid slug. The last flow pattern that was observed in this work is churn flow. This regime occurred in the larger column where a

wider range of gas flow rates was used.

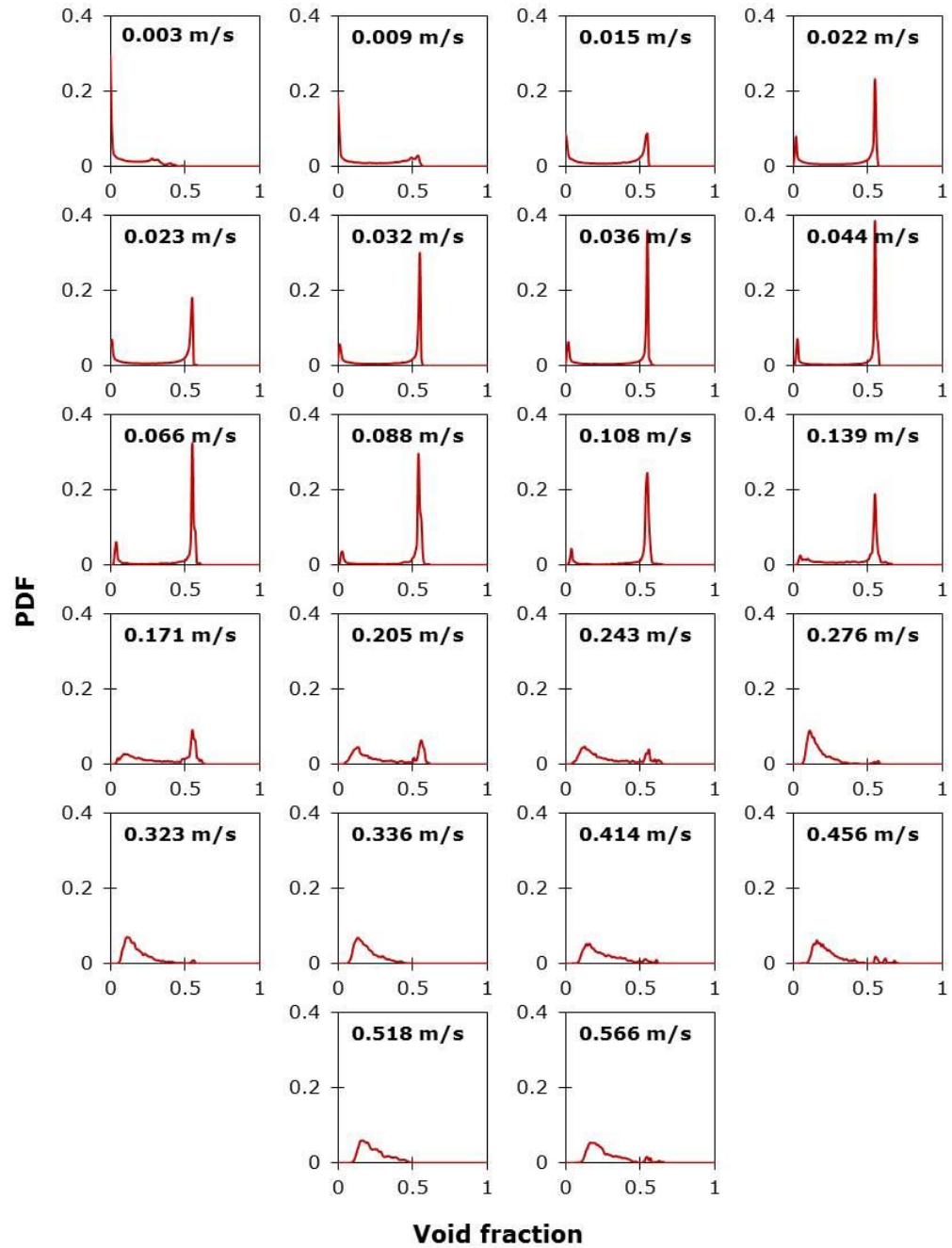


Figure 4.19: Probability density function profiles for air flow in stagnant 330Pa.s Silicone oil in 290 mm diameter column.

The churn flow can be seen in Figure 4.19 for the gas superficial velocities 0.276-0.566 m/s. It varies between one peak of low void fraction and one of low void and a number of small peaks of high void fraction. The first one can be explained by looking at the time series of void fraction in Figure 4.5.

4.2.6 Mean Void Fraction

Void fraction, which represents the cross-sectional area occupied by the gas, in gas-liquid flow, was obtained from the ECT measurements. The time averaged of the void fraction (here called the mean) from the time series data for a wide range of gas flow rates was determined in this work. Figure 4.20 shows the effect of increasing the gas flow rate on the values of the mean void fraction in two different diameter columns.

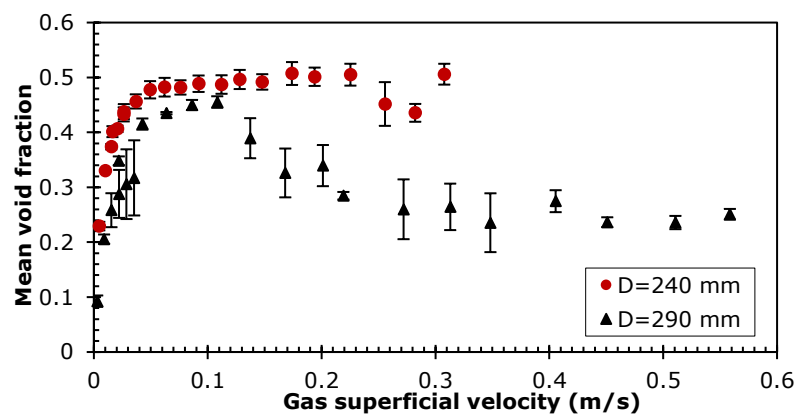


Figure 4.20: Time-averaged void fraction of air flow with different gas superficial velocities in stagnant high viscosity Silicone oil showing the standard errors for both 240 and 290mm diameter columns.

The error bars, in the graph, represent standard errors which were calculated from the standard deviation of the mean void fraction and the square root of the number of samples. The number of the samples in these experiments was 3; the runs were repeated 3 times for the whole range of the flow rates. First, in the smaller diameter column, the experiments were repeated with four different temperatures for the Silicone oil. Second, in the 290 mm diameter rig data were obtained by using three different acquisition frequencies. The details of these experiments can be found in Chapter 3.

In general, the figure shows an increase in void fraction with increasing gas superficial velocity (Kuncová and Zahradník (1995)). Then there is a sudden drop at higher values of gas superficial velocities. This potentially

corresponds to a transition in the flow pattern. The trend in both columns seems to be approximately similar in the bubbly and slug flow regimes. Void fraction increases sharply at lower flow rates (bubbly and beginning of the slug) and slightly in the second half of the slug flow regime. During the transition to churn flow, the values of void fraction showed different trends, so it increased slightly, then dropped suddenly to rise again in the last flow rate in the 240 mm diameter column. While in the larger column, it decreased gradually and continued to decrease to the next flow pattern to rise slightly at the end of the churn flow. The standard error showed a reasonable value of 0.2-4% in the most of the flow rates in the smaller diameter column. However, it seemed to be larger by 0.2-6.8% in the 290 mm diameter column. The highest percentage of the error appeared during the transition between the flow regimes where the flow is not structured.

Volume averaged void fraction from the level swell was determined and compared with values obtained from ECT. This can be calculated from the initial height of the oil (before air injection) and the final height of air-oil mixture assuming that the volume remains constant inside the column (Hills, 1976), See Section 2.5 in Chapter 2. Figure 4.21 illustrates this comparison. In general, maximum relative errors of 0.6 and 0.18 were determined between both methods in the 240 and 290 mm diameter columns in bubbly and slug flow regimes. However, the values of void fraction seem to scatter toward the churn flow region to value of 0.7 maximum relative error in the larger diameter column. The reason behind that is the mechanism of the eruption of the bubbles at the top surface of the liquid. It is quite difficult to select the position of the top surface after the bubble eruption due to the remaining liquid film from the previous erupted bubble. The eruption of the bubbles and the behaviour of the top surface of the oil are explained in Chapter 5.

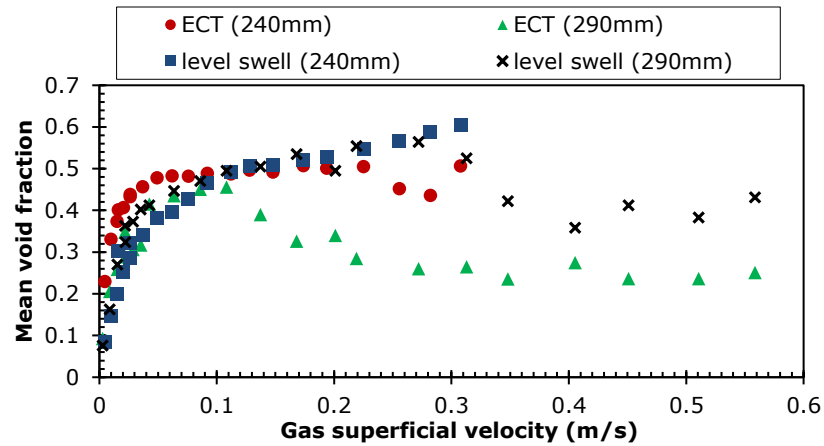


Figure 4.21: Comparing the time-averaged void fraction from the ECT and the level swell of air flow in high viscosity oil in different column diameters and gas superficial velocity.

4.2.7 Structure of the Flow in Viscous Oils

This section deals with the structure of the air-viscous oil flow according to the motion and the dimensions of the mixture. Structure velocity, Taylor bubble and liquid slug lengths, as well as liquid film thickness, can provide valuable information about the nature of the flow in the column. It might be worth using the term “large bubbles” rather than “Taylor bubble” in this section because it can be applied to all flow patterns and Taylor bubble as conventionally defined, is not found in transition to churn and churn flow patterns. A comparison between the structures of the flow in two different diameter columns beside theoretical approaches from literature can also be found below:

4.2.7.1 Motion of the Bubble and Liquid Slug

Structure velocities can be determined from cross-correlation of the void fraction measured by the two planes of the ECT. The time delay between both signals and the distance between both planes are used to calculate a structure velocity. This is the distance between the planes, divided by the time delay of the signals.

In this work, structure velocity was calculated for a wide range of gas superficial velocities in two large-scale columns. The experimentally determined structure velocities are compared with the velocity calculated by using a model proposed by Viana et al. (2003) and Guet et al. (2002), which was introduced in Chapter 2. The average data for the structure velocity in Figure 4.22 shows a linear trend for both columns at low gas flow rates. However, as the gas flow rate increases the rate of increase of structure velocity changes. It is higher in bubbly and slug flow regimes while it becomes less in the transition to churn and churn flow due to the characteristics of the flow in these regions.

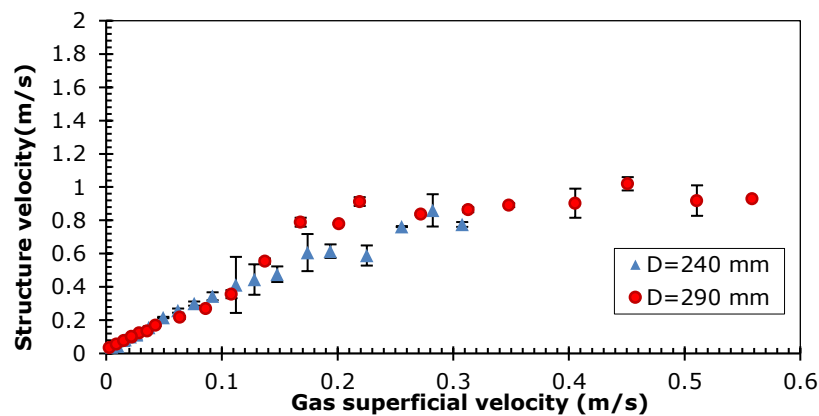


Figure 4.22: Structure velocity with the standard error comparing the 240 and 290mm diameter columns and high viscosity Silicone oil.

In bubbly and slug flow, where the Taylor bubble is dominating the flow structure, larger bubbles rise faster because of the balance between the buoyant weight and the drag force, (Feng (2008)). Then, with the higher gas flow rates, the increasing rate of the structure velocity decreases dramatically, due to the coalescence of the bubbles and the increasing of the lengths of the bubbles. Moreover, with changing the flow regime, the structure velocity indicates a completely different behaviour. It loses its trend and becomes almost random in churn flow at the higher gas flow rates studied. The error bars in the figure, calculated from the standard deviation, show a reasonable value for both columns. The values range between 0.03-

9% and 0.07-16.8% in the 240 and 290 mm diameter columns respectively. The maximum values of the error occur in the transition between the flow regimes.

In contrast, Taylor bubble velocity from the model by Viana et al. (2003) shows a linear trend with gas superficial velocity. Both values (experimental and numerical) show values of 0.5 and 2.2 maximum relative error in the 240 and 290 mm diameter columns respectively. The experimental values seem to scatter (decrease) during the transition to churn flow in the smaller column and in both transition to churn and churn flow in the larger diameter column. This is illustrated in Figure 4.23.

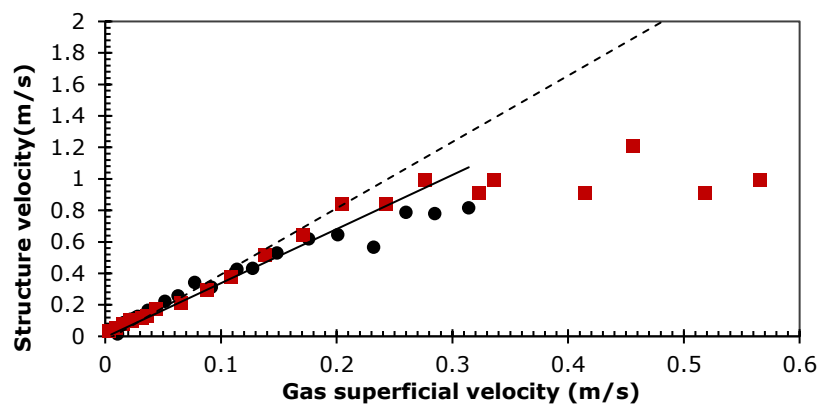


Figure 4.23: Theoretical and experimental structure velocities in viscous Silicone oil for both columns. •:240mm, ■: 290mm (experimental) and — 240mm, ---290mm (from Viana et al. (2003) and Guet et al. (2004)).

4.2.7.2 The Dimensions of the Bubbles and Liquid Slugs

The lengths of large bubbles and of liquid slugs in viscous Silicone oil were determined in the columns for a wide range of gas flow rates. The results were calculated from the time that each bubble/liquid slug passes through the ECT probes for each velocity. The lengths of the large bubble and liquid slug in this chapter represent mean values for each gas flow rates. They show an overview of the flow structure. The next chapters will present each flow pattern separately based on the accurate values of each dimension by

extracting the bubbles from the slug and calculating them individually.

Figure 4.24 presents the lengths of the large bubble and liquid slug in both columns. It is clear that in both columns, the length of the large bubble increases with increasing gas flow rate. It increases gradually at lower gas flow rates due to the structure of slug flow to reach the maximum 9.5 m at the start of the transition to churn flow; this will be explained in Chapter 7.

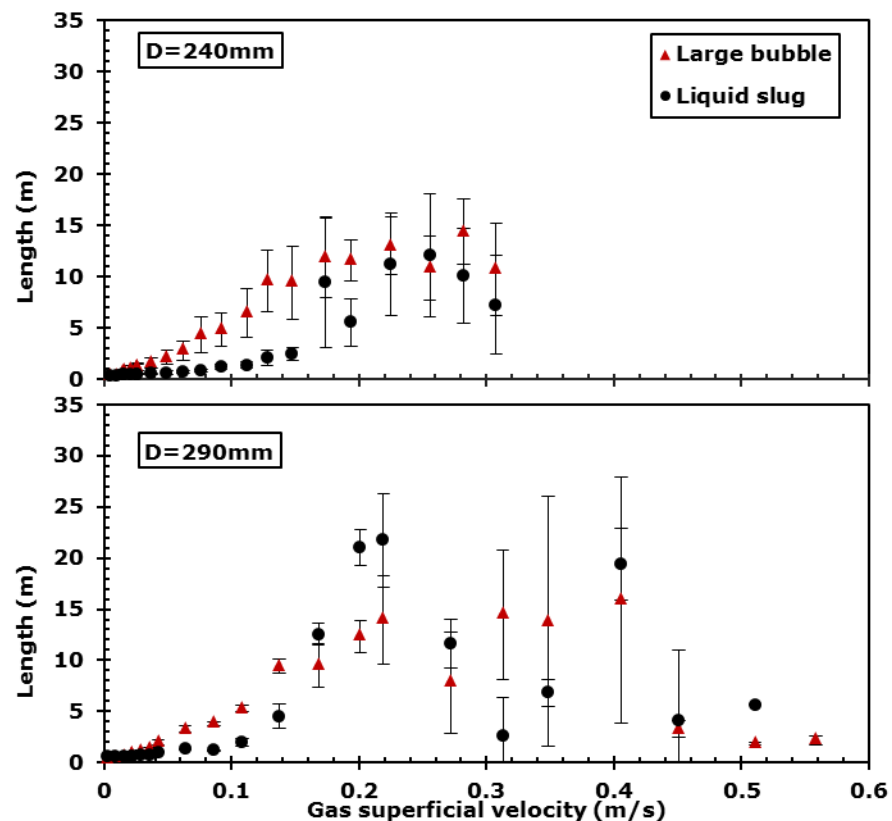


Figure 4.24: Comparing the standard error for the mean lengths of Taylor bubble and liquid slug for both 240 and 290mm diameter columns.

Then it starts to fluctuate for the rest of the transition in the smaller column. Similar behaviour is seen in the larger column. The bubble length increases to 13.9 m at a gas superficial velocity 0.219 m/s. This takes place at the beginning of the transition to churn flow regime. Then it starts to fluctuate as it does in the smaller column, but it drops suddenly at the end of the churn flow regime to reach approximately 3 m. To understand the results in these two graphs, it might be worth comparing each flow rate with the

corresponding time series of void fraction in Figure 4.4 and Figure 4.5. The liquid slug length shows a different aspect of changing the column diameter. It remains less than the large length for the whole ranges of flow rates in the smaller column. While it becomes longer than the bubbles in the 290 mm column in the transition to churn and churn flow regimes. This is due to the characteristic of churn flow where no defined Taylor bubble can be found. The last 7 plots in Figure 4.5 show the structure of churn flow. The error bars in Figure 4.24 that represent the standard error, calculated from the standard deviation, show a reasonable value in the low gas flow rates in both columns; then they increase in values in higher gas flow rates where the structure of the flow is not homogenous. In the small column, the values of the errors of the large bubble vary between 1.2-440 %, while they are equal to 1.7-629 % for the length of the liquid slug. In the 290 mm diameter column, the values of the standard errors recorded are 0.6-1200 % for the Taylor bubbles and (1-690)% for the liquid slug length. The high values of the standard errors in both columns appear in the very high gas flow rates due to the unstable structure in these flow regimes.

The frequency of the large bubbles was calculated from the time series data by counting the number of the bubbles per unit time for each gas flow rate. The frequency of the large bubbles appears to be affected by increasing the gas flow rate. It shows a different trend in each flow regime. In both columns, the frequency of the large bubbles increased steadily with increasing gas flow rate during the bubbly flow regime. Then it declined gradually during the slug flow regime due to the increase of coalescence between bubbles, which created longer bubbles. The minimum values of the frequency obtained at the beginning of the transition to churn flow regime are 0.029 and 0.023 Hz in 240 and 290 mm diameter columns respectively, as the bubbles are very long at this stage. The frequency starts to increase

with decreasing the length of the large bubbles by the end of the churn flow. The results from both diameters show that the frequency of the large bubbles is dependent on the gas flow rate. However, according to Azzopardi et al. (2014b), the frequency is not affected significantly by the gas flow rate. The standard errors showed a reasonable value of 0.04% to 5.7% in the larger diameter column, while it recorded 0.089-3.9 % in the 240 mm diameter column. See Figure 4.25.

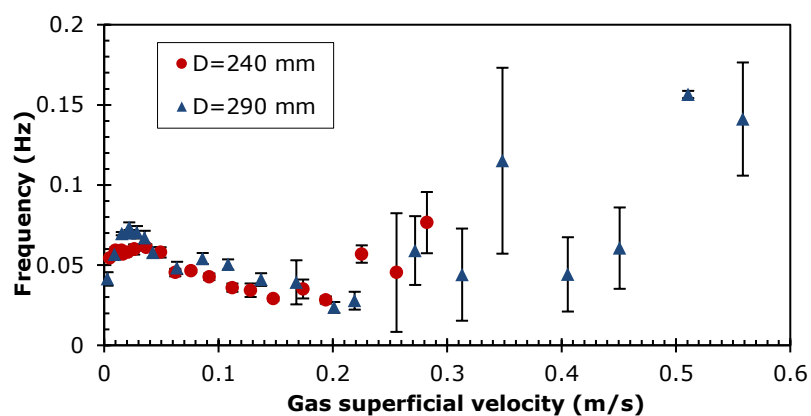


Figure 4.25: Frequency of large bubbles in a high viscosity oil and two different diameter columns. Error bars are the standard errors of the frequency.

A possible method of counting bubble frequency has been attempted. It is combined with the column diameter and the structure velocities to form a dimensionless group, the Strouhal number. In Figure 4.26, this is plotted against Froude number. An inverse relationship occurs in most flow patterns. However, it seems to increase in high Froude numbers. The last parameter studied to identify the structure of the flow in the column is the liquid film. The liquid film behaves as a falling film without interfacial shear in the case of bubbly and slug flow regimes. The characteristics of this liquid film change totally in the transition to churn and churn flow. The film becomes thicker and seems to travel in different directions due to the process of flooding.

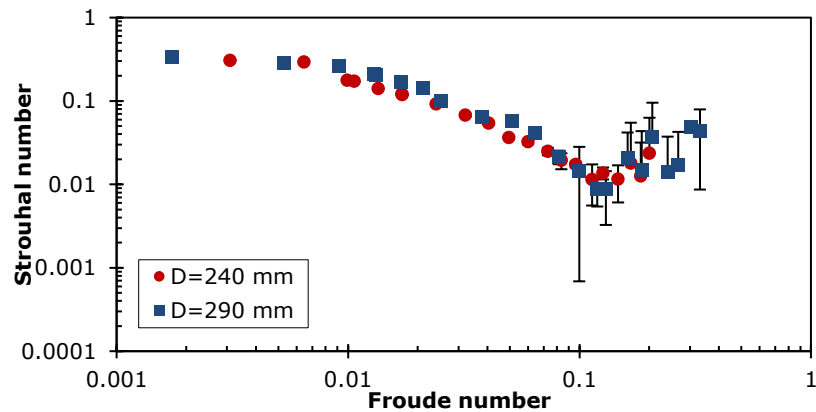


Figure 4.26: Effect of the gas superficial velocity on the Strouhal number in two different diameter columns.

The film thickness around the large bubbles was calculated from the mean void fraction of the gas-oil in the column. Figure 4.27 illustrates how mean film thickness depends on gas superficial velocity. It also presents two types of errors. First the error of the difference in film thickness between the two different diameter columns. Second, the standard error from the standard division of the film thickness of each gas flow rate.

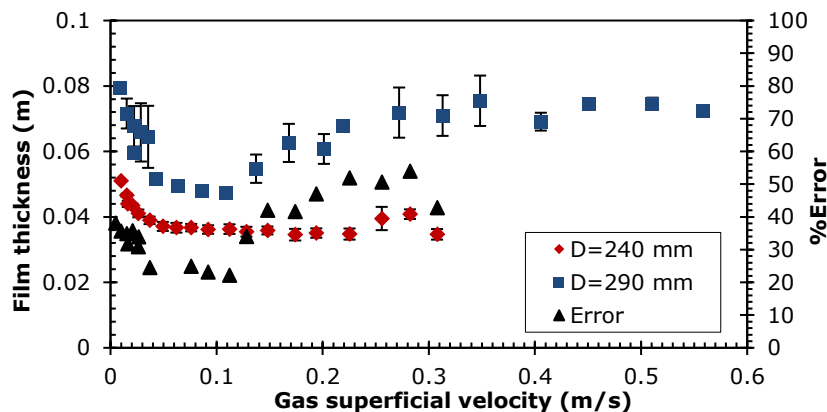


Figure 4.27: Film thickness around the large bubbles created in viscous oil in two different column diameters and a wide range of flow rates. The error presents the difference in percentage between the values of film thicknesses in both columns.

According to the results, the values of the film thickness are more affected by increasing the gas flow rate in the larger diameter column trend for the film thickness in both columns and the effect of the column diameter, which can be clearly seen in the graph. It decreases gradually by 32 mm in the bubbly and slug flow due to the increasing of the size of the large bubbles,

which occupy the column cross section. Then to increase slightly in the middle of the transition to churn regime to reach its maximum of 75 mm in the middle of the churn flow due to the flooding of the liquid film. The standard error seems to be very reasonable in this column; they start from 0% to 1.1%. However, in the 240 mm diameter column, the liquid film appears to be less affected by the gas flow rates. The film thickness decreases slightly with increasing the gas flow rates, then it remains approximately constant from the end of the slug flow regime because of the very large bubbles, which are almost equal in diameter to the column diameter. The standard error shows a very low value 0.1-0.2 %, which can be considered a good result in this case. The percentage of the error between the film thicknesses in both columns shows considerable results. The difference in the film thickness drops gradually from 38% to reach 22% in 0.076 m/s gas superficial velocities. Then it increases sharply to 51% at 0.225 m/s to rise again to 53% at 0.3 m/s. This is due to the different trend for the film thickness in both columns and the effect of the pipe diameter which appear clearly in the graph.

4.3 Summary

In general, the experimental results for air flow in high viscosities Silicone oils show a significant difference compared to the low viscosity liquids in literature. Four flow regimes with quite different structures have been found in the range of gas flow rates that have been used. The bubbly flow that consists of one spherical bubble rises in the core of the column. Then a long bubble with rounded nose and end, which occupies almost all of the cross-section area of the pipe (Taylor bubble), dominates the slug flow. Increasing the gas flow rate that leads to the creation a non-stable structure can be recognised as a transition to churn flow regime. In this flow pattern, Taylor

bubbles become very long accompanied by times of churn that increase with increasing the gas flow rate. Finally, churn flow, consists of a gas conduit passing through a column of liquid with unstable diameter and liquid film thickness. The high viscosity of the liquid and the high gas flow rate are the main factors that cause this flow structure. Mean void fraction of the mixtures, obtained from the ECT in both columns, increases with increasing gas flow rate as expected. Similarly, the structure velocity increases with increasing gas flow rate in both columns and shows similar trends with an approach from literature, except at very high gas flow rates. The length of the large bubbles increases with increasing gas flow rates in both columns to reach its maximum at the transition to churn flow regime. Then, in churn flow regions, the length decreases because of the changing in the structure of the flow. Contrary to the above, the liquid film thickness decreases with increasing gas flow rate, as the bubbles become larger, then it increases slightly towards churn flow. Moreover, the pressure gradient decreases with increasing gas flow rate in both columns, then it starts to increase at the transition to churn and churn flow regions.

Chapter 5

Large Bubbles in Viscous Liquids (Bubbly and Slug Flow)

5.1 Introduction

This chapter presents the results of experiments on the flow of bubbles in high viscosity liquids and large diameter pipes. The flow patterns at low gas flow rates consist of bubbly and slug flow. Ranges of 0.005-0.114 m/s and 0.003-0.088 m/s are used in the 240 and 290 mm diameter columns respectively to create these flow regimes. The Silicone oils of 360 P.s (smaller diameter column) and 330 Pa.s (larger diameter column), which are employed in this work, are shown in this chapter, in addition to the characteristics of the flow. Time series of void fraction, Probability Density Function (PDF) files, mean void fraction, large bubble and liquid slug lengths, large bubble velocity, liquid film thickness around the large bubbles, and the top surface oscillation are all measured. The results from the ECT are compared to the ones obtained using high-resolution, high-speed cameras. The behaviour of the bubble coalescence in high viscosity oils is shown for different gas flow rates and compared with liquids of lower viscosities from previous works. And finally, a full picture of the behaviour of the top section of the liquid is shown using a wide range of gas flow rates. Details about the experimental arrangements of each single experiment in this chapter are shown in the corresponding experiment set up in Chapter 3.

5.2 The Creation of the Bubbles

In this work, two types of bubbles (small and large) were observed to be flowing in high viscosity oils at low and intermediate flow rates and two different diameter columns. Section 3.4.1 in Chapter 3 describes the experimental arrangements used in this section. The large bubbles which dominate the structure of the flow, and contribute mainly to the void fraction in the air-Silicone oil mixture, are considered here. The lengths of these bubbles vary between 0.3 m to 8.29 m in the smaller diameter column and 0.3 -3.8 m in the 290 mm diameter column in the range of gas flow rates shown earlier. The small bubbles of (millimetres to centimetres diameter) creation, motion, and size are presented in details in Chapter 8.

It is most useful to start by defining the shape of the large bubbles. This depends mainly on the gas flow rate. Three main shapes were observed in this work:

1. Spherical shape bubbles, at gas superficial velocities of 0.005 and 0.003 m/s in the 240 and 290 mm diameter columns respectively.
2. Ellipsoidal bubbles, with the major axis vertically at gas superficial velocities of 0.011 m/s in the smaller diameter column and 0.009 and 0.016 m/s in 290 mm diameter column.
3. Long cylindrical, with rounded top and end, at ranges of gas superficial velocities of 0.016-0.114 m/s and 0.022-0.088 m/s in the 240 and 290 mm diameter columns respectively (these we call Taylor bubbles).

The first two bubble types can be recognised as bubbly flow. At low gas flow rates (flow conditions in No.1), the flow structure consists of individual large regular bubbles which rise at a constant velocity in the centre of the column

due to the wall effect. The bubbles are spherical in shape because of the surface tension of the oil, which has a significant effect only in low gas flow rates (Wraith, 1971, Feng, 2008). Furthermore, at this gas flow rate (in No.1. spherical bubbles) a minimum amount of coalescence was observed between the bubbles in the smaller diameter column while no coalescence occurs in the larger column. This could be due to, first, the uniform size of the bubbles which allow the bubbles to rise at a constant velocity. Second, the low frequency of the bubbles and the long liquid slugs between the bubbles.

At higher gas flow rates (in No.2. ellipsoidal bubbles), the spherical bubbles become larger and change to an ellipsoidal shape as the effect of surface tension is ceases to be so important (Feng, 2008). At This stage, the size of the bubbles depends on the resultant interaction between the upward buoyant force and the downward drag force. The amount of coalescence between the bubbles was observed to be higher compared with the previous flow rates in both columns. The bubbles rise at less uniform velocity as a result of the variation in the size of the bubbles (larger bubble rise at higher velocity). This is the result of the higher probability of coalescence with increasing gas flow rate. It has been observed that the large bubbles in the bubbly flow regime were followed by liquid slugs which contain a small amount of very small bubbles of millimetres. These small bubbles were seen to appear more in the larger diameter column as it appears in Figure 4.1 in Chapter 4.

The second flow structure can be considered as slug flow. Slug flow in viscous liquids, as described in Chapter 2, consists of Taylor bubbles which are long cylindrical bubbles with rounded top and end and a diameter approximately equal to the pipe diameter. Taylor bubbles are followed by liquid slugs containing very small bubbles. The amount of the small bubbles in the slug

flow appeared to be higher and the size larger compared with the corresponding small bubbles in the bubbly flow in both columns. Photos of bubbly and slug flow can be seen in Figure 4.1 and the image reconstruction of the internal structure of the air-Silicone oil for both flow regimes can be found in Figure 4.3 in the previous chapter.

The mean void fraction, which is the time averaged value of this parameter, is determined from the time series data obtained from the ECT measurements. Figure 5.1 shows the averaged void fraction for different gas flow rates (bubbly through to churn flow) in 360 and 330 Pa.s Silicone oils in the 240 and 290 mm diameter column respectively. In general, the void fraction increases with increasing gas flow rates in both columns as might be expected (Kuncová and Zahradnik, 1995). In the smaller diameter column, the mean void fraction values increase by only 0.26, while it increases by 0.348 in the larger diameter column for the same range of gas flow rate.

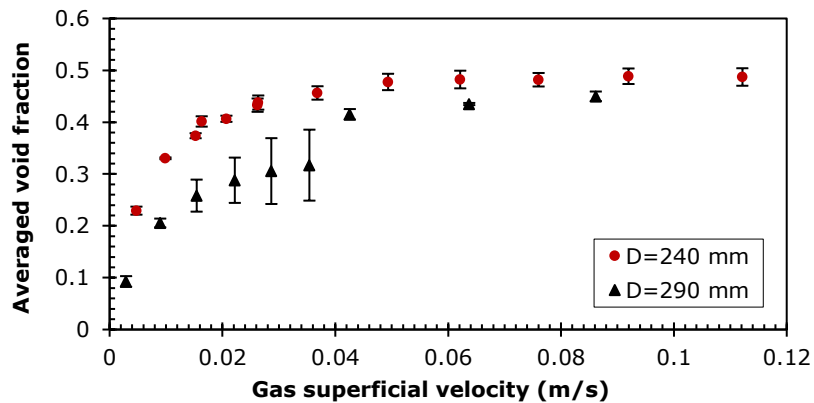


Figure 5.1: Effect of gas superficial velocity and pipe diameter on the time-averaged void fraction of air flow in stagnant high viscosity Silicone oil in two different large diameter column showing the standard errors.

This can be related to the diameter of the large bubbles which increases with increasing gas flow rates in both columns. The results of the mean void fraction could be compared to the liquid film thickness around the large bubbles in Section 5.4.2 which represents the liquid holdup. The liquid film thickness decreases as a result of increasing the bubble diameter.

Void fraction increases significantly in the bubbly flow region while it increases slightly in the slug flow region in both columns. It increases by 0.11 when the gas superficial velocity increases from 0.005 to 0.011 (bubbly flow region) m/s in the smaller diameter column. Whereas it increases by only 0.11 when the gas superficial velocity increases from 0.016 to 0.114 m/s (slug flow region) in the same column. Likewise, it increases by 0.158 and 0.16 for an increase in the gas superficial velocity from 0.003-0.016 m/s in the smaller and 0.022-0.088 m/s in the larger diameter column respectively. This could be related to the bubble diameters which seem to be more affected by the increasing gas flow rate at low gas flow rates, both for spherical and ellipsoidal bubbles. However, in high gas flow rates the length of the bubbles seems to be affected, the slug flow region. The sudden increase in the void fraction from 0.3 to 0.4 at gas superficial velocities of 0.035 and 0.04 m/s respectively might be due to the increase in the gas pressure and flow rate. These two gas superficial velocities correspond to a change from the low gas line of 2 bar to the high pressure of 6 bar. The error bars in the graph are the standard error of the mean void fraction as it is explained in Section 4.2.6 in Chapter 4. The percentage of the maximum error for the mean void fraction were 1.7% and 6.8% in the 240 and 290 mm diameter columns.

The averaged void fraction from the level swell was determined and compared with values obtained from ECT measurements in Figure 5.2. This can be calculated from the initial height of the oil (before air injection) and the final (height of air-oil mixture); assuming that the volume of liquid remains constant inside the column (Hills, 1976) See Section 2.5 in Chapter 2. In general, there is an agreement between these two methods over this range of gas flow rates in both columns. However, the results from the smaller diameter column seem slightly different compared to the larger one.

It starts with higher values of relative error between the two methods at low gas flow rates (0.6) and decreases to a minimum (0.008) at the maximum flow rate in the 240mm diameter column. On the other hand, the difference in values of averaged void fraction between the two methods does not show a specific trend. The value of relative error varies between 0.18 and 0.04 over the gas flow rates used in bubbly and slug flow. This is due to the difficulty in determining the position of the expanded height of the air-Silicone oil mixture.

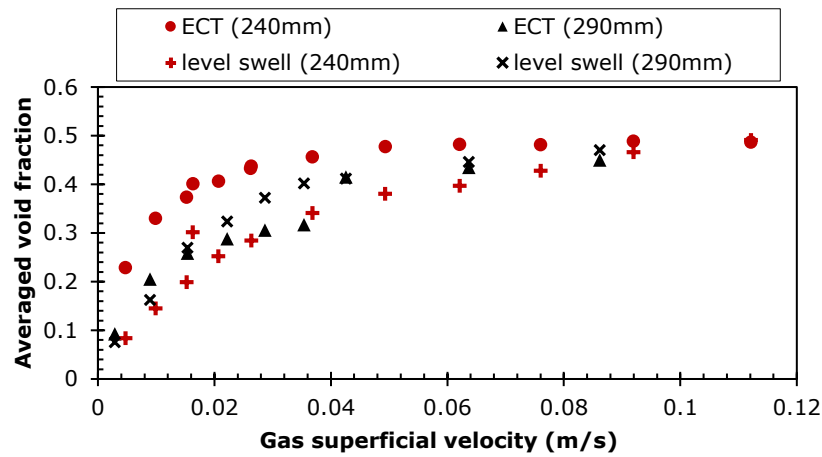


Figure 5.2: Comparing time-averaged void fraction obtained from ECT and level swell for air flow in high viscosity Silicone oil in different column diameters and gas superficial velocities.

5.2.1 Time Series of Cross-Sectional Averaged Void Fraction

The time series of the cross-sectional averaged void fraction were obtained using ECT, which provides time and cross-section resolved information about phase distribution inside the column. In this work, an acquisition time of 600 seconds and a sampling rate of 50 Hz were used for all runs, i.e., 2x30000 cross section per run. The time series of void fraction plots for each gas superficial velocity are illustrated in chapter 4 in Figures 4.4 and 4.5. They show changes in flow pattern from bubbly to slug flow in both columns. The plots include the 0.005-0.114 m/s gas superficial velocity in Figure 4.4

(240mm diameter column) and 0.005-0.088 m/s in Figure 4.5 (290mm diameter column). In general, the flow structure consists of a homogeneous bubble of low frequency and longer liquid slug in between at low gas flow rates (bubbly flow). Then, with increasing gas flow rates, the bubbles become longer, more frequent, exhibit more coalescence, and shorter liquid slug lengths (slug flow). At the end of the slug flow region, the bubbles become very long (approximately half of the column length). See the effect of gas flow rate on the length and the frequency of the bubbles in both columns in Section 5.4.

Bubbly flow occurs for gas superficial velocities between 0.005-0.011 and 0.003-0.016 m/s in the 240 and 290 mm diameter columns respectively. In the smaller diameter column, the bubbles rise approximately at a constant velocity at the centre of the column with almost uniform size. They also appear to be as about the same length as the liquid slug. Compared with the larger diameter column, it starts with smaller bubbles with low void fraction and lower frequency. The length of the liquid slugs in the larger diameter column at this flow regime is higher than the length of the bubbles. The bubbles, then become larger, the void fraction higher and the frequency increases with increasing gas flow rate in this range. There is no widespread evidence of bubble coalescence in both columns in this flow regime.

Slug flow appeared at gas superficial velocities of 0.016-0.114 m/s in the smaller diameter column (Figure 4.4) and 0.022-0.088 m/s in the 290mm diameter column (Figure 4.5). Slug flow in both columns showed similar behaviour. The bubbles become longer, but still have the uniform structure with less liquid slug. Then from 0.036 m/s gas superficial velocity in both columns, the length of Taylor bubbles increases dramatically due to the coalescence of these bubbles (increases with increasing the gas flow rate).

The flow becomes less structured in both columns, especially in the last two graphs in this range (0.092 and 0.114 m/s in Figure 4.4 and 0.066 and 0.088 m/s in Figure 4.5). At the same time, the length of the liquid slug decreases with increasing the gas flow rate, as the Figures clearly show, the time of the liquid slug decreases gradually.

The time series of the cross-sectional averaged void fraction data was compared with data collected by using a Dantec Dynamic high-speed camera. Information about the camera is presented in Chapter 3. In this experiment, 3 gas superficial velocities of 0.003, 0.006, and 0.009 m/s were used to create bubbly flow. This range was chosen in order to obtain proper bubbles suitable for the camera setup. The data was collected in synchronise between ECT and the high-speed camera. The camera was located about 2 m from the bottom of the column while the ECT sensor is about 2.5 m. Therefore, each bubble is captured by the camera first, then it rises through the ECT sensor. Moreover, this explains the existence of the first bubble in the time series in the 3 plots in Figure 5.3 without the corresponding photos. For example, in the first plot the first bubble, captured by the camera at time 0, is captured by the ECT after 50 seconds. Figure 5.3 compares between the bubbles obtained from the high-speed camera and the corresponding bubbles in the time series from the ECT. In general, both methods show a good agreement. The number and the size of the bubbles appearing in the time series match the bubbles in the photos. In the first plot (0.003 m/s), the bubbles are small separated by long liquid slugs in the photos. This corresponds to the low void fraction bubbles and the longer time of the liquid slug in the time series. In the second plot, the diameter of the bubbles becomes larger with increased gas flow rate and this corresponds to the increase in void fraction in the time series. At this stage, a number of coalescence between the bubbles started to occur. The circled bubbles at the

Chapter Five: Large Bubbles in Viscous Liquids (Bubbly and Slug Flow)

time series show the coalescence between the bubbles that occurred when the bubbles were passing through the ECT sensor.

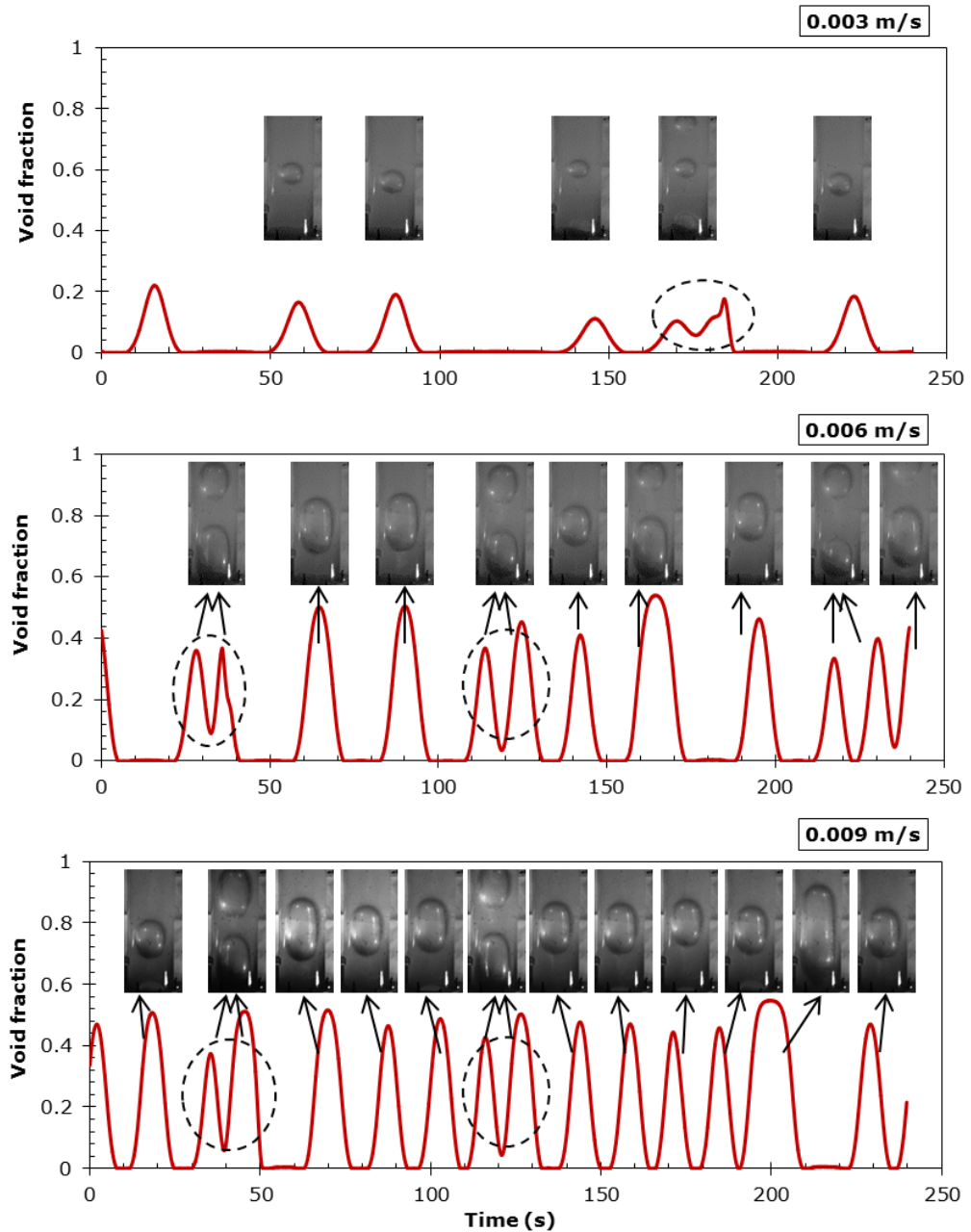


Figure 5.3: Comparing bubbles rising in stagnant 330 Pa.s and 290 mm diameter columns obtained from a high-resolution camera and ECT by comparing a single bubble image with the time series at 3 different gas flow rates.

Finally, the third plot at the higher flow rate in this experiment shows a good agreement between the other two methods as well. The higher void fraction corresponds to the increasing bubble diameters and the longer time at the time series to the longest bubbles in the photos.

5.2.2 Probability Density Function

The probability density function (PDF) is also used to indicate the flow structure in two-phase mixtures. It is used to present the void fraction data which is measured by the ECT in this work. As it is described in Chapter 2, Bubbly flow is recognised by a peak at low void fraction and spherical cap bubble flow by a peak and a tail towards high void fraction. Slug flow is characterised by the existence of two peaks. The small peak corresponds to the value of the liquid slug, while the large peak to the Taylor bubbles. In unstable slug flow, the value of the lower peak is significantly lower than it is in regular slug flow. This low peak starts to become less prominent until it disappears. In this work, PDFs of the high viscosity oils in both diameters show trends different from those reported by Costigan and Whalley (1997), see Chapter 2. The PDF files for bubbly and slug flow can be found in Figures 4.17 and 4.19 in Chapter 4. The plots include the 0.005-0.114 m/s gas superficial velocity in Figure 4.4 (240mm diameter column) and 0.003-0.088 m/s in Figure 4.5 (290mm diameter column).

Starting with bubbly flow, the PDF appears to have the same characteristics as spherical cap bubbles of water. It consists of two peaks, one with high void fraction and low PDF, which corresponds to the spherical large bubble, and another with zero void fraction and high PDF. This can be seen in gas superficial velocity of 0.005 m/s in Figure 4.18 and 0.003 m/s and 0.009 m/s in Figure 4.19. Increasing gas superficial velocities to 0.011 m/s in the 240 mm column and 0.015 m/s in the larger column increases the PDF of the large spherical bubbles.

In slug flow region, the PDFs for viscous oil appear to be more similar to the unstable slug than slug flow in water like the previous flow regime. The PDF shows two peaks of void fraction with a considerable difference. The low

peaks are at the lower values and the high peaks are at higher values of void fraction; Taylor bubbles, 0.016-0.114 m/s and 0.022-0.088 m/s in the small and the large diameter columns respectively. The slug flow in both columns records the highest PDF values for Taylor bubbles. The value of PDF decreases with increasing the gas superficial velocity. This indicates that the decreasing of the number of Taylor bubbles is due to the bubble coalescence and the creation of longer bubbles.

5.3 The Size of the Bubbles

The size of the bubbles in 330 Pa.s stagnant Silicone oil and 290 mm diameter column is determined by using the ECT. The experimental data was compared with a theoretical approach from literature. They were also compared with the results obtained from the high-resolution camera to determine the bubble size. In this experiment, the bubbles were created by using a single gas injection point and a low range of gas flow rate. This range has been chosen to create spherical bubbles with a maximum diameter of approximately $\frac{2}{3}$ of the column diameter in order to neglect the effect of the pipe wall. Then the results from the ECT could be compared with the equation by Gaddis and Vogelpohl (1986), [Equation 2.5](#), who considered a spherical bubble in space. This equation to predict the bubble diameter detachment has been validated over a wide range of gas flow rates and liquid viscosities by Gaddis and Vogelpohl (1986). The equation is illustrated in Chapter 2.

In general, data from ECT and Equation 2.5 showed maximum relative errors of 0.133 and 0.4 for the 240 and 290 mm diameter columns respectively within the selected range of gas flow rate. Gas superficial velocities of 0.0009 to 0.005 m/s and 0.003-0.026 m/s were used to create bubbly flow in 240 and 290 mm diameter columns respectively. First, the diameter of the

bubbles increases by 40 mm in both methods with increasing the gas flow rate to the maximum in the above range. Likewise, in the larger diameter column, the diameter of the bubbles increases by 90 and 82 mm in the case of ECT and the equation respectively for the gas flow rates shown in Figure 5.4.

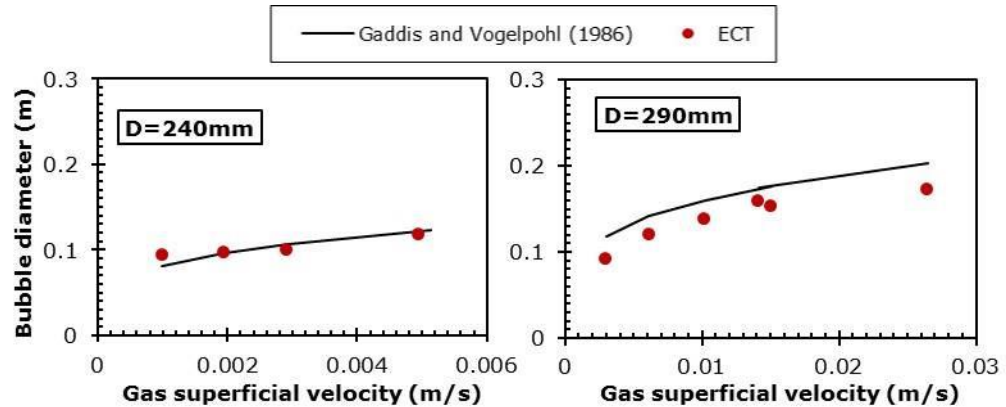


Figure 5.4: Comparing the diameter of the bubbles obtained from ECT with the theoretical approach by Gaddis and Vogelpohl (1986), Equation 2.5. The bubbles are rising in stagnant 360 and 330 Pa.s Silicone oil and 240 and 290 mm diameter columns respectively. One nozzle at the centre of the gas inlet section is used to create the bubbles.

The size of the bubbles depends mainly on the gas flow rate as the bubbles become larger with increasing gas flow rates. Figure 5.5 shows the increase in the size of the bubbles which determined using 3 different methods (data obtained from ECT, camera, and Equation 2.5). The figure also compares the volume of the bubbles in two different diameter columns. In the smaller diameter column, the maximum error calculated between the ECT results and theoretical values for the bubble size is 0.04%. In the larger diameter column, a high-resolution camera was used beside the ECT to validate Equation 2.5 by Gaddis and Vogelpohl (1986) in 330 Pa.s Silicone oil. The percentage of standard error between the 3 methods varies between 0.05%-1% at gas superficial velocities of 0.003-0.026 m/s.

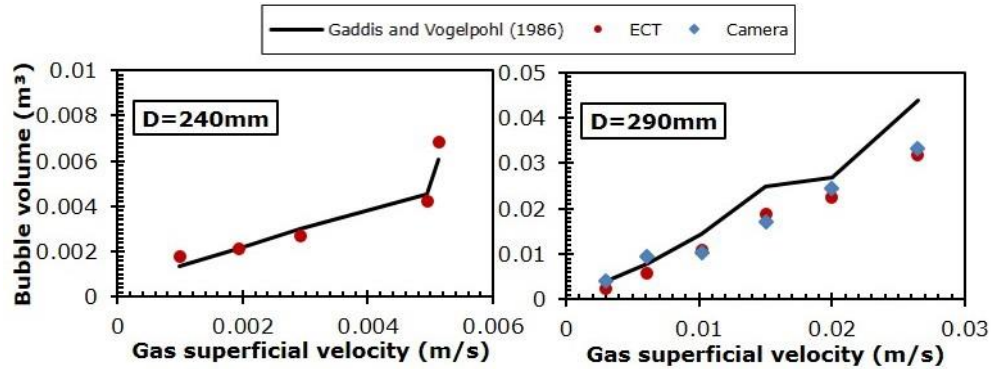


Figure 5.5: Comparing the volume of the bubbles rising in stagnant high viscosity oils. In the 240 mm diameter column, the comparison is between volume of the bubbles obtained from the ECT and theoretical approach by Gaddis and Vogelpohl (1986), Equation 2.5. In the larger diameter column, the comparison is between the volume of the bubbles obtained from the ECT, Equation 2.5 and results obtained using high-resolution camera. A single gas injection point is used at the centre of the gas inlet section.

Bubble diameter was also calculated using Davidson and Schuler (1997), [Equation 2.6](#), and Snabre and Magnifotcham (1998), [Equation 2.7](#), approaches. Bubble diameters, which were determined using Equations 2.6 and 2.7, showed standard errors of 24% and 71% in the 240 and 290 mm diameter columns respectively comparing with the data from ECT. Gaddis and Vogelpohl considered surface tension, viscous drag, and liquid inertia in the calculation of bubble diameter. This allows the equation to be applicable for a wide range of liquid viscosities and gas flow rates. On the other hand, in Equation 2.6 by Davidson and Schuler (1997), the surface tension force has not been taken into account in the calculations of the bubble diameter. Likewise, in the implicit equation (2.7) by Snabre and Magnifotcham (1998), the effect of liquid viscosity was not considered. Therefore, the last two methods might be applicable to a limited range of viscosity and surface tension. In the present study, high viscosity liquids of 330 and 360 Pa.s with low surface tension of 0.02 N/m were employed. This explains the significant difference in the bubble diameter values calculated by the last two methods with values calculated by Gaddis and Vogelpohl (1986) and measured by ECT. The theoretical approaches to calculating the bubble diameters are explained in Chapter 2.

Another experiment to determine the diameter of the bubbles was carried out on using a wide range of gas flow rate to create bubbly and a slug flow regime in 290 mm diameter column. The ECT, together with the High-resolution camera (Canon EOS 600D), was used for this purpose. ImageJ software was used to analyse the photo obtained from the high-resolution camera and compared with the ECT results. These photos also were corrected using a method by Rana et al. (2015), who employed a high-speed camera to study the phenomena of Taylor bubble bursting in three different viscosity liquids. According to the results in Figure 5.6, the photos corrected by Rana's method showed a standard error of 2.6% and 2.9% at bubbly and slug flow regimes respectively, when compared to the raw photos.

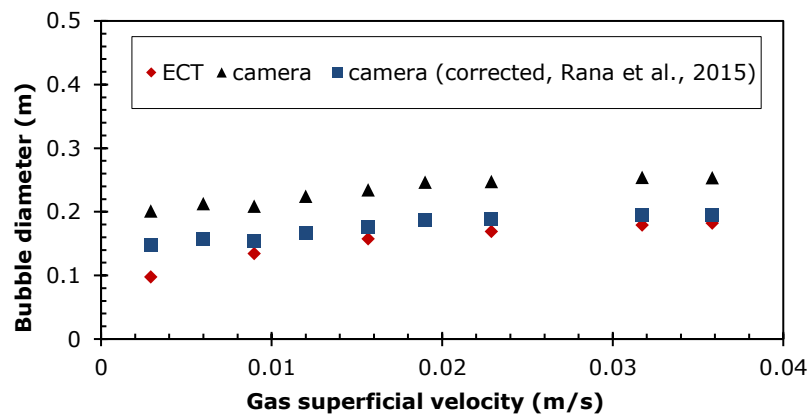


Figure 5.6: Comparing the bubble diameters obtained from the ECT with the values from high-resolution camera and the corrected images using Rana et al. (2015) method. The bubbles are rising in a 330 Pa.s Silicone oil and 290 mm diameter column. 5 nozzles are used to inject the air into the column.

On the other hand, the results from the ECT are more close to the corrected photos by Rana than the raw photos. For example, the percentage of the difference between the results of ECT and raw photos is between 5-3.5% at gas superficial velocities of 0.003 and 0.036 m/s respectively. Whereas, the difference is only 2.5-0.6% between the ECT and the corrected photos by Rana's method (at the same range of gas superficial velocities).

5.4 The Dimensions of the Slug Unit

This section presents the dimensions of the slug unit in bubbly and slug flow regions. The behaviour of bubbly flow in viscous liquids seems to be quite similar to slug flow in terms of flow characteristics. The only difference is in the bubble length/size. The dimensions of the slug unit, considered in this section, are large bubble length (spherical, ellipsoidal and Taylor bubble), liquid slug length, and liquid film thickness around the large bubbles.

5.4.1 Taylor Bubble and Liquid Slug Lengths

The length of large bubbles (Taylor bubbles) and liquid slugs was determined in stagnant 360 and 330 Pa.s Silicone oil in 240 and 260 mm diameter column. Gas superficial velocities of 0.005-0.114 m/s and 0.003-0.008 m/s are used in the 240 and 290 mm diameter columns respectively to create the bubbly and the slug flow regimes. The results were calculated from the time that each bubble/liquid slug passes through the ECT probes for each velocity. The experiments have been repeated to calculate the standard error for each flow rate in this range. Moreover, the experimental results obtained from the ECT were compared with theoretical approaches as well as other experimental methods. This section focuses on the length of "Taylor bubble" as the large bubbles in the bubbly flow seem to increase in size (spherical and ellipsoidal bubbles) with increasing the gas flow rate.

In general, the averaged lengths of Taylor bubbles increase significantly in comparing with liquid slug lengths which increase slightly with increasing the gas flow rate in both columns as in Figure 5.7. First, the length of Taylor bubbles increases significantly by 7.62 m in the small diameter column, whereas it increases by only 3.42 m in the larger by the end of the slug flow region. The range of slug flow regime is seen to be higher in the smaller

diameter column, it ends at a gas superficial velocity of 0.114 m/s before the transition to churn flow starts. This explains the significant difference in increasing the bubble lengths in both columns.

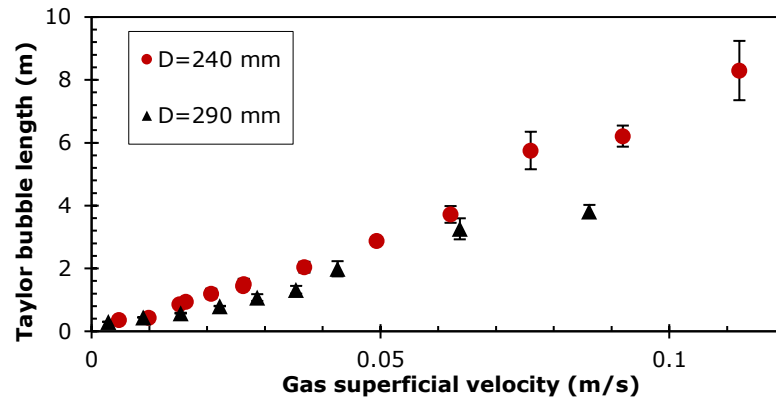


Figure 5.7: The length of the large bubbles in bubbly and slug flow regimes in viscous liquids and two different diameter columns.

However, the bubbles appear to be longer in the smaller diameter column for the most gas flow rates, because the rate of coalescence between the bubbles is higher in the smaller diameter column for the similar gas flow rates. The error bars in the figure represent the standard error which is calculated from the standard deviation. They vary between 2.2 to 94% from lower to higher gas flow rates respectively in the 240 mm diameter column. However, they appear lower in the larger diameter column for the same flow regimes. They vary between 0.6% minimum and 33% maximum errors in both flow regimes.

The second figure, which is Figure 5.8, compares large bubble and liquid slug lengths in both columns. According to the results, the length of the large bubbles increases significantly in the smaller diameter column by 7.62 m. However, liquid slug length increases slightly by only 0.99 m in the same column for the same range of gas flow rate. Likewise, in the 290 mm diameter column, Taylor bubble length increases significantly by 3.4 m while liquid slug length increases by only 0.61 m in the range of gas flow rate. The

length of liquid slug in the smaller diameter column seems to be slightly higher than the larger diameter column for most of the gas flow rates.

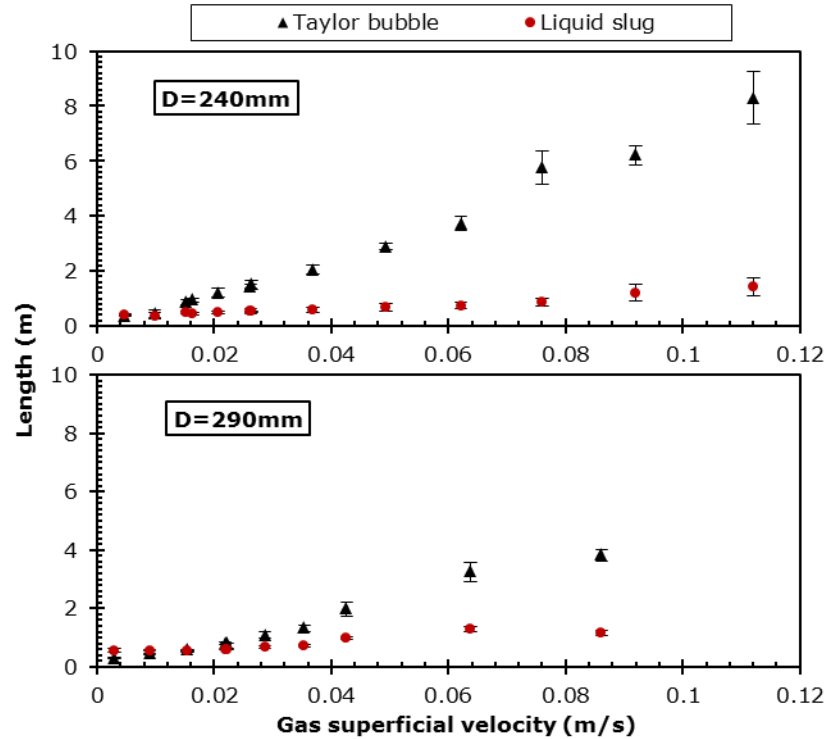


Figure 5.8: Comparing the length of Taylor bubbles and liquid slug in high viscosity oils in 240 and 290 mm diameter columns.

The experimental results obtained from the ECT were compared first with a theoretical approach by Khatib and Richardson (1984) which have been clarified in Chapter 2. In general, both results (experimental and theoretical) show a similar trend for a whole range of gas flow rate. This comparison can be seen in Figure 5.9. The maximum relative errors between the two methods are 0.2 and 0.9 for Taylor bubbles length in the 240 and 290 mm diameter column respectively. Similarly, maximum relative errors are 0.3 and 0.8 in the 240 and 290 mm diameter columns respectively for liquid slug lengths. The difference between the results from the ECT and Khatib's approach becomes high at higher gas flow rates, particularly in the smaller diameter column due to increasing coalescence between the bubbles. The difference in the values between both methods may belong to the methods of the determinations of the lengths and the structure of the flow. The values

obtained from the ECT are calculated from the structure velocity and the time of bubble/slug from the time series of void fraction. On the other hand, the numerical method by Khatib and Richardson (1984) uses the values of the large bubble and liquid slug void fractions from the PDF averaged void fraction.

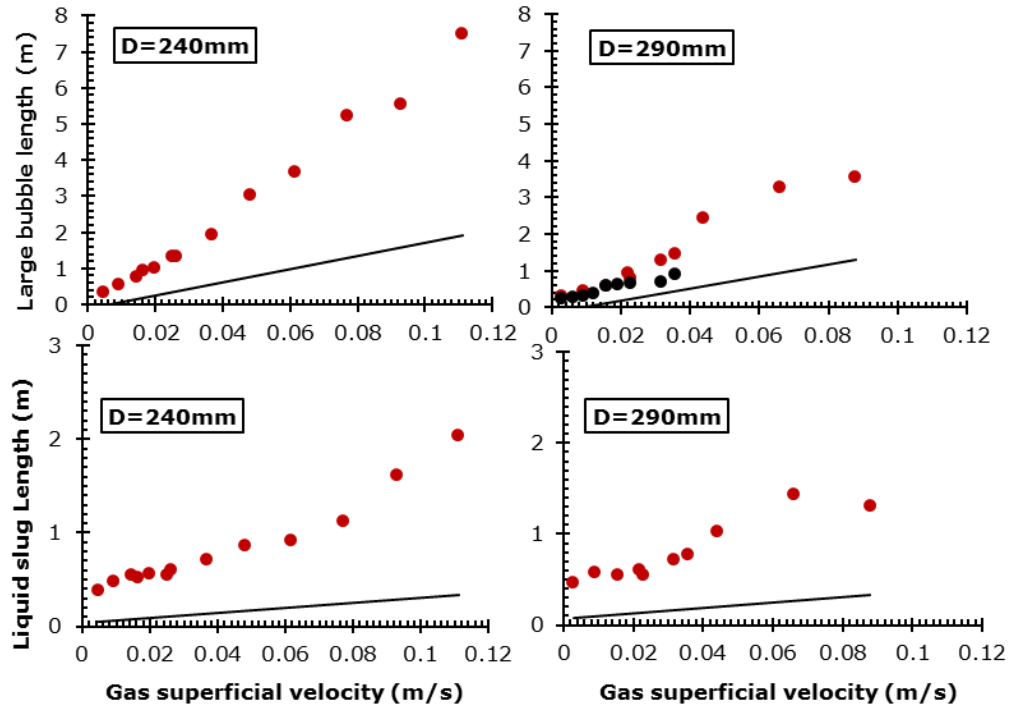


Figure 5.9: Experimental and theoretical lengths of Taylor bubble and liquid slug flow in high viscosity Silicone oil in two different diameter columns. • ECT, • High-resolution camera, — Khatib and Richardson (1984).

The second method could be applied to low gas flow rates when the flow is structured. The PDF-void fraction consists of two defined peaks in this range of flow rates. While in higher gas flow rates the flow become less structured due to the significant increasing the coalescence between the bubbles. In this case, more than one peak of high void fractions appear corresponding to the different PDF for the Taylor bubbles (see Figure 4.19 in Chapter 4). Therefore, it becomes difficult to define the void fraction and obtain results close to the ones obtained from the ECT. The same theory could be applied to the length of the liquid slug. In addition, a second approach was employed to determine bubbles length whereby videos were taken by using High-

resolution camera. ImageJ software was used to analyse the photographs obtained from the high-resolution camera. The images correction factors have not been used for the bubble length determination as these are not affected by the pipe curvature. The results obtained are plotted in Figure 5.9. The relative error between the two methods at lower gas flow rates is 0.04. The relative error between the two methods increases to 0.06 at the highest gas flow rate in this range. The difference between both methods is only 0.007 m at 0.003 m/s though it is 0.65 m at 0.035 m/s.

The length of the individual large bubbles was calculated from data of void fraction-time series obtained from the ECT. It was calculated by multiplying the time of each bubble by the structure velocity which was determined from cross correlating the void fraction from the two planes of the ECT. Figure 5.10 illustrates the individual bubble lengths versus the time of each bubble for different gas flow rates (bubbly to slug flow regime).

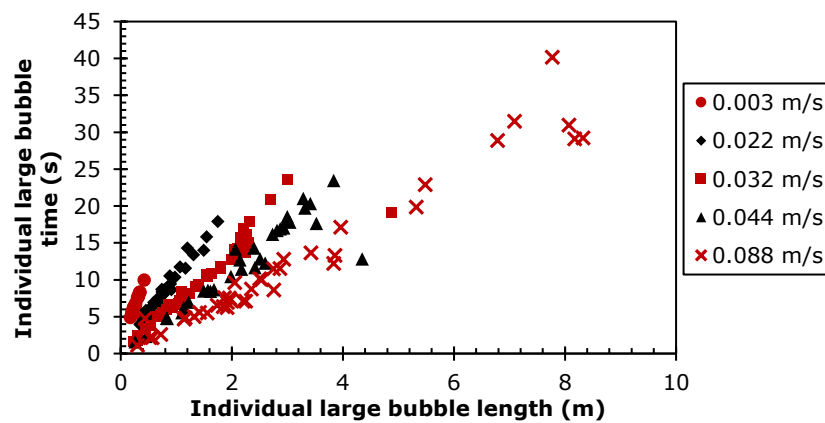


Figure 5.10: The lengths of the individual large bubbles at different gas flow rates. The bubbles are rising in 330 Pa.s Silicone oil and 290 mm diameter column.

The lengths of the bubbles are directly proportional to the time of all flow rates as expected. The figure also shows increasing the slope with increasing the gas flow rate due to increasing the bubble lengths and velocity. The frequency of the large bubbles was calculated from the time series data by counting the number of the bubbles per unit time for each gas flow rate. In

general, bubble frequency is seen to increase during bubbly flow until approximately the middle of the slug flow region, then it decreases gradually to the end of the slug flow region. The reason behind this is that the amount of the bubbles increase with increasing gas flow rate until it reaches a point when the bubbles become longer with increasing flow rate. At this point, the frequency starts to decline due to the increasing of the bubbles length, which dominates the structure of the flow, see Figure 5.11.

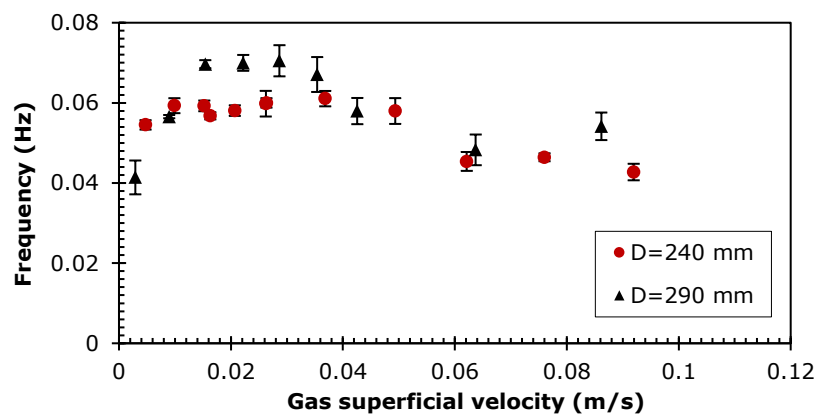


Figure 5.11: Frequency of large bubbles in high viscosity liquids and two large-scale pipes. The standard errors are equal to (0.08-0.3) % and (0.04-0.4) % for the 240 and 290 mm diameter columns respectively.

At the same time, the frequency of the bubbles appears to be more affected in the larger column by increasing the gas flow rate. It increases significantly from 0.04 to 0.07 Hz due to increasing gas flow rate, then declines to 0.05 Hz at gas superficial velocities of 0.003, 0.028, and 0.086 m/s respectively in the 290 mm diameter column. In contrast, it increases slightly from 0.054 to 0.06 Hz, then declines to 0.05 Hz at gas superficial velocities of 0.005, 0.049, and 0.11 m/s respectively in the smaller diameter column. This might be due to the structured flow in the larger diameter column, which contains more homogenous bubbles with less coalescence compared with the smaller diameter column. See the time series data in Chapter 4.

5.4.2 Liquid Film Thickness

Liquid film thickness around the large bubbles is calculated from the mean void fraction of air–Silicone oil in both columns. It is determined by the geometric relationship $\delta = (D / 2) (1 - \sqrt{\epsilon_g})$, where δ is the film thickness, ϵ_g is the void fraction obtained from the ECT. The range of gas superficial velocities is the same as the previous section. In general liquid film thickness decreased with increasing gas flow rates due to the increasing of the diameter of the bubbles. Figure 5.12 shows how mean liquid film thickness/pipe diameter is dependent on the gas flow rate. According to the results, the values of the film thickness are more affected by increasing the gas flow rate in the larger diameter column.

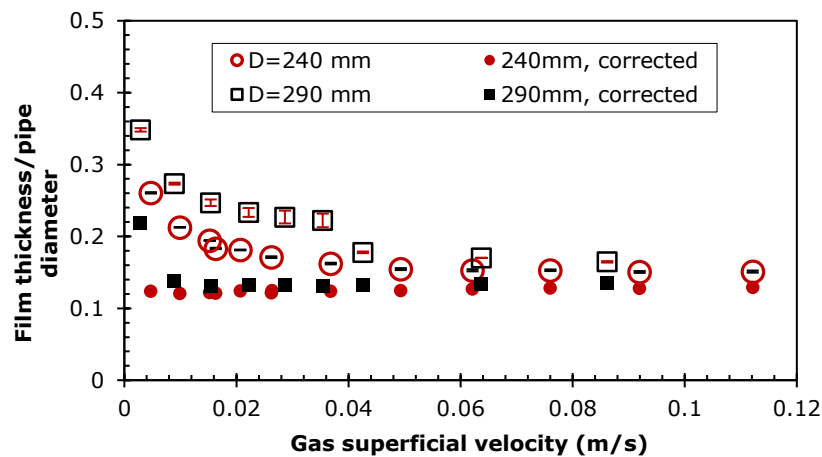


Figure 5.12: The falling film thickness of Silicone oil around the large bubble divided by the pipe diameter to convert it to dimensionless values, showing the standard errors for both 240 and 290 mm diameter columns. The corrected film thickness is calculated using Equation 2.13 by Azzopardi et al. (2014b).

Film thickness decreases by 26 mm in the smaller diameter column while it decreases by 52 mm in the larger one. At the same time, the amount of the decreasing in the film thickness changes between flow regimes in the same diameter column. It is higher in the bubbly flow region while it is slightly lower at slug flow. For example, in the 240 mm diameter column, it decreases by 16 mm at the bubbly flow region (0.005-0.011 m/s), while it

decreases by only 10 mm at slug flow (0.016-0.114m/s). Likewise, in the larger diameter column, it decreases by 32 mm by the end of bubbly flow while it decreases by only 20 mm during the slug flow. This might be due to characteristics of Taylor bubble, which seem to be affected in terms of length more than the diameter with increasing gas flow rate due to the effect of the pipe wall. See Figure 5.7, Taylor bubble length in both columns. The error bars (standard error) in the figure are calculated from the standard deviation and the square root of the number of the repetition of the experiments. The maximum errors that are recorded for the film thickness are 0.2% and 0.3% for the 240 and 290 mm diameter columns respectively. The experimental film thickness which was calculated by using the geometric relationship does not take in consideration the small bubbles which accompanied the large bubbles. These small bubbles are distributed in the falling film and the liquid slug and contribute the void fraction. The corrected film thickness was determined from the void fraction of Taylor bubble and liquid slug. The equation is shown in Chapter 2, Equation 2.13. This equation seems to have a maximum relative error of 0.1 in both columns with the geometrical relationship at the intermediate and high gas flow rates. However, the values of the relative errors appear to increase with decreasing the gas flow rates to 0.5 and 0.3 in the 240 and 290 mm diameter columns respectively. This might be due to the low density of the small bubbles at the low gas flow rates. The concentration of the small bubbles increases with increasing the gas flow rate and the time of gas injection, this is shown in Chapter 8, Section 6.2. The corrected film thickness does not show a significant change with increasing gas flow rate. This might not be applied for the low gas flow rates as the film thickness decreases significantly corresponding to the increasing of the bubbles diameter.

The film thickness around Taylor bubble was calculated using the Cylindrical

Nusselt equation which is explained in Chapter 2, Equation 2.11. The theoretical film thickness shows relative errors of 0.5 and 0.4 with the film thickness determined by using the ECT in the 240 and 290 mm diameter columns respectively. The reason behind that might be due to the equation, which assumes that the whole liquid film around Taylor bubble flows down and is displaced by the bubbles. In that case, F_D , which is a factor related the amount of the liquid film falling down, is equal to 1. Then the value of F_D for this experiments is less than 1 which is less than expected. This might belong to the difference between the flow rate of the upward and the downward liquid film. The quantity of the liquid film flowing down around Taylor bubbles is less than expected compared to the upward flow. The difference in film thickness between the two methods is displayed in Figure 5.13.

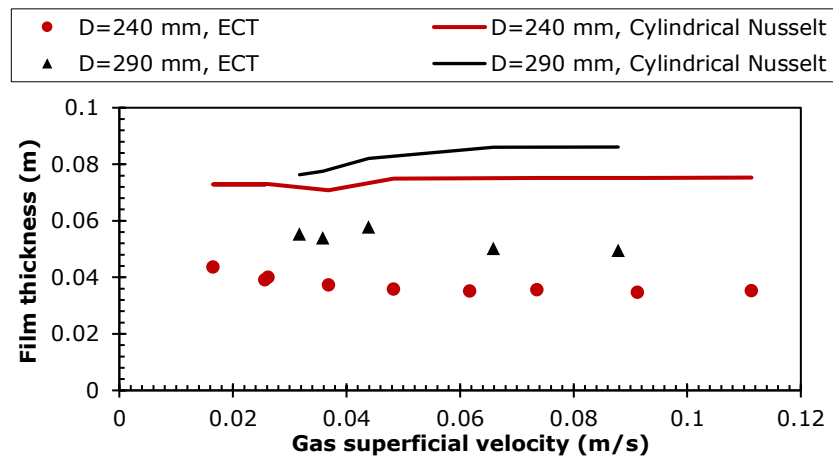


Figure 5.13: Comparing the corrected experimental (ECT) by Azzopardi et al. (2014b) and calculated (Cylindrical Nusselt equation) film thickness around Taylor bubbles rising in 360 and 330 P.s Silicone oils and 240 and 290 mm diameter columns respectively.

The experimental liquid film, which is used in this compression, is the corrected film thickness that determined using Equation 2.13. This equation also is used to estimate the value of F_D in this work. The values of the film thickness in this figure show similar trend with the film thickness/pipe diameter in Figure 5.12 and film thickness in Figure 4.28 in Chapter 4. The

methods show similar values for the mean film thickness at this range of flow rate.

The fraction of the falling liquid around the large bubbles in high viscosity oils was calculated for both columns. The methods which were applied by Nogueira et al. (2006) and James et al. (2008) were used to calculate the fraction of the liquid flowing down. In Figure 5.14, the fraction of the liquid flowing down is calculated by dividing the value of downward liquid flow by the upward flow of the liquid film. The upward flow is determined experimentally from the void fraction of Taylor bubble and liquid slug (Equation 2.13). The downward flow is determined from Equation. 5.1. This equation is solved from Equation 2.11 in Chapter 2 by using F_D from data of Nogueira et al. (2006) and calculating the velocities of the bubbles and the top surface from the data of James et al. (2009);

$$U_B \pi (R - \delta)^2 = U_B \pi R^2 + \left(\frac{\pi \rho_l g}{8l} \right) \left[-4(R - \delta)^2 R^2 + R^4 - 4(R - \delta)^4 \times \ln \left(\frac{R - \delta}{R} \right) + 3(R - \delta)^4 \right] . \quad (5.1)$$

The dimensionless group of $Fr Oh$, which consists of Froude number, $Fr = U_{gs} / \sqrt{gD}$, and Ohnesorge number, $Oh = \mu_l / \rho_l \sigma D$, is based on F_D as it is dependent on gas superficial velocity, pipe diameter, liquid viscosity, and surface tension.

$$F_D = \frac{0.36}{(Fr Oh)^{0.48}} , if Fr Oh > 0.12 . F_D = 1 , if Fr Oh < 0.12 . \quad (5.2)$$

The fraction of the liquid flowing down seems to decrease with increasing the group of $Fr Oh^{0.8}$ which consists of viscous force divided by the square root of the surface tension and the gravitational force. Azzopardi et al. (2014b) determined the fraction of fluid falling down by using liquid viscosities of 1, 5, and 300 Pa.s. Similar trends were obtained in both works.

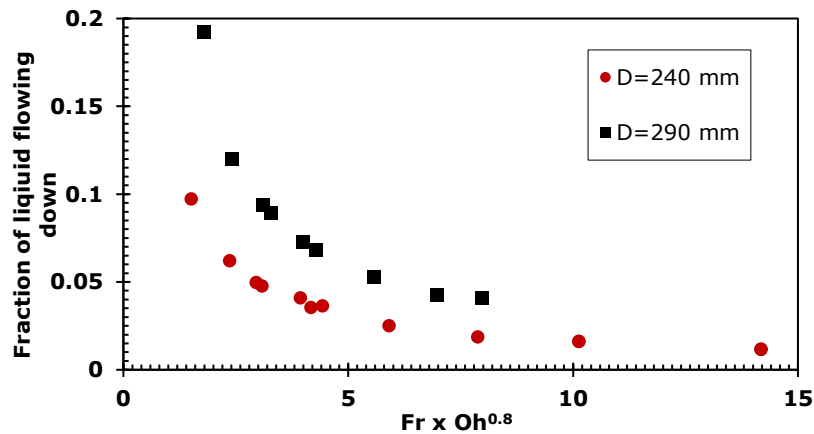


Figure 5.14: The fraction of the falling liquid film around Taylor bubbles rising in stagnant 360 and 330 Pa.s Silicone oil and 240 and 290 mm diameter columns respectively. $Fr \times Oh^{0.8}$ is a dimensionless group number represents Froude number and Ohnesorge number. The values were calculated using an approach by Azzopardi et al. (2014b).

5.5 Bubble Coalescence

The mechanism of the bubble coalescence in high viscosity oil was observed at different flow rates and compared to lower viscosity liquids from previous works. Photos for the coalescence between the bubbles were obtained using Canon EOS 600D high-resolution camera (18-megapixel resolution). The bubbles were rising in 330 Pa.s Silicone oil and 290mm diameter column at gas flow rates of 0.003, 0.006, 0.009 and 0.032 m/s. The photos were taken from about 0.4 to 1.4 m from the gas injection section.

Figure 5.15 shows the stages of coalescence of two bubbles in the 330 Pa.s Silicone oil and 290 mm diameter column at a gas superficial velocity of 0.003 m/s (bubbly flow). Coalescence occurs when there are two bubbles of different sizes (normally the upper bubble is smaller) rise with different velocities in the liquid. The larger bubble rises faster than the smaller bubble as the difference between the buoyant weight and the drag force is higher. The coalescence starts when the larger\ faster bubble comes very close to the upper bubble which starts to deform by creating a concave end at its

lower part. The lower bubble, then, starts merging with the upper one leaving a thin membrane of liquid between the bottom of the upper bubble and the top of the lower bubble. Later, the upper bubble expands and the lower one becomes smaller and elongated due to the merging between the bubbles, keeping the thin membrane between the original bubbles. Before the final stage, a new spherical bubble is created with a sharp cylindrical tail, the membrane between the original bubbles become very thin and disappear at the last stage. Finally, a new spherical bubble with a size corresponding to the combined volume of the two bubbles is created. The mechanism of coalescence of the bubbles in very viscous liquids was studied by Higuera (2005).

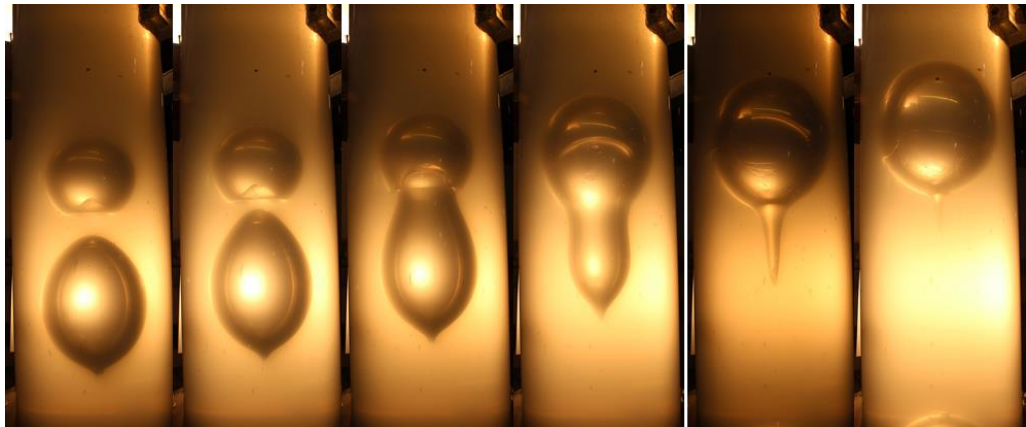


Figure 5.15: Bubble coalescence in 290 mm diameter column contains stagnant 330 Pa.s Silicone oil and flow at 0.003 m/s gas superficial velocity. The bubbles were created using single gas injection point. The time of the bubbles coalescence is 2 seconds.

Figure 5.16 is a photo of one stage before bubble coalescence in 330 Pa.s Silicone oil at a gas superficial velocity of 0.009 m/s. In this stage, the rounded bottom of the upper bubble deforms to an ununiformed concave shape. When the lower bubble rises, it pushes the liquid slug between the two bubbles to displace it by the lower bubble. The liquid flows downward as a falling film around the bubble to merge with the liquid slug behind the same bubble. This liquid slug between the two bubbles pushes the bottom of the upper bubble to prepare for the next stage (bubble merging). The

concave shape of the upper bubble becomes larger corresponding to the size of the lower bubble. The wake of the first bubble is laminar and no circulating region is created in such high viscosity oil. According to the visual observation, the velocity of the small bubbles in this area is virtually equal to the rise velocity of both large bubbles. The coalescence of the bubbles is considered to be another source for the small bubbles generation besides the top surface and the gas injection points.



Figure 5.16: A photo of the end and the top of two bubbles at one stage before coalescence in 330 Pa.s Silicone oil and 290 mm diameter column. The gas superficial velocity is 0.009 m/s.

The effect of the high viscosity of the oil appears in the bubble merging stage. A thin liquid film separates the end of the upper bubble from the nose of the lower bubble through all stages of coalescence. The required time for thinning then breaking this film is longer due to the high viscosity of the oil. See the film in the middle of the two bubbles in Figure 5.15 and Figure 5.17. The mechanism of the bubble coalescence seems independent of the size of the bubbles. The upper bubbles appear to follow the same scenario even if they are very small compared to the lower bubble. They are seen to expand and flatten in the first stages to correspond to the size of the nose of the lower bubble. This mechanism, too, is seen to be independent of the gas flow rate. Figure 5.17 compares the mechanism of bubble coalescence at two gas superficial velocities (0.006 and 0.032 m/s). In the below figure, (A) is bubbly flow and (B) is slug flow regime. Both bubbles at both flow rates (A

and B) are flowing the same steps and the liquid film between the bubbles requires similar time to break. Comparing both flow structures show increasing the density of the small bubbles at the higher flow rates due to increasing the gas flow rate and injection time, besides the coalescence of the bubbles.

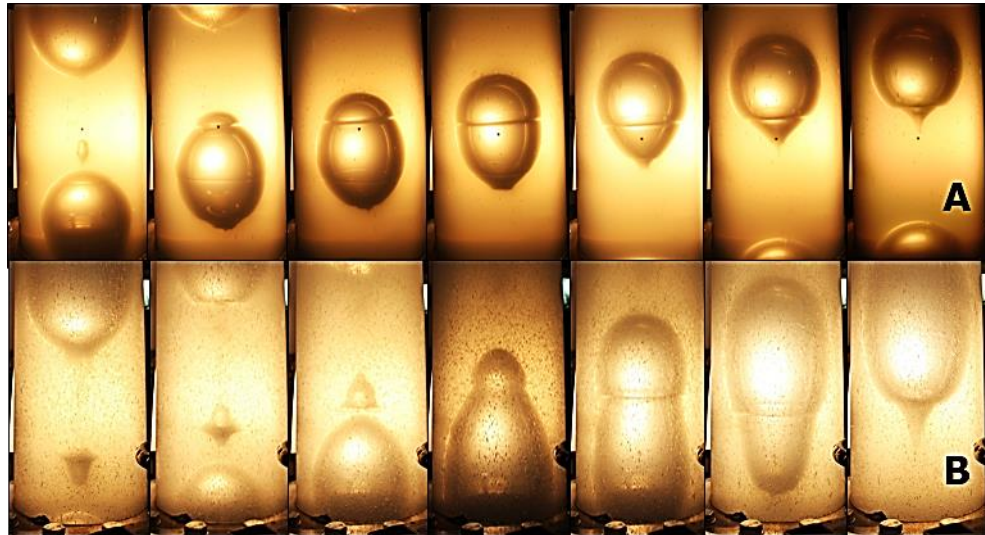


Figure 5.17: The stages of coalescence between two bubbles flowing at (A) 0.006 m/s and (B) 0.032 m/s in 330 Pa.s Silicon oil and 290 mm diameter column. The bubble created using single gas injection point. The time of the bubble coalescence for A and B is 2 seconds.

The mechanism of the bubble coalescence is different in bubbly flow, slug flow, and transition to churn flow. In slug flow, the mechanism is approximately similar, except the bubbles are larger and the time of the coalescence is longer. There is also the possibility of coalescence between more than two bubbles in rapid sequence. This was observed mostly in the transition to churn flow, which will be explained in Chapter 7. Taylor bubble becomes very long with small diameter open neck at the bottom of the bubble connected to the gas injection points.

The bubble coalescence in such high viscosity oil showed similar behaviour compared to the coalescence in lower viscosity liquids. In an experiment by Azzopardi et al. (2016), to study the coalescence between Taylor bubbles in 38 mm diameter pipe using different liquids viscosities, the mechanism was

similar. Figure 5.18 shows the coalescence of two bubbles rising in 0.108 Pa.s aqueous glycerol solution containing 1.3% potassium chloride with gas superficial velocity 0.035 m/s.

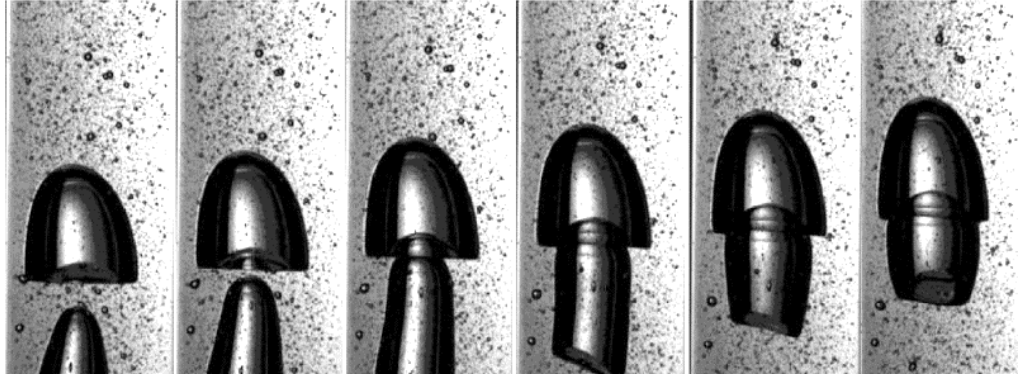


Figure 5.18: Coalescence of two bubbles rising in 0.108 Pa.s aqueous glycerol solution containing 1.3% potassium chloride with gas superficial velocity 0.035 m/s. (Azzopardi et al., 2016).

Figure 5.19 illustrates a similar coalescence between two bubbles in different conditions. The gas flow rate, liquid viscosity, surface tension, and pipe diameter seem not to have a significant effect on the mechanism of the bubble coalescence. The lower bubble merges with the upper bubble leaving the same thin membrane, which separates the volume of the gas inside each bubble until the last stage.

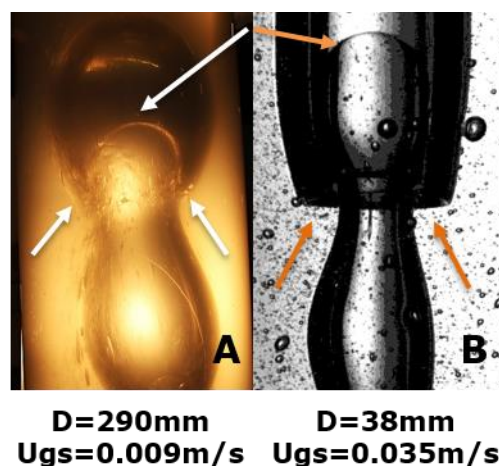


Figure 5.19: Comparing bubble coalescence in two different pipe diameter and liquid viscosities. (A): 330 Pa.s Silicon oil and surface tension of 0.02 N/m (B): from Azzopardi et al. (2016), 0.108 Pa.s aqueous glycerol solution containing 1.3% potassium chloride and surface tension 0.02 N/m.

5.6 Bubble Motion

The velocities of the large bubbles, in bubbly and slug flow, can be determined from the cross-correlation of the void fraction measured by the two planes of the ECT. The time delay between both signals and the distance between both planes were used to calculate the bubble velocity. This is the distance between the planes, divided by the time delay of the signals. In this work, bubble velocity was calculated for bubbly and slug flow in both columns. Information about experimental setup and gas flow rates is shown in Chapter 3. The experimental results obtained from the ECT were compared to another experimental method (camera) and to theoretical approaches from literature. In general, the average data for the large bubble velocity show a linear trend for both columns at both flow regimes, see Figure 5.20. The velocity of the bubbles increases due to the increasing of the bubble size with increasing the gas flow rates. Larger bubbles rise faster because of the balance between the buoyant weight and the drag force, (Feng, 2008). The error bars in the figure, calculated from the standard deviation, show a reasonable value for both columns. The values range between 0.02-2.6% and 0.06-1.2% in the 240 and 290 mm diameter columns respectively.

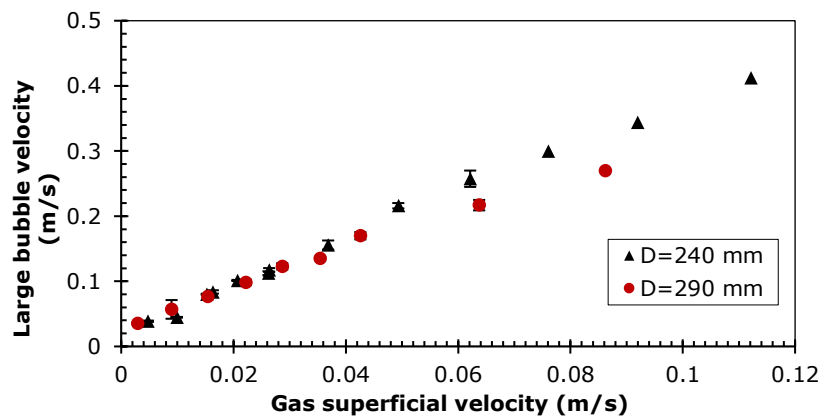


Figure 5.20: The velocity of the large bubbles with the standard errors in for air flow in viscous oils in two different diameter columns.

A comparison has been made between the velocity of the bubbles, obtained

from ECT, and other results from a high-speed camera. The data for both methods were captured at the same gas superficial velocities. The method of the analysis in this experiment was made by calculating the velocity of a single bubble from the time series of void fraction with the corresponding bubble obtained from a video from the high-speed camera. Both methods show the maximum relative error of 0.2, see Figure 5.21.

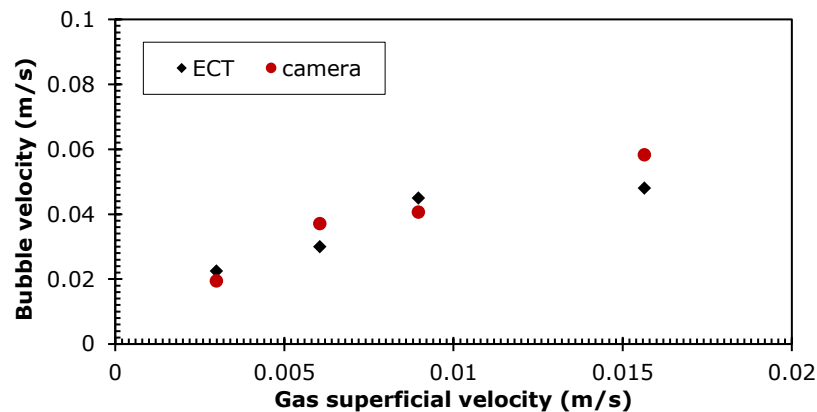


Figure 5.21: Comparing the velocity of a single bubble calculated from the time series obtained from the ECT with the corresponded bubble captured using the high-speed camera for each gas flow rate. The bubbles are rising in 330 Pa.s Silicone oil and 290 mm diameter column.

Figure 5.22 shows the comparison between the bubble velocities obtained from the ECT and the theoretical approach by Viana et al. (2003). The relative error between both methods is 0.5 to 0.02 for low and high flow rates respectively in the 240 mm diameter column. Similarly, relative errors are equal to 0.5 to 0.01 from low to high flow rates in the larger diameter column. Both values show similar trend particularly in the larger diameter column, the values seem to be very close. Although in the smaller diameter column the values seem to scatter with increasing the coalescence between the bubbles at the end of the slug flow region.

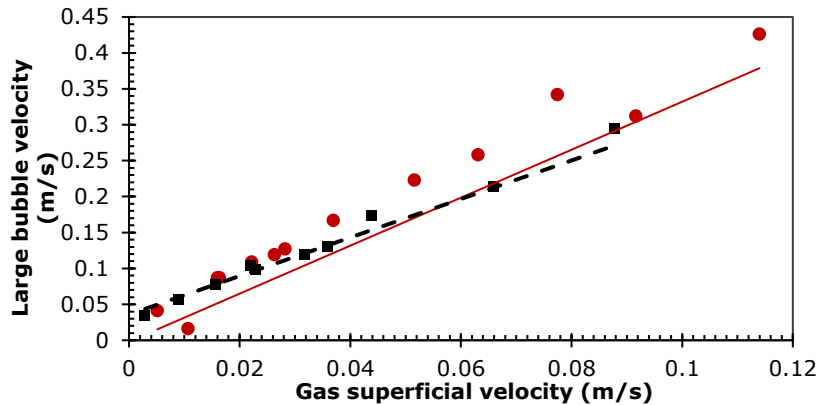


Figure 5.22: Theoretical and experimental large bubbles velocities in viscous Silicone oil for both columns. •: 240mm, ■: 290mm (experimental) and — 240mm, --- 290mm (from Viana et al. (2003), and Guet et al. (2004)).

5.7 Top Section and the Bursting of the Bubbles

The top section of the bubble column was observed so as to attain a better understanding of the behaviour of gas flow in viscous oils. The top surface oscillation was measured for different gas flow rates and was compared with the time series of the void fraction obtained from the ECT measurements. In bubbly and slug flow regimes, the top surface of the viscous oil in the column was observed to be rising and dropping periodically in a regular time and height. However, with increasing gas flow rates, at the transition to churn flow and at churn flow regimes, the top surface was seen to fluctuate in height and time of bubble bursting. The reason is that, at low gas flow rates (bubbly and slug flow), the bubbles rise with regular size and constant velocity. In contrast, at higher flow rates (transition to churn flow), the bubbles become very long and vary in lengths and time of bubble bursting. Finally, at very high gas flow rates (churn flow), the behaviour of the top surface changes significantly. The air flows in the core of the oil with a wavy passage, leaving the top surface open to the atmosphere with a possibility of existing a very long bubble. See the time series plots in Chapter 4. Defining an accurate value of the liquid height becomes difficult at high gas flow rate due to the liquid film, which remains on the pipe wall after the

bursting of the bubbles. This liquid film drains slowly around the ruptured bubble, merges with the liquid slug, and rises again by the next bubble. Moreover, the bubble bursting at the top section of the air-Silicone oil is considered as a major source of creating the small bubbles beside bubbles coalescence and the gas injection points. The mechanism of the generation of the small bubbles due to the bursting of the large bubbles was studied by Pandit et al. (1987).

Figure 5.23 shows the top section of the viscous oil column for low gas velocity (0.005 m/s), bubbly flow. The photos A-F illustrate the bursting of a single Taylor bubble at the top section of 360 Pa.s Silicone oil. The level of the top section changes from low to high corresponding to the passage of the Taylor bubbles.

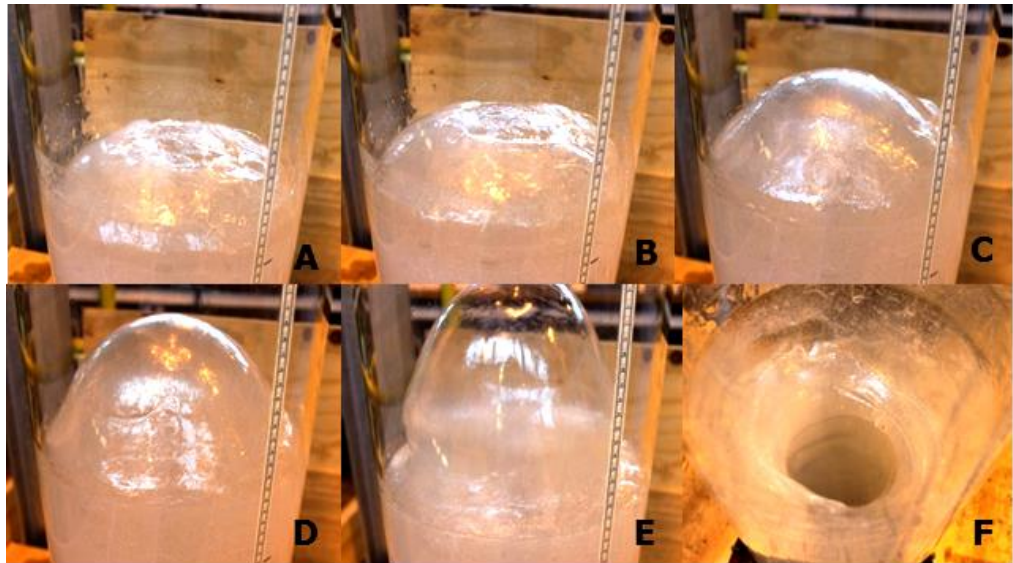


Figure 5.23: Photos of one bubble bursting at the top section of 240 mm diameter column and 360 Pa.s stagnant Silicone oil.

A single Taylor bubble, as seen in the photos, rises to the top section of the liquid until a thin film of oil (E) covers it. Then the bubble erupts leaving a thin liquid film on the pipe wall, this film falls down gradually around the ruptured bubble. Finally, (F) shows the inner section of the ruptured bubble which continues to rise.

To understand the behaviour of the top section of the gas-viscous oil mixture, the mixture height was measured at different flow regimes and was compared with the corresponding void fraction data. The effect of the gas injection time on the top surface was also studied in this experiment. Figure 5.24 compares: first, the top surface height of the mixture at 3 different gas flow rates/flow regimes; second, it compares the number of the bubble in each gas flow rate for equal periods. It also compares the behaviour of the top surface with the structure of the flow (the time series of void fraction).

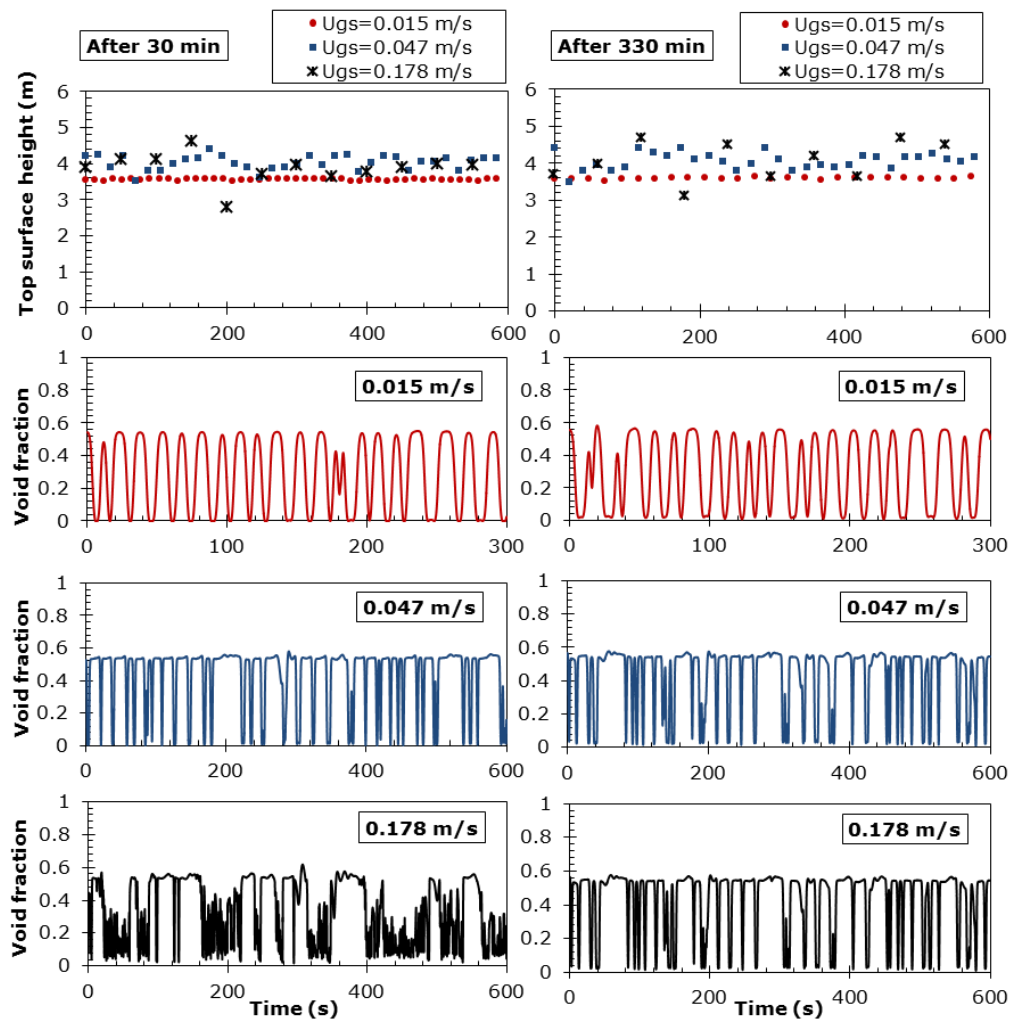


Figure 5.24: The position of the top section changing with time (counting the single bubble burst) during single runs for different gas superficial velocities in two different times during the day in 290 mm diameter column.

Finally, it compares the previous parameters in two different times: after 30

and 330 minutes of gas injection. Starting with the top surface height, the height of the mixture seems almost constant at bubbly flow (gas superficial velocity of 0.015 m/s). This is due to the structured bubbles which rise in constant velocity at bubbly flow. However, the top surface oscillation appears to fluctuate with increasing the gas flow rate. It fluctuates slightly at a gas superficial velocity of 0.047 m/s (slug flow) because of the variation of bubble length in this flow region, longer bubbles produce higher oscillation. Then at a higher gas superficial velocity (0.178 m/s) in the graph, which represents a transition to churn flow, the oscillation increases significantly. This is because of the structure of the flow, which contains very long bubbles, accompanied with the time of churning. The corresponding time series of void fraction in the same figure explain the behaviour of the top surface for each flow regime. Also, the number of the bursting bubbles was counted visually and compared to different gas flow rates and also compared with the corresponding bubbles obtained from the ECT. The results of this experiment agree with the measured bubble frequency for each flow rate, see Figure 5.11. The number of the bubbles decreases with increasing gas flow rate as a result of the increasing in the length of the bubbles. This can also be related to the bubble frequency, which in turn decreases with increasing bubble length. The time series of void fraction show how the length of the bubbles increases with increasing the gas flow rates (in this range of gas flow rates). Finally, the same data were collected after 5.5 hours to study the effect of gas injection time on the previous parameters. According to the results, the number of the bubbles/frequency decreases slightly with increasing gas flow rates (at this range) after injecting the gas for 5.5 hours. However, at the transition to churn flow, the frequency of the bubbles seems to increase when comparing the time series data with earlier time (after 30 minutes). At this flow regime, the observation at the top section of the liquid

might not be reliable due to the difficulty of locating the surface oscillation as it is mentioned earlier.

The top surface oscillation was also calculated and compared to the corresponding void fraction. An approach similar to the one applied by Grace et al. (1971) was used to calculate the positions of the top surface. As it is explained earlier, the top surface of the air-Silicone oil mixture has been seen to rise and fall periodically. It rises at velocity U_T equal to the difference between the Taylor bubble and the flow rate of the liquid which flow downward to merge with the liquid slug behind the Taylor bubble.

$$U_T = U_B \left(1 - \frac{2\delta}{D}\right)^2 (1 - F_D) \quad 5.3$$

The positions of the top surface were calculated by deriving an equation introduced by Azzopardi et al. (2014b) to locate the maximum positions-time for the top surface when the top of Taylor bubbles reach their maximum height just before the bursting stage. The n of each bubble can be calculated by deriving equations to solve the time t_n and position Z_{nH} for each corresponding bubble.

$$t_n = \frac{U_B}{(U_B - U_T)} \sum_{i=1}^{i=n} T_i + \frac{Z_o}{(U_B - U_T)} - \sum_{i=1}^{i=n-1} \frac{KL_{TBi}}{(U_B - U_T)} \quad , \quad 5.4$$

$$Z_{nH} = \frac{U_T U_B}{(U_B - U_T)} \sum_{i=1}^{i=n} T_i + \frac{U_B Z_o}{(U_B - U_T)} - \sum_{i=1}^{i=n-1} \frac{U_B KL_{TB(n-1)}}{(U_B - U_T)} \quad , \quad 5.5$$

$$Z_{nL} = Z_{nH} - KL_{TBn} \quad . \quad 5.6$$

Z_{nL} is the position of the top surface, K is a factor to allow for the draining of the falling film around Taylor bubbles. Calculations were made for the individual extracted Taylor bubbles and the mean values of the time T_i and the length of the bubbles, L_{TBi} . In the mean values, calculations of the value

of K was adjusted until constant results were obtained for the minimum and the maximum. The chosen values of K increased with increasing the gas flow rates. It started with 0.3 for gas superficial velocities of 0.022 and 0.032 m/s. Then it increased to 0.35145 and 0.41 for gas superficial velocities of 0.047 and 0.088 m/s respectively. Figure 5.25 displays the positions of the calculated top section using Equation 5.6. A comparison between the top surface position and the corresponding void fractions for different gas flow rates is shown as well. The difference between the maximum and the minimum positions increases significantly with increasing the gas flow rates in the individual.

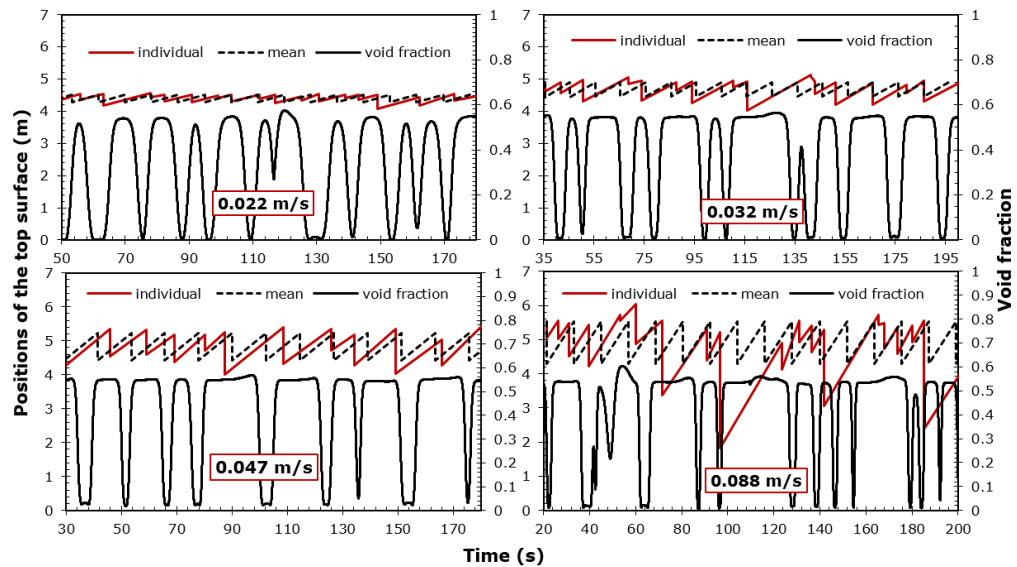


Figure 5.25: Comparing the positions of the top surface of the 330 Pa.s Silicone oil-air at different gas flow rates corresponding to the time series of void fraction obtained from the ECT at the same period of time and gas flow rates. The positions of the top surface were calculated using equations by Azzopardi et al. (2014b).

However, the mean values are seen to be slightly less affected. They are equal to 0.22, 0.45, 0.84, 1.17 m for the individual positions calculations from the lowest to highest gas superficial velocities in the graphs. And 0.2, 0.43, 0.78, and 1.18 m for the mean values of the top surface positions at the similar gas superficial velocities. This could be related to the lengths of Taylor bubbles, as they are more structured at lower gas flow rates. Then

with increasing gas flow rates and due to the coalescence of the bubbles, the length of the bubbles increases significantly causing higher fluctuation at the top surface. The graph of the gas superficial velocity of 0.088 m/s shows the synchronising between the lengths of a very long bubble and the corresponding top surface positions.

Comparing the positions of the top surface of the Silicone oil at gas superficial velocity of 0.047 m/s in Figure 5.24 and Figure 5.25 shows a similar trend. The range of liquid heights, which was determined experimentally, showed values close to the calculated heights by Azzopardi et al. (2014b) equation. Similarly, at lower gas superficial velocities (0.015 m/s experimental and 0.022 m/s calculated), the top surface of the liquid shows similar behaviour. It appears with less fluctuation for the top surface height due to the constant rise velocity for Taylor bubbles.

Figure 5.26 is a schematic drawing to show the overall picture of the flow pattern in high viscosity liquids and large diameter columns at low gas flow rates. The figure illustrates bubble creation, bubble coalescence and the bubble eruption at the top surface of the column. The videos of the top and the different parts of the column were collected together as a cartoon to illustrate the overall two-phase flow behaviour in the column. The letters (A-E) on the columns correspond to the gas-liquid flow behaviour with time at a constant gas flow rate. The numbers represent the level swell in the column which was discussed earlier.

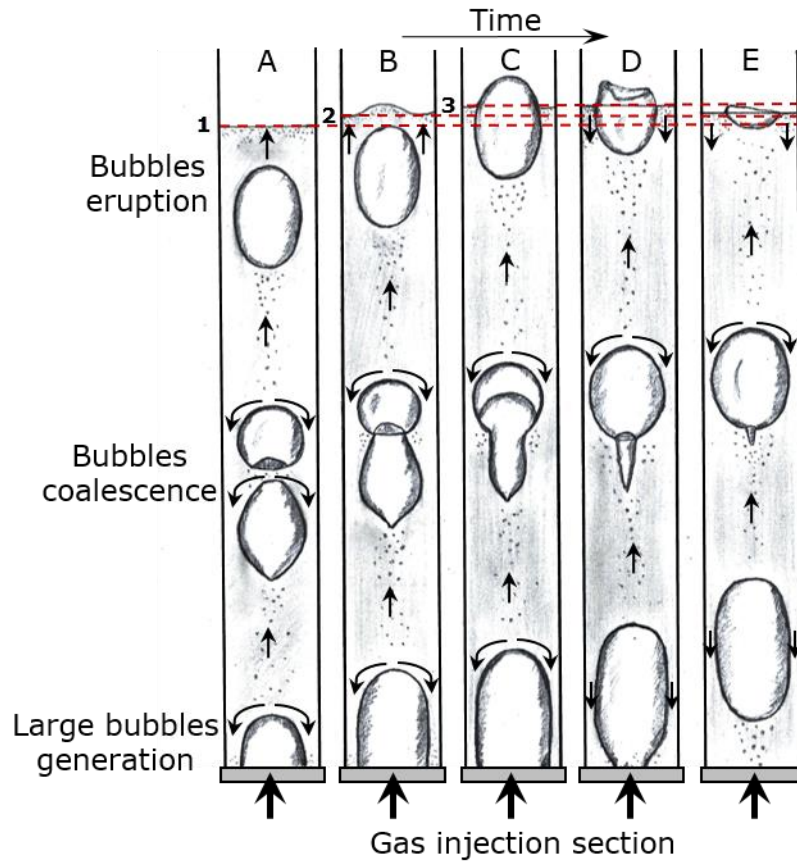


Figure 5.26: Schematic drawing showing the overall flow structure in high viscosity liquids and large diameter columns at low gas flow rates. The arrows in the figure correspond to the direction of the liquid flow, the numbers at the top section of the column correspond to the liquid levels in the column.

5.8 Summary

Bubbly and slug flow in high viscosity oils show completely different behaviours compared with low viscosity liquid. First, **bubbly flow** consists of a single large bubble that rises at a constant velocity at the centre of the column. These spherical and/or ellipsoidal bubbles, which are surrounded by a thick falling film, rise at low gas flow rates with a low density of small bubble in the liquid slug between the bubbles. Second, **slug flow** in viscous oils consists of long bubbles (Taylor bubbles) with rounded nose and end, occupies the cross section of the pipe, and is surrounded by a falling liquid film. Taylor bubbles that are followed by liquid slugs contain small bubbles

created at the top section of the liquid, gas injection points, and bubble coalescence and distributed all over the pipe.

The main findings from the experiments of bubbly or slug flow in 360 and 330 Pa.s stagnant Silicone oils in 240 and 290 mm diameter columns respectively are:

1. **Shape:**

- a. Spherical shape bubbles, at gas superficial velocities of 0.005 and 0.003 m/s in the 240 and 290 mm diameter columns respectively.
- b. Ellipsoidal bubbles, at gas superficial velocities of 0.011 m/s in the smaller diameter column and 0.009 and 0.016 m/s in the 290 mm diameter column.
- c. Long cylindrical bubbles with rounded top and end, at ranges of gas superficial velocities of 0.016-0.114 m/s and 0.022-0.088 m/s in the 240 and 290 mm diameter columns respectively.

2. Basing on the **time series of void fraction** signature, which is obtained from the ECT for the above range of gas flow rates in both columns. The flow structure consists of homogeneous bubbles of low frequency and longer liquid slug in between at low gas flow rates (bubbly flow). Then, with increasing the gas flow rates, the bubbles become longer, higher in frequency, more coalescence, and shorter liquid slug lengths (slug flow). At the end of the slug flow region, the bubbles become very long approximately half of the column length.

3. In bubbly flow, the **PDF** appears to have the same characteristics as spherical cap bubbles of water. It consists of two peaks, one with high void fraction and low PDF that corresponds to the spherical large bubbles and another with zero void fraction and high PDF. However, in slug flow region, the PDFs in viscous oil appear to be more similar to the unstable

slug than slug flow in water like the previous flow regime. The PDF shows two peaks of void fraction with a considerable difference. The lower peaks are at the low values and the higher peaks are at higher values of void fraction, Taylor bubbles.

4. **Dimensions:** the averaged **length of Taylor bubbles** increases significantly compared to the liquid slug lengths which increase slightly with increasing the gas flow rate in both columns. The bubbles appear longer in the smaller diameter column for the most gas flow rates because the rate of coalescence between the bubbles is higher in the smaller diameter column for the similar gas flow rates. On the other hand, the **length of liquid slug** in the smaller diameter column seems slightly higher than the larger diameter column for most of the gas flow rates. **Film thickness** decrease with increasing gas flow rates due to the increasing diameter of the bubbles. At the same time, the amount of the decreasing in the film thickness changes between different flow regimes in the same diameter column. It is higher in the bubbly flow region while it is slightly lower at slug flow.
5. The average data for the **large bubble velocity** shows a linear trend for both columns at both flow regimes and both columns. The velocity of the bubbles increases due to the increasing bubble size with increasing gas flow rates. Larger bubbles rise faster because of the balance between the buoyant weight and the drag force.
6. The **bubble coalescence** at different gas flow rates shows a similar mechanism. The larger bubble, which rises faster due to the difference between the buoyant weight and the drag force, merges with the upper bubble to create a new bubble with a size equivalent to the size of both bubbles. The rounded end of the upper bubble, deformed to a concave shape, correspond to the size of the nose of the lower bubble. Then, the

lower bubble merges with the upper one leaving a thin membrane between both bubbles. This membrane becomes thinner and bursts at the last stage of the coalescence. The same mechanism is seen in lower viscosities and smaller diameter pipes.

7. In bubbly and slug flow regimes, the **top surface of the viscous oil** in the column is observed to rise and drop periodically in a regular time and height. However, with increasing gas flow rates, at the transition to churn flow and at churn flow, the top surface is seen to fluctuate in height and time of bubble bursting. The reason behind that is that, at low gas flow rates (bubbly and slug flow), the bubbles rise with regular size and constant velocity. In contrast, at higher flow rates (transition to churn flow), the bubbles become very long and vary in lengths and time of bubble bursting of the bubbles. Finally, at very high gas flow rates (churn flow), the behaviour of the top surface changes significantly. The air flows in the core of the oil with wavy passage, leaving the top surface open to the atmosphere with a possibility of existing as a very long bubble
8. The comparison of the experimental results obtained from the ECT with the high resolution and high-speed cameras shows agreement in terms of bubble diameter and length. Likewise, comparing the ECT results for Taylor bubble and liquid slug length and large bubble velocity with the theoretical approaches by Khatib and Richardson (1984) and Viana et al. (2003) also shows agreement. In addition, validating the equation by Gaddis and Vogelpohl (1986) for 330 Pa.s Silicone oil shows good results.

Chapter 6

Gas Injection Methods

6.1 Introduction

This chapter includes results from a number of experiments, which has been carried out on the gas injection section at the bottom of 240 and 290 mm diameters columns. The effect of changing position, number, the distance between the gas injections points and gas injection time on the flow structure in high viscosity oils were studied in both columns. The rigs were designed to allow a full control on the 25 air injection points. Therefore, the number and the position of the air injection points in the column could be changed from the corresponding valves on the rig. The nozzles at the inlet section have been numbered and coloured to correspond to the coloured tubes on the rig; thus, each nozzle at the air inlet section can be controlled separately. More details about the design of the rig and the gas inlet section can be found in Chapter 3. Time series of void fraction, PDF, averaged void fraction, liquid slug length, film thickness large bubble velocity, frequency and lengths were determined and compared for different gas injection positions and flow rates in both columns.

6.1.1 Effects of Positions of Air Injection on the Flow Structure

This experiment focuses on the effect of the air injection positions in stagnant viscous liquids and the pipe diameter by using slug flow and large-scale pipes. A single injection point was used to inject air with a constant gas flow rate in both columns. The gas superficial velocity, for the 240 and 290 mm

rigs, were 0.04 and 0.03 m/s respectively; these were the closest values that could be arranged experimentally because of the different dimensions of the two rigs. In both rigs, 7 different positions were used to inject the air into the columns. The maps of the gas inlet section of the columns and the gas injections positions besides the procedure of the experiment are illustrated in Chapter 3.

In this range of flow rates, slug flow is observed to consist of large and small bubbles. Large bubbles (Taylor bubbles), of 1-1.5 m long, that occupies approximately the cross-section area of the pipe. These bubbles have rounded nose and end and are surrounded by a falling liquid film as described earlier. Furthermore, two more types of small bubbles are found. The one of centimetres diameter and the others are very small (few millimetres diameter) with a spherical shape. Full information about the small bubbles is shown in Chapter 8. Figure 6.1 shows the large bubble accompanied by small bubbles in different positions for the injection points.

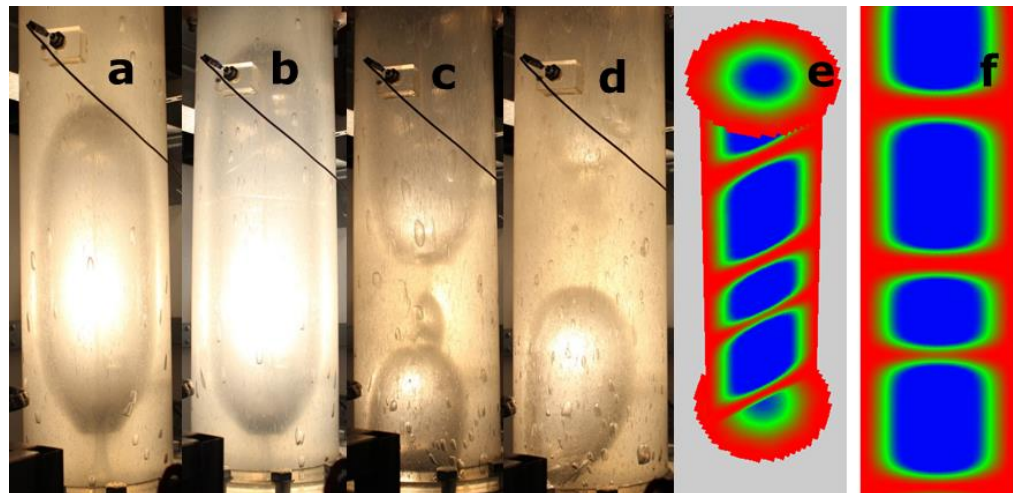


Figure 6.1: Taylor bubbles rising in 330 Pa.s Silicone oil and 290 mm diameter column using a single nozzle. Image (a) is where the nozzle at the centre of the column and (d) is (105mm) from the centre (next to the pipe's wall). (e) And (f) are the image reconstruction from the ECT software. The gas superficial velocity is 0.03 m/s.

The large bubble, from the nozzle next to the wall, moves to the centre of the column in approximately 0.7 m from the injection point, see (d) in

Figure 6.1. In the same figure, (e) and (f) are the reconstructed images from the ECT software (Plot3D [SVN 659]) that shows the flow structure.

The time series of the void fraction obtained from the ECT can show the behaviour of the Taylor bubbles for each position of the gas injection. In Figure 6.2 and 6.3, which show the flow structure in the 240 and 290 mm diameter columns respectively, the void fraction shows no significant effect of changing the position of the nozzles for either column. The exception is the nozzle nearest to the wall; the higher value of void fraction is due to an increase in bubble length. However, the flow structure seems affected in each position of the gas injection. Starting with the smaller diameter column (240 mm), the bubbles are more structured for all positions of the nozzles.

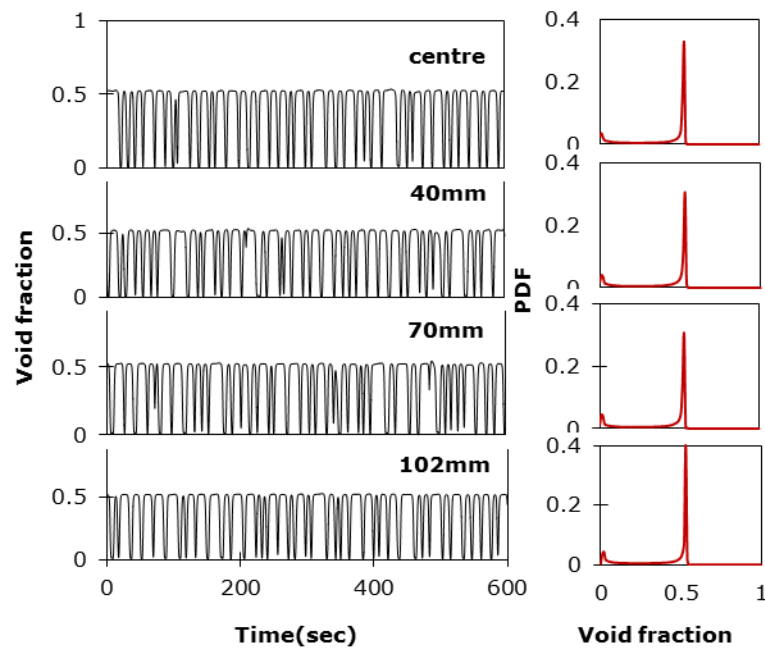


Figure 6.2: Time series of the cross-sectional averaged void fraction and the PDF files of slug flow in stagnant high viscosity oil at different positions for the gas injection in 240mm diameter columns, the gas superficial velocities are 0.04 m/s.

They have almost the same length with less bubble coalescence for each position, at the same time they become (0.3 m) longer in case of moving the nozzle toward the pipe wall. On the other hand, the larger diameter

column, Figure 6.3, the flow structure looks less organised than the smaller diameter column. The PDF plots in Figure 6.2 and Figure 6.3 display two picks. The one with higher void fraction and high PDF correspond to the Taylor bubbles, and the small pick with the smaller PDF represents the void fraction of the small bubbles in the liquid slug.

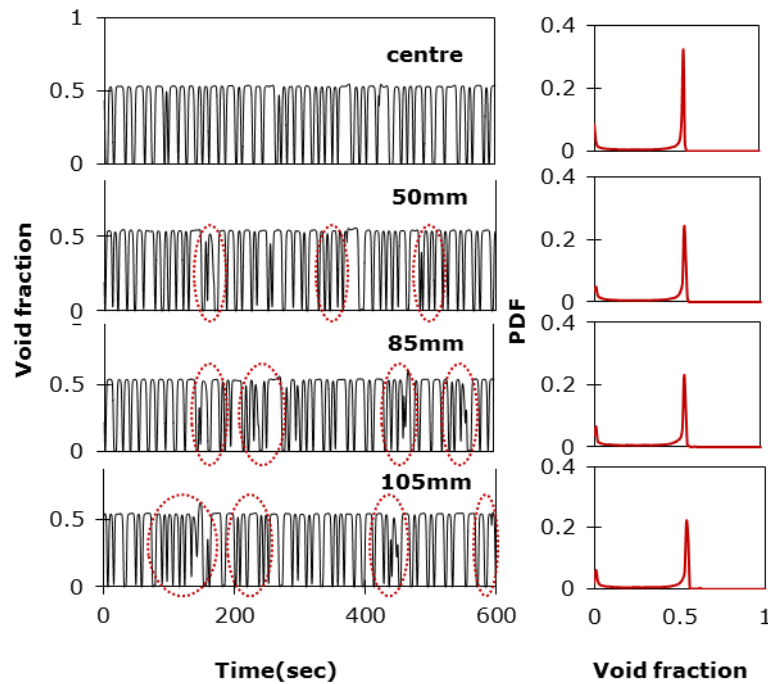


Figure 6.3: Time series of the cross-sectional averaged void fraction and the PDF files of slug flow in stagnant high viscosity oil at different positions for the gas injection in 290 mm diameter columns, the gas superficial velocities are 0.03 m/s.

The short bubbles, which dominate the structure, are accompanied by a number of quite long bubbles for the whole positions except the centre, which have a structure similar to the smaller diameter column. In general, the wall seems to have less effect in terms of bubble length with increasing the pipe diameter, whereas it slightly increases the frequency by 0.01 of the bubbles, which was calculated by counting the individual bubbles per time.

Figure 6.4 shows the effect of changing the position of gas injection points on the averaged void fraction in two different diameter columns. The mean void fraction presented in the figure shows no significant effects in both

cases, changing the position of the gas injection nozzles and the column diameter, as the void fraction is directly proportional to the gas flow rate. Standard errors of 0.5% minimum and 0.89% maximum are calculated for the 290 mm diameter column.

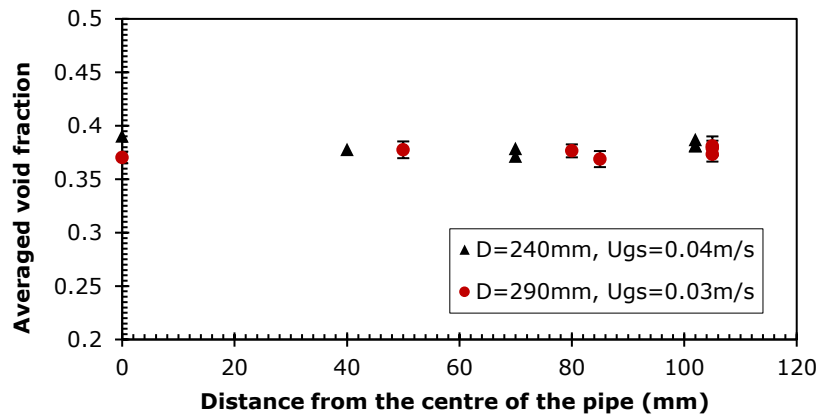


Figure 6.4: Averaged void fraction of slug flow in stagnant high viscosity Silicone oil using single nozzles and different positions of two different diameter columns.

Likewise, the bubbles velocity in Figure 6.5 did not show a significant effect by changing the positions of the gas injection points. The velocity of the Taylor bubbles was calculated from the cross-correlation of the void fraction signals from the two ECT planes. The time delay between the signals of the planes was used in conjunction with the distance between the planes to calculate the bubble velocity (velocity being equal length divided by time).

In considering the rise velocity of Taylor bubbles, it was noticed that the velocity of the bubble remains almost constant with changing the gas injection position in both columns. Nevertheless, the bubble velocity in the larger diameter column was slightly higher than the smaller one >0.02 m/s because of the collision between the bubbles (larger bubbles rise faster because of the balance between the buoyant weight and the drag force), Feng (2008). When compared with the analytical method by Viana et al. (2003), the ECT results showed maximum relative errors of 0.2 and 0.3 in the 240 and 290 mm diameter columns respectively.

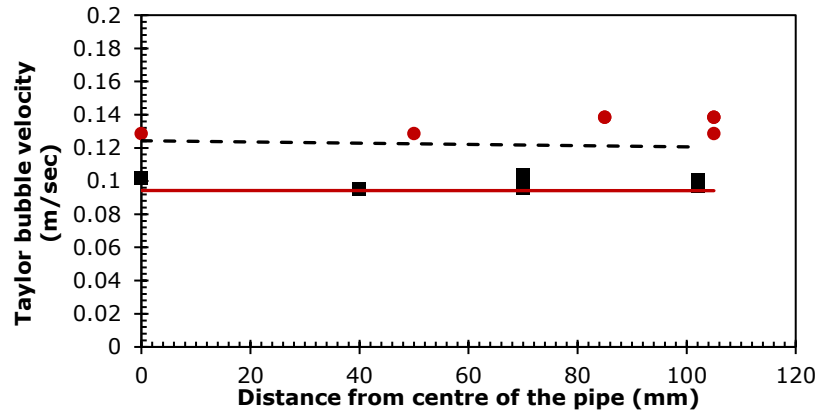


Figure 6.5: Comparing experimental and theoretical Taylor bubble velocity in high viscosity oils using single gas injection points and different positions in two different diameter columns. ■ from ECT, --- calculated from Viana et al. (2003), D=240mm, 0.04m/s, ● from ECT, — calculated from Viana et al. (2003), D=290mm, 0.03m/s.

The standard error, which was calculated from the standard deviation and the square root of the number of the repeated values, was determined for the Taylor bubble velocity in the 290 mm diameter column. The maximum value determined for the standard error for the bubble velocity was 0.75%, see Figure 6.6.

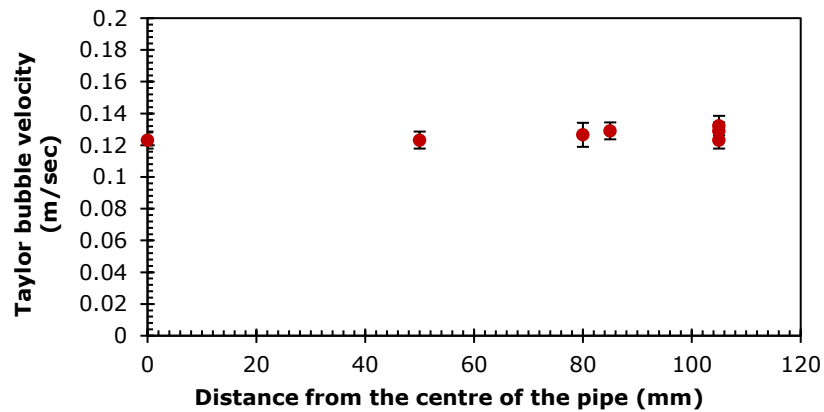


Figure 6.6: Taylor bubble velocity in 330 Pa.s Silicone oil and 290mm diameter column. Error bars represent standard errors.

Figure 6.7 illustrates the experimental and the calculated Taylor bubbles and liquid slugs lengths using an approach by Khatib and Richardson (1984). The length of Taylor bubble was calculated using the threshold of void fraction by relating the time of Taylor bubble, obtained from the time series, with the structure velocity obtained from the cross correlation of the void fraction

signals from the ECT planes. In general, the analytical results show no considerable change with moving the gas injection points, while the experimental results seem to change. The difference in the results between the two methods belongs to the methods of the determination of Taylor and liquid slug lengths as explained earlier in Chapter 5, Section 5.4.1. Liquid slug decreased significantly by 0.7 m in the 240 mm diameter column with moving the nozzle toward the pipe wall, whereas it increased only by 0.1 m in the 290 mm diameter column.

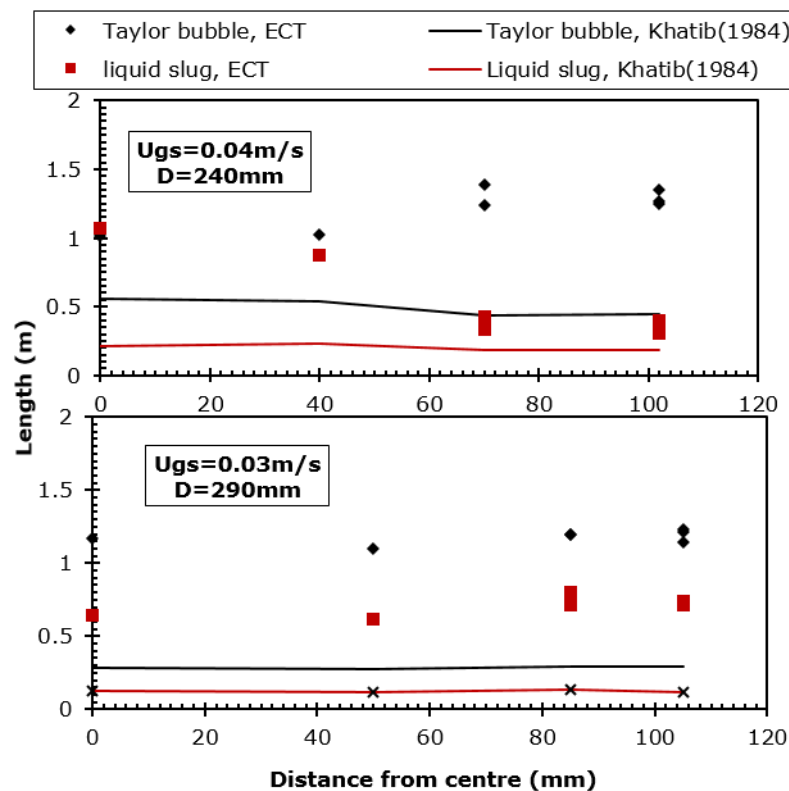


Figure 6.7: Experimental and analytical lengths of Taylor bubble and liquid slug using single nozzle and different positions for two different pipe diameters.

Figure 6.8 compares Taylor bubble lengths in 240 and 290mm diameter columns showing the error bars for the 290 mm column. The length of Taylor bubble appeared to be more affected by changing the position of the gas injection points in the smaller diameter column. Taylor bubble length increased significantly by 0.3 m by moving the gas injecting point by 102 mm from the centre of the smaller diameter column. While it increased

slightly by only 0.04 m in the larger diameter column by moving the gas injection point by 105 mm from the centre of the column.

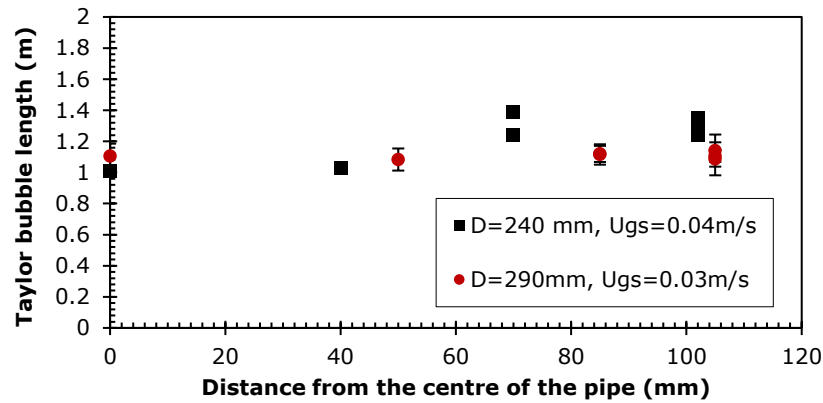


Figure 6.8: Taylor bubble length in high viscosity liquids and two different column diameters using single gas injection point and different positions. The maximum standard error is 10.6%.

Figure 6.9 shows the effect of changing the positions of the gas injection point on Taylor bubble frequency in two different diameter columns. In general, the frequency of Taylor bubbles is more affected by moving the gas injection toward the pipe wall in the smaller diameter column.

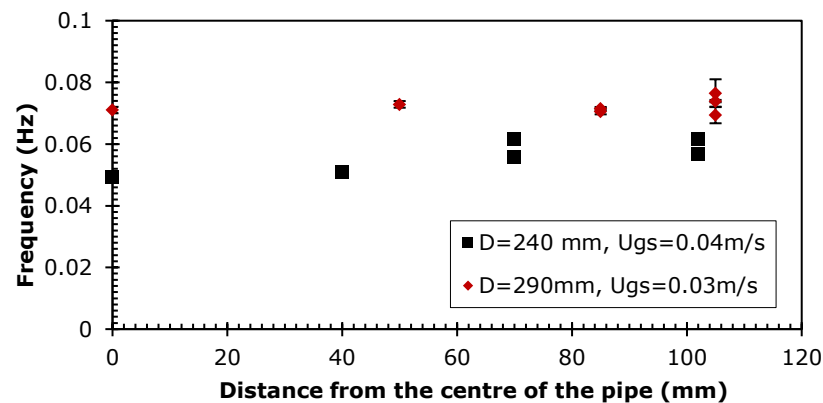


Figure 6.9: Frequency of Taylor bubble rising in two different pipe diameters using single injection points and different positions. The error bars represent the standard error (0.03-0.45) %.

It increased slightly by 0.01 Hz in the 240 mm diameter column, while no significant change was seen in the larger diameter column. On the other hand, the frequency in the 290 mm diameter column seemed higher by about 0.02 Hz for most of the gas injection positions corresponding to the

lower bubble lengths.

Another parameter, which also does not show any significant variation with change in the injection position for both columns, is the film thickness as displayed in Figure 6.10. Comparing film thickness over pipe diameter in the two columns shows an approximately 2% standard error which can be considered reasonable. The standard errors for the film thickness in the 290 mm diameter column were determined as well. It showed values of 0.06% and 0.1% minimum and maximum errors respectively.

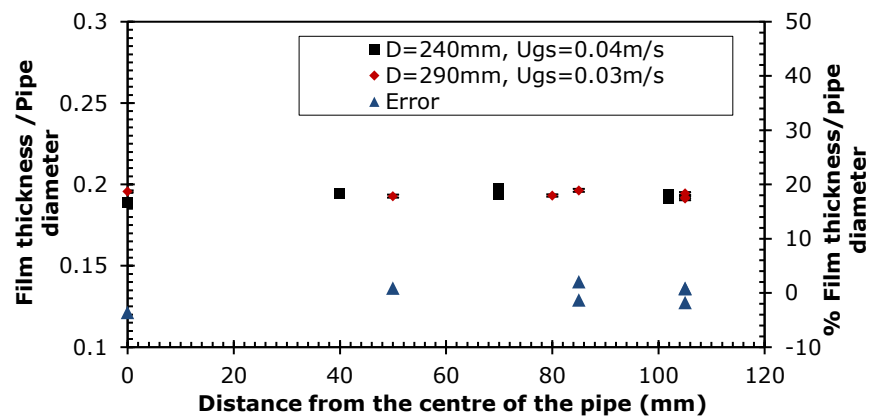


Figure 6.10: Liquid film thickness to the pipe diameter and the percentage of the error for both 240 and 290 mm diameter columns.

6.1.2 Number of Injection Points

In this set of experiments, two types of gas injection arrangements were used to study the effect of changing the number of air nozzles on the flow structure. First, air was injected with different gas superficial velocities using one injection point at the centre of the air inlet section. Second, air was also injected through 5 injection points distributed equally at the bottom of the columns. These experiments were conducted for both columns, using different ranges of gas superficial velocities. The values of the gas superficial velocities and more details about the experiments can be found in Section

3.4.4.2 in Chapter 3.

In general, two flow patterns were observed in both columns, for the range of flow rates presented, bubbly and slug flow. Time series of void fraction obtained from ECT can show the difference in the behaviour of the flow when using single or a number of injection points. Figure 6.11 and Figure 6.12 illustrate the time series of void fraction for air flow in 360 and 330 Pa.s silicone oil by using one and five nozzles in both the 240 and 290 mm diameter column respectively.

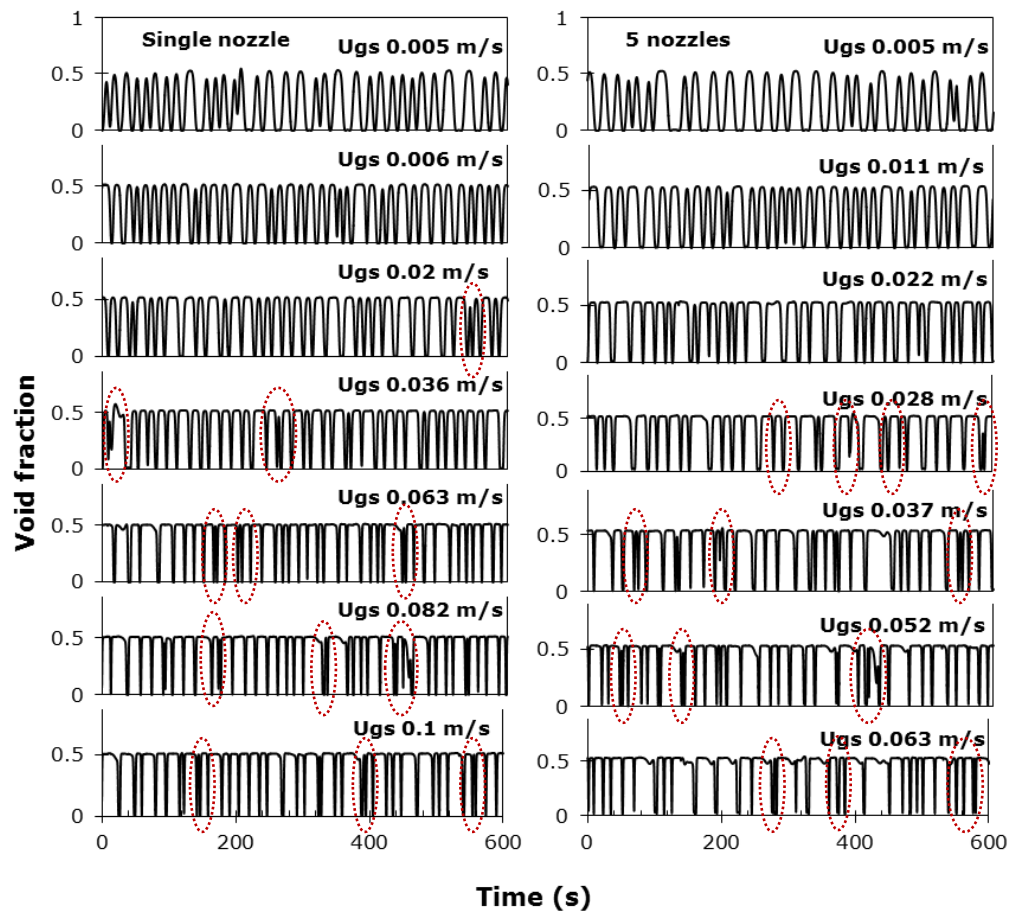


Figure 6.11: Comparing the Time series of void fraction of air flow in 360 Pa.s Silicone oil for 1 and 5 gas injection points in 240 mm diameter column the marked areas represents bubble coalescence.

Starting with the smaller diameter column, bubbly flow was observed at gas superficial velocities of 0.005 and 0.006 m/s when using a single gas nozzle, whereas it was observed at 0.005 and 0.011 m/s for the 5 gas nozzles. Slug

flow appeared at gas superficial velocities of 0.02 and 0.022 m/s for the single and 5 nozzles respectively.

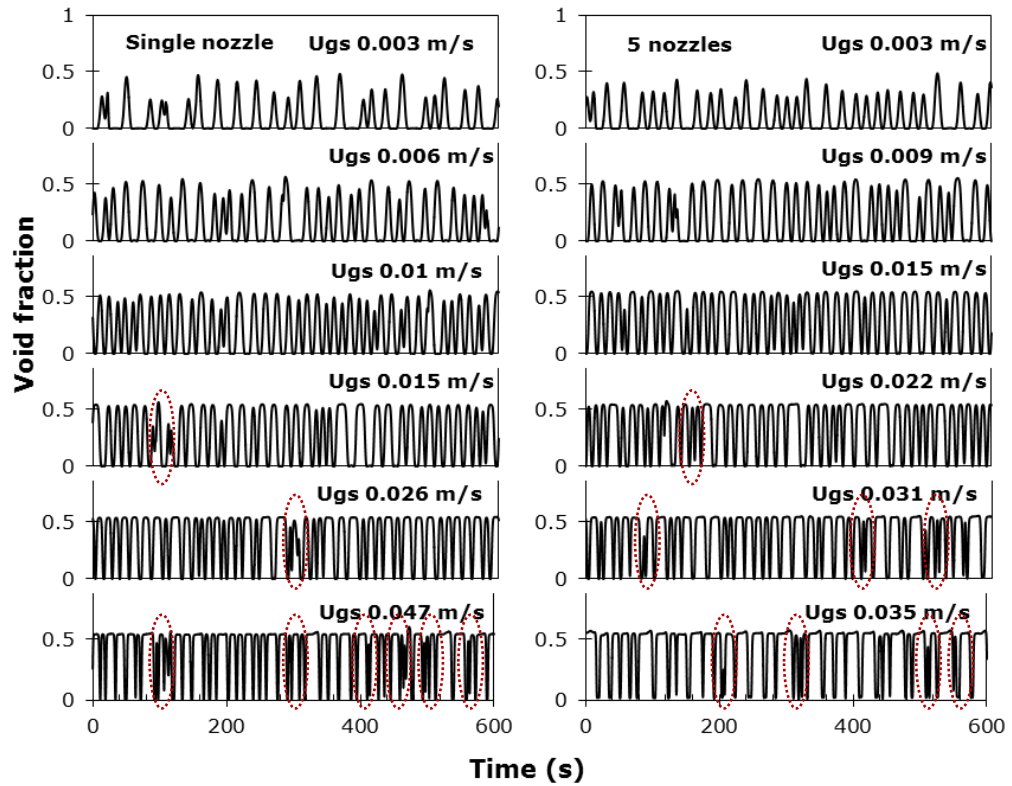


Figure 6.12: Comparing the Time series of void fraction of air flow in 330 Pa.s Silicone oil for single gas injection point with 5 injections in 290 mm diameter column the marked areas represents bubble coalescence.

In the 290 mm column, bubbly flow is observed at 0.003 and 0.006 m/s gas superficial velocities when using a single nozzle, while from 0.003 to 0.015 m/s in case of using 5 gas nozzles. This difference can be seen clearly from the photos of the larger diameter column showing a similar gas superficial velocity of air, but different flow patterns, see Figure 6.13. For example, at a gas flow rate of 30 L/min which corresponds to gas superficial velocities of 0.01 and 0.009 m/s in the single and 5 gas nozzles respectively. Slug flow appears with a long Taylor bubble in case of using one gas nozzle, while for the same flow rate bubbly flow appears with large spherical-ellipsoidal bubbles when using 5 nozzles. Similarly, in 0.015 m/s gas superficial velocity the flow is bubbly when using 5 gas nozzles to inject air inside the column, whereas it is slug flow when one air injection point was used.

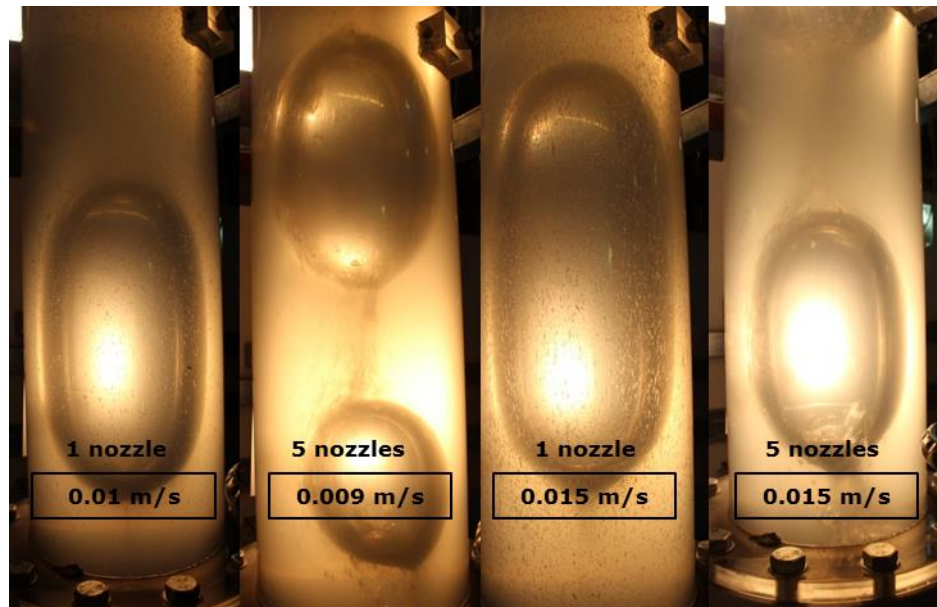


Figure 6.13: Comparing the large bubbles in two different gas inlet arrangements and two gas superficial velocities in 290 mm diameter column using 330 Pa.s Silicone oil.

First, comparing the time series of void fraction according to the number of gas injection points. In Figure 6.11, which shows the flow structure in the 240 mm diameter column, void fraction shows no significant effect by changing the number of gas injection points. The flow appears more structured at the bubbly flow region for both injection arrangements. However, at the slug flow region, the bubbles become gradually longer when using one nozzle with increasing gas flow rates, while in the case of 5 nozzles, they are much longer due to increasing the coalescence between the bubbles. This can be seen clearly when comparing with Figure 6.16 which shows the length of large bubbles for both arrangements and columns. The length of the large bubbles increases gradually from 0.36 to 2.67 m in the case of the single gas nozzle, while it increases significantly from 0.4 to 4 m when using 5 nozzles for the same range of gas flow rates. At the same time, the liquid slug is slightly longer in the case of using 5 gas nozzles and the bubble frequency less in this case.

On the other hand, the flow in the 290 mm diameter column seems to be similar to that of the smaller diameter column. The flow is more structured

at low gas flow rates, and then, with increasing the gas flow rate, the flow structure becomes less organized due to increasing the coalescence between the bubbles. The bubble coalescence is shown in Figure 6.12. The values of void fraction are at their minimum at the first gas superficial velocities (0.0029 and 0.003) m/s for both nozzle arrangements. Then they increase slightly and remain almost constant for the whole range of gas flow rates. There is no considerable change in bubble and liquid slug lengths with changing the number of air injection points, while bubble frequency is slightly higher when using 5 nozzles.

Since the range of gas superficial velocity is higher in the smaller diameter pipe, then the comparison of the time series of void fraction will consider 0.063 and 0.052 m/s as a maximum. Bubbly flow in the larger diameter column appears with lower void fraction, bubble length and frequency in addition to longer liquid slug length. However, in slug regime, and for the same range of gas superficial velocities, the bubble looks slightly longer in the smaller pipe with less coalescence. See Figure 6.11.

Figure 6.14 shows the change in mean void fraction in both columns with changing the number of the air injection points. In general, mean void fraction increases with increasing gas flow rates as expected. Mean void fraction is slightly higher when using 5 gas injection points in the smaller diameter column, while there is no significant change in the larger diameter column by changing the number of gas nozzles. However, void fraction in the smaller diameter column is higher than the larger diameter column, particularly in the lower flow rates. It is equal to 0.23 and 0.22 for the 5 and single air nozzles respectively in the 240 mm diameter column, while equal to 0.1 and 0.07 for the 5 and single nozzles respectively in the larger diameter column at the lowest gas flow rates. The reason for a higher void

fraction in the smaller diameter column, compared to the larger diameter column at the same gas flow rates, is because of the diameter of the bubbles. The ratio of the bubble to pipe diameter in the smaller diameter column is higher comparing to the 290 mm diameter pipe at the same gas flow rates.

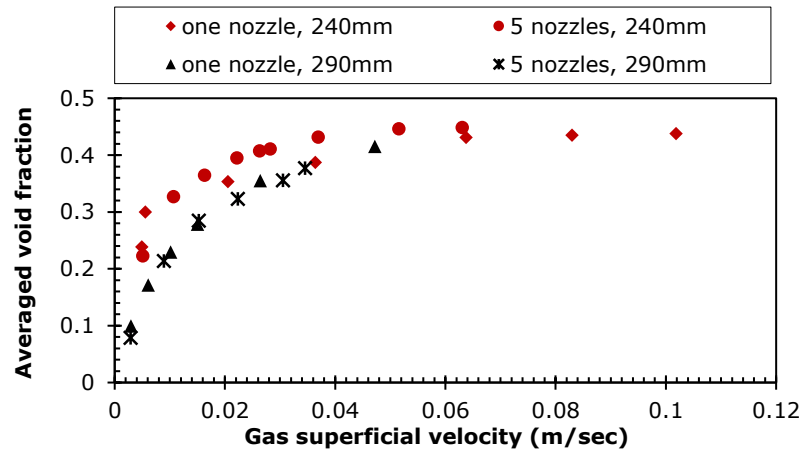


Figure 6.14: Comparing the effect of gas superficial velocity on the averaged void fraction for single nozzle at the centre of the gas injection section and five nozzles distributed equally at the injection section at the bottom of the column in viscous oils for both 240 and 290 mm diameter columns.

Unlike mean void fraction, the velocity of large bubble shows a slight difference in values between the two injection arrangements in the smaller diameter column, while the values in the larger diameter column are approximately the same. Figure 6.15 compares the velocity of the large bubbles in two different diameter pipe, using two different arrangements for the gas injection (single and 5 injections). The bubble velocity is slightly higher when using 5 air injection points. For example, at a gas superficial velocity of 0.063 m/s, bubble velocity is equal to 0.25 m/s in the 5 gas nozzles, whereas it is 0.18 m/s in case of injecting air with a single nozzle to the smaller diameter column. At the same time, the velocity of the bubbles seems not to be affected by the pipe diameter.

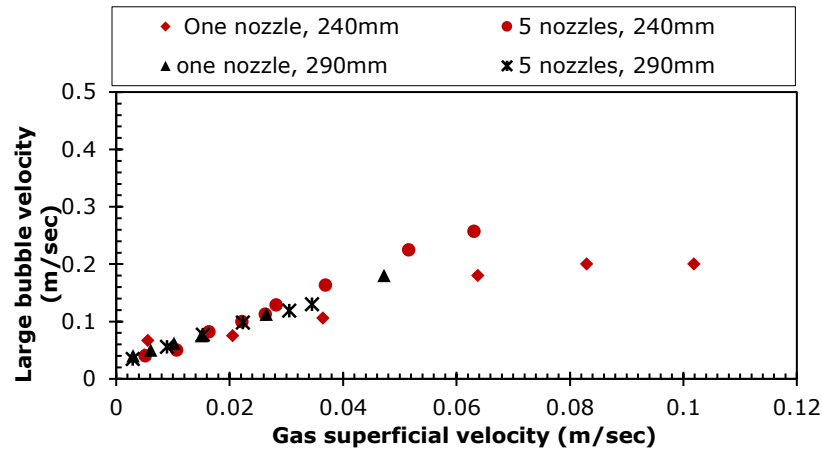


Figure 6.15: Large bubbles velocities in stagnant 300Pa.s Silicone oil using single and five injection points for both 240 and 290 mm diameter columns.

The length of large bubbles and liquid slug show similar behaviour for both columns and gas injection arrangements. They show no significant change in values of the large bubble and liquid slug lengths when changing the number of gas nozzles in the 290 mm diameter column, while the values are slightly different in the smaller diameter column. First, in the smaller column, the length of large bubbles increases significantly from 0.4 to 4 m in the case of 5 air nozzles. However, it increases steadily from 0.36 to 2.67 m when using single gas injection point in the same column for the same range of gas flow rates. Then, comparing the values show that the bubble lengths are approximately higher if using 5 gas injection points. Nevertheless, in the 290 mm diameter column, the bubble lengths increase from 0.3 to 1 m in both nozzle arrangements. Comparing bubble lengths in both columns show similar trend and values except when using 5 air injection points, the values are higher, see Figure 6.16. This might be due to the effect of the pipe wall which is seen more in the smaller diameter column and 5 gas injection points. This can be compared with the experiments of the location of the gas injection points. The length of the bubbles increases slightly when moving the gas injection points towards the pipe wall in both columns. However, this change in the bubble length is seen to be higher in the smaller diameter column due to the wall effect, see Figure 6.8. In the case of 5 gas injection

points, one point at the centre and 4 points distributed equally near the pipe wall were used to inject the gas. Then, not only the number of the gas injection points affects the length of the bubbles, but also the pipe wall which is seen clearly in Section 6.1.1.

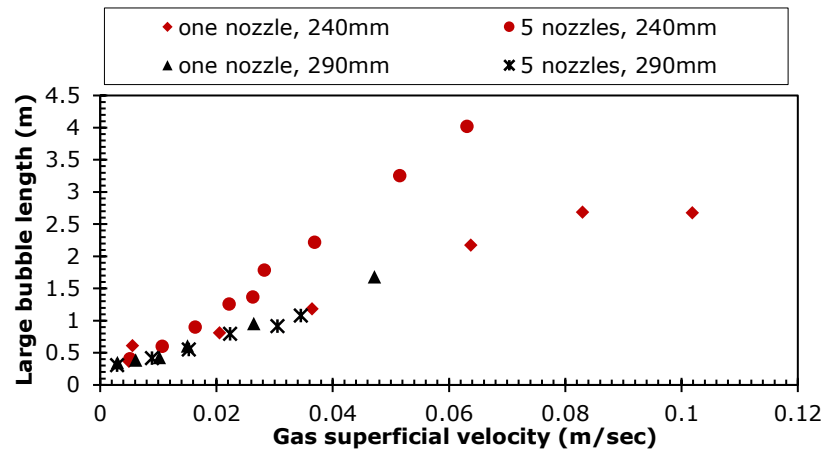


Figure 6.16: The effect of gas superficial velocity on large bubble length in viscous oils in 240 and 290mm diameter columns using single and five nozzles for air injection.

Likewise, the liquid slug length is slightly higher when using 5 gas injection points. It increases from 0.48 m to reach 1 m at a gas superficial velocity of 0.052 m/s, while the values are slightly lower when using single gas injection point (0.3-0.7) m. In the larger diameter column, the liquid slug is not showing a significant difference between the two arrangements except at a gas superficial velocity of 0.003 m/s, a single nozzle is higher by 0.3 m, (Figure 6.17).

Figure 6.18 shows the effect of changing the number of gas injection points on the frequency of large bubbles in high viscosity liquids by using different gas flow rates and different pipe diameters. In general, changing pipe diameter does not show a significant effect on the values of bubble frequency. At the same time, in the smaller diameter column, the bubble frequency seems slightly higher when using single gas injection (0.056 to 0.059 Hz), while it is between 0.04-0.05 Hz when using 5 gas injections. In

the 290 mm diameter column, bubble frequency increases by 0.03 Hz for both gas injection arrangements.

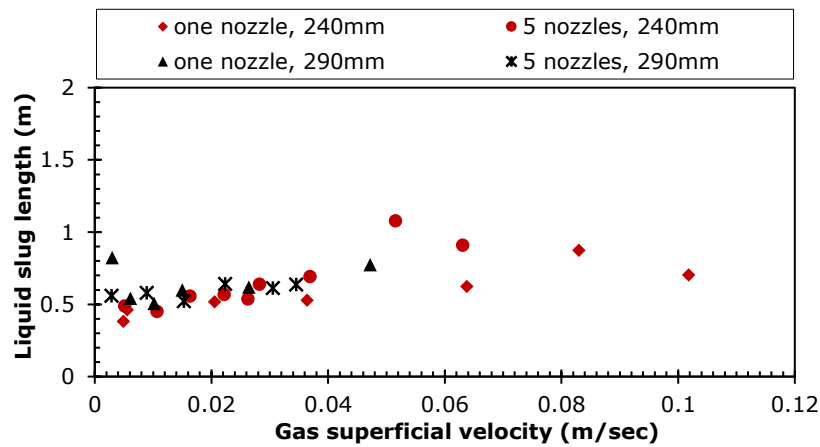


Figure 6.17: Liquid slug length in viscous liquids using different numbers of air injection points and different pipe diameters.

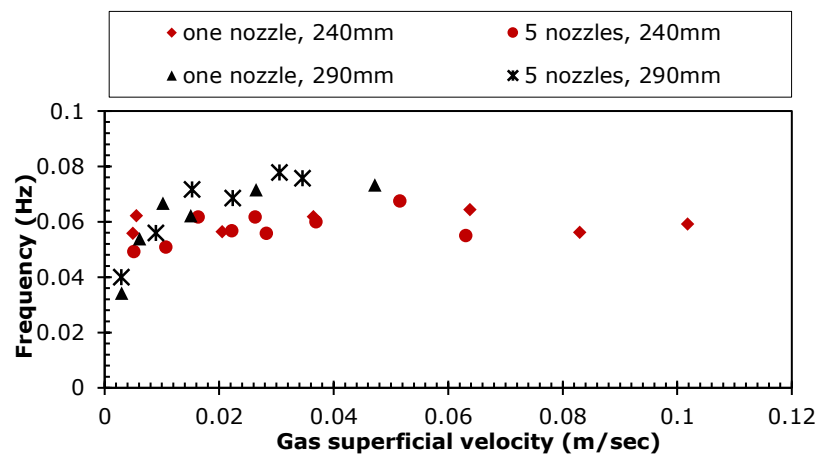


Figure 6.18: The frequency of the large bubbles in high viscosity liquids using a different number of gas injections and two large-scale pipes.

Finally, the film thickness around the large bubbles, which is calculated from mean void fractions, is illustrated in Figure 6.19 for both columns. Generally, the values of film thickness decrease with increasing gas flow rates in both columns and gas injection arrangements. This is due to an increase in the size of bubbles with increasing gas flow rates. At the same time, film thickness seems to be larger in the larger diameter column for the same range of gas flow rates. It is higher by 0.04 m at the lower gas superficial velocity and by 0.01 m at around 0.05 m/s in both gas injection arrangements.

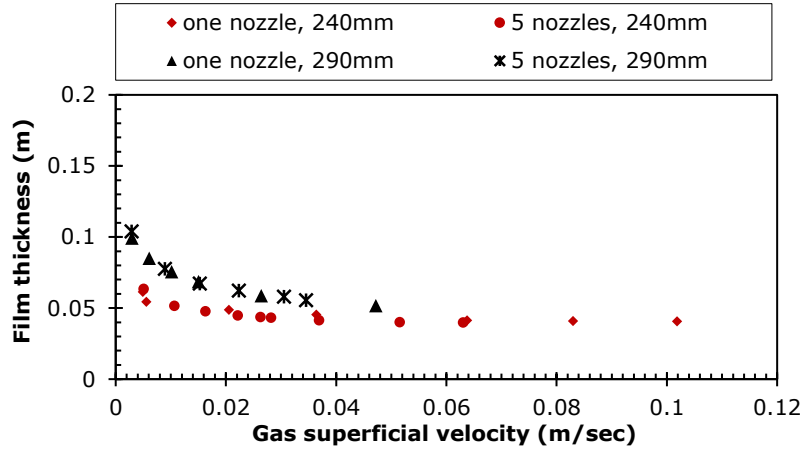


Figure 6.19: Comparing film thickness around the large bubble in high viscosity liquids using two different gas injection arrangements in two different diameter columns.

Comparing the effects of changing the number of the gas injection points, no significant change is observed in both columns. Film thickness decreases from 0.06 to 0.04 m and 0.06-0.039 m when using single and 5 air injection points respectively in the 240 mm diameter column. Similarly, it decreases from 0.1 to 0.05 m in both cases in the larger diameter column.

6.1.3 Distance between the Injection Points

In this section, the results show the effect of the space between the gas injection points on the structure of the flow in high viscosity liquids. Two gas injection points were used in both columns and the distance between the injection points was changed for each run. In total, 3 runs of 3 different gas flow rates and 3 different spacing between the gas injection points were used in the experiments in both columns. The gas superficial velocities in these experiments were 0.005, 0.01, and 0.1 m/s in the smaller diameter column and a spacing of 80, 140, and 204 mm was used for each gas superficial velocity. In the 290 mm diameter column 0.006, 0.016 and 0.1 m/s, gas superficial velocities were used with a spacing of 100, 170 and 210 mm for each gas superficial velocity. In general, in both columns, mean void fraction, bubble velocity, film thickness, and bubble frequency do not show any

significant variation in values. In contrast, the Taylor bubble and liquid slug lengths are seen to change.

Figure 6.20 shows the averaged void fraction and the film thickness of air flow in stagnant 360 and 330 Pa.s Silicone oils and 240 and 290 mm diameter columns respectively. The results report unfavourable effect of increasing distance between the 2 gas injection points on the previous parameters in both columns and at the existing range of flow rates.

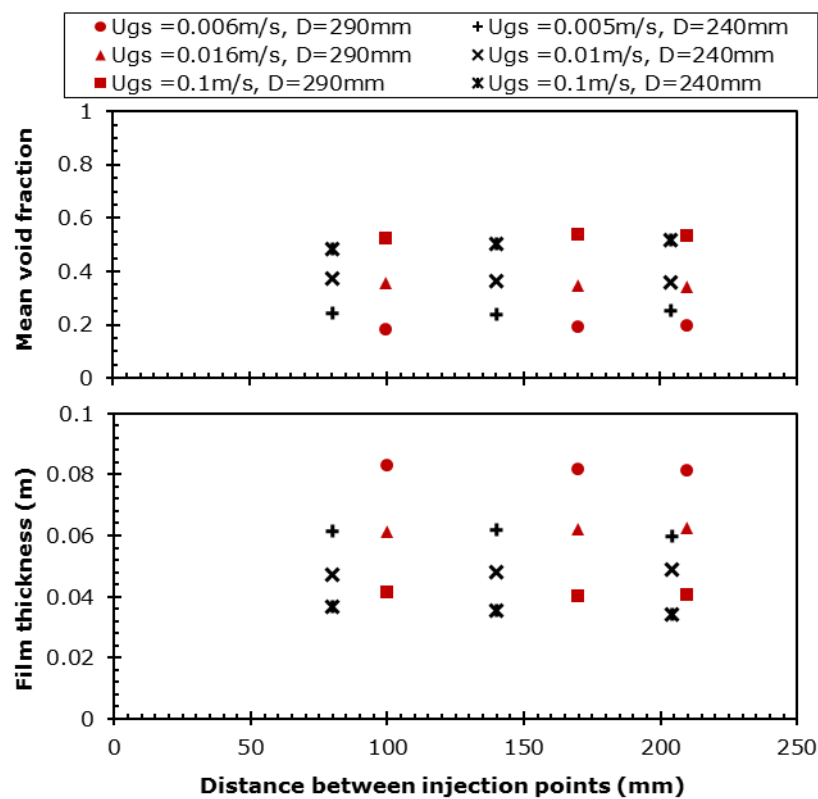


Figure 6.20: Comparing mean void fraction and film thickness in high viscosity oils using different gas superficial velocities and different spacing between the air injection points in two large diameter columns.

However, increasing the distance between both gas injection points seems to affect the length of the bubbles and liquid slug in both columns and bubble velocity and frequency in the smaller diameter column. Bubble length increases significantly by 1.2 m by increasing the spacing between the 2 gas injection points by 124 mm at the higher gas flow rate in the smaller diameter column. At the same time, it increases by 23 mm and decreases

by 99 mm at 0.005 and 0.01 m/s respectively in the same column.

On the other hand, bubble length increases slightly at gas superficial velocities of a 0.005 and 0.01 m/s by 21 and 12 mm respectively. It also increases significantly at 0.1 m/s gas superficial velocities by 1.2 m in the same column by increasing the distance between the gas injection points by 110 mm. Therefore, increasing the distance between both gas injection points does not show a clear trend for the bubble length in the smaller diameter column. While bubble length decreases with increasing spacing between both gas injection points in the larger diameter column, see Figure 6.21.

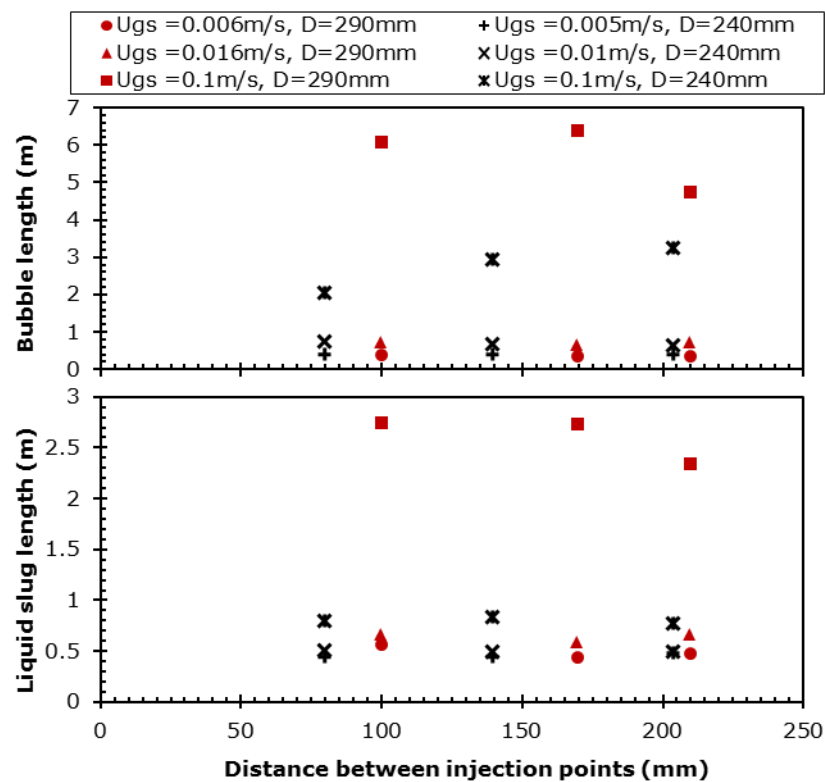


Figure 6.21: Comparing bubble and liquid slug lengths in high viscosity oils using different gas superficial velocities and different spacing between the gas injection points in two large diameter columns.

Likewise, liquid slug length seems to decrease with increasing spacing between the 2 gas injection points in the larger diameter column and fluctuate in the smaller diameter column. It decreases by 95, 2, and 397 mm

at gas superficial velocities of 0.006, 0.016, 0.1 m/s respectively by increasing the distance between both gas injections points by 110 mm in the larger diameter column. Whereas, it increases by 42 mm and decreases by 11.8 and 20 mm at gas superficial velocities of 0.005, 0.01, 0.1 m/s respectively by increasing the distance between the 2 gas injection point by 124 mm in the 240 mm diameter column. This behaviour for both parameters might be such due to the effect of the pipe wall as well as the difference in distance between gas injection points. Finally, bubble frequency and velocity in Figure 6.22 show different behaviour compared to the previous parameters. Bubble frequency shows the unfavourable effect of changing the spacing between both gas injection points in the smaller diameter column.

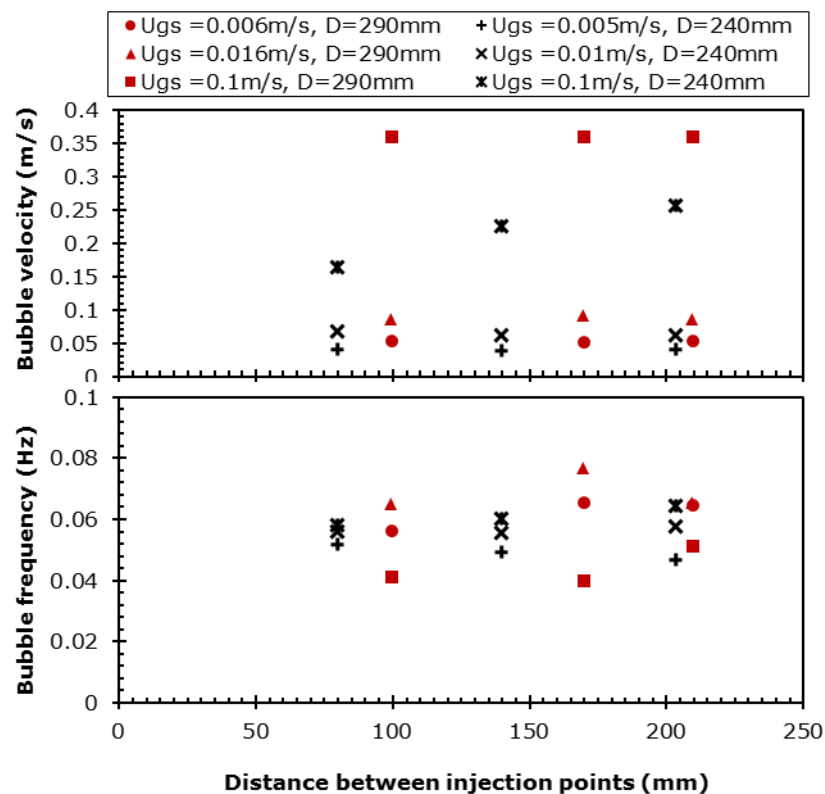


Figure 6.22: Bubble velocity and frequency in high viscosity oils using different gas superficial velocities and different spacing between the gas injection points in two large diameter columns.

While in the larger diameter column, it decreases slightly at the lowest gas flow rate and increases gradually at the higher gas flow rates. Likewise,

bubble velocity increases by 0.094 m/s by increasing the distance between both gas injection points of 124 mm in the 290 mm diameter column. Whereas, it seems almost constant in the larger diameter column with increasing distance between both nozzles.

6.2 The Effect of Gas Injection Time

The effect of the gas injection time on the structure of the flow in high viscosity oils was also studied in this work. The idea of this experiment is to inject gas with constant flow rate into the column for a certain time and collect the measurements regularly during that time. For example, in this experiment the total gas injection time was 6 hours and the data was collected every 30 minutes. The experiments were repeated using gas superficial velocities of 0.015, 0.047, and 0.178 m/s in order to compare the effect of increasing gas flow rates in different flow regimes. These gas superficial velocities represent bubbly, slug, and the transition to churn flow respectively. The average void fraction and the total liquid height in the column were measured in the 290 mm diameter column, in addition to a comparison between the two diameter columns. According to the visual observation, the amount of the small bubbles in the column increased significantly with time, even at a constant gas flow rate. Thus, the concentration of the small bubbles in viscous liquid is proportional not only to gas flow rate but also to the time of gas injection. This agrees with the results reported by Philip et al. (1990), which have been discussed in Chapter 2. Figure 6.23 shows 3 photos captured at the beginning, middle and at the end of the experiment. These photos will be explained in more detail in Chapter 8 with detailed information about the generation and dispersion of the small bubbles.

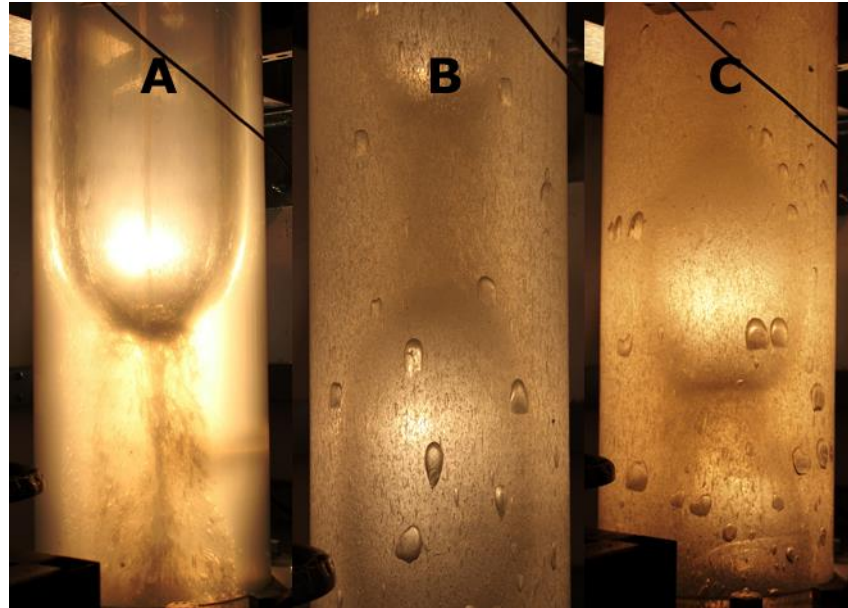


Figure 6.23: Photos of the large bubbles rising in 330 Pa.s stagnant silicone oil and 290 mm diameter column in at 3 different times with gas superficial velocity of 0.015 m/s. The photo A is obtained at the beginning of the experiment (time 0). B is after 3.5 hours. C is after 5.5 hours from the beginning of the experiment.

In general, the results seem to be approximately similar at bubbly and slug flow regime, although they seem to be quite different at the transition to churn flow. First, Figure 6.24 compares the mean void fraction for 3 different flow regimes at different times.

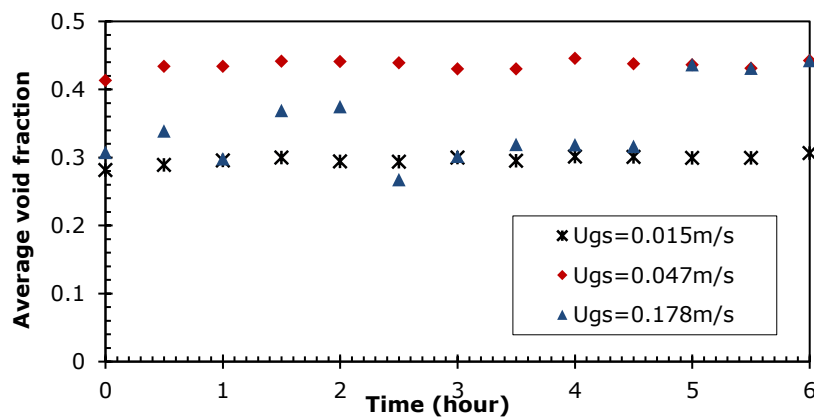


Figure 6.24: The effect of time and gas superficial velocity at the time-averaged void fraction of air flowing in stagnant viscous Silicone oil in 290 mm diameter column.

In bubbly and slug flow regimes, void fraction seems to be almost constant during the 6 hours of gas injection due to the constant gas flow rate. It increases by only 0.02 and 0.03 at gas superficial velocities of 0.015 and

0.047 m/s respectively. However, void fraction seems to fluctuate at the transition to churn flow during the 6 hours due to the unstable structure of the flow. It increases by 0.07 after 2 hours, then drops by 0.11 in 30 minutes and increases again by 0.04 by the end of the 6 hours. Similarly, the total height of the oil in the column appears relatively constant at bubbly and slug flow regimes and fluctuate at the transition to churn flow, during 6 hours of air injection. The height of the oil increases by only 10, 6, and 90 cm at gas superficial velocities of 0.015 and 0.047 and 0.178 m/s respectively, as in Figure 6.25. The effect of gas injection time on the large bubbles lengths and frequency besides the time series of void fraction for two different times are shown in Figure 5.24. The slight increase in void fraction and liquid height is due to the increasing density of the small bubbles with the time of gas injection. Furthermore, the rate of increasing this density seems to be higher at higher gas flow rates due to increasing bubble coalescence besides high gas flow rate.

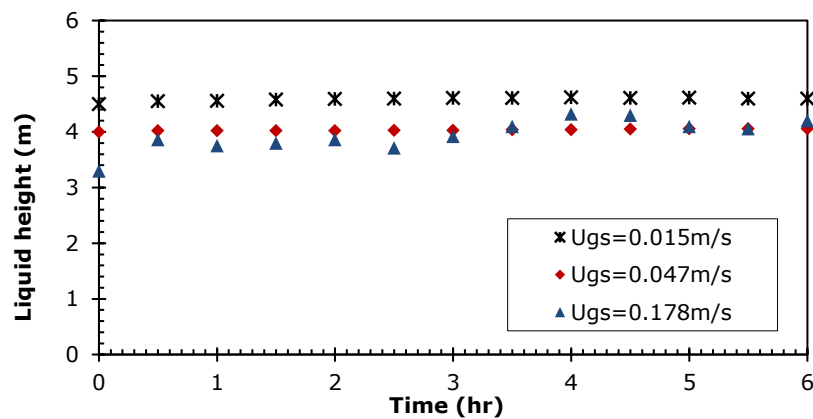


Figure 6.25: Comparing the level of the liquid height in a 290mm diameter column during 6 hours for three different gas superficial velocities.

The effect of the gas injection time on the flow structure in viscous oils was also studied in the 240 mm diameter column and compared to the previous results. The results from both columns seem to be similar in trend. See Figure 6.26 and 6.27.

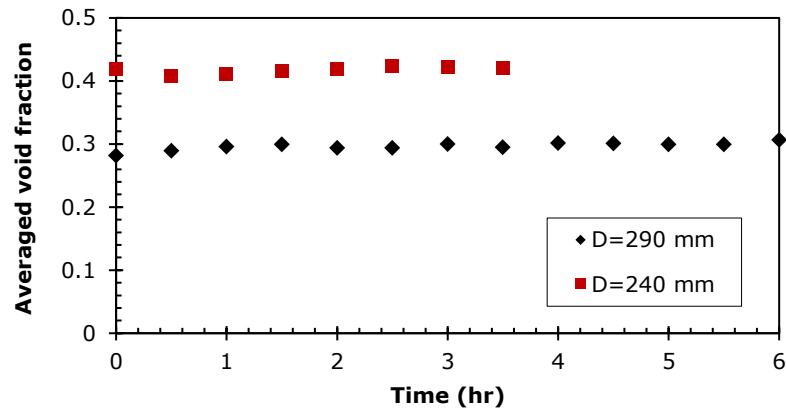


Figure 6.26: Comparing the time-averaged void fraction in 240 and 290 mm diameter columns changing with time for constant gas superficial velocities of 0.026 and 0.015 m/s respectively.

The mean void fraction in the smaller diameter column in the first figure is higher because of the difference in the pipe diameters and the gas superficial velocity, which is slightly higher by 0.011 m/s in the smaller diameter column. The rate of increasing the height of the liquid in the columns is 10 and 8 cm in the 290 and 240 mm diameter column respectively. This is due to increasing the density of the small bubbles inside the columns with increasing the time of gas injection.

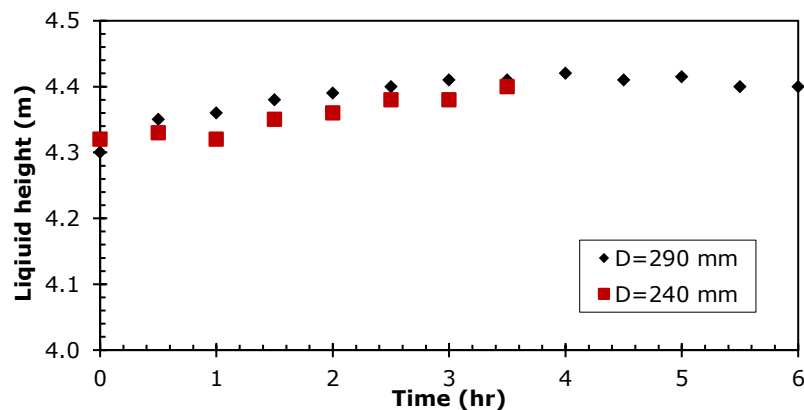


Figure 6.27: Comparing the change of the level of viscous Silicone oil during 6 hours in 240 and 290 mm diameter columns at 0.026 and 0.015 m/s gas superficial velocities respectively.

6.3 Summary

The effect of the methods of gas injection on the structure of the flow in 360 and 330 Pa.s Silicone oil in 240 and 290 mm columns respectively can be summarised as:

9. Studying the effect of the **position of the air injection** on the structure of the flow in the stagnant high viscosity liquids using two different large diameter columns has shown: first, In terms of changing the injection positions, the mean void fraction, structure velocity, frequency and film thickness did not show any significant variation in values. In contrast, the Taylor bubble and liquid slug lengths were seen to change. Second, the structure of the flow in the smaller diameter column (240 mm) seemed to be more affected by changing the position of the nozzle than the larger diameter column. Finally, the flow in the smaller diameter column appeared more structured than the larger one for the same gas superficial velocity. The first one contained homogenous bubbles with a lower amount of coalescence. However, the bubble velocity was slightly higher in the larger diameter column for the same gas superficial velocity compared to the smaller diameter column.
10. The study on the effect of changing the **number of the gas injection points** and pipe diameters has shown that: In the 240 mm diameter column, mean void fraction, bubble velocity and length, and liquid slug length were higher when using 5 gas injection points, while the frequency appeared to be higher when using single nozzle. However, in 290 mm diameter column, no significant changes appeared except in frequency. It was higher when 5 gas nozzles were used to inject air into the larger diameter column. On the other hand, changing the number of gas injection points seemed not to have an effect on the film thickness

around the large bubbles. Regarding the effect of pipe diameter, it seemed not to affect the bubble velocity and frequency. However, liquid slug length and film thickness were higher in the 290 mm diameter column. At the same time, mean void fraction and bubble velocity, when using a single gas nozzle, were higher in the smaller diameter column.

11. Studying the effect of the **distance between the air injection** points on the structure of the flow in high viscosity oils revealed that: Mean void fraction, bubble velocity, film thickness, and bubble frequency do not show any significant variation in values. However, the Taylor bubble and liquid slug lengths are seen to change.
12. The study on the effect of the **gas injection time** shows that the amount of the small bubbles in the column increases significantly with time, even at a constant gas flow rate. Thus, the density of the small bubbles in viscous liquid is not only proportional to gas flow rate but also to the time of gas injection. The mean void fraction increases slightly after a few hours of gas injection due to increasing the amount of the small bubbles in the column.

Chapter 7

Churn Flow in Viscous Liquids

7.1 Introduction

This chapter studies churn flow in high viscosity liquids and large diameter pipes. The characteristics of churn flow in stagnant high viscosity liquids have been studied for a wide range of gas flow rates using two different column diameters. The ECT was used to reconstruct the cross-sectional distribution of the gas-liquid phases inside the test section. Videos and photos were recorded and compared with the results from the ECT to obtain a full picture of the flow structure. The results from the time series of void fraction were used to determine the mean void fraction, PDF, structure velocity, large bubble length and frequency, and liquid film thickness. The film thickness and the wave velocities were also determined as they are considered a good indication of the two-phase flow pattern. And finally, the behaviour of the top surface of the column was also studied and compared with the time series of void fraction for the corresponding gas flow rates.

7.2 Churn Flow and the Transition to Churn

Churn flow which is an important intermediate regime occurs in between slug and annular flow patterns. The main characteristics of this flow pattern in vertical pipes are the existence of a downward falling liquid film, and very large waves travelling upwards. As eluded to earlier in the literature review, the majority of research studies on gas-liquid churn flow has been carried out mainly using water or liquids of low viscosities. Therefore, information about churn flow in very high viscosity liquids is still limited. In this work,

the transition to churn flow has been investigated together with the characteristics of churn flow regime in high viscosity oils and large diameter pipes. Two column geometries have been used one of 240 mm and the other is 290 mm internal diameter. The columns were initially filled with stagnant Silicone oil of viscosities 360 and 330 Pa.s respectively. In the 240mm column, gas superficial velocity was varied from 0.127 to 0.314 m/s to capture the transition to churn flow. While in the 290mm column the transition to churn flow region occurred in the gas superficial velocity range of 0.108 to 0.243 m/s. Churn flow was observed in gas superficial velocities ranging from 0.276 to 0.566 m/s. Gas is introduced through 25 different injection points located at the base of both columns. Two ECT sensors were used to collect information about the cross-sectional phase distribution inside the test section. Details of the pressure inside the columns were also evaluated for a wide range of gas flow rates in order to obtain more information about the two-phase flow regime. The pressure measurements are shown in Chapter 4. A high-resolution camera was employed to capture the flow structure in the 290 mm diameter column. The data of the time series of void fraction obtained from the ECT was processed and compared with literature. The next sections of this chapter present churn flow in high viscosity liquids in detail.

7.2.1 Time Series and Visual Observations

The transition to churn flow has been studied in more focus in the 290 mm diameter column in the range of gas superficial velocity of 0.108-0.566 m/s. A stagnant Silicone oil of 330 Pa.s viscosity was initially loaded on the column. Videos and photos were recorded and compared with the data of the time series of void fraction; also images were reconstructed from the recorded data using Plot3D software. It is worth mentioning that the

videos/photos of the column analysed in this chapter have been taken in the first few minutes following the gas injection. Each of the photos in this section has been taken on separate days to allow small bubbles trapped in the liquid to rise and therefore translucent oil is formed. For example, when comparing the photo of 0.108 m/s in Figure 4.1 with the photo of the same gas superficial velocity in Figure 7.2, the photos look different. The reason behind that is the small bubbles which cover the entire pipe wall and prevent the clear observation of the flow structure inside the column. Therefore, the photos were taken before the concentration of small bubbles increases and cloud cover the pipe wall. However, the photo in Figure 4.1 shows the general case encountered during experiments where small bubbles are present. In general, the characteristics of the flow of high viscosity oils in the transition to churn and also within churn flow regime differ compared with low viscosity liquids which have been reviewed in Chapter 2.

For a better understanding of the transition to churn flow, it is worth considering the preceding flow regime (slug flow). Figure 7.1 illustrates slug flow in 330 Pa.s Silicone oil and 290 mm diameter column. The corresponding gas superficial velocities for the figure are 0.066 and 0.088 m/s which represent the end of slug flow region. At this stage, the structure of the flow is dominated by Taylor bubbles about 3.2 to 3.8 m long. The images also show evidence of cross-section bridging by the liquid created from the accumulated falling film from the long Taylor bubbles. These liquid bridges are marked in red in the void fraction time series and labelled (B) in Figure 7.1. Therefore, at the end of slug flow region, the flow can be divided into two main structures. First, long Taylor bubbles dominating the gas structures ((A) in the time series and the photo). Second, a number of liquid bridges with lower void fraction and high frequency at the higher gas flow rates, the red marked regions in the time series and B in the photo.

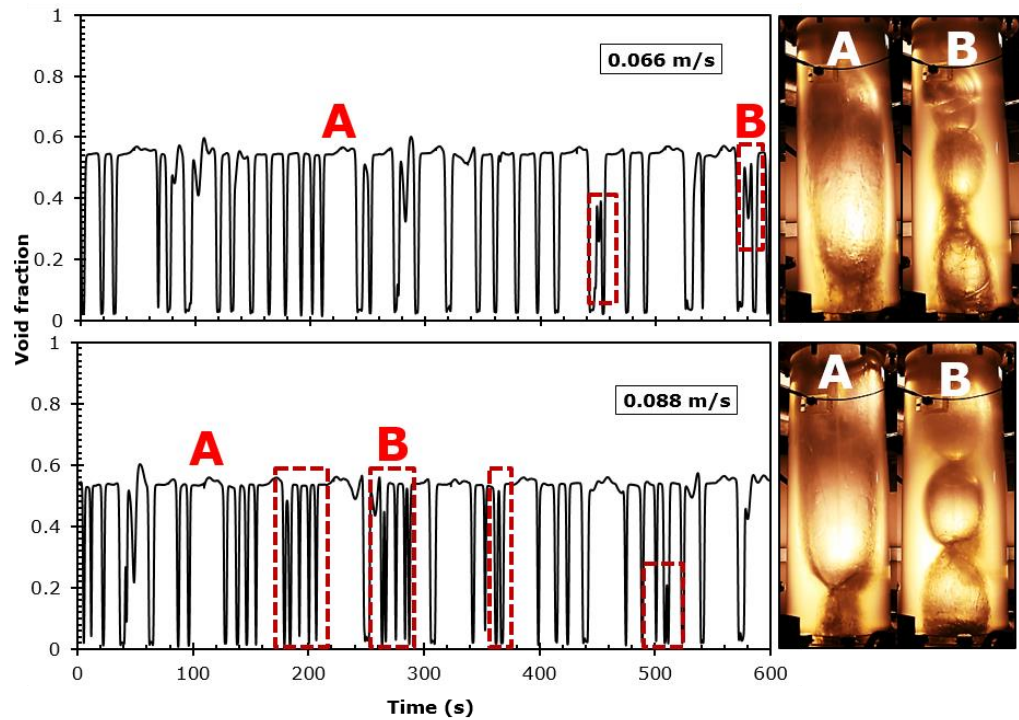


Figure 7.1: Time series of void fraction and photos showing the structure of slug flow in 330 Pa.s Silicone oil and 290 mm diameter column. A is Taylor bubble, B is the churn areas in both photos and the time series data.

With increasing gas flow rate, regions with high frequency activity appear. This extends systematically as the gas flow rate is further increased. These regions are shown as (B) on the time series of void fraction in the figures in this section also depicted on the photos of the column. Figure 7.2 displays the time series of void fraction and the equivalent photos which show the flow structure at the selected gas flow rates. The transition to churn flow regime starts when the length and the frequency of the regions of the high frequency activity, which appeared at the end of slug flow, increases. These regions of high frequency activities were observed as the liquid bridges flow up and down. The void fraction of these regions is lower than that in Taylor bubbles. The source of liquid bridging is believed to be from the drained falling film from the Taylor bubbles. With increasing gas flow rate the length of Taylor bubbles increases producing larger volumes of the falling film which lead to increase the length of the high frequency liquid bridges region. The high frequency liquid bridges region will be considered churn. This will be

explained in more detail in Section 7.2.4.

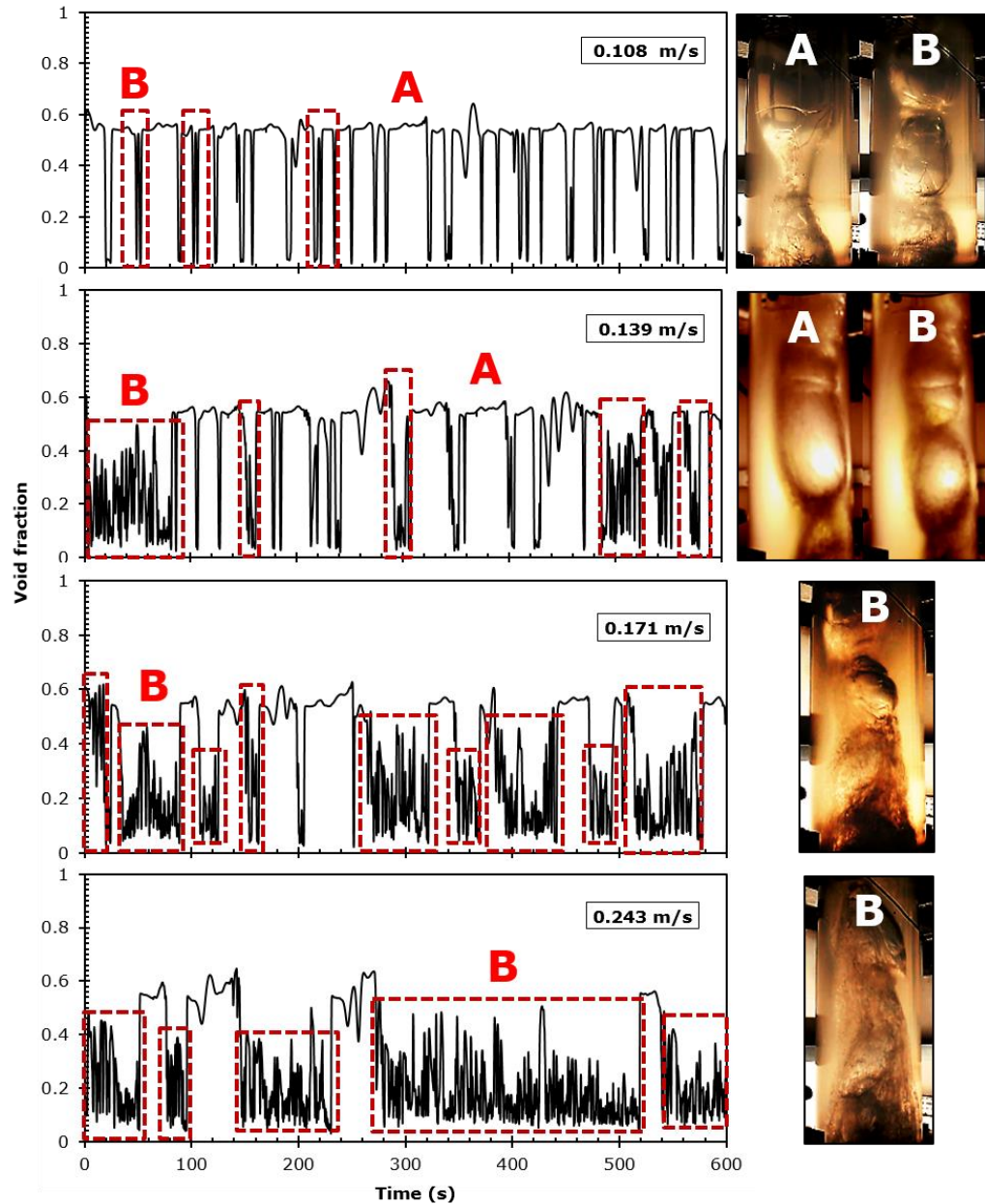


Figure 7.2: Time series of void fraction and photos of transition to churn flow in 330 Pa.s Silicone oil and 290 mm diameter column at different gas superficial velocities. A is the large bubbles, B is the churn areas in both photos and the time series data.

On the other hand, as the length of Taylor bubbles (A) is increased their frequency decreases. The maximum amount of the coalescence between Taylor bubbles occurs in this flow regime (transition to churn). The length of Taylor bubbles increases dramatically with increasing gas input till it reaches the total height of the column. This will be explained later in this chapter.

According to the visual observations, Taylor bubbles become deformed when the gas flow rate is increased. This is due to increasing the rate of the coalescence and therefore increasing the length of the bubbles. The Top and/or the bottom of the bubbles lose the round shape. The liquid slugs between the bubbles eventually disappear and get replaced by the churn areas (B). The churn areas are connected to the deformed Taylor bubble by a small neck (See (A) in Figure 7.2). The deformed long Taylor bubbles will be referred to as long bubbles at this flow regime. Coalescence between numbers of bubbles appears from about 100-150s and 230-280s at 0.243 m/s gas superficial velocity in the time series data. The photos of the column in (B) show the developing of the churn areas. With increasing gas flow rate, the structure of these regions becomes more complicated and deformed. By the end of this flow regime, only a few number of long bubbles appear.

The Silicone oil which used in this work showed the behaviour of Newtonian liquids during the viscosity bench test, See Section 3.2.4. The viscosity of the oils was measured by applying a certain shear rates using a rotating spindle. During the test, the viscosity of the oil remained constant with time. The viscosity of the oil seen to change with changing temperature, not with the amounts of applied shear force, Papanastasiou et al. (1999).

The time series data from the ECT besides the videos obtained from the top section and different positions at the column provide local information about the flow behaviour in the column. Therefore, an overall picture of the gas-high viscosity liquid behaviour in the column is presented in a cartoon drawing in Figure 7.3. This schematic drawing shows the overall flow structure in the column at the transition to churn flow regime. The letters A-E correspond to the behaviour of the flow structure at different times for constant gas superficial velocity (0.205 m/s.). The numbers 1-5 and the

dashed lines correspond to the oscillation of the top surface of the column which is explained in detail in Section 7.2.5 in this chapter. The liquid film was seen to travel upward and downward along the column, this is also explained in detail in Section 7.2.4.

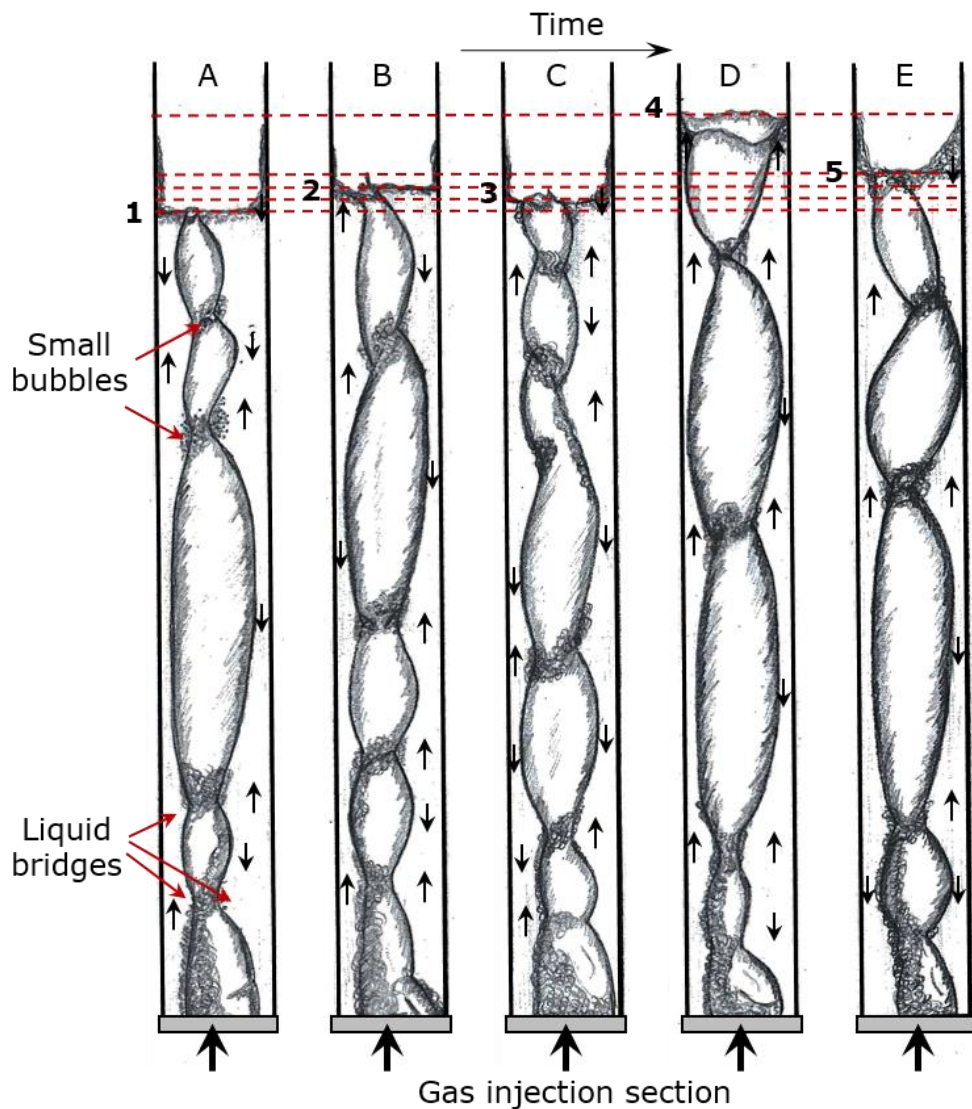


Figure 7.3: Schematic drawing showing the transition to churn flow pattern in high viscosity liquids and large diameter columns. The arrows in the figure correspond to the direction of the liquid flow, the numbers at the top section of the column correspond to the liquid levels in the column. The gas superficial velocity is 0.205 m/s.

The results of the pressure gradient measurements shown Figure 4.8 in chapter 4 came in agreement in trend with the work done by Owen (1986). Owen presented the change of pressure gradient data against dimensionless

gas velocity for a number of liquid flows in vertical pipes. According to his results, the pressure gradient values showed a sudden drop at the transition to churn flow due to the high frequency wave activity and friction. A similar increase in the pressure gradient values was observed in this work at the transition to churn flow. However, in Owen's work, the values of the pressure gradient decreased with increasing the gas flow rate at constant liquid flow. The decrease of interaction between the gas and the liquid is a reason of the decrease in the pressure gradient at the selected gas flow rate. Then by increasing the gas flow rate, the pressure gradients exhibits an increase again. In churn to annular flow region, the pressure gradient showed a different behaviour in the high viscosity oil. In churn flow region as shown in Figure 4.8, the pressure gradient increased gradually by 2 KPa/m from 0.276 to 0.322 m/s this is due to increase in frequency of the churn areas where the interaction is very intense between the oil and the air. Pressure gradient remained almost constant until a gas superficial velocity of 0.517 m/s was reached, where the transition to churn flow took place. The flow structure in churn flow regime was found more stable in terms of structure velocity, mean void fraction, and mean film thickness. This is explained in detail in the next sections in this chapter. The deviation in the pressure gradient behaviour can be attributed to the substantial difference in the liquid viscosities, surface tension, and the pipe diameters. Owen (1986) used air-water in a 32 mm diameter vertical pipe. Wallis (1962) found that for starting churn flow, entrainment of the droplets forms from the large liquid fractions which travel as large flooding waves along the column. This mechanism of churn flow formation is not applicable for viscosity liquids such as the oils used in the present work. No droplets were observed in the 330 and 360 Pa.s Silicon oils due to the very high viscosity.

At a gas superficial velocity of 0.276 m/s, the flow structure changed

significantly. The long bubbles disappeared and the churn areas dominated the flow structure. According to the time series data in figure 4.5, the photo shown in Figure 7.4, and the schematic drawing in Figure 7.5, churn flow consists of two regions. First an open gas core with a thick liquid film with the possibility of liquid bridging the cross-section creating series of short bubbles (columns D and E in Figure 7.5 and the photos in Figure 7.4). This gas core might not exist at the centre of the column.

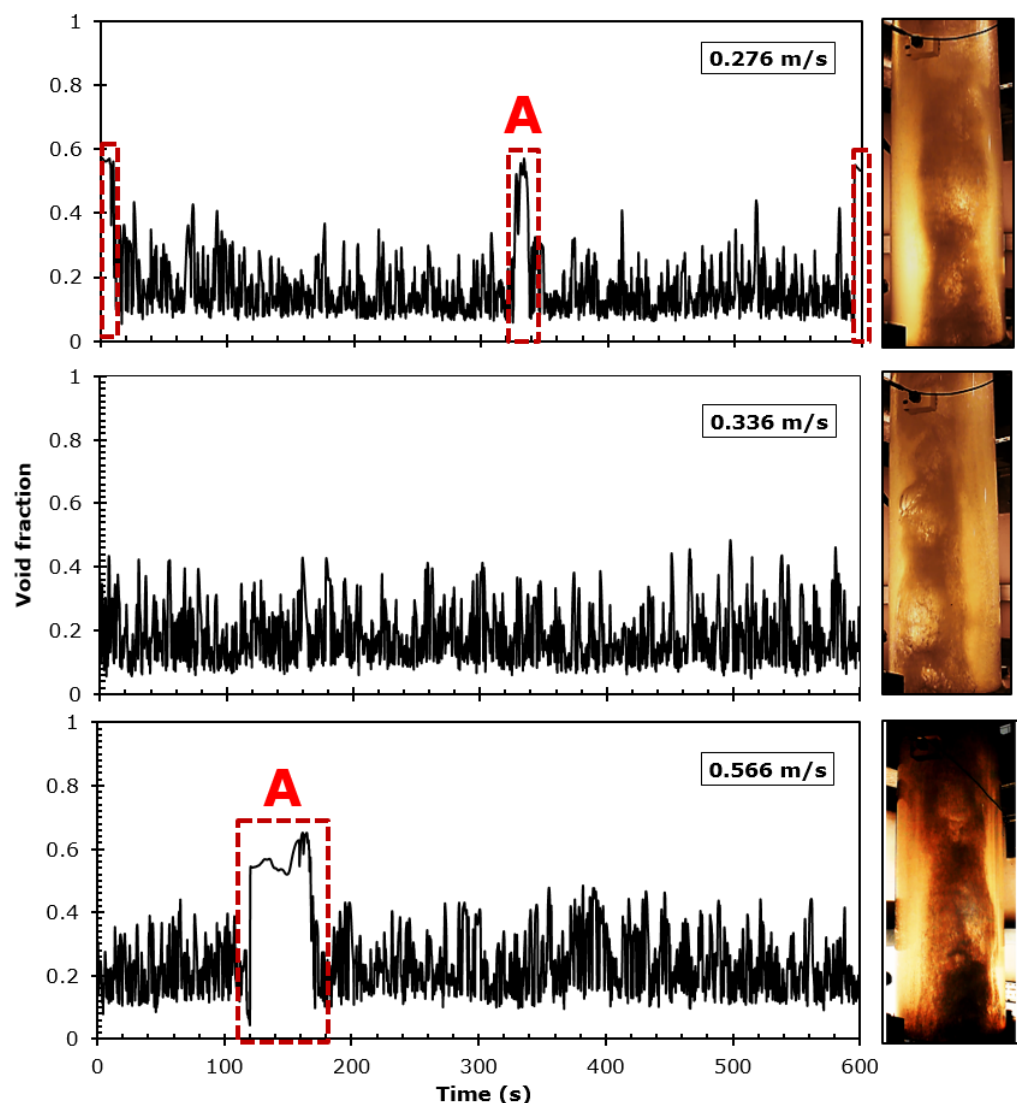


Figure 7.4: Time series of void fraction and photos showing churn flow in 330 Pa.s Silicone oil and 290 mm diameter column. The marked areas correspond to the large bubble and the photos show the churn regions.

This will be shown later in this chapter. Subsequently, with increasing gas flow rate, the gas build up and push the liquid (column A in Figure 7.5)

creating a very long bubble with a falling thin film. This long bubble bursts at the top section of the column and remain open to the atmosphere. This open large core is shown in Figure 7.4 at 120-180 sec (A) in 0.566m/s gas superficial velocity and (B) in Figure 7.5. Then the falling film, which drains along the pipe wall, accumulates at the bottom of the column (column C in Figure 7.5) and creates the next churn unit. See the long bubbles at gas superficial velocities of 0.414, 0.456 and 0.566 m/s in Figure 4.5.

The cartoon drawing in Figure 7.5 shows the overall behaviour churn flow regime along the column, as it described earlier, at a constant gas flow rate and at different times. Another schematic cartoon of these two flow regimes is displayed in Figure 4.2 "The flow patterns in 330 Pa.s Silicone oil and 290 mm diameter vertical pipe" in chapter 4. Three different gas flow rates are shown in this figure with the flow patterns which have been observed in the selected range. The figure also shows the flow pattern in the bubbly and slug flow regimes.

Churn flow can also be distinguished from the transition to churn flow regime by the sound the gas flowing inside the column. At the end of slug flow and the transition to churn flow regime the sound of the bubble eruption at the top surface could be heard with every single bubble. In churn flow regime, the voice generated by the flowing fluids in the rig changes completely. The open gas core sounds like the volcanic conduit during volcanic activity. This continuous sound of the air passage through the viscous oil stops when the fraction of the drained liquid increases at the bottom of the column. At this stage, the gas builds up and pushes the drained liquid until the liquid fraction becomes very thin and bursts at the top section. Then the sound from the open core stars again. The sounds of the air passing through the high viscosity liquid which used in this work were approximately similar to the

Arenal Volcano in Costa Rica. The volcano sounds were studied by Woulff and McGetchin (1958), Richards (1963), Garcés and McNutt (1997), and Dubosclard et al. (2004).

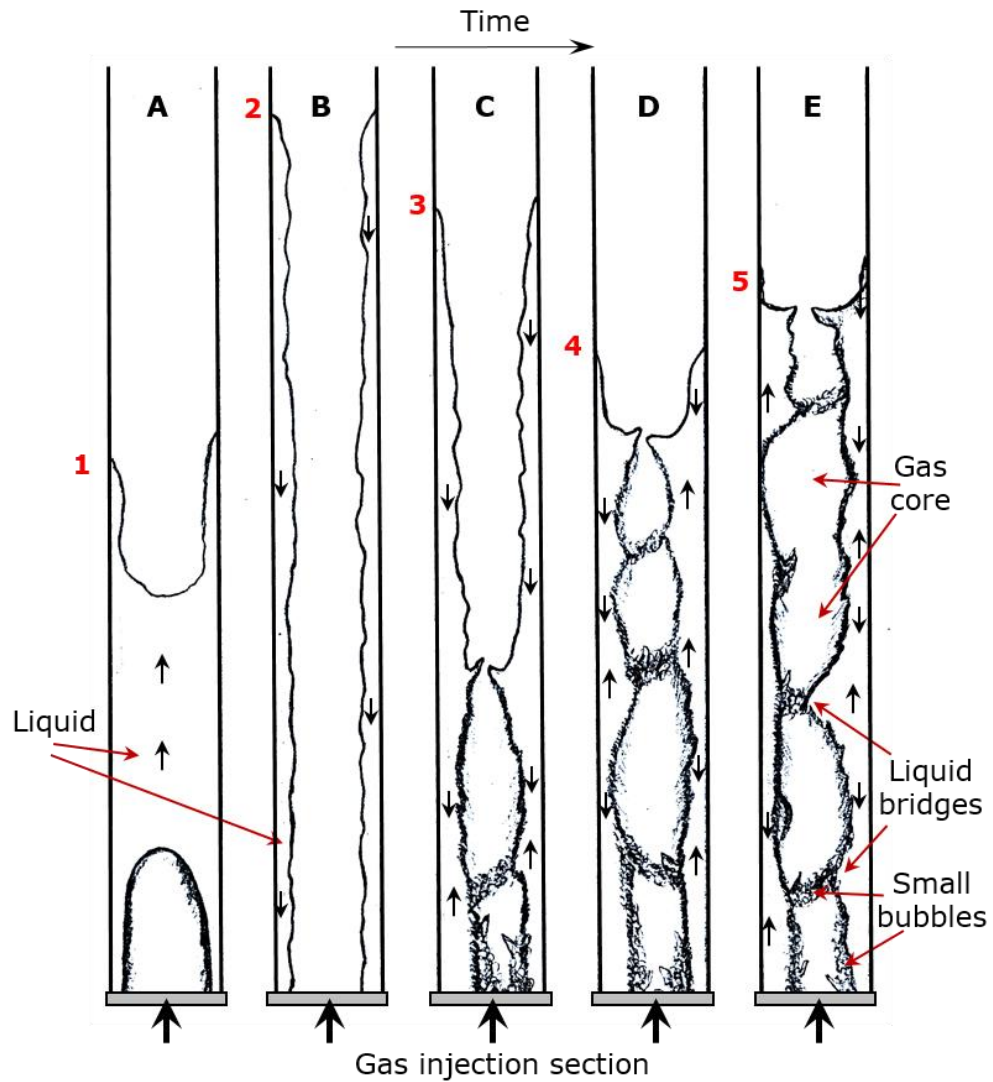


Figure 7.5: Schematic drawing showing churn flow pattern in high viscosity liquids and large diameter columns. The arrows in the figure correspond to the direction of the liquid flow, the numbers at the top section of the column correspond to the liquid levels in the column. D and E are the more common structure for this flow regime. A-C occur when the liquid accumulates at the bottom of the column and the gas build up and rise as one long bubble and carry the whole liquid up to drain again as a falling film. The gas superficial velocity is 0.566 m/s.

The percentage of churn was calculated from the time of churn and the overall flow of each gas flow rate. The time of churn can be defined as the time when the areas of the high frequency activity (churn areas) pass through the ECT planes. Figure 7.6 displays the effect of gas flow rate on the

time of churn in high viscosity liquids and large diameter pipes. In the 290 mm diameter column, the time of churn increased significantly with increasing the gas flow rate at the transition to churn regime (from 0-80%). While it decreased slightly from 96% to 91% in churn flow regime. This slight decrease in time of churn can be referred to the single long bubbles that appear at the end of this flow regime. See Figure 4.5. In the smaller diameter column, the churn time was found to be less than that in the larger diameter column at the same gas input. Churn time increased by approximately 22% over the range of gas superficial velocities at which ECT measurements were taken (0.127-0.314 m/s). Then from the data obtained using the pressure transducers for a wider range of gas flow rates, it continued increasing to about 79%.

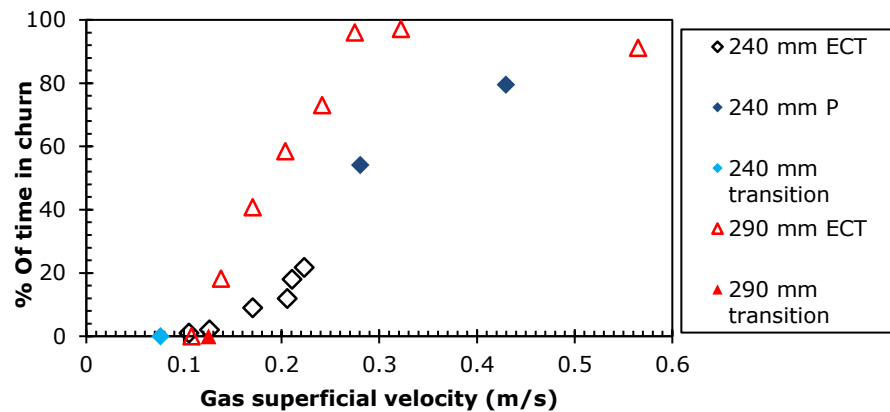


Figure 7.6: Time of churn calculated from the time series of void fraction of 240 and 290 mm diameter columns using the ECT, the pressure transducers in the smaller diameter column and the transition points are calculated from two straight lines on drift flux (u_{gs}/e_g vs u_{gs}).

Figure 7.7 compares the average void fraction in the transition to churn and churn flow regimes in the 290 mm diameter column. It also compares the values of void fraction at the transition to churn flow regime between the two columns. In general, mean void fraction decreases with increasing gas flow rate in both columns.

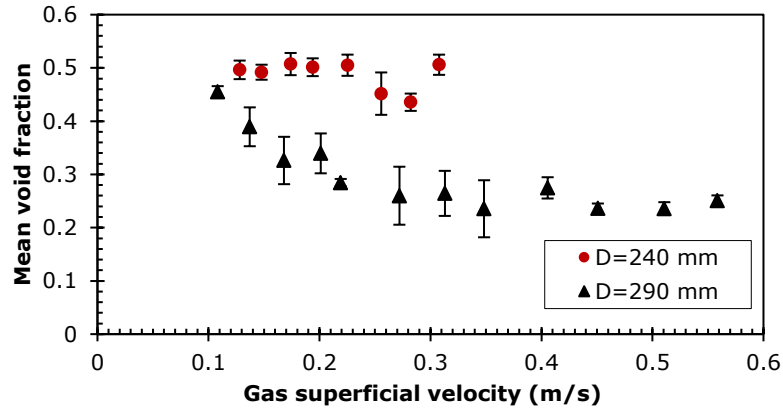


Figure 7.7: Effect of gas superficial velocity on the mean void fraction in high viscosity liquids and large diameter columns.

In the 290 mm diameter column, the mean void fraction continues to decrease in the transition to churn flow until it stabilises when churn flow is achieved. It decreases from 0.455 to 0.284 when gas superficial velocity was varied from, 0.108 to 0.243 m/s. This is due to the rise in the time of churn with increasing gas flow, as void fraction is lower at the churn sections. Then it varies between 0.259 to 0.250 from 0.276 m/s and higher gas superficial velocities (churn flow). In this flow regime, the structure is dominated by the low void fraction churn, and this explains the stability of void fraction values. Therefore, increasing gas flow rate at this flow regime does not limit the volume fraction of the gas which can pass through the viscous liquid column. This behaviour could be important for the volcanic degassing process. In this process, the rate of degassing from volcanic conduits might be controlled by the availability of the gas from the magmatic system. As a result, the rate of degassing might not be limited by the volcanic plumbing system.

On the other hand, the average void fraction showed a different behaviour in the 240 mm diameter column. Void fraction decreased slightly by 0.06 then it increased again by 0.07 at the last gas flow rate in this range. The difference in the void fraction between both columns at the transition to churn regime can be explained by comparing the data of the time series of void fraction in Figures 4.4 and 4.5. First, the transition to churn flow region

in the smaller diameter column consists of a longer bubble of higher frequency, higher coalescence rate, and lower frequency of churn regions compared to the case in the larger diameter column. This explains the higher values of void fraction which might be due to the difference in the column diameter and the viscosity. The gas volume fraction is larger in the smaller diameter column for the same gas flow rate due to the geometry. In addition, the void fraction increases with increasing liquid viscosity Philip et al. (1990). Second, the percentage of churn time increases gradually with increasing the gas flow rate in the smaller diameter column while it increases significantly in the larger diameter column. However, void fraction exhibits an increase at the highest gas flow rate in the smaller diameter column. This might be a beginning of fluctuation due to the significant increase in the coalescence between the bubbles before starting churn flow regime. The averaged void fraction data can be related to the film thickness in the next sections.

7.2.2 Probability Density Function

The Probability density Function (PDF) signature was used in combination with the time series of void fraction to identify the flow structure of churn flow in high viscosity oils. PDF is calculated from the void fraction time series data measured by the ECT. The shapes of the PDFs in these flow regimes and such viscous oils showed different trends from the characteristic shapes introduced by Costigan and Whalley (1997) in both columns, see Chapter 2. The PDF graphs for both columns are shown in Chapter 4 for the whole range of gas flow rates. This section is on the PDF profiles obtained at 4 gas flow rates from Figure 4.19 in Chapter 4 which represent transition to churn and churn flow in 290 mm diameter column.

Data from the void fraction time series where long bubbles occur was

extracted from the overall time series of void fraction and separated from that in the churn regime. The PDF was calculated for each segment separately. Figure 7.8 compares the average PDF which represents the whole flow structure with the PDFs of the churn and slug separately for two flow regimes. The first two graphs were carried out at 0.139 and 0.243 m/s gas superficial velocities which capture the transition to churn flow. While the last two PDFs at gas superficial velocities of 0.276 and 0.566 m/s show churn flow regime.

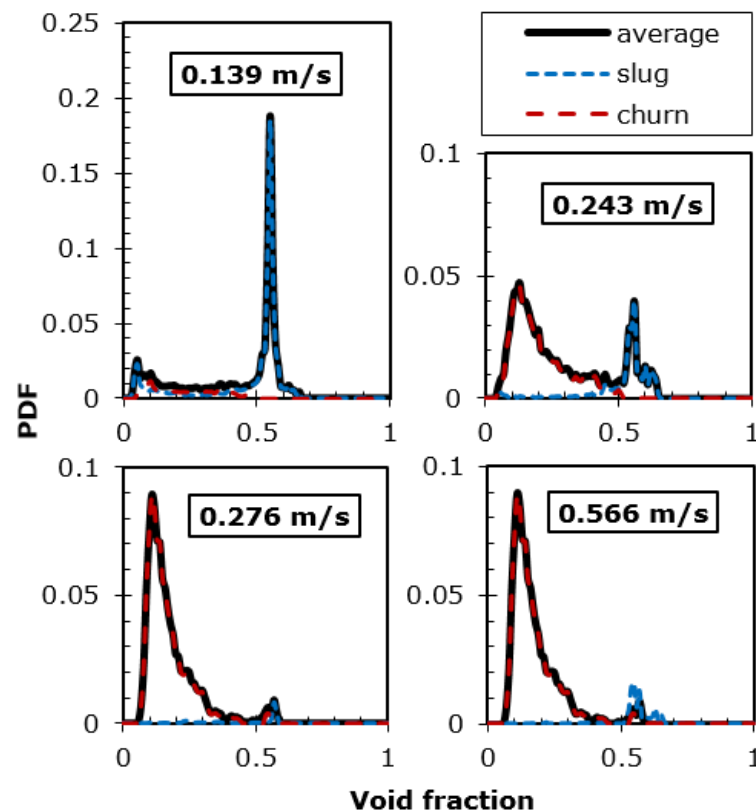


Figure 7.8: PDF for air flow in 330 Pa.s viscosity Silicone oil and 290 mm diameter column at different gas superficial velocities. The dashed lines in blue and red correspond to the slug and churn areas which were extracted from time series of void fraction of the 290 mm diameter column.

This figure can be divided into 4 parts. First, the beginning of the transition to churn flow (the lowest gas superficial velocity in the figure), when the structure contains long bubbles separated by regions of churning. At this gas flow rate, the long bubbles with high void fractions (PDF in blue) dominate the structure of the flow, while the probability of “churn” regions is still low

(PDF in red). Then, with increasing the gas flow rate, at the transition to churn flow, the frequency of churning regions increases. For example at a gas superficial velocity of 0.243 m/s the PDF of the higher void fraction in blue (the long bubbles) peaks at an approximately equal value to the PDF of the lower void fraction which is the churning regions (in red). Furthermore, at the beginning of churn flow, at gas superficial velocity of 0.276 m/s the frequency of churning regions (PDF in red) increases dramatically with a probability of appearing in a number of long bubbles. Finally, at the higher bound of gas flow rate, the Probability of presence of churning regions is almost similar to the lower gas flow rate with the existence of one long bubble, see the corresponding time series in chapter 4. However, at gas superficial velocities of 0.336 and 0.518 m/s (in the churn flow range) churn regions appear to dominate the whole flow structure with no evidence of any large bubbles, see Figures 4.5 and 4.19 in Chapter 4.

7.2.3 Structure, Motion, and Dimensions of the Flow

Structure velocity, long bubbles' length and frequency have been determined in the transition to churn and in the churn flow regimes in high viscosity oils and large diameter columns. They have been calculated from the time series data of void fraction obtained from the ECT. Structure velocity was calculated by cross-correlating the void fraction signal from both ECT planes. Figure 7.9 illustrates the effect of increasing gas flow rate on the structure velocity. In general, the structure velocity increases with increasing the gas flow rate in both columns at the transition to churn flow regime. While no considerable increase was seen in churn flow regime in the larger diameter column. It increased by 0.56 m/s in the larger diameter column and 0.33 m/s in the smaller diameter column.

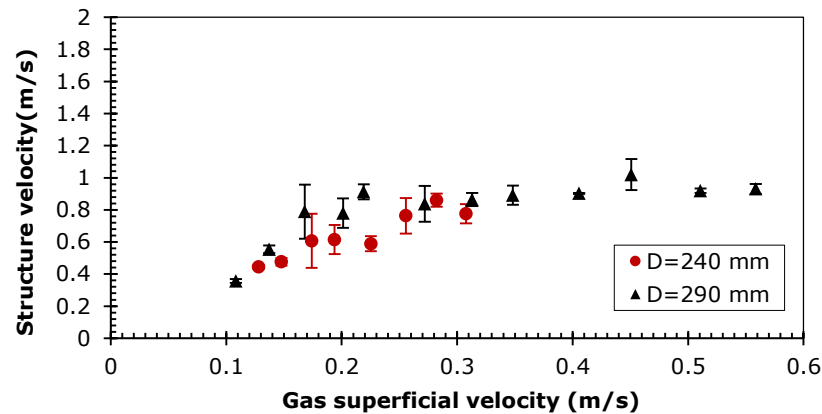


Figure 7.9: The structure velocity of gas flow in 360 and 330 Pa.s Silicone oil and 240 and 290 mm diameter respectively at the transition to churn flow and churn flow regimes.

The structure velocity is slightly lower in the smaller diameter column because the large bubbles are longer and the rate of coalescence is higher. On the other hand, the structure velocity in churn flow regime showed a slight increase by only 0.02 m/s in the 290 mm diameter column. This is due to the oscillating movement of the liquid bridges in this flow regime. The slight fluctuation in the values of the void fraction at this flow regime is caused by the single bubble that appears at some gas flow rates. The maximum standard errors calculated for the structure velocity are 16% and 9% for the 240 and 290 mm diameter columns respectively.

The experimental structure velocity was compared with a theoretical approach by Viana et al. (2003) and Guet et al. (2004) in Figure 7.10. Both data showed relative errors of 0.01 and 0.04 in the 240 and 290 mm diameter columns respectively at the transition to churn flow regime. However, in the smaller diameter column, the ECT measurement fluctuate due to the increasing coalescence rate between the bubbles. The calculated and experimental structure velocities show a big divergence in the churn flow regime (relative error of 1.3). The models used in these predictions are not applicable for churn flow regime, as slip velocity is calculated based on the velocity of large bubbles.

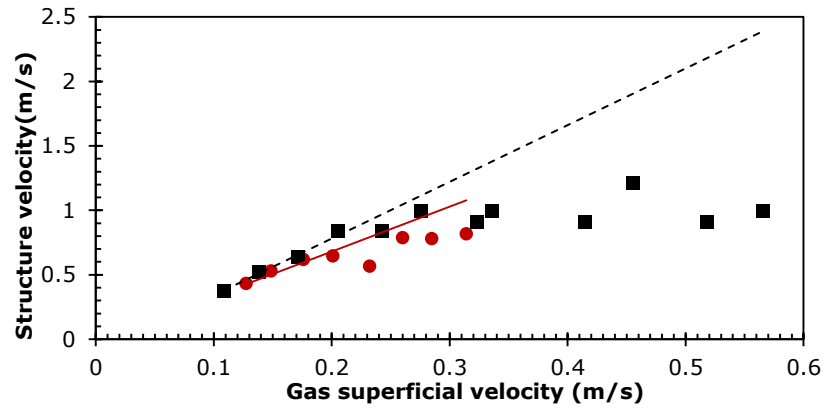


Figure 7.10: Theoretical and experimental large bubbles velocities in viscous Silicone oil for both columns. •:240mm, ■: 290mm (experimental) and — 240mm, ---290mm (from Viana et al. (2003), and Guet et al. (2004)).

The length of the large bubbles was calculated from the structure velocity and the time of the bubble passing from the ECT electrodes. The length of large bubbles flow in high viscosity liquids and two large diameter columns is displayed in Figure 7.11. In general, large bubbles' lengths increased significantly with increasing gas flow rate in the transition to churn flow region. Whereas, it decreased in the churn flow regime in the 290 mm column. Also, bubbles' lengths are higher in the smaller diameter column for the most of the gas flow rates.

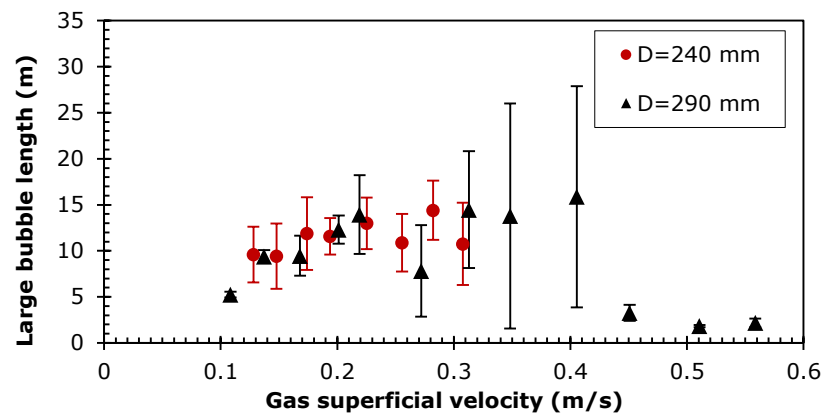


Figure 7.11: Lengths of the large bubbles in viscous oil and large diameter columns at the transition to churn flow and churn flow regimes.

In the 290 mm diameter column, the bubbles' length increases significantly from 5.2 to 13.9 m when the gas superficial velocity is increased from 0.108 to 0.243 m/s (transition to churn flow regime). While it increases gradually

from 9.5 to 14.4 m for gas superficial velocity increase from 0.127 to 0.314 m/s (churn flow). These high values of the bubble lengths are due to the high rate of coalescence between the bubbles which produces a very long bubble. At the same time, bubbles' frequency decreases due to the increasing frequency and length of the churn regions. It in turns increases due to the increasing fraction of liquid draining from the very long bubbles. The bubbles' length which appears in the figure is higher than the column's length at some gas flow rates. Some large bubbles that burst at the top section of the column in this flow regime remain as an open core for some time. Since the length of the bubble is calculated from the time of the bubble passing through the ECT sensor, therefore open core can be considered a very long bubble. This long bubble is not a Taylor bubble as it contains large waves on the falling film and does not have a rounded top and/or bottom as the Taylor bubbles that appear in slug flow.

At churn flow, the length of the bubbles fluctuates between 15.8 to 2 m for the whole range of gas flows. This is due to the complicated structure in this regime. The flow structure consists of an open gas core with liquid bridging of varying film thicknesses and a possibility of creating a very long bubble. As a result of this complicated structure, a very high percentage error is calculated in this regime. 122% standard error was calculated in churn flow in the 290 mm diameter column. While in the smaller diameter column 14% maximum standard error was calculated for the bubble length.

The large bubble frequency was also calculated from the time series of void fraction data by counting the number of bubbles over the gas injection time. Unlike bubble length, the frequency of the bubbles seems to decrease with increasing the gas flow rate in the transition to churn flow region. In the larger diameter column, the frequency decreases by 0.023 Hz in the

transition to churn flow region. This can be attributed to the increasing length of bubbles due to bubble coalescence. In churn flow regime, frequency increases by 0.113 Hz. See Figure 7.12.

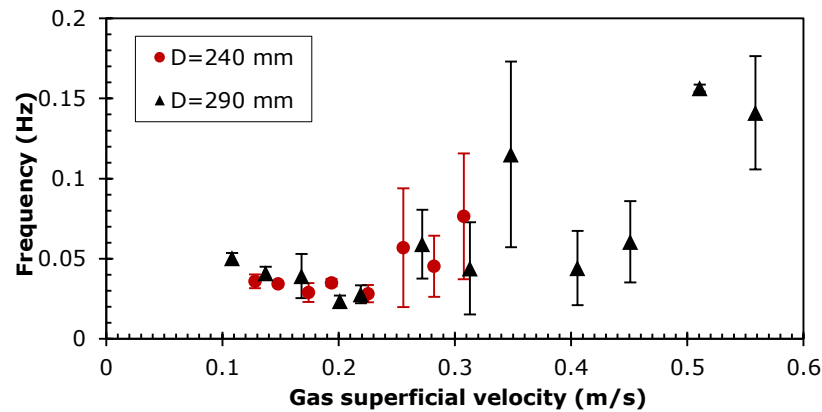


Figure 7.12: Frequency of the large bubbles at the transition to churn flow and churn flow regimes in high viscosity liquids and large diameter columns. The maximum errors are 3.9% and 5.7% for 240 and 290 mm diameter column.

Comparing frequency in this flow regime from the time series of void fraction data show a different trend. Bubble frequency increases when there is only one long bubble appearing in the void fraction signal at the same flow rate range. This might be due to the increase in the frequency of liquid bridges that behave like a series of bubbles connected to each other. Therefore, in the transition to churn flow region, the dominant frequency represents the frequency of long bubbles which decrease with increasing gas input. While in churn flow the dominant frequency represents that of short bubbles produced by liquid bridging which increases with increasing gas flow rate.

7.2.4 Film Thickness and Wave Velocity

Images of the phase distribution of air-oil and 330 Pa.s Silicone oil in 290 mm diameter were reconstructed using ECT32 and Plot3D software for the cross-sectional and the axial section of the column. The structure of flow in such viscous oils was studied by comparing the time series of void fraction, film thickness, and the reconstructed images of the phase distribution inside

the column. The average film thickness in the transition to churn and in churn flow regions was determined in the 240 and 290 mm diameter columns. It was calculated by the geometric relationship $\delta = (D / 2) (1 - \sqrt{\epsilon_g})$, where δ is the film thickness, ϵ_g is the void fraction from the ECT. The film thickness was estimated in both flow regimes using the void fraction data collected from the two ECT planes.

The reconstructed images of the cross-sectional distribution of air-Silicone oil in 290 mm diameter column are shown in Figure 7.13. In this figure, the phase distribution in the cross section from the upper plane of the ECT is displayed at different time instances. They have been generated using the ECT32 software at 0.566 m/s gas superficial velocity. The red colour corresponds to the high permittivity fluid which is the 330 Pa.s Silicone oil.

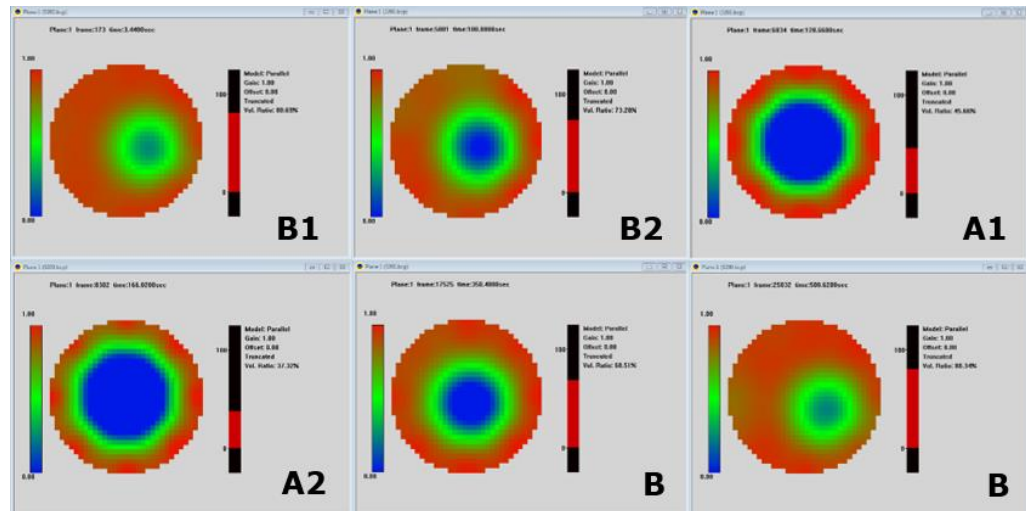


Figure 7.13: Image reconstructions of the cross-sectional distribution for the gas-330 Pa.s Silicone oil in 290 mm diameter column at 0.566 m/s gas superficial velocity using the ECT32 software. B1 and B2 represent the churn areas. A1 and A2 represent the large bubble in the column. The axial view, film thickness and void fraction of the selected area are displayed in Figure 7.14.

The blue colour corresponds to the low permittivity fluid, the air. The green and the yellow represent the interface between the two fluids. B1, B2, A1, and A2 in the next two figures in this section refer to a time or sections. For example, A1 presents the phase distribution of the large bubble at 120 s. It

also refers to the void fraction, film thickness, and the reconstructed image in Figure 7.14.

Figure 7.14 compares the reconstructed time-resolved images of the vertical section of the gas-liquid flow with the void fraction and the calculated film thickness for churn flow in the 290 mm diameter column. The Plot3D software was used to generate the reconstructed images. The three plots are at the same flow rate and the same time period. Moreover, they are corresponding to the plots in Figure 7.13.

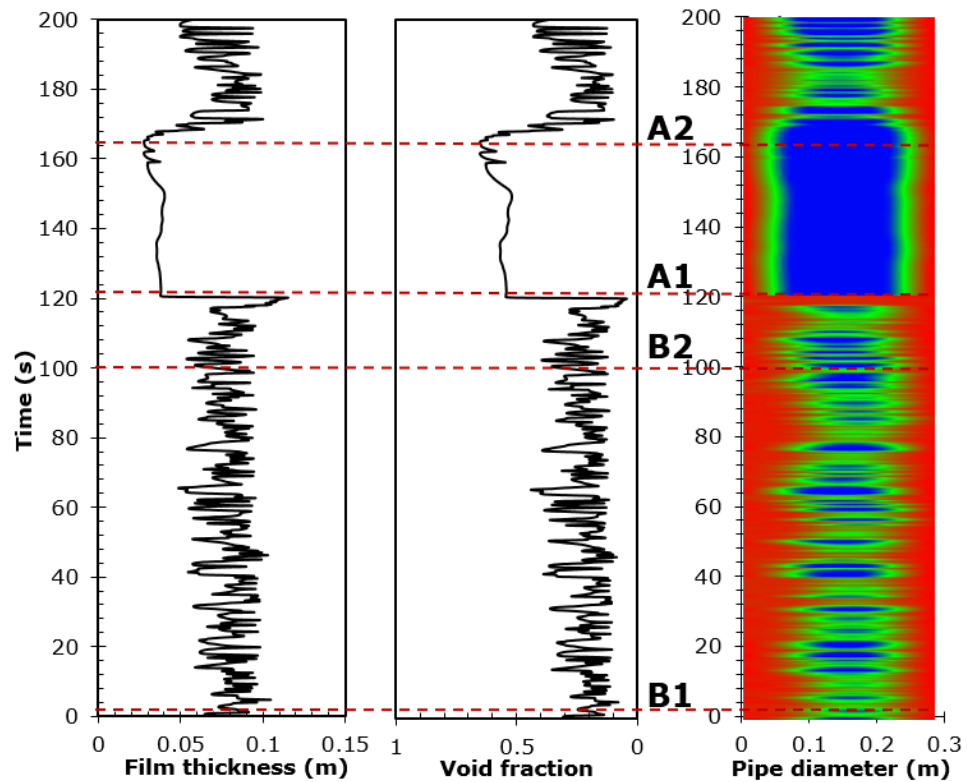


Figure 7.14: Film thickness, void fraction, and the reconstructed image using Plot3D software from the ECT for gas flow in 330 Pa.s Silicone oil and 290 mm diameter column. The 3 plots represent the flow structure for 200 s (10000 frames) at 0.566 m/a gas superficial velocity obtained from the upper plane of the ECT. B1 and B2 represent the churn areas. A1 and A2 represent the large bubble in the column. The cross section of the selected area is shown in Figure 7.13.

Comparing Figure 7.13 and Figure 7.14 gives a clear picture of the flow structure in churn flow regime. The churning regions exist between 3 and 100 s, corresponding to B1 and B2 in these two figures. Also (B) at 350 and

500 s at the time series in Figure 7.4 at the same flow rate. The gas core in churn flow regions seem to change its location from the centre to near the wall of the column unlike the case for low viscosity liquids where it always exists in the centre of the pipe. For example the gas core in B1 is next to the column's wall while in B2 is almost at the centre of the column. This also can be seen clearly from the photos of the column in Figure 7.4. This might be due to the effect of the very high viscosity and the very low surface tension (0.02 N/m) which is negligible in very high gas flow rates Snabre and Magnifotcham (1998). At about 120 and 166 s (A1 and A2) in both figures, the flow appears as a large bubble with a diameter approximately equal to the column diameter. The difference between this large bubble and the Taylor bubbles in slug flow regime is the top and the bottom shape of the bubbles. Large bubbles have almost a flat top and end and separated from the churn regions by a thin film. The length of these large bubbles in churn flow regime seem to be dependent on the gas flow rate. In the time series of void fraction in Figure 4.5 a single large bubble appears in churn flow regime, plots 0.276 to 0.566 m/s. the length of this single bubble increases with increasing the gas flow rate.

The liquid film in churn flow varies in direction and thickness in both sections A and B. The liquid film surrounding the large bubbles flows downward to merge with the liquid travelling upwards in the following churn region. While the film in the churn region flows in different directions and rests at some points. The average film thickness seems to decrease slightly at the same gas flow rate, this will be discussed in the next section. Figure 7.15 and Figure 7.16 show the film thickness of gas flowing in 330 Pa.s Silicone oil and 290 mm diameter column at two different flow regimes. The first figure is the transition to churn flow at gas superficial velocity of 0.243 m/s.

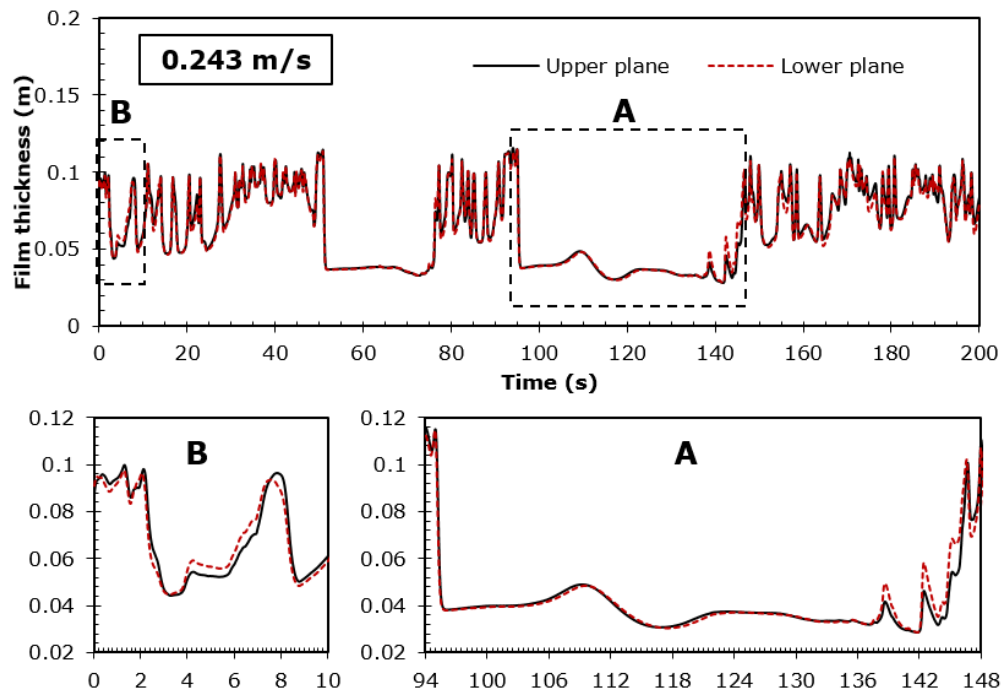


Figure 7.15: Film thickness at the transition to churn flow in 330 Pa.s and 290 diameter column at gas superficial velocity of 0.243 m/s.

The second figure (Figure 7.16) is at gas superficial velocity of 0.566 m/s which represents the film thickness in churn flow regime. In both figures, 200 s length of data is presented at acquisition frequency of 50 Hz. Both figures focus on the direction of the film in the large bubble and the churn sections. In the ECT sensor, the electrodes in plane 1 are located above the ones in plane 2. This makes the value of wave's velocity positive when they pass from plane 2 (lower plane) first. The values of the wave velocities were calculated from the time delay of each wave and the distance between the two planes. Velocity is equal to the distance between the planes (0.036m in this sensor) divided by the difference in time. Similarly, if the values of the calculated wave velocity are negative, this indicates downward film flow. Since the wave passes through the upper plane first P1 then to the lower plane P2.

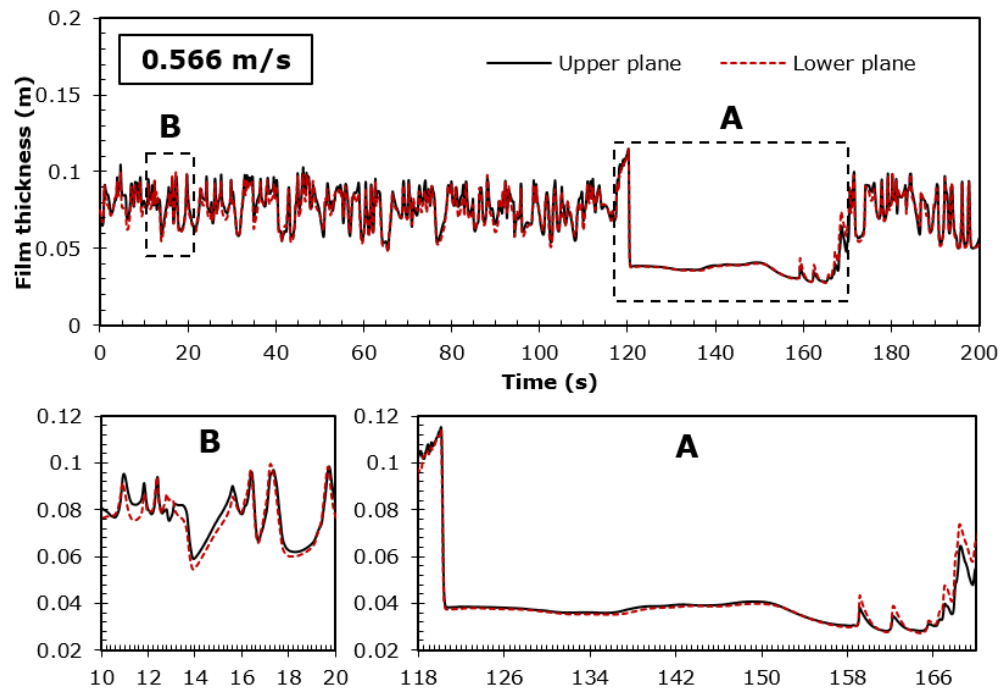


Figure 7.16: Film thickness at churn flow in 330 Pa.s and 290mm diameter column at gas superficial velocity of 0.566 m/s.

In general, the liquid film flow shows an average of downward movement in the large bubbles area ((A) in both figures) and upward flow in churn flow region (B) with some difference. Starting with the direction of liquid flow in the large bubbles (A) in both flow regimes (both figures). The waves' velocity showed negative values (downward flow) for all the waves in churn flow while it showed some upward flow in the transition to churn flow regime. For example, the calculated velocities for the wave on the film around the large bubble at the transition to churn flow were -0.078, -0.052, 0, 0.3, and 0.3 m/s at about 107, 120, 138, 142, and 146 s respectively. The velocity and the direction of the flow are not constant in this flow regime. The film flows downward between about 94 and 130 s. Then there is a time when the liquid holds/stagnates (between about 137-140 s) before it changes direction flowing upward from about 142 to 146 s. This area is followed by churn section. This upward direction for the film around the large bubble might be due to the development of this fraction to churn.

Different behaviour is exhibited by the liquid film around the large bubble

(A) in the churn flow is shown in Figure 7.16. In this flow regime, the calculated velocities of the waves were -0.045, -0.075, -0.22, 0, -0.225 m/s and hold at about 136, 152, 159, 162, 166, and 168 s respectively. This provides an evidence of downward direction of movement of the film around the large bubble.

The following churn section in churn flow regime did not show a significant effect on the direction and velocity of the film around the large bubble at the bottom. This might be due to the structure of flow in churn flow regime which seems to be more stable in terms of structure velocity, mean void fraction and mean film thickness. The wave velocities in the transition to churn flow and churn flow regimes were less than the values reported by Benjamin (1957). This can be referred to the very high viscosity of the liquid employed in this study. Benjamin studied the characteristics of the stability of laminar stream of viscous falling film on an inclined plane. According to his analysis, the velocity of the wave was equal to $-3 U_f$ for liquid film Reynolds number equal to 0. U_f is the velocity of the falling film.

The film thickness in churn sections -(B) in both flow regimes- shows different values in film thickness and directions compared to that in large bubbles. In general, the direction of flow fluctuates between up and liquid stagnation in churn flow regime, and upward flow in the transition to churn flow regime. The waves' velocity in the transition to churn flow is higher than the ones registered in churn flow. They vary from 0.22, 1.8, 4, and 0.1 m/s at about 0.2, 1.2, 2, and 8 s respectively (Figure 7.15). While in churn flow regime in Figure 7.16, the wave velocities were 0.6, 0, 0, 0, 0.3, and 0.9 at about 11, 12.4, 15.6, 17 and 19.7 s respectively. This times of liquid holding agree with the visual observation as the gas seems to move as pulses during experiments.

Comparing these results of the film thickness, reconstructed images and the photographs of these two flow regimes show two different flow characteristics. Both consists of two sections, one contains large bubbles which differ from Taylor bubbles but has a thin falling film. And second, sections of high frequency activity of liquid bridges with a thick film traveling in different directions. These two flow regimes can be distinguished by the frequency of the high activity sections (churn) which increases with increasing the gas flow rate. Also by the decrease of frequency of large bubbles with increasing gas input. Then, at a specific gas flow rate, the flow structure changes significantly (at 0.276 m/s gas superficial velocity). The flow at this gas flow rate is dominated almost completely by high frequency liquid bridging. A single large bubble might appear in a number of gas flow rates. Therefore, the flow regime from gas superficial velocity of 0.108 to 0.243 m/s can be characterised as transition to churn flow. While the flow regime from gas superficial velocity of 0.276 to 0.566 m/s is churn flow. Refer the time series of void fraction plots in Figure 4.5. The liquid film motion directions in churn areas are upwards and stagnation as indicated by the holdup which in turn refer to flooding. However, the waves in the flooding regions are not always showing the holdup effectively.

Seemingly, one of the known mechanisms for the transition to churn flow regime is flooding of the liquid film which occurs in slug flow inside the Taylor bubbles (Jayanti and Hewitt (1992) and Jayanti et al. (1993)). Govan et al. (1991) proposed that the transition to churn flow regime occurs due to the generation of "flooding-type waves". They also assumed the existing of the strong relation between churn flow and flooding phenomenon. The difference between the present work and the previous works from literature, which have been reviewed in Chapter 2, is the much larger pipe diameter and the much higher viscosity. Besides, the droplets which from the breaking up of

the flooding waves that occur in low viscosity liquids were not observed in the present study. In such high viscosity no entrained fractions of liquid were found due to the very high viscosity of the oil.

Figure 7.17 displays the effect of increasing the gas flow rate on average film thickness in the transition to churn and churn flow regimes. In general, gas flow rate shows more effect on film thickness in the larger diameter column. A change in the trend of the liquid film between the two flow regimes appears clearly in the larger diameter column. However, the film thickness in the smaller diameter column represents only one flow regime (the transition to churn flow).

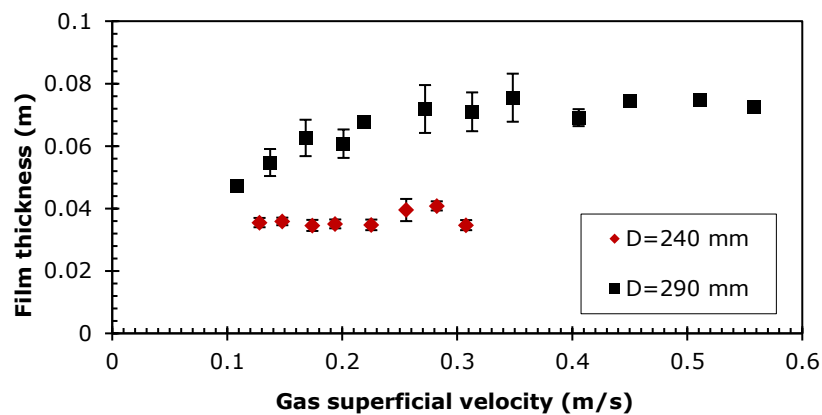


Figure 7.17: Averaged film thickness around the gas passing through 360 and 330 Pa.s Silicone oil in 240 and 290 mm diameter columns respectively at the transition to churn flow and churn flow regimes.

However, in the larger diameter column, the film thickness increased gradually by 0.2 m in the transition to churn flow regime while it increases slightly by only 0.005 m in the 240 mm diameter column. This increase in the film thickness values might be caused by the increasing frequency of the liquid bridges (churn) which have higher film thickness compared to the large bubbles region. Then when comparing the time series data in Figure 4.4 and 4.5, the frequency of the churn regions is higher in the larger diameter column. This explains the difference between the trends in both columns.

Also, in the larger diameter column, the frequency of the churning regions increases significantly in the transition to churn flow regime while it exhibits a slight increase in the smaller diameter column. This corresponds to the film thickness values in Figure 7.17.

In the churn flow regime, starting from 0.276 gas superficial velocity in the 290 mm diameter column, the film thickness appears to be more stable. It increases by only 0.003 m over a wide range of gas flow rates. This can be related to the time series data in Figure 4.5. The mean film thickness can also be linked to with Figure 7.7 the averaged void fraction. The error bars represent the standard error for the film thickness values. The maximum errors are 0.76% and 0.35% for the 290 and 240 mm diameter columns respectively.

7.2.5 Oscillations at the Top Section

The behaviour of the top surface of the bubble column was observed in transition to churn flow. The height of the liquid in the column was recorded at gas superficial velocity of 0.178 m/s for 10 minutes at 4 different times over 5.5 hours of the gas injection. In this experiment the gas was injected for 5.5 hours at constant flow rate to study the effect of the gas injection time on the flow structure in 330 Pa.s Silicone oil. This experiment has been explained in Section 6.2 Chapter 6. Figure 7.18 shows the top surface height obtained at 4 different times during 5.5 hours of constant gas injection to 290 mm diameter column and the corresponding time series of void fraction. In general, the top surface oscillation corresponds to the eruption of long bubbles shown in the time series of void fraction. The liquid height oscillates periodically in the data collected at 150 and 330 minutes from the start of the injection, while a different trend was obtained in the data collected at 30 and 240 minutes. This is due to the unstable structures in the transition to

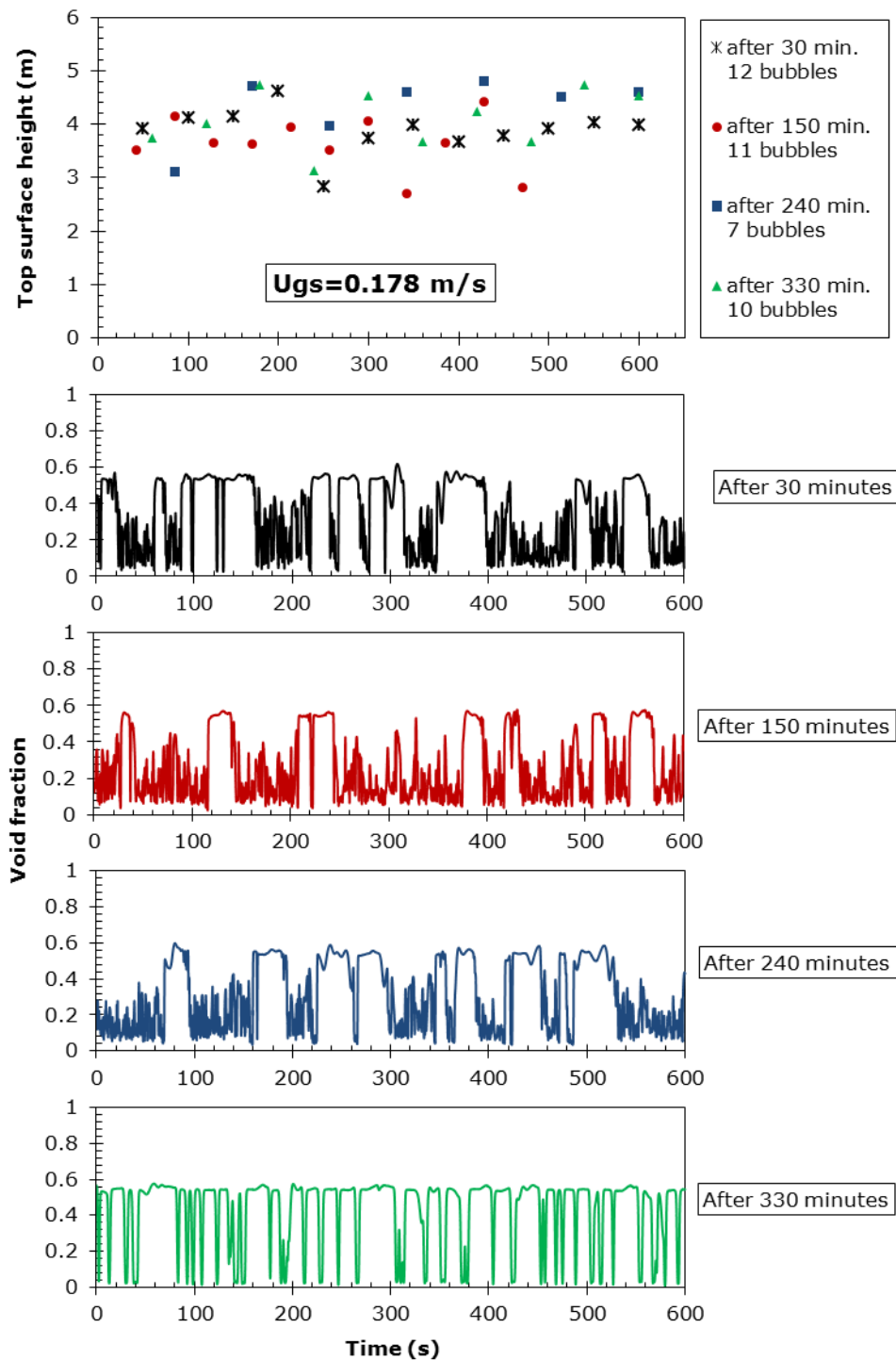


Figure 7.18: The change of the position of the top surface (liquid height per single bubble) during time for constant gas superficial velocity in four different times during the day in 290 mm diameter column. The colours of the time series of void fraction correspond to the height of the top surface for each time.

churn flow. It also might belong to the uncertainty of locating the liquid height due to the presence of liquid film on the pipe wall from the previous bubble's eruption. The difference in the average heights of the top surface

for the 5.5 hours is 140 mm. This is due to increase the concentration of the small bubbles (millimetres bubbles) in the column which also contribute to increasing the average void fraction as it shown earlier in Section 6.2. The fluctuation of the top surface decreases slightly from 1.8 to 1.6 m from 30 to 330 minutes. The reason behind that can be stemmed from the time series of the 330 minutes where the structure of the flow seems more organised. It reflects a lower rate of bubble coalescence.

7.3 Summary

Churn flow regime in high viscosity liquids has been studied in two large diameter columns over a wide range of flow rates. The **transition to churn flow** regime appeared in the range of gas superficial velocities between 0.127-0.314 m/s and 0.108-0.243 m/s in the 240 and 290 mm diameter columns respectively. **Churn flow** regime was observed in gas superficial velocity range of 0.276-0.566 m/s. The results of the time series of void fraction from the **ECT** and the visual observations from the **videos** have helped to distinguish between these two flow regimes. The characteristics of flow of high viscosity oils in the transition to churn and in churn flow regimes showed different behaviour compared to low viscosity liquids reported in the literature.

The **transition to churn** flow regime starts when the length and frequency of the regions of the high frequency activity- which appears at the end of slug flow region- increases. These regions of high frequency activities can be described as liquid bridges flowing upwards and then stagnates. The void fraction of these regions is lower than in Taylor bubbles. The source of the liquid bridge is from the drained falling film from the Taylor bubbles. With increasing the gas flow rate the length of Taylor bubbles increases producing a larger fraction of the film falling downwards which lead to increase in the

length of churn areas. The high frequency liquid bridges are named churn.

In **churn flow** regime, the structure of the flow changes significantly. The long bubbles disappear and the churn areas dominate the flow structure. Churn flow in 290 mm diameter column appears with two regions. First an open gas core with thick liquid film a possibility of the existence of liquid bridging creating a series of smaller bubbles. This gas core does not necessarily flow in the centre of the column. With increasing gas flow rate, the gas builds up and pushes the liquid creating a very long bubble with a falling thin film. This long bubble bursts at the top section of the column and remain open to the atmosphere. Then the falling film, which drains on the pipe wall, accumulates at the bottom of the column and creates the next churn unit.

The averaged **void fraction**, which was calculated from the time series data, seems to decrease with increasing gas flow rate in both columns. That is due to the increase of the frequency of churn units that have low void fraction. In the 290 mm diameter column, the mean void fraction continues to decrease at the transition to churn flow until it stabilises in churn flow. Unlike void fraction, the **structure velocity** increases with increasing the gas flow rate in both columns in the transition to churn flow regime. The bubble velocity increases with increasing the gas flow rate. Then no considerable increase was seen at churn flow regime as the flow structure becomes totally different. No fine bubbles were observed flowing in this flow regime. The **length** of large bubbles increases significantly with increasing gas flow rate in the transition to churn flow region. Whereas, it decreases in the churn flow regime in the 290 mm diameter column. Also, the length of the bubbles appears to be higher in the smaller diameter column for most of the gas flow rates. The **liquid film flow** shows an average downward flow movement in

the large bubbles area and upward flow in the churn regions. The direction of the flow fluctuates between upwards and liquid stagnation in churn flow regime, and upwards in transition to churn flow regime. Increasing the gas flow rate shows more increase in the **averaged film thickness** in the larger diameter column. A change in the trend of the liquid film between the two flow regimes was observed clearly in the larger diameter column. On the other hand, the film thickness in the smaller diameter column represents only one flow regime (the transition to churn flow). This increase in the film thickness values might be due to the increase in the frequency of the liquid bridges (churn) which have higher film thickness compared the large bubbles area. Finally, the **top surface oscillation** was also investigated in the transition to churn flow in the 290 mm diameter column. It seems to fluctuate according to the eruptions of long bubbles. The liquid height oscillates periodically for the 150 and 330 minutes, while a different trend was obtained in the data collected at 30 and 240 minutes. This is due to unstable structures in this flow regime and also the uncertainty of locating the liquid height due to the clouding by the liquid film on the pipe wall from the previous bubble's eruption.

Chapter 8

The Small Bubbles in Viscous Liquids

8.1 Introduction

This chapter focuses on the behaviour of the small bubbles, which appear and accumulate as the gas flows through high viscosity liquids in all flow regimes. "Small bubbles", in this work, refers to the bubbles of millimetres to centimetres diameter. The sources of these bubbles are from: the rupture of large bubbles at the top section of the column; at the gas injection points, and during the coalescence of the large bubbles. Two types of experiments were conducted and are presented in this chapter: the creation of small bubbles and their departure after stopping the gas injection. The rate of increase of the population of the small bubbles, their diameter and velocity were determined at two different locations in the 290 mm diameter column. Photos of the small bubbles were obtained at different times and gas flow rates to study the characteristics of the small bubbles in 330 Pa.s Silicone oil. Finally, the small bubbles were also observed in the stagnant oil after stopping the gas injection and were tracked over a period of days until the oil was completely free of bubbles.

8.2 Creation of Small Bubbles

Most of the research on clouds of small bubbles have been carried out using water or low viscosity liquids (Shah et al. (1982) and Lubbert et al. (1988)). However, a number of studies have used viscous liquids with viscosities about 1 Pa.s, (Schumpe and Deckwer (1987), Philip et al. (1990), Deckwer and Field (1992), Kawalec-Pietrenko (1992), and Kastánek et al. (1993a). Thus,

there is a lack of information about the characteristics of the small bubbles in very viscous liquids. In this work, 360 and 330 Pa.s Silicone oils in 240 and 290 mm diameter column were employed to study the behaviour of the small bubbles. Videos and photos of the small bubbles were obtained using the high-resolution camera, Canon EOS 600D, H&V resolutions of 72 dpi, through the transparent wall of the columns. The measured positions of the small bubbles were close to the column wall, as the diameter of the column is very large compared with the diameters of the bubbles. Therefore, the curvature of the wall was neglected as the distortion is very small. The observation of the small bubbles was achieved at two positions: first, at the top section of the column (at the top surface of the liquid) and, second, at about 0.6 m from the gas injection section at the bottom of the column. These experiments focused on the generation of the small bubbles, obtaining the distribution, the population, and the dimensions of the small bubbles.

Two classifications can be applied to categorise the small bubbles. The first category is their size since two main scales of the bubbles were observed (millimetres and centimetres). Second, the source of the bubbles, as they were created from three main sources. In this work, the second classification will be applied to identify the small bubbles in high viscosity liquids.

According to the visual observation, the small bubbles were generated from three main locations in the column. The size, the shape and the population of the small bubbles are mainly controlled by the location of the bubble generation. The structure of the flow also plays a considerable role in the creation of this type of bubbles. For example, the main source of the small bubbles is different in bubbly and slug flow than churn flow. In the bubbly and slug flow regimes, the main source is at the top surface of the liquid. While at churn flow regime it is seen at the churn regions due to the high

interaction activity between the two phases. Therefore, small bubbles in high viscosity liquids generate from: bubble eruption at the top surface, gas injection points at the bottom of the column, coalescences between the large bubbles, and finally the churn regions in the column.

At bubbly and slug flow regimes, the large bubbles rise in the column due to the balance between the buoyancy and drag force. The large bubbles push the liquid slugs at the top of the bubbles until a very thin film remains at the top of the bubble, see (C, D, and E) Figure 5.23 in chapter 5. Then the thin film collapses (bubble burst) and falls on the top surface of the liquid, while the next liquid slug still pushing the bottom of the erupted bubble (F) in Figure 5.23. In bubbly flow, when the diameter of the large bubbles are smaller, the collapsed film falls on two positions, the liquid ring around the erupted bubble and inside the erupted bubble (the bottom of the bubble). While at slug flow regime, when the bubble diameters are approximately equal to the column diameter, the film remains on the pipe wall and drains to merge with the liquid. The film, which results from the bubble eruption in both cases, traps air and creates a ring of unstable bubbles. These temporary bubbles, in turn, create smaller bubbles due to the shearing by the liquid motion on the pipe wall in case of higher gas flow rates. The amount of the small bubbles, which generates in higher gas flow rates, is higher not only because of increasing the surface area of the liquid film but also due to increasing the frequency of the bubble eruption. The mechanism of small bubble generation at the top surface of the liquid was studied by Pandit et al. (1987), Seyfried and Freundt (2000), and Bird et al. (2010). The small bubbles, which are created at the top surface, are smaller in diameter, relatively equal in size, homogenous, higher in concentration compared with the other small bubbles from different sources. They spread inside the column by the circulation motion of the liquid due to the rising of the large

bubbles, which push a fraction of the liquid down as a falling film. The diameter of the small bubbles ranges between about 0.1 to 5 mm during the bubble generation stage at the top section.

The small bubbles were also seen to generate from the gas inlet section at the bottom of the column. The bubbles at this area appeared with different characteristics. They were not homogenous, as the variation in bubble size is very large due to the mechanism of the bubble generation at this area. Figure 8.1 shows the small bubbles at about 0.4 m from the gas inlet point. In this figure, only one gas nozzle is employed to inject the air with two different positions and equal gas flow rates, one at the centre of the column and another at 105 mm from the centre (next to the pipe wall). In general, the small bubbles are created from the rapid break up or coalescence of the large bubbles at the gas nozzles. The large drag forces are increasing the possibility of the coalescence at the gas inlet points. However, the effect of the shearing at the pipe wall can be clearly seen in the figure, when the nozzle next to the pipe wall is used. The rate of the breaking up of the large bubbles increases due to the friction by the wall. The rate of the small bubble creation is higher, the shape and the size of the bubbles are not structured. Since 25 gas injection points were used in this work at high gas flow rates, the possibility of the interaction between both fluids increases at this position. Therefore, more friction, more coalescence and more breaking up of the large bubbles occur at the gas inlet section due to creating one bubble from 25 gas nozzles. The mechanism of the small bubble generation was produced by Pandit et al. (1987) and Philip et al. (1990). The shape, size, and rate of increasing of the small bubbles at this area do not depend only on the number and location of the gas injection points, but also on the time of gas injection. Then, in general, the bubbles are ellipsoidal with a length that varies from 4 cm to 1 cm and diameter from 2 cm to 2 mm. These small

bubbles either rise with the liquid slug then coalesce with other bubbles or remain on the pipe wall and shear to create smaller bubbles.

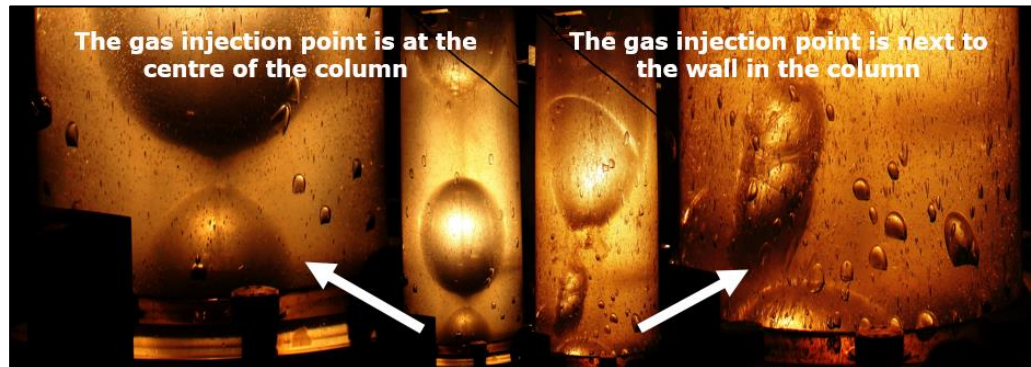


Figure 8.1: comparing the generation of the small bubbles at the gas inlet section using single gas injection point with two different positions at the same gas superficial velocity (0.03 m/s).

Another source for the small bubbles generation is the coalescence between pairs of large bubbles when rising in the column. The mechanism of the coalescence between the large bubbles is described in detail in Chapter 5. The small bubbles are produced mainly due to the entrainment of the two large bubbles of different rising velocity. This phenomenon was studied by Philip et al. (1990) and Manga and Stone (1994). It is also seen to produce due to the effect of the wake when the bottom of the upper bubble changes from rounded into concave. When the two large bubbles merge, the liquid at the bottom of the lower bubble remains as a thin liquid film until it collapses at the last stage of the bubble coalescence. This liquid film also generates small bubbles, (Bhaga and Weber (1981) and Colella et al. (1999)). Figure 5.16 in Chapter 5 shows the small bubbles generated from a coalescence of two bubbles flow at 0.009 m/s gas superficial velocity.

The small bubbles which produced at churn flow regime are relatively different due to the complicated structure in this flow regime besides the very high gas flow rate. Factors such as the high frequency and the unstable direction of the motion of the liquid bridge in addition to the high interaction between both fluids and the drainage of the liquid film play role in the

generation of the small bubbles. The small bubbles, in this case, are not uniform in size and shape. However, the rate of increasing the concentration and the distribution of the small bubbles is very high due to the very high gas flow rate and the motion of the liquid. Figures 7.2 and 7.4 in Chapter 7 display the beginning of the creation of small bubbles in churn flow regime.

8.2.1 Small Bubbles Concentration

The concentration of the small bubbles in high viscosity liquids is controlled by different parameters: gas flow rate, the location and the number of the gas injection points, column diameter, the viscosity of the liquid, and finally the time of gas injection. In this experiment, the rate of changing the concentration of the small bubbles in 330 Pa.s Silicone oil and 290 mm diameter column was determined at a constant gas flow rate. The gas was injected using 5 injection points at a gas superficial velocity of 0.006 m/s for 1 hour. Photos of the column were taken every 10 minutes from the top section of the liquid and the bottom of the column (60 cm from the gas inlet). The gas injection was stopped when the photos were taken to obtain a clear image of the small bubbles only and to avoid the effect of the large bubbles. The photos of the small bubbles were obtained using the high-resolution camera, Canon EOS 600D. In the 2D image processing, the distortion by the curvature of the cylindrical pipe was neglected as the bubbles were next to the pipe wall. More information about the experimental arrangements can be found in Section 3.4.6 in Chapter 3. Similar works in literature have been done to study the small bubbles behaviour in gas-liquid flow by Fraser et al. (1963), Azzopardi and Zaidi (2000), Bröder and Sommerfeld (2007), Mena et al. (2005), Majumder et al. (2006), and Schäfer et al. (2002). The difference between the previous works and the present work is that, in this work, a very high viscosity liquid is employed with a large diameter column.

Due to the high viscosity of the liquid, the generated small bubbles accumulate in the liquid increasing with the time of the gas injection as expected. Figure 8.2 shows increasing the concentration of the small bubbles with time at gas superficial velocity of 0.006 m/s. The population of the bubbles appears higher at the top section of the column. This is because a number of bubbles generated by the eruption of the large bubbles at the top are higher than those generated due to the effect of the gas injection points.

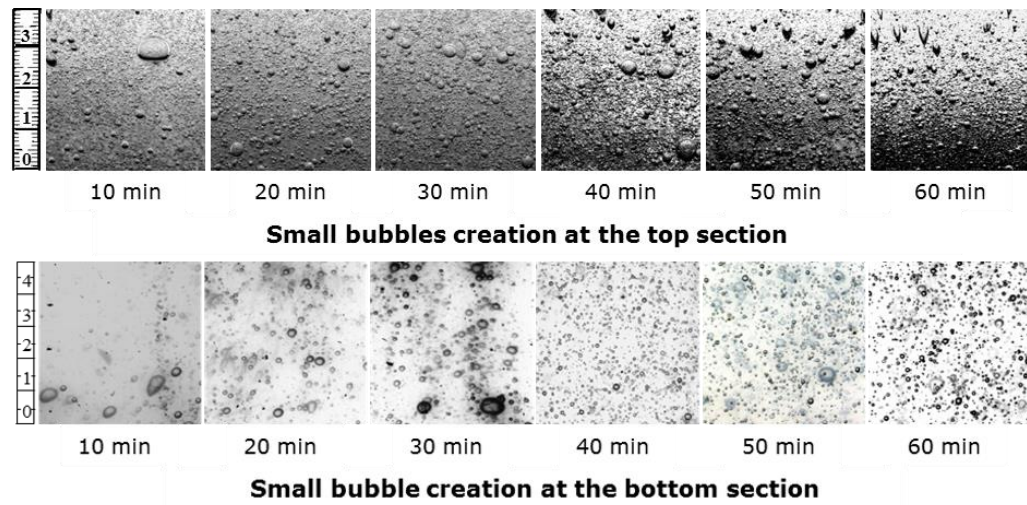


Figure 8.2: Small bubble creation in high viscosity liquid at two different positions of the column and different times of gas injection. The first photos from the top and the bottom are taken after injecting gas for 10 minutes and the last ones after one hour at a gas superficial velocity of 0.006 m/s. the gas injection stopped when the photos were taken. The resolution of the images is 28.85 Pixel/mm.

On the other hand, at low gas flow rate, and due to the slow motion of the liquid, mixing of the bubbles is not efficient at the first hour of gas flow rate. The efficiency of the mixing increases with increasing the gas flow rate as well as increasing the time of gas injection. Therefore, the rate of increasing the concentration of the small bubbles in this experiment is only applied for low gas flow rate. According to the visual observation the rate of increasing, the concentration of the small bubble was much higher at high gas flow rates. This is due to increasing the frequency of the large bubble eruption and coalescence. At both locations, the top section and the bottom of the column, the variation between the size and the shape of the small bubbles decreased

with time. For example, after 10 minutes in the top section, the diameters of the bubbles were about 5.8 to 0.2 mm. Then, after one hour, they became about 4.6 and 0.1 mm. The same scenario occurred at the bottom of the column; the diameter of the bubbles varied from about 4.6 to 0.3 mm and after one hour became about 1.49 and 0.17 mm. This variation is caused due to: first, the time, as more mixing between the bubbles achieves less variation in size. Second, the difference between the top and the bottom in term of the structured size is due to the nature of the mixing between both fluids in this area, which differs totally from the top section.

The volume fraction which is occupied by the bubbles was divided by the total tested volume at the bottom and the top of the column. The volume of the bubbles was determined by removing the image background using a photo of the bubble column in the absence of the bubbles as a thresholding, (Otsu (1979)). The details about this image analysis are shown in Chapter 3. The only limitation of this method is that it shows the volume occupied by the bubble approximately larger than the actual volume. This is because the overlap caused by the bubbles behind the bubbles which are positioned next to the pipe wall. The rate of the overlap increases with increasing the population of the small bubbles. Figure 8.3 shows the percentage of the volume occupied by the small bubble at the same location as Figure 8.2. The volume occupied by the small bubbles increased gradually as expected with increasing the time of gas injection at a constant gas flow rate (Philip et al. (1990)). Moreover, the volume occupied by the small bubbles at the top section of the column was seen larger than the bottom section. It increased from 12.3% after 10 minutes of gas injection to 24.7% after 60 minutes at the top section, whereas it increased from 4.3% to 13% at 10 and 60 minutes respectively. This agrees with the visual observations in the images in Figure 8.2.

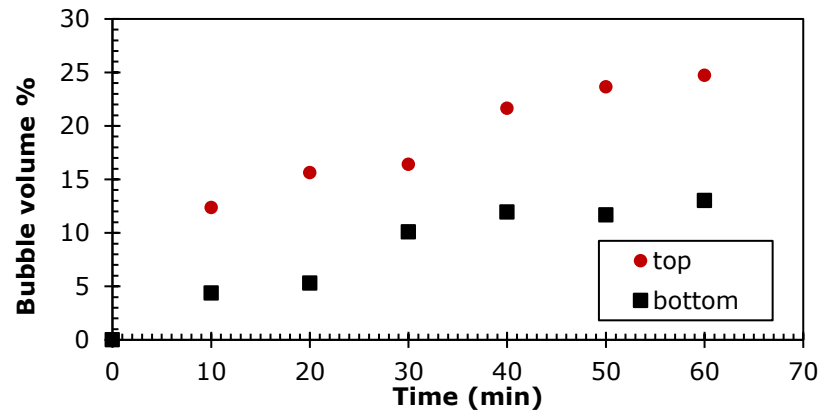


Figure 8.3: The concentration of the small bubbles during the creation stage at the top and bottom section of the column at a gas superficial velocity of 0.006 m/s.

The rate of the small bubbles formation in this work showed different results compared with the work presented by Pandit et al. (1987) due to the difference in viscosities of the liquids. The mechanisms of the large bubble eruption in such high viscosity liquid showed a different behaviour compared with the method described by Pandit et al. (1987). Figure 8.4 illustrates the mechanisms of the bubble eruption in 330 Pa.s.

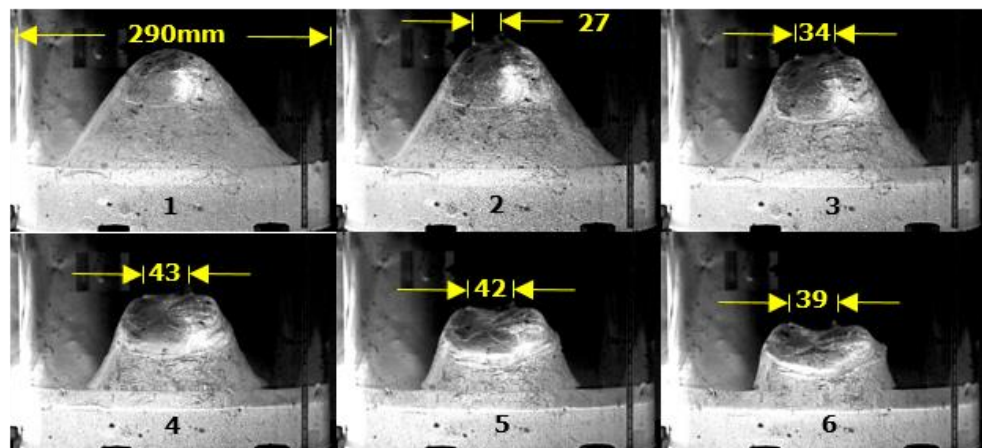


Figure 8.4: The large bubbles eruption at the top surface of a column of 330 Pa.s Silicone oil and 290 mm diameter. The dimensions in the photos 2-6 correspond to the diameter of the hole created from the ruptured film.

The liquid film above the bubble becomes thinner and rupture at the higher point when the pressure inside the bubble is higher than the atmospheric pressure due to the surface tension. The diameter of the hole created from the ruptured film remains smaller than the diameter of the ruptured bubbles.

The ruptured film then collapses and fall inside the bubble trapping air as temporary bubbles which break down to small bubbles due to the shearing by the liquid motion. The rim of the hole of the ruptured bubbles seen as the same thickness as the ruptured film unlike the 1 Pa.s glycerol which employed in work by Pandit et al. (1987). Therefore, comparing the rate of increasing the small bubble concentration in this work with Pandit et al. (1987) is not applicable.

8.2.2 Dimensions of Small Bubbles

The effect of the time on the diameter of the small bubbles, in high viscosity liquids, was determined at a constant gas flow rate. The same procedure in Section 8.2.1 was applied for the data acquisition. ImageJ software was used to scale and determine the bubbles diameters in both locations (top and bottom of the column). The bubble diameters shown in Figure 8.5 are an average of 95 and 57 bubbles at the top and the bottom of the column respectively. This figure can also be compared to the images in Figure 8.2, which displays the size of the bubbles. In general, the diameter of the small bubbles increased slightly in both locations to become approximately equal after one hour of gas injection. First, at the top section of the column, it increased significantly by 0.36 mm after 30 minutes then it decreased rapidly by 0.72 mm. While it increased slightly by 0.14 at the bottom of the column to become almost equal to the diameters from the top section. As it has been explained in the previous section, the mechanism of generating the bubbles controls the size and the shape of the bubbles. The small bubbles, which are created from the collapsed film from the large bubbles rupture at the top section, are relatively smaller and more structured.

On the other hand, the bubbles produced at the bottom, due to the gas injection effect, are larger and variable in size, but they become homogenous

by time due to the mixing. After 1 hour of gas injection, the average diameter in both locations becomes almost equal due to the mixing of the bubble inside the column. Therefore, at 0.006 m/s gas superficial velocity, the small bubbles, held up inside the column, become homogenous by 1 hour time. This amount of time changes proportionally when changing the gas flow rate. The bubble size distribution was also studied by Lage and Esposito (1999), Wongsuchoto et al. (2003), Mandal et al. (2005), Bordel et al. (2006), and Montante et al. (2008).

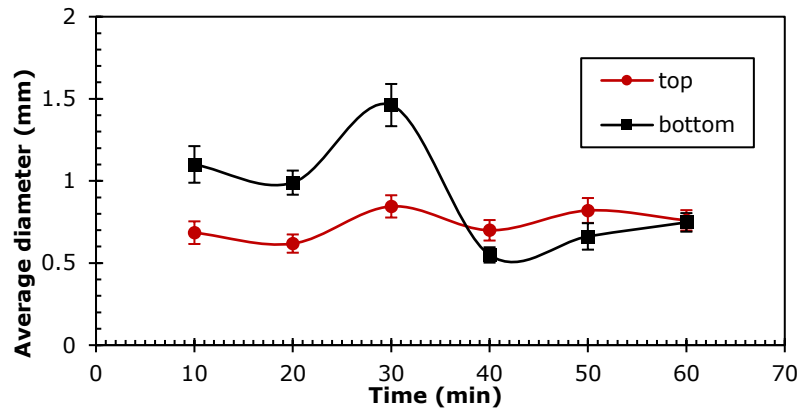


Figure 8.5: The diameter of the small bubbles during the creation stage at the top and bottom section of the column at a gas superficial velocity of 0.006 m/s.

8.3 The Characteristics of the Small Bubbles in Viscous Liquids

In gas-high viscosity liquids, the small bubbles, produced from the interaction between the two fluids, accumulate in the system for relatively a long time. This is due to the low velocity of the liquid circulation and the low motion of the liquid slugs. This delays the time of rising and releasing the small bubbles trapped by the liquid. This time is dependent on the viscosity of the liquids, Philip et al. (1990). In this work, the characteristics of the small bubbles were studied in high viscosity liquids and large diameter columns during the gas injections. The motion, direction, size and concentration of the small bubbles are the emphasis of this section. The high-

resolution camera, Canon EOS 600D and ImageJ software were used for this purpose. In general, the small bubbles were seen:

1. Flowing down with the falling film at the pipe wall around the rising large bubbles at bubbly and slug flow regimes. The bubbles in this area are elongated due to the liquid shearing at the wall. The velocity of the bubbles depends on the bouncy and shearing forces.
2. Flowing upward with the liquid slugs between the large bubbles at bubbly and slug flow regimes. The bubbles here have ellipsoidal shapes due to the surface tension of the liquid. The velocity of the rising small bubbles is a result of the buoyancy and drag forces. The bubbles can also be seen at churn flow regime in the draining film around the open gas core.
3. Oscillating up and down in the thick film and the liquid bridge in churn areas at churn flow regime. See 0.108 m/s in Figure 4.1 in Chapter 4. The bubbles are not structured in term of size, shape and distribution due to the complicated flow structure.

At low gas flow rates, the small bubbles appeared in low population, low rate of the population increasing, relatively equal in size and shape, less coalescence, small size, low velocity (both directions), and almost no bubbles at the pipe wall. With increasing the gas flow rate, the properties and the behaviour of the small bubbles change. They become higher concentration, a higher rate of increasing the concentration, variable in size and shape, a higher amount of coalescence, increasing the size, higher velocity (both directions), and increasing the number of the bubbles at the pipe wall.

Figure 8.6 shows the small bubbles rise in 330 Pa.s Silicone oil and 290 mm diameter column at a gas superficial velocity of 0.016 m/s using 5 gas nozzles. The photo was obtained after 40 minutes of gas injection at about 1m from the gas injection points. The photo is maximised and scaled to show

the actual sizes and shapes of the small bubbles. The bubbles are elongated due to the shearing caused by the liquid motion. They vary between about 10 to 0.5 mm length and 1.5-0.25 mm diameter. The density of the small bubbles in this photo is not high due to the low gas flow rate and the short time of gas injection. The large bubbles can be seen through the column in this case.

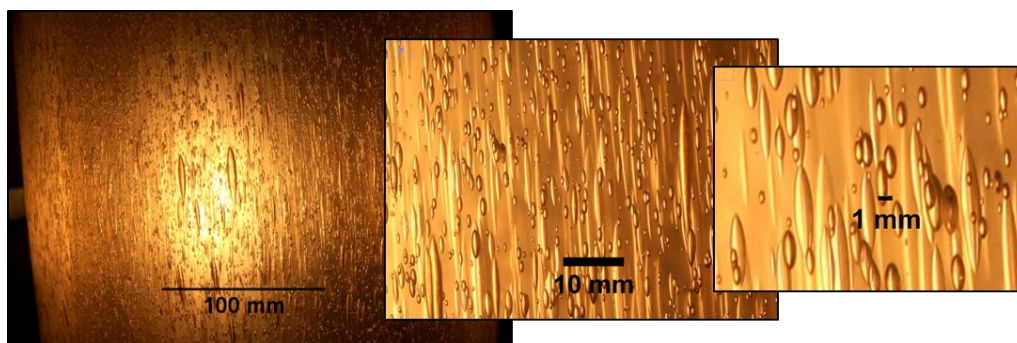


Figure 8.6: Photo of the small bubbles rising in 330 Pa.s Silicone oil and 290 mm diameter column after 40 minutes of injecting gas at 0.016 m/s gas superficial velocity.

The characteristics of the bubbles are not only affected by increasing the gas flow rate but also by increasing the time of gas injection. Therefore, the effect of the gas injection time on the behaviour of the small bubbles was involved in the study. According to the observation, increasing the time of the gas injection increases the population and the size of the bubbles.

In Figure 8.7, the photo was obtained after 3.5 hours of injecting gas at the same gas flow rate as the in Figure 8.6. The size and the population of the bubbles increased significantly compared with the previous photo. The concentration of the small bubbles increased as defining the boundaries of the large and small bubbles became difficult due to the milkiness of the liquid in this stage. The light scattering causes the milky colour. The milky colour appeared at the top section of the column first, and then it occupied the whole column gradually. This is because of the rate of the small bubble generation was higher at top section first. And second, the size of the small

bubbles was smaller in the top section. Then by the motion of the liquid, due to the large bubbles rising the small bubbles were transferred to the lower parts in the column. Very small bubbles of about $100\ \mu\text{m}$ in $300\ \text{Pa.s}$ and $240\ \text{mm}$ diameter column were reported by Azzopardi et al. (2014b).

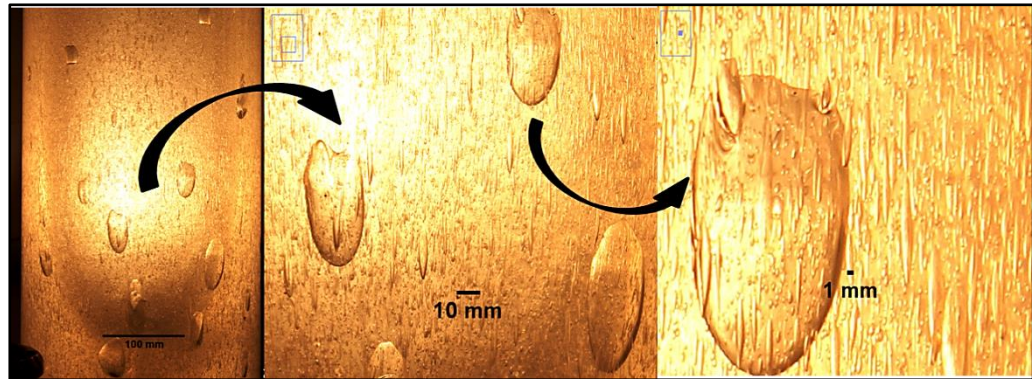


Figure 8.7: Photo of the small bubbles rising in $330\ \text{Pa.s}$ Silicone oil and $290\ \text{mm}$ diameter column during at gas superficial velocity of $0.016\ \text{m/s}$. The photo was taken after 3.5 hours of gas injection.

The size of the small bubbles also increased significantly to $4\ \text{cm}$ and $1\ \text{cm}$ length and $2\ \text{cm}$ and $0.2\ \text{cm}$ diameter. These bubbles of centimetres of size were seen on the wall of the lower part of the column. They appeared to elongate and break down to smaller bubbles due to the liquid shearing at the wall of the column. Coalescence was also observed to occur between the small bubbles due to the shearing by the liquid motion and the bouncy force. The rate of this coalescence increased with increasing the velocity of the large bubbles, which enhances the interaction and increases the shearing between the bubbles. These results agree with the work of Pioli et al. (2012), who used a wide range of liquid viscosities.

The contribution of the small bubbles in the void fraction was also studied. Philip et al. (1990) reported a 50% contribution of the small bubbles in the total void fraction using a wide range of liquid viscosities. They also found that a time of 15-60 minutes is needed to achieve stable void fraction values due to increasing the concentration of the small bubbles. However, Kuncová

and Zahradnik (1995), who employed liquids with viscosities of 1-30 mPa.s over a wide range of gas flow rates, showed quite different results. They found that the concentration and the void fraction of the small bubbles are independent of the time of gas injection at low and intermediate gas flow rates.

8.4 Removal of Small Bubbles in Stagnant Viscous Liquids.

In this Chapter, the behaviour of the small bubbles rising in 330 Pa.s Silicone oil, 290 mm diameter and 8.5 m column was studied. The initial height of the oil in the column is 3.60 m. The shape, size, concentration and motion of the small bubbles were determined at the bottom and the top section of the column. The same positions of the tested areas, as in the small bubble generation experiment, were used for the small bubbles removal. The small bubbles, which were accumulated in the column during the gas injection for 1 hour, were seen to remain trapped for days. The accumulated small bubbles appeared to rise at very low velocity due to the very high viscosity of the oil. The rising velocity of a single bubble in the stagnant liquid is controlled mainly by the buoyancy and drag force. The total drag force of sphere is mainly due to viscous force at very small Reynolds number $Re \leq 0.1$. The calculated Reynolds number for the small bubbles is about 2×10^{-7} near the top section of the liquid in the column. On the other hand, the buoyancy force, which is mainly a factor of the difference in density of both fluids and small bubbles diameter, is very small compared to the viscous force. Therefore, in the very high viscosity liquids, the terminal velocity of the individual small bubble is mainly controlled by the viscosity force. (Gaddis and Vogelpohl (1986) and Tomiyama (2002)).

In general, the small bubbles which rise in the very viscous oil appeared in

spherical shape due to the surface tension of the oil. No deformation of elongated bubbles has been seen. Less variation in the size distribution of the small bubbles was observed in the stagnant oil. Figure 8.8 displays the small bubbles, which are still rising in the Silicone oil after 3 days of stopping the gas injection. The photo was obtained near the top section of the column using a high-resolution camera. Diameters of the small bubbles ranged between 0.5-5 mm. No bubbles coalescence was observed during the rise of the small bubbles. Due to the very slow motion of the bubbles, it was difficult to observe the behaviour of the small bubbles. However, the coalescence between the small bubbles possibly happened due to the difference in the diameter of these bubbles. Further work can be done by recording videos for several hours to examine the possibility of coalescence of small bubbles rising in stagnant very viscous liquids. The shape of the small bubbles rise through a stagnant liquid was studied by Ellingsen and Risso (2001), Rastello et al. (2011), Keshavarzi et al. (2014), and Aoyama et al. (2016).

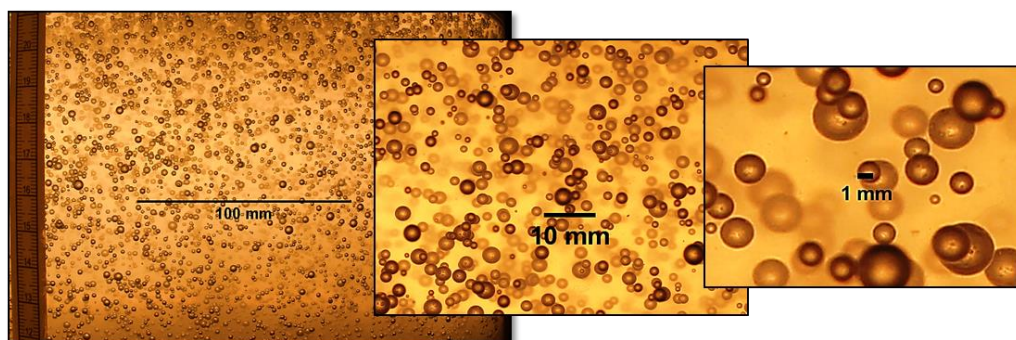


Figure 8.8: Small bubbles rising in stagnant 330 Pa.s Silicone oil and 290 mm diameter column at the top surface of the volume after 3 days from stopping the gas injection.

Figure 8.9 is a top view of the small bubbles rising in 330 Pa.s Silicone oil. The photo was obtained from the top section of the oil in the column after one day of stopping the gas injection into the column. The bubbles are in different sizes with a spherical shape.



Figure 8.9: The small bubbles at the top surface of high viscosity liquid (top view).

8.4.1 Small Bubbles Concentration

Removal of the small bubbles was observed in stagnant 330 Pa.s Silicone oil until oil was totally free of bubbles, which can be seen by naked eye. The microbubbles, in such viscous oil, were not studied at this stage. Further work into this area can be done using special visualisation methods. Photos of the small bubbles were obtained using high-resolution camera, Canon EOS 600D. During the processing of the 2D image, distortion caused by the curvature of the cylindrical pipe was neglected as examined bubbles were next to pipe wall. More information about experimental arrangements can be found in Section 3.4.6 in Chapter 3. Photos were taken from two positions, at the top section and about 60 cm from the gas inlet. After stopping gas injection in the previous experiment (bubble generation experiment), which was injected for 1 hour at constant flow rate, the first photos were obtained from both locations (top and bottom of the column) after 2 hours. Photos were obtained every 30 minutes over 2 hours at the bottom of the column until it became free of bubbles. Whereas, photos were taken for 5 days at the top section of the column. This was the required time for trapped small bubbles to rise in 3.60 m oil column. Figure 8.10 shows photos of the small

bubbles, which were taken at top and bottom sections of the column until the last bubble was released from the oil column.

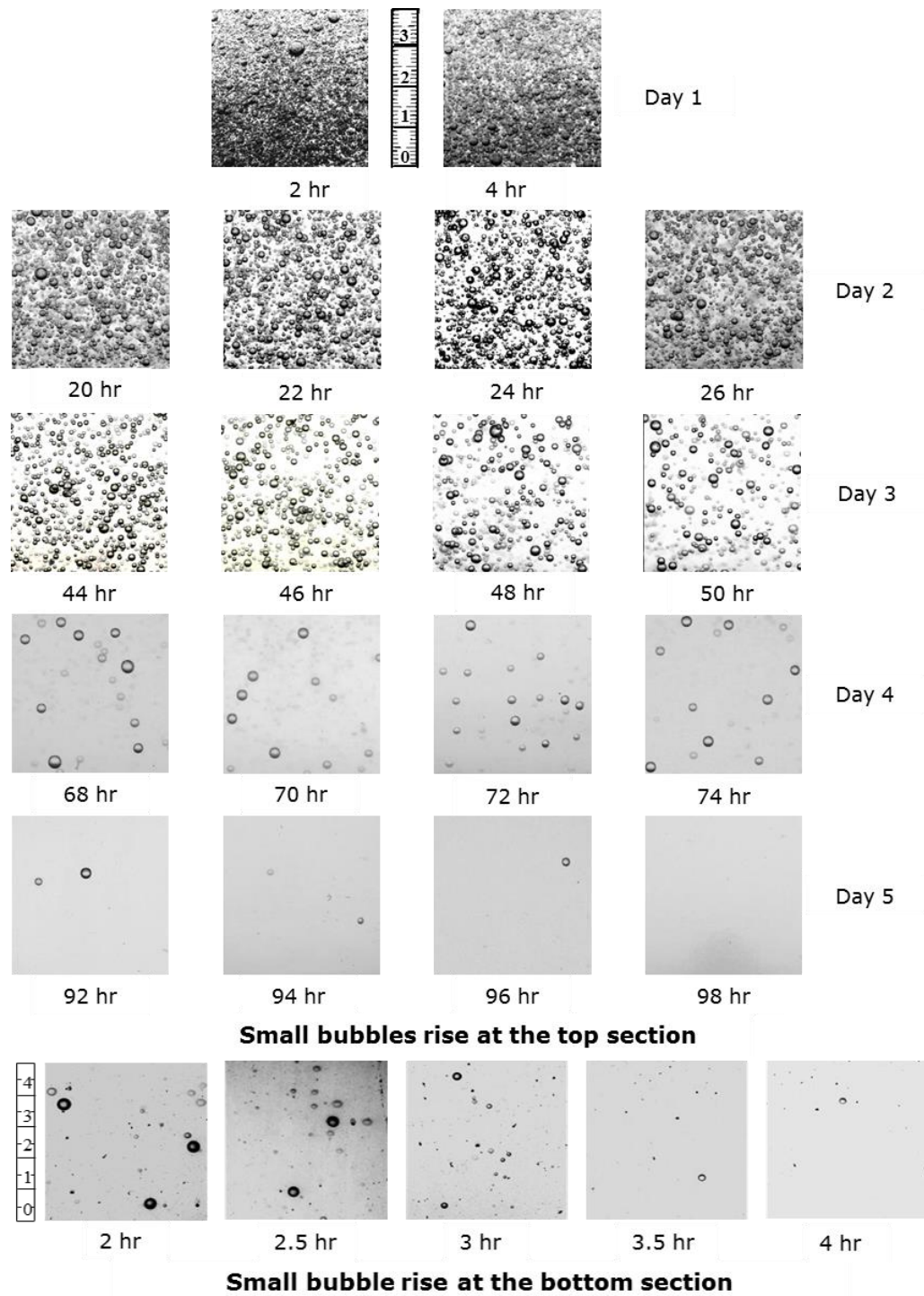


Figure 8.10: Small bubbles rise in stagnant 330 Pa.s Silicone oil at two different positions of the column. The photos are for five days for the top section and only 4 hours for the bottom of the column. The first photos of the top and bottom were taken after stopping gas injection by 2 hours. The resolution of the images is 28.85 Pixel/mm.

The concentration of the bubbles was seen to decrease as it is expected. In

general, the minimum size of the bubbles increased gradually from the second day (after 20 hours) of stopping the gas injection. They were about 0.14, 0.2, 0.57, and finally 0.3 mm from the 2nd to 5th day at the top section. On the other hand, the maximum size of the small bubbles decreased at the bottom section from 3.8, 3.5, 2.5, 2.11, and 1.85 mm in the 2, 2.5, 3, 3.5, and 4 hours respectively. This is due to the bouncy force, which is controlled by the bubble diameter (larger bubbles rise faster). Therefore, the population of the larger bubbles at the bottom decreased at the same time it increased at the top section. Another reason for increasing the size of the small bubbles, by the time they become close to the top section, is probably the coalescence between the bubbles due to the high population of the bubbles. The probability of the coalescence of the small bubbles was mentioned in the previous section.

The same method employed to determine the volume occupied by the bubbles during the bubble generation was applied to determine the same volume but for bubble removal. The method by Otsu (1979) was applied by changing the photos to binary images then by subtracting the background; the volume occupied by the small bubbles was determined too. Figure 8.11 shows the percentage of the volume occupied by the small bubbles rising at the bottom section of a column of 330 Pa.s Silicone oil. Two hours were required for the small bubbles to rise from the first 1 m in the bottom of 290 mm diameter column. The volume occupied by the small bubbles decreased steadily from 2.24 to 0.11 %. The only limitation of this method is the overlap, which occurs due to the high population of the small bubbles. This causes the higher value of the bubble volume, as the bubbles in the background merge with the front bubble and is considered then as one area. More information about this method is illustrated in Chapter 3.

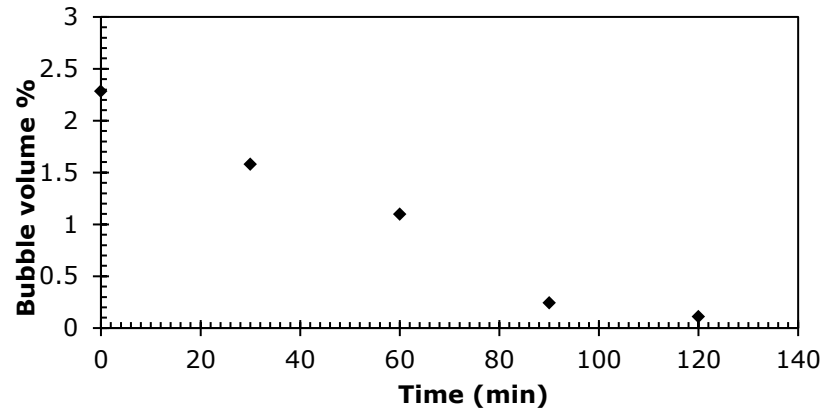


Figure 8.11: Concentration of small bubbles rise in stagnant 330 Pa.s Silicone oil at the bottom of the column.

On the other hand, at the top section of the column in Figure 8.12, the overall volume occupied by the bubbles decreased gradually from 36 to 0% in the 98 hours. It decreased slightly in the first 2 days from 36 to 27%. Then from the 3rd day, it decreased significantly from 27 to 0%. However, on the second day (between 20 to 24 hours), the volume occupied by the bubbles increased slightly 33 to 36%. This, probably, happened due to the accumulation of the small bubbles, which moved from the lower parts of the long column. The significant decrease in the volume occupied by the small bubble in the 3rd day is an indication of rising all the bubbles at the lower parts of the column.

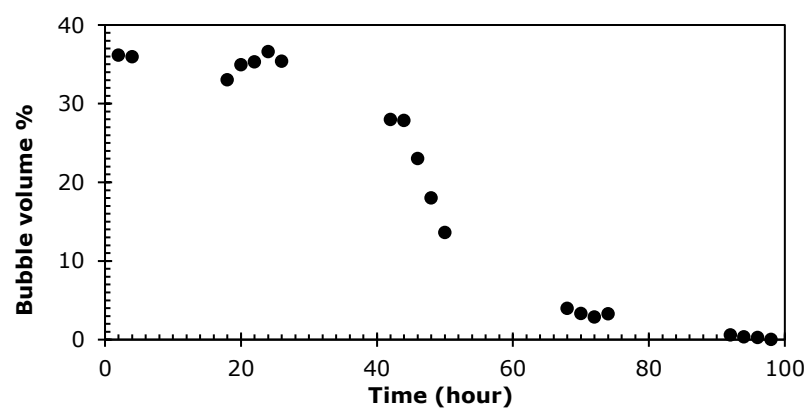


Figure 8.12: The Concentration of small bubbles rising in stagnant 330 Pa.s Silicone oil at the top section of the oil.

The void fraction of the small bubble was calculated from the total volume of the small bubbles inside the column, which was obtained from the level swell "oil-small bubbles". The total volume of the bubbles was obtained from

subtracting the initial volume of the liquid from the total volume of oil in presence of the small bubbles. Figure 8.13 shows the void fraction of the small bubbles over the 5 days of the bubble rising in a column of 3.6 m of viscous oil. The void fraction inside the column decreased indicating the decrease in the population of the small bubbles as in Figure 8.12. The void fraction of the small bubbles, during the generation stage, was not calculated from the level swell due to the difficulty of measuring the accurate height of the liquid. This is due to the falling film which remains on the column wall after the large bubbles eruption in the top section.

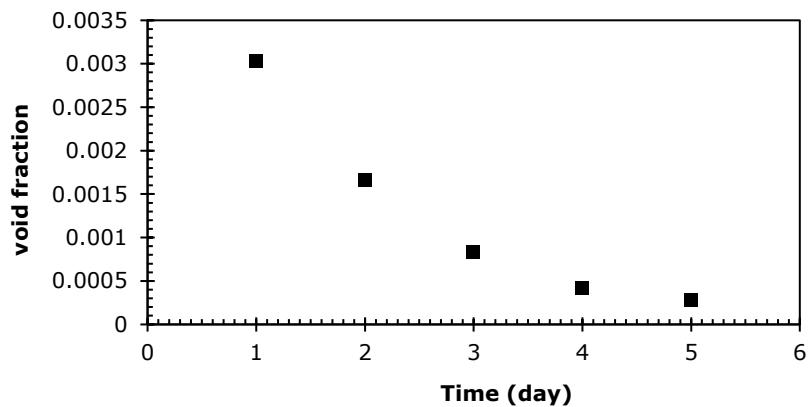


Figure 8.13: The void fraction of the small bubbles rise in a 330 Pa.s Silicone oil determined from the level swell.

The average suspension density, which represents the density of the oil in the presence of the small bubbles, was calculated by Stokes' law. It was calculated from the liquid volume fraction and the void fraction of the small bubbles. Figure 8.14 displays the average suspension density of Silicone oil containing the small bubbles. The density of the Silicone oil, inattentive of the small bubbles, is 950 Kg/m^3 . The average suspension density decreased due to the presence of the small bubbles and increased with decreasing the concentration of the small bubbles inside the oil, (Rhodes (2008)).

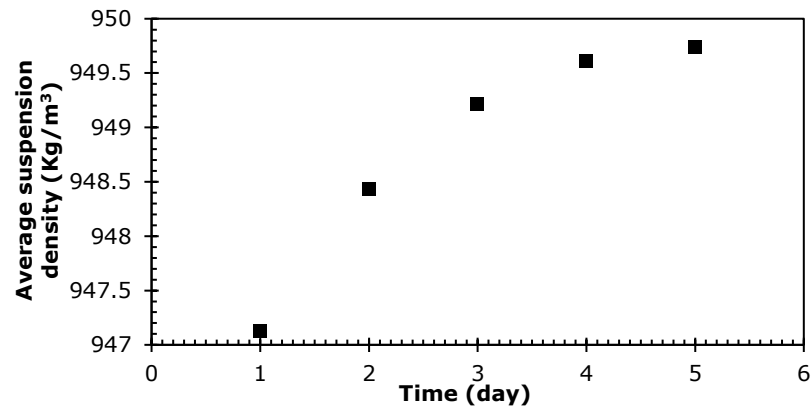


Figure 8.14: The average suspension density of air bubbles in 330 Pa.s Silicone oil over 5 days calculated by Stokes' law.

8.4.2 Dimensions Small Bubbles

The average diameter of the small bubbles was determined during the rise of the bubbles in the stagnant viscous Silicone oil at the top and the bottom section of the column. In general, the average diameter of the small bubbles increased slightly at the top section and decreased slightly at the bottom of the column.

Figure 8.15 shows the average diameter of the small bubbles rising at the top section of a column of 330 Pa.s Silicone oil. The values presented in the figure are an average of 58 to 12 bubbles for the whole time of data acquisition (5 days). The average diameter of the small bubbles was seen to decrease slightly by 0.01 mm over 20 hours. However, it appeared to increase by 0.25 mm on the 4th day. This is probably because of the coalescence which may occur between the bubbles throughout the 3 m oil column before arriving to the top section. The maximum standard error calculated for the diameter of the 58 bubbles was 5.94%.

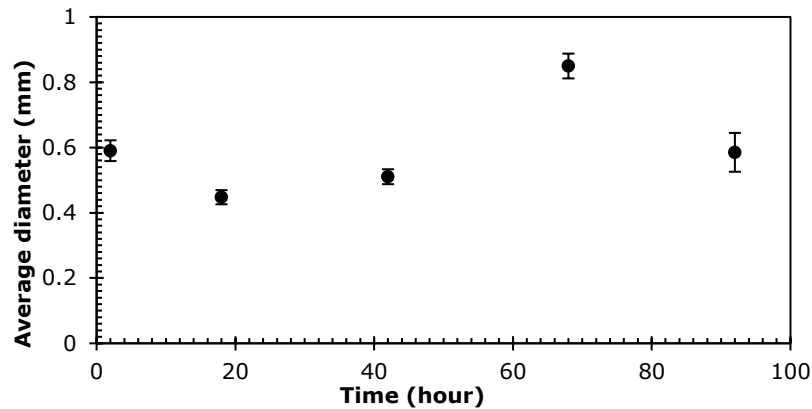


Figure 8.15: The average diameter of the small bubbles rises at the top section of column contain stagnant 330 Pa.s Silicone oil.

The average diameter of the small bubbles at the bottom section of the column in Figure 8.16 showed a slight increase by 0.23 mm. The values presented in the figure are an average of about 26 bubbles over the 2 hours. The average diameter of the small bubbles at the bottom of the column is larger than the top section due to the difference in the mechanism of the bubble generation in both positions. This was discussed in the small bubble generation section in this chapter. The maximum standard error calculated for the diameters of the 26 bubbles is 32%. This value is high due to the difference in the size distributions of the small bubbles at this section.

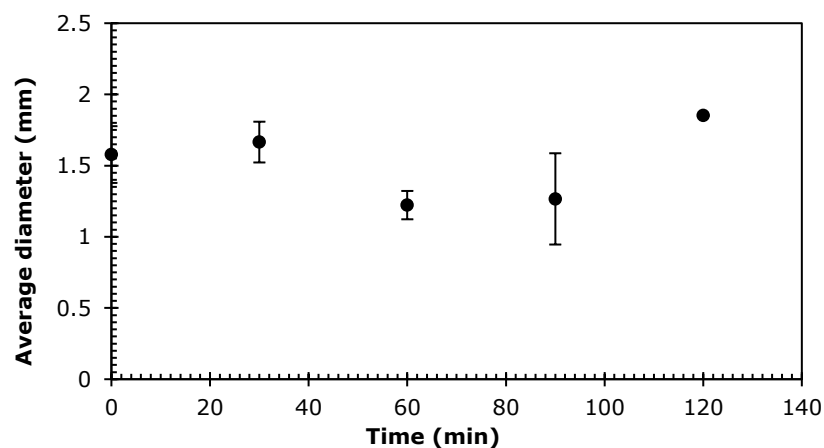


Figure 8.16: The average diameter of small bubbles rising in stagnant 330 Pa.s Silicone oil at the bottom of 290 mm diameter the column.

8.4.3 Velocity of Small Bubbles (rise velocity)

The terminal velocity of the small bubbles rising in very viscous liquid and large diameter column was determined and compared with the bubbles diameter. The velocity of the bubbles in stagnant liquids is basically a function of bubble diameter, the liquid viscosity, density, surface tension and gravity force (Ozdemir (2005)). In this work, since the viscosity of the liquid is very high compared with bubbles diameter, therefore the velocity is a function of the viscous force. The velocity was determined by tracing the selected bubbles periodically. Then, it was calculated from the values of changing the distance and time. See Chapter 3 for more information about the data analysis. The velocity of the small bubbles was studied by White and Beardmore (1962), Bhaga and Weber (1981), Tomiyama et al. (2002), and Talaia (2007).

In general, the bubbles were seen to rise in a very low velocity through the 3.60 m oil column. The rise velocity increased slightly in the first 3 days then it decreased gradually by the end of the 5th day. The values of the velocity at each time are an average of 8 bubbles. Each of the 8 bubbles was tracked for 6 positions to calculate the average velocity. The maximum standard error for this method is 0.13%. The average velocity of the small bubbles increased by 0.0054 mm/s over the first 3 days; then it decreased by 0.0055 mm/s from the 3rd day until the last day when the column became free of the small bubbles, see Figure 8.17. This might due to the effect of the change in concentration and the diameter of the small bubbles. The rise velocity of the small bubbles was hindered by the high concentration of the bubbles. Besides, the higher rising velocity bubbles increase the velocity of the slower bubbles, Acuña and Finch (2010).

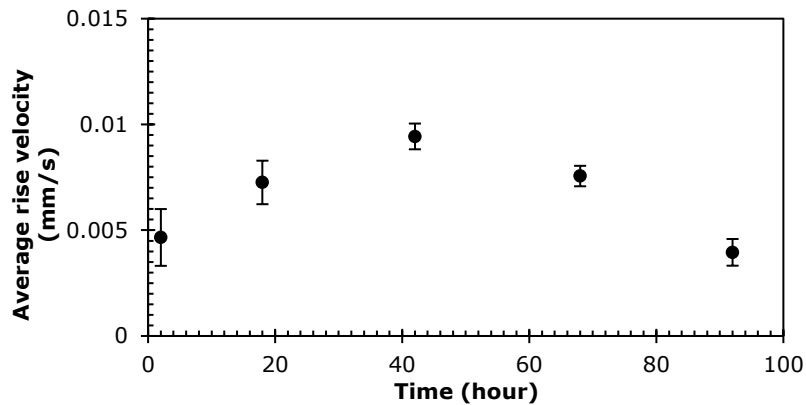


Figure 8.17: The averaged rise velocity of the small bubbles at the top section of a column of 330 Pa.s Silicone oil, 3.3 m liquid height, and 290 mm column diameter.

A slow recirculation in the oil could be driven by the difference in density of the oil. The temperature of the oil in the column changes due to the change in the ambient temperature that occurs through the 24 hours of the day. This variation in the oil temperature causes a change in the oil density in the column. The circulation velocity of the oil, which is taken as the intercept of the linear regression fit, is plotted versus the maximum temperature for each day in Figure 8.18. The circulation velocity increased with increasing the temperature.

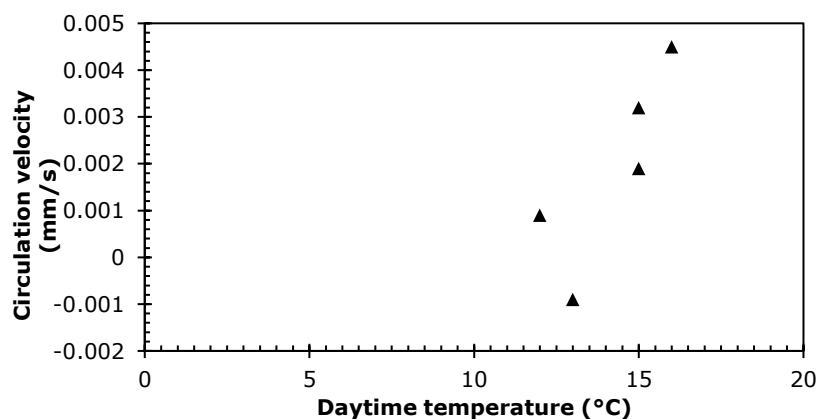


Figure 8.18: Effect of maximum daytime temperature on circulation velocity of 330 Pa.s stagnant Silicone oil.

Figure 8.19 shows the rise velocity of the single small bubbles increasing proportionally with increasing the diameters of the bubbles at different times. This behaviour is due to the effect of buoyancy force which is mainly

affected by the diameter of the bubbles. In general, the rise of bubbles in the viscous Silicone oil follows Stokes law. After all, the Reynolds numbers are $\ll 1$. If the measured data is plotted as rising velocity against diameter squared, they might be expected to lie on a straight line of slope $\rho_l g / 18 \mu_l$, the red line in the figure. The experimental data shows a significant discrepancy in this case. However, if a slow recirculation of the liquid is postulate, assuming a drift velocity of 0.006 mm/s, then the expected measured bubble rise velocities should lie on the dotted line. A better agreement with the experimental data could be obtained if the drift velocity is measured for every day considering the effect of temperature.

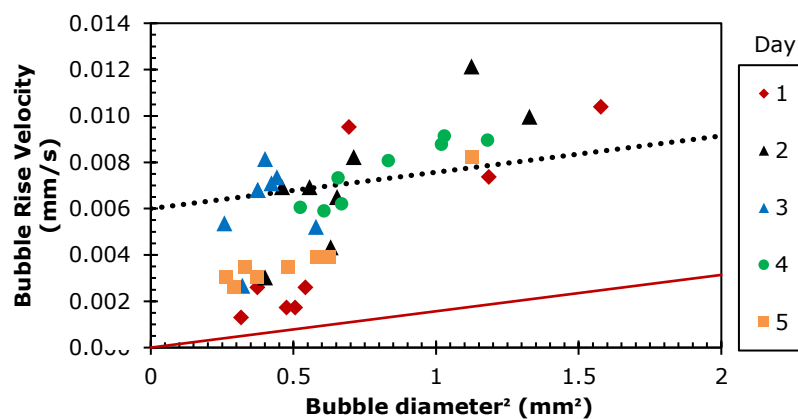


Figure 8.19: The effect of the bubbles diameter on the rise velocity of the small bubbles in stagnant 330 Pa.s Silicone oil.

8.5 Summary

In high viscosity liquids, the small bubbles are seen to accumulate in the column due to the low velocity of the liquid slug motion. The high viscosity of the liquid, besides the ineffective mixing of the bubble, are the reasons of trapping the bubbles for a long time inside the oil. The creation, mixing, and rise of the small bubbles was studied in 330 Pa.s Silicone oil and 290 mm diameter column. A high-resolution camera was used to study the characteristics of the small bubbles beside the ImageJ software used for the

image processing. Photos of the bubbles were obtained at the top and the bottom of the column during the generation and rise stages of the bubbles.

The characteristics of the small bubbles in high viscosity liquids and large diameter columns can be summarised in:

1. The small bubbles were seen in two main **shapes**; spherical in stagnant liquid and ellipsoidal in during the gas injection. The spherical bubbles became elongated due to the shearing by the liquid motion.
2. The **size** of the bubbles appeared to vary according to the gas flow rate, time of gas injection, and the location of the bubbles generation. The size of the small bubbles, generated at the top section of the column, was seen to be relatively smaller than the other parts of the column due to the mechanism of the bubble generation. The diameter of the small bubbles appeared between about 0.1 to 5 mm during the bubble generation stage in the top section. The length of these bubbles varied from 4 cm to 1 cm and diameter of 2 cm to 0.2 cm during the gas injection. On the other hand, the bubbles were seen in spherical shape at the bubble rise stage when the gas injection stopped. The diameter of the rising bubbles ranged between 1.1 to 0.2 mm at the top section. However, they ranged between 3.8 to 0.4 mm at the bottom of the column. The size of the small bubbles increased with increasing the gas injection time due to the increasing the population of the small bubbles. Increasing the population of the bubble, in turn, enhances the coalescence, which occurs due to the shearing by the liquid motion.
3. The small bubbles **generated** from the eruption of the large bubbles at the top surface, the gas injection points at the bottom of the column, the coalescence of the bubbles, and liquid bridge at churn flow regime. They accumulate in the column due to the low rise velocity of the liquid

slug, which contains the small bubbles as a reason to the very high viscosity of the oil. In bubbly and slug flow, the small bubbles are created from the **bubble eruption** at the top surface of the liquid and trap air to create a ring of unstable bubbles. These temporary bubbles, in turn, create smaller bubbles due to the shearing by the liquid motion on the pipe wall in case of higher gas flow rates. The amount of the small bubbles, which generate in higher gas flow rates, is higher not only because of increasing the surface area of the liquid film but also due to increasing the frequency of the bubble eruption. The second source for the small bubbles is the **gas injection points** at the bottom of the column. The bubbles were created from the rapid break up or coalescence of the large bubbles at the gas nozzles. The large drag forces increase the possibility of the coalescence at the gas inlet points. However, the effect of the shearing of the pipe wall can be clearly seen when the nozzle next to the pipe wall is used. The rate of the breaking up of the large bubbles increased due to the friction created by the wall. Another source for the small bubbles generation is the **coalescence of the large bubble** when rising in the column. The small bubbles are produced mainly due to the entrainment of the two large bubbles of different rising velocity. It is also seen to be produced due to the effect of the wake when the bottom of the upper bubble becomes concave shape from rounded. When the two large bubbles merge, the bottom of the lower bubble remains as a thin liquid film until it collapses at the last stage of the bubble coalescence. Further, this liquid film also generates the small bubbles. Finally, the small bubbles are generated from the **churn areas** in churn flow regime. The high frequency and unstable direction of the motion of the liquid bridge, in addition to the high

interaction between both fluids and the drainage of the liquid film, all these factors control the small bubble generation.

4. The **direction of the flow** of the small bubbles varied according to the location of the bubbles. They were initially seen flowing down with the falling film at the pipe wall around the rising large bubbles at bubbly and slug flow regime. Then, flowing upward with the liquid slugs between the large bubbles at bubbly and slug flow regimes. Also, they could be seen at churn flow regime in the draining film around the open gas core. And finally, oscillating up and down in the thick film and the liquid bridge in churn areas at churn flow regime.
5. The **concentration** of the bubbles is seen to be affected by gas flow rate, time of gas injection, number and location of the gas nozzles, and finally the location of the bubble generation.

Chapter 9

Conclusions and Recommendations

9.1 Conclusions

The characteristics of gas-high viscosity liquids in large diameter columns have been studied over a wide range of gas flow rates using ECT. Stagnant Silicone oil with viscosity of 360 Pa.s and air with a range of 0.005-0.314 m/s gas superficial velocity were employed in 240 mm diameter column. In addition, gas superficial velocities between 0.003 and 0.566 were injected into a 290 mm diameter column that contains 330 Pa.s Silicone oil. The gas injection system in these two rigs was designed to allow a full control of the number and the position of the gas injection points. The experiments, which were carried out in the main laboratory of the University of Nottingham to study the characteristics of gas flow in high viscosity liquids, were divided into: 1) the effect of gas flow rate, 2) The effect of the gas injection methods, 3) the effect of time, and 4) the small bubbles experiments.

In general, 4 flow patterns appeared in the 290 mm diameter column over the selected range of the gas flow rates. These flow regimes were bubbly, slug flow, transition to churn and churn flow regimes. However, in the smaller diameter column, bubbly, slug flow and transition to churn flow regimes appeared, as a smaller range of gas flow rate was employed. The main finding of this work can be summed up in:

9.1.1 Bubbly and Slug Flow Regimes:

Bubbly and slug flow in high viscosity liquids show completely different behaviours compared with low viscosity liquids. First, bubbly flow consists of a single large bubble rise at a constant velocity at the centre of the column. These bubbles are surrounded by a thick falling film rise at low gas flow rates with low frequency and velocity. They consist of 1) Spherical shape bubbles, at gas superficial velocities of 0.005 and 0.003 m/s in the 240 and 290 mm diameter columns respectively. And 2) Ellipsoidal bubbles, at gas superficial velocities of 0.011 m/s in the small column and 0.009 and 0.016 m/s in the 290 mm diameter column.

Second, slug flow in viscous oils consists of long bubbles (Taylor bubbles) with rounded nose and end, which occupy the greater part of the cross section of the column, and surrounded by a falling liquid film. The Taylor bubbles are followed by liquid slugs which contain small bubbles of sizes ranging millimetres/centimetres. Taylor bubbles are long cylindrical bubbles with rounded top and end, which appear at gas superficial velocities ranges of 0.016-0.114 m/s and 0.022-0.088 m/s in the 240 and 290mm diameter columns respectively. Slug flow regime appears at lower gas flow rate in the smaller diameter column and at a higher gas flow rate in the larger diameter column.

From data of time series of void fraction, length, velocity and frequency of the large bubbles increase significantly with increasing the gas flow rate in both columns. On the other hand, the liquid film thickness around the bubbles and the liquid slug length are seen to decrease with increasing gas flow rate. The coalescence between the large bubbles is also increased with increasing the gas flow rate until very long bubbles are created by the end of slug flow regime. The bubble coalescence at different gas flow rates shows

a similar mechanism. The larger bubble, which rises faster due to the difference between the buoyancy and the drag force, merges with the upper “smaller” bubble to create a new bubble with a size equivalent to the size of both bubbles. The top surface of the viscous oil in both columns is observed to rise and drop periodically in regular time and height as the bubbles rise with regular size and constant velocity. The Comparison of the experimental results obtained from the ECT, with the high resolution and high-speed cameras, shows agreement in terms of bubble diameter and length. Likewise, comparing the ECT results for bubble diameter, Taylor bubble, liquid slug length and large bubble velocity with the theoretical approaches by Gaddis and Vogelpohl (1986), Khatib and Richardson (1984) and Viana et al. (2003) also shows agreement. In addition, validating the equation by Gaddis and Vogelpohl (1986) for 330 Pa.s Silicone oil which shows good results.

Increasing the gas flow rate increases the void fraction and decreases the pressure gradient which reaches its minimum at the start of the transition to churn flow. Therefore, in bubbly flow, which consists of bubbles with low void fraction (thicker liquid film), the pressure gradients are relatively higher. Increasing the gas flow rate increases the void fraction and decrease pressure gradient due to the large Taylor bubbles which almost occupy the pipe cross-sectional area. Increasing the frequency of the bubbles and decreasing the liquid slug length are also considered factors that decrease the pressure gradient.

9.1.2 Transition to Churn and Churn Flow Regimes

The characteristics of gas flow in high viscosity liquids in transition to churn and churn flow regimes also show different behaviour compared with the low viscosity liquids, which have been reported in the literature. The transition

to churn flow regime appears at a range of gas superficial velocities 0.127-0.314 m/s and 0.108-0.243 m/s in the 240 and 290 mm diameter columns respectively. Churn flow regime appears at gas superficial velocities of 0.276-0.566 m/s in the 290 mm diameter column. The transition to churn flow appears at lower gas flow rate in the larger diameter column. Similarly, churn flow appears at lower gas flow rate in the larger diameter column.

The transition to churn flow regime starts when the length and the frequency of the regions of the high-frequency activity, which appeared at the end of slug flow, increase. These regions of high-frequency activities are observed as liquid bridges, flowing up and stagnant. The void fraction of these regions is lower in Taylor bubbles. The source of the liquid bridge is from the drained falling film around Taylor bubbles. With increasing gas flow rate, the length of Taylor bubbles increases producing a larger volume of falling film, which, in turn, lead to increase the length of the churn regions. The high frequency liquid bridges are named "churn". The top surface is seen to fluctuate in height and time of bubble bursting as the bubbles become very long and vary in lengths and time of bubble bursting. With increasing flow rate, where the transition to churn starts to appear, pressure gradient starts to increase and void fraction to decrease due to the increase of the liquid film thickness around the air conduit in the column. Structure velocity, large bubbles length and liquid film thickness increase gradually with increasing the gas flow rate. The liquid film flow shows an average of downward flow at the large bubbles area and upward flow at churn regions. The direction of the flow fluctuates between upward and upward in transition to churn flow regime.

In churn flow regime, the structure of the flow changes significantly. The long bubbles disappear and the churn regions dominate the flow structure. Churn flow in 290 mm diameter column appears with two patterns. First, an

open gas core of thick liquid film with the possibility of appearing liquid bridges to create series of smaller bubbles. Then with continuous gas injection, the gas builds up and pushes the liquid creating a very long bubble with a falling thin film. This long bubble bursts at the top section of the column and remain open to atmosphere. Then the falling film, which drains on the column wall, accumulates at the bottom of the column and creates the next churn region. The behaviour of the top surface at this flow regime changes significantly. The air flows at the core of the oil column with a wavy passage, leaving the top surface open to the atmosphere with the possibility of creating a very long bubble. With increasing the gas flow rate, pressure gradient increases gradually until the end of the churn flow region where the void fraction is lower than the values at slug flow regime. No considerable increase is seen in structure velocity as the flow structure in this flow regime is totally different. The liquid film flow shows an average of downward flow at the large bubbles area and upward flow at churn regions. The direction of the flow fluctuates between up and liquid hold.

9.1.3 Gas Injection Methods

The effect of gas injection methods on the structure of the flow in 360 and 330 Pa.s Silicone oil in 240 and 290 mm columns respectively can be concluded in:

9.1.3.1 Position of the Air Injection

In terms of changing the injection positions, the mean void fraction, structure velocity, frequency and film thickness do not show any significant variation in values. In contrast, the Taylor bubble and liquid slug lengths are seen to change. Second, the structure of the flow in the smaller diameter column (240 mm) seems to be more affected by changing the position of the nozzle than the larger column. Finally, the flow in the smaller diameter

column appears more structured than the larger one for the same gas superficial velocity.

9.1.3.2 Number of the Gas Injection Points

In the 240 mm diameter column, mean void fraction, bubble velocity and length, and liquid slug length are higher when using 5 gas injection points, while the frequency appears to be higher when using a single nozzle. However, in 290 mm diameter column, no significant changes appear except in frequency. It is higher when 5 gas nozzles are used to inject air into the larger column. On the other hand, changing the number of gas injection points seems not to have an effect on the film thickness around the large bubbles.

9.1.3.3 The Distance Between the Air Injection Points

Mean void fraction, bubble velocity, film thickness, and bubble frequency do not show any significant variation in values. However, the Taylor bubble and liquid slug lengths are seen to change.

9.1.3.4 Gas Injection Time

The amount of the small bubbles in the column increases significantly with time, even at a constant gas flow rate. Thus, the density of the small bubbles in viscous liquid is not only proportional to gas flow rate but also to the time of gas injection. The mean void fraction increases slightly after a few hours of gas injection due to increasing the amount of the small bubbles in the column.

9.1.4 The Small Bubbles

In high viscosity liquids, small bubbles are seen to accumulate in the column due to the low velocity of the liquid slug motion. The high viscosity of the liquid and the ineffective mixing of the bubble are the reasons of trapping

the bubbles for a long time inside the oil. The creation, mixing, and rise of the small bubbles have been studied in 330 Pa.s Silicone oil and 290 mm diameter column. The small bubbles are seen in two main shapes; spherical in stagnant liquid and ellipsoidal during the gas injection. The spherical bubbles become elongated due to the shearing by the liquid motion. The size of the bubbles varies according to the gas flow rate, time of gas injection and to the location of the bubbles generation. The diameter of the small bubbles appears between about 0.1 to 5 mm during the bubble generation stage in the top section. The length of these bubbles varies from 4 cm to 1 cm and diameter of 2 cm to 0.2 cm during the gas injection. On the other hand, the bubbles are seen in spherical shape at the bubble rise stage when the gas injection stopped. The diameter of the rising bubbles ranges between 1.1 to 0.2 mm at the top section, while they range between 3.8 to 0.4 mm at the bottom of the column.

The small bubbles generate from the eruption of the large bubbles at the top surface, the gas injection points at the bottom of the column, the coalescence of the bubbles, and liquid bridges at churn flow regime. They accumulate in the column due to the low rise velocity of the liquid slug, which contains the small bubbles because of the very high viscosity of the oil. The direction of the flow of the small bubbles varies according to the location of the bubbles. They flow down with the falling film at the pipe wall around the rising large bubbles at bubbly and slug flow regime. And flow upward within the liquid slugs between the large bubbles at the same flow regimes. The concentration of the bubbles is affected by the gas flow rate, time of gas injection, number and location of the gas nozzles, and finally the location of the bubble generation.

9.2 Recommendations

As mentioned previously, the hydrodynamics characteristics of the gas flow in high viscosity liquids differ from the lower viscosity liquids. Besides, knowledge about the hydrodynamics of gas flow in high viscosity liquids and large diameter columns is still limited in spite of the importance of this area. Therefore, more work can be undertaken to investigate the flow of gas-high viscosity liquids. The following are recommended for further work:

1. Study the effect of changing the parameter in Kaptitza number, $Ka = (\frac{\rho}{g\sigma^3})^{0.25}$ and Ohnesorge number, $Oh = \frac{\mu}{\sqrt{D\rho\sigma}}$ on the structure of the flow by using different liquids with different viscosities or surface tension.
2. Using higher gas flow rates to investigate annular flow in such high viscosity liquids.
3. Increase the liquid height to study the effect of the pressure head as well as flow structure throughout the long viscous liquid column. 2 ECTs might be used at different locations along the column.
4. Study the hydrodynamics of the bubble generation at the gas injection points by applying the visualisation technique at the gas injection section.
5. Study the hydrodynamics of the bubble eruption at the top section of the column in more detail, e.g., film thickness, using high-speed camera.
6. Study the mechanisms of generation of the small bubbles at the top section of the column using high-speed camera.

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Appendices

Appendix A. Experimental Set up Photos

1. High Viscosity Oil Rig

Photos of the 240 mm diameter high viscosity oil rig showing the flow facilities and the ECT system are shown in Figure A.1. The figure also shows the top view (inner section) of the gas inlet section with the 25 gas nozzles. And finally, photo of a side view of the gas inlet section including the 25 gas inlet tubes is shown in the same figure.

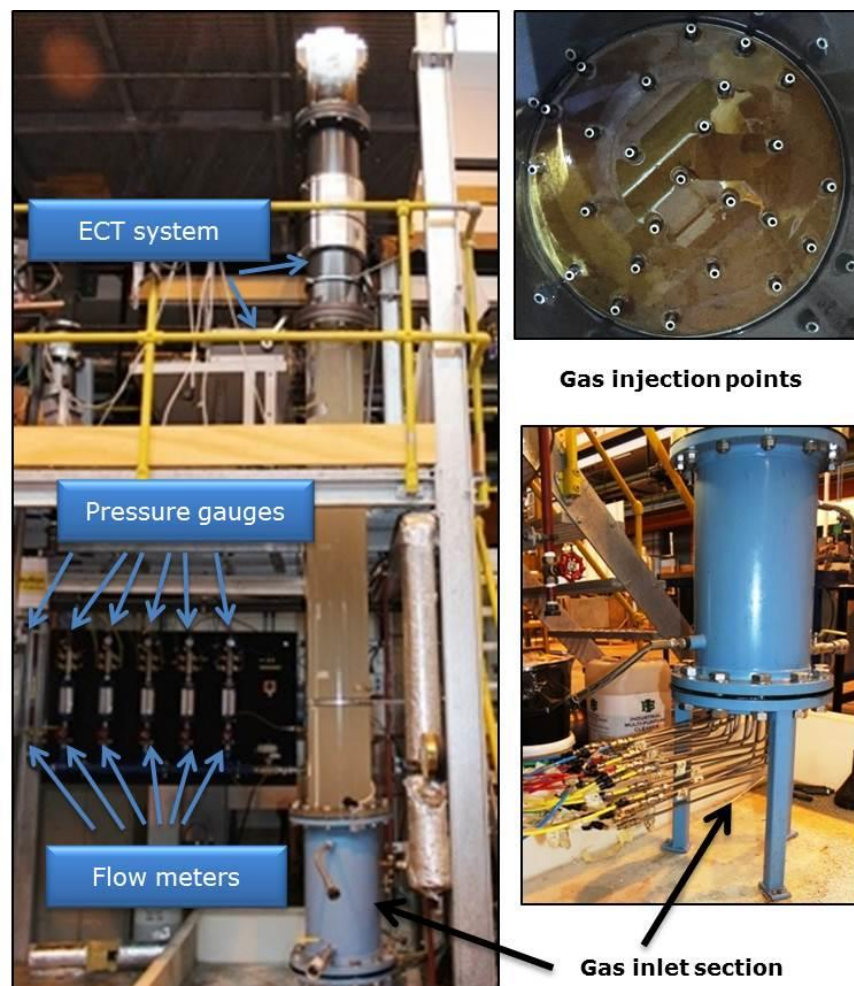


Figure A.1: Photos of the high viscosity oil rig showing the ECT, flow facilities, top view showing the 25 gas nozzles, and the gas inlet section.

2. Flow Facilities and Measurements

Figure A.2 illustrates the configuration of the gas lines applied in the gas-high viscosity liquid experiments. The mine gas line is divided first to 5 lines connected to 5 rotameters and pressure gauges, each of which can be controlled separately. Each line of the 5 is divided to more 5 lines with separate valves. This allows a full control for the number and the location of the gas injection point which can be used for the gas injection inside the column. The 25 small tubes are numbered and coloured to correspond to the numbering system in the gas inlet section at the bottom of the column.

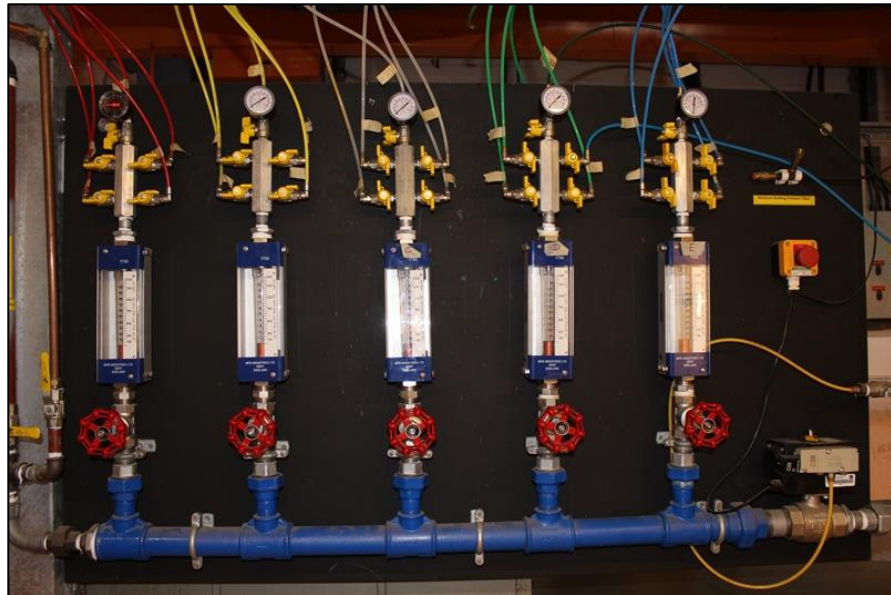


Figure A.2: Photos of the Rotameters and the pressure gauges in the 240 mm diameter column rigs showing the 25 gas lines.

3. Electrical Capacitance Tomography

The Electrical Capacitance Tomography System (ECT), which is used for the time series of void fraction measurements, consists of: an ECT (the sensor), an electronic box used for data acquisition, and finally a computer used for data processing and image reconstruction. This non-intrusive sensor measures the cross section distribution of capacitance in the column and which can be correlated to the cross section phase distribution for the non conductive liquids. Figure A.3 displays the ECT sensor and the TFLR 5000, ect 32V2, which is used for data acquisition in the 240 mm high viscosity oil column.

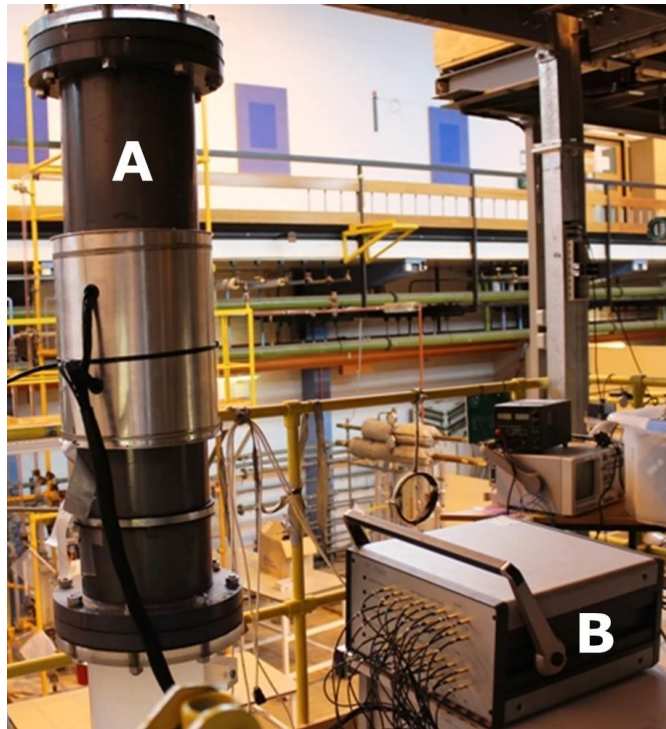


Figure A.3: ECT system connected to the 240 mm diameter column rig. A is the ECT sensor electrodes and B the electronic box.

4. Thermocouples Calibration

Temperature controller REX-D100, TC LTD was used to indicate the temperature of the air during the experiments. The temperature was used to correct the Rotameters' readings using a formula obtained from the manufacturer. The thermocouple was calibrated over a wide range of temperatures (0-100°C), which correspond to 1-5 V. Figure 4 shows the calibration setup (A) including the thermometer, the TENMA (72-7930.1) digital Multimeter, and the ice which used to obtain different range of temperatures. It also shows the thermocouple, which mounted on the rig to digitalise the temperature (B). Figure A.5 shows the results of the calibrations and best fitting line for the thermocouple calibration.

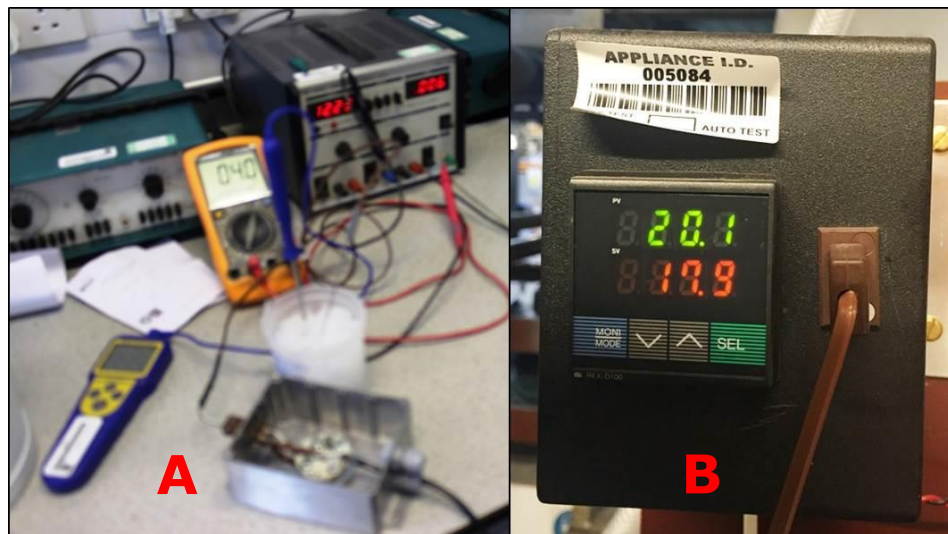


Figure A.4: A is the thermocouple calibration setup. B is the temperature controller REX-D100, TC LTD which employed for the temperature measurements.

Figure A.5 shows the results of the calibrations and best fitting line for the thermocouple calibration.

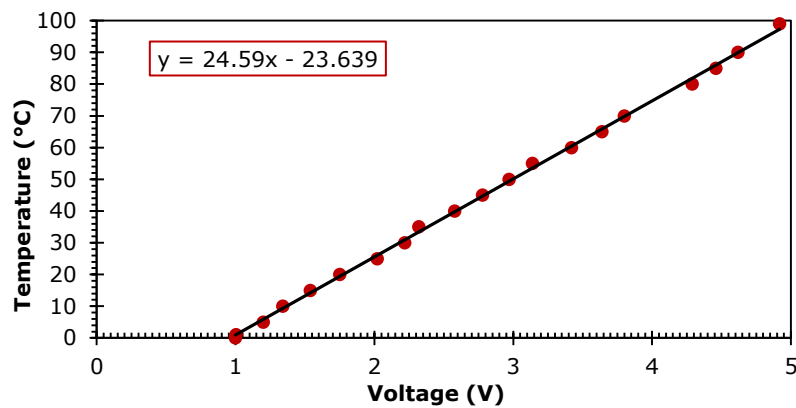


Figure A.5: The calibration line for the REX-D100, TC LTD temperature controller which used in the experiments.

5. Pressure Sensor Calibration

The pressure transmitters employed for the pressure gradients measurements in the high viscosity oil experiments, were calibrated before using them for the measurements. The Keller, PR-25/8797.1-5 transmitters, which are installed on the 290 mm diameter column, were calibrated within a range of 0-10 V. PV211 - pneumatic pressure and vacuum pump were used with TENMA (72-7930.1) digital Multimeter for this purpose. A number of selected pressures were applied by the pump to the pressure transmitters then with the resulted voltage from the Multimeter for each pressure the calibration curve was plotted. Figure A.6 displays the bench calibration setup.

The calibration equations were applied to correct the output voltage signals, which were converted to pressure (bar) using LabView software. Figure A.7 displays the calibration lines and the equations used to correct the signals obtained from the transmitters.



Figure A.6: The pressure sensors bench calibration setup. A is the pressure transmitters, B is the PV211-pneumatic pressure and vacuum pump, and C is the digital Multimeter.

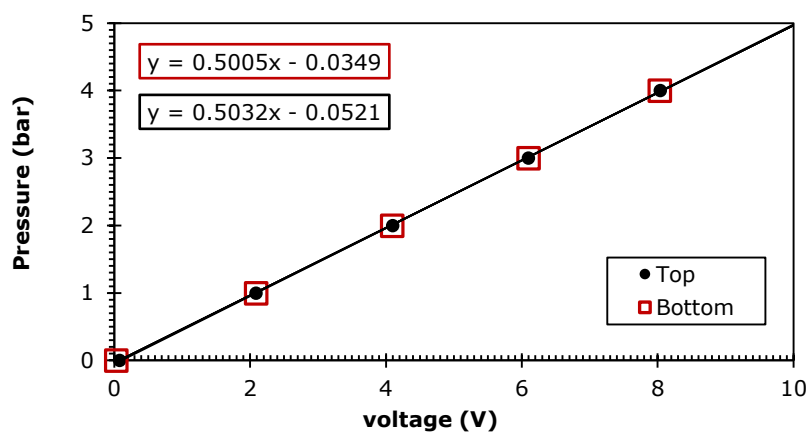


Figure A.7: The calibration lines of the pressure sensors which are installed on the 290 mm diameter column. The bottom and the top transmitters are installed at about 2.2 m and 3.6 m from the gas inlet points respectively.

6. High Resolution Visualization Technique

The Dantec Dynamics high speed camera (Vision research, Phantom V12:20-30 VDC), EX SIGMA 105mm 1:2.8 DG MACRO is shown in Figure A.8. This camera was used to determine the velocity and the size of the large bubbles rising in the 290 mm diameter column for different gas flow rates. The figure also shows the software used for data processing on the PC. A large spherical bubble, rising at flow rate of 10 L/min in the 330 Pa.s Silicone oil in 290 mm diameter column, is captured by the camera. The collected photos from this camera were analysed using ImageJ software. This software is able to determine the dimensions of the bubbles after setting up the calibration values for the images. The images were calibrated using a measuring tape placed on the columns, see the figure in Section 3.4.6.

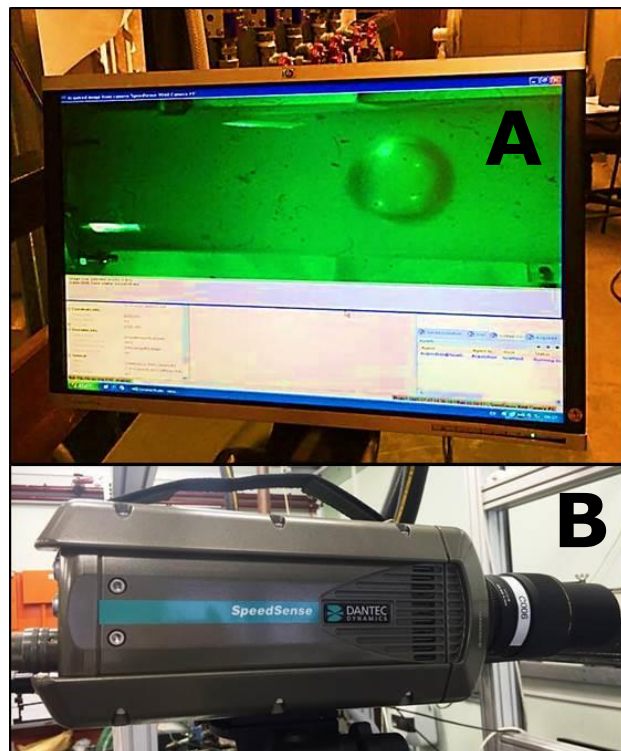


Figure A.8: A is the data processing software showing one bubble rising in 330 Pa.s Silicone oil and 290 mm diameter column. B is the Dantec Dynamics high speed camera.

7. Pressure Gradient Experiments

Figure A.9 shows the LabViwe block diagram used to convert the voltage signals from the transmitters, which were located on the column to pressure in bas. The block consists of a simple das assistant tool to convert the signal, waveform charts to display the signals from the 2 pressure transmitters. It also contains tools for data recording, saving, and data acquisition timing. More information about the experiment is shown in Section 3.4.3 in Chapter3.

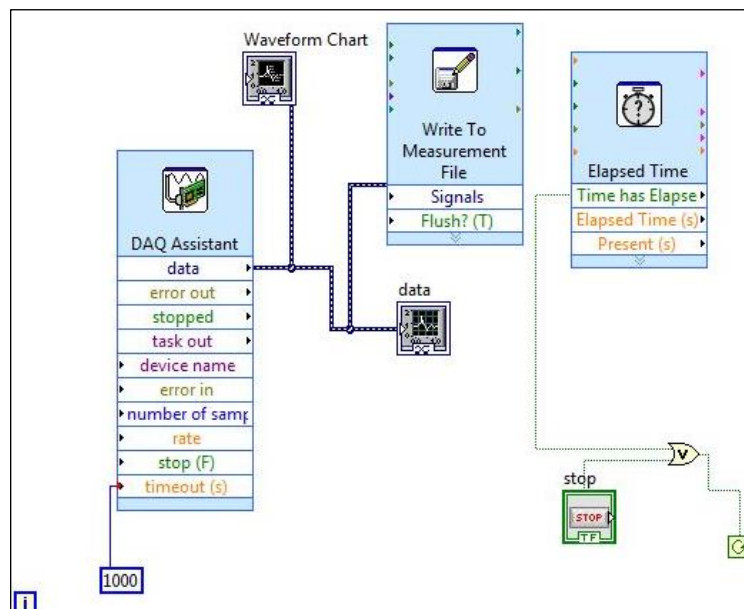


Figure A.9: The LabViwe block diagram which used to convert the voltage signals from the transmitters to pressure in bar units.

Appendix B. Uncertainty Analysis

The uncertainty analysis is used to estimate the limits of measurements error and to define the quality of the acquired experimental data, Holman (2002). The error was defined by Kirkup and Frenkel (2006) as the difference between the true and the measured values. In general, the errors can be divided to; systematic and random errors. Systematic errors are the difference between the true values and the average measured values. Random errors are a result of random fluctuations from the measurement system which employed in the experiments.

1. Systematic Error

In systematic errors, the measured values are consistently higher or lower than the true values. They caused by the limitation in measuring equipment and some physical effects. These errors appear when the experimental conditions change. Since the systematic errors remain consistent for the measurements, therefore, the experimental data cannot be used for this type of uncertainty. In the pressure and temperature measurements the systematic errors were determined from the calibration errors. The systematic errors for the void fraction could not be estimated. The gas superficial velocity, U_{sg} which is equal to the volumetric gas flow rate divided by the pipe cross section area, is obtained from the flow meter readings. Two types of gas flow meters with different numberings scales were used in the experiments. One of 2 L/min used for the low pressure line and other of 50 L/min connected to the high pressure gas line. The uncertainty in reading the gas flow rate is thus, is equal to 0.5 of the least count. This gives an error values of $U_{sg} = \pm 0.0063$ m/s and ± 0.00025 m/s for the high and the low pressure gas lines respectively.

2. Random Error

When the experiments are repeated using the same experimental setup then the difference between the random errors and the systematic errors is best observed. Therefore, repeating the experiment allows checking the errors by using the reproducible results. The standard deviation of the data population for N numbers points is calculated from the following equation:

$$\sigma_x = \sqrt{\frac{\sum_{i=1}^N (X_i - \bar{X})^2}{N - 1}} \quad B.1$$

Where: σ_x is the standard deviation of N data points.

The relative standard deviation about the mean, $\sigma_{\bar{x}}$ can be calculated from the scatter of the mean values from the following equation:

$$\sigma_{\bar{x}} = \frac{\sigma_x}{\sqrt{N}} \quad B.2$$

The relative standard deviation or the standard error is shown as error bars for all measured parameters in the result chapters. The standard error for the gas superficial velocity ranges between 0.00017% to 0.8% at the minimum and the maximum flow rates respectively in the 290 mm diameter column. While, it is vary between 0.019% and 0.41% at the minimum and maximum gas superficial velocities respectively in the 240 mm diameter column.