Pergamon

0301-9322(94)00067-0

Int. J. Multiphase Flow Vol. 21, No. 2, pp. 253–266, 1995

0301-9322/95 $9.50 + 0.00

EFFECTS OF THE LIQUID VISCOSITY ON THE PHASE DISTRIBUTIONS IN HORIZONTAL GAS–LIQUID SLUG FLOW

M. NÄDLER and D. MEWES
Institut für Verfahrenstechnik der Universität Hannover, Callinstrasse 36, D-30167 Hannover, Germany

(Received 23 May 1994; in revised form 25 September 1994)

Abstract—In the present paper, experimental results of the investigations of the effect of liquid viscosity on the phase distribution in two-phase air–liquid slug flow in horizontal pipes are represented. As the liquid phases, oil with viscosities in the range from 14 to 37 mPas and water are used. The experimental investigations are conducted in a horizontal pipe with an inner diameter of 59 mm. The superficial liquid viscosities and the superficial air velocities in the test section are varied in a range from 0.15 to 1.5 m/s and from 0.1 to 13.5 m/s, respectively. The liquid hold-up within the liquid slug zone, the liquid holdup in the elongated bubble zone and the flow averaged liquid holdup are measured by means of a multibeam γ-ray densitometer. The results indicate that there are significant differences in the distribution of liquid and gas when oil and air or water and air are used in the experiments. By increasing the liquid viscosity, increasing values for the flow averaged liquid holdup and the volume averaged liquid holdup inside the liquid slug and the elongated bubble zone are obtained.

Key Words: gas–liquid pipe flow, slug flow, liquid distribution, gas entrainment, liquid viscosity, γ-ray densitometer

1. INTRODUCTION

Intermittent flow of gases and liquids occurs over a wide range of volumetric flow rates in pipelines for multiphase flow transport. In particular, in the oil industry there are many pipelines operating under intermittent flow conditions. In order to design the transport and production systems, a knowledge of the effect of the fluid properties on the flow regime is required. The most common type of flow regime observed in oil production pipelines is the slug flow which is schematically shown in figure 1. In the slug flow regime, liquid slugs filling the whole cross-sectional area of the pipe are flowing where these slugs are separated by elongated gas bubbles.

Due to the momentum of the slugs, mechanical forces and vibrations can occur in the pipe system. Approaching the eigenfrequency of the pipe system, the slug frequency can cause resonance and may ultimately lead to damage. The knowledge of the liquid distribution and hence the void fraction within the liquid slug is important for the design of slug catchers for pipelines. Furthermore, input data for the void fraction within the liquid slug are required in available models as published by Dukler & Hubbard (1975), Nicholson *et al.* (1978), Kokal & Stanislav (1989), Taitel & Barnea (1990) or Andreussi *et al.* (1993) which permit the prediction of the pressure drop. The pressure drop and the characteristics of the slug flow are influenced by the fluid and rheological properties of the phases flowing in the multiphase pipeline. The knowledge of their effect on the slug characteristics is required for the exact design of multiphase pipeline systems. There are numerous investigations where the slug characteristics are measured. Most of these experiments have been carried out using water as the liquid phase and air as the gas phase. There are other investigations which have been carried out using air as the gas phase and oil as the liquid phase. However, only a few investigations have been made showing the effect of liquid viscosity on the characteristics and hence the phase distribution in horizontal gas–liquid slug flow.

Investigations of the influence of liquid viscosity on slug characteristics in horizontal, slightly inclined pipes have been carried out by Crowley (1984) and Sam & Crowley (1986). Both studies were conducted within the same test facility. The internal diameter of the multiphase pipeline is 171 mm. The experimental data are obtained using water, glycerin–water solutions and water–polymer solutions as the liquid phase and air or Freon 12 as the gas phase. According to

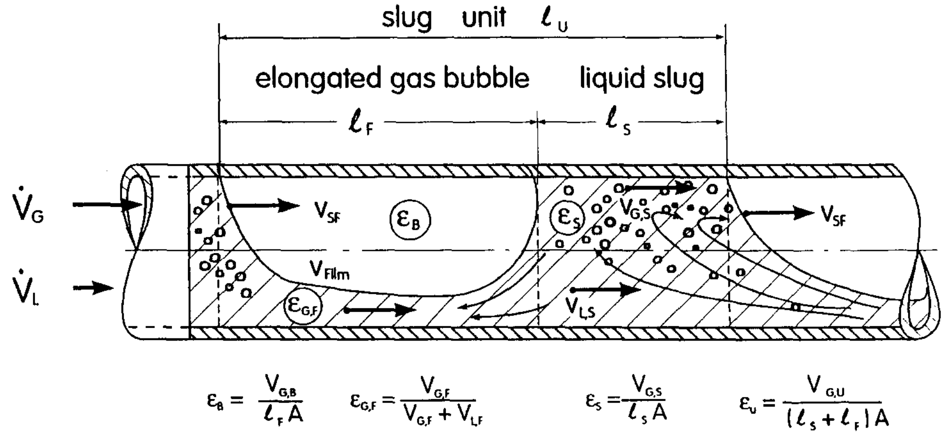

Figure 1. Simplified slug model.

the results of these investigations, increasing the liquid viscosity is expected to result in increasing slug velocities, increasing liquid film heights and increasing flow averaged liquid holdups.

Kago *et al.* (1986) investigated the effect of the liquid viscosity and surface tension on the flow averaged liquid holdup in a 51.5 mm inner diameter horizontal pipe. In these experiments results for the influence of liquid viscosity and surface tension were obtained using water, water–polymer solutions and slurries as the liquid phase and air as the gas phase. In the experimental investigations of the flow averaged void fraction, the viscosities of the liquid and the slurry were varied in the range from 0.8 mPas (water) to 55 mPas (slurry). The results of this study show that the flow averaged liquid hold-up increases by about 50–60% when the liquid viscosity increases by 45–55 times that of water, while, through adding surfactants to the water or the water–polymer solutions, no effect of the liquid surface tension on the void fraction was observed.

In the case of varying two fluid properties in an experimental investigation, e.g. the liquid viscosity and the liquid density by using water and glycerin–water solutions as the liquid phase, it is possible that no effect is observed because both fluid properties influence the slug characteristics and the two effects compensate each other. Otherwise a very strong effect may be observed due to the enhancement of two small effects.

Therefore, the objective of this paper is the experimental investigation of the effect of the liquid viscosity on the liquid and the gas distribution in slug flow in horizontal pipes with the other properties being kept constant.

2. EXPERIMENTAL SET-UP

In the experimental investigations the slug holdup distributions are measured. The measurements are conducted in the experimental set-up given in figure 2. The two-phase flows of oil and air and water and air are investigated by changing flow rates of the phases. The volumetric flow rates of the liquids are regulated by varying the number of revolutions of the centrifugal pumps. The gas phase is air which is supplied by a compressor network. The volumetric flow rates of all phases can be regulated independently and are measured by turbine flow meters and orifice gauges.

The experimental data are obtained using water and oil as the liquid phase. The oil, used as the second liquid phase, is an optically transparent mineral white oil (Shell Ondina 17) with Newtonian flow characteristics. The superficial liquid velocity and the superficial air velocity in the test section are varied in the range from $j_L = 0.15$ m/s to $j_L = 1.5$ m/s and from $j_G = 0.1$ m/s to $j_G = 13.5$ m/s, where the superficial velocities are the volumetric flow rates of liquid and the gas related to the cross-sectional area of the test section, respectively. For all test runs, the absolute pressure in the flow system is less than 0.5 MPa.

The viscosity of the liquids is measured by means of a Haake RV coaxial cylinder viscosimeter as well as a falling ball viscosimeter and the surface tension between the liquid phase and air is measured by means of a plate tensiometer.

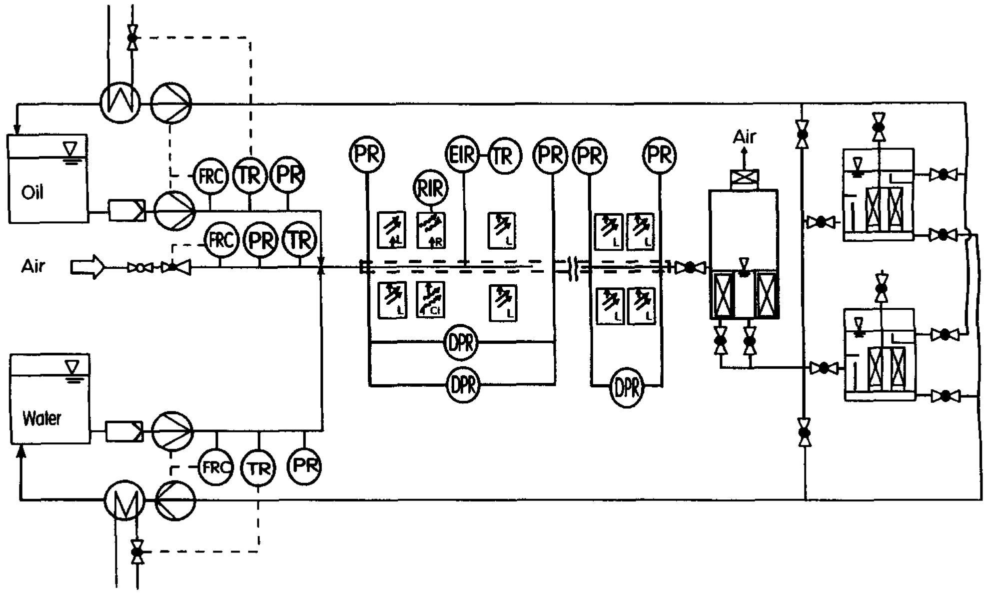

Figure 2. Experimental set-up.

In the experimental investigations, the oil viscosity is varied in the range from 14 to 37 mPas by changing the oil temperature by means of heat exchangers. For this viscosity range, the corresponding range for the densities of the pure oil is from 834 to 848 kg/m^3 and the corresponding range for the oil–air surface tension is from 29.6 to 31.2 mN/m. The change in the oil density is less than 2% and the change in surface tension is less than 3% so that an effect on the slug characteristics due to the change in oil density and surface tension is negligible in comparison to the changes in liquid viscosity.

Oil, water and gas are fed into the pipeline by a nozzle which is shown in figure 3. The nozzle is cone shaped. It has three sections separated by baffle plates. Each baffle plate consists of two parts, the second of which can be inclined upward or downward. This design has been selected to prevent the initiation of slug formation due to entrance effects, i.e. hydraulic jumps just behind the

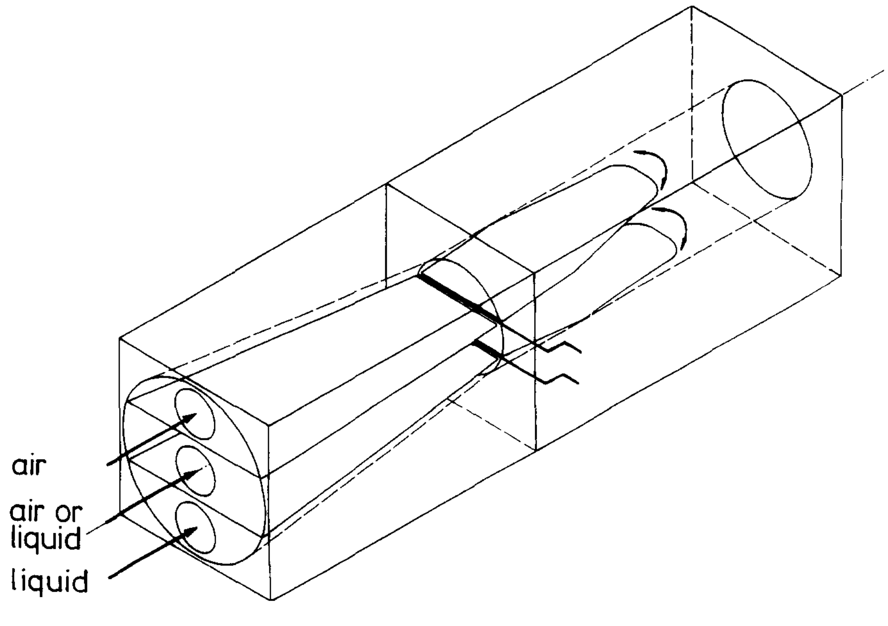

Figure 3. Schematic diagram of the entrance nozzle.

nozzle. The design of the nozzle used allows the investigation of several entrance situations and hence of different slug formation mechanisms.

By means of two baffle plates, the phases are fed into the pipeline in layers corresponding to their density. The pipeline is manufactured in acrylic glass with an internal diameter of 59.0 mm. The total length of the multiphase flow pipeline between the entrance nozzle and the separation unit is approximately 48 m. The multiphase pipeline consists of two legs connected by a U-bend with leg lengths of 25 and 23 m, respectively. The measurements of the present study are taken in the first leg behind an entrance length of the pipe of approximately 12 m.

A multi beam γ-ray densitometer is used to measure the liquid holdup in the slug and film zone and hence the flow averaged liquid holdup. The densitometer is shown schematically in figure 4. The beams are collimated and are traversing the pipe cross-section where the multiphase mixture is flowing.

For the measurements of the present study three beams are used. In order to minimize the error in the measurements of the liquid distributions an extensive test programme has been carried out. In these tests it is confirmed that the error is minimal if all three beams are located in one half of the pipe cross-sectional area and the angle between the longest beam and the vertical is 15 degrees. For this arrangement the three collimated beams are traversing nearly 70% of the selected half of the pipe cross-sectional area. The activity of the caesium-137 source used in this densitometer is 4 Ci. The period of measurement is 100 s for each test run at a sampling frequency of 1 kHz. By means of the high activity of the caesium-source and a high speed scintillation detector system a statistical error less 1% is obtained. A description of the scintillation detector system used for the measurements is given by Fortescue *et al.* (1994). In static and dynamic tests conducted with several plastic probes it is confirmed that the absolute error of the pipe cross-sectional averaged holdup measured by means of the γ-ray densitometer is less than 3%.

In order to obtain reliable mean values for the slug characteristics, the test run is repeated until a minimum number of 20 observed slugs is exceeded.

In order to be able to distinguish between the liquid slug zone and the film zone of the slug unit, the slug front as well as the end of the slug body has to be detected as accurately as possible.

One common method used in order to detect the slug front and hence the beginning of the liquid slug zone is based on determining rising flanks in the time trace of pressure transducers or in the

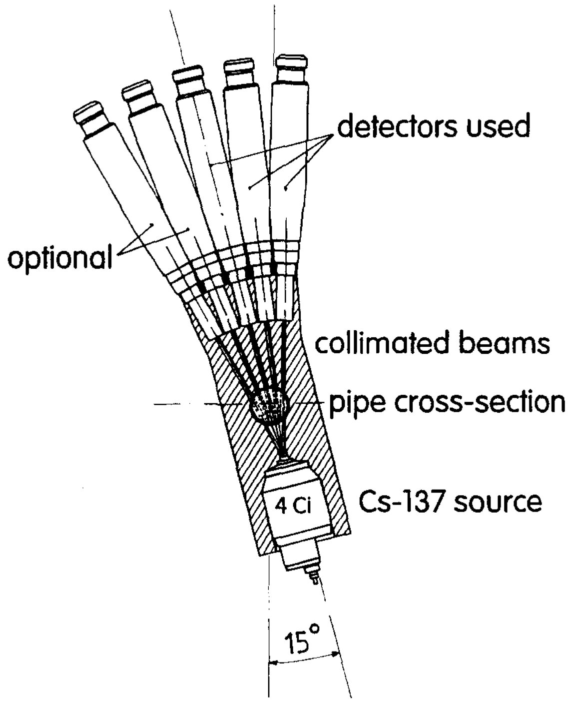

Figure 4. Schematic diagram of the γ-ray densitometer.

time trace of the liquid hold signal. It is confirmed by the present study that this method can fail. Particularly in the case of highly aerated mixing zones behind the slug front, it is not possible to detect the slug front only by means of rising flanks of the liquid holdup trace.

In case of waves flowing at the end of the film zone and hence just before the slug front it is possible that the liquid holdup in the pipe cross-sectional area containing these waves is of the same order as the liquid holdup within a highly aerated mixing zone (figure 1). Thus it is possible that the wave front or the end of the mixing zone is detected as a slug front. In the first case, the average liquid holdup within the slug zone is underpredicted. In the second case, only the high liquid holdup within the core of the slug body is detected. Hence the averaged liquid holdup is overpredicted.

Therefore, in addition to the holdup trace, liquid level sensors are used to detect the slug front. Each sensor consists of a LED light source and a receiver photodiode. The light beam of the LED light source traverses the test section horizontally in its upper third portion and illuminates the photodiode when the air passes through the transilluminated part of the tube. Due to the different refractive index of the gas and the liquid, the light path is traversed the very moment liquid passes through the test section so that the current of the photodiodes changes. Particularly in the case of highly aerated mixing zones, most light is scattered or reflected at the gas–liquid interface within the mixing zone so that there is an abrupt change in the voltage signal of the measurement circuit when the mixing zone passes the transilluminated part of the tube.

The end of the liquid slug zone is detected by means of the falling flanks in the liquid holdup trace. As mentioned by Nydal *et al.* (1992), in the intermittent flow regime regular slugs are to be distinguished from developing slugs which are highly aerated and shorter than regular slugs. In the present study, the developing slugs are identified by means of their length and the highly aerated core zone within these liquid slugs.

3. INTERMITTENT FLOW REGIMES

In the intermittent flow region, three flow regimes exist. These are distinguished by the variation of the liquid hold-up in the pipe cross-sectional area as shown in a previous paper (Nädler & Mewes 1992). These are the plug flow regime, the aerated slug flow regime and the regime of foam slugs.

In figure 5 the variation of the liquid hold-up measured by the γ-ray densitometer is shown. Figure 5 depicts the change in the variation of the oil holdup for oil–air slug flow for an increasing air velocity. The superficial oil velocity is constant $j_{L2} = 0.9$ m/s while the superficial air velocity is increased from $j_G = 0.6$ m/s in figure 5(a) to $j_G = 8.95$ m/s in figure 5(c).

The intermittent flow at the transition from the plug flow regime to the aerated slug flow regime is shown in figure 5(a). The plug flow is characterized by nearly unaerated liquid plugs and elongated gas bubbles. These bubbles are flowing in the upper section of the pipe under the top of the pipe. At the transition from the plug to the aerated slug flow regime, gas entrainment into the slugs occurs and the liquid plugs become aerated as shown in figure 5(a). The length of the elongated bubble is of the same order of magnitude as the liquid plug length.

By increasing the superficial air velocity, the gas entrainment into the liquid plug increases and the aerated slug flow regime will be observed [figure 5(b)]. Due to the gas entrainment into the slug, the liquid hold-up behind the slug front decreases and the void fraction within the liquid slug increases. Due to the lower density of the gas, the bubbles in the liquid slug rise along the slug zone. At the end of the slug zone the bubbles coalesce into the elongated gas bubble in the film zone of the slug unit. The average void fraction in the liquid slug increases with an increasing gas velocity to up to 52% which is the most compact arrangement of equally sized bubbles in the case of a cubical packing. The height of the liquid film is between 20 and 40% of the inner pipe diameter. The ratio between the length of the aerated liquid slug and the length of the film zone of a slug unit is less than 0.5.

Further increase of the superficial gas velocity to $j_G = 8.95$ m/s results in increasing gas entrainment into the liquid slug as shown in figure 5(c). There is a large scatter in the length and aeration of the slugs. There are slugs, where the average liquid holdup in the slug zone is lower than 48% so that the void fraction is higher than 52%. In the case of equally sized spheres, these void fractions are only possible for a rhombohedral packing. In this flow area, the slug is formed by a foam plug characterized by void fractions above 52% which is sliding over a thin liquid film.

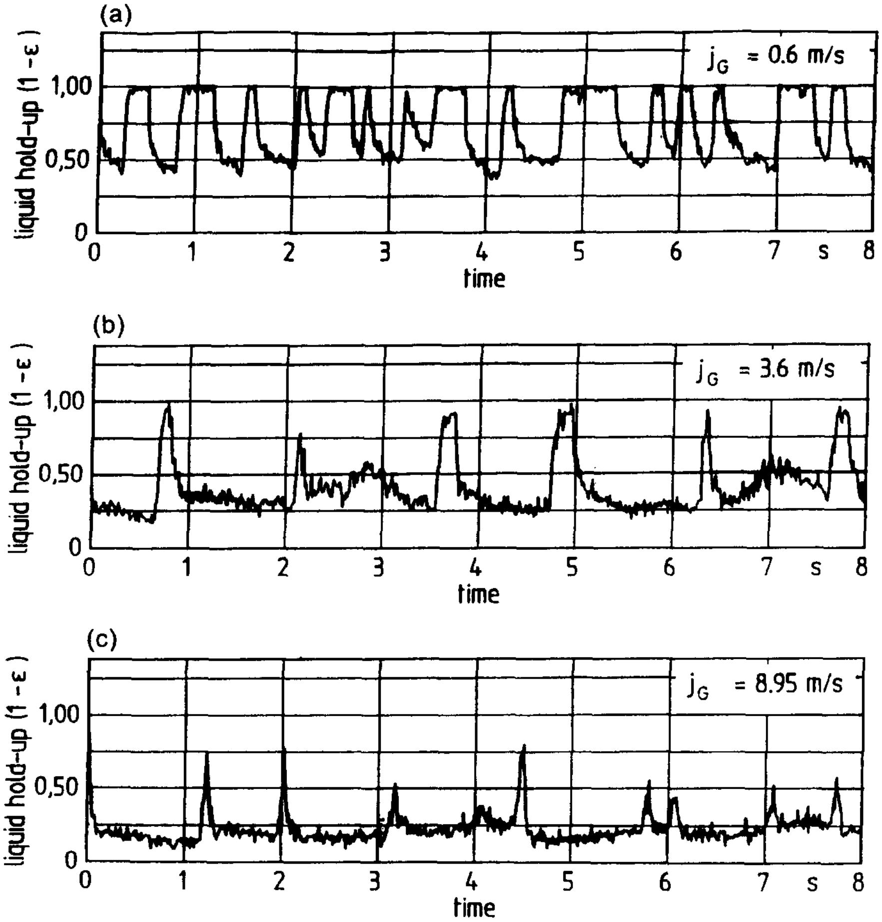

Figure 5. Variation of the liquid holdup measured by means of the γ-ray densitometer.

By means of visual investigations of the gas bubbles within the transparent oil it is confirmed that the void fraction in this liquid film increases with increasing slug velocity.

Probability density function of the liquid holdup trace

The probability density function of the area-averaged liquid holdup measured by the γ-densitometer is used to monitor the effect of the liquid viscosity on the liquid distribution in the slug flow.

The probability density function p of the liquid holdup $(1-\epsilon)$ is defined by

$$p\{(1-\epsilon)\} = \frac{dP\{(1-\epsilon)\}}{d(1-\epsilon)}, \qquad [1]$$

the derivative of the probability P with respect to the holdup $(1-\epsilon)$, where the quantity $p\{(1-\epsilon)\}$ $d(1-\epsilon)$ represents the probability that the holdup $(1-\epsilon)$ is between the values $(1-\epsilon)$ and $[(1-\epsilon)+d(1-\epsilon)]$. Liquid holdup increments of $\Delta(1-\epsilon)=0.01$ are used for estimating the probability.

In figure 6 the resulting probability density function of the holdup trace obtained for oil viscosities of $\eta_{L2} = 14$ mPas and $\eta_{L2} = 37$ mPas and water is illustrated, respectively. The results shown in figure 6 are obtained for a superficial liquid velocity of $j_L = 0.9$ m/s and superficial air velocities in the range of $j_G = 0.9$ m/s [figure 6(a)] to $j_G = 8.9$ m/s [figure 6(d)].

The results shown in figure 6(a) are obtained near the transition boundary between the plug flow and the aerated slug flow regime. The liquid slugs are much longer than the elongated gas bubbles which leads to a higher probability density in the range of high liquid holdups than in the range of lower liquid holdups associated with the elongated gas bubbles. The amount of gas entrained into the liquid slugs is less than 4%.

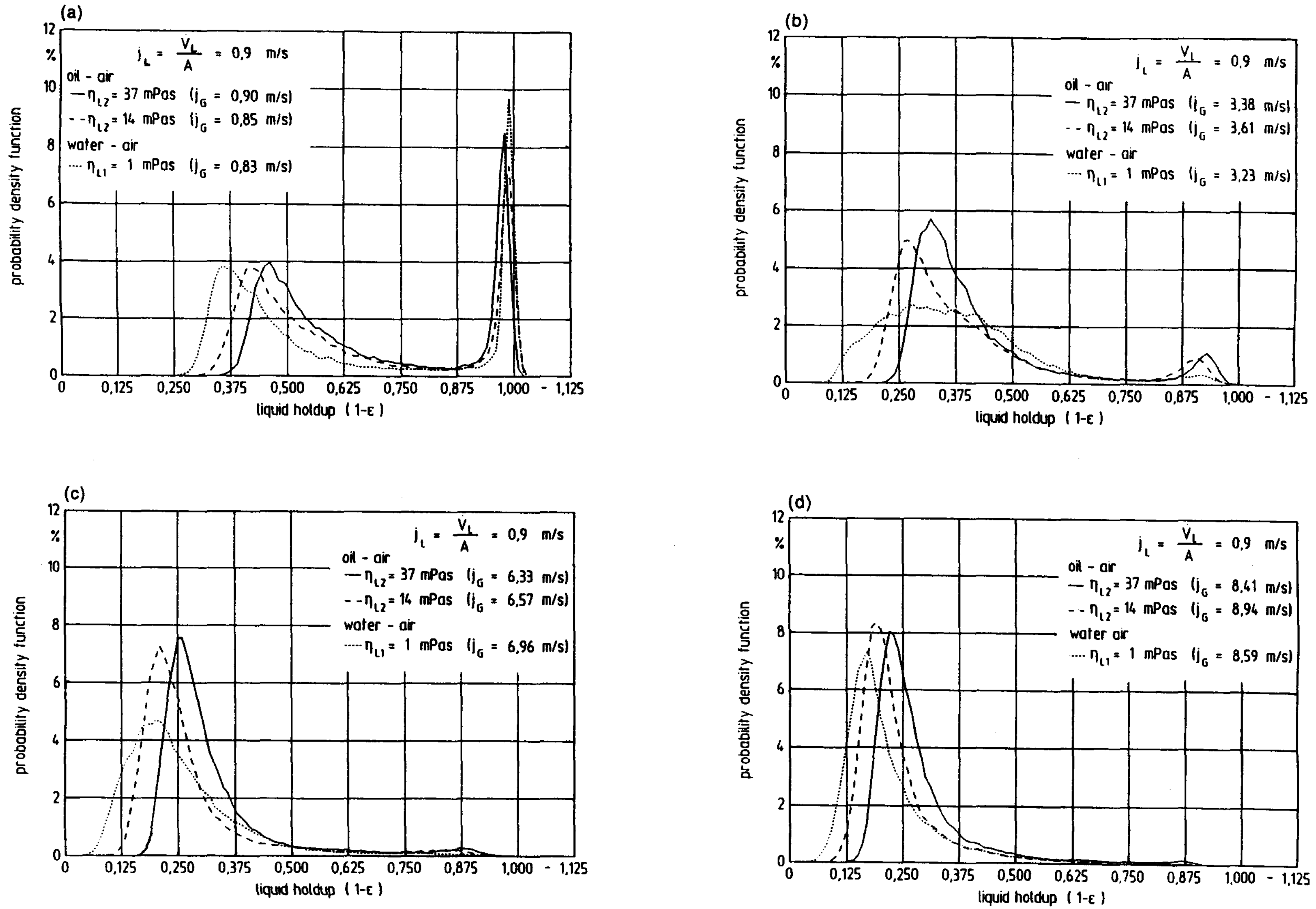

Figure 6. Probability density function of the liquid holdup measured by the γ-ray densitometer for a superficial liquid velocity of $j_L = 0.9$ m/s.

Increasing the superficial air velocity to $j_G \approx 3.5$ m/s, the gas entrainment into the liquid slugs increases and the height of the liquid film in the film zone decreases. This results in a shifting of both peaks in the probability density function to lower liquid holdups [figure 6(b)]. With increasing gas velocity, the length of the elongated gas bubbles increases and exceeds that of the liquid slugs. Therefore, the peak relative to the film zone becomes higher while the peak corresponding to the liquid slug zone decreases. In the case of the two-phase flow of oil and air distinct peaks are obtained while in the case of water and air no distinct peaks can be determined. In contrast to the slug flow of oil and air where a stratified flow is observed between the oil slugs, waves and developing slugs are present between the regular slugs in the case of the water–air slug flow.

Increasing the superficial air velocity above $j_G \approx 6$ m/s leads to an increasing value of the peak relative to the film zone. At the same time the peak for the liquid slug zone decreases and is no longer well-defined [figure 6(c) and (d)]. The increase in the gas velocity results in an increasing gas entrainment into the liquid slug where the ratio between the length of the mixing zone and the length of the core zone of the liquid slug increases.

Finally, for sufficiently high air velocities, first the slug-blow-through regime and then the annular flow regime occurs where only one peak in the probability density function at low liquid holdups is observed.

In the case of well-defined peaks in the probability function, the corresponding liquid holdups agree with the mean holdup in the smooth stratified film zone and the liquid slug zone. Increasing the gas velocity leads to probability density functions where the peaks cannot be determined accurately as shown in figure 6. In this case, the liquid holdup within the highly aerated liquid slugs approaches the liquid holdup within the film zone. Therefore, slugs cannot be further distinguished by means of the probability density function of the holdup signal neither from larger waves nor from short unstable slugs which are still growing or decreasing while travelling downstream as shown in figure 5(c).

If there are any liquid accumulations (e.g. like larger waves, developing or decaying slugs) between the regular slugs, these liquid accumulations are neither represented by the peak corresponding to the stratified film zone nor by the peak corresponding to the liquid slug zone. In this study it is found, that determining the average liquid holdup within the film zone and the slug zone is only accurate in the case of a smooth stratified film zone and low aerated slugs.

Therefore, in this study the method mentioned in section 3 is used in order to measure the liquid holdup within the slug and the film zone.

Nevertheless, the results shown in figure 6 indicate that the liquid holdup in the slug zone, the liquid holdup in the film zone and consequently the average liquid holdup within the slug unit increases with increasing liquid viscosity. This result is confirmed by the following results on the liquid distribution of the slug flow.

Averaged liquid holdup within the slug unit

The average liquid holdup within the slug unit $H_U = V_{L,U}/l_U A$ is defined as the volume $V_{L,U}$ of the liquid phase within a slug unit related to the total volume $l_U A$ of the slug unit. In figures 7 and 8 the effect of the liquid viscosity on the average liquid holdup $H_U = (1 - \epsilon_U)$ within the slug units is shown. In figure 7 results are given for the air–liquid slug flow at the superficial liquid velocity of $j_L = 0.15$ m/s and in figure 8 for the superficial liquid velocity of $j_L = 0.9$ m/s. The results are obtained for oil with viscosities of 14 and 37 mPas and for water as the liquid phases. For both liquids, the liquid holdup decreases with increasing superficial air velocities. The transition from the elongated bubble flow regime to the aerated slug flow regime is expected for superficial air velocities in the range from $j_G = 0.5$ m/s to $j_G = 1$ m/s as confirmed in a previous investigation (Nädler & Mewes 1992). By comparing the oil–air and the water–air data in the case of superficial air velocities lower than $j_G = 1$ m/s, the liquid holdup of the elongated bubble flow regime does not seem to be significantly influenced by the liquid viscosity. In the aerated slug flow regime at superficial air velocities greater than $j_G = 1$ m/s, the liquid holdup increases with increasing liquid viscosities which is in agreement with the available results mentioned above.

By increasing the superficial liquid velocity from $j_L = 0.15$ m/s to $j_L = 0.9$ m/s, the results shown in figure 8 are obtained. The same trend of increasing values for the liquid holdup with increasing liquid viscosity can be observed. Comparing the results for the different superficial liquid velocities,

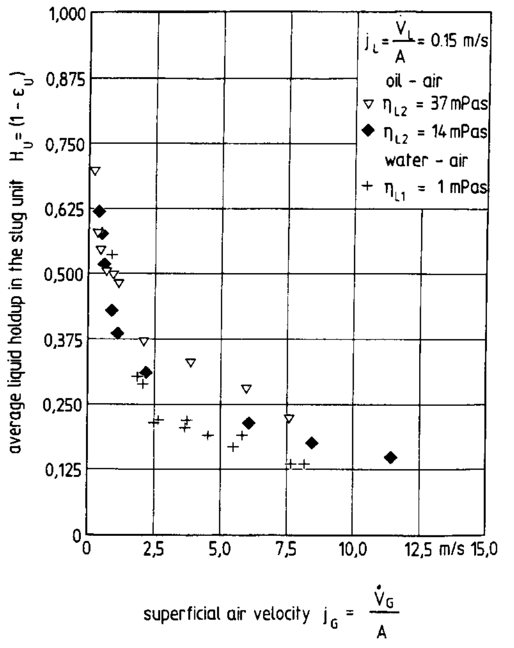

Figure 7. Measured liquid holdup in a slug unit for a superficial liquid velocity of $j_L = 0.15$ m/s.

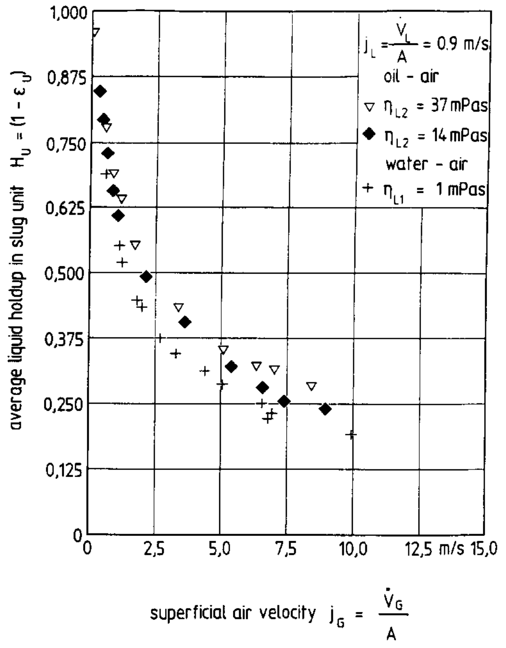

Figure 8. Measured liquid holdup in a slug unit for a superficial liquid velocity of $j_L = 0.9$ m/s.

there is a decreasing effect of the liquid viscosity on the liquid holdup with increasing superficial liquid velocity. In the case of oil–air slug flow the liquid holdup is significantly higher than in the case of water–air slug flow. This may not only be due to an effect of the liquid viscosity but also due to the different surface tensions and liquid densities.

Liquid holdup within the elongated bubble zones

The average liquid holdup within the elongated bubble zone $H_B = V_{L,B}/l_F A$ is defined as the volume $V_{L,B}$ of the liquid phase within the elongated bubble zone related to the total volume $l_F A$ of this zone. In figure 9 the average liquid holdup in the elongated bubble zone is shown as a

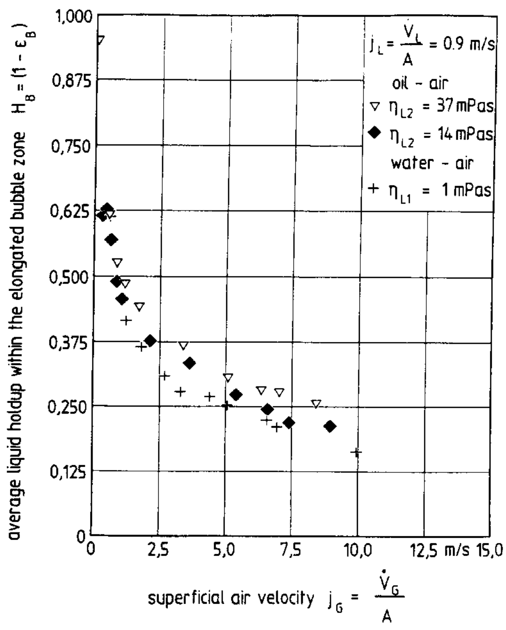

Figure 9. Measured liquid holdup in the elongated bubble zone for a superficial liquid velocity of $j_L = 0.9$ m/s.

function of the superficial air velocity. The liquid holdup in the elongated bubble zone increases with increasing liquid viscosity. The lowest values for the liquid holdup are obtained in the case of water being the liquid phase. The increase is due to the height of the liquid film. Therefore, the liquid holdup depends on the forces acting at the gas–liquid interface and the pipe wall. An increasing liquid viscosity causes higher shear stresses in the liquid phase. The liquid level rises until the shear forces acting at the pipe wall and at the gas–liquid interface are in equilibrium. The fact, that in the case of oil–air slug flow the liquid holdups are greater than in the case of water–air slug flow may not only be due to an effect of the liquid viscosity but also due to an effect of surface tension which is observed in the stratified wavy flow regime (Hand *et al.*, 1992; Hand & Spedding 1993).

The mean holdup values in figure 9 are higher than the corresponding peak values in the probability density function shown in figure 6. This is due to liquid accumulations in the film zone mentioned above. These accumulations are included in the mean holdup values in figure 9 while they are not represented by the peak values in the probability density function.

Liquid holdup within the liquid slugs

The average liquid holdup within the liquid slugs $H_S = V_{L,S}/l_S A$ is defined as the volume $V_{L,S}$ of the liquid phase within the liquid slug related to the total volume $l_S A$ of the liquid slug. In figures 10 and 11 the average liquid holdup in the slug zone $(1 - \epsilon_S)$ is shown as a function of the mixture velocity $j_m = (j_L + j_G)$. By increasing the mixture velocity, the values of the liquid holdup are decreasing.

The results given in figure 10 are obtained for several superficial liquid velocities in the range from $j_L = 0.15$ m/s to $j_L = 1.5$ m/s using oil with viscosities of 14 and 37 mPas and water as the liquid phase. Comparing the liquid holdup for oil–air and water–air slug flow, a difference can be distinguished.

In the case of oil–air the liquid holdups are lower than the corresponding liquid holdups in the case of water–air. As shown in several studies (Nydal 1991; Andreussi & Bendiksen 1989), the liquid holdup within the liquid slug increases with an increasing surface tension of the gas–liquid mixture. According to these results, the oil holdup is lower than the water holdup shown in figure 10 which is expected due to the different surface tension of the water ($\sigma_{L1} \approx 70$ N/m) and the oil ($\sigma_{L2} \approx 30$ mN/m) used in the present study.

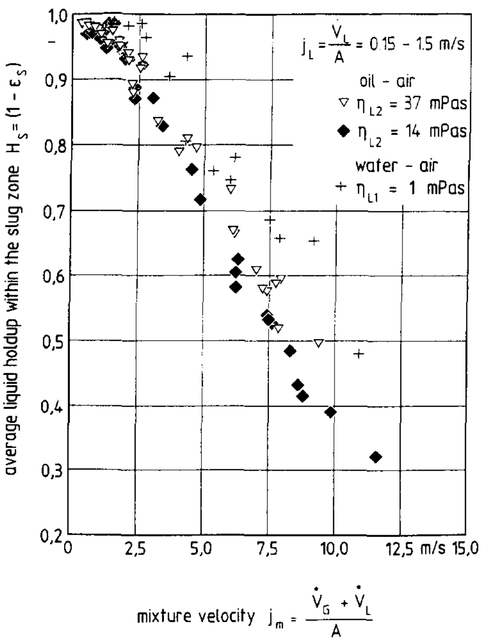

Figure 10. Measured liquid holdup within the liquid slug zone.

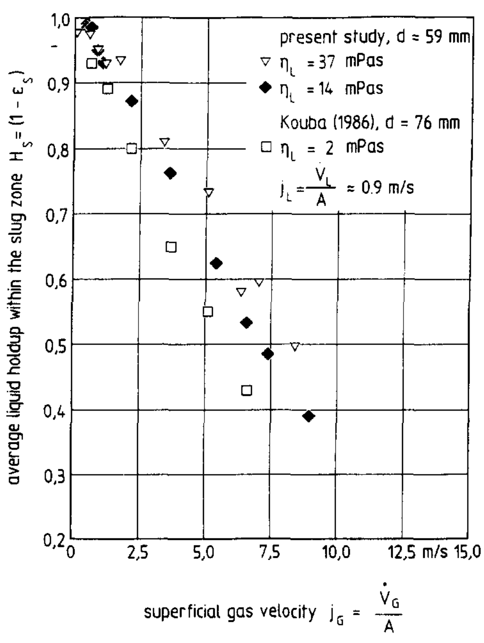

Figure 11. Comparison of the liquid holdup within the slug zone for different liquid viscosities.

The liquid holdup within the water slugs and hence the difference observed in the values for the liquid holdup between the water–air and oil–air slug flow is lower than that reported by Andreussi & Bendiksen (1989) and Nydal (1991). It should be mentioned that Andreussi & Bendiksen (1989) and Nydal (1991) obtained their results by means of conductivity probes while the results presented here are obtained by means of a multi-beam γ-ray densitometer. Heywood & Richardson (1979) also measured the liquid holdup within liquid slugs in water–air slug flow by means of a γ-ray densitometer. By comparing their results for the average liquid holdup within the water slugs in a horizontal 42 mm inner diameter pipe and the values given in figure 10, a good agreement is obtained. The different results for the liquid holdup within the water slugs obtained by means of conductance probes and γ-ray densitometers indicate that these measurements are very sensitive to the measurement method.

When considering only the oil–air data from figure 10, it is indicated, that the liquid holdup within the slug zone increases slightly by increasing the liquid viscosity.

The net gas entrainment into the liquid slug $\dot{V}_G = \dot{V}_1 + \dot{V}_2$ is the sum of the gas entrained with the aerated liquid film picked up at the slug front and the gas entrained from the elongated gas bubble. The rates of gas entrainment into the liquid slug are approximately proportional to the relative velocity $(v_{SF} - v_B)$ between the slug front v_{SF} and the average gas velocity in the elongated gas bubble v_B as well as proportional to the relative velocity $(v_{SF} - v_F)$ between the slug front v_{SF} and the average fluid velocity of the aerated liquid film v_F. Furthermore, the rate of gas entrainment is also proportional to the fractions of pipe cross-sectional area occupied by the elongated gas bubble and by the aerated film. The amount of gas bubbles within the film is investigated by means of visual observations and photographic techniques using a still camera and a video camera. It has been confirmed by these investigations that the void fraction $\epsilon_{G,F}$ within the liquid film just in front of the liquid slug is very low so that the amount of gas entrained by means of the aerated liquid film is negligible in comparison to the gas entrained from the elongated gas bubble, yielding $\dot{V}_1 \ll \dot{V}_2$.

The increase of the liquid viscosity results in an increase of the liquid holdup within the elongated bubble zone and hence according to figure 9 to a decrease in the void fraction ϵ_B. Due to continuity requirements, a decrease in the void fraction ϵ_B results in an increasing velocity v_B of the gas flowing in the elongated gas bubble. Both a decreasing void fraction ϵ_B as well as an increasing gas velocity v_B are expected to lead to decreasing gas entrainment rates into the liquid slug and hence to an increasing liquid holdup within the slug zone. Therefore, increasing the liquid viscosity may result in an increasing liquid holdup within the slug zone.

Increasing the liquid viscosity also results in increasing slug front velocities as shown in other investigations (Sam & Crowley 1986; Nädler & Mewes 1993). The results of previous investigations (Nädler & Mewes 1993) indicate that in the case of increasing viscosities, the change in liquid holdup is stronger than the change in the slug velocity. Therefore, the increase in liquid viscosity should finally result in increasing liquid holdup within the liquid slug zone.

The relation between the liquid holdup within the liquid slug zone $(1 - \epsilon_S)$ and the slug velocity v_{SF} can be expressed by

$$v_{SF} = \frac{(1 - \epsilon_S)v_{L,S} - j_L}{(1 - \epsilon_S) - (1 - \epsilon_U)} = \frac{j_G - \epsilon_S v_{G,S}}{\epsilon_U - \epsilon_S}, \qquad [2]$$

where $v_{L,S}$ and $v_{G,S}$ are the average velocities of the liquid and the gas phase within the liquid slug, respectively (Nädler & Mewes 1993).

Rearranging [2] results in

$$\epsilon_S = \frac{\epsilon_U v_{SF} - j_G}{v_{SF} - v_{G,S}}. \qquad [3]$$

The effect of the liquid viscosity on the parameters in [3] is considered with respect to the mechanisms shown in figure 12.

First the numerator $(\epsilon_U v_{SF} - j_G)$ in [3] is considered. By increasing the liquid viscosity $(\eta_L \uparrow)$ the average liquid holdup increases $((1 - \epsilon_U)\uparrow)$ (see figures 7 and 8) and hence the void fraction decreases $(\epsilon_U \downarrow)$, while the slug velocity increases $(v_{SF} \uparrow)$ whereas the superficial air velocity is unaffected $(j_G \rightarrow)$. As the decrease in the void fraction $(\epsilon_U \downarrow)$ is stronger than the increase in the

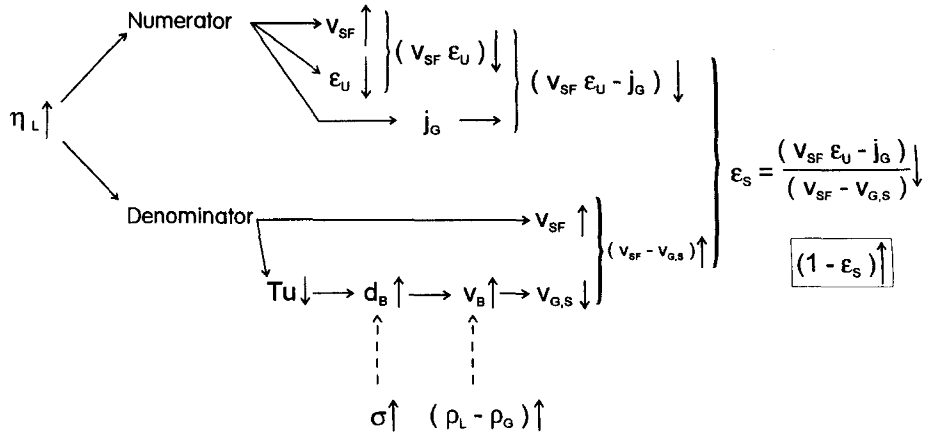

Figure 12. Schematic diagram of the possible effects due to changing liquid viscosity.

slug velocity ($v_{SF}\uparrow$), the product $((\epsilon_U v_{SF})\downarrow)$ and hence the numerator of [3] decreases $((\epsilon_U v_{SF} - j_G)\downarrow)$.

The denominator $(v_{SF} - v_{G,S})$ in [3] increases with an increase of the difference between the slug velocity v_{SF} and the gas velocity $v_{G,S}$. As mentioned above, by increasing the liquid viscosity ($\eta_L\uparrow$) an increasing slug velocity ($v_{SF}\uparrow$) is expected. Furthermore, by increasing the liquid viscosity ($\eta_L\uparrow$) the intensity of turbulence ($Tu\downarrow$) within the mixing zone is expected to decrease. The size of the gas bubbles created within the mixing zone depends on the intensity of turbulence within the mixing zone and the fluid and rheological properties of the liquid and the gas phase. Therefore, increasing the viscosity and decreasing the intensity of turbulence ($Tu\downarrow$) should result in a greater mean bubble diameter ($d_B\uparrow$). By increasing the mean bubble size ($d_B\uparrow$) the terminal bubble rise velocity ($v_0\uparrow$) will also increase. A greater rise velocity of the gas bubbles results in gas bubbles entrained from the mixing zone into the core zone of the liquid slug rising to the top of the pipe. The fluid flowing near the pipe wall moves slower than the average fluid velocity within the slug and will be shed from the rear of the slug. Therefore, by an increasing amount of gas bubbles rising to the upper pipe wall, the amount of gas flowing with low velocities will increase, too. The mean gas velocity within the slug zone ($v_{G,S}\downarrow$) decreases by the amount of gas flowing under the top of the pipe in the pipe cross-sectional area at low flow velocities. By increasing the slug velocity ($v_{SF}\uparrow$) and decreasing the gas velocity ($v_{G,S}\downarrow$) the difference between these velocities $((v_{SF} - v_{G,S})\uparrow)$ and hence the denominator of [3] will increase.

By increasing the liquid viscosity ($\eta_L\uparrow$) an increasing numerator and a decreasing denominator of [3] is obtained which leads to a decreasing fraction and hence a decreasing void fraction ($\epsilon_S\downarrow$) resulting in an increasing liquid holdup $((1-\epsilon_S)\uparrow)$.

Following the approach mentioned above, an increase of the surface tension ($\sigma\uparrow$) as well as an increasing difference of the fluid densities $((\rho_L - \rho_G)\uparrow)$ are also expected to result in lower void fractions ($\epsilon_S\downarrow$) within the liquid slug as shown in figure 12. Therefore, the liquid holdups in the case of oil–air should be lower than the corresponding liquid holdups in the case of water–air due to the lower surface tension and the lower difference of the fluid densities for oil–air mixtures. This is in agreement with the results shown in figure 10.

According to the mechanism shown in figure 12, in the case of results for the liquid holdup within the liquid slug zone obtained for different liquids it is possible that the effects of the liquid viscosity, surface tension and liquid density compensate each other.

Increasing liquid holdup with increasing liquid viscosity is also reported in investigations of jet loop reactors and bubble columns, i.e. by Padmavathi & Remananda Rao (1992), Eissa & Schügerl (1975), Bach & Pilhofer (1978), Godbole *et al.* (1982) and Zahradnik *et al.* (1987). The results of Eissa & Schügerl (1975) indicated, that in the case of constant volumetric flow rates of the fluids, an increase of the liquid viscosity from 1 to 3 mPas leads to decreasing fluid holdup where the minimum liquid holdup is obtained for a liquid viscosity of about 3 mPas. In contrast to that, a further increase of the liquid viscosity from 3 mPas leads to a continuously increasing liquid holdup where the liquid holdup in the case of a viscosity above 11 mPas is lower than in the case of a viscosity of 1 mPas. The effect of increasing liquid holdup with increasing liquid viscosity is found to be enhanced by increasing gas velocities. The enhancement of the increase of the liquid holdup due to the increase of the liquid viscosity by increasing the gas velocity is confirmed by the results of Bach & Pilhofer (1978), Godbole *et al.* (1982) and Zahradnik *et al.* (1987) and also by Padmavathi & Remananda Rao (1992). In these studies, the critical viscosity associated with a minimum of the liquid holdup is reported to be about 3 mPas.

A similar trend is observed in the results shown in figure 10. In the investigations of jet loop reactors and bubble columns the increasing liquid holdup with increasing liquid viscosity is attributed to bubble coalescence. According to these results, an increase in the liquid viscosity may result in higher bubble coalescence rates, larger gas bubbles and a decrease in the gas residence time and hence an increasing liquid holdup.

Comparing the results of these investigations and the results shown in figure 10, similar tendencies are observed. Furthermore, the results shown indicate that the effect of the surface tension on the liquid holdup within the slug zone is stronger than the effect of liquid viscosity.

In figure 11, data of this study for the liquid holdup within the slug zone are compared with the data reported by Kouba (1986) for mixtures of kerosene and air flowing in a 76 mm inner diameter pipeline. All data shown in figure 11 are obtained for a superficial liquid velocity in the order of $j_L \approx 0.9$ m/s.

The results reported by Gregory *et al.* (1978) for the liquid holdup in pipes with diameters of 25.8 and 51.2 mm as well as the results for the liquid holdup in pipes with diameters of 31, 53 and 90 mm reported by Nydal (1991, 1992) indicate, that in this diameter range the effect of the pipe diameter itself is only weak. Therefore, such a significant difference in the present data for the liquid holdup and the data reported by Kouba (1986), as shown in figure 11, seems to be affected by the difference in the liquid viscosities. This supports the trend of an increasing liquid holdup within the slug zone for increasing liquid viscosities. Nevertheless, further investigations are required to predict the effect of the liquid viscosity on the liquid holdup.

4. CONCLUDING REMARKS

In the investigations of two-phase air–liquid slug flow in horizontal pipes, the effect of the liquid viscosity on the slug characteristics is considered using oil of viscosities in the range from 14 to 37 mPas and water as the liquid phase. Comparing the results for oil–air and water–air slug flow, significant differences in the liquid distribution are determined.

The values of flow averaged liquid holdup and the liquid holdup within the film zone are significantly higher for oil–air slug flow than the corresponding values for water–air slug flow while the liquid holdup within the slug zone in oil–air slug is lower than the liquid holdup in water–air slug flow.

By increasing the liquid viscosity, increasing values for the volume averaged liquid holdup in the slug unit and the film zone as well as the slug zone are obtained in the aerated slug flow regime. While the effect on the liquid holdup in the film zone and the slug unit is significant, only a slight effect on the liquid holdup within the liquid slug is observed. According to the results mentioned above, the difference in the liquid holdup in the slug zone of oil–air and water–air may be caused by the difference in the surface tension or liquid density.

The results of the present study indicate trends which are observed for only a limited range in the liquid viscosity. Therefore, these results are not sufficient to generalize the effect of the liquid viscosity observed in the present study. In order to estimate the effect of the fluid properties on the slug characteristics and particularly, the effect on the liquid holdup within the slug zone, further investigations are required.

Acknowledgement—The authors wish to acknowledge the financial support of the BMFT (Bundesministerium für Forschung and Technologie) for a program on technical development of Multiphase Transportation Systems.

REFERENCES

ANDREUSSI, P. & BENDIKSEN, K. 1989 An investigation of void fraction in liquid slugs for horizontal and inclined gas–liquid pipe flow. *Int. J. Multiphase Flow* **15,** 937–946.

ANDREUSSI, P., MINERVINI, A. & PAGLIANTI, A. 1993 A mechanistic model of slug flow in near-horizontal pipes. *AIChE Jl* **39,** 1281–1291.

BACH, H. F. & PILHOFER, T. 1978 Variation of gas holdup in bubble columns with physical properties of liquid and operating parameters of columns. *Ger. Chem. Engng* **1,** 270–275.

CROWLEY, C. J., SAM, R. C., WALLIS, G. B. & METHA, D. C. 1984 Slug flow in large diameter pipe: I. Effect of fluid properties. *AIChE Annual Meeting*, San Francisco, CA.

DUKLER, A. E. & HUBBARD, M. G. 1975 A model for gas–liquid slug flow in horizontal and near horizontal tubes. *Ind. Engng Chem. Fundam.* **14,** 337–347.

EISSA, S. H. & SCHÜGERL, K. 1975 Holdup and back mixing investigations in cocurrent and countercurrent bubble columns. *Chem. Engng Sci.* **30,** 1251–1256.

FORTESCUE, T. R., LÖFFEL, R. & ROMANSKI, J. J. 1994 High precision gamma and x-ray densitometry using current mode. *Int. Conf. on New Trends in Nuclear System Thermohydraulics*, Pisa, Italy.

GODBOLE, S. P., HONATH, M. F. & SHAH, Y. T. 1982 Holdup structure in highly viscous Newtonian and non-Newtonian liquids in bubble columns. *Chem. Engng Commun.* **16,** 119–134.

GREGORY, G. A., NICHOLSON, M. K. & AZIZ, K. 1978 Correlation of the liquid volume fraction in the slug for horizontal gas–liquid slug flow. *Int. J. Multiphase Flow* **4,** 33–39.

HAND, N. P. & SPEDDING, P. L. 1993 Horizontal gas–liquid flow at close to atmospheric conditions. *Chem. Engng Sci.* **48,** 2283–2305.

HAND, N. P., SPEDDING, P. L. & RALPH, S. J. 1992 The effect of surface tension on flow pattern, holdup and pressure drop during horizontal air–water pipe flow at atmospheric conditions. *Chem. Engng Jl* **48,** 197–210.

HEYWOOD, N. I. & RICHARDSON, J. F. 1979 Slug flow of air–water mixtures in a horizontal pipe: determination of liquid holdup by γ-ray absorption. *Chem. Engng Sci.* **34,** 17–30.

KAGO, T., SARUWATARI T., KASHIMA, M., MOROOKA S., & KATO, Y. 1986 Heat transfer in horizontal plug and slug flow for gas–liquid and gas–slurry systems. *J. Chem. Engng Jap.* **19,** 125–131.

KOKAL, S. L. & STANISLAV, J. F. 1989 An experimental study of two-phase flow in slightly inclined pipes—II. Liquid holdup and pressure drop. *Chem. Engng Sci.* **44,** 681–693.

KOUBA, G. E. 1986 Slug flow modeling and metering. Ph.D. thesis, University of Tulsa, OK.

NÄDLER, M. & MEWES, D. 1992 Characteristics of gas–liquid and gas–liquid–liquid slug flow in horizontal pipes. *ASME FED Multiphase Flow Wells Pipe* **144,** 39–50.

NÄDLER, M. & MEWES, D. 1993 Multiphase slug flow in horizontal pipes. Paper I2, *30th European Two-phase Flow Group Meeting*, Hannover, Germany.

NICHOLSON, M. K., AZIZ, K. & GREGORY, G. A. 1978 Intermittent two phase flow in horizontal pipes: predictive models. *Can. J. Chem. Engng* **56,** 653–663.

NYDAL, O. J. 1991 An experimental investigation of slug flow. Ph.D. thesis, University of Oslo, Norway.

NYDAL, O. J., PINTUS, S. & ANDREUSSI 1992 Statistical characterization of slug flow in horizontal pipes. *Int. J. Multiphase Flow* **18,** 439–453.

PADMAVATHI, G. & RAO REMANANDA, K. 1992 Effect of liquid viscosity on gas holdups in a reversed flow jet loop reactor. *Can. J. Chem. Engng* **70,** 800–802.

SAM, R. G. & CROWLEY, C. J. 1986 Investigation of two-phase flow processes in coal slurry/hydrogen heaters. Final Report DOE/PC 800028-T1, Creare Inc., Hanover, NH.

TAITEL, Y. & BARNEA, D. 1990 A consistent approach for calculating pressure drop in inclined slug flow. *Chem. Engng Sci.* **45,** 1199–1206.

ZAHRADNIK, J., PETER, R. & KASTANEK, F. 1987 The effect of liquid phase properties on gas holdup in bubble column reactor. *Coll. Czech. Chem. Commun.* **53,** 335–347.