

SLUG FLOW AND ITS TRANSITIONS IN LARGE-DIAMETER HORIZONTAL PIPES

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Abstract—A flow regime map is compiled for the air/water two-phase flow in the Harwell 30 cm dia horizontal pipeline. The transitions differ substantially from those for small-diameter pipes and are not predicted accurately by any theoretical model. Data on slug frequency, length, holdup, translational velocity and pressure gradient are presented and compared with existing data. The results show that the pipe diameter has a large effect on all these properties. Several changes in the distribution of the phases in large-diameter pipes are reported. These are especially prominent in slug and annular flow.

Key Words: slug flow, horizontal flow, gas/liquid flow, transitions

1. INTRODUCTION

The simultaneous flow of gases and liquids occurs in many items of process equipment, e.g. boilers, reactors and oil and gas pipelines. Many workers have studied flow in small-diameter pipes, i.e. 2.54 and 5.08 cm dia pipes, but very little data is available from large-diameter pipelines, e.g. 15, 30 and 60 cm dia. The need for data on the latter is becoming increasingly important, because many offshore facilities are requiring that pipelines be combined into much larger pipes and the oil and gas be pumped ashore before separation. This work provides a flow regime map and slug flow characteristics—e.g. slug frequency, length, holdup, translational velocity and pressure gradient—for the flow of air and water in a 30 cm schedule 40 carbon steel commercial pipe and compares these with existing data and correlations.

Many flow regime maps have been produced for two-phase flow in horizontal pipes. Baker (1954) presented a map based on the flow in small-diameter pipes using several fluids. The axes of the map involved the mass fluxes of the phases together with the fluid properties, e.g. density and surface tension. Wallis & Dobson (1973) studied air/water flow in 2.54 and 30.5 cm channels and produced similar flow regime maps. Mandhane *et al.* (1974) studied two-phase flows in small-diameter pipes but constructed a map based on the superficial liquid and gas velocities, respectively. This type of map is now the most widely used.

Taitel & Dukler (1976) produced a theoretical, mechanistic flow regime map and this is still the one most often used but with modifications to the calculation of the interfacial friction factor. Simpson *et al.* (1981) looked at flows in 12.7 and 21.6 cm dia pipes and produced flow regime maps which showed similar effects to that of Wallis & Dobson (1973), but some of the data was inconclusive. Lin & Hanratty (1987a, b) used new techniques based on the cross-correlation signals from fast response pressure transducers to determine the flow regimes and produced a flow map for air/water flow in 2.54 and 9.53 cm dia pipes.

Several experimental and theoretical studies have also been carried out in horizontal pipelines. Gregory & Scott (1969) presented data on slug frequency and translational velocity in a 1.9 cm dia pipe. Dukler & Hubbard (1975) produced a realistic mechanistic model, which included correlations for holdup etc., developed from their experimental work on a 3.81 cm dia pipe. Some of their results are in terms of mass flow rates rather than volume flow rates and, since no density of the air was given, no direct comparison with this work was possible. Nicholson *et al.* (1978) extended the Dukler & Hubbard (1975) model and conducted experiments on 2.58 and 5.12 cm dia pipes, yielding data on pressure gradient, slug frequency, holdup and translational velocity.

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The Beggs & Brill (1973) correlation for pressure gradient was used for comparison with the present data. This correlation is based on experimental results for small-diameter pipes but is often extrapolated to large-diameter pipes.

Kouba (1987) provided data for slug flow in a 7.62 cm dia pipeline using kerosene and air as the fluids.

There are many other sources of data but many are difficult to access or to solve the relevant equations. Consequently, only the above are compared to the present work.

2. EXPERIMENTAL SETUP

The Harwell test facility is outlined in figure 1. Air from the atmosphere is drawn into two large Nokia centrifugal type HKXX-22 fans arranged in series. The fans are driven by 200 kW Hawker Siddeley electric motors. The fans can provide $3.3 \text{ m}^3/\text{s}$ of air at a maximum pressure of approx. 200 kPa absolute pressure. This gives a range of superficial air velocities from 0 to 30 m/s. The flow rate of air is controlled by the use of vanes on the fans. The air can be supplemented by the Harwell mains air supply.

The air is then cooled in a water-cooled heat exchanger and is then passed through an orifice plate and the pressure and temperature monitored. The mass flow rate of air is then calculated. Three different orifice plates are required to cover the full range of air velocities.

Demineralized water from a large storage/separator vessel is circulated using a Sigmund Pulsometer pump which can produce a flow rate of $0.265 \text{ m}^3/\text{s}$ against a 20 m head. The flow rate is controlled by a butterfly valve downstream of the pump. The superficial velocity of the water ranges from 0 to 3 m/s. The flow rate of the water is also measured using an orifice plate.

The water is then passed to the mixer where it is mixed with the air. The air passes axially through the mixer whilst the water is injected uniformly through an annulus into the air stream.

The air/water mixture then passes into a 60 m long, 30 cm schedule 40 carbon steel pipeline. There is a 1 m clear Perspex section 5 m from the end of the pipeline where flow visualization can be carried out. The carbon steel pipe had been treated with corrosion inhibitor. The air/water mixture then passed into the separator, where the air was vented to atmosphere and the water recirculated.

The temperature and pressure are measured using a thermocouple and a fast-response Druck pressure transducer. The flow rates of air and water are then calculated in the two-phase pipeline.

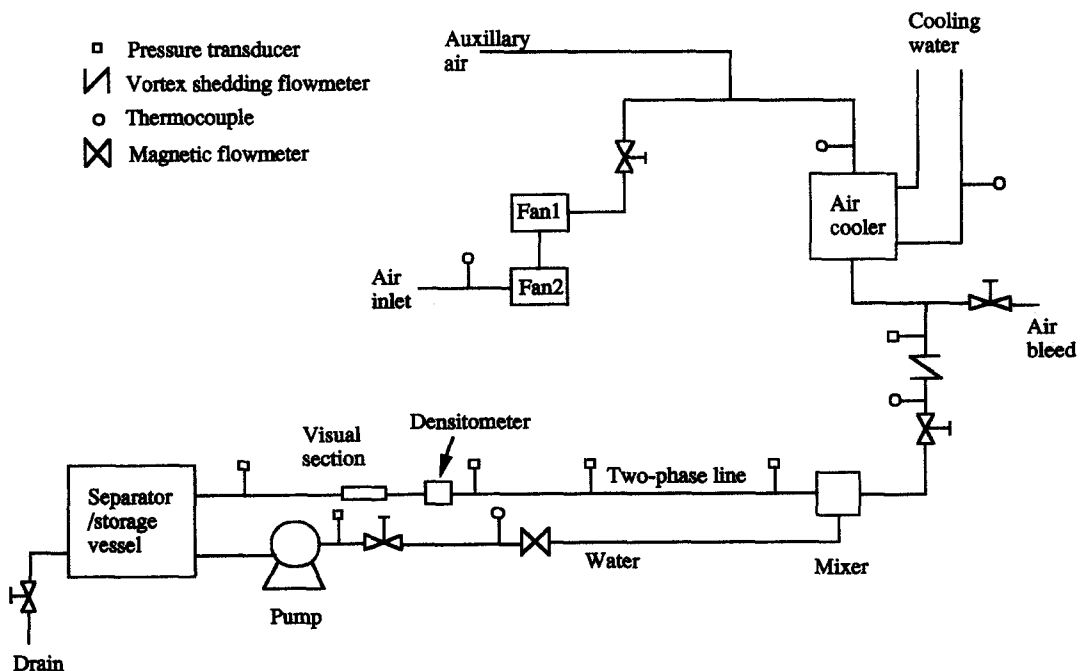


Figure 1. The Harwell 300 mm pipeline.

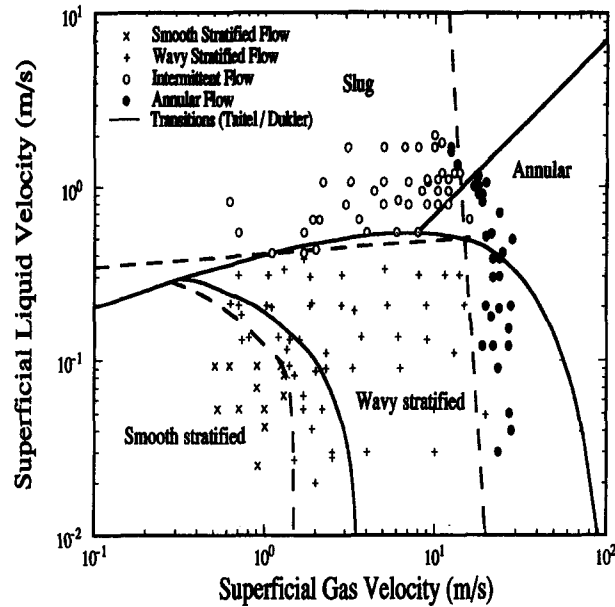


Figure 2. Flow regime map for the Harwell 300 mm pipeline.

A flow regime map was then compiled. A water superficial velocity was taken and kept constant and a range of air superficial velocities were considered. The flow regime determination was carried visually together with the use of two identical fast-response Druck pressure transducers and a Ramsay single-beam γ -densitometer. When the flow rates were selected, no measurements were taken until steady state had been reached, i.e. when there was no change in temperature, static pressure and holdup in the two-phase pipeline.

The fast-response pressure transducers were placed at 13.9 and 38.1 m from the mixer outlet, respectively. The γ -densitometer had a 550 mCi caesium-137 source and a sodium iodide scintillator detector and was positioned 39.1 m from the mixer outlet. The response time was approx. 100 ms but could be made faster by monitoring directly the count rate from the detector. The transducers were connected to a multi-pen chart recorder where the output was monitored. Direct readings or cross-correlations could be obtained from the transducers.

The flow pattern was determined visually and from the transducers in the way suggested by Lin & Hanratty (1987a, b).

After the flow regime transitions and map had been determined, a study of the characteristics of slug flow in the 300 mm pipeline was carried out.

From the output of either transducer or the γ -densitometer, the slug frequency was measured. The pressure drop over a slug body was measured using both transducers. The time taken for a slug to travel between the two transducers was noted and the translational velocity of the slug was calculated.

The output from the γ -densitometer provided information on the liquid film holdup in front of the slug and the holdup within the body of the slug. The slug length could also be estimated.

The results were taken over several slugs and the experiments were repeated to confirm reproducibility.

3. RESULTS

3.1. Flow Regimes

In the production of the flow regime map for the 300 mm dia pipeline, several differences in the distribution of the phases compared to a small-diameter pipe were observed.

Firstly, as one might expect for large-diameter pipes, the effect of stratification of the water phase is much greater and resembles channel flow. This was noticed not only in the stratified and wavy flow regimes but also in the annular flow regime, where the liquid film thickness at the bottom of

the pipe was large and of the order of several centimetres—even for superficial gas velocities around 30 m/s. The liquid film at the top of the pipe was very thin.

The droplets in the core seemed to be a mixture of large and small droplets near the bottom of the pipe but the large droplets did not seem to have enough momentum to carry them into the upper half of the pipe, which only contained small droplets.

In slug flow, the effect of stratification was also noted. At high enough air velocities, where air is entrained in the body of the slug, it was noted that the air was not distributed homogeneously in the slug body but the air seemed to be collecting mostly in the upper half of the pipe. There was little air in the lower regions of the pipe. Consequently, as the gas velocity increased, the upper part of the slug body became more “frothy” and eventually blow-through occurred, leading to the transition to annular flow.

The flow regime map for the 300 mm dia pipeline is shown in figure 2. The map is in the form of a Mandhane plot and the experimental transition boundaries are compared to those predicted by the Taitel & Dukler (1976) model. It was not possible to calculate the interfacial shear for inclusion in the Taitel & Dukler model, so the assumption was that the interfacial friction factor was approximately that for the air alone.

The plot shows that the transition from stratified to slug flow occurs at higher superficial liquid velocities, around 0.8 m/s, and that the transition from wavy to annular flow occurs at very much lower superficial gas velocities than those predicted by Taitel & Dukler (1976). A better agreement may be possible if other interfacial friction factors are used. However, the transition from slug to annular flow cannot be predicted by the Taitel & Dukler (1976) model. This is due to the distribution of the gas in the slug and the blow-through, as described earlier, which cannot be accounted for by the Taitel & Dukler model. The transition to annular flow occurs at superficial gas velocities between 13–18 m/s.

Figure 3 is also a Mandhane plot and compares the 300 mm dia data with that of: Lin & Hanratty (1987a, b), for 25.4 and 95.3 mm dia pipes; Wallis & Dobson (1973), for a 305 mm high channel; and the original Mandhane correlation.

It can be seen clearly that the transition from stratified to slug flow is greatly affected by an increase in pipe diameter. Much more liquid is required to obtain slug flow in large-diameter pipes, as expected. The transition to annular flow follows closely the transition lines suggested by Wallis & Dobson (1973) and Simpson *et al.* (1981). The data from the latter, however, is not shown as the plot would become very cluttered in this area. At low liquid superficial velocities, the Mandhane correlation closely follows the 300 mm data but begins to deviate at higher liquid velocities. Hill

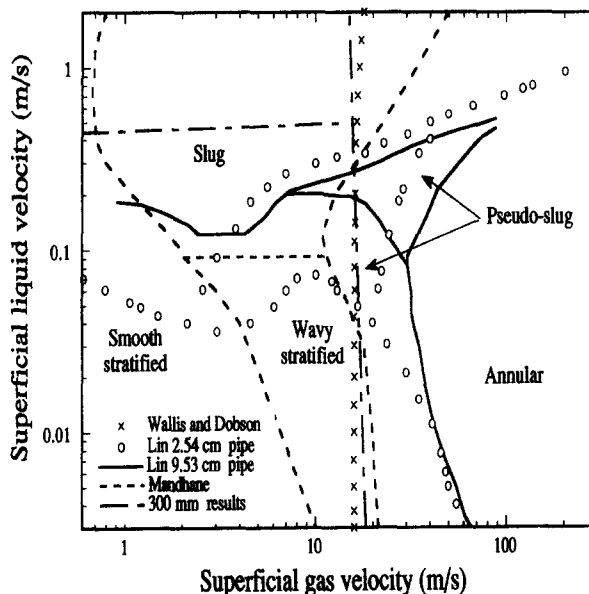


Figure 3. Comparison of the experimental results.

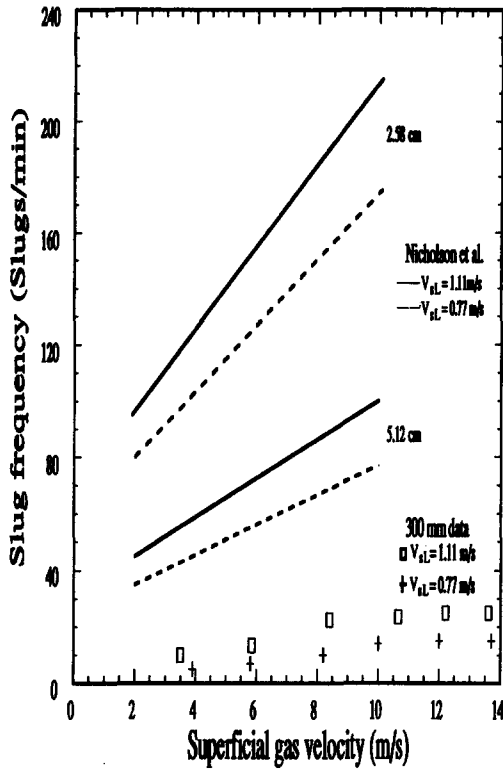


Figure 4. Variation of slug frequency with superficial gas velocity.

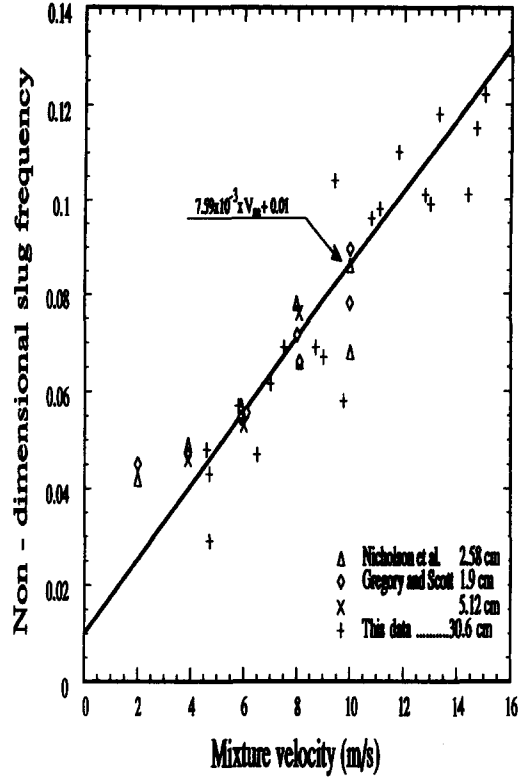


Figure 5. Non-dimensional slug frequency vs mixture velocity.

(1987) has noted similar annular transition lines when data on 60 cm dia pipelines was analysed. These pipelines were many kilometres in length.

3.2. Slug flow characteristics

The characteristics of the slug flow were studied and the results are shown in figures 4-14.

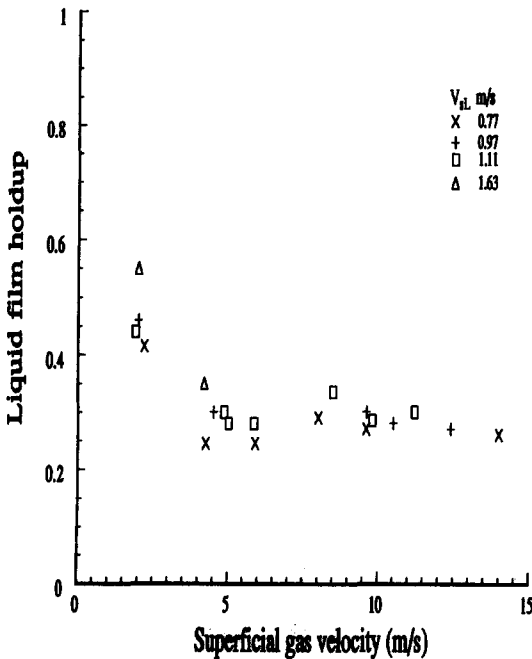


Figure 6. Liquid film holdup.

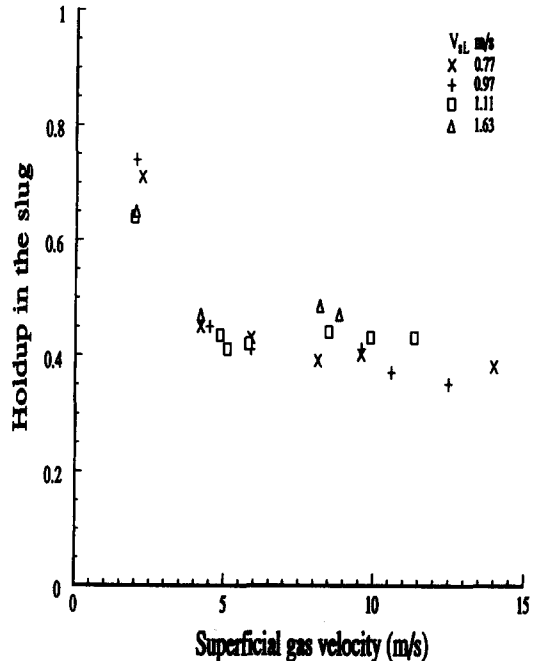


Figure 7. Holdup in the slug body.

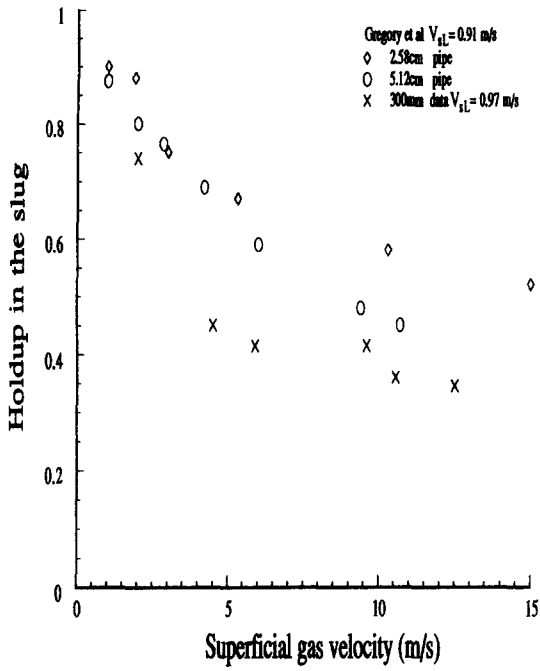


Figure 8. Holdup in the slug body.

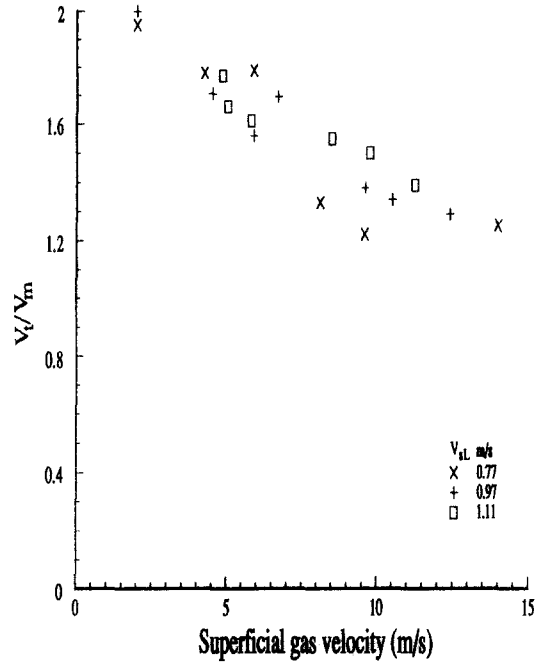


Figure 9. Variation of V_L/V_m with the superficial gas velocity.

Figure 4 shows a plot of slug frequency against superficial gas velocity for the 300 mm dia pipeline and compares the data with that of Nicholson *et al.* (1978), who considered 2.54 and 5.12 cm dia pipes, respectively. It shows that the frequency is reduced if the liquid superficial velocity is reduced. This is to be expected, since less liquid leads to less slugs. Further, increasing

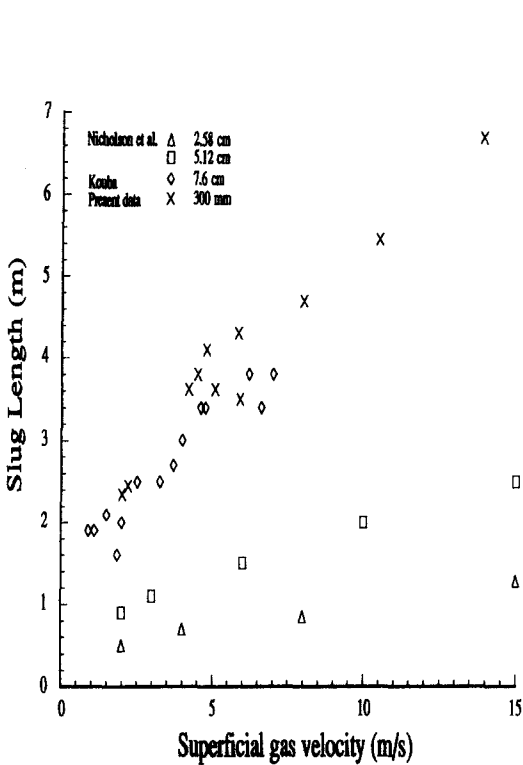


Figure 10. Effect of pipe diameter on the slug length.

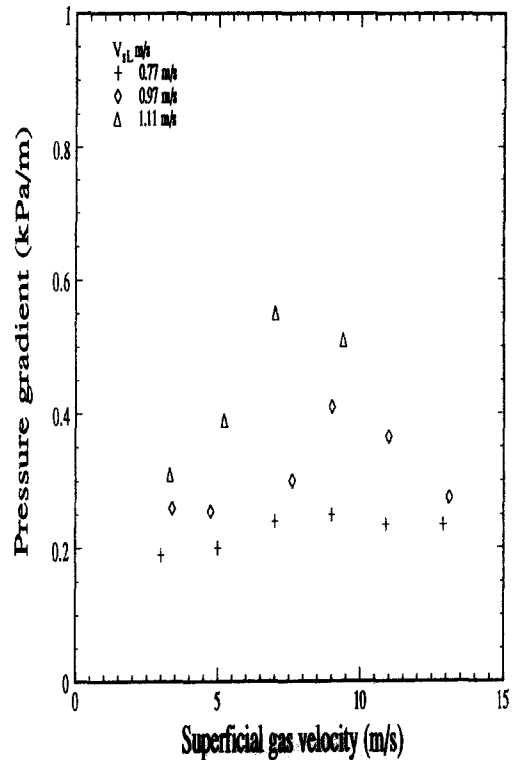


Figure 11. Pressure gradient in the 300 mm pipe.

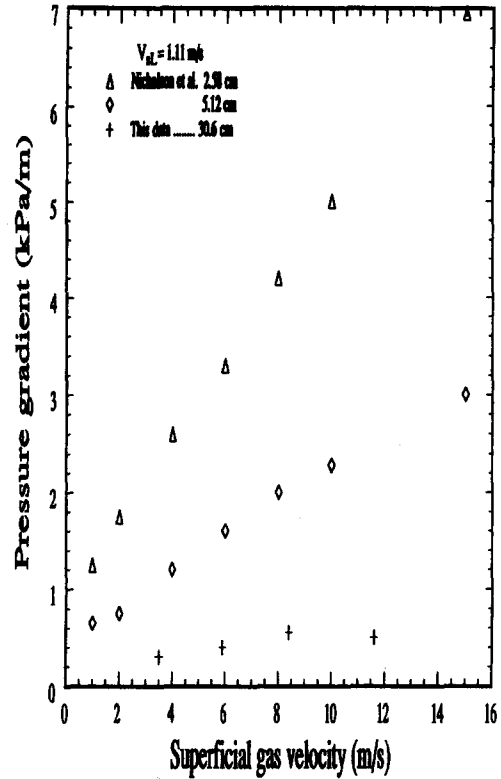
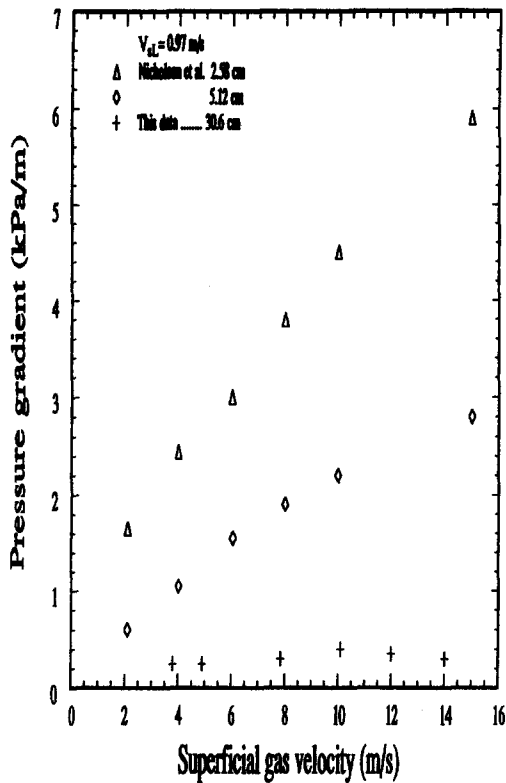


Figure 12. Effect of pipe diameter on the pressure gradient. Figure 13. Effect of pipe diameter on the pressure gradient.

the pipe diameter considerably reduces the slug frequency for a given gas velocity. For example, at 3 m/s superficial gas velocity, the slug frequency is approximately 100, 50 and 10 slugs/min for the 25.4, 51.2 and 300 mm dia pipelines, respectively.

In an attempt to incorporate the effect of pipe diameter, the non-dimensional group vD/V_{sl} is plotted against a mixture velocity, V_m , where v is the slug frequency (s^{-1}), D is pipe diameter (m), V_{sl} is the superficial liquid velocity (m/s) and the mixture velocity is the sum of the superficial liquid and gas velocities.

This plot is given in figure 5 and the 300 mm data is compared with that of Nicholson *et al.* (1978) and Gregory & Scott (1969). The plot shows a good correlation between vD/V_{sl} and the mixture velocity. The correlation is given by

$$\frac{vD}{V_{sl}} = 7.59 \times 10^{-3} V_m + 0.01.$$

Figure 6 shows the variation of the holdup on the liquid film ahead of the slug with superficial gas velocity, for superficial liquid velocities ranging from 0.77 to 1.63 m/s. It can be seen that at superficial gas velocities below approx. 5 m/s, the holdup increases with an increase in the superficial liquid velocity. However, at higher superficial gas velocities, the holdup tends to a constant value of 0.23 for all superficial liquid velocities. This phenomena has also been observed by Lin & Hanratty (1987a, b) in 2.54 and 9.53 cm dia pipes, where they report a value of approx. 0.2.

The variation of holdup in the body of the slug with superficial gas velocity is given in figure 7 for the same conditions as figure 6. As the superficial gas velocity approaches zero, the holdup approaches unity. The holdup decreases rapidly to about 0.45 as the superficial gas velocity is increased to 5 m/s. As the gas velocity is increased a constant holdup of 0.38 is reached, irrespective of superficial liquid velocity. This is much lower than that predicted by the Gregory *et al.* (1978) correlation and other experimental data for small-diameter pipes.

The effect of pipe diameter on holdup in the slug body is shown in figure 8. The actual experimental data of Gregory *et al.* (1978), for 25.4 and 51.2 mm dia pipes for a liquid superficial

velocity of 0.91 m/s, are compared with the 300 mm data for a liquid superficial velocity of 0.97 m/s. It is clearly seen that as the pipe diameter is increased from 25.4 to 51.2 mm and to 300 mm, the constant value of holdup at high superficial gas velocities decreases from 0.55 to 0.45 and then 0.38, respectively. At low superficial gas velocities, < 3 m/s, there seems little effect of pipe diameter on holdup.

The ratio of the translational velocity of the slug to mixture velocity, V_i/V_m , is plotted against the superficial gas velocity in figure 9. Many workers (e.g. Dukler & Hubbard, 1975) have reported constant values of V_i/V_m of 1.25–1.3 for slug flow in small-diameter pipes. Crowley *et al.* (1984) report values of V_i/V_m of 2.0 and 1.35 for slug flow on a 17 cm dia pipeline, depending on the fluid properties. Figure 9 shows that at low superficial gas velocities (< 2 m/s) $V_i/V_m = 2.0$, but as the gas velocity is increased V_i/V_m decreases continuously from 2.0 to a constant value of 1.25. This constant value is attained when the superficial gas velocity is greater than approx. 14 m/s. The change from 2.0 to 1.25 occurs more rapidly at lower superficial liquid velocities. This behaviour has also been suggested by Kouba (1987), who examined slug flow in a 76.2 mm dia pipe.

The effect of pipe diameter on the length of the slug is illustrated in figure 10. The slug lengths for the 300 mm dia pipeline are compared with those given by Nicholson *et al.* (1978) and Kouba (1987) for 25.8, 5.12 and 76.2 mm dia pipelines. Figure 10 shows that, for all pipe diameters, as the superficial gas velocity is increased, the slug length increases. Further, as the pipe diameter increases, the slug length increases. However, at superficial gas velocities < 5 m/s, the data for the 300 mm dia pipe is very close to that of Kouba for the 76.2 mm pipe. This could be due to the fact that Kouba used kerosene and air rather than water and air but, in some recent qualitative studies on the new Harwell 150 mm dia clear Perspex pipeline, at superficial gas velocities < 5 m/s, the slugs seemed to be growing as they moved down the pipeline. At higher gas velocities, the slugs did not seem to grow.

The pressure gradient calculated from the pressure drop over a slug unit and the length of a slug unit are shown in figures 11–14. A slug unit is made up of the slug body itself plus the length of pipe occupied by the gas bubble between subsequent slugs. In each case, a data point represents an average of 50 measurements of pressure drop.

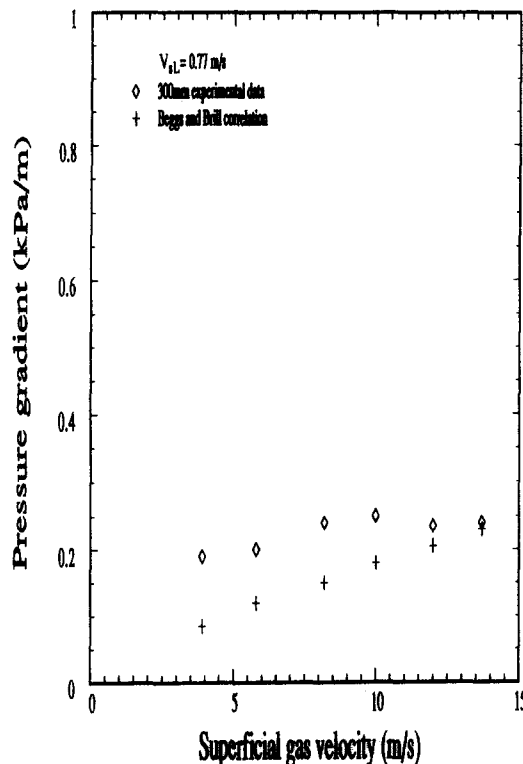


Figure 14. Comparison of the pressure gradient data with the existing correlation.

Figure 11 shows that the pressure gradient initially increases with an increase in both the liquid and gas superficial velocities, as expected. However, at higher superficial gas velocities, there is an indication of a decrease in the pressure gradient. This is believed to be due to the high degree of aeration and the start of blow-through.

Figures 12 and 13 compare the data for the 300 mm dia pipe with that of Nicholson *et al.* (1978) for superficial liquid velocities of 0.97 and 1.11 m/s, respectively. Both clearly show that increasing pipe diameter substantially decreases the pressure gradient when the diameter is changed from 25.4 to 51.2 mm, but the rate of decrease is not so great as when the pipe diameter is increased from 51.2 to 300 mm. This is probably due to the fact that the pipes used by Nicholson *et al.* were smooth plastic pipes, whilst the present pipes are commercial carbon steel pipes that have a larger roughness.

Figure 14 compares the pressure gradient in the 300 mm dia pipeline at a superficial liquid velocity of 0.97 m/s with that calculated by the correlations of Beggs & Brill (1973). There is reasonable agreement with the experimental data, even though the correlation does not take into account the actual length of a slug unit. It gives an average pressure gradient per unit length of pipeline, irrespective of the number of slugs in the pipeline. In the 300 mm pipeline, because of the low frequency of slugs, there were only two or three slugs in the pipeline at any instant. Consequently, the pressure gradient can only be measured over a slug unit rather than averaged over a long length of pipe. Further, the above correlations are derived mostly from data from small-diameter pipelines often made of smooth plastic rather than commercial metal pipes. The friction factor for metal pipes can be much larger than that of smooth bore tubes. There is no other readily available data relating to large-diameter pipelines.

4. CONCLUSIONS

There is a large pipe diameter effect on flow transitions. As the pipe diameter is increased, an increase in the superficial liquid velocity is needed to attain slug flow. Annular flow is attained at lower superficial gas velocities for large pipe diameters. The Taitel & Dukler (1976) model does not predict the effect of pipe diameter. The transition to annular flow agrees well with the Wallis & Dobson (1973), Simpson *et al.* (1981) and BP field trails. The stratified to slug flow transition follows the same trends as described by Lin & Hanratty (1987a, b) and Taitel & Dukler (1976).

Slug frequency decreases with an increase in the pipe diameter. The effect of pipe diameter can be accounted for by using the dimensionless group vD/V_{sL} when plotted against the mixture velocity.

The liquid film holdup tends to a limit of 0.24 as the superficial gas velocity is increased. Increasing the superficial liquid velocity increases the film holdup at lower superficial gas velocities.

The holdup in the liquid slug tends to a limit of 0.38 as the superficial gas velocity is increased. An increase in the pipe diameter decreases this limit. Increasing the superficial liquid velocity increases the slug holdup at lower gas velocities.

The ratio of the translational velocity to the mixture velocity, V_l/V_m decreases continuously from 2.0 at low gas velocities to 1.25 at high gas velocities. Increasing the superficial liquid velocity increases V_l/V_m at lower gas velocities.

The slug length increases with an increase in the pipe diameter.

The pressure gradient increases with increasing superficial gas and liquid velocities up to a point where the slugs become extremely frothy. Here the pressure gradient decreases. The pressure gradient decreases with an increase in the pipe diameter.

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