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LIQUID-PHASE AXIAL MIXING IN TWO-PHASE HORIZONTAL PIPE FLOW

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Abstract—The liquid-phase axial dispersion coefficient and volume-averaged fractional phase hold-ups have been measured in two-phase horizontal pipe flow. Radioactive ^{99m}Tc—technetium-99 metastable—(as an aqueous solution of sodium pertechnate) was used as a tracer. The pulse technique with two-point measurement was employed. Superficial gas (air) and liquid (water) velocities were varied in the range 20–2300 and 30–800 mm/s, respectively. The flow regimes covered were bubbly, elongated bubbly, stratified, wavy and slug. Experiments were also performed using single-phase pipe flow. The liquid-phase dispersion coefficient has been shown to depend upon the flow regime and the superficial gas and liquid velocities.

INTRODUCTION

Two-phase gas–liquid cocurrent flow occurs in many instances (Barnea & Taitel 1985; Delhaye 1983; Doraiswamy & Sharma 1984; Govier & Aziz 1972; Hetsroni 1982; Hewitt *et al.* 1982; Kakac & Mayinger 1977). From a reaction engineering point of view, pipeline reactors are useful when reactions are very fast. There are several examples of industrial importance, e.g. sulphonation, hydrohalogenation, halogenation, ammonolysis, absorption of NO_x, SO_x in alkaline solutions, ozonolysis etc., which can be conveniently carried out in pipeline reactors.

The fractional phase hold-up directly decides the volume of the reactor depending upon the phase in which the rate controlling step occurs. Indirectly, it governs the flow pattern and turbulence structure in the multiphase system. It is known that the turbulence structure is the fundamental parameter which controls other design variables.

Knowledge of liquid-phase axial mixing is important in addition to the kinetics and mass transfer coefficients. The degree of axial mixing decides the concentration profile within the reactor and hence the reactor volume.

In the present paper, a radioactive tracer technique has been used for measurement of liquid-phase hold-up and the liquid-phase dispersion coefficient. For the latter, radiotracers are convenient because: (i) this technique does not disturb the flow as the measurement is performed outside the equipment; (ii) the response time is very low as compared with the residence time. The second condition is not usually fulfilled by conventional tracers. For the measurement of hold-up, however, several techniques are available: these have been reviewed by Hewitt (1978).

In the present work a radioactive tracer technique has been used to measure the volume-averaged hold-ups of the phases and to determine their respective volume-averaged true velocities.

As regards liquid-phase axial mixing, although substantial information is available for the case of single-phase pipe flow, there is no data in the case of two-phase pipe flow. Therefore, investigation of liquid-phase axial mixing in two-phase pipe flow, covering most flow regimes, was undertaken.

EXPERIMENTAL SETUP

The experimental studies were performed on a 3.25 m long Perspex circular pipe of uniform i.d. 32 mm. The entry section to the pipe was a T-joint (figure 1) to which the lines for water and air were connected. The air supply was provided via a compressor through calibrated capillarimeters. Water was introduced through calibrated rotameters. The end of the test pipe section was

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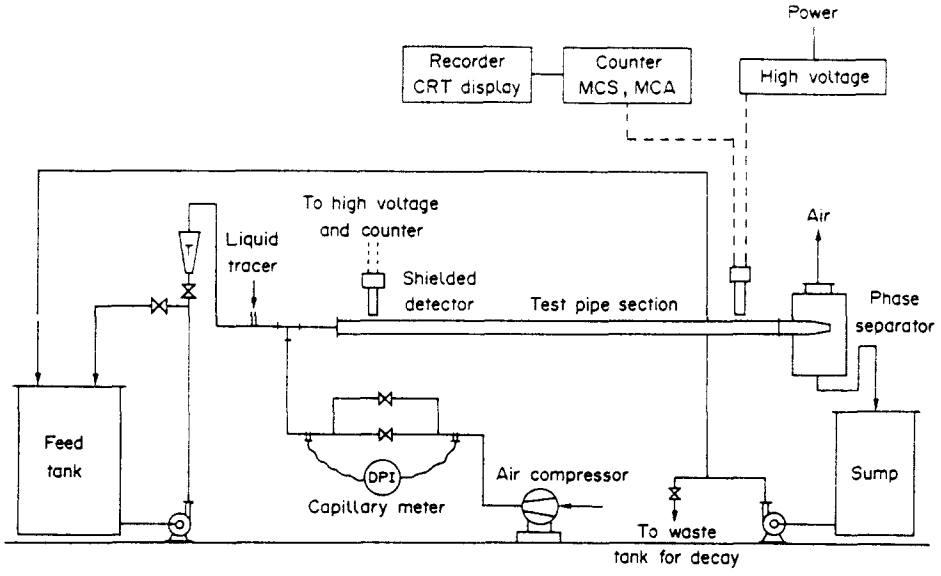

Figure 1. Experimental setup.

connected to a flow separator, wherein the gas disengaged from the liquid and exited via the top. The drained water was available for recirculation or disposal.

An injection point, a rubber cap on a nozzle, was provided downstream of the gas inlet to the T-joint. This point was used to inject desired quantities of the liquid tracer (a radioactive isotope).

Technetium-99 metastable (^{99m}Tc), as sodium pertechnate salt, was used as the radioactive tracer. Sodium pertechnate was extracted from the solution of sodium molybdate (^{99}Mo). ^{99m}Tc in the form of sodium pertechnate has a more convenient half-life ($t_{\frac{1}{2}} = 6$ h) than ^{99}Mo ($t_{\frac{1}{2}} = 66.7$ h). ^{99}Tc was "milked" from ^{99}Mo solution using methylethyl ketone as solvent. An aqueous solution of sodium pertechate (^{99m}Tc) was then used as the radioactive tracer.

^{99m}Tc is primarily a γ-ray emitting source of convenient energy—0.14 MeV. γ-Rays emitted from the tracer source in the system were counted using sodium iodide (thallium impurity) scintillation detectors.

The counting apparatus consisted of two BICRON scintillation detectors (model G1) and two BICRON "Analyst" counter-ratemeters with suitable recorder outputs. Output signals proportional to count rates from both counter-ratemeters were collected by a MINIGOR 50 recorder.

The two detectors were placed along the length of the test pipe section. Both detectors were shielded against any background activity using lead bricks. Lead collimators were used in order to expose only a slit length of the pipe section. The detectors were placed such that their central axis was normal to the pipe axis. The first detector was placed downstream of the tracer injection point; the second detector was further downstream (2.7 m) along the pipe length. After injection the tracer passed the first detector and then the second, resulting into two peak curves on the recorder. The two peaks were then analysed for mean residence time and residence time distribution. Typical recorder outputs are shown in figure 2.

DATA ANALYSIS

The output curves corresponding to concentration vs time curves from the two fixed-point detectors were analysed for residence time distribution on the basis of the dispersion model. The resultant net dispersion in flow between the two points was related to dispersion by the following equation:

$$\frac{\sigma_2^2 - \sigma_1^2}{\bar{t}^2} = \sigma_\theta^2 = 2\left(\frac{D}{uL}\right), \qquad [1]$$

where

$$\bar{t} = \bar{t}_2 - \bar{t}_1, \qquad [2]$$

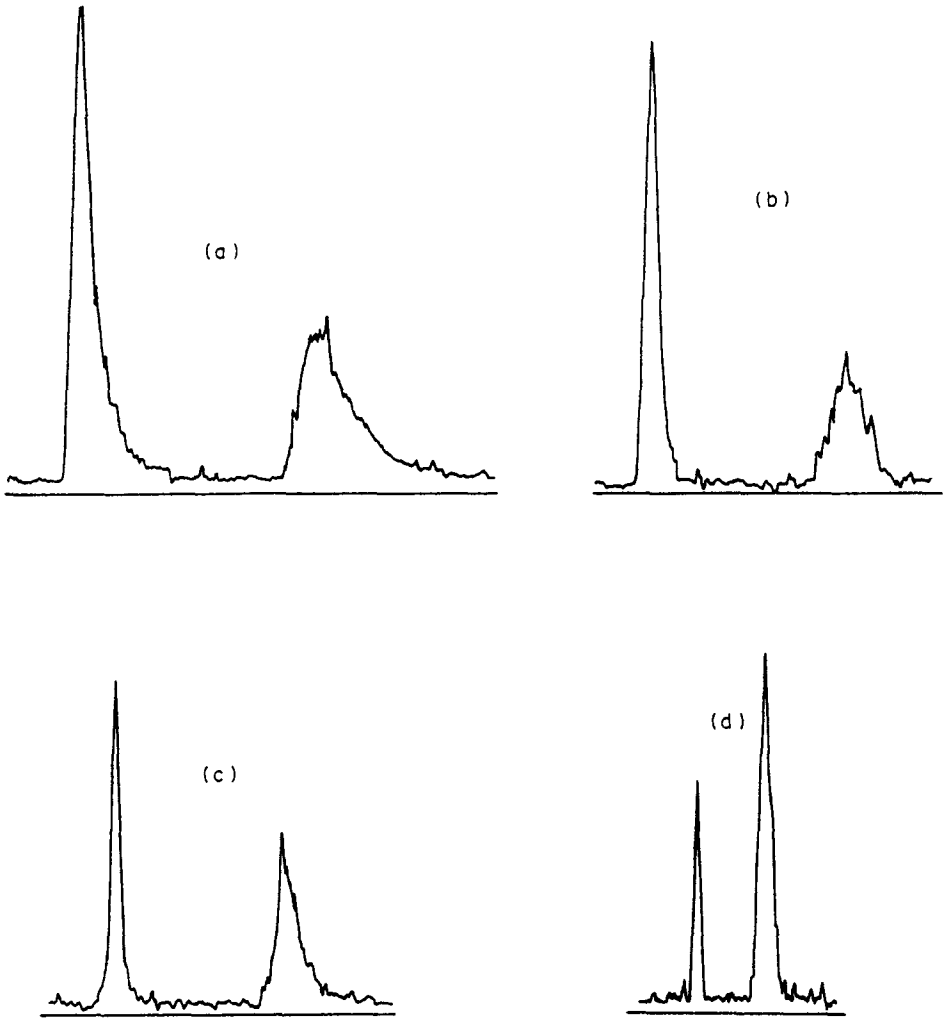

Figure 2. Typical response curves (at two locations) to pulse input: (a) stratified flow; (b) wavy flow; (c) high-amplitude wavy flow; (d) slug flow.

$$\bar{t}_i = \frac{\sum C_i t}{\sum C_i}, \qquad i = 1 \quad \text{for peak 1} \qquad [3]$$

$$i = 2 \quad \text{for peak 2}$$

and

$$\sigma_i^2 = \frac{\sum C_i t^2}{\sum C_i} - (\bar{t}_i)^2. \qquad [4]$$

The mean residence time was used to estimate the volume-averaged true flow velocities of both (gas and liquid) phases:

$$\frac{L}{\bar{t}_L} = V'_L \qquad [5]$$

and

$$\frac{L}{\bar{t}_G} = V'_G. \qquad [6]$$

The liquid-phase volume (v_L) was also estimated for each run,

$$Q_L \times \bar{t}_L = v_L. \qquad [7]$$

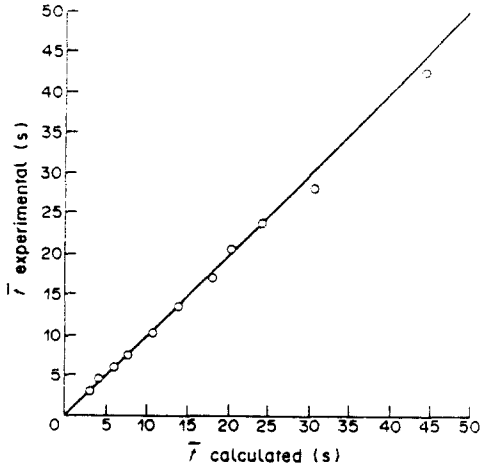

Figure 3. Mean residence time for single-phase liquid flow: comparison between the experimental and calculated values.

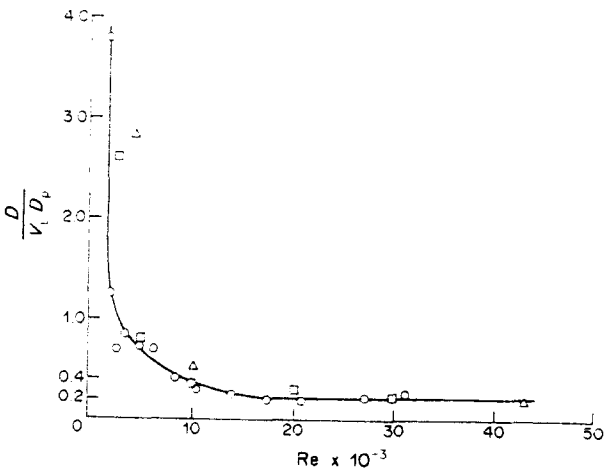

Figure 4. Liquid-phase axial dispersion coefficient in single-phase pipe flow: ○, experimental data from the present work; △, data from Tichacek *et al.* (1957); □, data from Levenspiel & Bischoff (1962).

Hence, the gas-phase volume (v_G),

$$v_G = v_T - v_L \qquad [8]$$

and

$$v_T = \frac{\pi}{4} D_p^2 L; \qquad [9]$$

and the gas mean residence time,

$$\bar{t}_G = \frac{v_G}{Q_G}. \qquad [10]$$

In [1]–[10] the subscripts 1 and 2 represent the response curves at the two points (figure 2); σ^2 and $\bar{t}$ are the variance and mean, respectively; L is the pipe length; $\bar{t}_L$ and $\bar{t}_G$ are the liquid- and gas-phase residence times, respectively; V'_L and V'_G are the true velocities of the liquid and gas phases, respectively; D_p is the pipe diameter; D is the liquid-phase axial dispersion coefficient; and Q_L and Q_G are the volumetric flow rates of the liquid and gas, respectively.

SINGLE-PHASE LIQUID FLOW

Axial dispersion and mean residence times were estimated for single-phase water flow through the test pipe section. Mean residence times based on average velocity, and on the mean residence time from the two detector output curves, are compared in figure 3. The axial dispersion number, $D/V_L D_p$, determined experimentally using a two-point measurement technique, is compared with data from Tichacek *et al.* (1957) and Levenspiel & Bischoff (1962) in figure 4. The agreement is excellent. Thus, figures 3 and 4 indicate the suitability of using the present experimental procedure for the measurement of average residence time and the dispersion coefficient.

TWO-PHASE AIR–WATER FLOW

Two-phase gas–liquid experiments were carried out in a number of sets. In each set the gas volumetric flow rate was kept unchanged while the liquid volumetric flow rate was varied in fixed-range intervals. The superficial gas (air) velocity was thus varied in the range 20–2300 mm/s and the superficial liquid (water) velocity was varied over 30–800 mm/s in each set.

Figure 5 shows fractional liquid-phase hold-up (ϵ_L) data vs superficial liquid velocity (V_L) with superficial gas velocity (V_G) as a parameter. Up to $V_G = 286$ mm/s, the flow pattern was observed to be stratified up to a certain value of V_L. In this region ϵ_L was noted to vary almost linearly with

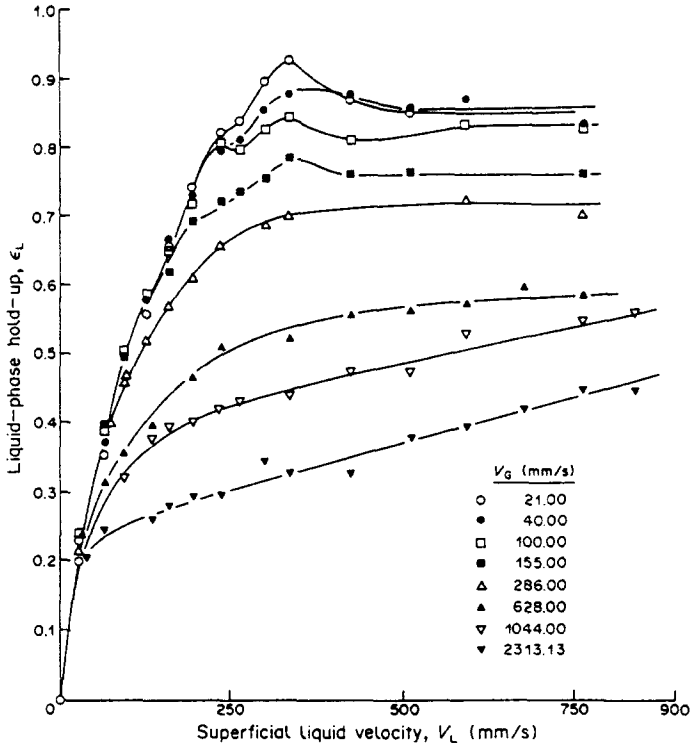

Figure 5. Effect of superficial liquid and gas velocities on fractional liquid-phase hold-up.

V_L. This is interesting since it means that the true liquid-phase velocity V'_L remains practically constant with increasing V_L, whereas V'_G increases. Increasing V_L beyond a certain value cannot keep V'_L constant and a flow transition occurs.

It can be seen from figure 5 that, whenever the final prevailing regime is that of bubbly or plug flow, ϵ_L seems to remain practically constant and independent of V_L.

At very high gas velocities, such as 628, 1045 and 2313 mm/s, stratified flow did not occur. At lower values of V_L, intense wavy flow was observed which transformed into plug flow for $V_G = 628$ mm/s and slug flow for $V_G = 1045$ and 2313 mm/s, at higher liquid velocities. It may be of note that in slug flow (▲, ▽ and ▼ plots in figure 5) ϵ_L increases linearly with V_L. It is further interesting to note that the slope of ϵ_L vs V_L at different gas velocities is almost constant. The flow patterns observed in this work agree with the flow map reported by Spedding & Nguyen (1980).

LIQUID-PHASE AXIAL DISPERSION IN TWO-PHASE FLOW

Liquid-phase axial dispersion was computed using the axial dispersion model. The dispersion number, $D/V_L D_p$, was estimated from the variance data on two output curves from two detectors placed along the length of the test pipe. The convolution technique, as discussed by Levenspiel (1972), was used to obtain the net variance of dispersion, and hence the dispersion number.

Figures 6–8 illustrate the behaviour of the dispersion number, $D/V_L D_p$, with respect to the liquid-phase superficial Reynolds number (Re_L) with superficial gas velocity (V_G) as a parameter. These figures depict the experimental observations.

The value of the dispersion number was found to be strongly dependent on the flow regime. In particular, the following observations were made:

(1) At low gas and liquid superficial velocities, when the flow is stratified, the dispersion number decreases with V_L.
(2) The beginning of the transition to bubble, plug or wavy flow is marked by a sudden jump in the dispersion number.
(3) For $V_G = 21$ and 40 mm/s, when the flow pattern smoothly transforms from

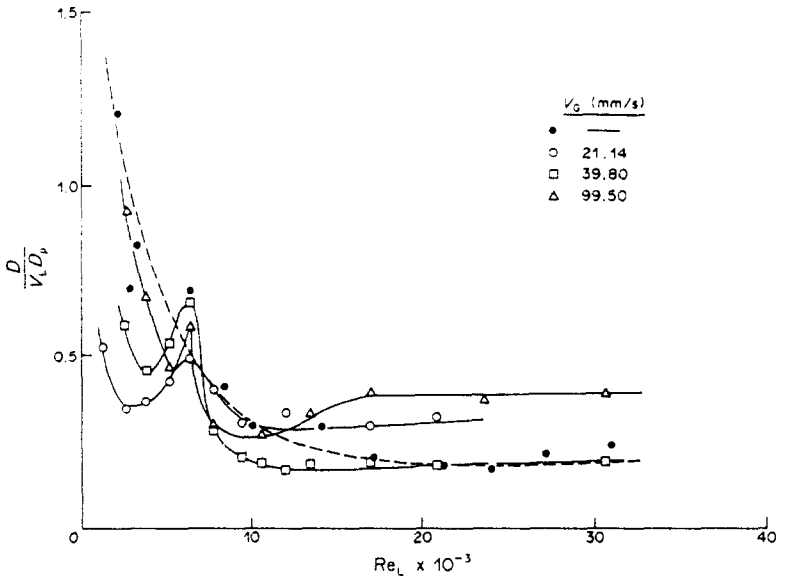

Figure 6. Effect of superficial gas and liquid velocities on liquid-phase axial dispersion.

stratified into bubbly or plug flow, the dispersion number decreases with V_L after the maximum and finally attains a constant value (figure 6).

(4) At $V_G = 100$, 155 and 286 mm/s, the transition from stratified to plug occurs via wavy flow. For instance, in figure 7, portion AB depicts the stratified, BC depicts the change from stratified to wavy and CD shows the development of plugs. Once the plug flow is developed, $D/V_L D_p$ remains constant and independent of V_L (portion DE in figure 7).

(5) At $V_G = 628$, 1044 and 2312 mm/s, stratified flow did not occur. Therefore, the first maximum, which corresponds to the transition from stratified flow, is not observed at these high gas velocities. A typical behaviour at these gas velocities is shown in figure 8. The portion AB corresponds to wavy flow where $D/V_L D_p$ decreases with an increase in V_L. The development of slugs occurs during the region BCD where the point C shows a maximum value. Once the slug flow is developed (point D), the dispersion number is independent of V_L and practically remains constant, as shown by portion DE in figure 8.

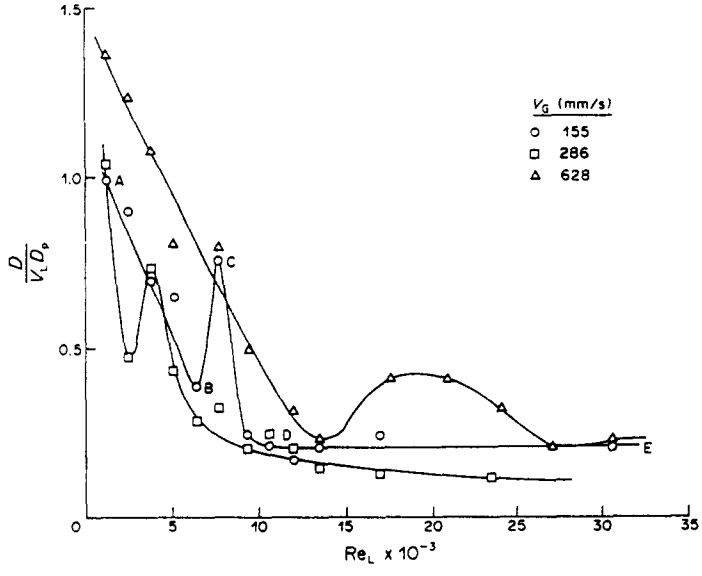

Figure 7. Effect of superficial gas and liquid velocities on liquid-phase axial dispersion.

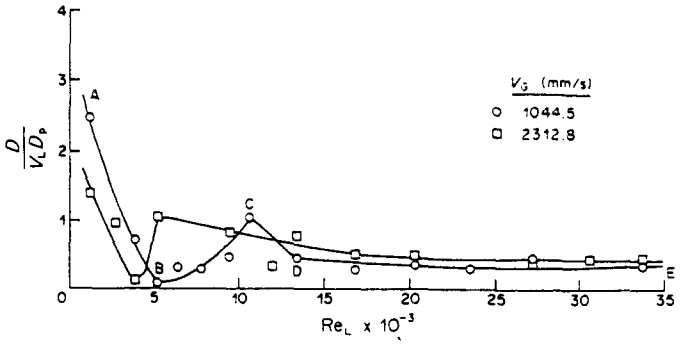

Figure 8. Effect of superficial gas and liquid velocities on liquid-phase axial dispersion.

(6) Whenever, wavy flow occurred, dispersion was found to be higher as compared with that observed in any other regime.
(7) In plug flow and slug flow, $D/V_L D_p$ is constant at high values of V_L. Similar observations are made in single-phase pipe flow as well. However, the final slug flow dispersion number (0.4) is higher than that for single-phase pipe flow, whereas the final plug flow dispersion number (0.2) is of the order of that for single-phase pipe flow.

CONCLUSIONS

(1) A radioactive tracer technique has been used successfully for the measurement of phase hold-ups and the liquid-phase axial dispersion coefficient.
(2) The liquid-phase axial dispersion coefficient depends on the flow regime. Explanation has been provided for the dispersion behaviour in different flow regimes.

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