Int. J. Multiphase Flow Vol. 18, No. 6, pp. 847–860, 1992
0301-9322/92 \$5.00 + 0.00

PREDICTION OF THE SLUG-TO-CHURN FLOW TRANSITION IN VERTICAL TWO-PHASE FLOW

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(Received 1991; in revised form 14 August 1992)

Abstract—An assessment is made of the various viewpoints on the slug-to-churn flow transition in vertical upward flow in the light of recent experimental results obtained at Harwell Laboratory. It is found that the flooding model of McQuillan & Whalley and the bubble entrainment model of Barnea & Brauner give satisfactory results at low and high liquid flow rates, respectively. An improved model for flooding, which takes account of the effect of the falling film, has been proposed. It is shown that this new model is in good agreement with experimental results at both low and high liquid flow rates.

Key Words: vertical gas–liquid flow, slug flow, churn flow, transition mechanism, flooding, bubble entrainment

1. INTRODUCTION

As the name implies, churn flow (also referred to as froth flow and semi-annular flow) is a highly disturbed flow of gas and liquid in which the liquid motion is oscillatory, going up and down alternately although not in a periodic and regular manner. The net flow of liquid is generally in the direction of the flow of gas, although it may be zero or even negative. This flow regime normally occurs between the slug and annular flow regimes in vertical or near-vertical upward flow. However, it may not occur at very high liquid flow rates, where a direct transition to an annular flow pattern may occur either from slug flow or bubbly flow patterns.

Churn flow is unusual among the flow regimes in that several schools of thought exist as to the mechanism of transition to this flow regime. It is also comparatively difficult to predict its occurrence. For example, McQuillan & Whalley (1985) reported only a 36% success rate in predicting it, whereas other flow regimes were predicted with an accuracy of up to 80%. Others, for example, Mishima & Ishii (1984) and Brauner & Barnea (1986) have not used such an index to measure the efficacy of their models, but even in their results, several instances of incorrect prediction of this flow regime can be found, and their success rate is also rather poor. It is possible that this poor agreement is due to one or both of the following: the predictive methods do not model the underlying mechanisms to a sufficient degree of sophistication; and secondly, there may be more than one mechanism, over a range of flow conditions, governing this transition. To the best of our knowledge, neither of these possiblities has yet been investigated rigorously, and it is the purpose of the present paper to examine both possibilities in the light of previoulsy unpublished experimental results obtained at Harwell Laboratory. These data take the form of pressure gradient measurements and visually-observed flow regime transitions in air–water flow at near-atmospheric pressures in a long vertical tube, and are eminently suited to test the efficacy of the various models and predictive methods.

The organization of the rest of the paper is as follows. In section 2, four models for the slug-to-churn transition are described. The predictions of these models are compared in section 3 against the experimental results of Owen (1986). In section 4, the models are reassessed in the light of this comparison, and some modifications to the models are investigated to improve their predictions. The overall conclusions are given in section 5.

2. VARIOUS MECHANISMS OF TRANSITION

There is extensive literature on the transition criteria for various flow patterns. For the transition from slug to churn flow, there appear to be four major schools of thought. These are:

(a) Entrance effect mechanism.
(b) Flooding mechanism.
(c) Wake effect mechanism.
(d) Bubble coalescence mechanism.

There have been several attempts at modelling these mechanisms, and four recent models are taken here for comparison with experimental results. The various mechanisms are described below with reference to these models.

2.1. Entrance effect mechanism (Dukler & Taitel 1986)

Dukler & Taitel (1986) as well as Taitel *et al.* (1980) treat churn flow as an entrance phenomenon. They see it as part of the process of the formation of stable slug flow further downstream in the pipe. According to them, at the inlet where gas and liquid are introduced, short liquid slugs and Taylor bubbles are formed. However, since short liquid slugs are unstable, they collapse down the tube, coalesce with the next slug and form a bigger slug which is able to retain its identity longer before it too collapses. If the pipe is long enough, this collapse and merging of successive slugs will ultimately lead to liquid slugs long enough to become stable. The flow in the intervening region (i.e. that between the entrance and the point where stable slugs are formed) appears to be churn flow because of the oscillatory motion of the rising and collapsing liquid slugs. The basis for this scenario of Dukler and co-workers are the experimental studies of the slug and churn flows conducted in their laboratories in pipes of diameters of 0.025 and 0.05 m.

Dukler & Taitel (1986) also present a formula to evaluate the entrance length, l_e, required to form stable slugs in a given flow situation:

$$\frac{l_e}{D} = 42.6\left(\frac{U_m}{\sqrt{gD}} + 0.29\right), \tag{1}$$

where U_m is the mixture velocity, i.e. the sum of the superficial velocities of the two phases, g is the acceleration due to gravity and D is the tube inner diameter. If the actual pipe length is less than l_e calculated using [1], then churn flow may be observed in the entire pipe; otherwise, slug flow will be observed at the end.

2.2. Flooding mechanism (McQuillan & Whalley 1985)

McQuillan & Whalley (1985) and several others, including Nicklin & Davidson (1962), Wallis (1969) and Govan *et al.* (1991), attribute the slug-to-churn transition to the flooding of the liquid film surrounding the Taylor bubble in slug flow (see figure 1). Flooding is a phenomenon in which the liquid film in countercurrent flow of gas and liquid breaks down due to the formation of large interfacial waves. The link between flooding and churn flow has been established experimentally in at least some cases. For example, Wallis (1961) observed that the air velocity required to initiate flooding in a falling film of very small flow rate is roughly the same as that at the slug-to-churn flow transition. Similarly, in a flooding experiment in which the water feed was at a midway point rather than at the top, Chaudhry *et al.* (1965) observed that the flow above the point of water injection was typical of churn flow when flooding occurred in the pipe.

Several correlations/calculation methods exist for the prediction of flooding velocities (e.g. Wallis 1969; Govier & Aziz 1972; Bankoff & Lee 1986). One of these is the following semi-empirical equation proposed by Wallis (1961) and Hewitt & Wallis (1963):

$$\sqrt{U^*_{Gs}} + \sqrt{U^*_{Ls}} = C, \tag{2}$$

where C is a constant close to unity and U^*_{Gs} and U^*_{Ls} are the non-dimensionalized gas and liquid superficial velocities, U_{Gs} and U_{Ls}, respectively, and are defined as

$$U^*_{Gs} = U_{Gs}\frac{\sqrt{\rho_G}}{\sqrt{gD(\rho_G-\rho_L)}} \quad \text{and} \quad U^*_{Ls} = U_{Ls}\frac{\sqrt{\rho_L}}{\sqrt{gD(\rho_L-\rho_G)}}. \qquad [3]$$

Here ρ_G and ρ_L are the densities of the gas and the liquid phases, respectively. In view of the link between flooding and the slug-to-churn transition, McQuillan & Whalley (1985) proposed that the latter can be calculated using [2] with the characteristic phase velocities as those of the superficial Taylor bubble (U_{bs}) and the superficial liquid film (U_{fs}) (because flooding would occur in this region) evaluated as follows:

$$U^*_{bs} = U_{bs}\frac{\sqrt{\rho_G}}{\sqrt{gD(\rho_G-\rho_L)}} \quad \text{and} \quad U^*_{fs} = U_{fs}\frac{\sqrt{\rho_L}}{\sqrt{gD(\rho_G-\rho_L)}}. \qquad [4]$$

U_{bs} and U_{fs} are calculated using the following relations obtained using the simple model shown in figure 1:

$$U_{bs} = \left(1-4\frac{\delta}{D}\right)\left[1.2(U_{Gs}+U_{Ls})+0.35\sqrt{\frac{gD(\rho_L-\rho_G)}{\rho_L}}\right] \qquad [5]$$

and

$$U_{fs} = U_{bs} - (U_{Gs} + U_{Ls}), \qquad [6]$$

where δ is the liquid film thickness in the Taylor bubble region, and is evaluated using Nusselt's expression for a laminar falling film (Nusselt 1916):

$$\delta = \left[\frac{3U_{fs}D\mu_L}{4g(\rho_L-\rho_G)}\right]^{1/3}, \qquad [7]$$

where μ_L is the dynamic viscosity of the liquid. The constant C in [2] depends on many factors, such as the tube end conditions and tube length [see, for example, the review article by Bankoff & Lee (1986)], but is normally around unity. McQuillan & Whalley propose that C = 1.

2.3. *Wake effect of Taylor bubbles (Mishima & Ishii 1984)*

Mishima & Ishii (1984) attribute the collapse of the liquid slug to the wake effect of Taylor bubbles. At a point close to the slug–churn transition, the liquid slug would be very short, and the Taylor bubbles would be very close to each other. This would create a strong wake effect which would destabilize the liquid slug and destroy it. Under these conditions, the mean void fraction over the Taylor bubble region would be equal to the average void fraction in the pipe. Mishima & Ishii (1984) therefore state their criterion as follows: the transition from slug flow to churn flow

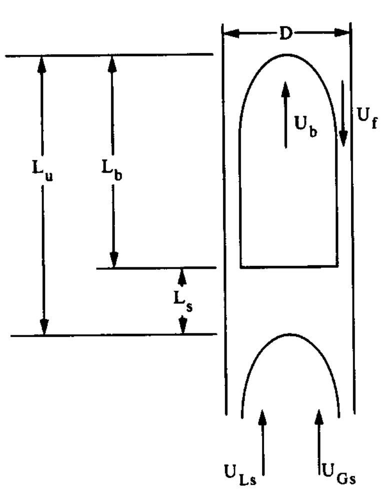

Figure 1. Model of slug flow of McQuillan & Whalley (1985).

occurs when the void fraction in the pipe (ϵ_{avg}) is just greater than the mean void fraction over the Taylor bubble region (ϵ_b). They evaluate the former using the standard drift flux model, and obtain a new expression for the latter using potential flow analysis:

$$\epsilon_{avg} = \frac{U_{Gs}}{\left[C_0 U_m + 0.35\sqrt{\frac{gD(\rho_L - \rho_G)}{\rho_L}}\right]} \quad [8]$$

and

$$\epsilon_b = 1 - 0.813\left[\frac{(C_0 - 1)U_m + 0.35\sqrt{\frac{\Delta\rho g D}{\rho_L}}}{U_m + 0.75\sqrt{\frac{\Delta\rho g D}{\rho_L}}\left(\frac{\Delta\rho g D^3 \rho_L}{\mu_L^2}\right)^{1/18}}\right]^{3/4}; \quad [9]$$

where

$$\Delta\rho = \rho_L - \rho_G,$$

$$C_0 = 1.2 - 0.2\sqrt{\frac{\rho_G}{\rho_L}} \quad \text{for round tubes} \quad [10a]$$

and

$$C_0 = 1.35 - 0.35\sqrt{\frac{\rho_G}{\rho_L}} \quad \text{for rectangular ducts.} \quad [10b]$$

2.4. Slug collapse by bubble coalescence (Brauner & Barnea 1986)

Brauner & Barnea (1986) attributed the transition to churn flow from slug flow to the formation of highly aerated liquid slugs. According to them, the gas phase is entrained in the liquid slugs and is kept in the form of dispersed bubbles because of the turbulence within the slug. When the void fraction in the liquid slug reaches a value of 0.52, the separation between bubbles becomes too small for them not to collide with adjacent bubbles, and as a result they coalesce. This destroys the identity of the liquid slug and leads to churn flow. Thus, transition to churn flow occurs when the void fraction in the liquid slug is >0.52. It should be noted that this limiting void fraction is much less than the maximum packing fraction of 0.74 for spherical bubbles of uniform size. It is assumed that bubble coalescence will occur well before this limiting void fraction is reached because the probability of collision between, and of the coalescence of, bubbles increases rapidly as the bubbles come closer.

Brauner & Barnea (1986) used a model developed by Barnea & Brauner (1985) for predicting the liquid holdup in the slug. It is based on the assumption that the gas in the liquid slug behaves as dispersed bubbles on the verge of transition to slug flow. The final expression for the void fraction in the slug (ε_s) for given flow conditions is as follows:

$$\varepsilon_s = 0.058\left[d_c\left(\frac{2f_m U_m^3}{D}\right)^{0.4}\left(\frac{\rho_L}{\sigma}\right)^{0.6} - 0.725\right]^2, \quad [11]$$

where d_c is the characteristic bubble size which for vertical flow is given by (Barnea 1987):

$$d_c = 2\sqrt{\frac{0.4\sigma}{g\,\Delta\rho}}. \quad [12]$$

Here σ is the surface tension and f_m is the friction factor based on the mixture velocity, U_m, and is defined as

$$f_m = \frac{2\tau_w}{\rho_L U_m^2},$$

where τ_w is the wall shear stress. The friction factor is obtained from the usual f vs Re relationships, with the Reynolds number being defined as

$$\mathrm{Re} = \frac{\rho_L U_m D}{\mu_L};$$

$$f = \frac{16}{\mathrm{Re}} \qquad \text{if Re} \leqslant 2100,$$

$$f = 0.046\mathrm{Re}^{-0.2} \quad \text{if Re} > 2100.$$

2.5. *Comments on the proposed mechanisms and models*

There are significant differences in the four views on the transition from slug flow to churn flow. Briefly, Dukler & Taitel (1986) see churn flow as a transient, developing phenomenon which will eventually lead to stable slug flow in a pipe of sufficient length. McQuillan & Whalley (1985) attribute it to the flooding of the liquid film in the Taylor bubble region and present a model to evaluate the corresponding flooding velocities. Both Mishima & Ishii (1984) and Barnea & Brauner (1985) cite excessive bubble entrainment as the reason for the transition. However, the former propose a limiting value for the void fraction in the pipe, whereas the latter propose one for the void fraction in the slug. In the next section, we examine if and to what extent these mechanisms and models are consistent with some recent experimental results.

3. COMPARISON WITH EXPERIMENTAL RESULTS

3.1. *Description of the experimental results*

The experimental results used to test the various models were obtained by Owen (1986) as part of his Ph.D. studies at Harwell Laboratory. The experiments were conducted on the LOTUS facility, the main feature of which is a 20 m long test section of 0.0318 m i.d. Facilities were available to measure the inlet flow rates of air and water, the total pressure gradient and other important flow parameters such as the void fraction, wall shear stress, film thickness, film flow rate in the test section. A visualization section was located at about 18 m from the point where the two-phase flow began.

Owen (1986) conducted flow visualization experiments over a range of air and water flow rates at a nominal test section pressure of 2.4 bar. He identified visually five flow regimes, namely, bubbly, slug, churn, annular, wispy-annular flows. The results are given in table 1 in the form of the range of gas flow rates over which flow regime transitions occur at a given liquid flow rate. It is seen that the transitions from bubbly and slug flow regimes are fairly sharp, while those from churn to annular or to wispy annular are spread over a broad range of gas flow rates.

Owen (1986) also measured the total pressure gradient in the pipe; these results are reproduced in figures 2–4. In these figures, the non-dimensionalized pressure gradient is plotted against the non-dimensionalized superficial air flow rate for low, medium and high water flow rates. Also marked on the figures are the various flow regimes detected visually. It can be seen that in the case of churn flow, which is of particular interest to us, there are significant differences in the three figures. At low liquid flow rates (figure 2), the transition to churn flow is accompanied by a large

Table 1. Owen's (1986) results of flow regime transitions in air–water vertical upward flow at a test section pressure of 2.4 bar; tube i.d. = 0.0318 m

Liquid mass flux ($\mathrm{kg/m^2 \cdot s}$)	Gas mass flux ($\mathrm{kg/m^2 \cdot s}$) at the flow regime transition between				
	Bubbly & slug	Slug & churn	Churn & wispy annular	Churn & annular	Wispy annular & annular
5.3	5.8	6.5–8.0	No transition	22.7–46.1	No transition
10.5	5.9	7.1–9.0	No transition	19.4–51.9	No transition
49.5	5.9	8.2–9.6	No transition	22.1–40.1	No transition
111.8	5.8	9.6–11.3	No transition	25.1–49.9	No transition
199	4.6	6.6–8.4	No transition	18.2–49.9	No transition
297	4.6	9.0–14.3	18.9–38.2	No transition	50.7–80.6
399	—	11.3–13.4	20.9–54.9	No transition	59.9–110

and sudden increase in the pressure gradient. (This may be significant because a similar jump in the pressure gradient occurs after flooding in countercurrent flow.) Such a jump is present in the intermediate liquid flow rate case (figure 3); however, it is much less pronounced. In the high liquid flow rate case, shown in figure 4, there is no jump in the pressure gradient in this region and the transition is much more gradual. In the rest of this section, we will examine if the various models are consistent with these observations.

3.2. Model of Dukler & Taitel (1986)

According to this model, churn flow is essentially a developing slug flow, and stable slug flow will be formed if the length of the pipe is greater than that given by [1]. For the case of Owen's (1986) experiments, this length is shown in table 2 for the several combinations of air and water flow rates. The required developing length varies between 200 and 360 pipe diameters (6.3–11.5 m); beyond this length, the flow should be slug flow. However, in all these cases, Owen observed churn flow at a distance of 560 diameters (18 m) downstream of the beginning of two-phase flow. Since this is far greater than the developing length required, we conclude that the model of Dukler & Taitel (1986) is incorrect, at least for these data.

It may be argued that while the model is quantitatively incorrect, the underlying scenario may yet be correct qualitatively in the sense that the churn flow observed at the end of the 18 m long section may still be developing, and that given sufficient further length, stable slug flow will prevail.

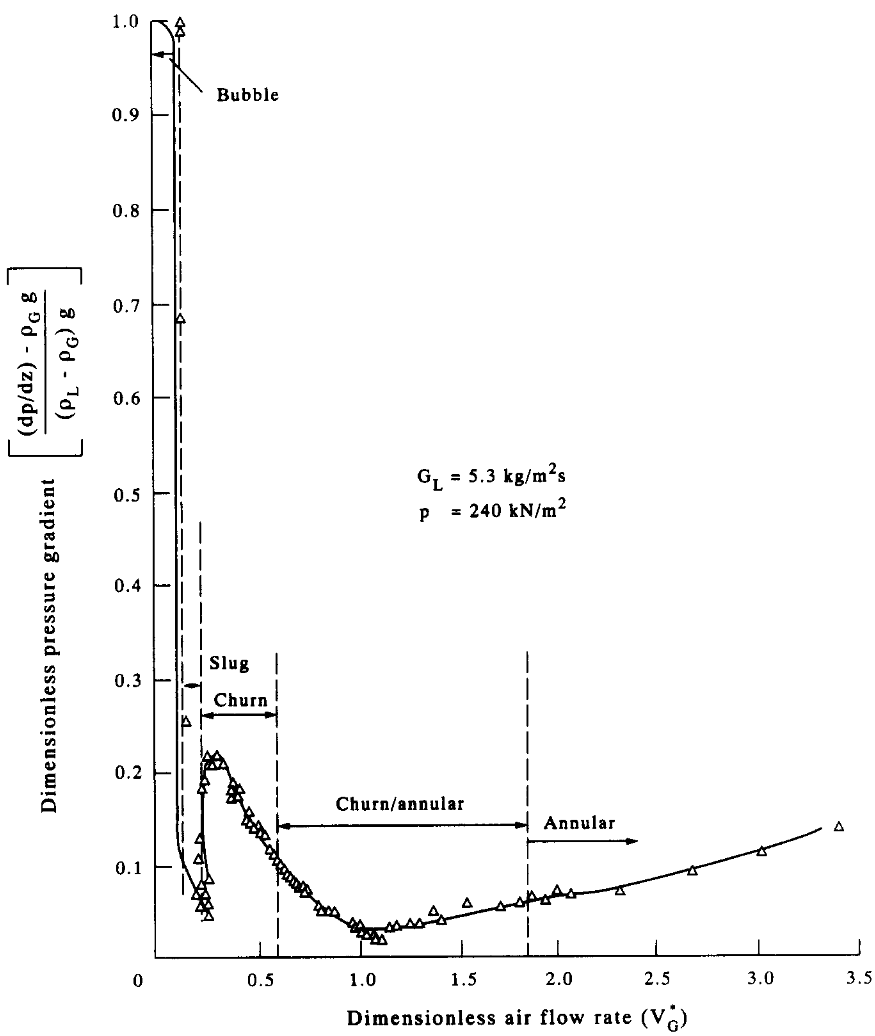

Figure 2. Measured pressure gradient as a function of gas flow rate in vertical two-phase flow; experimental results of Owen (1986) at low liquid flow rates.

Table 2. Developing length for slug flow according to the model of Taitel & Dukler (1986)

Liquid mass flux ($kg/m^2 \cdot s$)	Gas mass flux at transition ($kg/m^2 \cdot s$)	l_e/D	l_e (m)
5.3	7.2	203	6.5
10.5	8.0	225	7.2
49.5	8.9	252	8.0
111.8	10.4	284	9.0
199	7.5	214	6.8
297	11.6	342	10.9
399	12.2	366	11.6

Owen's (1986) results do not throw any light on this, although it may be pointless, from a practical point of view, to consider the flow to be still developing even after 600 diameters because such long vertical pipe sections are rarely encountered in real situations.

3.3. Model of McQuillan & Whalley (1985)

McQuillan & Whalley (1985) associate the transition to churn flow with the flooding of the liquid film in the Taylor bubble. The calculation of the critical gas velocity for transition is summarized in table 3. The flooding correlation [2] contains a constant C, which should have a value of about unity. Table 3 compares the critical gas velocities calculated for $C = 0.88$, 1.0 and 1.2. It can be

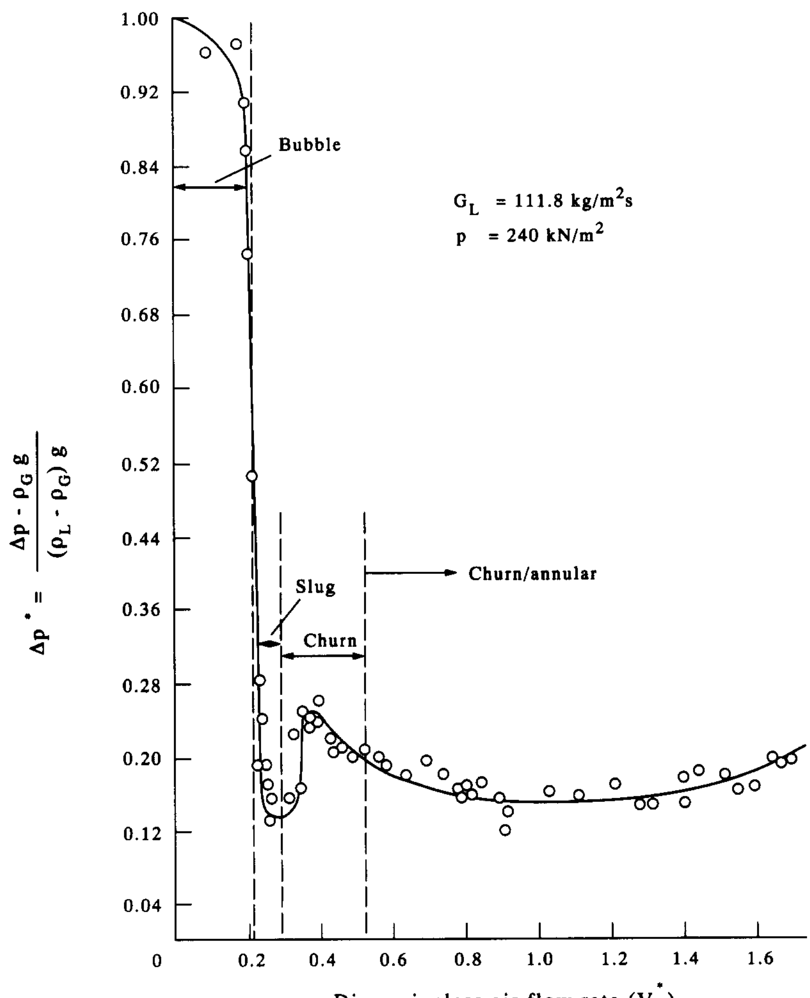

Figure 3. Measured pressure fradient as a function of gas flow rate in vertical two-phase flow; experimental results of Owen (1986) at intermediate liquid flow rates.

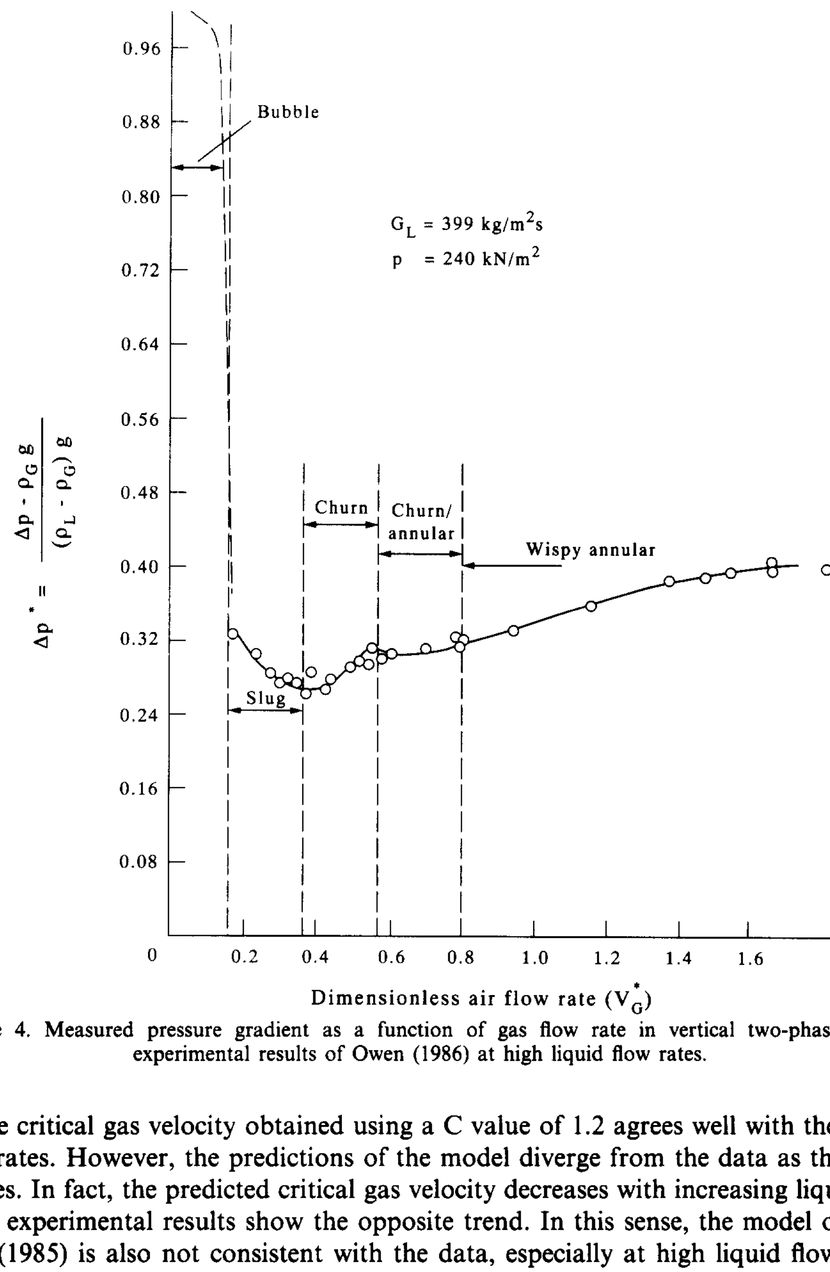

Figure 4. Measured pressure gradient as a function of gas flow rate in vertical two-phase flow; experimental results of Owen (1986) at high liquid flow rates.

seen that the critical gas velocity obtained using a C value of 1.2 agrees well with the data at low liquid flow rates. However, the predictions of the model diverge from the data as the liquid flow rate increases. In fact, the predicted critical gas velocity decreases with increasing liquid flow rate, whereas the experimental results show the opposite trend. In this sense, the model of McQuillan & Whalley (1985) is also not consistent with the data, especially at high liquid flow rates.

However, this should not rule out flooding as a probable mechanism of the transition to churn flow. Flooding is a complicated phenomenon and it depends on several factors not taken in the quantitative model used by McQuillan & Whalley. Examples of these are the inlet and outlet configurations and the effect of pipe length, both of which are relevant to the present case. A

Table 3. Comparison of the experimental gas mass flux at the transition to churn flow with that predicted by the flooding model of McQuillan & Whalley (1985)

Liquid mass flux (kg/m$^2\cdot$s)	Gas mass flux (kg/m$^2\cdot$s) at transition obtained			
	From data	With C = 1.2	With C = 1.0	With C = 0.88
5.3	7.2	5.8	2.8	1.5
10.5	8.0	5.7	2.7	1.5
49.5	8.9	5.6	2.6	1.4
111.8	10.4	5.5	2.5	1.2
199	7.5	5.2	2.3	0.9
297	11.6	4.9	2.0	0.65
399	12.2	4.6	1.7	0.35

discussion of these factors is postponed to section 4. It is only noted here that a point in favour of the flooding mechanism is the sudden increase in the pressure gradient following (or close to) the transition to churn flow (see figures 2 and 3). This occurs only at low liquid flow rates, and makes it plausible that the slug-to-churn transition is related to flooding. Its effect, if any, at high liquid rates, appears to be masked by other factors/mechanisms.

3.4. Model of Mishima & Ishii (1984)

Mishima & Ishii (1984) propose that the transition to churn flow occurs when the mean void fraction over the entire slug unit (consisting of a liquid slug and a Taylor bubble) exceeds that in the Taylor bubble region. The calculations of this model are summarized in table 4. It is seen that the predicted gas flow rate for the transition agrees well with experimental data for low and medium liquid flow rates, but that it is overpredicted at high liquid flow rates, and that the critical gas flow rate increases monotonically with increasing liquid flow rate.

While the agreement between experiment and theory is excellent (except at high liquid flow rates) as far as the critical gas flow rate is concerned, there is a more fundamental deficiency in this model: the transition criterion implies that the average void fraction in the liquid slug must be significantly greater than that in the Taylor bubble region. This can be seen from the following relation among the void fraction in the slug unit, ε_u, that in the liquid slug section, ε_s, and that in the Taylor bubble region, ε_b:

$$\varepsilon_u = \varepsilon_s(1-\beta) + \varepsilon_b\beta; \quad \text{where } \beta = \frac{L_b}{L_b + L_s}. \qquad [13]$$

Since $0 < \beta < 1$, ε_u can be greater than ε_b only if $\varepsilon_s > \varepsilon_b$. This appears unlikely in reality because the latter itself is generally very high. For example, typical values of this quantity for Owen's (1986) experimental conditions are shown in the last column of table 4. These have been evaluated using an expression developed by Mishima & Ishii (1984) ([9] above). It can be seen that the void fraction in the Taylor bubble region is about 0.78 in all cases. [This is slightly less than the value of about 0.85 reported by Orrel & Rembrand (1986).] This means that the void fraction in the liquid slugs must be higher than this (because the Taylor bubbles are normally longer than the liquid slugs). However, according to Brauner & Barnea (1986), among others, the limiting void fraction is 0.52 for liquid slugs. Indeed, it is physically impossible to have a volume packing fraction of 0.78 for fairly spherical bubbles of uniform diameter. This inconsistency casts doubt on the accuracy of the model. It should be noted that Mishima & Ishii (1984) do not make any attempt to evaluate the void fraction in the liquid slug, and instead use the standard drift flux model to evaluate the mean void fraction in the pipe even at the slug-to-churn transition point. Their model to determine the void fraction in the Taylor bubble has also been questioned by McQuillan & Whalley (1985). Thus this model, while showing good superficial agreement with the experimental data of Owen (1986), lacks a direct mechanism as basis (the criterion expresses a void fraction limit and is not related explicitly to the wake effect of Taylor bubbles), and is inconsistent in its detail.

3.5. Model of Brauner & Barnea (1986)

In this model, it is assumed that a transition to churn flow occurs when the void fraction in the liquid slug reaches a limiting value of 0.52. The predicted critical gas flow rates for the various liquid

Table 4. Comparison of the gas mass flux ($kg/m^2 \cdot s$) at the transition to churn flow with the prediction of Mishima & Ishii (1984)

Liquid mass flux ($kg/m^2 \cdot s$)	Gas mass flux ($kg/m^2 \cdot s$) at transition obtained from: Experiment	Theory	ε_b at transition
5.3	7.2	6.0	0.778
10.5	8.0	6.5	0.777
49.5	8.9	7.5	0.776
111.8	10.4	9.5	0.775
199	7.5	12.0	0.773
297	11.6	15.0	0.772
399	12.2	18.0	0.771

Table 5. Comparison of the gas mass flux ($kg/m^2 \cdot s$) at the transition to churn flow with the prediction of Brauner & Barnea (1986)

Liquid mass flux ($kg/m^2 \cdot s$)	Gas mass flux ($kg/m^2 \cdot s$) at transition obtained from: Experiment	Theory
5.3	7.2	13.0
10.5	8.0	12.9
49.5	8.9	12.8
111.8	10.4	12.6
199	7.5	12.4
297	11.6	12.1
399	12.2	11.8

flow rates in Owen's (1986) experiments are compared with experimental results in table 5. It is seen that the transition gas velocity is overpredicted at low liquid flow rates, but that it is in line with the data at high liquid flow rates. Also, the predicted value decreases slightly with increasing liquid flow rate whereas Owen's data show a more pronounced and opposite effect of the liquid flow rate. Thus, the predictions of this model are consistent with experimental data at high liquid flow rates, but not at low liquid flow rates.

A possible reason for this lack of agreement at low liquid flow rates may be that the model to predict the slug void fraction (ε_s) is inaccurate for these conditions. The model of Barnea & Brauner (1985) is based on the assumption that the gas in the slug behaves as dispersed bubble flow. It does not consider the process of bubble entrainment at the front of the slug, and the subsequent dispersion of the bubbles through the body of the slug. It is possible therefore that the simplified model for slug void fraction may be in error.

4. MODELS REVISITED

4.1. *Results of the comparison with Owen's (1986) data*

From the above comparison, it appears that the entrance effect mechanism of Dukler & Taitel (1986) can be discounted as a satisfactory explanation of the slug-to-churn transition. The model of Mishima & Ishii (1984), while giving good results for low and medium flow rates, is found to be inconsistent in its detail and its good agreement with data is probably fortuitous. Both the flooding mechanism model of McQuillan & Whalley (1985) and the excessive bubble entrainment model of Brauner & Barnea (1986) are based on phenomena normally associated with slug flow, and show good agreement with experimental results albeit in a limited range, namely, at low and high liquid flow rates, respectively. In addition, they do not reflect correctly the trend of the critical gas velocity with increasing liquid flow rate. Both models show that the gas flow rate for the transition decreases, whereas experimental data suggest the opposite trend. There is thus clearly a need for improving both models. However, the scope for improving the prediction of the slug void fraction is limited because the underlying processes are poorly understood. We therefore restrict our attention to the flooding mechanism, and investigate the effect of accounting for two of the major shortcomings of the model of McQuillan & Whalley (1985).

4.2. *Improvements to the modelling of the flooding mechanism*

There are two major shortcomings in the McQuillan & Whalley (1985) model of this mechanism as it might happen in the Taylor bubble in vertical slug flow. The first of these is that the film surrounding the Taylor bubble is assumed to be laminar, and the Nusselt relation [7] between the film thickness and film velocity is used in the analysis. It is proposed that this be replaced by the empirical correlation of Brotz (1954), which has been found to be applicable over a wide range of film Reynolds numbers (Fulford 1964):

$$\delta\left[\frac{g\Delta\rho}{\nu_L^2\rho_L}\right]^{1/3} = 0.1719\mathrm{Re}_f^{2/3}; \quad \text{where } \mathrm{Re}_f = \frac{\Gamma}{\mu_L}. \tag{14}$$

Here Γ is the film flow rate (by mass) per unit wetted perimeter, and ν_L is the kinematic viscosity of the liquid. This relation has been derived by Brotz (1954) from experimental results in the range of $100 < \mathrm{Re}_f < 4300$, but comparison with data shows that it is valid over a much wider range of Re_f (Fulford 1964). This relation can be rewritten in terms of the superficial film flow rate as follows:

$$U_{fs} = 9.916(1-\varepsilon_b)\sqrt{\frac{gD\Delta\rho(1-\sqrt{\varepsilon_b})}{\rho_L}}; \quad \text{where } \varepsilon_b = 1 - 4\frac{\delta}{D}. \tag{15}$$

It is proposed that [15] be used in place of [7] to better represent the flow in thin, turbulent films.

The second major shortcoming with the McQuillan–Whalley model of flooding is the neglect of the effect of the length of the falling film on the flooding velocity. This is illustrated in figure 5 [adapted from Hewitt & Hall-Taylor (1970)], which shows the variation of the flooding gas flow rate as a function of the liquid flow rate for a range of countercurrent flow columns, all of which have identical inlet and outlet configurations. The flooding velocity increases rapidly as the column

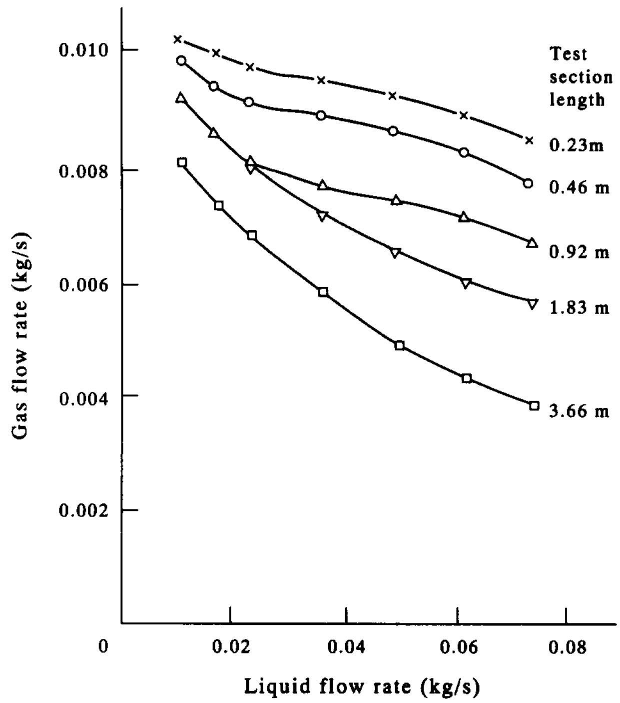

Figure 5. Effect of test section length on flooding velocity in air–water countercurrent flow [adapted from Hewitt & Hall-Taylor (1970)].

length, and hence, the falling film length, decreases. This may explain some of the inconsistency in the predictions of McQuillan & Whalley's (1985) model in the following way.

Flooding correlations of the form of [2] with a constant C of around unity are strictly valid for long falling films. In slug flow, this condition may be satisfied at low liquid flow rates, but not necessarily at high liquid flow rates. This is because the Taylor bubble length decreases as the liquid flow rate increases. This is demonstrated in table 6 which shows L_b close to the critical gas velocity for the cases listed in table 3. Here L_b is calculated from the following liquid phase mass balance equation over the slug unit:

$$U_{Ls} = U_{sL}(1-\varepsilon_s)(1-\beta) - U_{fs}\beta, \qquad [16]$$

where U_{Ls} is the superficial velocity of the liquid through the pipe, U_{sL} is the velocity of the liquid in the slug body and ε_s is the void fraction in the slug body. In these calculations, U_{sL} is taken to be equal to the mixture velocity, $\varepsilon_s = 0.50$ and $L_s = 12D$, all of which are typical values in vertical slug flow. It can be seen immediately that L_b is very high at low liquid flow rates but that it decreases rapidly as the liquid flow rate increases. Thus, it may be expected that the Taylor bubble will experience very different flooding conditions at low and high liquid flow rates, and that the transition velocity may therefore be higher than that predicted by the McQuillan–Whalley model.

Table 6. Length of the Taylor bubble in vertical slug flow close to the transition to churn flow with C = 1.2, $L_s = 12D$ and $\varepsilon_s = 0.50$

Liquid mass flux (kg/m$^2\cdot$s)	Gas mass flux (kg/m$^2\cdot$s)	Length of the Taylor bubble (m)
5.3	5.5	9.4
10.5	5.5	8.3
49.5	5.3	4.1
111.8	5.1	2.2
199	4.8	1.2
297	4.6	0.8
399	4.3	0.5

Unfortunately, there are no standard correlations that take proper account of the falling film length on the flooding velocity (see Bankoff & Lee 1986). We therefore propose an empirical correlation using a limited set of experimental data obtained at Harwell Laboratory (Hewitt *et al.* 1965). These data were obtained in an air–water countercurrent flow column of 0.0318 m i.d. Five different column lengths were investigated with identical inlet and outlet geometries (porous sinters were used for liquid injection and removal): 0.23 m, 0.46 m, 0.91 m, 1.83 m and 3.66 m. In each case, the flooding gas velocity was determined for a set liquid flow rate. The experiments were conducted at three pressures: 1 bar, 1.35 bar and 2.7 bar. The non-dimensional phase velocities (defined in [3]) under flooding conditions are plotted in figure 6(a). It is seen that there is a roughly linear relation between $\sqrt{U^*_{Gs}}$ and $\sqrt{U^*_{Ls}}$ for a given pressure and length of the falling film. There is a slight pressure effect, but the length effect is much more pronounced and is similar to that shown in figure 5. Neglecting the pressure effect, the general form of the correlation is as follows:

$$\sqrt{U^*_{Gs}} + m\sqrt{U^*_{Ls}} = \mathrm{C}, \qquad [17]$$

where the coefficient m is a function of L/D [see figure 6(b)] and is given by

$$m = 0.1928 + 0.01089\left(\frac{L}{D}\right) - 3.754 \times 10^{-5}\left(\frac{L}{D}\right)^2 \quad \text{if } \frac{L}{D} \leqslant 120, \qquad [18a]$$

$$m = 0.96 \approx 1 \qquad \text{if } \frac{L}{D} > 120. \qquad [18b]$$

Equations [17] and [18a,b] constitute a new design equation for vertical flooding columns, and take account of the length effect. It is suggested, based on the data shown in figure 6, that C = 0.88. The general validity of this equation for columns outside the range of the present study is yet to be established (all variation in L/D is achieved by varying L alone while keeping D constant); however, it is noted that the data on which this correlation is based were obtained in experimental conditions close to those of Owen (1986) described above. It is therefore argued that the above correlation would describe the flooding conditions for Owen's data, and that they could be used to test the validity of the flooding mechanism in the slug-to-churn transition.

The effect of the length of the falling film on the flooding gas velocity can now be taken into account by using [17] and [18a,b] instead of [2]. There remains, however, the value of the constant C in [17]. In flooding columns, this value depends on the end conditions. For example, it is said to be 0.75 for sharp flanges into the tubes and 0.88 for rounded flanges (Hewitt *et al.* 1965) and

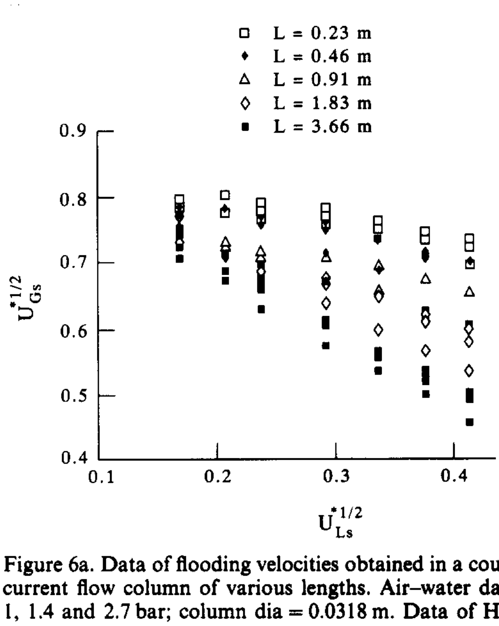

Figure 6a. Data of flooding velocities obtained in a counter-current flow column of various lengths. Air–water data at 1, 1.4 and 2.7 bar; column dia = 0.0318 m. Data of Hewitt *et al.* (1965).

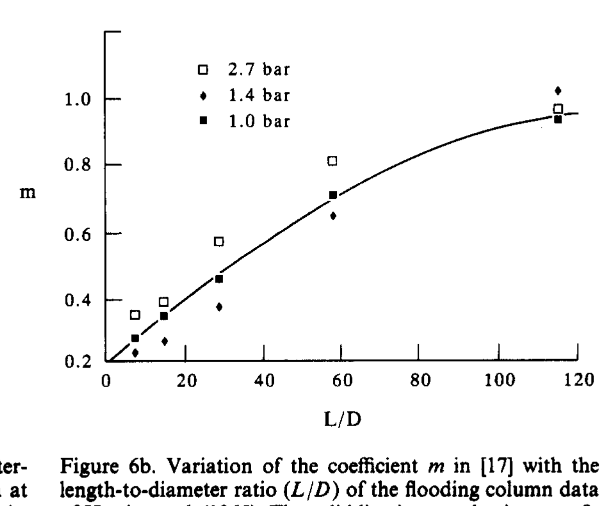

Figure 6b. Variation of the coefficient m in [17] with the length-to-diameter ratio (L/D) of the flooding column data of Hewitt *et al.* (1965). The solid line is a quadratic curve-fit given by [18a].

Table 7. Comparison of the gas mass flux ($kg/m^2 \cdot s$) at the transition to churn flow with the predictions of the old flooding model [2] and the modified flooding model [17]

Liquid mass flux ($kg/m^2 \cdot s$)	Gas mass flux ($kg/m^2 \cdot s$) at transition obtained from		
	Experiment	Old model	New model
5.3	7.2	5.8	7.4
10.5	8.0	5.7	7.5
49.5	8.9	5.6	8.1
111.8	10.4	5.5	8.9
199	7.5	5.2	9.8
297	11.6	4.9	10.6
399	12.2	4.6	11.3

a value of unity fits the data for smooth inlet and outlet conditions given by Hewitt & Wallis (1963). There is thus a degree of variability in this constant. In view of this, it is proposed that its value should be fixed such that the slug-to-churn transition velocity is correctly predicted at very low liquid flow rates (case 1 in table 1) where the length effect would not be present. This gives $C \approx 1$, which is reassuringly close to that for the data of Hewitt *et al.* (1965).

The following equation is therefore proposed to predict the slug-to-churn transition using the flooding phenomenon as the underlying mechanism:

$$\sqrt{U^*_{Gs}} + m\sqrt{U^*_{Ls}} > 1. \qquad [19]$$

Here U_{bs} and U_{fs} are determined by solving simultaneously [5], [6] and [15]; m is obtained from [18] by setting $L = L_b$, which itself is calculated using [16]. In the following calculations, ϵ_s is taken to be 0.5 [Brauner & Barnea (1986), among others, suggest that the void fraction in the slug body increases rapidly as the slug-to-churn transition is approached], and L_s is set to be $12D$ [Moissis & Griffith (1962) suggest that L_s is between $8D$ and $16D$].

The transition gas velocities predicted using this method are compared in table 7 with experimental data as well as with the values predicted earlier without the corrections for the length effect. It is clearly seen that the new method gives vastly improved predictions, and reflects the experimental trend very well. It is therefore concluded that the flooding mechanism is capable of predicting the slug-to-churn flow transition over the full range of liquid flow rates.

5. CONCLUSIONS

It is concluded that of the four models investigated in this study, only the flooding mechanism is capable of predicting the gas velocity required for the transition from slug to churn flow in vertical tubes. While this in itself does not demonstrate that the transition is caused by flooding of the liquid film in the Taylor bubble, it now appears to be the most likely cause of the transition. More direct evidence is required to establish that this is so.

It has been shown that it is essential to consider the effect of the length of the Taylor bubble on the flooding velocity in order to predict the correct trend in the experimental data of transition to churn flow. A model incorporating this has been proposed, and is shown to give satisfactory results.

Acknowledgements—S. Jayanti would like to acknowledge the financial support received from the Science and Engineering Research Council of the United Kingdom during the course of the work reported in this paper. The authors would also like to thank Dr N. Brauner of Tel-Aviv University and the anonymous referees for useful comments and discussions.

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