ONE DIMENSIONAL DRIFT FLUX MODEL APPLIED TO HORIZONTAL SLUG FLOW

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Abstract. *The slug flow is better described by a succession of a liquid pistons trailed by elongated gas bubbles which are not periodic neither in time nor in space. The prediction of the pressure drop as well as of the other flow properties such as sizes and velocities are still a challenge to the flow models including: the unit cell models, the two-fluid model and the mixture models. The objective of the present work is to adapt one of the variants of the mixture model, the Drift Flux model, to estimate the horizontal slug flow properties. It is well recognized that the Drift Flux model captures the features of the vertical slug flow. But there is no available work extending its application to horizontal slug flow. In fact in the former case the pressure gradient is dominated by the weight of the mixture column while in the last case the frictional forces define the pressure drop. To predict accurately the pressure drop in horizontal flows one needs to estimate the lengths of the liquid piston and of the elongated bubble (or liquid film length) and also the liquid holdup on the bubble zone (or liquid film zone). The present work combines the features of the Drift Flux model with a unit cell model to estimate extra information necessary to solve the Drift Flux model. The numerical output is compared against experimental data taken in a horizontal line for air and water flow mixture in a 26 mm internal diameter pipe with* $900D$ *length.*

Keywords: *Slug Flow, Drift Flux, Liquid Film.*

1. INTRODUCTION

The mixture model is a simplification of the two-fluid model, and presents the advantage of enabling the solution of various problems with relative ease of numerical solution. However, it requires knowledge of the phases velocities in relation to the mixture, expressed thru a kinematic relation (Zuber and Findlay, 1965), which depends on the flow pattern of existing.

The 3D transport equation of the mixture model are reduced to a 1D transport equation taking the average along pipe's cross section. This averaging process condenses all the information along the cross section to a point constraining the mass and momentum balances to the axial direction.

This work employs the Drift Flux model, one of the most popular versions of the mixture model. The transport equations are expressed in terms of the mixture properties and the mass center velocity of mixture. The model is solved in terms of the mixture velocity once the drift velocity is supplied by a constitutive equation.

This work uses the Drift Flux model to get estimates of the pressure gradient in horizontal air-water flows. This flow pattern is better described by a succession of liquid pistons trailed by elongated air bubbles which are not periodic in time or in space, see Fig. 1. There is a successful history of application of the Drift Flux model to get estimates of the pressure gradient in upward vertical gas-liquid mixture flowing in the slug regime (Padki et al., 1991; Hasan and Kabir, 1992; Hibiki and Ishii, 2003; Goda et al., 2003; Lima and Rosa, 2008). The success of this application is in the part that the pressure gradient is dominated by the gravitational forces, i.e.

$$\frac{dP}{dz} \cong -\rho g, \tag{1}$$

where ρ is the mixture density defined in Eq. (11). But it is not known applications of the Drift Flux model to horizontal slug flows. In this scenario the gravitational forces do not acts, the pressure gradient is defined by the friction forces acting on the walls. To accurately represent the friction forces on the Drift Flux model it required to acknowledge the intermittent nature of this flow and considers the friction forces due to the liquid slug as well as the forces due to the liquid film and associated gas bubble, see Fig. 1.

The objective of this work is to add a sub-model to the Drift Flux model capable to properly represent the friction forces per unit volume. The development of the sub-model is based on the unit cell model (Taitel and Barnea, 1990).

2. MODEL

The one dimensional formulation of the mixture model condenses the cross section information into a single point through the averaging process represented as:

$$\langle \Psi \rangle = \frac{1}{A} \int \Psi \, dA, \tag{2}$$

where Ψ is a generic quantity and A is the cross section area. Complementary, the cross section average, weighted by the phase volumetric concentration α_k, is represented as:

$$\langle \Psi_k \rangle^{\alpha} = \frac{(1/A) \int_A \alpha_k \, \Psi_k \, dA}{(1/A) \int_A \alpha_k \, dA} = \frac{\langle \alpha_k \, \Psi_k \rangle}{\langle \alpha_k \rangle}, \tag{3}$$

where the subscribed k indicates the phase.

Since the average processes is implicit through all the developments of the 1D mixture model, its representation will be simplified to $\langle \Psi \rangle = \Psi$ and $\langle \Psi_k \rangle^{\alpha} = \Psi_k^{\alpha}$.

2.1 Conservation equations

The gas phase and the liquid phase are represented by the subscripts G and L, for a two components mixture. The mass conservation of the mixture and of the gas phase, as well as the mixture momentum conservation are defined in Eqs. (4), (5) and (6) (Ishii and Hibiki, 2006).

$$\frac{\partial \rho}{\partial t} + \frac{\partial}{\partial z} (\rho \, U) = 0, \tag{4}$$

where ρ is the mixture density and U is the average mixture velocity, defined in Eq. (14).

$$\frac{\partial}{\partial t} (\alpha \, \rho_G) + \frac{\partial}{\partial z} (\alpha \, \rho_G \, U) = \Gamma_G - \frac{\partial}{\partial z} \left(\alpha \, \frac{\rho_G \, \rho_L}{\rho} \, V_{G,J} \right), \tag{5}$$

where α is a gas phase volumetric concentration (void fraction), Γ_G is the mass flow due to phase change and $V_{G,J}$ is the average drift velocity.

$$\begin{aligned} \frac{\partial}{\partial t} (\rho \, U) + \frac{\partial}{\partial z} \left(\rho \, U^2 \right) \quad &= \quad -\frac{\partial P}{\partial z} - \mathbf{T}_W - \rho \, g \, \sin(\theta) - \frac{\partial}{\partial z} \left(\frac{\alpha}{1-\alpha} \, \frac{\rho_G \, \rho_L}{\rho} \, V_{G,J}^2 \right) \\ &- \quad \frac{\partial}{\partial z} \left[(C_V - 1) \left(\rho \, U^2 + \frac{\alpha}{1-\alpha} \, \frac{\rho_G \, \rho_L}{\rho} \, V_{G,J}^2 \right) \right], \end{aligned} \tag{6}$$

where P is the mixture pressure, $\mathbf{T}_{\mathrm{W}}$ is the wall friction term per unit volume, g gravity acceleration, θ is the pipe inclination and C_V is the covariance coefficient of the mixture velocity. In the absence of forces due to surface tension, the pressures that act on liquid and gas phases are equal, so that $P = P_k$.

3. SOLUTION

In this formulation, the following assumptions are adopted: stationary state and isothermal flow without phase change, uniform density and pressure of each phase throughout the cross section, ideal gas and constant viscosity.

3.1 Steady Drift Flux model

In permanent regime, the Drift Flux model equations are substantially simplified. The mass conservation of the mixture, Eq. (4), reduces to:

$$\frac{d}{dz} (\rho \, U) = 0 \quad \therefore \quad \rho \, U = \mathrm{constant}, \tag{7}$$

i.e. the mixture mass flow remains constant throughout the duct. The gas phase mass conservation, Eq. (5), in the absence of mass transfer, $\Gamma_k = 0$, simplifies:

$$\frac{d}{dz} (\alpha \, \rho_G \, U_G^{\alpha}) = 0 \quad \therefore \quad \alpha \, \rho_G \, U_G^{\alpha} = \mathrm{constant}, \tag{8}$$

so that the gas mass flow also remains constant throughout the duct. Note that this result is consistent, therefore ensuring the mass conservation of the mixture and of one the phases, mass conservation of the other phase is implicitly satisfied.

Finally, the momentum conservation equation of the mixture, Eq. (6), reduces to balance among the pressure gradient and frictional and gravitational forces, and a source term associated with the change in momentum due to the expansion of the mixture and distribution of phases in the cross section:

$$\frac{dP}{dz} = -\mathbf{T}_W - \rho\, g\, \sin(\theta) - \frac{d}{dz}\left[C_V\left(\rho\, U^2 + \frac{\alpha}{1-\alpha}\,\frac{\rho_G\,\rho_L}{\rho}\,V_{G,J}^2\right)\right]. \tag{9}$$

Grouping the terms with gradients in the axial direction, the Eq. (9) takes the form:

$$\frac{d}{dz}\left[P + C_V\left(\rho\, U^2 + \frac{\alpha}{1-\alpha}\,\frac{\rho_G\,\rho_L}{\rho}\,V_{G,J}^2\right)\right] = -\mathbf{T}_W - \rho\, g\, \sin(\theta). \tag{10}$$

The Drift Flux model is represented by Eqs. (7), (8) and (10). The boundary conditions at the pipe outlet are the superficial velocities of gas J_G and liquid J_L, and the mixture pressure P. The values of ρ_L, g and θ are known. The others variables are resolved by auxiliary and constitutive equations presents in the next section. The numerical solution was based on a fourth order Runge-Kutta method (Press et al., 1992). The numerical integration advances of the pipe outlet to the pipe inlet.

3.2 Auxiliary and constitutive equations for slug flow

The average mixture density in the cross section is defined by:

$$\rho = \alpha\,\rho_G + (1-\alpha)\;\rho_L, \tag{11}$$

where $P_G/\rho_G = \mathrm{constant}$ (ideal gas law for isothermal flow).

The void fraction (gas phase volumetric concentration) is determined by the Zuber and Findlay relation (1965):

$$\alpha = \frac{J_G}{C_0\,J + V_\infty}, \tag{12}$$

where C_0 is the distribution parameter and V_∞ is the drift velocity.

The mixture superficial velocity is defined by:

$$J = J_G + J_L, \tag{13}$$

where $J_G = Q_G/A$ and $J_L = Q_L/A$ are the superficial velocities of gas and liquid phases. Q_G and Q_L are the volumetric flow of gas and liquid.

The average mixture velocity (Ishii and Hibiki, 2006) is defined by:

$$U = J + \alpha\,\frac{\rho_L - \rho_G}{\rho}\,V_{G,J}, \tag{14}$$

where $V_{G,J}$ is the average drift velocity defined by:

$$V_{G,J} = (C_0 - 1)\,J + V_\infty. \tag{15}$$

For the definition of constitutive equations, it is necessary to determine the distribution parameter C_0 and the drift velocity V_∞. Moreover, to the closing of the mixture momentum conservation equation, it is also necessary to define the friction term and the covariance coefficient of the mixture velocity C_V. All these parameters are dependent on the flow pattern.

In horizontal flow, the distribution parameter C_0 and the drift velocity V_∞ are defined in Tab. 1. These values were established by studies conducted by Nicklin et al. (1962) and subsequently Bendiksen (1984) in slug flow with a low viscosity fluids.

The covariance coefficient of velocity (Ishii and Hibiki, 2006) is defined by:

Table 1. Values of the C_0 and V_∞ for horizontal slug flow.

Regime	C_0	V_∞
Laminar	1.8 to 2.0	0
Turbulent	1.1 to 1.2	0, if $\mathrm{Fr}_J \geq 3.5$
		$0.54\sqrt{g\,D\,\dfrac{\rho_L-\rho_G}{\rho_L}}$, if $\mathrm{Fr}_J < 3.5$

where $\mathrm{Fr}_J = J/\sqrt{gD}$.

$$C_V = \frac{\rho_G\,\alpha\,C_{V,G} + \rho_L(1-\alpha)C_{V,L}}{\rho}, \tag{16}$$

where $C_{V,G}$ and $C_{V,L}$ are the covariance coefficients of the velocities of gas and liquid, respectively. For slug flow, does not have a specific definition for the covariance coefficient. However, for the turbulent regime is expected that $C_{V,G} = C_{V,L} = C_V = 1$, since the velocities profiles and volumetric concentration is relatively flat due to turbulent regime. For the laminar regime, this may not be true, requiring further study on the velocity covariance coefficient.

The wall friction force term per unit volume $\mathbf{T}_W$ correspond to the contribution of the liquid slug $\mathbf{T}_{W,S}$ as well as of the liquid film which, if the gas phase is in contact to the pipe wall, has two components one corresponding to the liquid phase and the other to the gas phase. This term is modeled as:

$$\mathbf{T}_W = \beta(\mathbf{T}_{W,G} + \mathbf{T}_{W,L}) + (1-\beta)\mathbf{T}_{W,S}. \tag{17}$$

The components of friction force per unit volume are written in terms of the shear stresses as:

$$\begin{aligned} \mathbf{T}_{W,G} &= \tau_G\,S_G/A, \\ \mathbf{T}_{W,L} &= \tau_L\,S_L/A, \\ \mathbf{T}_{W,S} &= \tau_S\,S/A, \end{aligned} \tag{18}$$

where S_G, S_L and S are the wetted perimeters of elongated gas bubble, liquid film and and liquid slug. τ_G, τ_L and τ_S are the shear stresses of elongated gas bubble, liquid film and liquid slug, respectively, see Fig. 1. By its turn the shear stresses are evaluated employing Fanning friction factors for smooth pipes:

$$\begin{aligned} \tau_G &= C_{f,G}\,\rho_G\,U_G\,|U_G|/2, \\ \tau_L &= C_{f,L}\,\rho_L\,U_F\,|U_F|/2, \\ \tau_S &= C_{f,S}\,\rho_L\,J\,|J|/2, \end{aligned} \tag{19}$$

where:

$$C_{f,k} = \begin{cases} 16\,\mathrm{Re}_k^{-1} & \text{(laminar)}, \\ 0.046\,\mathrm{Re}_k^{-0.2} & \text{(turbulent)}, \end{cases} \qquad k = G, L, S \tag{20}$$

and Re_k is the phase Reynolds number defined according to:

$$\begin{aligned} \mathrm{Re}_G &= \rho_G\,U_G\,D_G/\mu_G, \\ \mathrm{Re}_L &= \rho_L\,U_F\,D_L/\mu_L, \\ \mathrm{Re}_S &= \rho_L\,J\,D/\mu_L, \end{aligned} \tag{21}$$

where μ_k is the phase viscosity.

The phase hydraulic diameter D_k, within the elongated gas bubble region, is defined according to:

$$\begin{aligned} D_G &= 4\,(1-R_F)\,A/(S_G+S_I), \\ D_L &= 4\,R_F\,A/S_L, \end{aligned} \tag{22}$$

where R_F is the liquid holdup in the film zone provided by a unit cell model.

In order to estimate the friction force per unit volume, Eqs. (17) thru (22), are required the estimates of the intermittence factor β, the wetted perimeters and the gas and liquid velocities at the film zone and at the liquid slug zone. These informations are provided by employing a sub-model, which is described on the next section.

3.3 Sub-model

The function of the sub-model is to provide at the outlet of the domain the required film and slug lengths as well as the phase velocities and the wetted perimeters. It is based on the unit cell model proposed by Taitel and Barnea (1990). Once R_F, L_S and L_F are known at the outlet, the mixture model can propagate them downstream as long as the pipe inclination remains constant. For example, if L_S and L_F are known at the outlet the estimates at any point upstream the pipe are determined:

$$
\begin{aligned}
R_F &= R_{F,outlet} = \text{constant}, \\
L_S &= L_{S,outlet} = \text{constant}, \\
L_F &= L_{F,outlet}\,(P_{outlet}/P), \\
\beta &= L_F/(L_F + L_S).
\end{aligned}
\tag{23}
$$

The following velocities definitions, based on volumetric balances of the gas and liquid phases, are useful to the development of the unit cell and are introduced here for convenience.

The gas velocity in the elongated gas bubble zone is defined by (Taitel and Barnea, 1990):

$$
U_G = \frac{J - R_F\,U_F}{1 - R_F}. \tag{24}
$$

The liquid film velocity is defined by (Taitel and Barnea, 1990):

$$
U_F = U_T + (U_L - U_T)\frac{R_S}{R_F}, \tag{25}
$$

where U_L is the liquid velocity in the slug zone, defined by (Taitel and Barnea, 1990):

$$
U_L = \frac{J - (1 - R_S)\,U_B}{R_S}. \tag{26}
$$

The elongated gas bubble propagation velocity U_T, the gas velocity (dispersed gas bubble velocity) in the liquid slug zone U_B and the liquid holdup in the slug zone R_S will be defined in the next section.

3.4 Liquid film model (LFM)

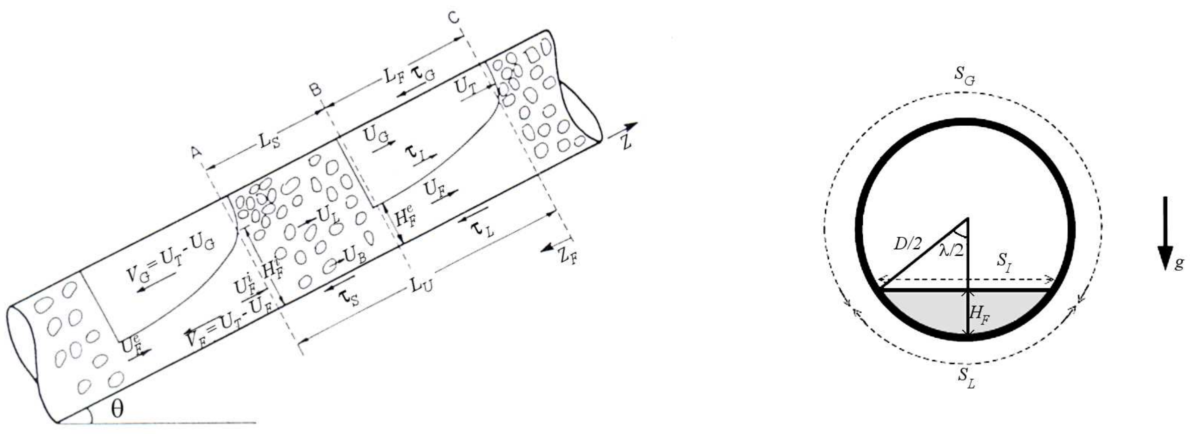
The prediction of the pressure drop as well as of the other flow properties such as sizes and velocities are still a challenge to the flow models including: the unit cell models, the two-fluid model and the mixture models. The unit cell concept (Wallis, 1969) bred a number of models for calculating the slug hydrodynamics parameters. Dukler and Hubbard (1975) developed the first comprehensive model. Others liquid film models also was developed by Nicholson et al. (1978), Kokal and Stanislav (1989) and Taitel and Barnea (1990). The liquid film models derived from the one dimensional steady state momentum equations applied to the gas and liquid phases. The differences among the models arise by neglecting some terms on the momentum balance and also on the closure relations.

Figure 1(a) shows the unit cell schematic geometry for the slug flow. The unit cell is subdivided into the liquid slug zone of length L_S and the film zone of length L_F. Although the liquid slug zone can be aerated by dispersed bubbles, it forms a competent bridging and gas cannot penetrate through the slug zone (Taitel and Barnea, 1990). The liquid hold-up within the liquid slug zone is designated as R_S. In the liquid slug zone, U_L is the average axial liquid velocity and U_B is the average axial velocity of dispersed bubbles. In the film zone, U_T is the nose's propagation velocity of elongated bubble, U_G is the gas velocity in the bubble and U_F is the liquid velocity in the film. V_F and V_G are the relative velocities of liquid film and gas in the elongated bubble. The liquid and gas velocities in the film zone vary along the pipe due the variation of the film thickness H_F.

The film zone consists of a liquid film and an elongated gas bubble. For horizontal and inclined pipes, the bubble is in the upper part of the pipe (Fig. 1(b)). The interface is considered to be plane with internal angle λ which is related with the liquid film height H_F through the relation:

$$
\lambda = 2\ \arccos\left(1 - 2\,\frac{H_F}{D}\right). \tag{27}
$$

Due to the hypothesis of plane interface the liquid film holdup R_F is expressed as:



(a) Unit cell geometry. (b) Film zone cross section.

Figure 1. Schematic diagram for slug flow zones.

$$R_F = \frac{\lambda - \sin(\lambda)}{2\,\pi}. \tag{28}$$

The wetted perimeters of elongated gas bubble, liquid film and interface are defined as function of internal angle λ:

$$\begin{aligned} S_G &= D\,(\pi - \lambda/2)\,, \\ S_L &= D\,\lambda/2, \\ S_I &= D\,\sin(\lambda/2). \end{aligned} \tag{29}$$

The liquid film shape is determined through the film dynamic equation (Taitel and Barnea, 1990):

$$\frac{dH_F}{dz_F} = \frac{\dfrac{\tau_L\,S_L}{A} - \dfrac{\tau_I\,S_I}{A}\left(1 - \dfrac{R_F}{1-R_F}\right) - \dfrac{R_F}{1-R_F}\,\dfrac{\tau_G\,S_G}{A} + \left(1 - \dfrac{\rho_G}{\rho_L}\right) R_F\,\rho_L\,g\,\sin(\theta)}{\left(1 - \dfrac{\rho_G}{\rho_L}\right) R_F\,\rho_L\,g\,\cos(\theta) - \rho_L\,V_F^2\left[1 - \left(\dfrac{R_F}{1-R_F}\,\dfrac{\rho_G}{\rho_L}\,\dfrac{V_G^2}{V_F^2}\right)\right]\,\dfrac{S_I}{A}}, \tag{30}$$

where $V_G = U_T - U_G$ and $V_F = U_T - U_F$.

The shear stresses are expressed in terms of the absolute velocities:

$$\begin{aligned} \tau_G &= C_{f,G}\,\rho_G\,U_G\,|U_G|/2, \\ \tau_L &= C_{f,L}\,\rho_L\,U_F\,|U_F|/2, \\ \tau_I &= C_{f,I}\,\rho_G\,(U_G - U_F)\,|U_G - U_F|/2, \end{aligned} \tag{31}$$

where $C_{f,G}$ and $C_{f,L}$ are the Fanning friction factors for the elongated gas bubble and liquid film, previously defined, and $C_{f,I} = 0.014$ is the interface friction factor. The absolute velocities U_G and U_L also are previously defined.

3.4.1 Liquid film equation: closure relations and integration

The input variables for the liquid film profile integration are J_G, J_L, ρ_G, ρ_L, μ_G, μ_L and D. The elongated gas bubble propagation velocity U_T and the dispersed gas bubble velocity U_B in the liquid slug zone can be determined by the closure equations (Taitel and Barnea, 1990):

$$\begin{aligned} U_T &= C_0\,J + V_\infty, \\ U_B &= J + V_D, \end{aligned} \tag{32}$$

where V_D is the drift velocity of dispersed bubbles in the slug zone (Taitel and Barnea, 1990):

$$V_D = 1.54\,\sin(\theta)\left(\sigma\,g\,D\,\frac{\rho_L - \rho_G}{\rho_L^2}\right)^{1/4}, \tag{33}$$

where σ is the gas-liquid surface tension.

The condition of non-aeration in the slug zone, for horizontal flows, can be considered as a good approximation to determine the liquid holdup in the slug zone, i.e. $R_S = 1$.

For horizontal and near horizontal flows, one starting integrating from the bubble nose at $z_F^i = 0$ with $H_F^i \approxeq D$, since that $dH_F/dz_F < 0$ evaluated at the initial condition is satisfied. This condition neglects the bubble nose region which typically extends 3 to 4 pipe diameter. For long bubbles it is an accurate approximation but for short bubbles it is not satisfactory. The liquid film length is incremented by:

$$z_F^{i+1} = z_F^i + \Delta H_F \left| \frac{dH_F}{dz_F} \right|^{-1}. \tag{34}$$

The geometrical parameters, velocities and shear stresses are calculated at each H_F step. For step sizes smaller than $10^{-2}D$ changes on the liquid film profiles does not observed. This work employs a step size of $10^{-4}D$. The numerical integration is carried out until the mass balance for gas (Eq. (35)) is satisfied and, thus, the film zone length L_F is obtained. similar mass balance for the liquid can also be obtained.

$$J_G = (1 - R_S)U_B + \left[(1 - \overline{R_F}) - (1 - R_S)\right] U_T \frac{L_F}{L_U}, \tag{35}$$

where $\overline{R_F}$ is the average liquid holdup in the film zone:

$$\overline{R_F} = \frac{1}{L_F} \int_0^{z_F} R_F \, dz_F. \tag{36}$$

The unit length L_U is determined by:

$$L_U = \frac{U_T}{f}, \tag{37}$$

where f is the unit frequency obtained from experimental data.

The slug length L_S can be obtained by the the mass balance:

$$L_S = L_F \left\{ \frac{\left[(1 - \overline{R_F}) - (1 - R_S)\right] U_T}{J_G - (1 - R_S)U_B} - 1 \right\}. \tag{38}$$

4. RESULTS

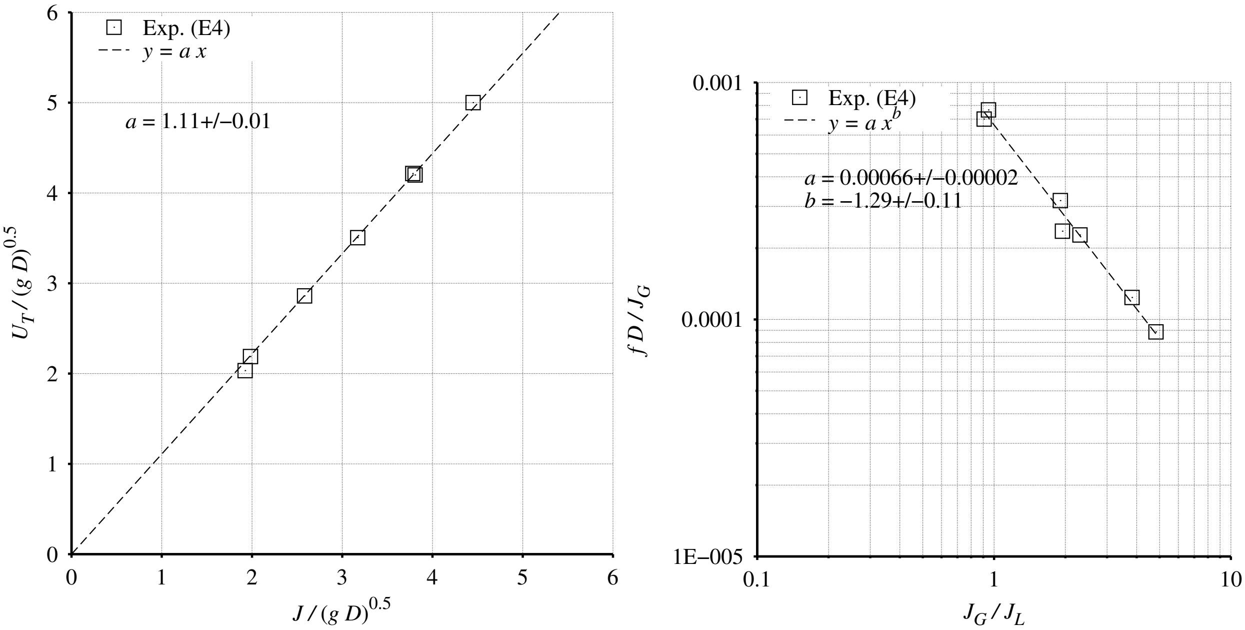
Table 2 shows the test grid and the experimental measurements in a horizontal line of the pressure gradient, bubble velocity and unit frequency.

This experimental results are compared against the numerical results obtained in this work. The circuit is a transparent horizontal acrylic straight pipe with 26 mm of internal diameter D and the total length of approximately $900D$. The measurement stations E1, E2, E3 and E4 are located at $127D$, $265D$, $495D$ and $777D$ of the downstream gas-liquid mixer, respectively. The values presented in Tab. 2 applies only to the measured at station 4 (E4).

Table 2. Experimental tests grid.

Test [#]	Flow Pattern [name]	J_G [m/s]	J_L [m/s]	J [m/s]	∇P [mbar/m]	U_T [m/s]	f [Hz]
1	Slug	0.64	0.33	0.97	1.52	1.03	0.58
2	Slug	1.27	0.33	1.60	2.38	1.77	0.60
3	Slug	1.59	0.33	1.92	2.88	2.12	0.54
4	Slug	0.48	0.53	1.00	2.56	1.11	1.29
5	Slug	0.63	0.67	1.30	4.33	1.44	1.87
6	Slug	1.25	0.66	1.91	5.96	2.13	1.53
7	Slug	1.57	0.68	2.25	6.66	2.53	1.37

The coefficient C_0 and drift velocity V_∞ were obtained from the data fit of elongated gas bubble propagation velocity U_T against the mixture superficial velocity J, as seen in Fig. 2(a). The unit frequency of tests can be observed in Fig. 2(b). From Fig. 2(a), the values obtained were $C_0 = 1.11$ and $V_\infty = 0$, from the set of experimental data. These values are used as input parameters in the model, as well as the unit frequency data for determination of unit length L_U. The unit length together with the mass balance allows the determination of the length of the liquid film L_F and the liquid slug L_S.



(a) Gas bubble propagation velocity vs. mixture superficial velocity.

(b) Unit frequency data of tests.

Figure 2. Experimental data.

Figure 3(a) shows the values of the pressure gradient of the mixture ∇P obtained experimentally (Exp.) against values of the numerical data. For mixture pressure gradient, it is possible to see that there an agreement between numerical and experimental results better 10 %. As expected in horizontal flow, the mixture pressure gradient is resultant of the friction of the fluid with the pipe wall.

The value of pressure gradient mean error was around 7.7 % for the tests. The pressure gradient error is given by the difference between the numerical and experimental pressure gradient, divided by the experimental pressure gradient, in percentage. The mean error is determined by the root mean square of tests errors, i.e. the quadratic mean of tests errors.

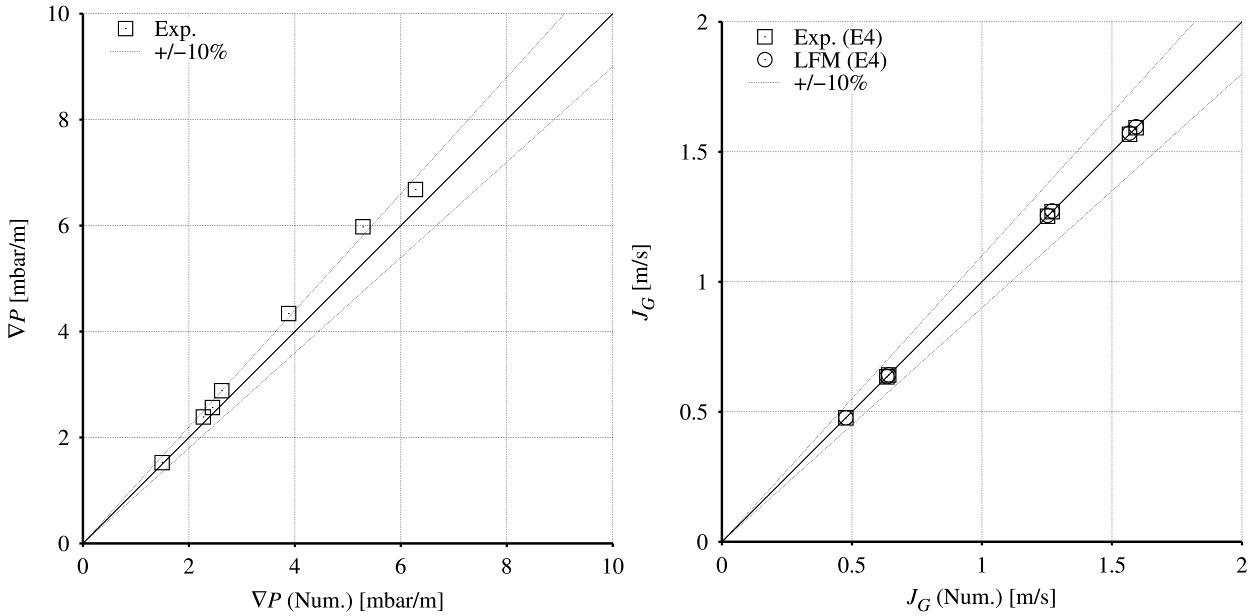
One can observe that the results obtained for the gas superficial velocity using liquid film model satisfies the gas mass balance, as observed in Fig. 3(b). A similar result can be obtained using the liquid mass balance. This is important because allows a correct evaluation of the lengths of the liquid film and L_F liquid slug L_S.

5. CONCLUSIONS

The Drift Flux model combined with a liquid film model presents captures the mixture pressure gradient trend. The differences between the numerical and experimental pressure gradient is associated with several factors, which can include:

- Uncertainties in determining the experimental pressure gradient. The pressure gradient due to friction has order of magnitude smaller than the hydrostatic pressure gradient, therefore, the uncertainties in pressure gradient measurements become far significant in horizontal flow because are dominated by friction forces.
- Uncertainties in determining input parameters in the Drift Flux model. For example, elongated bubble propagation velocity, holdup in the liquid slug zone, unit frequency and superficial velocities, from experimental data or correlations in literature.
- Appropriate relations to the closing integration of the liquid film model and consequently a more precise determination of the lengths of the liquid film and the liquid slug.

The liquid film model provides a better determination of the pressure gradient due to friction in the unit cell composed of liquid slug followed by elongated gas bubble. It is known that in horizontal flow, the frictional forces is crucial for the mixture pressure gradient. Therefore, the combination of the model of liquid film with the Drift Flux model provides a better estimate of the mixture pressure gradient.



(a) Mixture pressure gradient.

(b) Gas superficial velocities.

Figure 3. Comparison between experimental and numerical data.

This combined modeling of Drift Flux model with a liquid film model can be extended to inclined slug flow. As the inclination of the pipe increases the trend is the mixture pressure drop to be less dependent on the frictional forces and more dependent on the mixture weight. The Drift Flux model gives better results when the weight of the mixture becomes more important that the friction.

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