A COMPARITIVE STUDY OF CLOSURE EQUATIONS FOR GAS-LIQUID SLUG FLOW

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Abstract. *The simultaneous flow of gas and liquid in pipelines, over a wide range of flow rates, is characterized by a succession of aerated liquid pistons followed by elongated gas bubbles recognized as slug flow pattern. The main feature of this flow pattern is the intermittency. The gas and liquid structures do not repeat in space or time. Modeling this type of flow has been a challenge during the last four decades. Nowadays there are available steady state one dimensional models based on the unit-cell concept and more accurate physical representations based on two-fluid models or slug tracking models which embody transient flow capabilities. Despite of the modeling efforts the one-dimensional nature of these models still requires closure equations based on experimental data. They are needed to physically represent three dimensional flow features into the one-dimensional model. The objective this the article is to present a state-of-the review of closure correlations concerning: front bubble velocity, slug frequency and liquid slug holdup. An analysis on the accuracy of these correlations is performed based on direct comparison against experimental slug flow data taken by the 2PFG/DE/UNICAMP in horizontal lines.*

Keywords: *slug flow, slug length, slug frequency, bubble velocity.*

1. INTRODUCTION

The flow of gas and liquid in pipelines is characterized by patterns in which the phases become spatially arranged. Over a wide range of gas and liquid flow rates the slug flow pattern occurs. Thus, understanding slug flow is of major importance to the oil and gas as well as to nuclear industries.

The most important characteristic of slug flow is its intermittency. Two flow structures repeat randomly in space and time: separated and disperse. In separated flow, a large bubble bridges almost all pipe cross section while a thin liquid film, free of gas, flow around it. In disperse flow, a great amount of liquid separates two consecutive large bubbles. This portion of liquid is called slug, and may contain some dispersed bubbles. Figure 1 represents a slug flow unit cell containing a large bubble with length L_B and a liquid slug with length L_S .

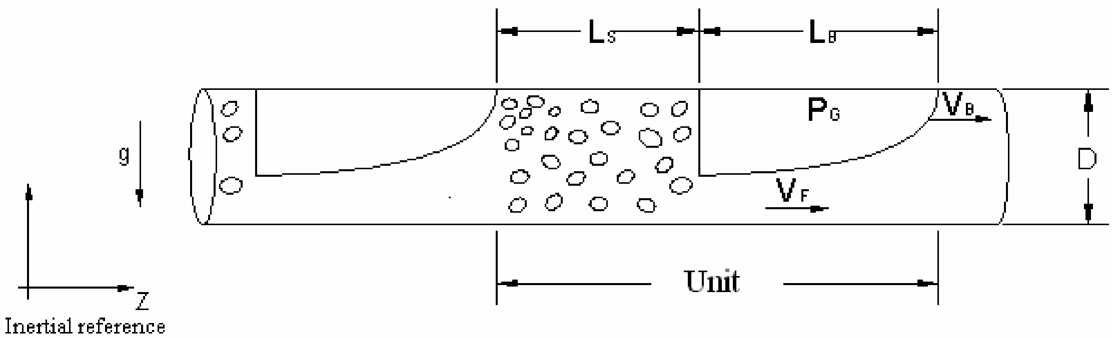

Figure 1. Representation of a horizontal gas-liquid slug flow unit cell (elongated bubble and liquid slug)

Wallis (1969) was one of the firsts to study slug flow. The author developed one model in which one unit cell develops periodically through the pipe. Dukler and Hubbard (1975) and Taitel and Barnea (1990) developed predictive models based on the flow hydrodynamics. However, these models predict slug characteristics for only one unit cell and generalize to the whole pipe. Since they ignore the flow intermittency they are called steady-state models. In addition, the model has more unknowns than equations and some closure relations based on experiments are needed. Consequently the model becomes weak to explore new scenarios. However, steady-state models are still useful in some applications and as a first guess.

As a result of computer development, new methodologies to model slug flow appeared, such as the one-dimensional two-fluid and slug tracking models. These models embody transient capabilities, thus, capturing flow intermittence in both space and time.

In comparison to two-fluid, slug tracking models present low computational cost and no numerical diffusion. Slug tracking models arrive from mass and momentum balances in elongated bubble and liquid slugs regions. Coupling the equations for all cells along the pipe, one obtains an equation system which governs the flow for each time step (Franklin, 2004; Rodrigues et al., 2006). As the control volumes containing elongated bubbles and liquid slugs travel along the pipe, the model is Lagrangean. The major advantage of slug tracking is its potential to predict interaction between bubbles (even coalescence) and progression of pressure drop, bubble velocity, bubble and slug lengths and slug frequency all along the pipe.

In addition to mass and momentum balance equations, the two-fluid and the slug tracking models need some closure equations. These equations must be function of the known parameters such as inlet flow rates, geometric configurations and fluids rheological properties.

The most frequent closure equations are for bubble velocity, slug frequency and liquid slug holdup. The bubble velocity gives information of the front bubble displacement in time. Besides, in slug tracking models, the bubble displacement on each time step is given as an integration of bubble velocity in time (Franklin, 2004; Rodrigues et al., 2006). In steady-state models slug frequency is used to calculate some unit-cell parameters (Dukler and Hubbard, 1975; Taitel and Barnea, 1990), while in slug tracking models it acts as a inlet condition to the simulation. Slug liquid holdup is relevant to the mass balance, as it influences the pressure drop due to gravitational and friction forces, as it changes liquid slug mean density and viscosity.

The evaluations of the closure equations are uncoupled to mass and momentum balances. They arrive from experimental correlations or simple analytical models. The present work presents a review of the closure equations to evaluate elongated bubble velocity, slug frequency and liquid slug holdup. The performance of bubble velocity and slug frequency correlations is analyzed through comparison with experimental data.

2. CLOSURE RELATIONS TO SLUG FLOW MODELS

This section reviews the models to calculate the bubble velocity, the slug frequency and the liquid slug holdup.

2.1. Elongated bubble velocity

In slug flow experiments, bubble velocity may be measured with good precision. As a primary variable, it is useful to determine other parameters in the experiment. In practical applications, bubble velocity is an important measure of the flow rates and time of a slug unit passage along the pipe.

Nicklin et al. (1962) were the first to study elongated bubble motion within a flowing liquid. The authors stated that bubble velocity could be calculated as a linear superposition of the velocity in stagnant liquid and the influence of the moving liquid and proposed:

$$V_B = C_0 j + C_1\sqrt{gD} \tag{1}$$

where V_B is the bubble translational velocity, j is the superficial mixture velocity, C_0 is a constant that measures the influence of mixture velocity in bubble velocity, C_1 is a constant to evaluate the drift velocity, g is the gravity acceleration and D the pipe diameter. Through experiments in vertical turbulent flow Nicklin et al. (1962) established the values of C_1=0.351 and C_0=1.2. The authors also identified that in laminar flow C_0 is a function of the liquid velocity profile ahead of a bubble, which depends of fluid properties and flow configuration. Since this work was accepted that C_0=C_0(Re,Fr,Eo,β) and:

$$\mathrm{Re}_j = \frac{\rho_L D j}{\mu_L} \;; \qquad \mathrm{Fr_j} = \frac{j}{(gD)^5}, \qquad Eo = \frac{\Delta\rho g D^2}{\sigma} \tag{2}$$

where Rej, Frj and Eo represents the mixture Reynolds number, Froude number and Eotvos number, respectively, ρ_L the liquid phase density, $\Delta\rho$ is the difference of liquid and gas densities, μ_L is the liquid phase viscosity, σ is the surface tension and β is the pipe inclination from horizontal.

Through their experiments with one bubble rising in stagnant liquid Davies and Taylor (1950) proposed the value of C_1=0.328. Zukoski (1966) carried out experimental studies to determine viscosity, surface tension and pipe inclination influence on C_1 value. The author presented a curve of C_1 as a function of pipe inclination from the horizontal, β, and the inverse viscosity, defined as $Nf = D^{1,5}\sqrt{\rho_L \Delta\rho g}\,/\mu_L$. Benjamin (1968) developed an analytical approach to horizontal flows and obtained C_1=0.542. However, the author used the inviscid theory neglecting the influence of both viscosity and surface tension.

Weber (1981) analyzed Zukoski (1966) experimental data and claimed that for small diameter pipes surface tension is rather important. Therefore the author proposed the following relation as a function of Eo:

$$C_1 = 0,54 - \frac{1,76}{Eo^{0,56}}\,. \tag{3}$$

There was a discussion regarding the existence of C_1 in horizontal flows. Dukler and Hubbard (1975) proposed C_1=0 in horizontal flows, since there is no buoyancy force acting in the pipe axis direction. However, Zukoski (1966) and

Benjamin (1968) experiments showed that there is a bubble velocity in stagnant liquid in horizontal pipes, also called drift velocity. Later, Bendiksen (1984) and Fabre and Liné (1992) reviewed Benjamin (1968) studies and concluded that the drift velocity exists in horizontal pipes and is also a function of Re, Fr and Eo numbers. Table 1 presents nine bubble velocity correlations, all of them are based on Eq. (1) but they differ on the values for C_0 and C_1.

Table 1. Correlations for calculate the elongated bubble translational velocity, V_B

Author	$V_B = C_0 j + C_1\sqrt{gD}$		Re	Geometry
	C_0	C_1		
Nicklin et al. (1962)	1,2	0,351	> 8000	Vertical
Dukler e Hubbard (1975)	$1,022 + 0,021\ln(\mathrm{Re}_j)$	-	30000 to 400000	Horizontal
Ferré (1979)	1,10 $Fr_j \le 2,26$ 1,30 $2,26 < Fr_j < 8,28$ 1,02 $Fr_j \ge 8,28$	0,44 $Fr_j \le 2,26$ 0 $2,26 \le Fr_j < 8,28$ 3 $Fr_j \ge 8,28$		Horizontal
Bendiksen (1984)	$1,05 + 0,15(\mathrm{sen}\beta)^2$ $Fr_{j_L} < 3,5$ 1,2 $Fr_{j_L} \ge 3,5$	$0,54\cos\beta + 0,35\mathrm{sen}\beta$ $Fr_{j_L} < 3,5$ $0,35\mathrm{sen}\beta$ $Fr_{j_L} \ge 3,5$		Horizontal and Vertical
Dukler et al. (1985)	1,225	-	> 8000	Horizontal and Vertical
Théron (1989)	$1,3 - \frac{0,23}{\Gamma} + 0,13(\mathrm{sen}\beta)^2$	$\left(-0,5 + \frac{0,8}{\Gamma}\right)\cos\beta + 0,35\mathrm{sen}\beta$		Horizontal and Vertical
Manolis (1995)	1,033 $Fr_j < 2,86$ 1,216 $Fr_j \ge 2,86$	0,477 $Fr_j < 2,86$ 0 $Fr_j \ge 2,86$		
Woods e Hanratty (1996)	1,1 $Fr_j < 3,1$ 1,2 $Fr_j \ge 3,1$	0,52 $Fr_j < 3,1$ 0 $Fr_j \ge 3,1$		Horizontal
Petalas e Aziz (1998)	$\frac{1,64 + 0,12\mathrm{sen}\beta}{\mathrm{Re}_j^{0,031}}$	-		Horizontal and Vertical

Where: $\mathrm{Fr}_{j_L} = \frac{j_L}{\sqrt{gD}}$; $\Gamma = 1 + \left(\frac{\mathrm{Fr_j}}{\mathrm{Fr_{Crit}}}\cos\beta\right)$, with $\mathrm{Fr_{crit}} = 3.5$; $\mathrm{Re_j} = \frac{\rho_L jD}{\mu_L}$; β = Pipe inclination from horizontal

2.2. Models for frequency calculation

The unit slug frequency is defined as the reciprocal of the elapsed time to the bubble front travel a unit cell distance or:

$$f = \frac{V_B}{L_B + L_S} \tag{4}$$

There are many models to predict slug frequency as a function of known flow parameters, some based on j_L and j_G while others based on geometric configurations or fluid properties. The great majority of the slug frequency models are based on experimental data (Gregory and Scott, 1969; Greskovich and Shrier, 1972; Heywood and Richardson, 1979; Zabaras, 2000; Shell - appud Zabaras, 2000). Although these models are easy to use (i.e. a simple algebraic equation) their usefulness may be restricted in new scenarios. Sakaguchi et al. (2001) studied the influence of tube diameter and fluid properties in their experiments with vertical flow. The authors proposed a correlation based on regression functions. Some models were presented which are based on the mechanics of slug formation from a horizontal stratified flow (Tronconi, 1990; Hill e Wood, 1990; Hill e Wood, 1994). Thus, the frequency is related to fluid properties. In these models one needs to calculate parameters of the steady-state stratified flow such as the gas velocity and film thickness, thus their implementation becomes more difficult. Table 2 presents the closure equations proposed to calculate slug frequency.

Table 2. Slug frequency Closure equations

Author	f
Gregory e Scott (1969)	$f=0{,}0226\left[\dfrac{j_L}{gD}\left(\dfrac{19{,}75}{j}+j\right)\right]^{1{,}2}$
Heywood e Richardson (1979)	$f=0{,}0434\left[\lambda\left(\dfrac{2{,}02}{d}+Fr_j\right)\right]^{1{,}02}$
Shell correlation	$f\sqrt{\dfrac{D}{g}}=Fr_{Min}+A\left[\left(Fr_{j_L}+Fr_{j_G}\right)^{0{,}1}-1{,}17\left(Fr_{j_L}\right)^{0{,}064}\right]^2$
Zabaras (2000)	$f=\left[0{,}836+2{,}75\mathrm{sen}^{0{,}25}(\beta)\right]0{,}0226\left[\dfrac{j_L}{gD}\left(\dfrac{19{,}75}{j}+j\right)\right]^{1{,}2}$
Sakaguchi et al. (2001)	$f=\left[16100\left(\dfrac{D}{j}\right)\left(\dfrac{j_G}{j}\right)^{1{,}38}\left(\dfrac{j_L}{j}\right)^{-0{,}166}\left(\dfrac{j}{\sqrt{gD}}\right)^{-0{,}317}\left(\dfrac{\rho_L Dj}{\mu_L}\right)^{1{,}61}\left(\dfrac{\rho_L Dj^2}{\sigma}\right)^{-0{,}564}\left(\dfrac{\mu_G}{\mu_L}\right)^{0{,}333}\left(\dfrac{\rho_G}{\rho_L}\right)^{3{,}04}+0{,}087\right]^{-1}$
Tronconi (1990)	$f=0{,}61\dfrac{\rho_G}{\rho_L}\dfrac{u_G}{h_G}$, u_G velocidade e h_G altura do gás do escoamento estratificado
Hill e Wood (1990)	$f=\left(\dfrac{j_L+j_G}{D}\right)\dfrac{0.275}{3600}\cdot 10^{2.68\cdot(1-\alpha_e)}$, α_e calculado a partir de h_G para o escoamento estratificado
Hill e Wood (1994)	$f=\dfrac{f'}{3600}\left(\dfrac{j_L+j_G}{D}\right)\dfrac{1}{(1-0.05j_G)D^{0.3}}$
Where: $Fr_{j_L}=\dfrac{j_L}{\sqrt{gD}}$; $Fr_{j_G}=\dfrac{j_G}{\sqrt{gD}}$; $\lambda=j_L/j$; $N_{Fr_{\min}}=0{,}048\left(Fr_{j_L}\right)^{0{,}81}$; $A=0{,}73\left(Fr_{j_L}\right)^{2{,}34}$ $f'=-24.729+0.00766e^{9.91209H'}+24.721e^{0.20524H'}$, $H'=(1-\alpha_e)\left(1-\dfrac{0.068}{j_L}\right)$	

2.3. Slug liquid holdup

The slug liquid holdup is the ratio between the volume of liquid in the slug and the slug volume. This variable is required to perform the mass balance for liquid and gas flows. Besides, the mean density is important to the gravitational forces evaluation and is highly influenced by the gas fraction within the slug.

The presented relations assume that the gas distribution within the slug is uniform which is not true. In fact, visualizations of horizontal slug flow show that the bubbles concentrate in the upper portions near the pipe wall and mostly in the slug front region due to the elongated bubble turbulence.

Gregory et al. (1978) correlated slug holdup only to the mixture velocity. Later, Malnes (1982) used Gregory et al. (1978) experimental data and obtained a correlation which takes into account fluid properties. Since then, Fershneider (1983), Barnea and Brauner (1985), Andreussi and Bendiksen (1989), Marcano et al. (1998), Gomez et al. (2000), Abdul-Majeed (2000) and Zhang et al. (2003) proposed correlations based on new experimental data. These correlations are shown in Tab. 3.

3. ACCURACY OF THE CLOSURE EQUATIONS

Bubble velocity and slug frequency predictions are compared to experimental data. Slug liquid holdup results are compared between the models since there is no experimental data for this variable.

The experimental data regarding the horizontal gas-liquid slug flow were obtained at the 2PFG/DE/UNICAMP as well as at the NUEX/BR circuits which are described by Rosa (2006). The fluid properties and geometrical configurations are shown in Tab. 4.

The relative error between the experimental and predicted values is evaluated as:

$$\varepsilon(\%)=\frac{\left|x-x_{Exp}\right|}{x_{Exp}}.100 \tag{5}$$

where ε is the relative error, x is the calculated and x_{Exp} is the measured value.

Table 3. Correlations for calculate slug liquid holdup

Author	α_{LS}
Gregory et al. (1978)	$\alpha_{LS} = \dfrac{1}{1+\left(\dfrac{j}{8,66}\right)^{1,39}}$
Malnes (1982)	$\alpha_{LS} = 1-\dfrac{1}{1+\left(\dfrac{83}{Fr_j Bo_L^{0,25}}\right)}$; onde $Bo_L = \dfrac{gD^2\rho_L}{\sigma}$
Fershneider (1983)	$\alpha_{LS} = \dfrac{1}{\left[1+\left(\dfrac{j}{\sqrt{gD(\Delta\rho/\rho_L)}}\right)^2 \Big/ \left(\dfrac{A}{Bo^\beta}\right)^2\right]^2}$
Barnea e Brauner (1985)	$\alpha_{LS} = 1-0,058\left[0,605\left(\dfrac{Bo}{\mathrm{Re}_j}\right)^{0,1} \hat{Fr}_j^{1,2} - 0,725\right]^2$; $\hat{Fr}_j = j\Big/\sqrt{gd\rho_L\Delta\rho}$
Andreussi e Bendiksen (1989)	$\alpha_{LS} = \dfrac{(F0+F1)}{(Fr_j+F1)}, \begin{cases} F0 = \text{máx}\left(0; 2,6\left[1-2\left(\dfrac{d_0}{D}\right)^2\right]\right) & ; \quad d_0 = 2,5\ cm \\ F1 = 2400\left(1-\dfrac{\text{sen}\beta}{3}\right)Bo^{-3/4} \end{cases}$
Marcano et al. (1998)	$\alpha_{LS} = \dfrac{1}{1,001+0,0179\,j+0,0011\,j^2}$
Gomez et al. (2000)	$\alpha_{LS} = \exp\left[-\left(0,45\beta + 2,48\cdot 10^{-6}\,\mathrm{Re}_j\right)\right]$
Abdul Majeed (2000)	$\alpha_{LS} = (1,009 - C\cdot j)A, \begin{cases} C = 0,006+1,3377\,\mu_G/\mu_L \\ A = 1 & \text{se } \beta \le 0 \\ A = 1-\text{sen}\beta & \text{se } \beta > 0 \end{cases}$

Table 4. Experimental configurations

Fluids	*Air and water*	*Air and glycerin solution*	*Natural gás and oil*	*N_2 and Oil A*	*N2 and Oil B*	*N_2 and SAE 20-50 oil*
Number of data points	*72*	*23*	*40*	*34*	*32*	*32*
Pipe diameter	*0,026 m*	*0,026 m*	*0,154 m*	*0,0288 m*	*0,0288 m*	*0,0288 m*
Pipe length	*20,098 m*	*20,098 m*	*175 m*	*26.266 m*	*26.266 m*	*26.266 m*
Liquid phase average viscosity	*1 cP*	*30 cP*	*30 cP*	*392,5 cP*	*268,8 cP*	*293,6 cP*
Liquid phase average density	*1000 kg/m3*	*890 kg/m3*	*900 kg/m3*	*917 kg/m3*	*920 kg/m3*	*880 kg/m3*
Superficial liquid velocity (j_L)	*0,33-1,378 m/s*	*0,326-0,673 m/s*	*0,327-1,318 m/s*	*0,17-0,67 m/s*	*0,17-0,66 m/s*	*0,17-0,685 m/s*
Superficial gás velocity (j_G)	*0,422-1,678 m/s*	*0,415-1,551 m/s*	*0,514-2,99 m/s*	*0,37-1,557 m/s*	*0,663-2,00 m/s*	*0,673-2,017 m/s*

Figure 2 presents the predicted bubble velocity against the experimental data for horizontal flow. The dispersion can be as high as 100%, but the majority of the models have relative errors of ±20%. This is confirmed by the average error and the standard deviation presented in Tab. 5 which varies from 13 to 28% accordingly to the fluid viscosity. It should be noted that the correlation from Dukler and Hubbard (1975) result in greater errors for high viscosities since this model was developed based on turbulent flow analysis. Ferre (1979), Bendiksen (1984) and Woods and Hanratty (1996) correlations follows the same pattern and their errors decrease when viscosity increases, unlike from Manolis (1995) correlation which demonstrate no significant error variation for all viscosities.

The calculated slug frequency variation with respect to j_G is shown in Fig. 3a and 3b for constant j_L of 0.5m/s and 1m/s, respectively. For both values of j_L the frequencies calculated by the correlations present similar characteristics, excluding Gregory and Scott (1969) and Shell (appud Zabaras, 2000). The former is nearly 15% greater for j_L=1m/s than for j_L=0.5m/s while the later predicts lesser values with j_L=0.5m/s and j_G<1m/s. The methods based on slug formation (Tronconi, 1990; Hill and Wood, 1990; Hill and Wood, 1994) show a singular aspect: there is a sudden change in the frequency value for j_G=1.5m/s in both j_L=0.5 and 1m/s. In addition, when j_L=0.5m/s the 3 models are quite alike, whereas when j_L=1m/s they tend to predict different results.

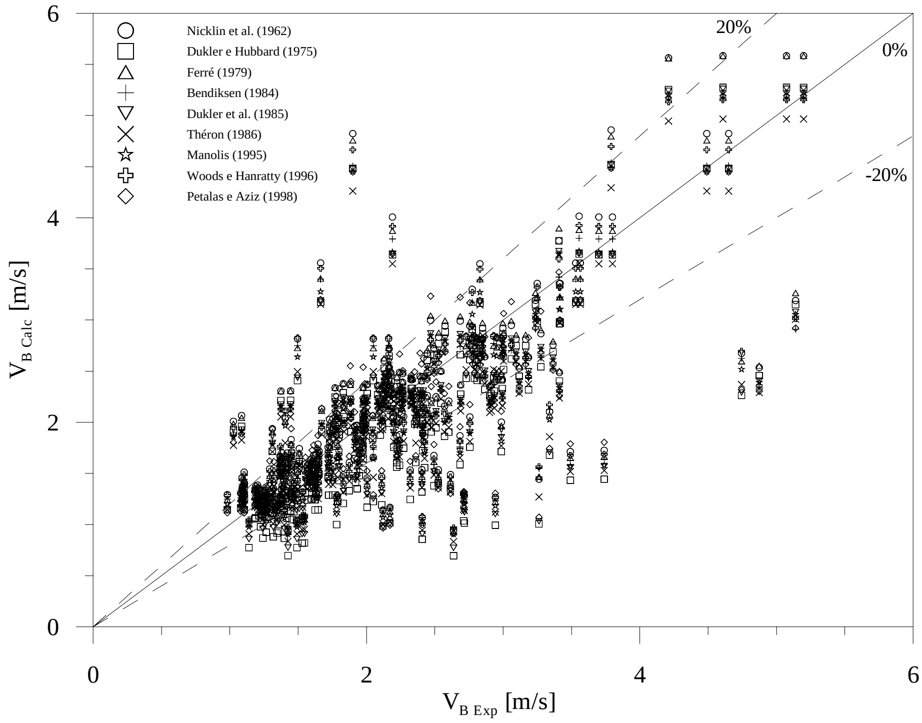

Figure 2. Comparison between measured ($V_{B\ Exp}$) and calculated ($V_{B\ Calc}$) bubble velocity

Table 5. Bubble velocity error and standard deviation

	Average absolute error (%)			Standard deviation (%)		
Liquid viscosity (cP)	**1**	**30**	**> 200**	**1**	**30**	**> 200**
Nicklin et al. (1962)	20.92	19.30	15.31	18.50	28.82	13.66
Dukler e Hubbard (1975)	17.62	21.22	24.49	18.10	23.99	16.07
Ferré (1979)	20.11	18.59	16.22	18.25	27.58	14.04
Bendiksen (1984)	18.14	17.20	16.55	17.41	26.55	13.88
Dukler et al. (1985)	17.14	20.04	19.57	17.86	24.48	15.30
Théron (1986)	16.09	19.17	20.11	17.24	22.55	14.97
Manolis (1995)	17.35	17.79	17.97	17.35	24.89	14.27
Woods e Hanratty (1996)	18.90	17.67	15.47	17.35	28.14	13.65
Petalas e Aziz (1998)	16.26	19.11	15.90	17.72	24.78	13.78

The experimental results of non-dimensional slug frequency ($f.D/j_G$) are presented in Fig. 4 with respect to superficial velocities ratio (j_G/j_L). The data obtained for grater liquid viscosities is separated from the low viscosity ones, which reveals a dependence of slug frequency on viscosity. This aspect was noticed by Rosa (2006). The correlation proposed by Sakaguchi et al. (2001) calculates slug frequency as a function of fluid properties. Since this correlation was obtained in vertical flow, it can not be compared to the experimental data presented in this work.

Table 6 presents the average errors and the standard deviation for calculated slug frequency relative to experimental results. The errors from the models based on experimental data (Gregory and Scott, 1969; Heywood and Richardson, 1979; Shell – appud Zabaras, 2000; Zabaras, 2000) reflect the fact that the experimental fluids were of low and medium viscosities, while the errors for Tronconi (1990) model at high viscosity fluids are huge since this model relies on inviscid non-linear instability theory. Hill and Wood (1990) and Hill and Wood (1994) models present lower errors because their experimental data had larger viscosity variations.

The calculated slug liquid holdup, α_{LS} , is shown in Fig. 5 as a function of the mixture velocity, j, for the air-water condition reported in Tab. 4. There are no experimental data available to check the closure equations accuracy for the liquid slug holdup. All models predict similar characteristics: slug holdup increases while j decreases. This is physically consistent because when the mixture velocity decreases, the turbulence following the elongated bubble, which is responsible for slug aeration, decreases as well. The great differences among the models are the magnitude of α_{LS} . Barnea and Brauner (1985), Fershneider (1983), Gregory et al. (1978) and Malnes (1982) predict a large α_{LS} variation. In fact, these models predict values of α_{LS} lower than 0.7 for high mixture velocities which is physically inconsistent.

The models from Gomez et al. (2000), Abdul-Majeed (2000), Andreussi and Bendiksen (1989) and Marcano et al. (1998) predict a slight decrease on α_{LS} as j increases which is more consistent.

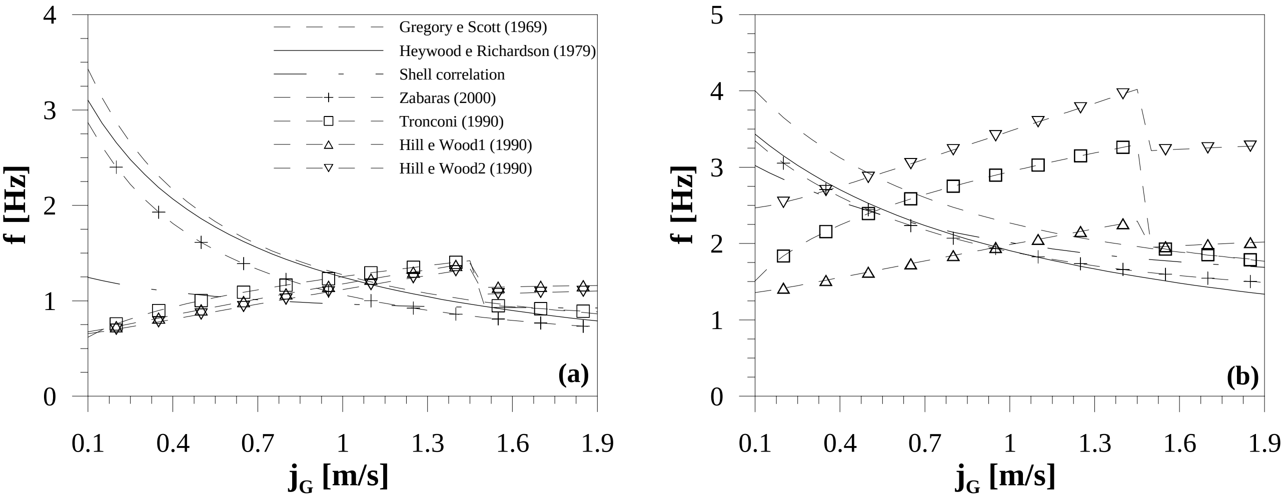

Figure 3. Calculated slug frequency with respect to gas superficial velocity with liquid superficial velocity a) j_L=0.5m/s and b) j_L=1m/s

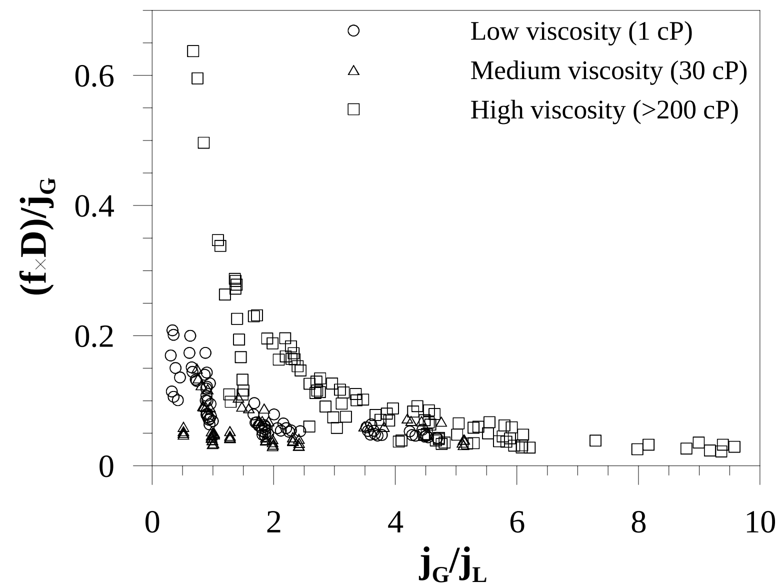

Figure 4. Experimental dimensionless slug frequency with respect to dimensionless gas superficial velocity

Table 6. Slug frequency absolute error and standard deviation

	Average absolute error (%)			**Standard deviation (%)**		
Liquid viscosity (cP):	**~ 1**	**~ 30**	**> 200**	**~ 1**	**~ 30**	**> 200**
Gregory e Scott (1969)	36,88	27,84	74,10	34,26	34,28	34,21
Heywood e Richardson (1979)	41,65	39,70	71,88	34,61	34,61	34,38
Shell correlation	30,14	35,68	79,24	20,60	20,69	21,04
Zabaras (2000)	37,07	29,26	78,35	26,44	26,42	26,47
Tronconi (1990)	26,62	77,68	270,32	19,94	19,95	20,06
Hill e Wood (1990)	37,03	43,98	46,39	24,54	25,07	26,14
Hill e Wood (1994)	28,22	32,88	75,16	17,67	17,51	17,44

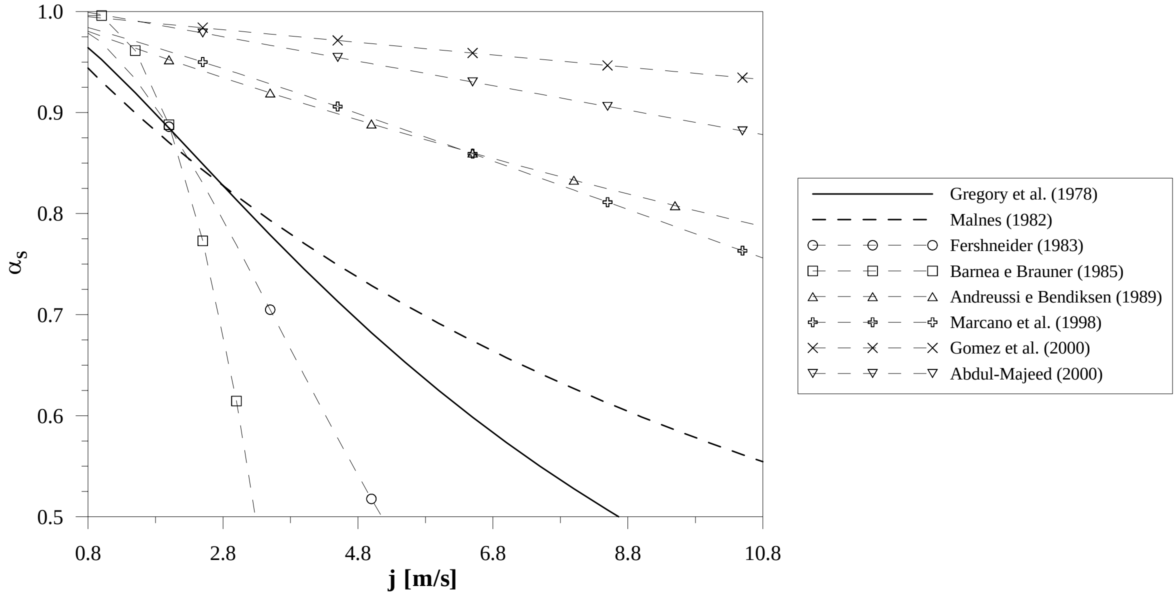

Figure 5. Calculated slug liquid holdup with respect to mixture velocity

4. CONCLUSIONS

A review of the existent models used for estimate elongated bubble velocity, slug frequency and liquid slug holdup in two-phase gas-liquid slug flow was presented. The results of elongated bubble velocity and slug frequency models were compared with experimental data.

All the elongated bubble velocity correlations are based on the bubble drift velocity and mixture velocity influences, thus, predicting the values of C_0 and C_1. Few correlations estimate the influence of fluid properties and pipe inclination and no correlation was proposed concerning the bubble velocity within a train of bubbles. The relative errors obtained by elongated bubble calculations varied between 15-20%.

The slug frequency experimental data showed a significant dependence of slug frequency on the liquid viscosity, although only a few models recognize it. Because of that slug frequency errors were small (30%) for low viscosity fluids and increased considerably (to approximately 70%) to fluids with high viscosity.

The liquid slug holdup was not experimentally measured and the models predictions are similar (α_{LS} decreases while j increases). In the comparison between the models, some presented inconsistent result such as extremely low liquid slug holdups.

Slug flow parameters demonstrate a large randomness due to its intermittent nature and no model is capable to predict these parameters with good accuracy.

5. ACKNOWLEDGEMENTS

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