ISSN 2176-5480

22nd International Congress of Mechanical Engineering (COBEM 2013)
November 3-7, 2013, Ribeirão Preto, SP, Brazil

EVALUATION OF MECHANISTIC MODELLING OF TWO-PHASE GAS-LIQUID FLOW IN PRODUCTION OIL WELLS

André Gonçalves Medeiros
Artur Posenato Garcia
E&P Science and Technology School, PETROBRAS University
andremedeiros@petrobras.com.br, posenatogarcia@petrobras.com.br

Abstract. *Two-phase flow of oil and gas is frequently encountered in production oil wells, not only in vertical and horizontal pipe segments but also in inclined pipes with several different angles. Besides that, sometimes gas is injected in the tubing in order to increase oil production rate. Therefore, it is of utmost importance to determine two-phase flow behaviour in pipes with every conceivable inclination and combination of mass fraction of each phase. Historically, empirical correlations were developed to predict flow conditions in production oil wells, but the reliability of such approach is limited when extrapolated to conditions with flow parameters outside the set used to adjust the correlations. This limitation occurs due to the suppression of some physical mechanisms involved in two-phase flow while developing the correlations. In order to allow a more reliable extrapolation, some models that take into account the two-phase flow mechanisms were published. Several mechanistic models can be found in technical literature, but two of them were chosen for their applicability in a large range of inclination angles. One of them was proposed by Zhang* et al. *(2003) and the other was completely presented by Shoham (2006), despite being a compilation of several previous publications.*
Hence, this work's goal is to evaluate comparatively these two mechanistic models with respect to their considerations, difficulties found during implementation and convergence, discontinuities and computational costs. Conjointly, the flow predictions are compared with some traditional correlations (Beggs & Brill (Beggs and Brill, 1973), Hagedorn & Brown (Shoham, 2006)), commercial softwares (OLGA ©) and field data. The model has to be precise in liquid holdup and pressure gradient calculations for every flow pattern (bubble, annular, stratified and slug) in order to be considered satisfactory. From this consideration, for each flow patter, different pipe inclinations are selected for comparison purposes of the predicted values obtained from mechanistic flow models and what is considered reference for each situation. Through the obtained results, it is possible to determine which model is more precise in the scenarios evaluated and, eventually, combine the best functions in a new model.
As both the pressure gradient and holdup are dependent upon the flow pattern, it is also necessary to evaluate the models capacity of precisely predict the flow pattern, in other words, determine the flow pattern maps. Thus, the maps are also compared in this work. Finally, the mechanistic models need closure relations to correctly represent the two-phase flow phenomena. Consequently, this work proposes some modifications in the closure relations to improve the obtained results with the references. As an example, the proposition of a distribution coefficient C_0 can be mentioned and also some interfacial friction factor correlations.

Keywords: *Two-Phase Flow; Mechanistic Models; Empirical Correlations; Field Data Validation*

1. INTRODUCTION

In order to design and operate oil production systems in an optimized manner, it is necessary to accurately predict the behaviour of two-phase flow of oil and gas in pipes with several different angles. A unified model which could be applied to every conceivable practical condition (inclination angles, pipe diameters, pressure and flow rates) would be desired and a few of such models are available in the technical literature.

One of the early predictive methods commonly used was published by the Beggs & Brill (Beggs and Brill, 1973), which was developed in the 1970's in order to determine the effect of inclination angle on liquid holdup and pressure loss. Empirical correlations for liquid holdup and friction factor were developed for predicting pressure drop at all angles for many flow conditions (Beggs and Brill, 1973) based on experimental data. Despite that, according to Zhang *et al.* (2003) the generalization and accuracy improvement of empirical correlations are limited due to simplifications of the complex physical mechanisms involved in two-phase flow.

Although mechanistic models for gas-liquid two-phase flow have been developed since the mid 1970's (Zhang *et al.*, 2003), there are still several recent developments. The recent approach, based on physical phenomena, can be verified and improved with limited experimental data and, since these models incorporate the physical phenomena and the important flow variables, they can be extended to different operational conditions (Shoham, 2006).

Generally, in a mechanistic model, the flow pattern must be predicted first using simple models in order to determine the specific flow pattern momentum equations and closure relationships which have to be used to predict the multiphase flow behaviour (Zhang *et al.*, 2003). Another possible approach is to determine the flow pattern directly by solving the momentum equation for gas-liquid two-phase flow. The idea is that pattern transitions are hydrodynamic phenomena and

ISSN 2176-5480

should be treated as such. The transitions occur as the flow rates change and the momentum exchange between the phases and the boundaries require a different geometric distribution.

This work will briefly present two mechanistic models and evaluate comparatively with respect to their considerations, difficulties found during implementation and convergence, discontinuities and computational costs. There are several mechanistic models published, but this work will focus on two, which will be called as Mechanistic Modelling of Gas-Liquid and Unified Model via Slug Dynamics.

1.1 Mechanistic Modelling of Gas-Liquid

In multiphase flow the phases might be distributed in several different geometric configurations depending (Stratified, Annular, Bubble, Slug) on the flow conditions, liquid and gas flow rates, pressure, temperature, fluid properties among others. Every flow pattern has its own respective mathematical model to allow calculation of liquid holdup and pressure drop. Therefore, it is important to identify first the flow pattern.

1.1.1 Flow Pattern Identification

The mechanistic model presented by Shoham (2006) for stratified flow (horizontal and near-horizontal flow) uses the Taitel & Dukler separated flow model to calculate the equilibrium holdup. Then, a Kelvin-Helmholtz stability criterion is used to determine the minimum gas velocity for which stratified flow is not stable.

$$v_G \geq \left(1 - \frac{h_L}{D}\right)\left[\frac{(\rho_L - \rho_G)\, g \cos\theta A_G}{\rho_G S_I}\right]^{0.5} \tag{1}$$

The dispersed bubble flow occur when neither agglomeration nor "creaming" of the bubbles occur. According to Shoham (2006), the mechanism of bubble agglomeration is based on turbulence forces balancing surface tension forces. If the first are greater, then bubbles do not agglomerate. The expression proposed by Barnea (Shoham, 2006) is based on Hinze (Shoham, 2006) study of 1955 resulting in the following expression for the maximum diameter of the dispersed bubbles:

$$d_{MAX} = \left[0.725 + 4.15\left(\frac{v_{SG}}{v_m}\right)^{0.5}\right]\left(\frac{\sigma}{\rho_L}\right)^{0.6}\left(\frac{2 f_M v_M^3}{d}\right)^{-0.4} \tag{2}$$

And the critical diameter, above which bubbles start to coalesce (Shoham, 2006):

$$d_{CD} = 2\left[\frac{0.4\sigma}{(\rho_L - \rho_G)\, g}\right]^{0.5} \tag{3}$$

"Creaming" is termed by Shoham (2006) as the case in which the buoyancy forces overcome the turbulence forces and the bubbles concentrate near the (upper) pipe wall. The critical bubble diameter for this transition is:

$$d_{CB} = \frac{3}{8}\frac{\rho_L}{(\rho_L - \rho_G)}\frac{f_M v_M^2}{g\cos\theta} \tag{4}$$

For the dispersed-bubble flow to occur, the maximum bubble diameter d_{MAX} should be less than both d_{CD} and d_{CB} and the void fraction should be less than 0.52.

Annular flow is based on Alves' model (1991) according to Shoham (2006), where the momentum balance is used to determine the liquid film from an equilibrium between an "available" and "required" surface tension.

$$\tau_{Ireq} = g(\rho_L - \rho_G)\, d \sin\theta \left(\tilde{\delta}_L - \tilde{\delta}_L^2\right)^2\left(1 - 2\tilde{\delta}_L\right) + \frac{1}{32} C_L \rho_L \left(\frac{\rho_L d}{\mu_L}\right)^{-n}(v_{SL})^{2-n}\left[\frac{\left(1 - 2\tilde{\delta}_L\right)}{\left(\tilde{\delta}_L - \tilde{\delta}_L^2\right)^2}\right] \tag{5}$$

$$\tau_{Iavail} = \frac{1}{2} f_I \rho_G \frac{v_{SG}^2}{\left(1 - 2\tilde{\delta}_L\right)^4} \tag{6}$$

ISSN 2176-5480

Annular flow occurs if $\tilde{\delta}_{L,OPR} < 0.065$.

For the bubble flow to occur it is necessary the the pipe inclination remains between 60^o and 90^o and that the pipe diameter be:

$$d \geq 19 \left[\frac{(\rho_L - \rho_G)\,\sigma}{\rho_L^2 g} \right]^{0.5} \tag{7}$$

And it is also necessary that the liquid superficial velocity is greater than the following (for a void fraction less than 0.25):

$$v_{SL} = 3.0 v_{SG} - 1.15 \left[\frac{g\,(\rho_L - \rho_G)\,\sigma}{\rho_L^2} \right]^{0.25} \sin\theta \tag{8}$$

Slug flow occur if no other previously mentioned criteria is satisfied.

1.1.2 Pressure Drop Calculation

- **Stratified:**
  Stratified flow is characterized by the complete segregation of the phases, being more likely to occur in downward (inclined) flow. Shoham (2006) recommends the Taitel & Dukler model which consists in obtaining the liquid holdup which makes the pressure drop in the gas phase equals to the pressure drop in the liquid phase. Neglecting the acceleration term, the following equation is obtained:

$$\tau_{WG} \frac{S_G}{A_G} - \tau_{WL} \frac{S_L}{A_L} + \tau_I S_I \left(\frac{1}{A_L} + \frac{1}{A_G} \right) - (\rho_L - \rho_G)\, g \sin\theta = 0 \tag{9}$$

  And the interfacial friction factor is considered equal to the gas friction factor directly taken from the Moody diagram. Cohen and Hanratty (Shoham, 2006) suggest a constant value of 0.0142.

- **Slug:**
  Slug flow is modelled as a slug unit composed by a Taylor bubble and a liquid (gasified) piston unit, according to the Taitel & Barnea model suggested by Shoham (2006). From the mass balance in the slug unit, the following relation is obtained:

$$H_{LSU} = \frac{v_{TB} H_{LLS} + v_{GLS}\,(1 - H_{LLS}) - v_{SG}}{v_{TB}} \tag{10}$$

  The liquid film height in the Taylor bubble region is calculated as it was calculated in the stratified flow, using the following relations for the gas and liquid velocities in this region:

$$v_{TB} - v_{LTB} = \frac{(v_{TB} - v_{LLS})\, H_{LLS}}{H_{LTB}} \tag{11}$$

$$v_M = v_{LTB} H_{LTB} + v_{GTB}\,(1 - H_{LTB}) \tag{12}$$

  A few closure relations are still necessary for the Taitel & Barnea model. The Taylor bubble is calculated from Bendiksen:

$$v_{TB} = C_0 v_m + 0.54\sqrt{gd}\cos\theta + 0.35\sqrt{gd}\sin\theta \tag{13}$$

  The liquid piston holdup is calculated from Gomez:

$$H_{LLS} = e^{-7.85\times10^{-3}\theta + 2.48\times10^{-6} Re_{LS}} \tag{14}$$

ISSN 2176-5480

And the piston length from Scott:

$$\log\left(L_S\right) = -25.4 + 28.5\left[\log\left(d\right)\right]^{0.1} \tag{15}$$

The gas velocity in the liquid piston is calculated with a modified Harmathy correlation:

$$v_{GLS} = cv_M + 1.53\left[\frac{g\sigma\left(\rho_L - \rho_G\right)}{\rho_L^2}\right]^{0.25} H_{LLS}^{0.5}\sin\theta \tag{16}$$

The value of "c" is a little controversial, for the literature suggests something between 1 and 1.5. This is further discussed in the results' section.

- **Annular:**
  The annular modelling is very similar to the stratified model, in the sense that in order to obtain the equilibrium holdup a separated two-phase flow model is used. The difference lies in the geometry, where the gas is considered to flow in the centre (core) of the pipe and the liquid as an annular film in the pipe wall.

  The model suggested by Shoham (2006) is Alves' model, with the following momentum equation:

$$-\tau_{WL}\frac{S_I}{A_F} + \tau_I S_I\left(\frac{1}{A_F} + \frac{1}{A_C}\right) - \left(\rho_L - \rho_G\right)g\sin\theta = 0 \tag{17}$$

  The entrainment liquid fraction f_E in the gas core is calculated with Oliemans' correlation:

$$\frac{f_E}{1 - f_E} = 10^{-2.52}\rho_L^{1.08}\rho_G^{0.18}\mu_L^{0.27}\mu_G^{0.28}\sigma^{-1.8}d^{1.72}v_{SL}^{0.70}v_{SG}^{1.44}g^{0.46} \tag{18}$$

- **Bubbles:**
  The bubble flow has liquid as its continuous phase, and the gas phase is dispersed as bubbles. Using a correlation for the gas velocity (the same used for slug flow) it is possible to determine the liquid holdup for the hydraulic calculations:

$$v_0 = v_G - c_0 v_M \;\; with \;\; v_0 = 1.53\left[\frac{g\left(\rho_L - \rho_G\right)\sigma}{\rho_L^2}\right]^{0.25}\left(1 - \alpha\right)^{0.5} \tag{19}$$

1.2 Unified Model via Slug Dynamics

The unified model proposed by Zhang *et al.* (2000) (REF) differs from previous studies in the way which the slug flow is related to all other flow patterns. According to this model, every flow pattern is represented in slug flow. The film zone of slug flow resembles stratified or annular flow and the slug body resembles bubbly or dispersed bubble flow (Zhang *et al.*, 2000). Besides that, the slug flow lies always in the central part of the flow pattern map, surrounded by the other flow patterns. Therefore, the flow was investigated based on slug dynamics as starting point.

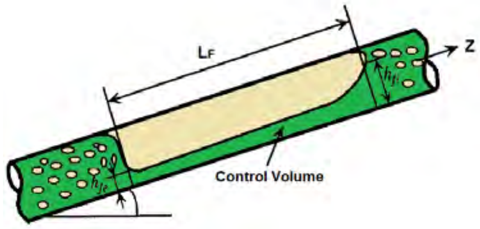

Figure 1. Control volume proposed by Zhang (REF) - modified.

The entire liquid film and the gas pocket in the film zone of a slug unit are used as the control volume, as observed in Fig. 1, with a coordinate system moving with velocity v_t. The resulting volumetric continuity equations for liquid and gas in the entire slug unit are:

ISSN 2176-5480

$$L_U v_{SL} = H_{LS} L_S v_S + H_{LF} L_F v_F \;\; and \;\; L_U v_{SG} = (1 - H_{LS}) L_S v_S + (1 - H_{LF}) L_F v_F \tag{20}$$

And the slug unit length given by

$$L_U = L_S + L_F \tag{21}$$

is not independent from the continuity equations for steady flow, therefore only two of them can be used at the same time (Zhang *et al.*, 2000).

1.2.1 Flow Pattern Transitions

The algorithm was implemented in a way that it sequentially tests the transition boundaries in order to detect the flow pattern for the input given data.

The transition between slug and dispersed bubble flow is calculated, and the transition liquid superficial velocity is determined for each given gas superficial velocity (Zhang *et al.*, 2003). Thus, knowing that $v_m = v_{SG} + v_{SL}$, or that the volumetric mixture velocity is the sum of the gas and liquid superficial velocity, it is clear that $v_{SL} = v_m - v_{SG}$ and as v_{SG} is an input, v_m must be determined. In order to calculate v_m, (Zhang *et al.*, 2003) came up with the following expression for liquid holdup in the slug body:

$$H_{LS} = \frac{1}{1 + \frac{T_{sm}}{3.16[(\rho_L - \rho_g) g\sigma]}} \tag{22}$$

For dispersed bubble flow, the liquid holdup can be assumed to be equal to the no-slip holdup, the ratio between the liquid superficial velocity (that is to be calculated) and the mixture velocity. Since the objective is to calculate v_m it is possible (although not necessary) to make it explicit from the previous equation:

$$T_{sm} = \frac{f_S}{C_e}\frac{1}{2}\rho_S v_m^2, \;\; C_e = \frac{2.5 - |\sin(\theta)|}{2} \Rightarrow v_m = \sqrt{\frac{2C_e\left(\frac{1}{H_{LS}-1}\right) 3.16\sqrt{(\rho_L - \rho g G) g\sigma}}{f_m \rho_S}} \tag{23}$$

Therefore, from that equation, it is possible to calculate v_m iteratively. The initial guess for liquid holdup can be made by use of the Gregory correlation (Zhang *et al.*, 2003) and, consequently the liquid superficial velocity. If the actual liquid velocity is higher than the one calculated from this iterative procedure, then the flow pattern is dispersed bubble. If the liquid velocity is lower, then it is still necessary to determine the boundary between slug and stratified/annular flow.

The transition from slug flow to stratified/annular flow occurs when the liquid film becomes infinitely long. This transition can be roughly approximated by a vertical strip in the flow-pattern map, and this transition is better visualized in terms of the horizontal axis variable, which is the gas superficial velocity. Therefore, for a given liquid superficial velocity, the gas superficial velocity must be determined.

The annular transition can be given by the relation derived by Zhang *et al.* (200)

$$v_A = (1 - H_{LF}) v_C - F_E v_{SL} \tag{24}$$

Where the entrainment factor is calculated by the Oliemans (Zhang *et al.*, 200) correlation. The transition has to satisfy the momentum equation for the annular/stratified flow which is given by:

$$-\frac{\tau_F S_F}{H_{LF} A} + \frac{\tau_C S_C}{(1 - H_{LF}) A} + \tau_I S_I \left(\frac{1}{H_{LF} A} + \frac{1}{(1 - H_{LF}) A}\right) - (\rho_L - \rho_C) g \sin(\theta) = 0 \tag{25}$$

And substituting the shear stresses, it is possible to isolate v_C from this momentum equation:

$$v_c = \sqrt{\left[S_f \frac{f_f}{A_f}\frac{1}{2}\rho_L v_f^2 - S_c \frac{f_c}{A_c}\frac{1}{2}\rho_G v_c^2 + (\rho_L - \rho_C) g \sin(\theta) \frac{2}{S_i f_i \rho_G \left(\frac{1}{A_f} + \frac{1}{A_c}\right)}\right]} + v_f \tag{26}$$

ISSN 2176-5480

But first, it is necessary to calculate the slug translational velocity, which can be expressed as a function of the mixture velocity as proposed by Nicklin (Zhang *et al.*, 200):

$$v_t = Cv_m + \left(0.54\sqrt{gD}\cos\left(\beta\right) + 0.35\sqrt{gD}\sin\left(\beta\right)\right) \tag{27}$$

Then, the liquid holdup of the film can be solved:

$$H_{LF} = \frac{\left(H_{LS}\left(v_t - v_s\right) + v_{SL}\right)\left(v_{SG} + v_{SL}F_E\right) - v_t v_{SL} F_E}{v_t v_{SG}} \tag{28}$$

Since H_{LS} can be calculated with the Gregory correlation (Zhang *et al.*, 200), for the initial guess and for every iteration it must be calculated from the energy relations proposed by Zhang *et al.* (200), without the dispersed bubbles simplifications. This leads to an iterative algorithm to calculate the slug to stratified/annular boundary.

Actually, from this boundary alone it is not possible to determine whether the flow is stratified or annular. Zhang and Sarica (2011) presented a model for the interfacial geometry of the stratified or annular flow. From this model, the transition from stratified to annular flow is smooth since the interface is not flat. The interface bends according to the flow condition and the wetted-wall fraction changes accordingly.

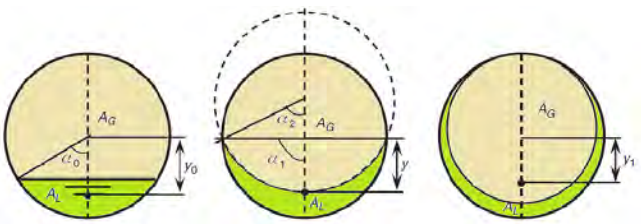

Figure 2. Wetted wall fraction proposed by Zhang (REF) - modified.

Zhang (REF) proposed, according to Fig. 2, the following model to predict the wetted-wall angle (divide by π to obtain the wetted-wall fraction):

$$\alpha_1 = \frac{\pi + \alpha_0}{2} + \frac{\pi - \alpha_0}{3.464}\tan\left[\frac{\pi\left(2y - y_1 - y_0\right)}{3\left(y_1 - y_0\right)}\right] \tag{29}$$

With:

$$\alpha_0 = \pi H_{LF} + \frac{\sin\left(2\alpha_0\right)}{2},\ \frac{y_1}{d} = 0.215H_{LF} - 0.25,\ y_0 = -\frac{d\sin^3\left(\alpha_0\right)}{3\pi H_{LF}}\ and\ y = \frac{y_0}{e^{\left[\ln\left(\frac{y_0}{y_1}\right)\frac{\frac{\rho_g\left(v_g - v_f\right)^2}{\left(\rho_L - \rho_g\right)gd\cos\left(\theta\right)}}{0.7\left(\frac{1-H_{LF}}{H_{LF}}\right)^{0.5}}\right]}} \tag{30}$$

The stratified to annular transition criterion adopted by Zhang *et al.* (2003) was that it can be considered that the flow pattern is annular for wetted wall fractions greater than 90%. The correlation for y_1 does not seem correct for its extreme values and an alternative should be proposed.

1.2.2 Pressure Drop Calculation

- **Stratified/Annular:**
  The stratified model proposed by Zhang *et al.* (2003) is a little different from the one suggested by Shoham (2006), since it considers that the gas-liquid interface is not flat. The friction factor uses Andritsos and Hanratty correlation, nevertheless when the original momentum equation is simplified to the stratified flow it becomes very similar to Taitel & Dukler model. It would also be easy to include an acceleration term.

$$f_{Istrat} = f_C\left(1 + 14.3H_{LF}^{0.5}\left(\frac{v_{SG}}{v_{SG,t} - 1}\right)\right) \tag{31}$$

Where $v_{SG,t}$ is approximated by:

ISSN 2176-5480

$$v_{SG,t} = 5\left(\frac{\rho_{G0}}{\rho_G}\right)^{0.5} \tag{32}$$

Actually, the only difference between the stratified and the annular flow pressure drop calculation in the Zhang's model is the friction factor. The geometric transition, differently from the Alves' model, is smooth but the use of different friction factor correlations might incur discontinuities in the pressure drop.

$$f_{Iann} = f_G\left(1 + 13.8We_G^{0.2}Re_G^{-0.6}\left(h_F^+ - 200\sqrt{\frac{\rho_G}{\rho_L}}\right)\right) \tag{33}$$

- **Slug:**
  The momentum equation for the liquid film for the slug dynamics is:

$$\frac{\rho_L\left(U_T - U_F\right)\left(U_S - U_F\right) - \rho_G\left(U_T - U_G\right)\left(U_S - U_G\right)}{L_F} - \frac{\tau_F S_F}{H_{LF}A} + \frac{\tau_G S_G}{\left(1 - H_{LF}\right)A} + \tau_I S_I\left(\frac{1}{H_{LF}A} + \frac{1}{\left(1 - H_{LF}\right)A}\right) - \left(\rho_L - \rho_G\right)g\sin\theta = 0 \tag{34}$$

Where it is important to note the moment flux that has been considered and differs from the model proposed by Shoham (2006). Despite this equation for the liquid film, when the entire slug unit is considered for the pressure drop calculation, the moment fluxes cancel each other.

$$\left[\frac{\partial P}{\partial z}\right]_U = -\frac{L_F}{L_U}\frac{\tau_F S_F + \tau_G S_G}{A} - \frac{L_S}{L_U}\frac{\tau_S S_S}{A} - \frac{g\sin\theta}{L_U}L_F\left[\rho_L H_{LF} + \rho_G\left(1 - H_{LF}\right)\right] - \frac{g\sin\theta}{L_U}L_S\rho_S \tag{35}$$

Nevertheless, the equilibrium holdup is determined considering the moment fluxes and differ from other models. According to Zhang *et al.* (2000), this makes the model more reliable with experimental results.

- **Bubble:**
  For bubble flow, the bubble rise velocity is considered using the Harmathy correlation in the same way as the model proposed by Shoham (2006). Dispersed flow can be calculated assuming homogeneous no-slip flow.

2. PRESSURE DROP AND HOLDUP VALIDATION

The two mechanistic models presented in this work were implemented in Scilab and the results are compared with Olga © and with classical correlations commonly used in field applications (Beggs & Brill and Hagedorn & Brown). This section presents the obtained results.

2.1 Stratified Flow

Figure 3 presents a comparison between the Taitel & Dukler model, proposed by Shoham (2006), where the interfacial friction factor is equal to the friction factor of the gas with the pipe wall, with the results obtained by using Olga ©. The upper and lower straight lines correspond to a percentile error of 10% and the central straight line correspond to a exact match with the result obtained from Olga ©. The average holdup error is 3.6% and the average pressure drop error is 5.2%.

Figure 4 presents a comparison between the results obtained from Olga © and the results obtained from the Unified Model via Slug Dynamics. The average holdup error is 27.3% and the average pressure drop error is 26.7%.

The results obtained by the mechanistic model proposed by Taitel & Dukler are very correlated to the results from Olga ©, with errors below 10%. The Unified Model via Slug Dynamics resulted in poor correlation with Olga ©, specially in the pressure drop (although better than Beggs & Brill, as can be seen in Fig. 5). The difference between the two mechanistic models lies in the friction factor, because they are separated flow models (for stratified flow). Therefore, it becomes evident the importance of this parameter. Besides that, the interfacial surfaces are different for each model and it can also contribute to this difference.

Figure 5 presents a comparison between the results obtained from Olga © and the results obtained from the Beggs & Brill correlation. The average holdup error is 47.6% and the average pressure drop error is 236%.

ISSN 2176-5480

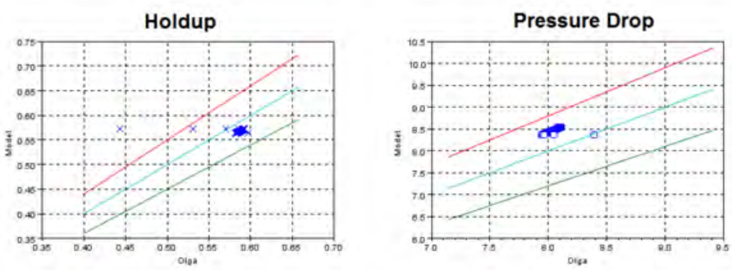

Figure 3. Taitel and Dukler stratified model.

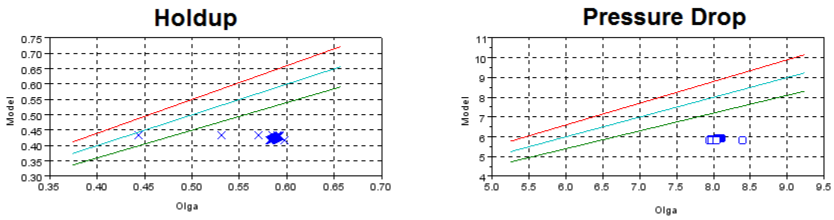

Figure 4. Unified Model via Slug Dynamics - Stratified Flow.

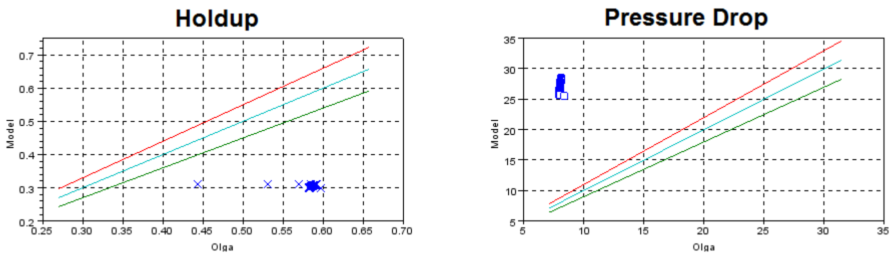

Figure 5. Beggs & Brill - Stratified Flow.

2.2 Slug Flow

The comparisons between the mechanistic model proposed by Taitel & Barnea and OLGA © are presented in Fig. 6 for the horizontal case, for which the C parameter for the Harmathy correlation was considered equal to 1.2. The average holdup error is 8.6% and the average pressure drop error is 5.3%, which has a good adherence to OLGA ©.

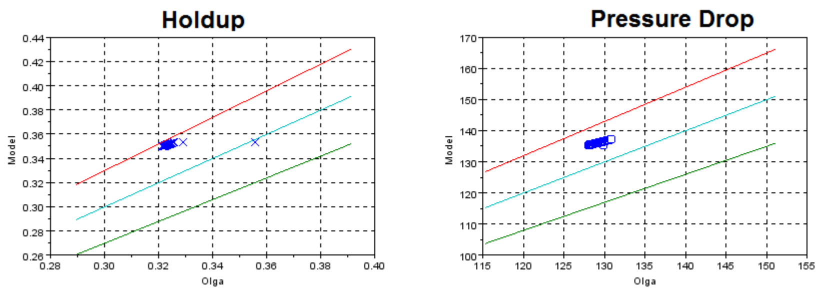

Figure 6. Tatiel & Barnea - Horizontal Slug Flow, $C = 1.2$.

However, the results obtained for the vertical case with $C = 1.2$ were unsatisfactory (the average holdup error is 36.2% and the average pressure drop error is 39.1%). Therefore, an optimization procedure was performed with respect to the holdup, using results obtained from OLGA © and the it was then considered for vertical flow that $C = 1.0$ with the results presented in Fig. 7. The average holdup error is 1.9% and the average pressure drop error is 3.4% for the vertical case, and if $C = 1.0$ would be considered for the horizontal case, the average holdup error would be 1.8% and the average pressure drop error would be 10.6%.

The results obtained from the Unified Model via Slug Dynamics are compared with OLGA © in Fig. 8 for Vertical Slug Flow. The average holdup error is 5.3% and the average pressure drop error is 4.5% for the vertical case and the average holdup error is 19.9% and the average pressure drop error is 36.1% for the horizontal case.

The results obtained from the Hagedorn & Brown correlation are compared with OLGA © in Fig. 9 for Vertical

ISSN 2176-5480

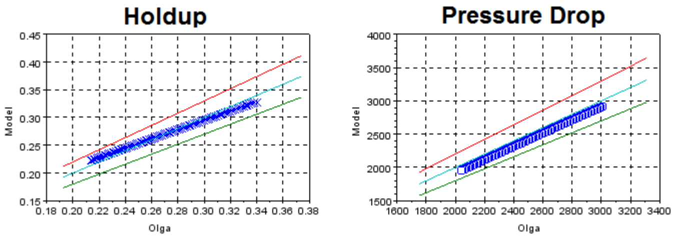

Figure 7. Tatiel & Barnea - Vertical Slug Flow, $C = 1.0$.

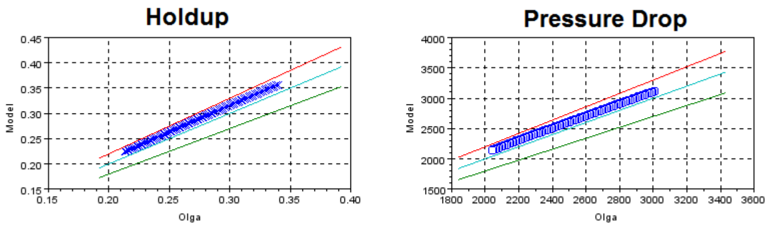

Figure 8. Unified Model via Slug Dynamics - Vertical Slug Flow.

Slug Flow. The average holdup error is 13.8% and the average pressure drop error is 11.7%. The results obtained from the Beggs & Brill correlation are not shown but were calculated and the average holdup error was 27.9% and the average pressure drop error was 29.2%.

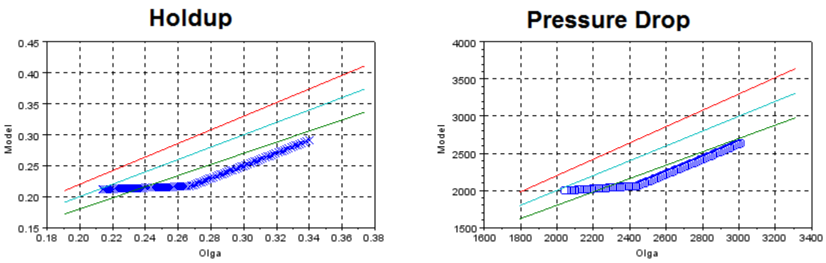

Figure 9. Hagedorn & Brown - Vertical Slug Flow.

From the analysis of the results presented by the Taitel & Barnea model, the great influence of the parameter C (gas velocity in the liquid piston) of the Harmathy correlation to estimate the liquid holdup for the vertical case can be observed, thus the importance of its adequate determination. The Unified Model via Slug Dynamics has shown a good adherence with OLGA © for the vertical case, however it did not present the same adherence for the horizontal flow. Considering only the vertical flow, Beggs & Brill presented again a great deviation from the reference and the Hagedorn & Brown has presented a better performance, even though inferior to the mechanistic model.

2.3 Bubble Flow

The Mechanistic Modelling of Gas-Liquid bubble pattern also uses the Harmathy correlation to calculate the gas velocity. Since $C = 1.0$ was considered to improve the reliability of the slug model, the same value is used for bubble flow. Figure 10 presents the results, where the average holdup error is 2.0% and the average pressure drop error is 1.1%.

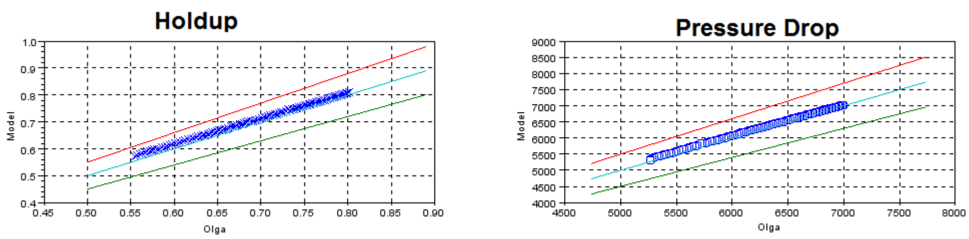

Figure 10. Mechanistic Modelling of Gas-Liquid - Vertical Bubble Flow.

For Beggs & Brill correlation, the average holdup error is 2.8% and the average pressure drop error is 2.5% and for Hagedorn & Brown the average holdup error is 3.4% and the average pressure drop error is 3.8%. Every model analyzed presented results very similar to the results obtained from OLGA © probably because bubble flow is almost homogeneous and similar to one-phase flow.

The Unified Model via Slug Dynamics obtained the best results, seen in Fig. 11, where the average holdup error is 1.1% and the average pressure drop error is 0.9%, and the mechanistic models are still better correlated to OLGA © than the classical correlations.

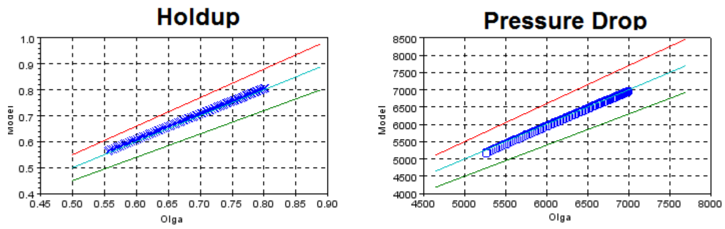

Figure 11. Unified Model via Slug Dynamics - Vertical Bubble Flow.

2.4 Annular

The results for annular flow using Alves' model, suggested by Shoham (2006) are presented in Fig. 12, where the average holdup error is 18.3% and the average pressure drop error is 2.7% with respect to OLGA ©. It must be noted, however, that despite the high relative error, the absolute error is extremely low since the absolute value is also extremely low.

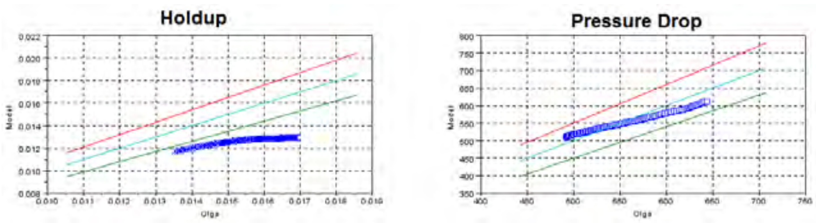

Figure 12. Mechanistic Modelling of Gas-Liquid - Vertical Annular Flow.

The classical correlations were also simulated, but only the values are presented here, not the graphics. For Beggs & Brill correlation, the average holdup error is 54.4% and the average pressure drop error is 205% and for Hagedorn & Brown the average holdup error is 129% and the average pressure drop error is 11.2%.

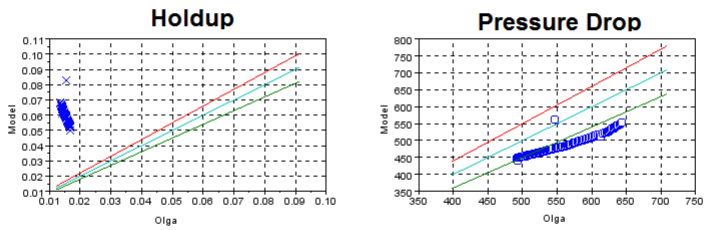

Figure 13. Unified Model via Slug Dynamics - Vertical Annular Flow.

The results for annular flow using the Unified Model via Slug Dynamics (Zhang *et al.*, 2003) are presented in Fig. 13, where the average holdup error is 284% and the average pressure drop error is 12.5% with respect to OLGA ©.

Alves' model holdup forecast was less precise than the adjustments used for other flow patterns, even so the overall result was satisfactory. The pressure drop has shown a good agreement with OLGA ©. The annular flow employs several closure correlations (interfacial friction factor, liquid film friction factor, entrainment fraction) which must be closely investigated to improve the adjustments.

Beggs & Brill has shown large errors both for liquid holdup and pressure gradient. Hagedorn & Brown correlation and Unified Model via Slug Dynamics (Zhang *et al.*, 2003) presented large errors for liquid holdup, but a satisfactory result for pressure drop.

2.5 Concluding Remarks

Mechanistic models were implemented for all flow patterns (stratified, annular, slug and bubbles). In all simulated cases, numerical errors or convergence problems were not observed. In practically every case studied the mechanistic

ISSN 2176-5480

model presented lower errors, comparing with OLGA ©, than the classical correlations Hagedorn & Brown and Beggs & Brill. The models proposed by Shoham (2006) presented in all cases the highest correlation with the reference. Despite exploring more physical aspects of multiphase flow than the correlations, mechanistic models still need several closure empirical correlations. The closure correlations employed in this work seem to be adequate to adjust with OLGA ©.

3. FLOW PATTERN PREDICTION

In the case of flow pattern prediction, the mechanistic models are only compared with each other. For the vertical flow condition ($+90^o$), Fig. 14, both flow regime maps implemented are extremely similar. The greatest difference lies in the bubble region, which occupies a larger area in the model suggested by Shoham (2006). Some differences are also present in the dispersed bubble flow, that in the Unified Model via Slug Dynamics occupies a larger area. For the later model, the criterion for dispersed bubble transition is only valid for superficial liquid velocities greater than 0.1 m/s, thus for velocities below this value, the model suggested by Shoham (2006) is recommended.

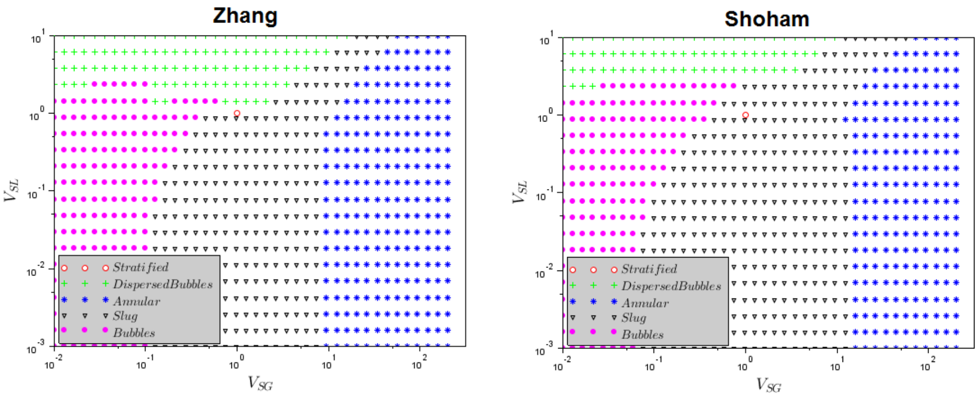

Figure 14. Flow Regime Map for Vertical Flow.

An anomaly can be observed in the Unified Model via Slug Dynamics for pipes with inclination between $+1^o$ to $+10^o$ (Fig. 15). The flow regime map predicts the occurrence of stratified flow in conditions that it should not occur, but should be annular flow. However, it must be noted that in this model the distinction between stratified and annular flow is based on an estimation of wetted perimeter of the pipe and it only indicates annular flow for wetted perimeter larger than 90%. Thus, it is possible that the difference is in flow characterization and not in flow prediction or that the wetted perimeter criterion could be modified.

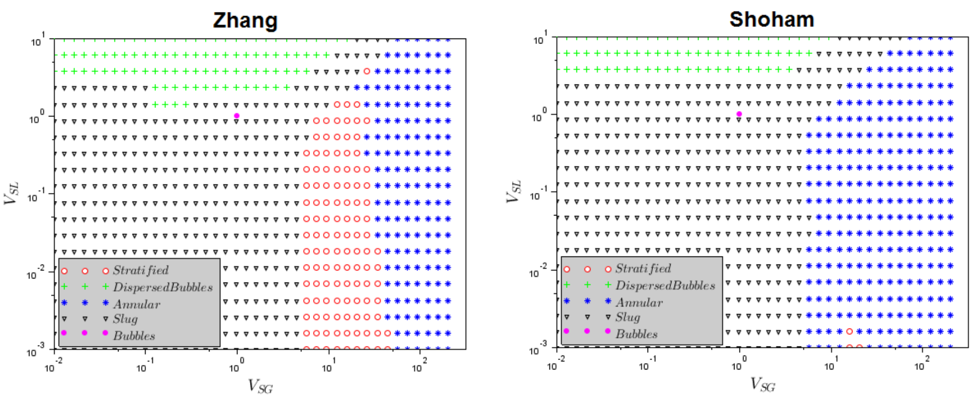

Figure 15. Flow Regime Map for $+10^o$ Inclined Flow.

It is also important to observe that if annular and stratified flow regimes are grouped together, the flow regime maps for both mechanistic models implemented would be extremely correlated for pipes with inclination between $+1^o$ to $+10^o$.

For the horizontal case, Fig. 16, the stratified flow occupies a larger area in the Unified Model via Slug Dynamics flow regime map than in the corresponding map from the model suggested by Shoham (2006). In this case, despite the differences in characterization of stratified and annular flow, even grouping them together, the differences between the two models is noticeable.

The flow regime maps for -5^o inclined pipes, both flow regime maps become extremely correlated if the annular and

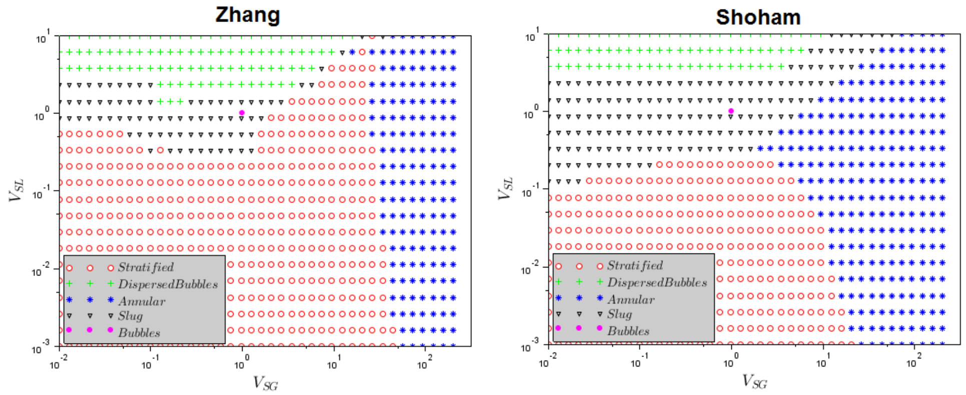

Figure 16. Flow Regime Map for Horizontal Flow.

stratified flow are grouped together. The area occupied by the dispersed bubble flow is larger in the Unified Model via Slug Dynamics flow regime map and the slug flow is smaller.

4. CONCLUSIONS

The Unified Model via Slug Dynamics model (Zhang *et al.*, 2003) is documented only in published papers and the information gathering to perform the simulation is still harder than the model suggested by Shoham (2006). Therefore, it is not possible to ensure that the model implemented is actually the same one developed by Zhang *et al.* (2003), but an effort was employed to make it as close as possible. This model presents some closure equations more recent than the equivalent that are used by the other mechanistic model, therefore when it is compared with OLGA ©, it might be the case that the equations used by the commercial software are not as up to date as Zhang's model. In general, the model suggested by Shoham (2006) is the one which was more closely related with OLGA ©.

The empirical correlations presented larger errors and in general are not recommended, but each application must be individually analyzed. If the production system's characteristics are very similar to the ones used to develop the correlation, probably the empirical model will present good results. It is better to have different models available to evaluate the most precise one for each application. It might also be interesting to eventually compare the different results from different models to make a sensibility analysis.

From this work it is clear that the hydraulic models are extremely important, but equally important is the fluid characterization and also reliable parameters from the production system. The best hydraulic model will be of no help without correct thermodynamic and transport properties. Other effects, such as non-equilibrium effects in steady-state might also be relevant.

5. ACKNOWLEDGEMENTS

The author would like to thank Petrobras for permission to publish this work, Karen Kiyomi Shimabukuro for all her help and support.

6. REFERENCES

Beggs, H.D. and Brill, J.P., 1973. "A study of two-phase flow in inclined pipes". *Journal of Petroleum Technology*.

Shoham, O., 2006. *Mechanistic Modeling of Gas-Liquid Two-Phase Flow in Pipes*. Society of Petroleum Engineers, Richardson - TX, 1st edition.

Zhang, H.Q., Jayawardena, S.S., Redus, C.L. and Brill, J.P., 2000. "Slug dynamics in gas-liquid pipe flow". *Transactions of ASME*, Vol. 122.

Zhang, H.Q. and Sarica, C., 2011. "A model for wetted wall fraction and gravity center of liquid film in gas/liquid pipe flow". *SPE Journal*.

Zhang, H.Q., Wang, Q., Sarica, C. and Brill, J.P., 200. "A unified mechanistic model for slug liquid holdup and transition beteween slug and dispersed bubble flows". *Internation Journal of Multiphase Flow*, Vol. 29, pp. 97–107.

Zhang, H.Q., Wang, Q., Sarica, C. and Brill, J.P., 2003. "Unified model for gas-liquid pipe flow via slug dynamics - part 1: Model development". *Transactions of ASME*, Vol. 125.

7. RESPONSIBILITY NOTICE

The authors are the only responsible for the printed material included in this paper.