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A simple mathematical model for slug liquid holdup in horizontal pipes

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Abstract This paper presents a methodology for calculation of a slug, two-phase flow hold-up in a horizontal pipe. The advantage of this method is that the slug unit hold up can be calculated directly from the solutions of flow field equations with no need to use correlations. An experimental apparatus to measure air–water hold up was setup. The flow pattern and liquid holdup in horizontal and inclined pipes, from angles 5° to 40°, for air–water two-phase flow, are experimentally observed. The test section, with an inside diameter of 30 mm and 3 m in length, was made of plexy-glass to permit visual observations of the flow patterns. The proposed model was tested extensively against experimentally collected data. Furthermore, other data sources for slug flow in horizontal pipes, for air–water and air–oil systems, were also used for comparison. The presented methodology was compared against four recently developed models of a two phase, slug flow holdup in horizontal pipes. Not only does the presented model demonstrate good agreement, with less than 6.8% error, compared to the experimental data, but also has less error compared to other models. These results substantiate the general validity of the model.
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1. Introduction

Slug flow can be observed in many two-phase flow engineering applications, such as flow in oil and gas pipelines and other process industries. The formation of the slug regime is transient in nature, passing from stratified to wavy flow and then onto slug flow. The existence of an interface between the two phases that will change easily is a challenge for modeling [1]. Initial attempts to model slug flow neglect its non-deterministic nature by considering the alternating liquid pistons and gas bubbles in an orderly periodic way.

The flow is reduced to periodic cells moving downstream composed of a liquid piston trailed by a gas bubble, also named a unit cell. Based on this approach, Taitel and Barnea [2] concluded that across the boundary between the liquid slug and the long gas bubble region, flow appears steady, with mass and momentum conserved. The unit cell concept bred a number of models for calculating slug hydrodynamic parameters [3].

The first comprehensive model was developed by Dukler and Hubbard [4]. Further papers employing the unit cell concept improved the steady state slug flow model: Nicholson et al. [5]; Kokal and Stanislav [6] and, later, De Henau and Raithby [7] developed a steady-state two-fluid model for the pressure gradient and liquid holdup.

They tested their model against a limited number of experimental data. Good agreement was reported for air–water and air–oil slug flow in horizontal pipes.

One of the most significant studies in this field was carried out by Taitel and Barnea published in their second paper [8]. They employed a consistent method for calculating the pressure gradient in steady state slug flow in inclined pipes. For calculation of slug holdup, they considered variations of film thickness and, based on this assumption, momentum balance was developed. Then, the length of liquid film holdup, and the velocity at the end of the liquid film were the model outputs. Taitel and Barnea [8] used the holdup to calculate pressure, while assuming that the film thickness was uniform, and

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simplified the momentum balance. The reformulated version of the simple sub model of Taitel and Barnea [8] for calculating pressure drop was presented by Orell [9]. Orell [9] reformulated the Reynolds number for calculating the friction factor, and improved the accuracy of the Taitel and Barnea [8] model for slug pressure drop in horizontal pipes.

However, Orell [9] had to use empirical correlation for slug holdup calculation, and the correlation of Andreussi and Bendiksen [10], as reformulated by Andreussi et al. [11], was used to calculate the slug holdup for both air–water and air–viscous oil systems.

In the present work, the Orell [9] model has been developed for calculation of average liquid holdup in a slug unit. In this study, unlike Orell [9], the liquid film length in the present model is evaluated when deriving the slug liquid holdup through solving the equations. So, the present model enables direct calculation of the slug holdup from the solution of flow field equations. It is independent of any specific experimental correlation and can be used for different fluids (high viscose or low viscose with high or low velocity).

On the other hand, compared to Taitel and Barnea [8], the liquid film thickness in the present model is assumed uniform for slug liquid holdup calculation, so the momentum equation is simplified. Fanning friction factors of the liquid film and gas bubble are calculated by the correlation used by Orell [9] instead of the correlation employed by Taitel and Barnea [8].

Also, liquid holdup in the film zone is defined by one half of the angle that subtends the liquid–gas interface. The present work has a much simpler structure and equations are solved analytically for horizontal pipes. Nowadays, attempts to estimate liquid holdup by slug tracking can be seen [12]. Generally, numerous correlations and models have been developed to predict the hold up of slug flow, some of the newest being: Zhang et al. [13]; Gomez et al. [14]; Issa and Kempf [15]; Abdul-Majeed [16]; Soleimani and Hanratty [17].

2. Experiments

The experimental facility is shown schematically in Figure 1. The main components of the system consist of a test section, air supply, water supply and an instrumentation system. The test section has a fixed and a movable part. Each part, with 1.5 m length and 30 mm diameter, is made of plexy-glass to permit visual observations of the flow pattern. A digital camera is used to record the two-phase flow regimes under different conditions. The experiments are run for different inlet gas flow rates (0–200 l/min), liquid flow rates (150–5500 l/h) and different slopes of the tube, ranging from 0° (horizontal) to 40° (inclined upward).

The water was at room temperature, and the atmospheric pressure was pumped from the storage tank through two rotameters, and then to the air–water mixer and the test section.

Surplus water was sent back to the storage tank. Air was supplied to the test section by a compressor. In a mixer, both the air and water streams were brought together. This part is designed according to previous work [18,19]. The mixer consists of a PVC pipe (6 mm inside diameter) inserted into the liquid stream by means of a compression fitting. The end of the tube was soldered. Four holes (three rows of 1.5 mm, four rows of 3 mm and eight rows of 4 mm) were positioned at 90° intervals around the tube perimeter.

The hole centers are distanced at 1.3 cm. Passing air through the PVC pipe mixes with the water that exists in the plexy-glass pipe. In order to prevent water from entering into the air flow,

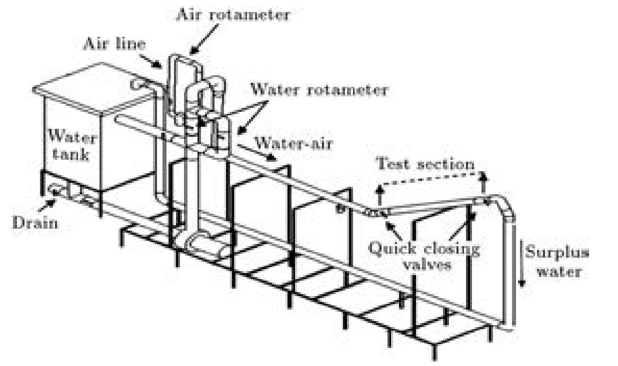

Figure 1: Schematic diagram of the test facility.

the size of the holes diminishes through the internal tube. After mixing gas and liquid flow, the mixed flow passes through the test section.

The air inlet flow rate was measured by a digital flow meter within the range of 0–12 000 l/h. Water flow rate was measured by two rotameters, within ranges of 0–600 and 1000–6000 l/h. During the experiments, the superficial gas velocity was maintained at a set value.

Corresponding to literature, no previous study provides conclusive experimental evidence or estimates of the minimum required length for the flow development section [20]. Also, some of the reported experimental observations are for temporary or developing flow patterns [21]. Based on these lengths, the setup is set at 3 m. This provides around 2 m for the flow to develop. Observations show that flow almost reaches development at this stage and hold up measurements could proceed.

The liquid holdup was measured once the system had reached steady state conditions, having changed the superficial liquid velocity. The liquid holdup was measured by isolating the test section with two rapidly closing valves. Having isolated the test section, the first valve is opened, measuring the amount of water isolated between the two valves; the holdup being calculated through the division of this amount of water by the total volume.

Four differential pressure transducers of range 0–100 psig were installed to measure the two phase pressure drop across the test section. The system was allowed to approach steady state conditions before the air and water flow rates, flow pattern and holdup were recorded. Each data was repeated three times to ensure repeatability and the average was reported. A total number of 720 experiments were carried out to measure the two-phase flow pressure drops, liquid holdup and flow regimes in different gas/liquid flow rates and tube slopes.

2.1. Experimental result

The liquid holdup was measured by two quick closing valves in horizontal and inclined pipes. The experimental results were shown in Figures 2–6 for inclination angles: 0°, 10°, 20°, 30°, 40° (the experimental results for 5°, 15°, 25°, 35° are not displayed as they are very similar to these). In order to ensure result repeatability, each measurement was repeated three times; the average being reported here. In doing so, the resulting outliers have been eliminated. Figures 2–6 show that in horizontal and inclined pipes, liquid holdup tends to go in the same direction as gas and liquid flow rates. In other words, naturally, by increasing the liquid flow rate (at constant gas flow rate), the

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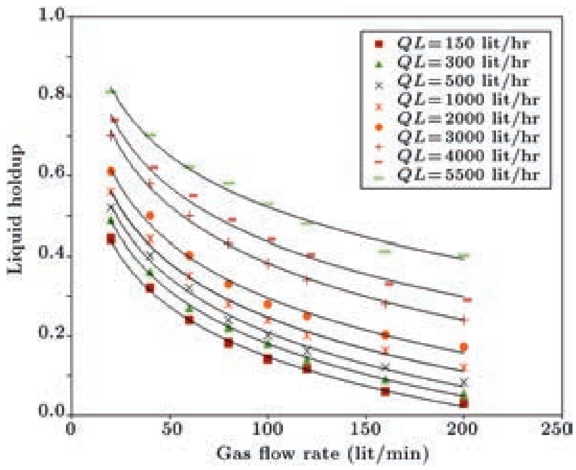

Figure 2: Variation of liquid holdup by gas flow rate in a horizontal pipe with $D = 0.03$ m.

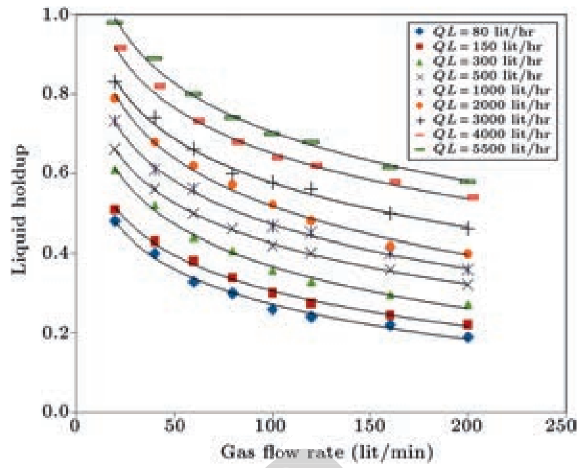

Figure 5: Variation of liquid holdup by gas flow rate in a pipe with $\Theta = 30°$ and $D = 0.03$ m.

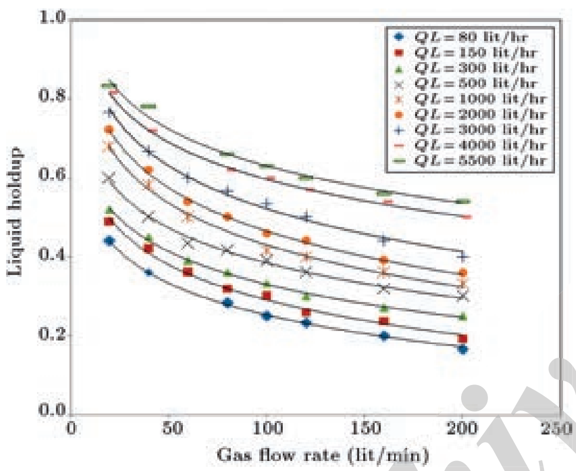

Figure 3: Variation of liquid holdup by gas flow rate in a pipe with $\Theta = 10°$ and $D = 0.03$ m.

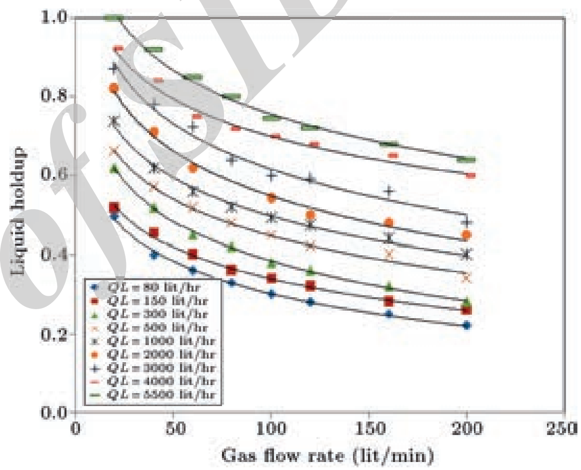

Figure 6: Variation of liquid holdup by gas flow rate in a pipe with $\Theta = 40°$ and $D = 0.03$ m.

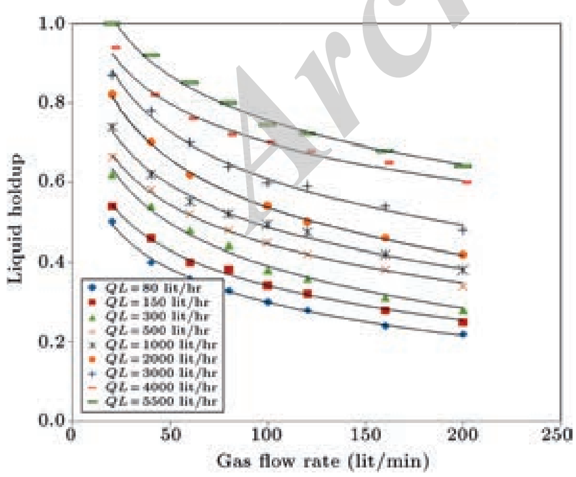

Figure 4: Variation of liquid holdup by gas flow rate in a pipe with $\Theta = 20°$ and $D = 0.03$ m.

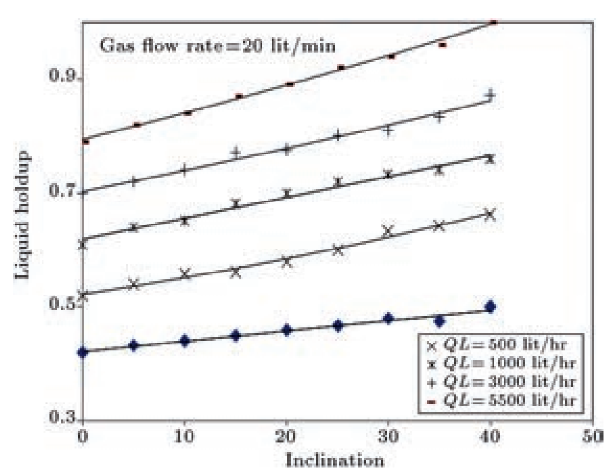

Figure 7: Variation of liquid holdup with inclination angle at GF = 20 l/m and $D = 0.03$ m.

liquid slippage increases, and as the gas inertia forces are not enough to overcome it, the gas phase cannot carry over the liquid, resulting in an increase in liquid holdup. The value of liquid holdup seems to increase when the inclination angle increases, because the liquid is naturally pulled to the bottom by gravity. Figure 7 shows this change at 20 l/min gas flow rate. In other gas flow rates, the changes of liquid holdup by the inclination angle are the same.

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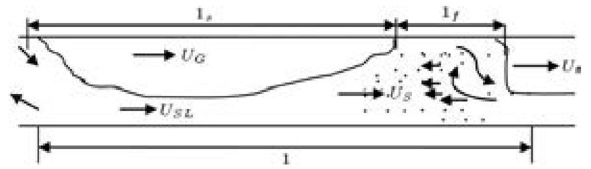

Figure 8: The physical model for slug flow.

The experimental results show that the inclination angle has a significant effect on liquid holdup. In horizontal and inclined pipes, liquid holdup goes in the same direction as gas and liquid flow rates. The value of liquid holdup seems to increase when the inclination angle increases, because the liquid was naturally pulled to the bottom by gravity when the inclination angle increased.

3. Mathematical model

3.1. The model governing equations

The first step in evaluating the average liquid holdup of slug flow is to predict the major variables affecting it. So, a sketch of an idealized slug has been presented in Figure 8. In this figure, l is the length of the slug unit that consists of two sections: a liquid slug zone of length l_S and a film zone of length l_f [3].

Assume a liquid film of a uniform thickness overlaid by an elongated gas bubble. The liquid and gas mass balances over the slug unit yield:

$$U_{SL} = U_S H_S \frac{l_S}{l} + U_f H_f \frac{l_f}{l} \tag{1}$$

$$U_{SG} = U_S(1 - H_S)\frac{l_S}{l} + U_G(1 - H_f)\frac{l_f}{l}, \tag{2}$$

where H_s and H_f are the liquid holdups in the slug and film zones, and U_{SL} and U_{SG} are the superficial liquid and gas velocity, respectively. Following Taitel and Barnea [8], the aerated liquid slug velocity, U_S, was evaluated by:

$$U_S = U_m = U_{SL} + U_{SG}, \tag{3}$$

where U_m is the superficial velocity of the gas–liquid mixture. The value of l was evaluated by the Dukler and Hubbard [4] correlation. They presented this correlation based on experimental pressure drop data for a 0.0381 m diameter smooth pipe.

$$l = \frac{U_t}{\upsilon_S}. \tag{4}$$

Gregory and Scott [22] used slug data in 0.75 and 1.5 inch pipes and correlated the slug frequency, υ_S, by the slug Froude number, Fr_s, where, in the derived equation, g is the gravity acceleration constant and D is the internal pipe diameter:

$$\upsilon_S = 0.0226\ \mathrm{Fr}_s^{1.2}, \tag{5}$$

$$\mathrm{Fr}_s = \left(\left(\frac{U_{SL}}{gD}\right) / (19.75/U_m + U_m)\right), \tag{6}$$

where U_m, the mixture velocity, is in ft/s and the slug frequency is in s^{-1}. Following Taitel and Barnea [8], U_t in Eq. (4), which is the translational velocity, is calculated by the following equation for slug flow in horizontal pipes:

$$U_t = 1.2\ U_m + 0.54\sqrt{gD}. \tag{7}$$

Xin et al. [23] concluded that when the gas velocity is between 3 and 6 m/s and the liquid velocity is lower than 2 m/s (the present data bank relies on this condition), the slug length is between 10 and 20 D, the average of which (15 D) is considered the fixed slug length in the present model. The value of slug length for other different conditions can be estimated from previous studies, such as Xin et al. [23].

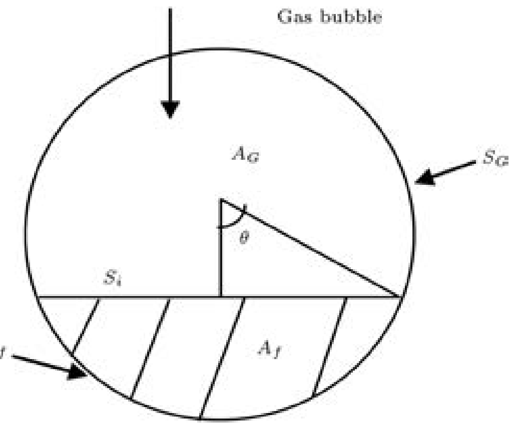

Figure 9: Film zone flow geometry.

Liquid mass balance, relative to a coordinate system that travels at the translational velocity of the slug unit, yields [3]:

$$(U_t - U_f)H_f = (U_t - U_S)H_S. \tag{8}$$

Unlike the Taitel and Barnea [8] model, the film region thickness is assumed uniform and end effects have been neglected, so the film zone can be considered as stratified flow. Figure 9 displays the film zone flow geometry. A momentum balance on the liquid and gas yields:

$$-A_f\frac{dP}{dx} = \tau_f S_f - \tau_i S_i, \tag{9}$$

$$-A_G\frac{dP}{dx} = \tau_G S_G + \tau_i S_i. \tag{10}$$

In the above equations, A_f & A_G refer to the film and gas cross-section areas, respectively, with τ_f, τ_G & τ_i referring to the film, gas and interfacial zone shear stresses, respectively, with P standing for pressure.

The pressure gradient at the pipe cross-section within the film zone is uniform. Simplification of Eqs. (9) and (10) results in:

$$\frac{(\tau_G S_G + \tau_i S_i)}{A_f} = \frac{(\tau_f S_{f-\tau_i S_i})}{A_G}. \tag{11}$$

The shear stresses, τ_f, τ_G and τ_i, are given by:

$$\tau_f = \frac{1}{2} f_f \rho_L U_f^2, \tag{12}$$

$$\tau_G = \frac{1}{2} f_G \rho_G U_G^2, \tag{13}$$

$$\tau_i = \frac{1}{2} f_i \rho_G (U_G - U_f)^2 \tag{14}$$

where f_i, f_G and f_f are the Fanning friction factors of the liquid film, gas bubble and gas–liquid interface, respectively, ρ_L and ρ_G being liquid and gas density. For slug flow in a smooth pipe, f_f and f_G are expressed in terms of the Reynolds number of each phase and are given by Orell [9]:

$$f_f = \frac{C_f}{(\rho_L U_f D_{hf}/\mu_L)^n}, \tag{15}$$

$$f_G = \frac{C_G}{(\rho_G U_G D_{hG}/\mu_G)^m}. \tag{16}$$

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The hydraulic diameters (D_{hf}, D_{hG}) of the liquid film and gas bubbles are [9]:

$$D_{hf} = \frac{4A_f}{S_f}, \tag{17}$$

$$D_{hG} = \frac{4A_G}{(S_G + S_i)}. \tag{18}$$

For laminar flow of both phases:

$$C_f = C_G = 16 \qquad n = 1. \tag{19}$$

For turbulent flow of both phases:

$$C_f = C_G = 0.046 \qquad m = 0.2. \tag{20}$$

The interfacial friction factor suggested by Hanratty and Cohen [24] correlation:

$$f_i = 0.0142. \tag{21}$$

The above value can be used for slug flow in inclined pipes too. The film zone flow geometry is shown in Figure 9 and the liquid holdup in the film zone is defined by one half of the angle that subtends the liquid–gas interface, Θ [9]:

$$A_f = \frac{D^2}{4}(\theta - 0.5\sin 2\theta), \tag{22}$$

$$A_G = \frac{D^2}{4}(\pi - \theta + 0.5\sin 2\theta). \tag{23}$$

The wet perimeter of film, gas and interfacial zones (S_f, S_G, S_i) are expressed by:

$$S_f = D\theta, \qquad S_G = D(\pi - \theta), \qquad S_i = D\sin\theta, \tag{24}$$

$$H_f = \frac{\theta - 0.5\sin 2\theta}{\pi}. \tag{25}$$

3.2. Computational procedure

The simple sub model of Taitel and Barnea [8] for horizontal gas–liquid slug flow has been reformulated by Orell [9]. This model could not predict the slug unit liquid holdup theoretically, so the experimental correlation of Andreussi and Bendiksen [10] was used for slug holdup calculations.

Also, the slug length is not obtainable from this model. In the present study, no experimental correlations for film and slug holdups are used, the slug unit liquid holdup being calculated by the equations referred to in the following solution procedure:

1. Following parameters e.g. ρ_L, ρ_G, μ_L, μ_G, D, U_{SL}, U_{SG} should be defined as inputs.
2. Eqs. (1), (2) and (11) constitute the basic equations of the model. Substituting the required parameters from Eqs. (8) and (12)–(25), result in a set of equations.
3. The equations are solved analytically providing the slug flow characteristic variables, U_f, U_G, H_f, H_S. A computer program was used to solve the equations;
4. Using the equations below, the average liquid holdup in slug unit, H_{SU}, and the average void fraction in the slug unit, E_{SU}, were defined:

$$H_{SU} = H_S\frac{l_S}{l} + H_f\frac{l_f}{l}, \tag{26}$$

$$E_{SU} = (1 - H_S)\frac{l_S}{l} + (1 - H_f)\frac{l_f}{l}. \tag{27}$$

4. Results and discussion

The proposed model was tested against the experimental liquid holdup data of air–water from an available setup and other air–water and air–oil data sources in horizontal pipes. The specification of these data sources are recorded in Table 1.

The developed model was compared with Gomez et al. [14], Leung [25], Abdul-Majeed [16] and Zhang et al. [13] models. Parameter ranges of these empirical correlations are shown in Table 2.

The Leung [25] model was evaluated based on a wide experimental database by Ghajar and Woldesemayat [26]; they suggested this correlation is an accurate empirical correlation for holdup calculations in horizontal and inclined pipes.

The parameter range for which this assessment was carried out is shown in Table 2. The Zhang model [13] is a mechanistic model for inclined and horizontal pipes, which is valid for gas velocities over 0.1 m/s.

The accuracy of the models was evaluated by calculating the percentage error of individual data points (IPE), the Average Percentage Error (APE) and Absolute Average Percentage Error (AAPE) of each data source, which is defined below [9]. The regression coefficient (R^2) of all data sources is also reported.

$$\mathrm{APE} = \frac{\sum_{i=1}^{n}\mathrm{IPE}_i}{n}, \qquad \mathrm{AAPE} = \frac{\sum_{i=1}^{n}|\mathrm{IPE}_i|}{n}, \tag{28}$$

$$\mathrm{IPE} = \frac{\text{Predicted} - \text{Measured}}{\text{Measured}} \times 100. \tag{29}$$

The result of this comparison has been shown in Table 3. This table shows that the theoretical model matches the holdup of 125 data points within an average percent error of 6.72 and absolute average percent error of 11.14. As demonstrated in Table 3, the error of presented model was the lowest when compared to other methods studied.

By contrast, Abdul-Majeed's [16] model, developed to calculate slug liquid holdup in horizontal and slightly inclined two-phase slug flow, has shown scattered results. This empirical model is not valid for superficial liquid velocities higher than 2.316 m/s, so outlier data are eliminated for comparison.

Abdul-Majeed [16] and his colleagues did not provide verification results of their model for horizontal pipes. They mixed horizontal data with upward and downward inclined flow data ($-5 < \Theta < 10$), then compared and calculated the APE of their model with horizontal-upward inclined flow and horizontal-downward inclined flow data sources.

The error was lower than $\pm 10\%$, but comparison of Abdul-Majeed's [16] model to the experimental data of horizontal pipes shows that this empirical model is not accurate enough for horizontal pipes.

One reason for this problem could be that the empirical correlation presented by Abdul-Majeed [16] only contained liquid to gas velocity ratios. Corresponding to literature, the empirical correlations which used other significant flow characteristics (e.g. gas density, liquid density, surface tension, etc.) are more accurate.

Gomez et al. [14] developed a dimensionless correlation for calculating slug liquid holdup in horizontal and upward vertical pipes. The dimensionless groups of this correlation are inclination angle and Reynolds number.

The Reynolds number was defined by a mixture velocity without considering the slippage of phases, and the effect of the gas phase was ignored for calculation of density and viscosity. Thus, it is not surprising that the recent model, which considers the slippage of phases and effects of both phases in flow characteristics, provides more accurate results.

Armand and Massima's model [25] was considered in the category of $K\varepsilon_H$ correlations by Ghajar and Woldesemayat [26].

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Table 1: Details of data bank used for model validation.

Data source	Diameter (m)	Gas velocity (m/s)	Liquid velocity (m/s)	Fluids	Data points
Available set up	0.03	1.7–4.2	0.39–1.8	Air–water	25
Heywood and Richardson [27]	0.042	0.3–4.5	0.48–1.47	Air–water	43
Abdul-Majeed [16]	0.051	1.98–3.07	0.22–1.52	Air–oil	9
Nicholson et al. [5]	0.0512	0.02–4.0	0.12–1.86	Air–oil	31
Kokal and Stanislav [6]	0.0763	0.1–5	0.5–1	Air–oil	17

Table 2: The range of flow parameters of empirical correlations.

Empirical correlation	Angle (°)	Diameter (cm)	Fluids
Gomez et al. Model [14]	0–90	7.6, 17.8, 20.3, 5.1	Air–kerosene, freon–water, air–water, nitrogen–diesel, air–oil
Leung [25]	0–90	5.25, 10.22, 2.54, 3.81, 4.55, 7.80, 1.9, 5.08, 1.27	Air–water, air–kerosene, natural gas–water
Abdul-Majeed [16]	−10–+9	7.62, 5.10, 17.145, 20.32, 2.58	Air–kerosene, air–light oil, freon–water, air–water, nitrogen–diesel

Table 3: Comparison of slug unit liquid holdup predictions with data sources.

Data source	Present model		Gomez et al. model [14]		Leung [25]		Abdul-Majeed [16]		Zhang et al. [13]	
	APE	AAPE	APE	AAPE	APE	AAPE	APE	AAPE	APE	AAPE
Available set up	0.394	1.44	33.69	33.69	26.42	26.42	56.01	56.01	26.53	26.53
Heywood and Richardson [27]	11.62	15.93	29.57	29.57	1.75	3.89	56.27	56.27	38.48	38.48
Abdul-Majeed [16]	14.61	14.61	8.90	8.90	41.67	41.67	12.61	12.61	43.08	43.08
Nicholson et al. [5]	11.49	12.80	34.69	36.56	1.03	15.44	36.14	36.14	38.45	38.45
Kokal and Stanislav [6]	6.24	8.41	24.19	32.30	27.76	27.76	34.67	34.67	36.43	36.43
Total error	6.72	11.14	29.12	30.72	12.46	16.87	44.26	44.26	36.13	36.13

Table 4: Comparison of slug unit liquid holdup predictions with all data sources.

	Presented model	Gomez et al. model [14]	Leung [25]	Abdul-Majeed [16]	Zhang et al. [13]
R^2	0.94	0.83	0.89	0.78	0.82

This correlation consists of constant (K) and some functional multiple of the no-slip void fraction, ε_H. This model was not developed specifically for slug flow, but it was based on a wide range of data bank for horizontal two phase flow [26]. This has better results compared to other empirical models.

It can be concluded that the empirical models which are based on a wide range of data are more reliable than other empirical correlations. However, they depend on their data banks. The present model, which is a combination of some correlations and theoretical approaches, is still more accurate.

In the Zhang model [13], slug liquid holdup can be calculated by a balance between the turbulent kinetic energy of the liquid phase and the surface free energy of dispersed spherical gas bubbles. To evaluate slug liquid holdup in various inclination angles, these two terms are entered in the Zhang model [13]. As Table 3 shows, the Zhang et al. [13] model does not produce better accuracy compared to the present model.

One of the important reasons can be found in the slug length term. In the Zhang model [13], 32D is used as the slug length value (l_s), but as mentioned above, Xin et al. [23] concluded that when the gas velocity is between 3 and 6 m/s and the liquid velocity is lower than 2 m/s (the conditions of the present data bank), the slug length is between 10 and 20 D. So, it seems that the Zhang model [13] is not appropriate for the presented data and the high error of the model might be the result of this fact. Hence, for horizontal pipes, the recent model is suggested.

For a better comparison, the proposed model results are compared to all experimental data sources, the results being shown in Figure 10. The regression coefficient (R^2) is also calculated for all models in Table 4.

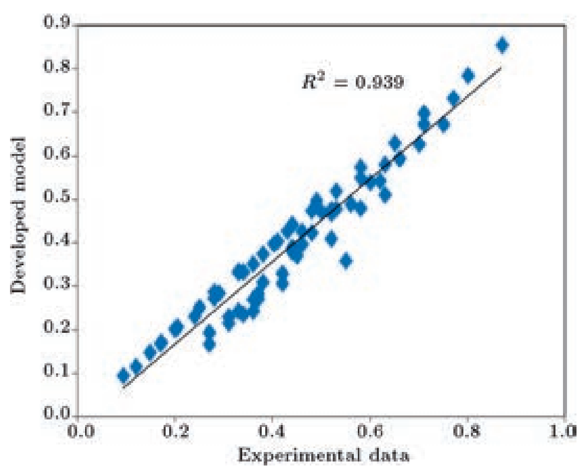

Figure 10: Comparison of developed model to all experimental data.

The presented model matches the data sources perfectly with a regression coefficient of 0.94. Meantime, the results of the developed model versus gas superficial velocity for air–water and air–oil data sources are shown in Figures 11 and 12, confirming the model validity.

5. Conclusions

In the present model, the value of slug unit length (l) was calculated by the Dukler and Hubbard [4] correlation. On the

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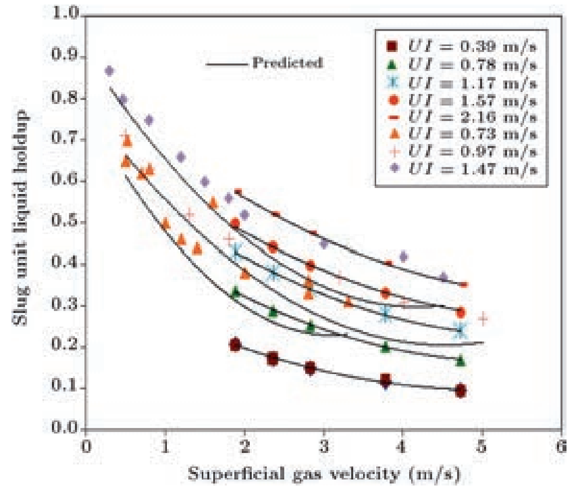

Figure 11: Comparison of predicted slug unit liquid holdup and air–water experimental data.

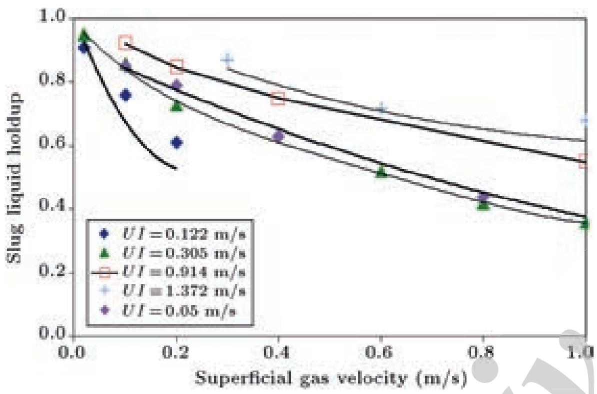

Figure 12: Comparison of predicted slug unit liquid holdup and air–oil experimental data.

other hand, liquid slug length (l_s) was estimated by the Gregory and Scott [22] correlation. Two critical unknowns for calculating slug unit liquid holdup (H_S, H_f) are found by solving momentum equations.

So, unlike the Orell model [9], the calculation of slug liquid holdup is not dependent on a specific range and is not as sophisticated as the Taitel and Barnea [8] model.

The experimental work was carried out to gather liquid holdup air–water data in horizontal and inclined pipes. The proposed model was tested extensively against 125 experimental data sets. The average percentage error and absolute average percentage error were calculated for model comparison.

The results show that the total average percentage error of the present model was 6.72 and the average absolute percentage error was 11.14, both of which are lower than other empirical and mathematical models. These results substantiate that the new modified model is independent of various physical properties of flowing fluid, so it has an adequate prediction potential for a broad range of situations for slug liquid holdup calculations in horizontal pipes.

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