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MEASUREMENT OF PRESSURE LOSS IN GAS-LIQUID TWO-PHASE FLOW IN A HORIZONTAL PIPELINE

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Abstract. Multiphase flow occurs often in nature as in industry applications, and its knowledge is of particular importance. One of the most common types of multiphase flow is the two-phase gas-liquid flow, who is frequently found within the oil and natural gas industry. The behavior of the gas-liquid flow depends on liquid and gas flow rates at the inlet of the pipe, besides the type of the fluid and geometric properties. Two-phase flow in pipeline can be present different geometric distributions named flow patterns that depend mainly on oil and gas superficial velocities and pressure. In this work has been developed an experimental study about air-water two-phase flow in a horizontal pipeline to measure the pressure drop for various volumetric flow conditions on pipe entrance, both to gas as liquids, comparing later with the Moody diagram and the Lockhart-Martinelli's separated flow model. The flow range analyzed in this work corresponding to slug flow pattern, which is one of the most common patterns for oil extraction. To accomplish this goal, an experimental facility at GPEM-UniBrasil lab was used. The analysis performed in this paper seeks a best understanding of the flow behavior, which is of great importance in the industry.

Keywords: multiphase flow, pipe, bubble, pressure drop

1. INTRODUCTION

Multiphase flow can be defined as any flow containing more than one phase or component (Brenen, 2003). They are found in various areas of nature as well as in various technological applications, and are used in industrial processes involving boiling of liquids and condensation of vapors, applied in petroleum refineries, refrigeration systems, chemical processes and nuclear energy. In the case of oil, the multiphase flow is found in the wells production column and in the production pipelines where the phases are transported over long distances. The knowledge and determination of flow characteristics, such as pressure drop or friction factor, becomes one of the key points for determining the design of the facilities and carrying out the activities of these industries, because these flows determine the flow pattern which occurs at a particular location in the extraction or transportation pipeline.

One of the main multiphase flow patterns analyzed in the literature is the two-phase gas-liquid slug flow in pipelines, whose geometrical distribution depends on fluid properties, flow channel size, pipe inclination and orientation, and fluid flow (Julia et al., 2012) is characterized by the intermittent passage of a unitary cell composed of a swirling liquid with a large amount of movement followed by a compressible gas bubble which slides on the liquid film. According to Coates and Pressburg (1959), with the presence of a second phase, there is an increase in the loss of charge when compared to the resultant if only one phase were flowing with equal mass flow in the flow.

The methodology of correlating the properties of two-phase flow to the flow of each phase separately is widely used by researchers in the field, and has as one of the precursors the researchers Lockhart & Martinelli (1949). In their work, the pressure drop in horizontal ducts can be determined from the pressure drops that would have the phases flowing separately, considering a correction factor. Chisholm (1967), based on the researches of Lockhart and Martinelli, presented a simple correlation to determine this correction factor as a function of the ratio of liquid and single-phase flow pressure drops.

The experimental measurements showed in this work is based in Darcy's friction factor and Lockhart & Martinelli parameter, with the objective to study and analyze the pressure drop and the friction factor of the two-phase flow in a horizontal pipeline. To achieve the objectives of this work, the experimental facilities of the Mechanical Engineering Research Group (GPEM) of UniBrasil was used.

2. METHODOLOGY

According Shoham (2006), the pressure drop in the two-phase flow can be calculated with the sum of the friction, gravity and acceleration components.

$$\left.\frac{dP}{dL}\right)_{TP} = \left.\frac{dP}{dL}\right)_F + \left.\frac{dP}{dL}\right)_G + \left.\frac{dP}{dL}\right)_A \tag{1}$$

where the index TP, F, G and A correspond to two-phase flow, frictional, gravitational and accelerational components of the pressure drop.

In the case of flow in a horizontal pipeline, the gravitational term can be neglected. Thus, the equation for frictional pressure drop can expressed as

$$\left.\frac{dP}{dL}\right)_F = \left.\frac{dP}{dL}\right)_{TP} - \left.\frac{dP}{dL}\right)_A \tag{2}$$

In turn, the acceleration pressure drop can be calculated using the homogeneous model given by Eq. (3).

$$\left.\frac{dP}{dL}\right)_A = G^2 x \frac{d\upsilon_G}{dP} \left.\frac{dP}{dL}\right)_{TP} \tag{3}$$

where G is the mass flow of the mixture, x is the quality and υ_G is the specific volume of the gas. On the other hand, one of the methods to calculate the pressure drop by friction is the model of Lockhart & Martinelli (1949) for separated flow, who proposed a model where a correction factor for a single-phase pressure drop is used, considering the interaction between the two phases of the pressure drop.

The pressure drop is calculated considering that single-phase flow in the pipeline and from surface velocities, where in most cases it is higher than the average flow velocity. The equation for the liquid or gas surface velocity (J_φ) and the mixture (J) are given, respectively, by

$$J_\alpha = \frac{Q_\alpha}{A} \tag{4}$$

$$J = J_L + J_g \tag{5}$$

where A is the cross-sectional area of the tube, α refers to gas or liquid and Q_α are the volumetric flow rates of the liquid or the gas. To determine the behavior of the flow in the pipeline, it is necessary to calculate the Reynolds number of the mixture, which is defined by Garcia et al. (2003) as follows:

$$Re_m = \frac{\rho_L JD}{\mu_L} \tag{6}$$

where ρ_L is the density of the liquid, μ_L is the viscosity of the liquid and D is the internal diameter of the pipe. To $Re_m < 2300$ consider laminar flow and $Re_m > 2300$ turbulent flow. Garcia et al. (2003) also indicates the equation of the load loss using the Fanning friction factor, which can be adapted to the Darcy-Weisbach friction factor. The pressure drop in a pipe (ΔP) can be calculated by the equation:

$$\frac{\Delta P}{L} = \frac{f}{D} \frac{\rho V^2}{2} \tag{7}$$

where f is the Darcy's friction factor, L is the length of the pipe, V is the velocity of the phase and ρ is the density of fluid.

By relating the Darcy-Weisbach's friction factor to the Reynolds number and the relative roughness we can obtain the graph proposed by Moody (1944). The Moody diagram is widely used to determine the friction factor in pipes based on load losses, flow behavior and roughness of the pipe through which the fluid flows (Garcia, 2016).

Lockhart & Martinelli (1949) consider that the loss of charge of the two-phase flow is the same if only one of the phases flows through the pipeline with the same flow and conditions of temperature and pressure, which is correlated with the dimensionless parameter X, showed in Fig. 1 and given per:

$$X^2 = \frac{f_L \rho_L J_L^2}{f_g \rho_g J_g^2}$$

(8)

The relationship between two-phase pressure drop in each phase is given by:

$$\phi_\beta^2 = \frac{\left.\dfrac{dP}{dx}\right|_M}{\left.\dfrac{dP}{dx}\right|_\beta} = \frac{\text{Frictional pressure gradient of the mixture}}{\text{Pressure gradient of the } \beta\text{-phase}}$$

(9)

In Eq. (9), the variable φ_β is a dimensionless parameter that are a function of the dimensionless variable X and the pressure drop of the β-phase can be the pressure gradient that would exist if only liquid or gas flowed through the pipeline. In 1967 Chisholm transformed into equations what was previously graphically demonstrated by Lockhart and Martinelli (1949), as follows:

$$\varphi_g^2 = 1 + CX + X^2$$

(10)

$$\varphi_L^2 = 1 + \frac{C}{X} + \frac{1}{X^2}$$

(11)

The parameter C is dependent on the behavior of the flow, turbulent or laminar. Table 1 shows the suggested values for the parameter C by Chisholm (1967), according to the regime of each phase of the flow.

Table 1. Values of C for pipes (Chisholm, 1967)

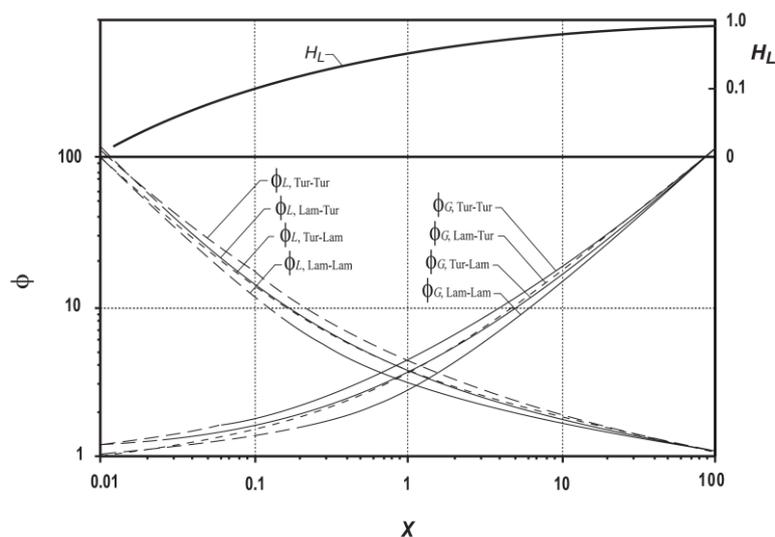| Liquid phase	Gas phase	C
Turbulent	Turbulent	20
Laminar	Turbulent	12
Turbulent	Laminar	10
Laminar	Laminar	5

Figure 1. Diagram of the parameter X developed by Lockhart & Martinelli (Shoham, 2006)

3. EXPERIMENTAL PROCEDURE

To obtain the pressure drop of a liquid-gas flow in a horizontal pipeline, the experimental facilities appropriate in GPEM-UniBrasil were used, whose operating scheme is shown in Fig. 2, consisting of an acrylic pipe with internal diameter of 25.8 mm and a test line with 11 meters of length, having rotameters and manometers for measuring gas and liquid flow.
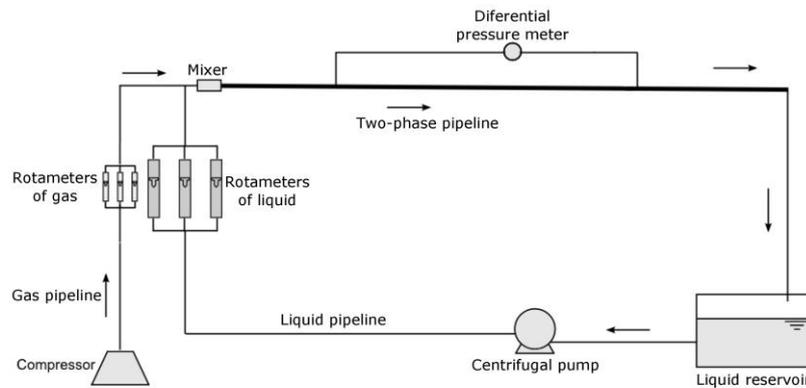

Figure 2. Schematic of the GPEM/UniBrasil experimental facilities to study the two-phase flow characteristics

Figure 3. Photo of the experimental facilities present in GPEM/UniBrasil

The pressure drop was measured in two sections of the experimental facilities, which are 5.15m and 7.93m at downstream from the beginning of the horizontal two-phase pipeline. These pressure drop measurements were performed between different volumetric flow rates (Q), which were measured with the gas and liquid rotameters.

The combination of flow rates used in this work is shown in Tab. 2, where are proportional to the values of the liquid and gas surface velocities in the inlet section and the differential pressure between the two sections mentioned above.

Table 2. Liquid / gas volumetric flow rates and respective pressure drop in cmH2O

Q_L (LPM) \ Q_G (LPM)	0	5	10	15	20	25	30	35	40
10	0,200	0,365	0,420	0,480	0,555	0,640	0,695	0,820	0,935
15	0,375	0,590	0,692	0,772	0,917	1,086	1,405	1,804	2,080
20	0,630	0,844	0,991	1,209	1,468	1,666	2,518	2,834	3,017
25	0,893	1,066	1,344	1,593	1,937	2,130	2,603	3,038	3,384
30	1,191	1,403	1,708	2,098	2,425	2,817	3,068	3,405	3,751

To obtain the pressure difference between the two measurement sections a digital differential pressure sensor of up to $150 cmH_2O$ column height was used.

By performing a rearrangement of Eq. (6) and considering the distance between the pressure ports, the internal diameter of the pipe and the surface velocity of the mixture of each measurement point, the values of the Darcy's friction factor can be calculated, which were compared with the values obtained from the Moody diagram (1944). To determine the Lockhart-Martinelli's dimensionless number, a flow with both turbulent phases was considered.

4. RESULTS

The pressure drop was measured by setting the liquid flow rate and changing the gas flow rate. Using the Reynolds number equation it can be seen that the flow under analysis is turbulent. Both the friction factor and the Reynolds number are dependent on the surface velocity of the mixture, which was calculated with the liquid and gas flow data.

In Fig. 4, the values obtained for the pressure drop with respect to the superficial velocity of the gas, calculated through the Eq. (7), are shown.
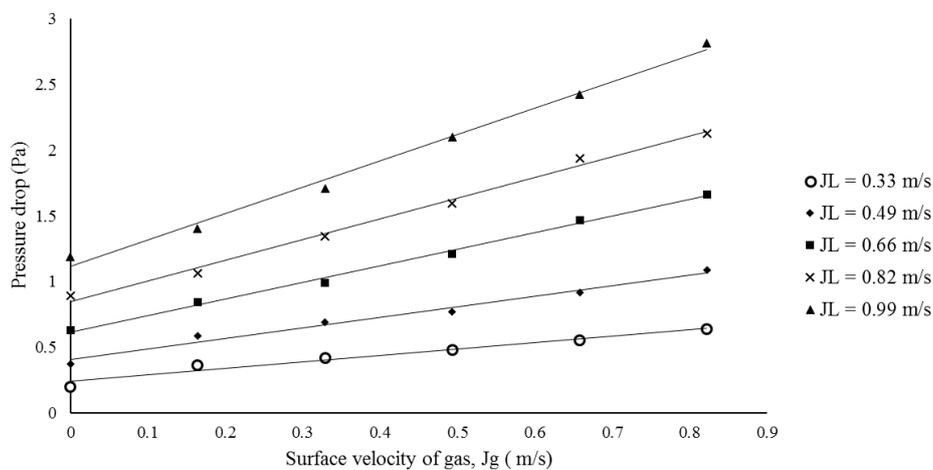

Figure 4. Pressure drop in relation of the surface velocity of gas

The increase in pressure drop has been observed as the gas flow rate is increased. It can be explained as the more superficial velocity of gas, the more superficial velocity of the mixture. This increase of superficial velocity of the mixture also increase the shear stresses in the walls of the pipe as observed in Eq. (7).

When applying the surface velocities of the mixture and its respective pressure drops in Eq. (7), we obtain the graph referring to the Moody diagram, shown in Fig. 5.
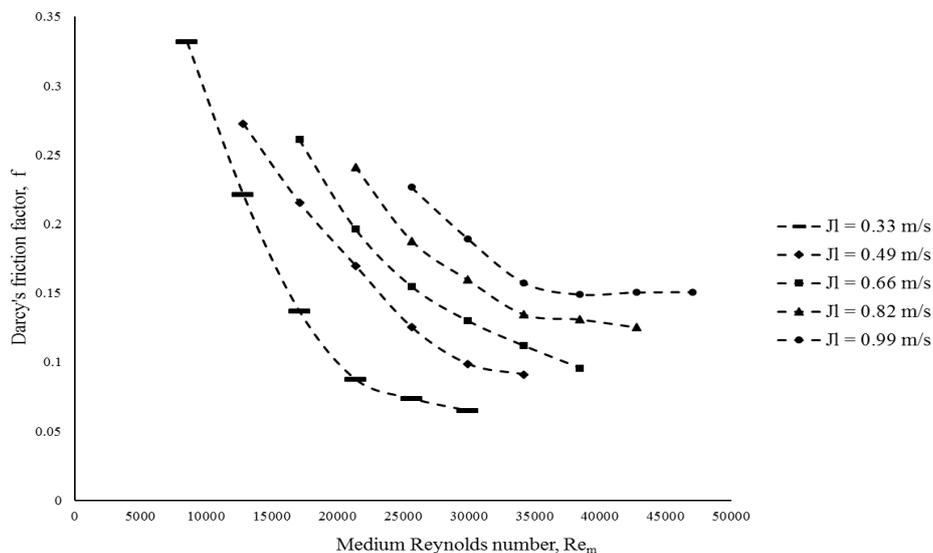

Figure 5. Darcy's friction factor in relation of the mixture Reynolds

In Fig. 5, it can be observed that for constant liquid surface velocity values, the surface velocity of the blend too increases, i.e. the Reynolds number of the blend causes a decrease in the friction factor. This is due to the increase in flow turbulence, as can be observed in the Moody diagram.

For moderated Reynolds number, the friction factor is high, corresponding to the beginning of the turbulent region in the Moody diagram. For greater values of the Reynolds number, the friction factor decreases until the region where it depends only on the roughness of the pipe.

It has also been observed that by varying the gas flow rates to constant values of the surface velocity of the mixture, the friction factor decreases as the gas ratio increases, as can be seen in Fig. 6.
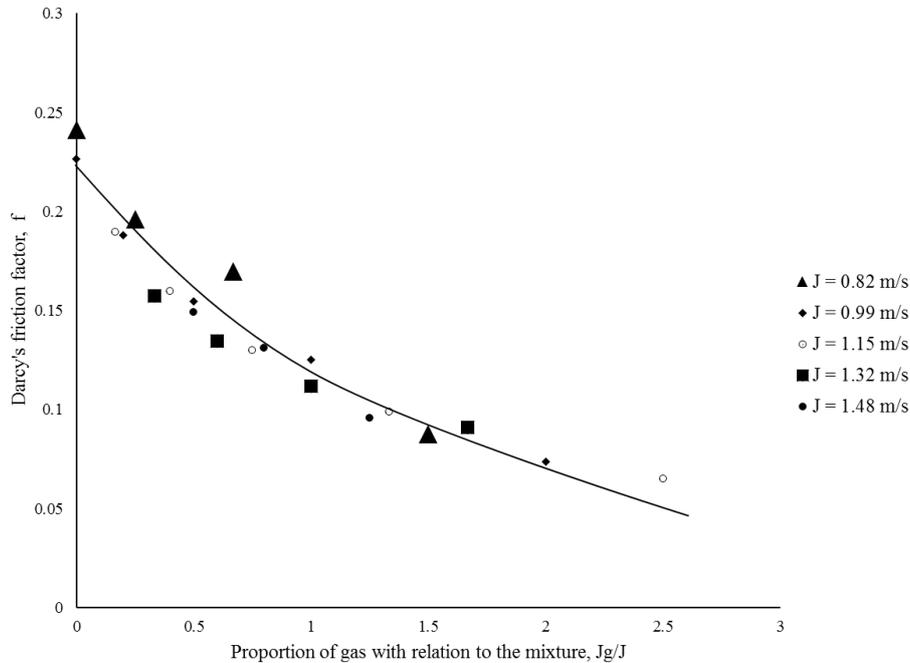

Figure 6. Darcy's friction factor in proportion of the gas in the mixture

The higher the amount of gas the lower the friction in the flow, because the gas has a viscosity lower than the water, causing lower shear stresses in the wall for the gas than for the liquid.

An important evaluation of the Lockhart-Martinelli's separated flow model is to compare with experimental data. This comparison can be observed in Fig. 7, for every superficial velocity of liquid for different superficial velocities of gas.

In Fig. 7, the experimental data of frictional pressure drop can be obtaining by applying Eq. (2). The frictional pressure drop has been calculated by homogeneous model from Eq. (3). For experimental data, the accelerational pressure drop has been calculated, giving values lower than 2% of total pressure drop. By subtracting these accelerational pressure drop, the values of frictional pressure drop from experiments and Lockhart-Martinelli's model can be compared.

For J_L=0.33m/s (Fig. 7a), the behavior of experimental data is similar to the separated model for low superficial velocity of gas, while the separated model is between 7% and 25% from experiments for high gas flow rates. As a preliminary understanding, it can occur due to the higher gas flow rate the higher effects of mixture and effects of the interface between the phases, which results in more interaction between the phases, contrary to the hypotheses of separated flow.

For moderated liquid flow rates (Fig. 7b,c), it can be observed that the experimental data is under the model for low gas flow rate and above the model for high gas flow rate. Following the reasoning of the previous paragraph, for higher and lowers gas flow rates, the behavior of separated model is more deviated from the experiments because a great difference between liquid and gas can cause more interactions between the phases, misrepresenting the hypotheses of separated flow. The error of this data is between 6% and 28%.

Finally, the previous observations and preliminary understanding can also be observed in Fig. 7d,e for high liquid flow rates. If the gas flow rate is too high, the difference between experiments and model is lower than for low gas flow rates, with error between 1% and 38%.
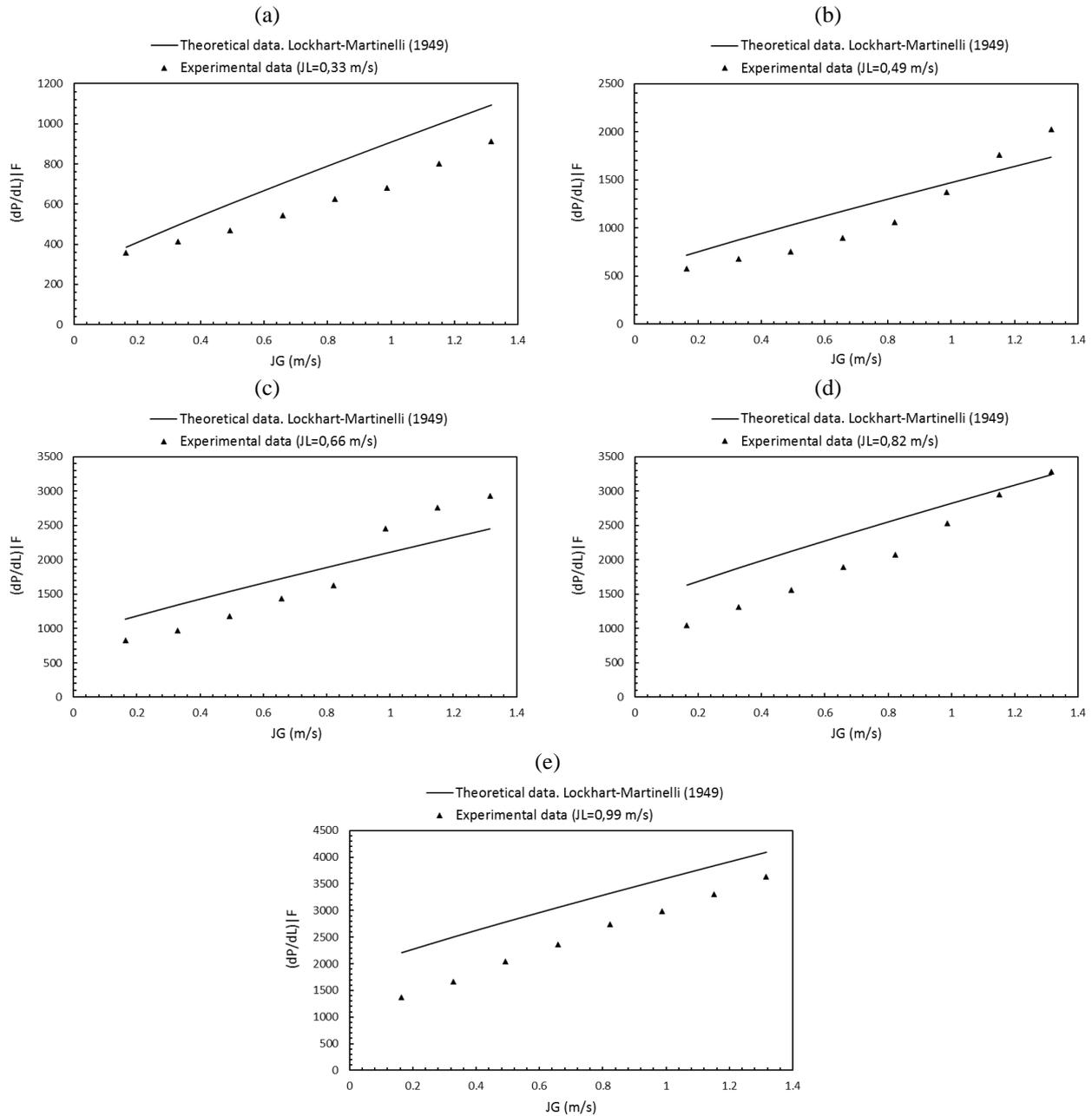
Figure 7. Comparison between the values of the friction factor loss from the experiments performed and the one proposed by Lockhart & Martinelli (1949) for (a) J_L=0.33 m/s, (b) J_L=0.49 m/s, (c) J_L=0.66 m/s, (d) J_L=0.82 m/s, and (e) J_L=0.99 m/s.

Another evaluation data is to analyze the behavior of the parameter ϕ_L^2 with respect to the X Lockhart-Martinelli's parameter. Theoretically, this behavior is given by the curve shown in Fig. 1, with the experimental data analyzed in this work in the lower right region of the curve, for X-values between 6 and 108, and values of ϕ_L^2 between 0.74 and 3.82. Thus, the experimental data is shown in Fig. 8, with a curve fitting of power function to compare the data obtained between the model and the experiments.

In Fig. 8, can be observed that the values are similar to the model. For example, for Fig. 8a can be calculated values of ϕ_L^2 for different values of X. For X-values of 10, 20, 30 and 40 can be obtained ϕ_L^2-values of 2.47, 1.45, 1.43 and 1.24. These and other values of ϕ_L^2 and X can be observed in Tab. 3, where can be compared with the values obtained –approximately– from Fig. 1.

Table 3. Values of X and ϕ_L^2 obtained from the experiments for comparison with Lockhart-Martinelli's chart.

X \ ϕ_L^2	ϕ_L^2 (Lockhart-Martinelli's chart)	ϕ_L^2 (J_L=0.33m/s)	ϕ_L^2 (J_L=0.49m/s)	ϕ_L^2 (J_L=0.66m/s)	ϕ_L^2 (J_L=0.82m/s)	ϕ_L^2 (J_L=0.99m/s)
10	1.90	2.47	2.86			
20	1.50	1.75	1.80	1.98	1.79	1.69
30	1.40	1.43	1.37	1.47	1.40	1.36
40	1.25	1.24	1.13	1.19	1.17	1.17
50	1.20		0.98	1.02	1.02	1.04
60	1.20			0.89	0.91	0.94
70	1.15			0.79	0.83	0.87
80	1.15				0.76	0.81
90	1.10				0.71	0.76
100	1.10					0.72

In Fig. 8, can be observed that the predict values for the model differ considerably from experimental data for lower and higher X-values. Nevertheless, these values have better approximation with experiments for X-values between 20 and 40. These divergences may occur because the curve fitting has limitations to represent lower X-values. It could be improved by use a better equation to represent the curves.

However, in Fig. 8 also be noted that the minimum values of ϕ_L^2 for the experiments realized in this work are in the order of 0.75, which differs anyway from the model, whose minimum value in this range is in the order of 1.10.
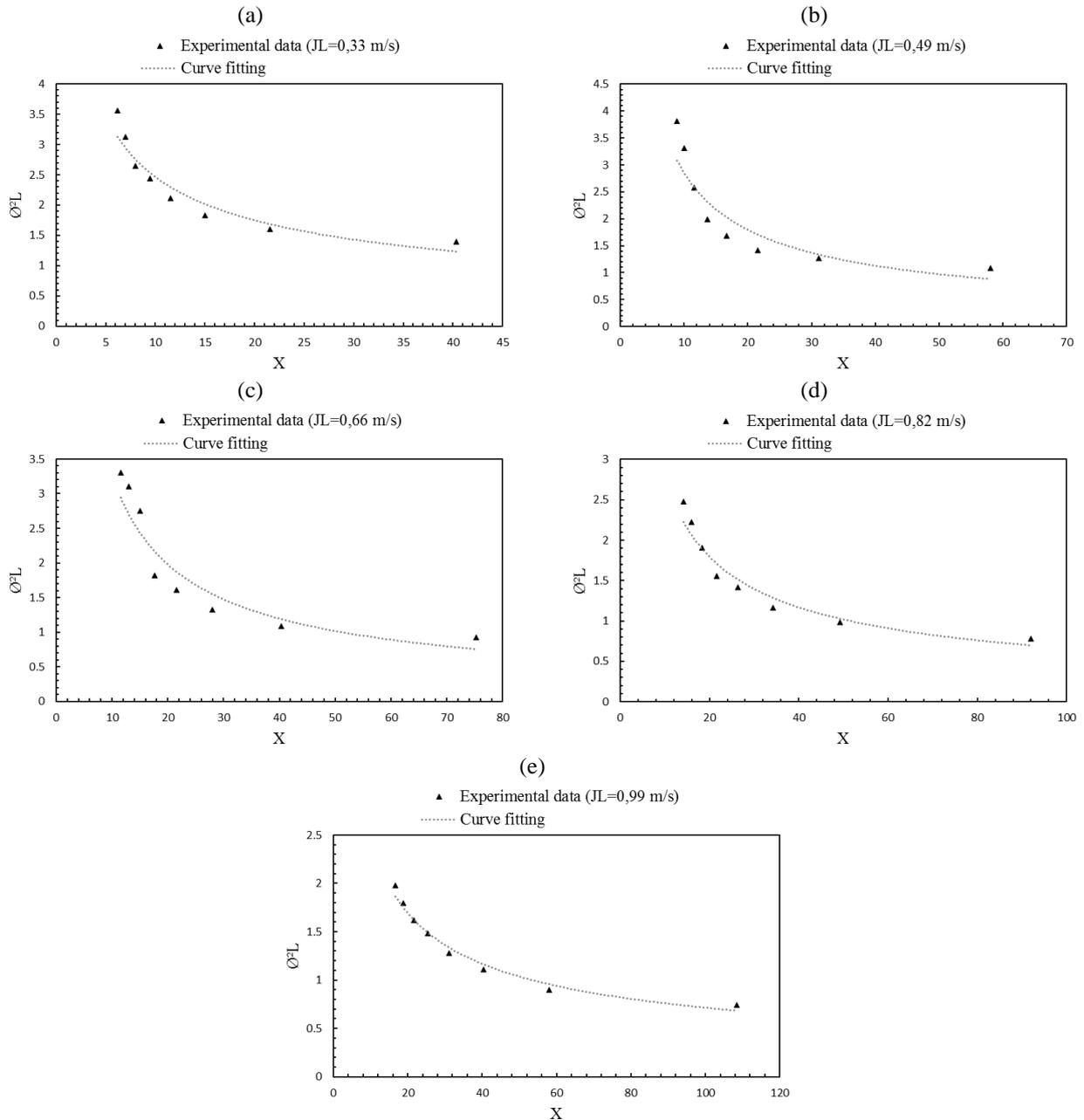
(a)

(b)

(c)

(d)

(e)

Figure 8. Comparison between the parameters obtained from the experiments performed and the one proposed by
Lockhart & Martinelli (1949) for (a) J_L=0.33 m/s, (b) J_L=0.49 m/s, (c) J_L=0.66 m/s,
(d) J_L=0.82 m/s, and (e) J_L=0.99 m/s.

5. CONCLUSION

In this work an experimental study was developed to analyze the behavior of a flow in a biphasic system of liquid and gaseous phases and to obtain an analysis of the loss of charge of the system during the flow through the pressure drop, in order to verify the parameters obtained with other correlations found in the literature.

When analyzing the pressure drops when using different liquid and gas flows, it is possible to observe that the increase of the surface velocity of the gas produces a greater pressure drop and a lower friction factor, because the wall of the acrylic tube has more contact with the than the liquid.

When comparing the experimental data obtained with the Darcy and Reynolds correlations, with theorists based on the Moody Diagram, it is noted that there is a similar tendency according to Fig. 5.

The frictional pressure drop from experiments and from the Lockhart-Martinelli's separated flow model has been compared, indicating that for lower difference between gas and liquid flow rates, giving clues that the model can represent

adequately the experimental data. On the other hand, for great differences between gas and liquid flow rates, the model is very inaccurate.

Another comparison carried out in this work is the relation between $\phi_L{}^2$ and *X*. The results indicate that the model can represent approximately the experiments for moderated values of X (between 20 and 40), and that for lower or higher X-values, the model is inadequate. However, it should be evaluated better approximation functions to represent the experimental data, because the power function utilized in this work has limitations for higher and lower values. Anyway, the experiments shown a difference for *X*-values more than 50, while the model indicates values greater than 100, the experiments indicates values in the order of 0.75.

Broadly, more experiments are needed to be more conclusive about the behavior of the Lockhart-Martinelli's separated flow model.

6. ACKNOWLEDGEMENTS

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