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COMPARISON BETWEEN EXPERIMENTAL PRESSURE DROP AND PREDICTIONS BY DIFFERENT SLUG UNIT CELL APPROACHES FOR HORIZONTAL AIR-SHEAR THINNING FLUID FLOW

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Abstract. *The present work has the main objective of validating a unit cell model against experimental data for total pressure drop available in the literature. For this purpose, a horizontal slug flow with shear-thinning liquid is studied. Additionally, the performance of the model implemented in this work is also evaluated in comparison to other model taken from the literature. Two-phase gas/shear-thinning liquid flows are often encountered in oil and gas industries. Nevertheless, only a few research in the literature have focused on gas/liquid flow in which the liquid has a non-Newtonian behaviour. The pressure drop is one of the main parameters in hydraulic transport, and its evaluation is beneficial for equipment design. The unit cell approach is a mechanistic model that consists of performing mass and momentum balances across an assumed intermittent length. Several numerical simulations were performed for a wide range of superficial velocities and three different carboxymethyl cellulose solutions. The predictions of the mechanistic model were validated against the experimental data taken from the literature. The model predicts the total pressure drop with better overall performance than the literature model .*

Keywords: *Two-Phase Flow, Pressure Drop, Slug Flow, Unit Cell Model, Shear-Thinning Fluid*

1. INTRODUCTION

In oil and gas industry, fluids with shear thinning and yield stress characteristics are often encountered. Mechanistic analysis has been widely used in industrial simulations to predict two-phase flows in pipelines, saving computational and practical time compared to three-dimensional and experimental procedures, respectively. For two-phase gas/Newtonian liquid flow, a variety of numerical and experimental analyses were performed and validated. However, for two-phase air/shear-thinning liquid, only a few authors have performed experiments and numerical analyses.

By performing an empirical flow pattern visualisation, Chhabra and Richardson (1984) modified the flow map of Mandhane *et al.* (1974) for the gas/non-Newtonian case. This was the first description of a flow pattern map involving a non-Newtonian phase.

The mechanistic model of Taitel and Barnea (1990) for steady slug flow was extended to account for a non-Newtonian rheology by Xu *et al.* (2007). Three different CMC-water solutions were tested over wide ranges of inclination angles and pipe diameters. The predicted dimensionless pressure drop fit the experimental data with less than 20% as average error.

Xu *et al.* (2009) performed an experimental and theoretical study of gas/non-Newtonian fluid flow through horizontal pipes to assess the drag reduction by gas injection in a power-law fluid flow in stratified and slug flow patterns. The results indicates a reasonably good agreement for liquid hold-up and dimensionless pressure drop for both flow regimes, when compared with experimental data.

For horizontal and inclined pipes, Xu (2013) studied the slug flow regime of gas/shear-thinning fluids. In this work, a new empirical correlation for liquid slug hold-up was developed as a function of the liquid Reynolds number and inclination angle. The proposed equation is based on measured data consisting of 271 data points taken from the literature. This research was extended by Xu *et al.* (2014) to predict the pressure drop of gas/non-Newtonian power-law fluids in horizontal pipes for slug flow regime. The authors compared unit cell results and predictions given by the empirical models of Dziubinski (1996) and Dziubinski and Chhabra (1989) with the experimental data of Chhabra *et al.* (1983), Chhabra and Richardson (1984), Xu *et al.* (2007), Xu *et al.* (2009), and Ruiz-Viera *et al.* (2006). Xu *et al.* (2014) concluded that these empirical correlations fails to predict the data, presenting large errors. In opposition, the implemented unit cell model presented better predictive ability compared to the experiments, with an absolute average error of 14%.

Flow pattern transition maps for two-phase gas/shear-thinning fluid were built in the work of Picchi and Poesio (2016) for horizontal and slightly inclined pipe. From the steady and fully developed 1D Two-fluid model governing equations, the boundaries of the transition from stratified to slug flow regime were carried out by performing a linear stability analysis.

For stratified, plug and slug flow in horizontal and inclined pipes, Picchi *et al.* (2015) performed experiments and modeling a flow of a gas/power law liquid at three different carboxymethyl cellulose (CMC) solutions as a liquid fluid. For slug flow, a new unit cell model was proposed by extending the model of Orell (2005), and the predictions were compared to the experimental data. The model predicts the experimental pressure drop for downward and horizontal flows with a satisfactory agreement: 78% of the experimental data within $\pm$ 30% error.

In the electronic annex of Picchi *et al.* (2015), the experimental data for total pressure drop in the horizontal slug flow regime is presented. By using the unit cell model proposed by Dukler and Hubbard (1975), we conducted numerical simulations in the LedaFlow software, and the results were compared with the unit cell results and the experimental data given by Picchi *et al.* (2015). This methodology can also be applied in the future to test the implemented unit cell model for more complex non-Newtonian behaviours, such as Herschel-Bulkley fluids.

2. METHODOLOGY

2.1 Data characteristics

The experimental total pressure drop used to validate the unit cell model proposed in this work is obtained from Picchi *et al.* (2015). In this study, the authors used a facility with a 9 m long glass pipe, with an inner diameter of 22.8 mm.

In the present work, we assessed and compared the total pressure drop only for the horizontal case (0^o), although the authors have evaluated other inclinations. The air properties are $\mu_g = 1.8 \cdot 10^{-5}$ Pa.s, and $\rho_g = 1.2$ kg/m³ (dynamic viscosity and density, respectively). This properties were measured at 25^oC and atmospheric pressure.

The superficial velocities ranges from 0.05 up to 1.4 m/s for the liquid phase, and from 0.1 up to 2 m/s for the gas phase. The rheology of the three CMC-water solutions are presented in Tab. 1. These solutions shows a shear-thinning fluid behaviour, which can be described by a power law model. In this model, the shear stress τ is related to the shear rate $\dot{\gamma}$ as $\tau = K\dot{\gamma}^n$, where K and n are the fluid consistency index and the flow behaviour index, respectively.

Table 1: Properties of the CMC concentrations, given by Picchi *et al.* (2015).

	Conc. (%w/w)	ρ_l (kg/m³)	K (Pa.s^n)	n (-)
CMC-1	1	998.0 ± 0.5	0.007 ± 0.001	0.942 ± 0.010
CMC-3	3	999.0 ± 0.5	0.061 ± 0.002	0.875 ± 0.011
CMC-6	6	1002.0 ± 0.5	0.264 ± 0.010	0.757 ± 0.010

(1) Measured at 25 ^{o}C.

2.2 Dukler and Hubbard model

Figure 1 shows the physical model that leads to the mathematical model developed by Dukler and Hubbard (1975) (DH model). It represents the schematic diagram of the unit cell model (UCM) covered in this work, in which the total length (L_U) can be divided in two main regions, called the liquid slug region (L_S), and the separated region (L_F), that includes the liquid film at the bottom and the gas phase at the top. At the bubble tail, a turbulent mixture zone can be found (L_M).

In the slug region, H_{LLS} is the liquid holdup, V_{LLS} and V_{GLS} are the velocities of the liquid and gas phases at this region, respectively. The no slip condition is applied in this region ($V_{LLS} = V_{GLS} = V_S$). In the separated region, H_{LTB} is the liquid holdup, V_{LTB} and V_{GTB} are the liquid and gas velocities, respectively; and V_{LTBe} is the minimum liquid velocity, since the liquid velocity in the separated region is not constant. At the beginning of the Taylor bubble, V_{TB} is the translational velocity.

The liquid slugs fill the whole pipe cross-sectional area, which implies a high velocity for the slug. The liquid film in front of the slug moves slower. As a result, the liquid slug overtakes the slow film in front of it, and increases it velocity to the slug velocity. Thus, a turbulent zone is created in the bubble tail. In addition, the gas bubble passes into the slug region, making the slug to release liquid from its back, which creates the film region.

The total pressure drop across the unit cell ($-\Delta p_U$) is composed by the accelerational pressure drop ($-\Delta p_A$) and the

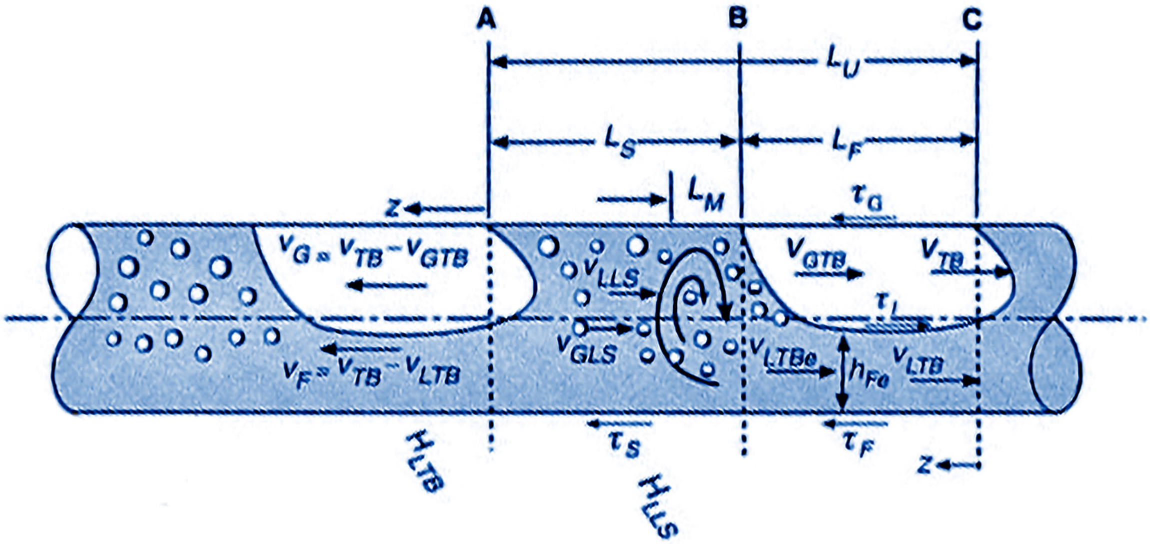

Figure 1: Unit cell model proposed by Dukler and Hubbard (1975).

frictional pressure drop ($-\Delta p_F$). The accelerational component emerges due to the velocity difference between the slug region and the liquid film. Meanwhile, the frictional component is caused by the shear between the liquid phase and the pipe wall. Then, the total pressure drop is given by:

$$-\frac{dp}{dL} = -\frac{\Delta p_U}{L_U} \tag{1}$$

The accelerational component of the pressure drop is written as follows:

$$-\Delta p_A = \frac{x}{A_p}(V_S - V_{LTBe}) \tag{2}$$

where x is the rate in which mass is gathered from the film zone to the slug region, the part $x(V_S - V_{LTBe})$ represents the force acting on the gathered mass by the slug region, and A_P is the pipe cross sectional area.

The frictional pressure component can be calculated according to the average slug properties, as shown in Eq. (3).

$$-\Delta p_F = \frac{2 f_S \rho_S V_S^2 L_S}{d} \tag{3}$$

where ρ_S is the density in the slug, calculated on the basis of the holdup in the slug, and f_S is the Fanning friction factor, that can be determined by the Blasius equation (Eq. (4)),

$$f_S = C_S Re_S^{-m_S} \tag{4}$$

The density in the slug is calculated by the following equation:

$$\rho_S = \rho_L H_{LLS} + \rho_G (1 - H_{LLS}) \tag{5}$$

where C_S and m_S are constants that depends on the flow regime, and Re_S is the Reynolds number for a power law fluid ($\tau = K\dot{\gamma}^n$) considering mixture conditions, yielding:

$$Re_S = \frac{d^n V_S^{2-n} \rho_S}{K 8^{n-1} \left(\frac{1+3n}{4n}\right)^n} \tag{6}$$

As proposed by Taitel and Barnea (1990), the slug zone velocity V_S is supposed to be the same as the mixture velocity V_M, as follows:

$$V_S = V_M = V_{GS} + V_{LS} \tag{7}$$

where V_{GS} and V_{LS} are the gas and liquid superficial velocities, respectively.

By applying a mass balance on a control volume limited by the Taylor bubble tail (section B in Fig. 1) and an arbitrary section in the slug region, the liquid film moves backward of the section B, and the liquid slug moves backward of the arbitrary section. This mechanism allows to perform a continuity analysis for the liquid phase, as mathematically defined by:

$$x = (V_{TB} - V_{LTBe})\rho_L A_P H_{LTBe} = (V_{TB} - V_S)\rho_L A_P H_{LLS} \tag{8}$$

By making an algebraic manipulation of the right side of the above equation, the translational velocity can be found:

$$V_{TB} = V_S + \frac{x}{\rho_L A_P H_{LLS}} \tag{9}$$

The translational velocity is the sum of the slug velocity (or mixture velocity as explained) and the additional velocity gained by the gathered process. Defining c as in Eq. (10), yields to the bubble velocity in Eq. (11).

$$c = \frac{x}{\rho_L A_P H_{LLS} V_S} \tag{10}$$

$$V_{TB} = V_S + cV_S = (1 + c)V_S = C_0 V_M \tag{11}$$

Note that the Eq. (11) is a deduction proposed in the original paper. Several investigations, such as Dukler and Hubbard (1975), did not consider the drift velocity in horizontal flow, arguing that gravity can not affect the flow in horizontal pipes. However, this hypothesis is flawed because of the hydrostatic difference between the nose of the bubble and the liquid film. For translational velocity calculation, the Nicklin (1962) equation is the currently most widely used relation:

$$V_{TB} = C_0 V_M + u_d \tag{12}$$

where C_0 is the distribution parameter and u_d is the drift velocity. Bendiksen (1984) obtained experimentally this constants over several geometric configurations, superficial velocities, and fluids properties. LedaFlow utilises the Dukler and Hubbard (1975) model and the experiments conducted by Bendiksen (1984) to perform the unit cell approach (Kjølaas *et al.* (2013)).

The cross-sectional area of the liquid film decreases with the z-direction, as can be seen in Fig. 1. As a consequence, the relative velocity increases. Therefore, the film and slug velocities can be related to a specific location in the slug unit cell, in other words, to the z-coordinate. The liquid film flows backward at a constant mass flow rate. It is required to obtain the liquid film velocity profile $V_F(z)$, the holdup profile $H_{LTB}(z)$, and the liquid height $h_F(z)$. The velocity of the liquid film is:

$$V_F = V_{TB} - V_{LTB} = \frac{x}{\rho_L A_p H_{LTB}} \tag{13}$$

At an arbitrary control volume in the separated region, with a width of dz, a mass balance must be applied. The resultant equation consists of a first order ordinary differential equation, that should be numerically solved.

$$\frac{dh_F}{dz} = \frac{-\tau_F S_F}{\frac{x^2 H'_{LTB}}{\rho_L A_P H_{LTB}^2} - A_P H_{LTB} \rho_L g} \tag{14}$$

where $H'_{LTB} = dH_{LTB}/dh_F$, S_F is the wetted perimeter, and τ_F is the shear stress, given by:

$$\tau_F = \frac{f_F}{2}\rho_L V_{LTB}^2 = \frac{f_F}{2}\rho_L (V_{TB} - V_F)^2 \tag{15}$$

The film friction factor f_F is calculated based on the hydraulic diameter. The length of the slug film is defined as:

$$L_F = L_U - L_S = \frac{V_{TB}}{\Omega_S} - L_S \tag{16}$$

where Ω_S is the slug frequency. The liquid slug length is determined by means of a mass balance on the liquid, yielding:

$$W_L = \left(V_M \rho_L A_P H_{LLS} T_S + \int_0^{T_F} V_{LTBe} \rho_L A_P H_{LTBe} dt \right) \frac{1}{T_U} \tag{17}$$

The equation above consists of an integration in time and must be transformed into the space domain, by changing variables:

$$T_S = \frac{L_S}{V_{TB}} \qquad T_F = \frac{L_F}{V_{TB}} \qquad T_U = \frac{1}{\Omega_S} \qquad d_t = \frac{d_L}{V_{TB}} \tag{18}$$

An expression for $V_{LTBe} H_{LTBe}$ can be obtained from Eq. (8), combining it with the Eq. (12). Thus, Eq. (17) becomes:

$$\frac{W_L}{\rho_L A_P} = \frac{\Omega_S}{V_{TB}} \left\{ V_M H_{LLS} L_S + L_F \left[H_{LTBe}(C_0 V_M + u_d) + H_{LLS}(V_M(1 - C_0) - u_d) \right] \right\} \tag{19}$$

Equation (19) must be solved for the slug length L_S. By means of an initial guess for L_S, Eq. (16) should be solved in order to obtain the L_F value. After that, Eq. (14) must be numerically integrated. Finally, L_S will be obtained from the results of the previous step, by making it explicit in Eq. (19). This iterative procedure is repeated until convergence is achieved.

2.3 Picchi model

The Orell model consists of a reformulation of the Taitel and Barnea (1990) submodel for horizontal gas/Newtonian liquid slug flow. Picchi *et al.* (2015) extended the Orell model to inclined flows, also considering the rheology of the power law fluid. The Picchi model, such as the Dukler and Hubbard model, is drawn up by performing a mass and momentum balance over the slug unit cell. However, in the film zone (L_F), the assumption of uniform film thickness is applied. As a consequence, the gas-liquid flow in the film region is treated as stratified smooth flow.

In summary, the liquid height in the separated region is a constant value, and it should be calculated by geometrical relations at the pipe cross section (see Orell (2005)). On the other hand, the Dukler and Hubbard model requires the calculation of a differential equation.

A further important difference between the models relies on the required closure correlations. Dukler and Hubbard model used in this work requires three closure relations: the liquid holdup in the slug region (H_{LLS}), the slug frequency (Ω_S), and the translational velocity (V_{TB}). Note that the slug length can also be provided, instead of the slug frequency. As for the Picchi model, two correlations are required: one for the liquid holdup in the slug body (H_{LLS}), and another one for the translational velocity (V_{TB}).

In addition, the velocities of the liquid and gas phases in the slug body are not the same in the Picchi model. Thus, the mass balance over the slug unit cell can be written in terms of the superficial velocities of the liquid and the gas phase, given respectively by:

$$V_{LS} = V_S H_{LLS} \frac{L_S}{L_U} + V_F H_{LTB} \frac{L_F}{L_U} \tag{20}$$

$$V_{GS} = V_S (1 - H_{LLS}) \frac{L_S}{L_U} + V_G (1 - H_{LTB}) \frac{L_F}{L_U} \tag{21}$$

where V_F is given by Eq. (13), and V_G is the gas velocity in the film zone, expressed by $V_G = V_{TB} - V_{GTB}$.

A liquid mass balance relative to a coordinate system that travels at the translational velocity of the slug unit cell yields:

$$(V_{TB} - V_F)H_{LTB} = (V_{TB} - V_S)H_{LLS} \tag{22}$$

In the view of the assumption of uniform film thickness, the momentum balance in the film zone can be written as for stratified flow. By eliminating the pressure gradient, the momentum balance leads to:

$$-\tau_L \frac{S_L}{A_L} + \tau_i S_i \left(\frac{1}{A_G} + \frac{1}{A_L}\right) + \tau_G \frac{S_G}{A_G} = 0 \tag{23}$$

where $A_{G,L}$, $S_{G,L}$, and $\tau_{G,L}$ are the cross section, the wetted perimeter, and the wall shear stress of the gas and liquid. τ_i is the interfacial shear stress, and S_i is the interfacial perimeter.

Equations (20), (21), (22), and (23) need to be simultaneously solved. Therefore, the total pressure drop, considering horizontal flow, can be calculated by:

$$-\frac{dp}{dL} = \frac{2 f_S \rho_S V_S^2}{d} \frac{L_S}{L_U} + \frac{4}{\pi d^2}(\tau_L S_L + \tau_G S_G)\frac{L_F}{L_U} \tag{24}$$

where ρ_S is given by Eq. (5), and f_S is given by (4).

3. RESULTS AND DISCUSSION

In order to evaluate and compare the two unit cell models with the experimental results, the average relative error (e_r) is defined:

$$e_{rPM} = \frac{1}{N}\sum_{i=1}^{N}\sqrt{\left(\frac{dp/dL_{PM_i} - dp/dL_{exp_i}}{dp/dL_{exp_i}}\right)^2} \tag{25}$$

$$e_{rDHM} = \frac{1}{N}\sum_{i=1}^{N}\sqrt{\left(\frac{dp/dL_{DHM_i} - dp/dL_{exp_i}}{dp/dL_{exp_i}}\right)^2} \tag{26}$$

where the subscripts PM_i and DHM_i indicates the Picchi model and the Dukler and Hubbard model, respectively. dp/dL_{PM_i}, dp/dL_{DHM_i}, dp/dL_{exp_i}, and N are the pressure drop of the Picchi's model, the pressure drop of the Dukler and Hubbard's model, the experimental pressure drop, and the number of experiments, respectively.

In a recent study, Baungartner *et al.* (2023) conducted an experimental analysis for horizontal gas/power law liquid slug flow for three different carboxymethyl cellulose solutions: 0.05, 0.1, and 0.2 (%w/w). They also compared the same two unit cells models approaches assessed in this work, except for the fact that they used the original concept for the translational velocity in the Dukler and Hubbard (1975) model (Eq. (11)). In their research, different combinations of closure terms were tested, isolating the effects of each one. The authors concluded that the Dukler and Hubbard (1975) model clearly outperformed the extended Orell power law model for pressure drop predictions, when compared with experiments. Despite that, they did not carried out an extensive discussion about the differences between the models results.

Figure 2 shows the results for pressure drop as a function of the mixture velocity, on the left, and the results for experimental pressure drop comparing the models predictions, on the right.

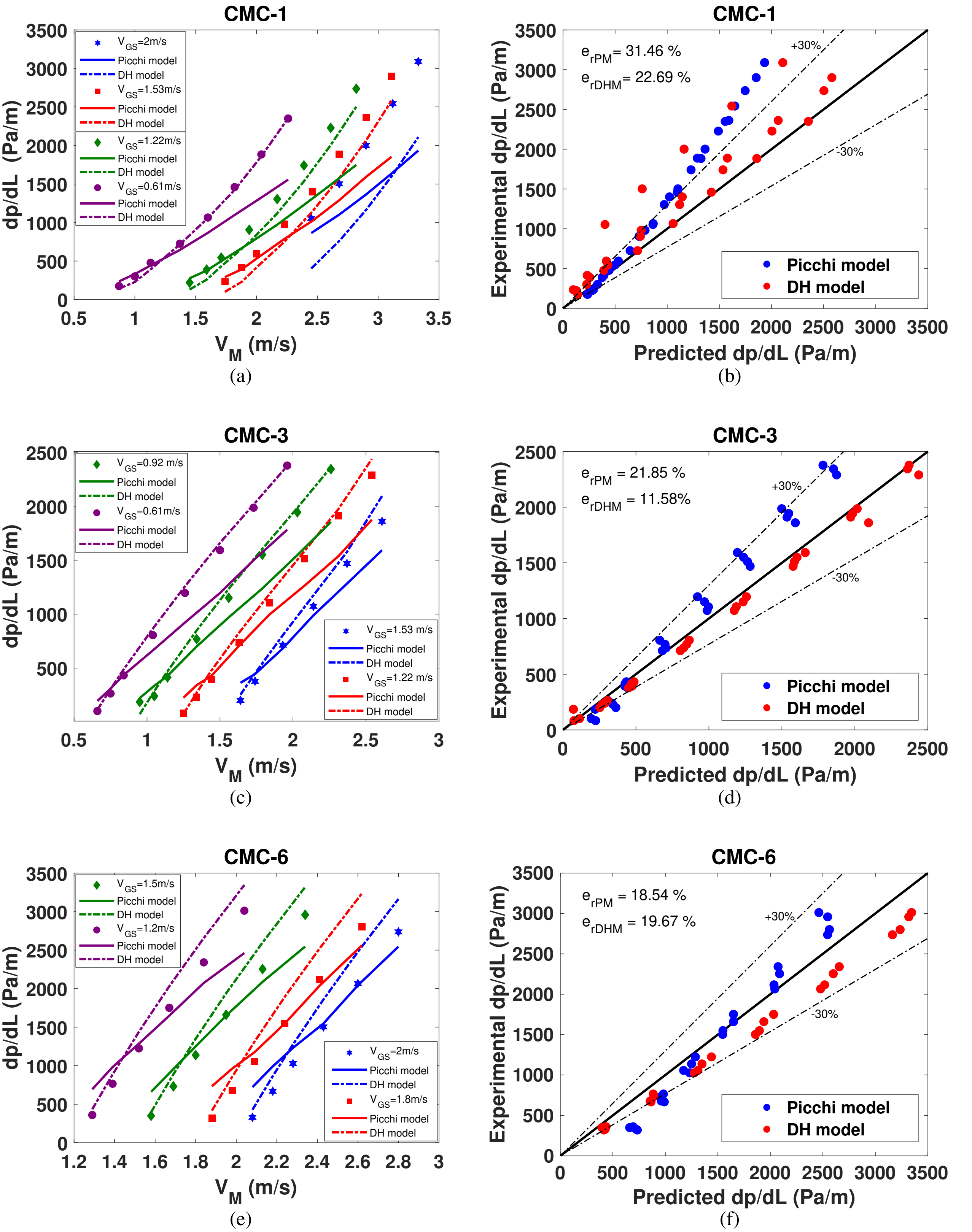

Figure 2: Experimental pressure drop of the Picchi *et al.* (2015) work for three CMC solutions compared to the prediction of the Dukler and Hubbard (1975) model and of the Picchi *et al.* (2015) model. Dashed lines represent $\pm$ 30%. The left side corresponds to the predictions as a functions of the mixture velocity, and the right side represents the models predictions against the experimental data.

For the CMC-1 solution, both models presented a similar behaviour when the gas superficial velocity increases. From the lowest gas superficial velocity (V_{GS}) to the highest one, the difference between models predictions and experiments is increased. The mean error value of the DH model at V_{GS}=$0.61m/s$ is around 8.28%. This value gradually increases when the gas superficial velocity increases, and it hits a mean error of 44.21% at V_{GS}=$2m/s$. These points are shown in Fig. 2b, and consequently they are the farthest points from the centerline. As for the Picchi model, at V_{GS}=$0.61m/s$ the mean error is about 25.6%, and it is 43.79% at V_{GS}=$2m/s$. The Picchi model predicts 48% of the experimental data within $\pm$ 30% error, while the DH model predicts 69%.

The results for pressure drop predictions by the DH model are very encouraging for the CMC-3 solution. At V_{GS}=$0.61m/s$, the best performance was reached. The mean error for these operational condition was around 7.46%. Note that this value is very close to the error of CMC-1 at the same superficial velocity. In contrast, the better performance for the Picchi model was achieved at V_{GS}=$1.53m/s$, with an error of 17.33%, and the worse performance of 26.67% at the lowest superficial velocity. The Picchi model predicts 77% of the experimental data within $\pm$ 30% error, while the DH model predicts 97%. It is worth mentioning that 80% of the data was within $\pm$ 15% error in the DH model.

A reasonably good performance in representing the experimental data was noted for the DH model in CMC-6, which predicts 96% of the data within $\pm$ 30% as error. The mean errors gradually increases with the increasing of the gas superficial velocity. These values ranges from 15.70 to 23.75%. The performance of the Picchi model for pressure drop predictions was practically the same for all gas superficial velocities, since the lowest mean error is about 17.38%, at V_{GS}=$1.5m/s$, and the highest one was 19.39%, at V_{GS}=$1.2m/s$. Picchi model was able to fit exactly 50% of the total data within $\pm$ 10% as error. This model predicts 79% of the data within $\pm$ 30%.

A very distinct behaviour between the two models can be stated, with regard to the performance of both in relation to the increase in the gas superficial velocity. The performance of the DH model for the three solutions decreases by increasing the gas superficial velocity. On the other hand, each one of the three solutions of the Picchi model showed contrasting features by increasing the gas superficial velocity. CMC-1 presented the same characteristic described above by the DH model. In opposition, CMC-3 showed the best performance at the highest gas superficial velocity, and the worse performance at the lowest one. Lastly, CMC-6 displayed practically the same mean error for all gas velocities.

By rising the CMC concentration, the power law consistency index (K) increases, and the fluid behaviour index (n) decreases. Assessing the same mixture velocity for the three concentrations, the pressure drop increases when n decreases, this was also noted by Baungartner *et al.* (2023). Although the work of Baungartner *et al.* (2023) addressed other concentrations, pipe geometry and operational conditions, the overall poor results for pressure drop predictions by the DH model relies on the lower concentration, as also observed in this work. Nevertheless, the overall poor predictions given by the extended Orell (2005) model was found for the middle concentration in the work of Baungartner *et al.* (2023). This conclusion can not be stated for this work, once the worse predictions relies on the CMC-1 concentration.

4. CONCLUSION

In this work, experimental pressure drop taken from the work of Picchi *et al.* (2015) for three carboxymethyl cellulose solutions were compared with the model of Dukler and Hubbard (1975) and the model of Picchi *et al.* (2015), for a wide range of operational conditions.

A robust mathematical description of the Dukler and Hubbard (1975) was made.

Different prediction behaviours for each model was noted as the gas superficial velocity increases. For the Dukler and Hubbard (1975) model, the better fitting occurred at the CMC-3 concentration, with an average error of 11.58%, followed by the CMC-6 solution (19.67%) and by the CMC-1 solution (22.69%). As for the Picchi *et al.* (2015) model, a better agreement between experimental and predicted values was achieved for the CMC-6 solution (error about 18.54%), followed by the CMC-3 solution (21.85%), and by the CMC-1 concentration (31.46%). For both models, the worse performance relies on the CMC-1 solution. At the CMC-3 solution, the Dukler and Hubbard (1975) model achieved a noticeable performance fitting 80% of the data within $\pm$ 15% of error, while the Picchi *et al.* (2015) model was able to fit 50% of the data within $\pm$ 10% as error, at the CMC-6 solution.

The mechanistic models assessed in this paper are based on physical insights and empirical correlations to describe a particular phenomenon. The Picchi *et al.* (2015) model requires two closure relations and has a lower mathematical robustness in comparison with the Dukler and Hubbard (1975) model, which in turn requires three closure correlations. Overall, Dukler and Hubbard (1975) model was more suitable to fit the experimental data for pressure drop. Therefore, with the results presented, it is concluded that a complex phenomenon such as a two-phase flow involving a gas and a shear-thinning liquid requires a more sophisticated mathematical description.

5. ACKNOWLEDGEMENTS

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