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BRAZILIAN JOURNAL OF PETROLEUM AND GAS
Journal Homepage: www.portalabpg.org.br/bjpg

A publication of the
ABPG
BRAZILIAN ASSOCIATION OF R&D IN
PETROLEUM AND GAS

STUDY ON PRESSURE DROP AND LIQUID VOLUME FRACTION OF THE OIL-GAS FLOW IN A VERTICAL PIPE USING CFX AND THE BEGGS AND BRILL CORRELATION: VISCOSITY EFFECTS

[a] Silva, L. D. S.[1]; [b] Marinho, J. L. G.

[a] School of Chemical Engineering, Federal University of Alagoas - UFAL, Maceió – AL, Brazil
[b] School of Petroleum Engineering, Federal University of Alagoas - UFAL, Maceió – AL, Brazil

Received: 21.09.2015 / Revised: 02.12.2015 / Accepted: 07.12.2015 / Published on line: 06.04.2016

ABSTRACT

Two-phase liquid-gas flows are common in several industrial processes. Since oil and gas are simultaneously produced in most petroleum reservoirs, the two-phase flow occurs in petroleum transport as well. Depending on the petroleum viscosity, a large amount of energy is needed to move the oil-gas mixture, resulting in significant expenses. The present work aims to investigate the influence of oil viscosity on pressure drop and liquid volume fraction of the upward two-phase flow of different types of oil in a vertical pipe. This study was accomplished using Computational Fluid Dynamic techniques, and Beggs and Brill correlation. The numerical simulations were performed using the application Ansys CFX 13.0, in which governing equations were solved utilizing the finite volume method. The results of pressure drop and liquid volume fraction obtained by both methods were analyzed and discussed. The numerical results for the pressure drop show that the CFX value was approximately 24% lower than that predicted by the Beggs and Brill correlation in the worst case. The liquid volume fraction decreased along the pipe length due to the viscosity effects of the oil.

KEYWORDS

CFX; Beggs and Brill correlation; pressure drop; viscosity

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[1] To whom all correspondence should be addressed.
Address: Unidade Acadêmica Engenharia Química, Centro de Tecnologia, UFAL, Av. Lourival Melo Mota, s/n Tabuleiro, Maceió – AL – Brazil.
ZIP Code: 57072-970 | Telephone: +55 82 99600-3504 |e-mail: lucas.ds25@gmail.com

doi:10.5419/bjpg2016-0001

1. INTRODUCTION

Two-phase liquid-gas flows often occur in several industrial processes. In general, the petroleum flow also involves two phases, since most petroleum reservoirs simultaneously produce oil and gas **(Albuquerque et al., 2007; Bangtang et al., 2014; Ozbayoglu et al., 2012)**.

Oils produced in different reservoirs have different characteristics such as density, and viscosity **(Thomas et al., 2001)**. Depending on the oil viscosity, the expenses with petroleum transport may be affected adversely. For instance, heavy oil flow leads to a large pressure drop due to high oil viscosity, resulting in a significant amount of energy required to pump the fluid and an increase in operating costs **(Andrade et al., 2013; Dunia et al., 2011; Hasan et al., 2010; Gadelha et al., 2013; Martínez-Palou et al., 2011)**. In addition, the extraction of petroleum in deep waters, where the water temperature is approximately 5 °C, is still challenging since the oil viscosity is a function of the temperature **(Albuquerque et al., 2007; Marinho, 2012)**. As a result, studies on pressure drop prediction for petroleum flow have been important from an economic standpoint. However, because the petroleum flow is usually characterized as a two-phase flow, the presence of a second phase makes the calculation utilized to predict the pressure drop more complex **(Chang et al., 2008; García et al., 2007; Jahanandish et al., 2011)**.

Many authors have sought to understand the two-phase flow phenomena by applying numerical techniques such as Computational Fluid Dynamics (CFD), which uses a mathematical modeling based on the conservation of mass, momentum, and energy. Because their results are of great reliability, and can be obtained quickly and with lower cost, the numerical methods are usually more advantageous over analytical and experimental methods **(Silva et al., 2014; Souza et al., 2011).**

Nevertheless, many empirical correlations have been utilized to predict pressure gradient because of their usability. The Beggs and Brill correlation is one of the many empirical models cited by several authors as a classic correlation, for it is employed in any pipe inclination and flow pattern. Since the correlation is based on the flow pattern that the pipe would present if it were horizontal, corrections are made in order to consider the actual pipe inclination. Because of its reliable predictions, this correlation has been widely used in the oil and gas industry **(Beggs and Brill, 1991; Sarah et al., 2014; Souza et al., 2010; Zhao et al., 2013)**.

The present work aims to investigate the influence of oil viscosity on pressure drop and liquid volume fraction of the upward two-phase flow of oil and gas in a vertical pipe. Different oils were utilized in this study, which was performed by using Computational Fluid Dynamics techniques and the Beggs and Brill correlation. The numerical simulations were performed employing the application Ansys CFX 13.0. The governing equations were solved by the finite volume method. The results obtained using both methods were compared and analyzed.

2. METHODS

To solve the problem proposed, a 10-m segment length of a representative vertical pipe used in offshore petroleum production was utilized (Figure 1(a)). The pipe had an inner diameter of 0.18 m. For the simulations in Ansys CFX, a

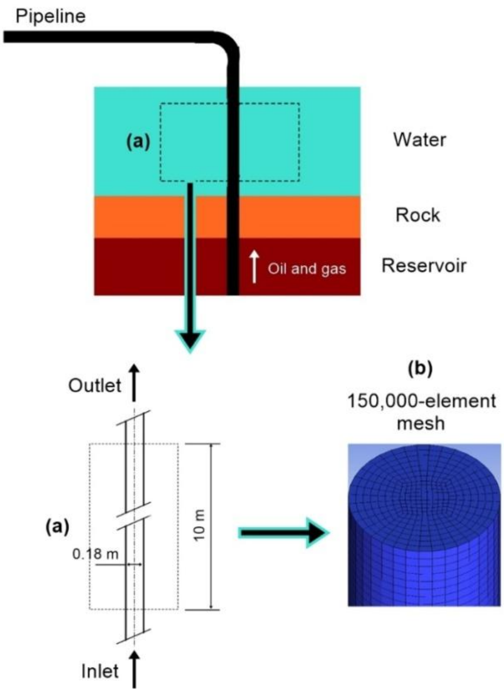

Figure 1. Representative schematics of the domain in study (a) and view of the pipe outlet after the mesh generation (b).

numerical mesh with 150,000 elements was generated using the application ICEM CFD. Figure 1(b) shows the view of the pipe outlet of the mesh utilized in all the simulations.

2.1 CFD mathematical modeling

For the simulations in CFX, the following assumptions were considered: isothermal and incompressible flow, smooth pipe, gravitational effect, tridimensional domain, constant physical-chemical properties, and no chemical reactions. To simplify the problem proposed, tridimensional instabilities that continually occur in the two-phase flow were neglected so that a steady-state condition was assumed. A dispersed flow using an Eulerian-Eulerian model was adopted, in which the gas was considered as dispersed particles with diameter of 2.0×10^{-3}m. According to **Marinho (2012)** and **Silva et al. (2014)**, based on these assumptions, the main governing equations that describe the problem in study are Equations 1 and 2, which represent the mass and the momentum conservation equations, respectively:

$$\nabla\cdot\left(f_\alpha\rho_\alpha\vec{U}_\alpha\right)=0 \qquad (1)$$

$$\begin{aligned}\nabla\cdot\left[f_\alpha\left(\rho_\alpha\vec{U}_\alpha\otimes\vec{U}_\alpha\right)\right]&=-f_\alpha\nabla p\\&+\nabla\cdot\left[f_\alpha\left(\tau_\alpha+\tau_\alpha^{turb}\right)\right]+\vec{S}_{MS\alpha}+\vec{M}_\alpha\end{aligned} \qquad (2)$$

The momentum balance on the right side of the Equation 2 is a function of pressure gradient, shear stress, external forces, and interfacial forces that act on α phase. The interfacial force $\vec{M}_\alpha$ was assumed to be a function of the drag coefficient, the mixture density, and the interfacial area per unit volume, as shown in Equation 3:

$$\vec{M}_\alpha=C_D\rho_{\alpha\beta}A_{\alpha\beta}\left|\vec{U}_\beta-\vec{U}_\alpha\right|\left(\vec{U}_\beta-\vec{U}_\alpha\right) \qquad (3)$$

The drag coefficient C_D = 0.44 was adopted in all numerical simulations presented here. The κ-ε turbulence model was utilized in turbulent flow cases. Equations 4 and 5 describe the κ-ε turbulence model:

$$\begin{aligned}&\nabla\left(\rho\vec{U}\kappa\right)=\\&=\nabla\left[\left(\mu+\frac{\mu_t}{\sigma_\kappa}\right)\nabla\kappa\right]+\left(P_\kappa-\rho\varepsilon\right)\end{aligned} \qquad (4)$$

$$\begin{aligned}\nabla\left(\rho\vec{U}\varepsilon\right)&=\nabla\left[\left(\mu+\frac{\mu_t}{\sigma_\varepsilon}\right)\nabla\varepsilon\right]+\\&+\frac{\varepsilon}{\kappa}\left(C_{\varepsilon1}P_\kappa-C_{\varepsilon2}\rho\varepsilon\right)\end{aligned} \qquad (5)$$

A root mean square (RMS) equal to 10^{-6} was utilized as a convergence criterion. Details regarding the mathematical modeling can be found in **Ansys CFX 13 (2011)** and **Marinho (2012)**.

2.2 The Beggs and Brill correlation

The Beggs and Brill correlation is utilized to predict pressure gradient, and is based on the flow pattern that the pipe would present if it were horizontal. The pressure gradient calculation initially involves obtaining the horizontal flow regime, which depends on the points of the Froude number, and input liquid content in the flow pattern map of the correlation. Then, the horizontal liquid hold-up is calculated and corrected to consider the actual pipe inclination. The corrected liquid hold-up, or simply liquid hold-up, is the liquid volume fraction referred to as the total segment volume.

The next step of the correlation involves the two-phase friction factor calculation. The no-slip friction factor is obtained from the curve on the Moody diagram using the corresponding Reynolds number of the flow. With the no-slip friction factor and the corrected liquid hold-up values, the two-phase friction is calculated. The frictional pressure gradient is a function of the two-phase friction factor f_{TP}, as shown in Equation 6:

$$\left(\frac{dp}{dz}\right)_f=\frac{f_{TP}\rho_n v_m^{\,2}}{2d} \qquad (6)$$

where ρ_n is the mixture density weighed by the input liquid content, v_m is the mixture velocity, and d is the pipe diameter. The gravitational pressure gradient is calculated as a function of the mixture density weighed by the corrected hold-up ρ_s, and the pipe inclination from the horizontal θ (Equation 7). The acceleration pressure gradient was also included for better accuracy, even though this term is not significant, except for high velocity flow (Equation 8):

$$\left(\frac{dp}{dz}\right)_{el} = \rho_s g \sin\theta \qquad (7)$$

$$\left(\frac{dp}{dz}\right)_{acc} = E_K \frac{dp}{dz} \qquad (8)$$

The acceleration term E_K is a function of the mixture velocity v_m and the superficial velocity of the gas phase v_{SG} (Equation 9). The total pressure gradient, which is composed of the frictional, gravitational, and the acceleration parcels, is shown in Equation 10:

$$E_K = \frac{\rho_S v_m v_{SG}}{p} \qquad (9)$$

$$\frac{dp}{dz} = \frac{\left(\frac{dp}{dz}\right)_f + \left(\frac{dp}{dz}\right)_{el}}{1 - E_K} \qquad (10)$$

The differential equation shown in Equation 10 was solved for the pressure p by using second-order Runge-Kutta method and the boundary conditions shown in the section 2.3. Full details about the correlation can be found in **Beggs and Brill (1991)**.

2.3 Boundary conditions

Table 1 shows the boundary conditions to the cases presented in this paper. Since the no-slip condition was adopted, the relative velocity between the fluids and the wall was 0 m/s. The average gas volume fraction at the inlet was 5%. Although the gas pressure inside the pipe usually reaches large values, the gas was assumed to be under atmospheric conditions since the domain in study is near the surface.

Table 1. Boundary conditions.

	T (K)	v_S (m/s)	p (Pa)
Inlet	298.15	$v_{SL} = 0.95$ $v_{SG} = 0.05$	—
Wall		$v_{SL} = v_{SG} = 0$	—
Outlet		—	101,325

Table 2. Physical properties of the fluids.

	μ (N.s/m²)	ρ (kg/m³)	Source
Gas	0.7236	1.08×10^{-5}	**Silva et al., 2014**
Oil #1	0.021	855	**Silva et al., 2014**
Oil #2	0.5	951	**Barbosa et al., 2012**
Oil #3	3.184	963.6	**Souza et al., 2011**
Oil #4	42.44	989	**Santos et al., 2010**

2.4 Physical properties of the fluids and cases evaluated

Table 2 shows the physical properties of the fluids. The data shown in Table 2 was arranged so that the least viscous fluids are at the top.

The cases evaluated consist of two-phase flow of each of the oils with the same gas, totaling four cases:

- Case 1: gas and oil #1;
- Case 2: gas and oil #2;
- Case 3: gas and oil #3;
- Case 4: gas and oil #4.

The results for pressure drop and liquid volume fraction obtained for each case were analyzed to investigate the influence of oil viscosity. The results obtained by both methods presented were compared as well.

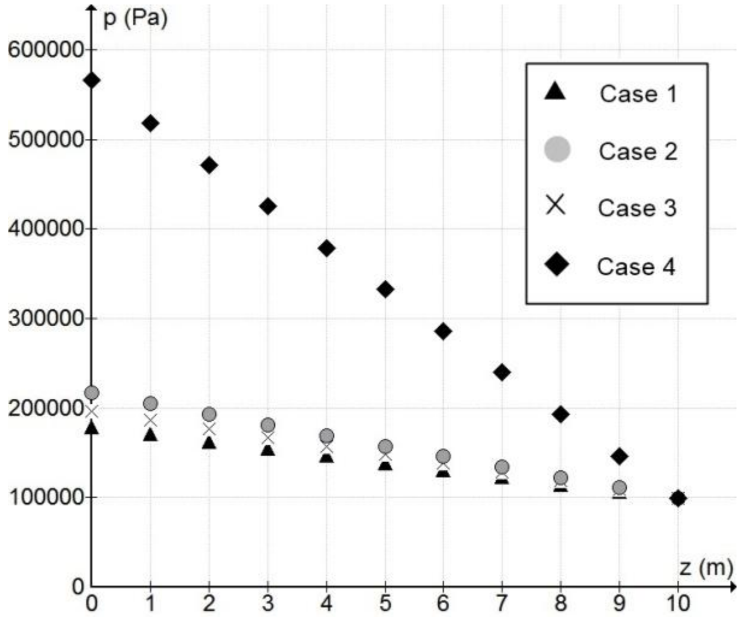

Figure 2. Pressure distribution along the pipe length obtained by using CFX.

3. RESULTS AND DISCUSSIONS

The results for the pressure distribution obtained with the CFX Postprocessor are shown in Figure 2. The pressure drop for all cases presented a linear behavior along the pipe length. Case 1 presented the lowest value of pressure drop along the segment. This was due to the low viscosity and density of the oil utilized, which required less pumping energy. On the other hand, case 4 presented the largest value of pressure drop because of the high viscosity and density of the fluid utilized.

Table 3 shows the results for the pressure drop obtained using CFX and the Beggs and Brill correlation. As demonstrated in Table 3, both methods presented increasing values of pressure drop (Δ*p*) from case 1 through case 4, in which the least and the most viscous oil were utilized, respectively. Table 3 also shows the relative error, in absolute value, between the results for pressure drop obtained by both methods, adopting the Beggs and Brill values as reference. The relative error decreased as the pressure drop increased. This was true for cases 1 through case 3, which presented the lowest relative error. The highest relative error obtained for case 4 can be attributed to the fact that the Beggs and Brill correlation may provide overestimated values of liquid hold-up, thus increasing the pressure drop.

The results for the liquid volume fraction along the pipe length obtained with Ansys CFX are shown in Figure 3. At the pipe inlet (*z* = 0 m), the oil volume fraction is approximately 0.96, which is close to the CFX set value of the average liquid volume fraction at this position. As illustrated in Figure 3 for all cases, as the mixture of oil and gas flows through the pipe, the liquid volume fraction decreases. Since the gas has a lower viscosity and density than the oil, the former flows more rapidly than the latter. As a result, the gas phase

Table 3. Results for the pressure drop predicted by CFX and the Beggs and Brill correlation.

	CFX *Δp* (Pa)	Beggs & Brill *Δp* (Pa)	Relative error (%)
Case 1	79,760	70,542	13.1
Case 2	93,290	84,520	10.4
Case 3	117,600	119,159	1.3
Case 4	466,000	610,386	23.7

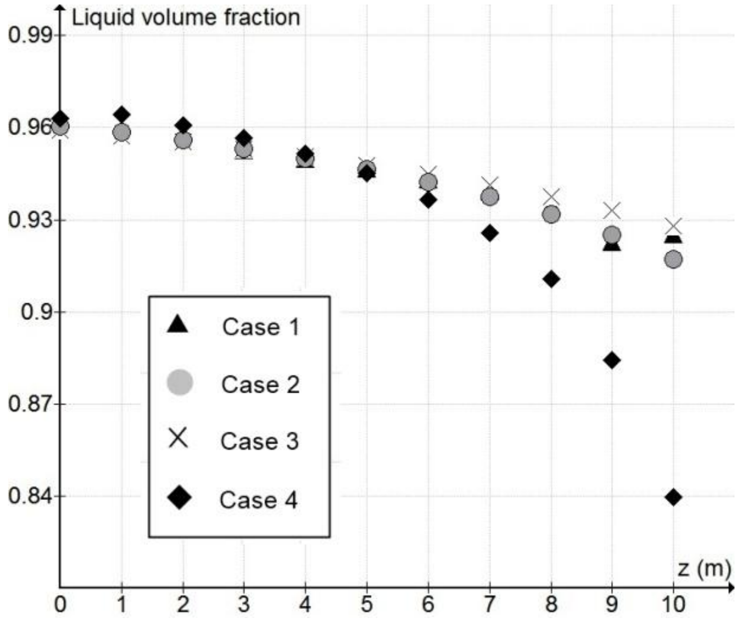

Figure 3. Liquid volume fraction along the pipe length obtained from the numerical simulations.

accumulates at the pipe outlet, thus, decreasing the liquid volume fraction along the pipe length. Figure 3 also shows that case 4 presented the higher drop in liquid volume fraction along the segment, in which the oil volume fraction is approximately 0.84 at *z* = 10 m.

4. CONCLUSIONS

Based on the results, the numerical simulations can be utilized properly to obtain the pressure drop for some of the cases presented in this study. The simulations for cases 1 through 3 predict a pressure drop with up to 14 percent relative error when compared to the results obtained from the correlation. However, the simulations for case 4 presented a relative error of approximately 24 %, which corresponds to a pressure drop 150 kPa lower than the results predicted by the correlation. This difference is significant and cannot be ignored in petroleum and gas production projects, which usually involve larger values of pressure drop.

The results show the influence of oil viscosity on the pressure drop. Case 1 used low viscosity oil and presented the lowest pressure drop value for both methods. The pressure drop increased as the oil viscosity increased from cases 1 through 4. Case 4 used high viscosity oil and presented the greatest pressure drop value.

The oil viscosity also affected the liquid volume fraction along the segment. It happened because the gas is less dense and viscous than the oil, causing the gas to flow more rapidly than the liquid phase. As a result, the gas accumulates at the pipe outlet, thus decreasing the liquid volume fraction along the pipe length. This effect is evident in case 4, in which the most viscous oil presented led to the highest liquid volume fraction drop when comparing with the other cases.

NOMENCLATURE

Latin letters:

$A_{\alpha\beta}$	Interfacial area per unit volume between phases α and β	m^{-1}
C_D	Drag coefficient	—
d	Pipe diameter	m
E_K	Acceleration term	—
f_α	Volume fraction of phase α	—
f_{TP}	Two-phase friction factor	—

g	Acceleration of gravity	m/s²
M_α	Interfacial forces on phase α	N/m³
p	Pressure	Pa
$S_{MS\alpha}$	External forces on phase α	N/m³
T	Temperature	K
$\vec{U}$	Velocity vector	m/s
V_G	Gas mean velocity	m/s
V_L	Liquid mean velocity	m/s
v_m	Mixture velocity	m/s
v_S	Superficial velocity	m/s
v_{SG}	Gas superficial velocity	m/s
v_{SL}	Liquid superficial velocity	m/s
z	Vertical position	m

Greek letters:

α, β	Phases	—
θ	Angle with the horizontal plane	—
μ	Dynamic viscosity	N.s/m²
ρ	Density	kg/m³
τ^{turb}	Turbulent Reynolds stresses	kg/(m.s²)

Symbols:

$\nabla\cdot$	Divergence operator	—
∇	Gradient operator	—
$\otimes$	Tensor product	—

ACKNOWLEDGEMENTS

The authors thank Petrobras and Coordenação de Aperfeiçoamento de Pessoal de Nível Superior (CAPES) for supporting this project.

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