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Article

Evaluation of the Behavioral Characteristics in a Gas and Heavy Oil Stratified Flow According to the Herschel–Bulkley Fluid Model

Lin-tong Hou, Shuo Liu, Jian Zhang, and Jing-yu Xu*

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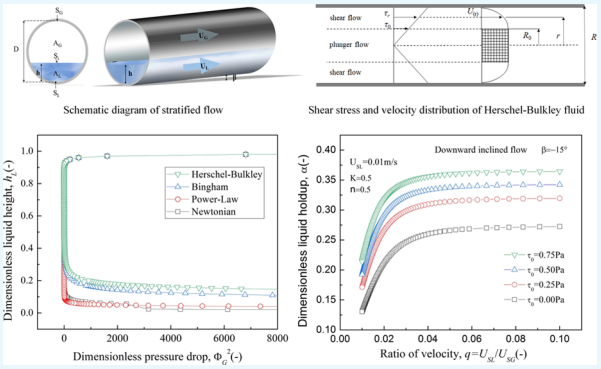
ABSTRACT: At present, most researches on gas–liquid two-phase flow use a power-law fluid model. However, with the development of unconventional petroleum resources and the restarting of heavy oil, the fluid showed strong yield characteristics. The power-law constitutive will not be able to express the yield-pseudoplastic fluid rheological properties. In order to make the study applicable to a larger range of shear rates, this study used the Herschel–Bulkley fluid model to discuss the gas–liquid flow characteristics. Based on the Herschel-Bulkley fluid constitutive, a two-fluid model, combined with dimensionless and iterative calculation methods, was used to theoretically derive the prediction model of liquid holdup and pressure drop for gas–liquid stratified flow. The effects of non-Newtonian fluid rheological parameters, flow conditions, and pipeline geometry on Herschel–Bulkley fluid and gas stratified flow were further analyzed. The results show that the power-law index n and the yield stress τ_0 (characterizing the rheological characteristics of the liquid phase) have significant effects on the gas–liquid two-phase stratified flow. Specifically, the enhanced liquid yield and shear thinning characteristics will lead to an increase in liquid holdup and a decrease in pressure drop. Comparing with the experimental data, the calculation model proposed in this work has a good prediction effect and provides new insights into the flow behavior of gas and waxy heavy oil with yield stress.



1. INTRODUCTION

In the flow process of multiphase flow pipelines, there are many different flow patterns affected by factors such as phase materials and flow conditions. Among them, stratified flow is common in horizontal or inclined multiphase flow systems under the gravity field. Moreover, the flow information of stratified flows can be used as the initial conditions for flow pattern transition in the flow process.[1] Based on this, the flow characteristics of different flow patterns can be further studied.

Stratified flow can be divided into gas–liquid stratified flow, gas-mixture liquids stratified flow, and gas–liquid–solid stratified flow according to the different phases. The research of Lockhart and Martinelli[2] was the first to study the gas–liquid flow in a horizontal pipe, introduced the Lockhart–Martinelli parameter (X^2), and established a general equation for predicting the liquid holdup and pressure drop at the interface. A one-dimensional two-fluid model was established, and it is still widely used in the research of the fully developed steady state of multiphase flow.[3] Gas-mixture liquid flow can generally be considered as a special two-phase flow of gas and non-Newtonian fluids. The study of Zhang and Xu[4] indicates that oil–water mixtures have shear thinning rheological properties of non-Newtonian fluids through rheological test. The flow parameters can be calculated by measuring the apparent viscosity of the non-Newtonian fluid. Feng et al.[5] and others carried out indoor simulation experiments to study the liquid properties of heavy oil–water and the stability of the system, which promoted the study of liquid–liquid mixtures flow. In addition, Cao et al.[6] conducted a numerical study on bubble movement and mass transfer characteristics in a gas-yield fluid two-phase flow and discussed the influence of fluid yield stress and other factors on the mass transfer rate.

In experimental research and practical applications, gas is generally air or natural gas in stratified flow, which is a typical Newtonian fluid, while the liquid phase, liquid–liquid mixed phase, and liquid–solid mixed phase can be regarded as a non-Newtonian fluid. The abovementioned multiphase stratified flow can be summarized as Newtonian fluid and non-Newtonian fluid two-phase flow. The constitutive relations commonly used to characterize the rheological properties of

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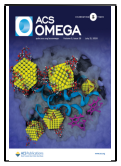

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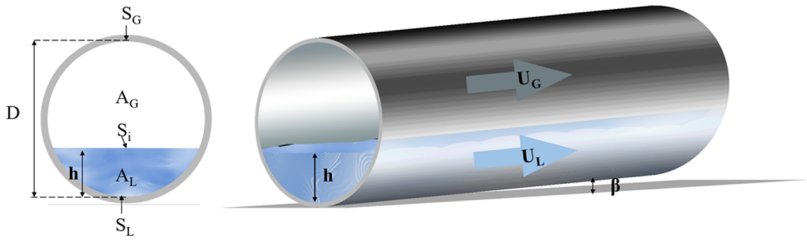

Figure 1. Schematic diagram of stratified flow.

such non-Newtonian fluids are as follows: power-law model, Bingham model, and Herschel−Bulkley model. In the study of gas−liquid stratified flow in the medium-low viscosity liquid phase range, Newtonian and power-law fluid models are mostly adopted. Yu and Han[7] used power-law models to solve the equation of motion in the two-phase flow process by theoretical research and determined the velocity distribution for the fully develop flow. In recent years, more progress has been made in the study of Newtonian/power-law fluid two-phase stratified flow. Some scholars used the power-law fluid constitutive to conduct the gas−liquid stratified flow in horizontal pipes and discussed the effect of rheology on the stratified flow characteristics. Based on the theoretical derivation of the two-fluid model, the dimensionless parameters for calculating the liquid holdup, pressure drop, drag reduction effect, and interface shear stress were obtained. The results were extended to micro-inclined pipe conditions, and a large number of experimental verifications were performed.[8−10] In order to research the universal applicability of the gas/power-law fluid flow prediction model, Picchi et al.[11] extended the mechanical steady-state model of the above study to other flow patterns and confirmed it with laboratory simulation experiments. Picchi and Poesio[12] derived equations of a gas−liquid stratified flow and obtained the effect of liquid phase rheology on flow through dimensionless treatment.

However, more attention needs to be paid to the complex fluid problems brought about by technological progress. Shear thinning fluid with yield stress plays an important role in industrial applications. Research on such fluids is not yet systematic, but there are some valuable studies to learn from the work of Li et al.,[13] which researched the yield stress of waxy crude oil under different conditions. Moreover, Bingham fluid, as a constitutive model with yield stress, has received increasing attention since the middle of the last century. Dean et al.[14] provided a review of the Bingham fluid theory and numerical calculation methods for the study of the application of elastoplastic fluids in multiphase flows in recent years. Firouzi and Hashemabadi[15] used the Bingham fluid constitutive to solve the kinematic equation of two-phase stratified flow and found that the rheological characteristics have a significant impact on the flow pressure drop and velocity distribution. Recently, Picchi et al.[16−18] have found the limitations of the widely used power-law fluid model to describe the stratified flow of gas/high-viscosity shear thinning fluid. Therefore, an iterative algorithm based on the Carreau model is proposed to calculate the exact solution of Newtonian/non-Newtonian shear thinning fluid pipe flow to correctly describe the behavior of the liquid phase at a high shear rate and then obtained the effect of shear thinning fluid rheology on the two-phase stratified flow.

In order to expand the research work of stratified flow of gas/non-Newtonian fluid, this work presents theoretical equations of the fully developed gas−liquid stratified pipe flow in horizontal and inclined pipelines with low phase velocities to predict liquid holdup and two-phase pressure drop. Considering the limitations of the constitutive model, power-law and Bingham fluid models cannot be used well for heavy oils. Therefore, this work uses the Herschel−Bulkley model constitutive to more accurately describe the above-mentioned non-Newtonian fluid. The Herschel−Bulkley fluid model can characterize the yield and pseudoplasticity of the fluid and can describe the rheological characteristics of the fluid under a large shear rate range. This work comprehensively considers the effect of rheological properties on the flow characteristics of two-phase stratified pipe flow, which can also provide theoretical guidance in engineering applications.

2. THEORETICAL MODEL DESCRIPTION

2.1. Problem Formulation and Rheology Model for Herschel−Bulkley Fluid. Taitel and Dukler[3] derived a two-phase horizontal stratified flow model of gas/Newtonian fluid, which was extended by Heywood and Charles[19] to non-Newtonian liquid phase (Figure 1). It is assumed that two-phase fluid flows in one dimension along the pipe length, without considering the heat and mass transfer and phase transformation and ignoring the effects of acceleration and hydraulic gradient in the liquid phase. The momentum balance of steady-state flow of each phase in the two-fluid model is as follows:

$$-A_L\left(\frac{dP}{dx}\right)_{TP} - \tau_L S_L - \tau_i S_i - \rho_L A_L g \sin\beta = 0 \tag{1}$$

$$-A_G\left(\frac{dP}{dx}\right)_{TP} - \tau_G S_G - \tau_i S_i - \rho_G A_G g \sin\beta = 0 \tag{2}$$

where A is the area, τ is the shear stress, S is the wetted periphery, ρ is the density, g is the acceleration due to gravity, and β is the angle of inclination from the horizontal. Subscripts TP, L, G, and i refer to the two-phase, liquid phase, gas phase, and interface, respectively. x is the axial coordinate, and P is the static pressure. Eliminating the two-phase pressure gradient, $(dP/dx)_{TP}$, gives

$$\tau_G\frac{S_G}{A_G} - \tau_L\frac{S_L}{A_L} + \tau_i S_i\left(\frac{1}{A_L} + \frac{1}{A_G}\right) - (\rho_L - \rho_G) g \sin\beta = 0 \tag{3}$$

The wall shear stress and interface shear stress are defined, respectively, as

$$\tau_G = f_G \cdot \frac{\rho_G u_G^2}{2} \tag{4}$$

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$$\tau_L = f_L \cdot \frac{\rho_L u_L^2}{2} \tag{5}$$

$$\tau_i = f_i \cdot \frac{\rho_G (u_G - u_L)|u_G - u_L|}{2} \tag{6}$$

where u is the average velocity and f is the friction factor in a smooth pipe of each phase. f can be defined as

$$f_L = C_L Re_L^{-n} \tag{7}$$

$$f_G = C_G Re_G^{-m} \tag{8}$$

where $C_L = C_G = 16$, $n = m = 1$ for laminar flow, which also appeared in the research of Xu[10] and Bishop and Deshande.[20]

As for gas−liquid interface stress, in concurrent flow, the volumetric flow rates of both phases are positive $q_{L,G} > 0$. The interface friction is usually approximately equal to the wall friction of the layer with fast flow velocity $|u_G| > |u_L|$, $f_i = f_G$, which meets the conditions of gas−liquid two-phase flow velocity.

The Reynolds number for the gas phase as Newtonian fluid is defined by

$$Re_G = \frac{\rho_G u_G D_G}{\mu_G} \tag{9}$$

$$D_G = \frac{4A_G}{S_G + S_i} \tag{10}$$

For non-Newtonian materials, the rheological behavior obeys the Herschel−Bulkley model, $\tau = \tau_0 + K\dot{\gamma}^n$, which can be transformed into

$$-\frac{du}{dr} = \begin{cases} \left[\frac{1}{K}\left(\frac{\Delta P}{2L} r - \tau_0\right)\right]^{1/n} & R_0 \le r \le R \\ 0 & r < R_0 \end{cases} \tag{11}$$

The appropriate Reynolds number is defined by eq 12 described from the work of Chhabra and Richardson:[21]

$$Re_L = \frac{8\rho_L u_{ann}^2}{\tau_0 + K\left(\frac{8u_{ann}}{D_{shear}}\right)^n} \tag{12}$$

The Herschel−Bulkley fluid velocity distribution discussed in this work is shown in Figure 2. There are additional flow

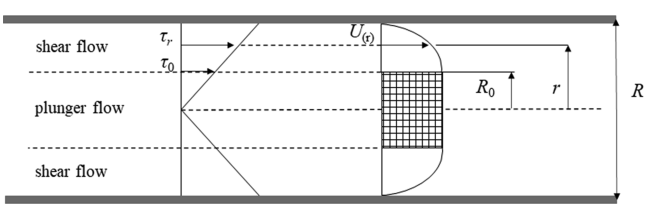

Figure 2. Shear stress and velocity distribution of Herschel−Bulkley fluid.

conditions in the Herschel−Bulkley fluid pipeline flow: plug flow and laminar flow. This study does not discuss the above two cases; the applicable conditions of the three cases are derived in the Appendix section:

$$u_{ann} = \frac{Q - Q_{plug}}{\pi(R^2 - R_0^2)} \tag{13}$$

$$D_{shear} = 2(R - R_0) = D_L(1 - \phi) \tag{14}$$

$$D_L = \frac{4A_L}{S_L} \tag{15}$$

where D_L and D_G are the equivalent diameter for the liquid and gas phases, respectively. τ_0 and τ are the yield stress and wall shear stress, respectively. $\dot{\gamma}$ is the strain rate, K is the consistency coefficient, and n is the power-law index. $\Delta P/L$ is the liquid pressure gradient.

In the expression of Reynolds number, u_{ann} is the velocity of liquid-velocity-gradient zone, and D_{shear} is the size of liquid-velocity-gradient zone, which can be calculated by liquid equivalent diameter. ϕ is the ratio of plug-flow zone radius, R_0, to the liquid equivalent radius, R. $Q - Q_{plug}$ is the liquid flow of liquid-velocity-gradient zone, which can be calculated by integrating the liquid phase constitutive equation:

$$Q - Q_{plug} = \pi \frac{nR^{(3n+1)/n}(1-\phi)^{(2n+1)/n}(1 + 2n + \phi + 4n\phi)\left(\frac{\Delta P}{2KL}\right)^{1/n}}{(3n+1)(2n+1)} \tag{16}$$

2.2. Problem Normalization. Stratified two-phase incompressible laminar flow in a pipe under steady and fully developed conditions, when non-Newtonian fluid flows at the bottom of a circular tube, the dimensionless momentum equation for the stratified gas−liquid two-phase flow reads

$$\begin{aligned} &\frac{\pi \tilde{S}_G(\tilde{S}_G + \tilde{S}_i)}{2\tilde{A}_G^3} \cdot \tilde{W} \\ &- \frac{2^{4/n}(1-\phi^2)^2(3n+1)^2(2n+1)^2}{n^2(1-\phi)^{(2/n)+4}(1+2n+\phi+4n\phi)^2} \\ &\cdot \left[\phi + 4^n \frac{n^2(1-\phi)^{n+1}(1+2n+\phi+4n\phi)^n}{(3n+1)^n(2n+1)^n(1-\phi^2)n}\right] \cdot \\ &\frac{\tilde{u}_L^2 \tilde{S}_L q^2}{\tilde{D}_L^{(2/n)+1}\tilde{A}_L (X^2)^{(2/n)-1}} \cdot \tilde{P} \\ &+ \frac{\pi \tilde{S}_i(\tilde{S}_G + \tilde{S}_i)(\tilde{A}_G + \tilde{A}_L)\left(\tilde{A}_G^{-1} - \frac{\tilde{A}_L^{-1}}{q}\right)^2}{\tilde{A}_L \tilde{A}_G} \cdot \tilde{W} - \tilde{Y} \\ &= 0 \end{aligned} \tag{17}$$

where X^2 is the Lockhart−Martinelli parameter[2] defined as

$$X^2 = \frac{(dP/dx)_{SL}}{(dP/dx)_{SG}} \tag{18}$$

where dP_F/dx refers to the frictional pressure gradient and subscripts SG and SL refer to the superficial gas and liquid phase for either phase flowing alone in the channel, respectively.

The dimensionless parameters and variables used are

$$\tilde{P} = \frac{U_{SG}^2 K^{2/n}}{D^{(2/n)+2}(dP/dx)_{SG}^{2/n}} \tag{19}$$

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$$\tilde{W} = \frac{U_{SG}{}^{n}\mu_G}{D^{n+1}(dP/dx)_{SG}} \tag{20}$$

$$\tilde{Y} = \frac{(\rho_L - \rho_G)g\sin\beta}{(dP/dx)_{SG}} \tag{21}$$

$$\tilde{D}_G = D_G/D \tag{22}$$

$$\tilde{D}_L = D_L/D \tag{23}$$

$$\tilde{S}_G = S_G/D \tag{24}$$

$$\tilde{S}_L = S_L/D \tag{25}$$

$$\tilde{S}_i = S_i/D \tag{26}$$

$$\tilde{A}_G = A_G/D^2 \tag{27}$$

$$\tilde{A}_L = A_L/D^2 \tag{28}$$

where the liquid height is assumed to be h_L, whose dimensionless number $\tilde{h}_L$ is normalized with respect to the pipe diameter, D. The local phase velocity, $\tilde{u}_{G,L}$ is normalized with respect to the superficial velocity of each phase, U_{SG}, $_{SL}$, and $q = U_{SL}/U_{SG}$. All the dimensionless quantities with a tilde (~) in the above equations are functions of the dimensionless liquid height. With the help of the mass conservation equation, dimensionless velocity can be expressed as follows: $\tilde{u}_G = u_G/U_{SG} = \frac{\pi}{4}\tilde{A}_G{}^{-1}$ and $\tilde{u}_L = u_L/U_{SL} = \frac{\pi}{4}\tilde{A}_L{}^{-1}$.

2.3. Phase Holdup and Pressure Gradients. *2.3.1. Liquid Holdup.* According to the definition of liquid holdup

$$\alpha = \frac{A_L L}{AL} = \frac{A_L}{A} \tag{29}$$

where L is the length of the control body and A is the cross sectional area of pipeline. The holdup can be simplified as

$$\alpha = \frac{1}{\pi}[\cos^{-1}(1 - 2\tilde{h}_L) - (1 - 2\tilde{h}_L)\sqrt{1 - (1 - 2\tilde{h}_L)^2}\,] \tag{30}$$

2.3.2. Pressure Gradient. Eliminating the interfacial term from the gas and liquid phase momentum equations gives

$$\left(\frac{dP}{dx}\right)_{TP} = \frac{-1}{A_L + A_G}[\tau_L S_L + \tau_G S_G + (\rho_L A_L + \rho_G A_G)\, g\sin\beta] \tag{31}$$

Non-Newtonian liquid phase pressure gradient can be calculated by the formulation proposed by Metzner and Reed:[22]

$$(dP/dx)_{SL} = \frac{4}{D}K'\left(\frac{8V}{D}\right)^{n'} \tag{32}$$

where $V = U_{SL}$ and parameters K' and n' are the modified consistency coefficient, K, and power-law index, n, respectively. As for the Herschel–Bulkley model, K' and n' can be expressed as

$$K' = K\left(\frac{3n+1}{4n}\right)^n\left(\frac{1}{1-\phi}\right)\left(\frac{1}{1 - a\phi - b\phi^2 - c\phi^3}\right)^n,\ n' = n \tag{33}$$

Solving ϕ with the help of Bingham number, Bingham number characterizes the ratio of yield effect to viscosity effect, defining a function as follows:

$$Bn = \frac{\tau_0}{K\left(\frac{U_{SL}}{D}\right)^n} \tag{34}$$

$$F(\phi) = \left(\frac{\phi}{1-\phi}\right)\left(\frac{d}{1 - a\phi - b\phi^2 - c\phi^3}\right)^n - Bn,\ \phi \in [0, 1] \tag{35}$$

$$a = \frac{1}{2n+1} \tag{36}$$

$$b = \frac{2n}{(n+1)(2n+1)} \tag{37}$$

$$c = \frac{2n^2}{(n+1)(2n+1)} \tag{38}$$

$$d = 6 + \frac{2}{n} \tag{39}$$

where parameters a, b, c, and d are only related to power-law index, n:

$$(dP/dx)_{SL} = \frac{4}{D}K\left(\frac{3n+1}{4n}\right)^n\left(\frac{1}{1-\phi}\right)\left(\frac{1}{1 - a\phi - b\phi^2 - c\phi^3}\right)^n\left(\frac{8U_{SL}}{D}\right)^n \tag{40}$$

The gas frictional pressure gradient can be calculated by the gas phase momentum equation:

$$(dP/dx)_{SG} = \frac{\tau_G S}{A} = \frac{32\mu_G U_{SG}}{D^2} \tag{41}$$

The two-phase total pressure drop is composed of the gravitational pressure drop and the frictional pressure gradient, which can be determined by the liquid height. The dimensionless pressure drop is given by

$$\begin{aligned}\Phi_G{}^2 &= \frac{(dP/dx)_{TP}}{(dP/dx)_{SG}} \\ &= \frac{2^{4/n}(1-\phi^2)^2(3n+1)^2(2n+1)^2}{n^2(1-\phi)^{4+(2/n)}(1+2n+\phi+4n\phi)^2} \cdot\left[\phi + 4^n\frac{n^n(1-\phi)^{n+1}(1+2n+\phi+4n\phi)^n}{(3n+1)^n(2n+1)^n(1-\phi^2)^n}\right]\cdot \\ &\quad \frac{\tilde{u}_L{}^2\tilde{S}_L q^2}{\tilde{D}_L{}^{1+(2/n)}(\tilde{A}_L + \tilde{A}_G)(X^2)^{(2/n)-1}}\cdot\tilde{P} \\ &\quad + \frac{\pi\tilde{\mu}_G\tilde{S}_G(\tilde{S}_G + \tilde{S}_i)}{2\tilde{A}_G{}^2(\tilde{A}_G + \tilde{A}_L)}\cdot\tilde{W} + \tilde{Y}\cdot\frac{(\rho_L\tilde{A}_L - \rho_G\tilde{A}_G)}{(\tilde{A}_G + \tilde{A}_L)(\rho_L - \rho_G)}\end{aligned} \tag{42}$$

$$\Phi_L{}^2 = \frac{(dP/dx)_{TP}}{(dP/dx)_{SL}} = \frac{\Phi_G{}^2}{X^2} \tag{43}$$

2.4. Solution Algorithm. The solution of two-phase momentum (eq 3) needs to be obtained by numerical calculation. Its form is transformed into dimensionless (eq

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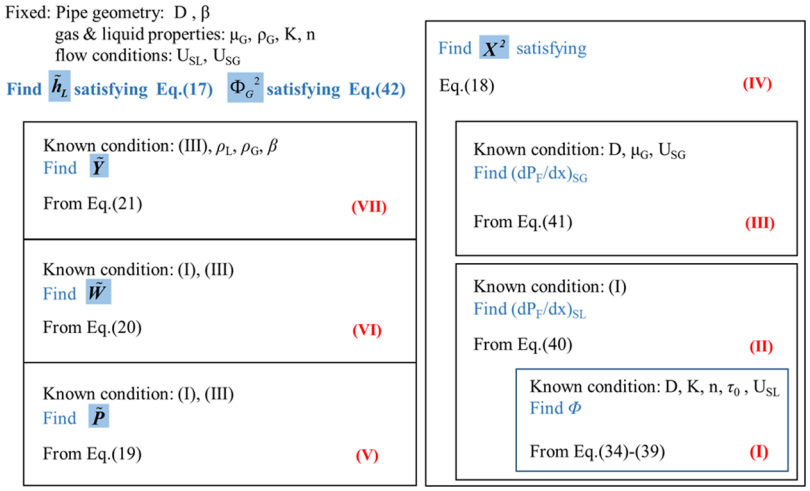

Figure 3. Block diagram of the algorithm to compute the laminar solutions.

17) with only one unknown parameter (dimensionless liquid height), which can be obtained by iterative calculation. First of all, the given pipeline conditions, phase material properties, and flow parameters are brought into the corresponding equations (eqs 40 and 41) to obtain the friction pressure gradient of each phase; then, the three redefined dimensionless quantities are solved by eqs 19, 20, and 21 to characterize the factors of two-phase stratified flow; subsequently, the quantities are brought back to the momentum equation of two-phase flow, and the dimensionless liquid level height can be solved by iterative single variable function using eqs 30 and 42; the calculation results of liquid holdup and pressure gradient can be obtained. The implemented iterative scheme is shown in Figure 3.

3. RESULTS AND DISCUSSION

In order to validate the method of two-phase stratified flow, the experimental data are compared with the results predicted from eqs 17, 30, and 42, as shown in Figure 4. The results of these comparisons indicate good agreement for the liquid holdup data. The theoretical prediction model of gas−liquid two-phase laminar liquid holdup derived in this work is compared with the experimental results of some scholars, including horizontal and inclined flows. Among the experimental data published in the field of gas−liquid two-phase flow, some original experimental results of Xu et al.[9] and Bishop and Deshande[20] were selected.

In order to verify the theoretical model of the gas−liquid stratified flow of Herschel−Berkeley fluid, the verification data of the horizontal pipe flow adopted the research results of two scholars. Among them, in the work of Xu et al.,[9] the inner diameter of the test pipeline was 25 mm, and the superficial liquid velocity is fixed at 0.015 m/s. The experimental working medium is a water-containing polymer, and the power-law fluid has a fluidity index of 0.85. In the work of Bishop and Deshande,[20] the diameter of the tube is 50 mm, the superficial velocity of the liquid phase is 0.1769 m/s, and the liquid phase is a polymer solution with a power-law index of 0.952. The figure shows that the prediction curve of the liquid retention rate deduced in this work is in good agreement with the experimental data of the two horizontal tubes. In the horizontal stratified flow, the gas−liquid flow calculation model derived

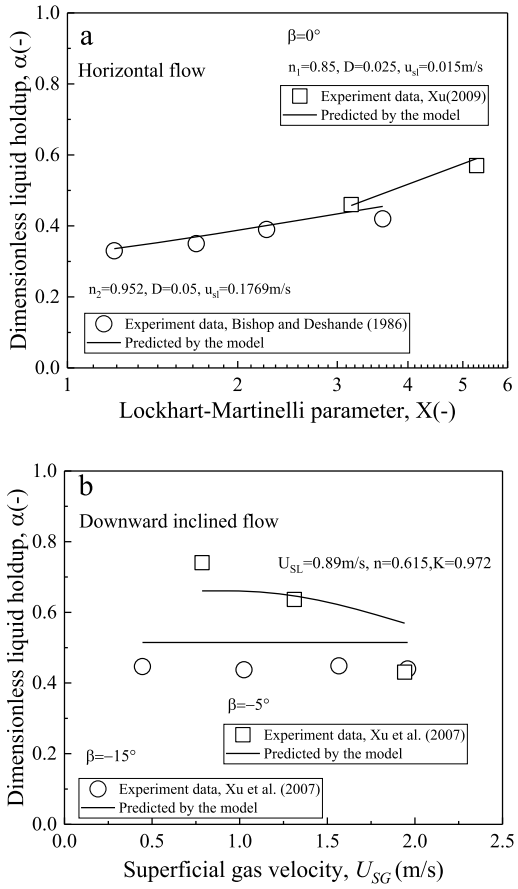

Figure 4. Comparison of the theoretical predictions obtained for the liquid holdup with experimental data from the works of Xu et al. (2007) and Bishop and Deshande (1986) in (a) horizontal and (b) inclined stratified flow regimes.

from the Herschel−Bulkley fluid constitutive model has a good prediction effect.

In the inclined gas/liquid two-phase pipe flow, the experimental data of 5° and 15° downward in Xu et al.[9] were selected for verification. The experimental working condition was 50 mm inner diameter of the pipeline, the superficial liquid velocity was 0.89 m/s, and the experimental working fluid used a polymer aqueous solution with a

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consistency coefficient of 0.972 and a fluidity index of 0.615. Its characteristics were similar to crude oil. The figure shows that there is a gap between the prediction model and the experimental data for two reasons: First, the flow pattern formed by the higher superficial velocity of the liquid phase is not a stable stratified flow but the middle of the transition from the stratified flow to intermittent flow pattern. Second, in the downward inclined pipe, the gravity effect increases with the angle change, leading to the unstable development of the flow pattern during the gas−liquid flow.

3.1. Liquid Holdup. As mentioned above, the dimensionless liquid holdup can be iteratively calculated by eq 17 for gas/non-Newtonian liquid stratified flow in an inclined pipe. Our aim is to investigate the effect of the rheology of the non-Newtonian liquid on the two-phase flow characteristics, focusing mainly on the trends of integral variables (e.g., liquid holdup and pressure gradient). The study is carried out for gas and Herschel−Bulkley fluid two-phase stratified flow.

3.1.1. Effect of Liquid Rheological Properties. Using the prediction model proposed in this work, four different fluid constitutive relationships (Newtonian fluid, power-law fluid, Bingham fluid, and Herschel−Bulkley fluid) are obtained by changing the liquid phase characteristic parameters. It is now assumed that the characteristic parameters of four different constitutive fluids are shown in Table 1.

Table 1. Characteristic Parameters of Four Different Constitutive Fluids

fluid type	τ_0 (Pa)	K (Pa·s^n)	n (−)
Newtonian	0	0.5	1
power-law	0	0.5	0.5
Bingham	0.5	0.5	1
Herschel−Bulkley	0.5	0.5	0.5

Figure 5 shows the effect of fluid type on gas−liquid two-phase stratified flow. The Herschel−Bulkley fluid has a higher

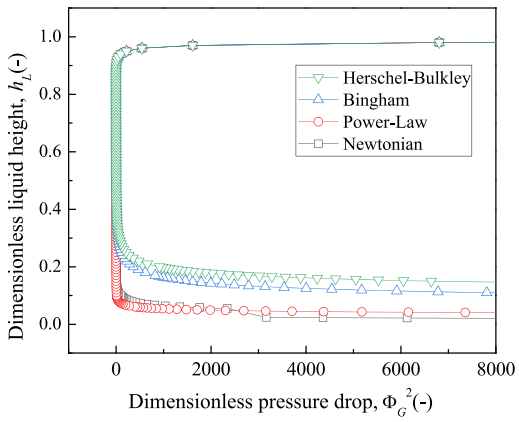

Figure 5. Differences in liquid level of liquid constitutive relationships for gas/liquid stratified flow.

liquid holdup than other fluid constitutive types in the theory of gas/liquid two-phase stratified flow. The difference between the Herschel−Bulkley fluid and the power-law fluid is that the Herschel−Bulkley fluid has an initial yield stress, which makes the fluid characteristics between the liquid phase and the solid phase, and is more suitable for the characterization of ultrahigh viscosity oil. The Herschel−Bulkley fluid is compared with the Bingham fluid containing the original shear stress. Its characteristic is that the Herschel−Bulkley fluid has a power-law index that allows the fluid to exhibit more accurate fluid properties when subjected to shear. The Herschel−Bulkley fluid constitutive contains both power-law index and initial dynamic shear force, which can better describe the rheological characteristics of non-Newtonian fluids to a certain extent.

In the existing studies of gas−liquid stratified flow, the specificity of the Herschel−Bulkley fluid constitutive is rarely discussed. The fluid can be regarded as a power rate constitutive with yield stress in essence. Power-law index and yield stress are important rheological parameters describing the characteristics of the fluid. Figure 6 presents the relationship

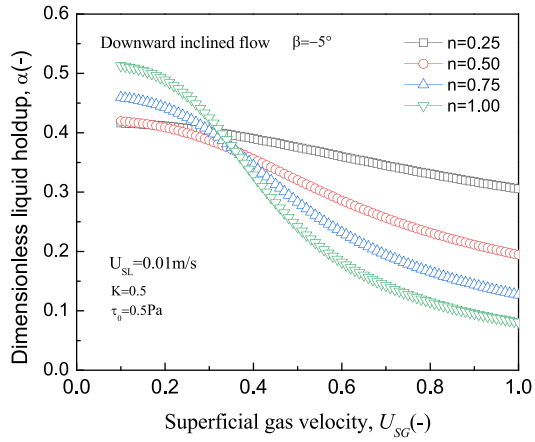

Figure 6. Relationship between the superficial gas velocity and liquid holdup for different rheological properties of non-Newtonian liquid.

between the superficial gas velocity and liquid holdup for different rheological properties of non-Newtonian liquid. As known in the figure, when the superficial gas velocity is small, the effect of power-law index, n, on holdup does not become a law. The small superficial gas velocity results in the weak interaction between the two phases, which means weak momentum exchange and interference between gas and liquid phases. When the gas velocity increases to a certain value, the flow shows an obvious rule: there is an increase in liquid holdup with the decrease in power-law index. Because the viscosity of liquid increases with the decrease in n, the liquid fluidity is weakened, which shows the accumulation of liquid level height. As the superficial gas velocity increases, the liquid with a smaller power-law index has stronger non-Newtonian characteristics, and the liquid holdup decreases more slowly. In addition, it can be observed in Figure 7 that when the pipeline is horizontal, the liquid holdup suddenly drops. The reason is that the gas/Herschel−Bulkley fluid stratified flow needs a suitable range for the value of the gas−liquid ratio. When the liquid−gas velocity ratio, q, is small, the stratified flow pattern studied can be satisfied.

Comprehensively considering the influence of gas−liquid viscosity ratio, power-law index, and yield stress on the two-phase flow, the Lockhart−Martinelli parameter, X^2, can be introduced. Equation 44 can be obtained by deriving eqs 18, 32, and 41:

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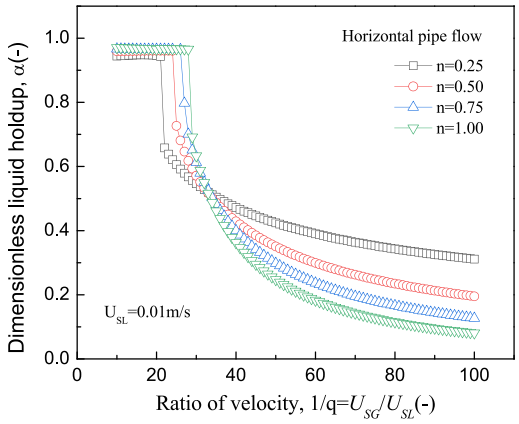

Figure 7. Effect of power-law index, n, on the liquid holdup of the gas/non-Newtonian liquid stratified flow.

$$X^2 = q\frac{K}{\mu_G}\left(\frac{3n+1}{4n}\right)^n\left(\frac{1}{1-\phi}\right)\left(\frac{1}{1-a\phi-b\phi^2-c\phi^3}\right)^n \left(\frac{8U_{SL}}{D}\right)^{n-1} \quad (44)$$

Figure 8 depicts the effect of power-law index, n, on the dimensionless liquid height in downward inclined stratified flows. It shows the liquid height for non-Newtonian liquid as a function of the Lockhart−Martinelli parameter, X^2, in stratified flows for various power-law index values corresponding to the shear-shinning fluid behavior as predicted from eq 11. As a whole, the liquid level decreases with the increase in $1/X^2$. Different power-law exponent n values form the difference in the speed of decrease in liquid level height, which shows that the larger the value of n is, the higher the decrease speed of liquid level height. The power-law index, n, has a great influence on the height of liquid surface when $1/X^2$ is small. In the two-phase downward flow, when $1/X^2 > 1/2000$, the liquid height increases obviously with decreasing n. In the horizontal and upward flow, when $1/X^2 < 1/2000$, the liquid level height is very high, but the liquid level height is almost unchanged. Because a smaller $1/X^2$ corresponds to a larger gas−liquid ratio q, under this condition, the two-phase flow has exceeded the range of stratified flow, and this section will not be discussed.

Figure 9 shows that the dimensionless liquid holdup increases with increasing τ_0 and the speed of holdup increases gradually. The viscosity of the liquid increases with increasing yield stress, which shows the poor follow-up ability. In the process of flow, the interface friction pressure between phases is not enough to drive the high viscosity liquid, which results in a big liquid phase height and the increase in liquid holdup. As shown in Figure 10, with increasing τ_0, the holdup becomes larger gradually and the slope of this curve gradually decreases, that is, the growth rate of the liquid holdup gradually slows down. In this figure, the black line represents the function

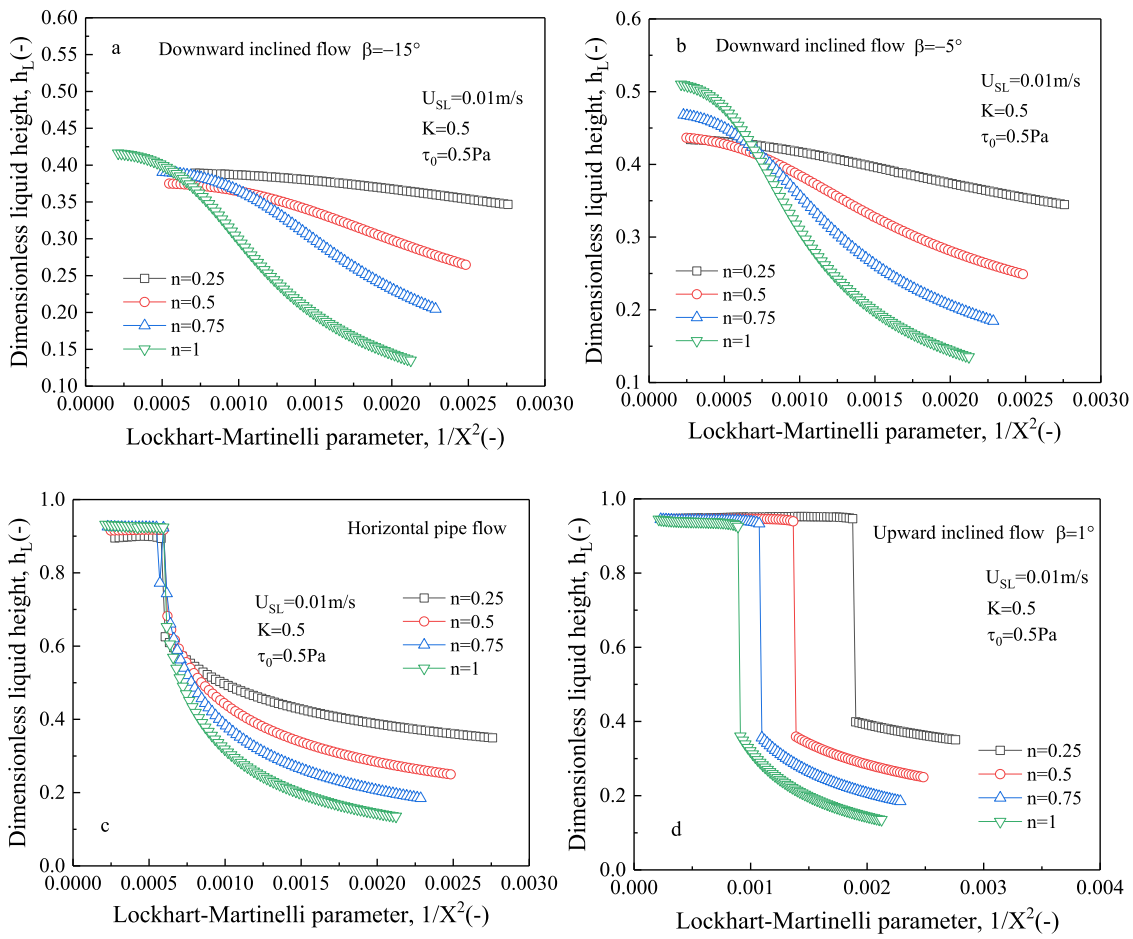

Figure 8. Effect of power-law index, n, on the dimensionless liquid height in inclined stratified flows for a gas/non-Newtonian liquid flow: (a) $\beta = -15°$, (b) $\beta = -5°$, (c) $\beta = 0°$, and (d) $\beta = 1°$.

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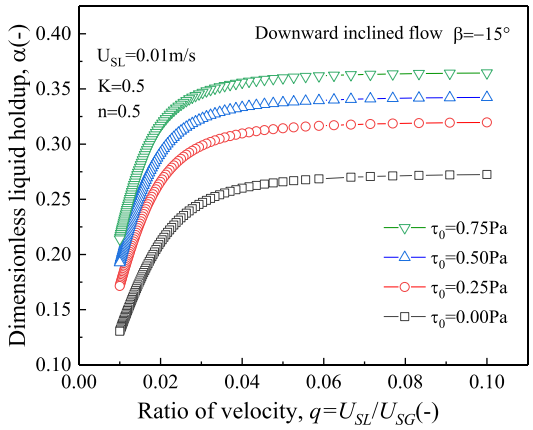

Figure 9. Dimensionless liquid holdup, α, as a function of the velocity ratio q for stratified flow of gas/non-Newtonian liquids in downward pipeline.

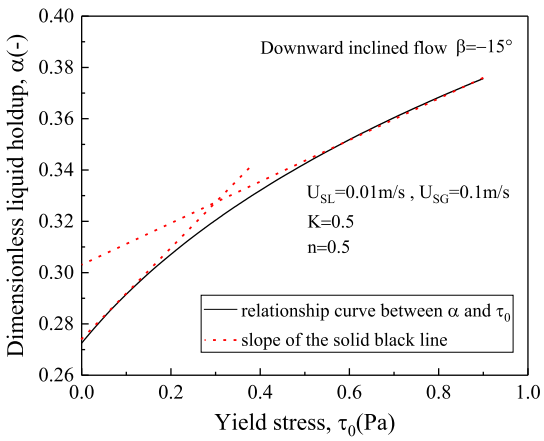

Figure 10. Effect of yield stress, τ_0, on the dimensionless liquid holdup, α, for stratified flow of gas/non-Newtonian liquids in downward co-current pipe flow.

curve of the dimensionless liquid holding capacity and the yield stress. The two sets of red lines describe the slope of the above curve under high and low yield stress, respectively, and the lines representing the slope approximately intersects at the position of τ_0 = 0.3. When τ_0 increases to around 0.3, the slope hardly changes. The above conclusion can also be confirmed by the distance of each curve in Figure 9 being reduced first and then unchanged.

Liquid holdup is related to the increase in superficial gas velocity with different yield stress values. As shown in Figure 11, when the superficial gas velocity is not greater than 0.6 m/s, the difference in yield stress value will cause a large difference in liquid holdup. As the superficial gas velocity increases, the effect of yield stress on the two-phase stratified flow decreases and it is reasonable to predict that, when the superficial velocity of the gas phase increases to a certain value, this effect can be ignored.

In order to further explore the influence of liquid phase rheological properties on gas−liquid two-phase flow, the Bingham number (eq 34) is used for characterization. The definition of Bingham number includes the power-law characteristics and yield characteristics of the liquid medium. Figure 12 shows the liquid holdup for Herschel−Bulkley liquids as a function of the Lockhart−Martinelli parameter, X^2, in stratified horizontal flow for various Bingham number values

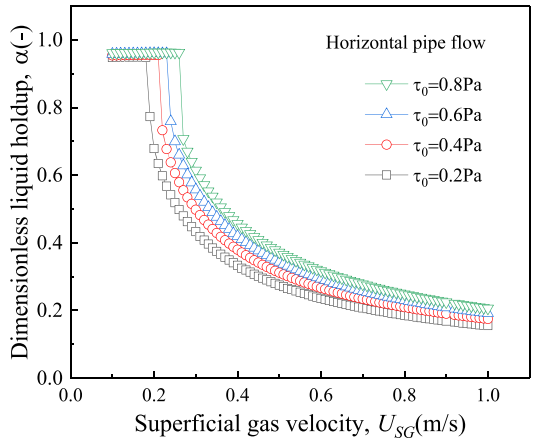

Figure 11. Variation of liquid holdup with gas velocity at different yield stress values.

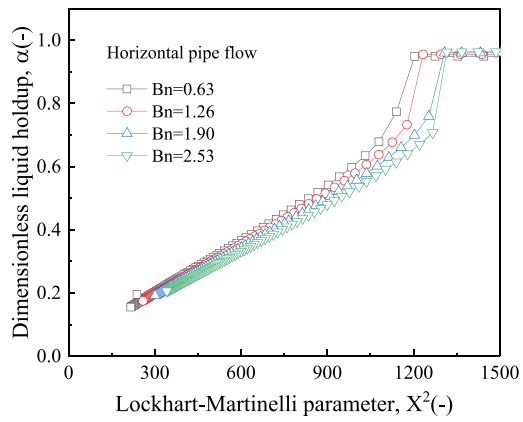

Figure 12. Liquid holdup for Herschel−Bulkley liquids as a function of the Lockhart−Martinelli parameter, X^2, in stratified horizontal flow.

corresponding to the Herschel−Bulkley fluid behavior as predicted from eq 17. When X^2 is constant, as the Bingham number increases, the liquid holdup decreases. In addition, when the value of X^2 is small, the gas−liquid ratio is relatively large, and the influence of gas on the two-phase wave surface is dominant. The effect of the Bingham number, Bn, on the liquid holdup, α, is weaker. Over the range of X^2 from 0 to 1500, α increases for a given Bn from about 0.1 to 1.0, which is in agreement with Xu et al.[10] Moreover, Figure 12 also illustrates that, in the range of relatively high X^2 values, the effect of Bn values on α is of main importance in a gas/liquid stratified flow. When X^2 exceeds 1200, because the gas−liquid ratio reaches a high value, the two-phase flow pattern is no longer a stratified flow, and the theoretical formula derived in this work is no longer applicable, resulting in a sudden step of liquid holding rate to a high value.

Figure 13 shows the effect of the Bingham number on the two-phase stratified flow. It can be seen that, when the value of ${\Phi_G}^2$ is less than 1000, an inflection point appears in the dimensionless pressure drop curve, and the rate of pressure drop decline slows down. Near the inflection point, the value of the Bingham number has the most obvious influence on the pressure drop.

3.1.2. Influence of Incoming Flow Conditions. In downward flow, the liquid holdup, α, increases obviously with the increase in the superficial velocity of liquid phase, U_{SL}, as be shown in Figure 14. When the superficial liquid velocity, U_{SL}, is

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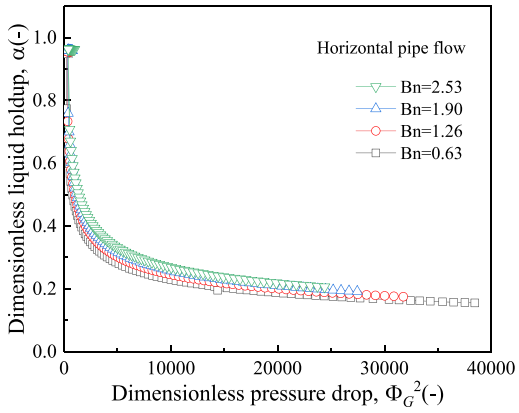

Figure 13. Liquid holdup versus the dimensionless pressure drop, Φ_G^2, in stratified horizontal flow with various Bingham number values.

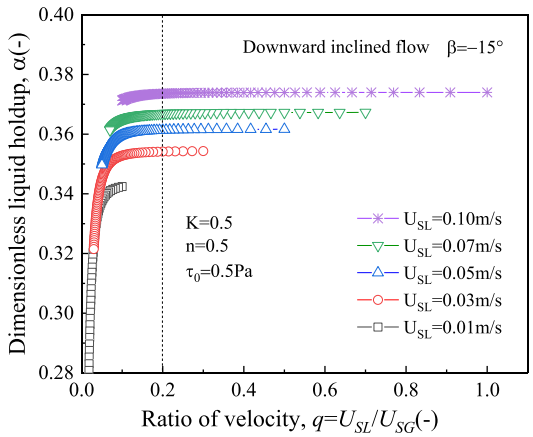

Figure 14. Dimensionless liquid holdup, α, as a function of the superficial liquid velocity, U_{SL}, of downward stratified flow.

fixed, the effect of q on holdup can be negligible in the condition of a high value of q (a low value of U_{SG}). When $q < 0.2$, which means that the superficial velocity of the gas phase, U_{SG}, increases to nearly five times of that of liquid phase, U_{SL}, the holdup decreases significantly.

Figure 15 shows the effect of the velocity ratio, q on the liquid height in inclined pipe for downward stratified flows. As it can be seen, the liquid height increases gradually with

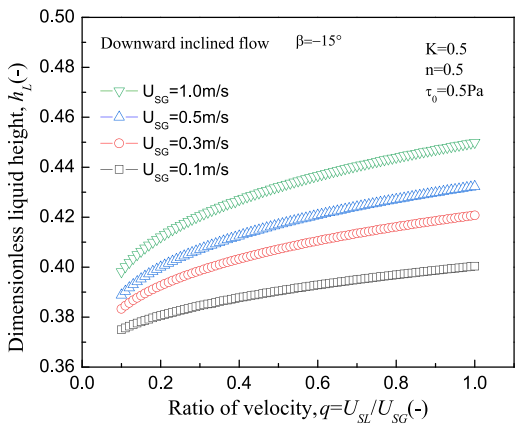

Figure 15. Dimensionless liquid height, h_L, as a function of the velocity ratio, q, of gas/non-Newtonian liquid stratified flow in different superficial gas velocity conditions.

increasing superficial liquid velocity when the superficial gas velocity is a constant. However, the liquid height, h_L, still shows a gradually increasing trend with the increase in U_{SG} and a constant value of q. The reason is that the liquid phase velocity plays a decisive role in the change of liquid holdup during two-phase stratified flow while increasing the superficial gas and liquid velocities by the same multiple at the same time.

3.1.3. Role of Pipeline Inclination in Two-Phase Flow. The pipe geometry directly affects the gas−liquid two-phase flow behavior. Different inclination angles of pipes will form different flow patterns and flow characteristics. Figure 16 shows the effect of two-phase velocity ratio and pipe inclination on the liquid holdup. It can be seen that, with the increase in the ratio of the superficial gas and liquid velocities, $1/q$, the liquid holdup decreases gradually. The difference in Figure 16a,b is as follows: the inclination angle of the pipeline is negative, that is to say, the gas/liquid two-phase flow is stratified downward. With the increase in $1/q$, the liquid holdup decreases in the form of power exponent and remains as a small constant at last. When the gas/liquid two-phase flow is horizontal or upward, the liquid holdup is very high but decreases sharply. The reason is that, in the horizontal and oblique upward flow, when the gas−liquid ratio exceeds the appropriate range, it is easy to form a non-stratified flow, and the liquid holding rate step appears in the figure. The stepped part with the higher liquid level and liquid holding rate does not meet the research conditions of this study, which can be seen in Figures 16b and 17b.

3.2. Dimensionless Two-Phase Pressure Gradient. The factors and changes of pressure drop of gas−liquid two-phase stratified flow in inclined pipe and horizontal pipe are discussed.

3.2.1. Horizontal Pipe Flow. The dimensionless pressure drop, Φ_G^2, of the gas−liquid two-phase stratified flow in the horizontal pipe flow is significantly affected by the parameters of the liquid medium. As shown in Figure 18, when the superficial velocity of the gas phase is large, the pressure drop of the yield stress is more obvious. As the yield stress value increases, the fluidity of the liquid phase is poor, which will cause the flow pressure drop to decrease. When the apparent velocity of the gas phase is small, the effect of yield stress on the pressure drop of the two-phase flow is almost negligible. The superficial gas velocity is positively correlated with the dimensionless pressure drop. The gas phase plays a lubricating role in the two-phase flow system. The increase in the gas phase velocity enhances the fluidity and the pressure drop.

Figure 19 shows the effect of the power-law behavior index, n, on the dimensionless pressure gradient in a stratified flow. In the prediction model proposed in this study, when the power index $n = 1$, the liquid phase behaves as Bingham fluid, and the pressure drop changes drastically with the gas phase velocity. When the power-law index gradually decreases, the characteristics of the Herschel−Bulkley constitutive model are obvious, and the viscosity of the liquid phase increases. As a result, the pressure drop tends to be flat.

Considering the liquid phase power-law characteristics and yield characteristics comprehensively, the influence of the change of Bingham number on the pressure drop is discussed. As shown in Figure 20, in the relationship curve of dimensionless pressure drop, Φ_G^2, and Lockhart−Martinelli parameter, X^2, the change of Bingham number has a significant effect on the pressure drop of the two-phase flow. When X^2 is

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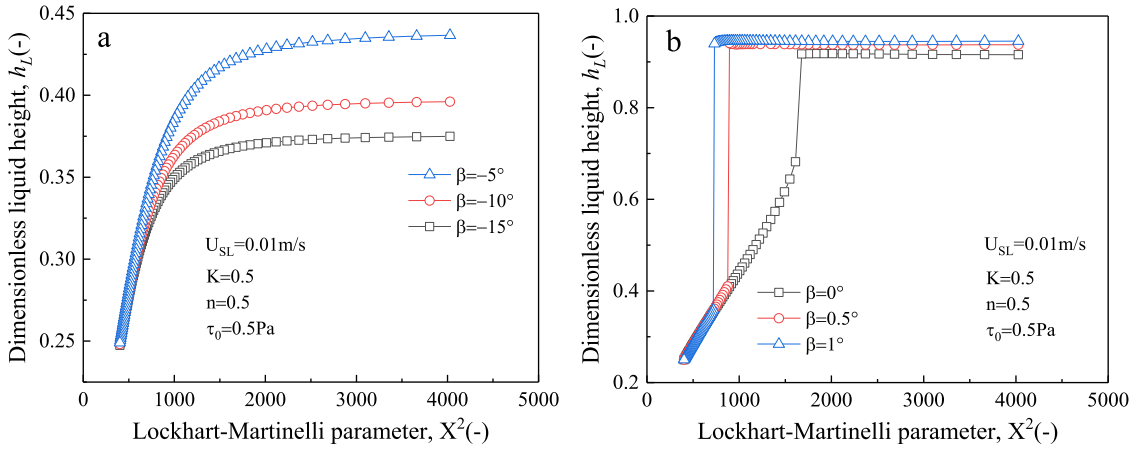

Figure 16. Effect of pipe inclined angles, β, on the dimensionless liquid height, h_L, for stratified flow: (a) downward flow and (b) horizontal and upward flow.

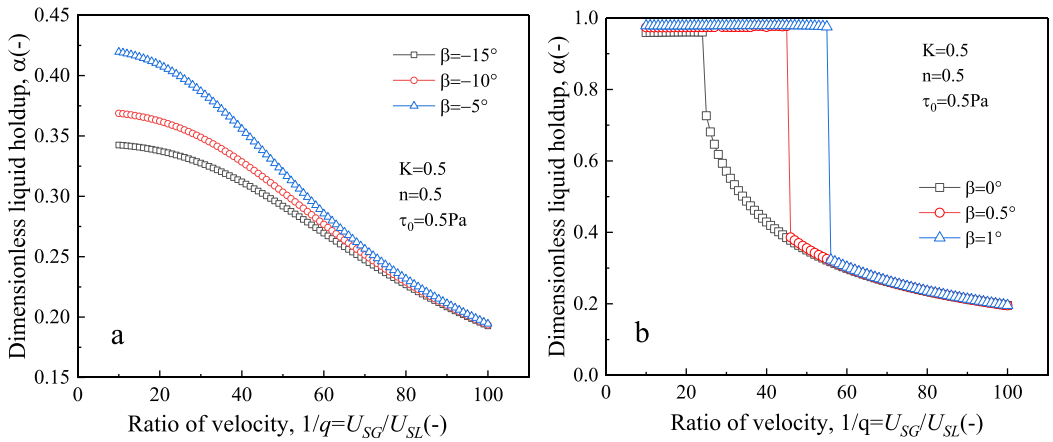

Figure 17. Holdup curve for horizontal and inclined stratified flow: (a) downward flow and (b) horizontal and upward flow.

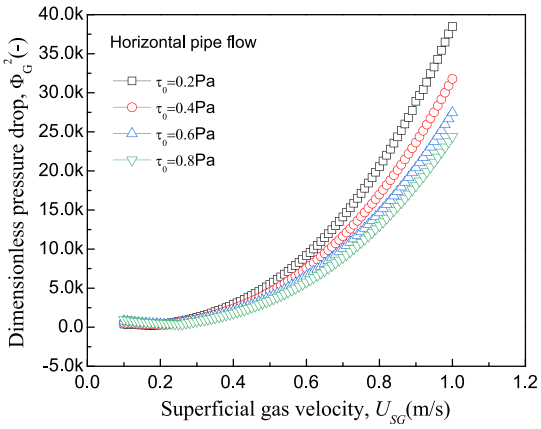

Figure 18. Relationship between the superficial gas velocity and dimensionless pressure drop for different yield stress values of Herschel−Bulkley liquid.

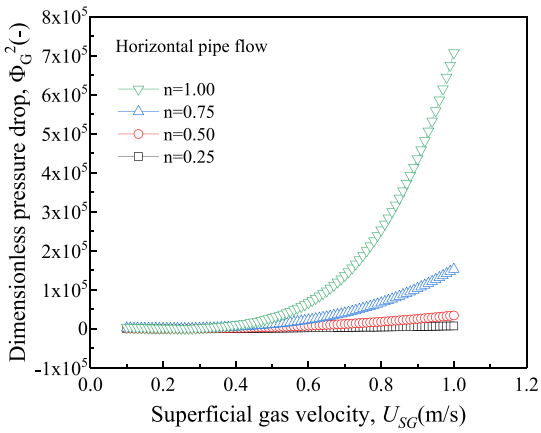

Figure 19. Variation of dimensionless pressure drop with superficial gas velocity at different power-law index values.

constant, the dimensionless pressure drop increases accordingly as the Bingham number increases.

3.2.2. Inclined Pipe Flow. Figure 21 shows the effect of the yield stress, τ_0, the power-law behavior index, n, and inclination angle, β, on the dimensionless pressure gradient in a stratified flow as predicted from eq 26. A laminar gas and liquid flow were assumed. It can be seen in Figure 21a that the dimensionless pressure gradient decreases as τ_0 increases, but for a high value of q, the effects of liquid phase properties on Φ_G^2 may be negligible. Because the gravitational and frictional terms in the total pressure gradient equation have opposite signs for downward inclined flows, two-phase flows may experience either pressure-gain or pressure-loss, depending upon the physical properties, input fluxes of the two phases, and the size and orientation of the pipe. Figure 21b illustrates the effects of the power-law behavior index, n, on drag reduction in a stratified downward flow as predicted from eq

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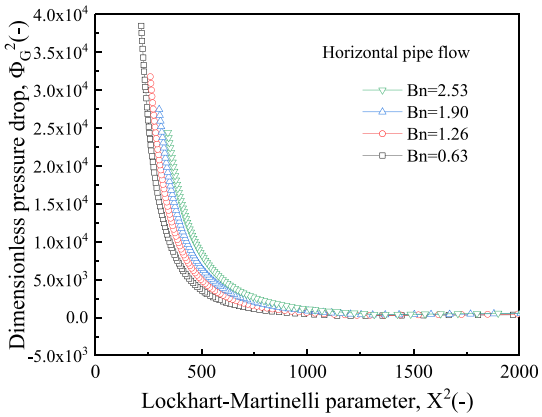

Figure 20. Dimensionless pressure drop for Herschel–Bulkley liquids as a function of the Lockhart–Martinelli parameter, X^2, in stratified horizontal flow.

26. The dimensionless frictional pressure gradient, Φ_G^2, as a function of the ratio of superficial liquid and gas velocity, q, for stratified flow of non-Newtonian liquids increases for a given q as n is increased from 0.25 to 1 over the range of $0.01 < q < 0.1$.

The effects of pipe inclination angles, β, on the dimensionless frictional pressure gradient in inclined stratified flow are shown in Figure 21c,d. For a downward inclined flow ($\beta < 0$), Φ_G^2 is generally much higher due to the increasing superficial gas velocity, and Φ_G^2 increases with the increasing inclination angle due to the hydrostatic term being positive. However, for an upward flow, there are still multiple solutions as shown in Figure 17b, and the solution obtained, when the gas–liquid ratio is relatively small, should be discarded to apply to the gas/Herschel–Bulkley fluid stratified flow.

4. CONCLUSIONS

Laminar solutions of stratified flow for horizontal and inclined pipes, in cases where one of the two-phase is a non-Newtonian shear thinning fluid, are presented. The Herschel–Bulkley constitutive model is considered to model the rheology of the shear thinning fluid due to its capability to describe the power-law and yield properties of liquids. An algorithm based on iterative calculation is presented to obtain the laminar solution for all the feasible flow conditions. The main purpose of this work was to investigate the effect of the non-Newtonian liquid constitutive relations on two-phase flow characteristics while referring to cases relevant to gas–liquid applications.

The results are discussed in terms of holdup curves and dimensionless pressure gradient to demonstrate the effect of the constitutive relations of non-Newtonian liquids on the two-phase flow characteristics. A mechanistic model for predicting the liquid holdup and pressure gradient for gas–liquid flow in horizontal and inclined pipes has been developed and verified

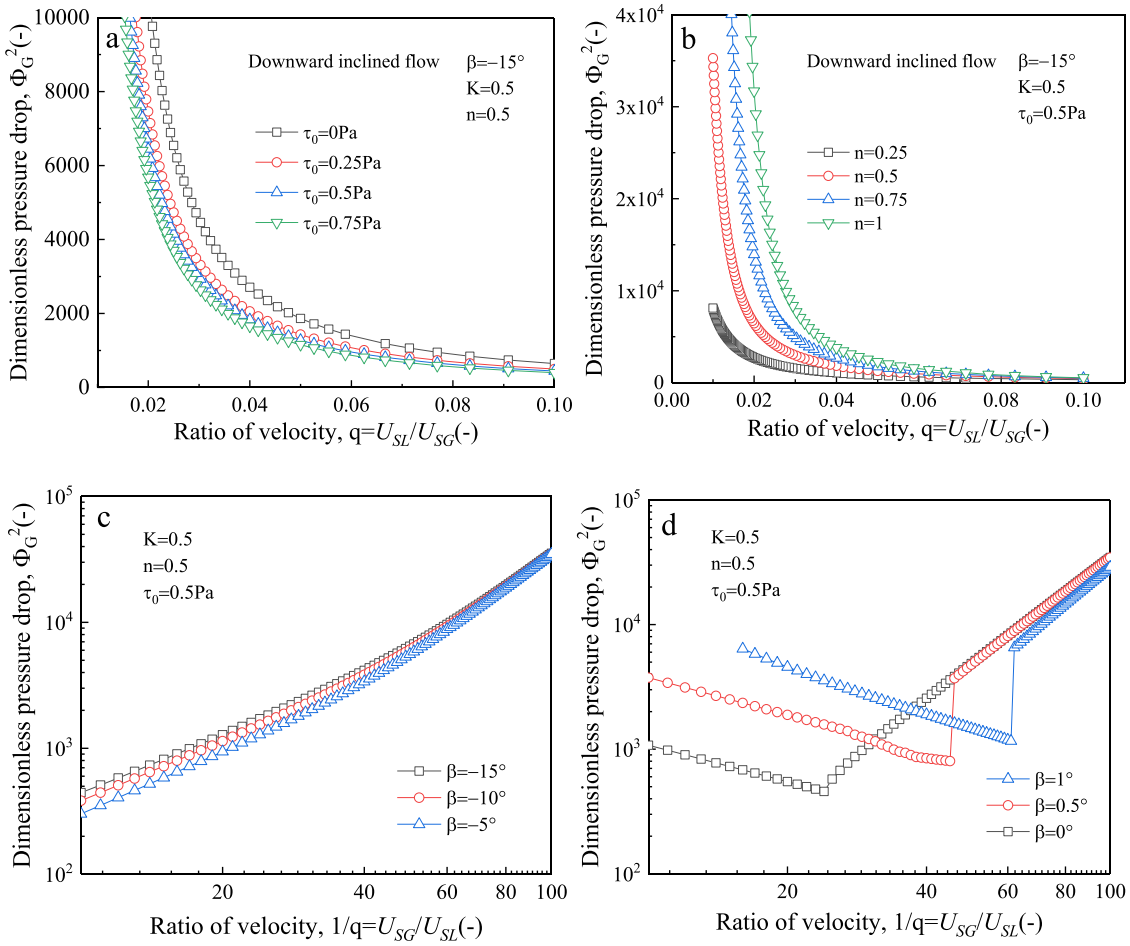

Figure 21. Dimensionless two-phase pressure gradient, Φ_G^2, as a function of the ratio of superficial liquid and gas velocity, q, for different values of (a) yield stress, τ_0, (b) power-law index, n, and (c, d) inclination.

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with experimental measurements. The results show that fluid physical properties, flow conditions, and pipe inclination angles have significant effects on the liquid holdup and pressure gradient of gas–liquid stratified flow. The influence of power-law index, n, and yield stress, τ_0, which characterize the rheological properties of liquid phase, on the flow of gas–liquid two-phase stratified pipe is noteworthy. The increase in shear thinning and yield properties will lead to the increase in liquid holdup and decrease in pressure drop.

There is a positive correlation between the superficial liquid velocity and the liquid holdup of the two-phase stratified pipe flow, and the effect is much greater than that of the superficial gas velocity. In a downward inclined two-phase stratified pipe flow, as the angle increases, the liquid holdup decreases but the friction pressure drop increases. In a horizontal pipe and upward inclined two-phase stratified flow, the effect of angle is almost negligible as the superficial gas velocity increases. When the gas–liquid ratio changes beyond the stratified flow range, a non-stratified flow will occur. The stratified flow prediction model derived in this work is not applicable, resulting in a sudden breakthrough in liquid holdup and pressure drop curves. The calculation model proposed in this work has a good prediction effect and provides new insights into the flow behavior of gas and waxy heavy oil with yield stress.

■ APPENDIX

Evolutionary Conditions of Herschel–Bulkley Fluid

There are three velocity distributions of Herschel–Bulkley fluid flowing in the pipeline: plunger flow distribution, plug-shaped + velocity ladder distribution, and laminar flow velocity distribution (Figure A.1).

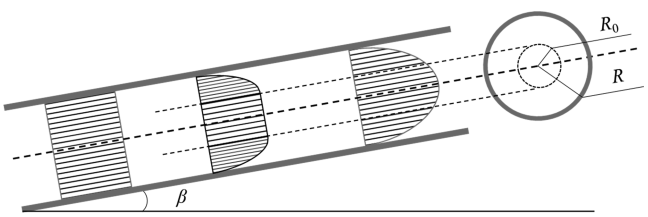

Figure A.1. Velocity distribution diagram of three stages of Herschel–Bulkley fluid.

Appendix 1. The plunger flow is essentially a shear-free flow, so the constitutive equation can be expressed as

$$f(\tau) = 0 \tag{A.1}$$

The condition for satisfying this flow state is that the velocity distribution v_z should be constant:

$$v_z' = 0 \tag{A.2}$$

$$R_0 = R \tag{A.3}$$

$$\phi = 1 \tag{A.4}$$

The speed distribution at this time is

$$v_z = \frac{n}{n+1}\left(\frac{\Delta p}{2KL}\right)^{1/n}[(R-r)^{(n+1)/n} - (r-R)^{(n+1)/n}] \tag{A.5}$$

The pressure gradient is

$$\frac{\Delta p}{L} = \frac{2\tau_0}{R_0} \tag{A.6}$$

Substitute eq A.6 into eq A.5 to get

$$v_z = \frac{n}{n+1}\left(\frac{\tau_0}{R_0}\right)^{1/n}\left(\frac{1}{K}\right)^{1/n}[(R-r)^{(n+1)/n} - (r-R)^{(n+1)/n}] \tag{A.7}$$

$$v_z' = \left(\frac{\tau_0}{R_0}\right)^{1/n}\left(\frac{1}{K}\right)^{1/n}[(R-r)^{1/n} + (r-R)^{1/n}] \tag{A.8}$$

$$[(R-r)^{1/n} + (r-R)^{1/n}] = 0 \tag{A.9}$$

It can be concluded that the value of n is the main factor that determines the occurrence of slug flow.

When $n < 1$, it is represented by shear thinning fluid. The condition for a complete plug flow velocity distribution is that $1/n$ is not an even number.

When $n > 1$, it is represented by shear thickening fluid. The condition for a complete plug flow velocity distribution is that n is not an even number.

2. The flow state with the velocity distribution types in the slug area and the speed ladder area is the part considered in this study:

$$0 < \phi < 1 \tag{A.10}$$

$$\phi = \frac{R_0}{R} = \frac{2\tau_0}{\frac{\Delta p}{L}R} = \frac{2\tau_0}{R}\frac{D}{4K}\left(\frac{4n}{3n+1}\right)^n(1-\phi)(1-a\phi-b\phi^2-c\phi^3)^n\left(\frac{D}{8U_{SL}}\right)^n \tag{A.11}$$

$$\frac{\phi}{(1-\phi)(1-a\phi-b\phi^2-c\phi^3)^n} = \frac{2\tau_0}{R}\frac{D}{4K}\left(\frac{4n}{3n+1}\right)^n\left(\frac{D}{8U_{SL}}\right)^n \tag{A.12}$$

Parameters a, b, and c in the formula are calculated as follows:

$$a = \frac{1}{2n+1} \tag{A.13}$$

$$b = \frac{2n}{(n+1)(2n+1)} \tag{A.14}$$

$$c = \frac{2n^2}{(n+1)(2n+1)} \tag{A.15}$$

Equation A.12 can be reduced to a functional relationship like eq A.16:

$$\phi = f(U_{SL}) \tag{A.16}$$

Need to guarantee: if $0 < f(U_{SL}) < 1$, then the range of liquid phase velocity is obtained, which is the applicable condition for this case.

3. Laminar flow velocity distribution means that the flow core disappears (no plunger). The following conditions must be met: $R_0 = 0$, $\phi = \frac{R_0}{R} = 0$.

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The velocity distribution can be simplified as

$$v_z = \frac{n}{n+1}\left(\frac{\Delta p}{2KL}\right)^{1/n}[(R-r)^{(n+1)/n} - r^{(n+1)/n}] \tag{A.17}$$

$$\frac{\Delta p}{L} = \frac{4}{D}K\left(\frac{3n+1}{4n}\right)^n\left(\frac{8U_{SL}}{D}\right)^n \tag{A.18}$$

$$v_z = \frac{n}{n+1}\left(\frac{1}{2K}\right)^{1/n}\left[\frac{4}{D}K\left(\frac{3n+1}{4n}\right)^n\left(\frac{8U_{SL}}{D}\right)^n\right]^{1/n} [(R-r)^{(n+1)/n} - r^{(n+1)/n}] \tag{A.19}$$

$$v_z = \frac{(n+1)(3n+1)}{2^{(1/n)-1}}D^{-(1+(1/n))}[(R-r)^{(n+1)/n} - r^{(n+1)/n}]U_{SL} \tag{A.20}$$

when $r = 0$, there is

$$v_z = v_{z\max} = (n+1)(3n+1)\left(\frac{D}{2}\right)^{2/n}U_{SL} \tag{A.21}$$

To ensure that the flow core disappears, the condition to be met is that the Bingham number should approach 0, which means

$$\frac{\tau_0}{K\left(\frac{U_{SL}}{D}\right)^n} \to 0 \tag{A.22}$$

$$\tau_0 \ll \frac{1}{1000}K\left(\frac{U_{SL}}{D}\right)^n \tag{A.23}$$

The condition for simplifying to ensure that the flow core disappears is

$$U_{SL}{}^n \gg \frac{1000D^n\tau_0}{K} \tag{A.24}$$

AUTHOR INFORMATION

Corresponding Author

Jing-yu Xu − *Institute of Mechanics, Chinese Academy of Sciences, Beijing 100190, China; School of Engineering Sciences, University of Chinese Academy of Sciences, Beijing 100049, China;* orcid.org/0000-0002-1058-2257; Phone: 0086-10-82544179; Email: xujingyu@imech.ac.cn

Authors

Lin-tong Hou − *Institute of Mechanics, Chinese Academy of Sciences, Beijing 100190, China; School of Engineering Sciences, University of Chinese Academy of Sciences, Beijing 100049, China*

Shuo Liu − *Institute of Mechanics, Chinese Academy of Sciences, Beijing 100190, China*

Jian Zhang − *Institute of Mechanics, Chinese Academy of Sciences, Beijing 100190, China*

Complete contact information is available at: https://pubs.acs.org/10.1021/acsomega.0c02482

Author Contributions

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Notes

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NOMENCLATURE

A cross-sectional area, m^2
D pipe diameter, m
S wetted parameter, m
U mean velocity, m/s
u phase velocity, m/s
h_L liquid height, m
P pressure, Pa
Q flow rate
Re Reynolds number
Bn Bingham number
f friction factor
K consistency coefficient
n power-law index
R_0 plug-flow radius, m
g acceleration of gravity, m/s^2
x axial distance, m
X^2 Lockhart−Martinelli parameter

Greek Letters

ρ density
μ viscosity
α liquid holdup
τ shear stress
β pipe inclination angle, rad

Subscripts

G gas
SG superficial gas
L liquid
SL superficial liquid
TP two phase
i interface
plug liquid plug-flow zone
ann liquid-velocity-gradient zone

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