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Experimental and Modeling Studies for Gas-liquid Two-phase Flow at High Pressure Conditions

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Experimental and model studies were performed on two-phase flow behavior at high-pressure conditions. The experiments were conducted using nitrogen and water in a test loop of 106.4 mm diameter pipe with inclination angles of 0°, 1°, and 3° at 2060 kPa. The liquid holdup data of 81 runs for each inclination angle were analyzed to identify the flow pattern.

The mechanistic model developed for low pressures was modified for high-pressure conditions. The model first detects the flow pattern, and then calculates liquid holdup and pressure drop based on the flow pattern. For dispersed-bubble flow, the critical bubble size mechanisms were also applicable at high pressures to predict a flow region in the flow pattern map, and the slip model of liquid holdup showed better matches with the experimental data than the non-slip model. For stratified flow, the flow region in the flow pattern map extended to higher liquid flow rates than at low pressures. Sequential application of the Taitel–Dukler and Bendiksen–Espedal criteria could correctly identify the stratified and non-stratified flow transition, and the Lockhart–Martinelli correlation based on the shear stresses could evaluate the liquid holdup much better than the common correlation based on the material balance. Elongated-bubble flow changed directly into dispersed-bubble flow as the liquid flow rate increases. Excellent performance of the model was demonstrated by error analyses of liquid holdup and pressure drop calculations.

Keywords

High pressure condition, Large diameter pipe, Flow pattern, Pressure drop, Liquid holdup, Mechanistic model

1. Introduction

Two-phase flow is commonly encountered in the production and transportation of oil and gas in the petroleum industry. Similar conditions are found in the chemical and nuclear industries. Accurate prediction of liquid-holdup distributions and pressure drops are extremely important for the optimum design of pipeline systems in these industries. For a two-phase mixture of gas and liquid flowing together in a channel, different internal flow geometries or structures can occur depending on the flow rates of the two phases and the geometrical variables of the pipes, together with the flow conditions and physical properties of the phases. The different interfacial structures are called flow patterns or flow regimes. Knowledge of the flow patterns that could occur under a given set of conditions leads to better predictions of two-phase flow

behavior.

The two-phase flow study has a long history, and much recent effort has been directed toward the development of general mechanistic models for steady-state two-phase flow. These models should be valid for all inclination angles, and have been evaluated with a large number of experimental data. However, some of the empirical correlations used in the models were derived from data acquired at small-scale facilities, and have concentrated on the effects of the inclination angle on the flow-pattern transition boundaries at low-pressure conditions, mostly less than 980 kPa, and in small-diameter pipelines of 2 inches. Little study of two-phase flow has been conducted in large-diameter pipes at pressure conditions higher than 1961 kPa (20 kgf/cm²) except for a model evaluation of flow in 2-inch diameter pipes at high pressures (1379 kPa and 3100 kPa)¹⁾.

The present study acquired new experimental data of liquid holdup and pressure drop for different flow patterns at high pressures, using horizontal and slightly

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Fig. 1 Schematic Diagram of Two-phase Flow Test Facility

upward-inclined, large-diameter pipelines, and developed a mechanistic model to predict flow pattern, liquid holdup and pressure drop. Since different flow mechanisms are dominant in different flow patterns, correct identification of the flow pattern is the basis for prediction of liquid holdup and pressure drop as in the case of low pressures.

2. Experimental Study

2.1. Test Facility

The main objective of the experimental program was to acquire data of gas-liquid two-phase flow in slightly inclined, large diameter pipes at high pressures. The large-scale multiphase flow test facility of Japan National Oil Corporation at Kashiwazaki was used for this purpose. A schematic diagram of the test facility is shown in Fig. 1.

Nitrogen and water were used as test fluids of the gas and liquid phases, respectively. The gas phase was supplied into the 152.4 mm (6 inch) gas supply line by a two-stage compressor, and the water phase stored in a steel tank of 27.5 m³ capacity was supplied by a pump into the 203.2 mm (8 inch) water supply line. The mixing tee mixes gas and liquid, and the mixture was injected into the test lines of 106.4 mm (4.19 inch) ID and 100 m total length. Each test line consists of an upstream horizontal inlet section of 32.5 m length, an upward slightly-inclined test section of 20 m length, a slightly downward inclined section of 20 m length, and a 27.5 m downstream section. The inclined angle of the slightly inclined section was changed to 0°, 1° and

3° for this study. The gas-water two-phase mixture flowed into the separator at the end of the test lines. After separation, gas was returned to the compressor and water to the water storage tank.

2.2. Test Program

Eighty-one runs were conducted for each inclined angle. Liquid holdup, pressure, temperature, and gas and liquid flow rates were measured during each run. Gas and liquid flow rates were converted to the superficial gas and liquid velocities using 2060 kPa (21 kgf/cm²) pressure and atmospheric temperature. Seven conductance-type high-pressure sensors were installed on the upstream horizontal section and the test section for measurements of the average liquid holdup. The conductance sensor consists of a pair of ring electrodes mounted on a short section of an opaque plastic pipe which measures the conductance of the fluid between the electrodes that is proportional to the liquid holdup. A total of 3800 holdup measurements were acquired at a sampling rate of 120/s for each case. The electronic diaphragm-type variable-reluctance sensor was used to measure pressures and pressure drops at the test section. The test section of 20 m length allowed the flow to be developed and stabilized. This was followed by two pressure-drop sections of 10 m and 7.7 m lengths. The pressure drops were measured at fixed intervals *via* pressure taps placed along the test section. The pressure in the test section was adjusted using a control valve located on the gas outlet line from the separator.

2.3. Data Analysis and Flow Pattern Determination

A TEXELA™ computer data-acquisition system was

Fig. 2 Typical Liquid Holdup Signal of Dispersed Bubble Flow, Horizontal Pipe

used to generate experimental data sets with unique characteristics. Information on the flow structure was extracted from these data sets. For pipe flow at low pressure, the definition of the flow pattern can usually be based on visual observation of the gas-liquid distribution within the pipe through a transparent plastic section. However, for high-pressure conditions, the flow cannot be visually observed. Therefore, the flow patterns were determined on the basis of changes in the void fraction or the liquid holdup within the cross section of the pipe. The flow patterns that occurred in the experiments are classified as below.

2.3.1. Dispersed Bubble Flow

Numerous tiny bubbles are transported by the liquid phase with little relative motion between the two phases. **Figure 2** shows typical signals for horizontal dispersed-bubble flow at three gas flow rates with a common liquid flow rate of $v_{SL} = 5.63$ m/s. At relatively low gas rates, bubbles are located near the top of the pipe due to gravity segregation as shown in **Fig. 2 (a)**. When the gas flow rate increased, the bubbles are dispersed more uniformly as shown in **Figs. 2 (b) and (c)**. The continuous phase of conducting liquid contains many dispersed bubbles that hit the probes at high frequency and cause very crowded pulses. When the pipe is inclined at 1° and 3° upward, the gas bubbles tend to flow in a smaller part of the upper cross-section than for the horizontal case.

2.3.2. Stratified Flow

Stratified flow occurs at relatively low gas and liquid flow rates. Low flow rates of gas and liquid result in two phases separated by gravity, *i.e.* the liquid phase flows at the bottom of the pipe and the gas phase at the top. This flow pattern can be further sub-classified into stratified-smooth, stratified-wave, and roll-wave flow types. In stratified-smooth flow, the gas-liquid interface is smooth. Stratified-wave flow occurs at relatively higher gas rates, at which stable waves form on the interface. In roll-wave flow, more liquid climbs up the pipe wall, and significantly concave-down curvature occurs at the gas-liquid interface. The stratified flow pattern can be extended with the liquid level lower than the pipe centerline at low pressures in small diameter pipes. However, the liquid holdup data as shown in **Fig. 3 (b)** demonstrated stratified flow with the liquid level higher than the centerline of the pipe. Since the experiments were conducted at high-pressure conditions, the wave cannot form liquid-bridges due possibly to pressure from the upper part, which made stratified flow occur at higher liquid flow rates than at low-pressure conditions.

Figure 3 displays typical liquid holdup data obtained for horizontal flow. **Figures 3 (a) and (c)** show typical signals for smooth stratified flow that are characterized by smooth and flat surfaces of the liquid layer generated by low rates of gas and liquid. **Figures 3 (b) and (d)** show typical signals for stratified-wave flow or roll-wave flow that are characterized by rugged and wavy surfaces. Since the flow could not be visually observed, it is difficult to identify the difference between stratified-wave and roll-wave only by the changes in the void fraction or the liquid holdup within the cross section of the pipe. **Figures 3 (e) and (f)** show signals in the transition region near the annular flow. At these experimental conditions, since it was difficult to distinguish stratified flow and annular flow in this flow region, the flow pattern was determined on the basis of the criteria in the model.

Usually, stratified flow is most sensitive to the inclination angle. Even small inclinations from the horizon have major effects on the transition from the stratified to the intermittent or annular pattern. **Figures 4 (a) and (b)** show signals for stratified flow or annular flow at 1° upward inclination, with increasing liquid flow rate and about the same gas rate of $v_{SG} = 3.10$ m/s.

2.3.3. Intermittent Flow

The intermittent flow pattern is characterized by alternate flow of liquid and gas. Slugs of liquid separated by gas pockets fill the entire pipe cross-sectional area. The liquid slugs also contain small bubbles. The intermittent pattern is usually subdivided into elongated-bubble, slug, and froth flow patterns. In elongated bubble flow, plugs of gas, formed by coalescing many bubbles, flow in a continuous liquid phase.

Fig. 3 Typical Liquid Holdup Signal of Stratified Flow, Horizontal Pipe

When the flow is calm, the plugs tend to travel in the upper half of the flow channel. **Figure 5** shows typical signals for elongated-bubble flow in the horizontal and inclined cases. The horizontal case, **Fig. 5 (a)**, has elongated bubbles of varying sizes flowing along the top of the channel. Such elongated bubbles may have been caused by the low gas flow rate and high

Fig. 4 Typical Liquid Holdup Signal of Stratified Flow, 1° Inclined Pipe

flow pressure in the pipe. The liquid bridges are separated with elongated gas bubbles less frequently than in **Figs. 5 (b) and (c)**, which show the inclined cases, where the upper surface of the pipe remains wet during gas passage. When liquid slugs contain very few entrained bubbles, the flow tends to become slug flow. If the liquid flow rate increases in the inclined cases, liquid slugs tend to merge allowing these elongated gas bubbles to flow at the top of the pipe as shown in **Figs. 5 (d) and (e)**.

Slug flow is one of the most common patterns encountered in pipelines. At a high flow rate, when the liquid is aerated and contains gas bubbles, the flow pattern becomes slug flow. A complete cycle of the flow pattern, called a slug unit, consists of the slug body and the film or gas pocket zone. The slug body also contains a mixing zone. Flow in the slug body can be assumed to be homogeneous flow. The signals shown in **Fig. 6** correspond to slug flow in the horizontal and inclined cases.

Figure 7 shows typical signals for froth flow in the horizontal and inclined cases, where the liquid holdup in the slug body is lower than 70% and the film height is lower than slug flow. The cause is more gas occupying the slug body, and the slug body begins to lose its shape, or the liquid in the slug body will start to flood the gas plug region. This flow pattern occurs in the intermediate region between slug flow and annular or dispersed-bubble flow.

2.3.4. Annular Flow

The annular flow pattern occurs at very high gas flow rates. The gas flows along the center of the pipe in annular flow. The liquid flows partially as a thin film around the walls of the pipe, and partially as droplets in the central gas core. At a high gas flow rate, more liquid is dispersed in the core, leaving a very

Fig. 5 Typical Liquid Holdup Signal of Elongated Bubble Flow

thin liquid film flowing along the wall. In the experiments, the flow types that resembled annular flow were generated in the transition region from stratified flow to annular flow as shown in Figs. 3 (e) and (f) and Figs. 4 (a) and (b). However, it was difficult to discriminate annular flow based only on the experimental data.

3. Modeling Flow-pattern Transitions

Understanding the flow mechanisms and the internal geometry is very important for modeling two-phase flow. Different flow patterns are characterized by the relative location of interfaces between the phases. As the pressure drop also changes with differences in the

Fig. 6 Typical Liquid Holdup Signal of Slug Flow

Fig. 7 Typical Liquid Holdup Signal of Froth Flow

flow pattern, correct identification of the flow pattern is crucial for accurate estimation of pressure drops as well as liquid holdup. Four major flow patterns encoun-

tered in the experiments are modeled in the following sections.

3.1. Dispersed-bubble Flow

The boundary of dispersed-bubble flow is located at higher liquid rates, where the turbulent forces overcome the interfacial tension dispersing the gas phase into small bubbles transported by the liquid phase. At this transition boundary, the maximum bubble size in the flow region, $d_{\max}$, and the critical bubble sizes, d_{CD} and d_{CB} , are given by the following equations²⁾:

$$d_{\max} = (0.725 + 4.15\alpha^{0.5}) \left(\frac{\sigma}{\rho_L} \right)^{\frac{3}{5}} \left(\frac{2f_M v_M^3}{d} \right)^{-\frac{2}{5}} \quad (1)$$

$$d_{CD} = 2 \left[\frac{0.4\sigma}{(\rho_L - \rho_G)g} \right]^{0.5} \quad (2)$$

$$d_{CB} = \frac{3}{8} \frac{\rho_L}{(\rho_L - \rho_G)} \frac{f_M v_M^2}{g \cos \theta} \quad (3)$$

Two mechanisms, agglomeration of large distorted bubbles and migration of bubbles to the upper part of the pipe due to buoyancy can be considered. If the maximum bubble size becomes greater than the critical bubble size, d_{CD} , the turbulent breakup process cannot prevent the bubbles from agglomerating. The latter critical bubble size, d_{CB} , is based on the balance of buoyant force and turbulent force on a single bubble. For the model in this study, the transition boundary was predicted by conditions such that the maximum bubble size, $d_{\max}$, calculated by Eq. (1), which is valid for a gas void fraction of less than 0.52, is smaller than the critical bubble sizes, d_{CD} .

3.2. Stratified Flow

Using sequentially two criteria proposed by Taitel and Dukler³⁾ and Bendiksen and Espedal⁴⁾, Abdouy et al.⁵⁾ have obtained good results for identifying the stratified flow pattern and its transition at low pressure conditions in 54.9 mm (2.16 inch) ID pipes. These criteria also provide satisfactory agreements with experimental results for high pressures and a large diameter 106.4 mm (4.19 inch) ID pipe as shown in Fig. 8.

The dimensionless liquid level, h_L/d , is an important parameter for the criteria. This value can be obtained from solution of the momentum balance equations for the gas and the liquid phases as

$$-A_L \frac{dp}{dL} - \tau_L S_L + \tau_I S_I - \rho_L A_L g \sin \theta = 0 \quad (4)$$

$$-A_G \frac{dp}{dL} - \tau_G S_G - \tau_I S_I - \rho_G A_G g \sin \theta = 0 \quad (5)$$

Eliminating the pressure gradient from these equations, the combined momentum equation is given as follows:

$$\tau_L \frac{S_L}{A_L} - \tau_G \left[\frac{S_G}{A_G} + \frac{\tau_I}{\tau_G} \left(\frac{S_I}{A_G} + \frac{S_I}{A_L} \right) \right] + (\rho_L - \rho_G) g \sin \theta = 0 \quad (6)$$

The wall shear stress for each phase and the interfacial shear stress are determined by

Fig. 8 Flow Pattern Map for Nitrogen-Water System in 106.4 mm ID Horizontal Pipe at 2060 kPa

$$\tau_G = f_G \frac{\rho_G v_G^2}{2} \quad (7)$$

$$\tau_L = f_L \frac{\rho_L v_L^2}{2} \quad (8)$$

$$\tau_I = f_I \frac{\rho_G (v_G - v_L)^2}{2} \quad (9)$$

where f_G , f_L , and f_I are the gas-wall friction factor, liquid-wall friction factor and interfacial friction factor, respectively. The gas-wall friction factor was evaluated by the Vlachos *et al.*⁶⁾ correlation, whereas the correlation proposed by Liang-Biao and Aziz⁷⁾ was used for the liquid-wall friction factor.

$$f_G = 0.046 N_{ReG}^{-0.2} \quad (10)$$

$$f_L = \frac{1.6291}{N_{ReL}^{0.5161}} \left(\frac{v_{SG}}{v_{SL}} \right)^{0.0926} \quad (11)$$

As the interfacial friction factor greatly influences the transition boundary between stratified and intermittent flow, several correlations were necessary for satisfactory results. When $N_{ReG} \geq N_{ReG,t}$, the following correlation according to Paras *et al.*⁸⁾ was used:

$$f_I = 0.022 \quad (12)$$

When $N_{ReG} < N_{ReG,t}$, the Ellis and Gay⁹⁾ correlation as the function of the gas-phase Reynolds number was used.

$$f_I = 1.3 N_{ReG}^{-0.57} \quad (13)$$

If the value of f_I calculated by Eq. (13) was smaller than f_G , the value of f_G was used for f_I . For the upward inclined flow, the constant interfacial friction factor proposed by Amaravadi *et al.*¹⁰⁾ was used.

$$f_i = 0.030 \quad (14)$$

For the critical gas Reynolds number, $N_{ReG,t}$, the following correlation proposed by Lee and Bankoff⁽¹¹⁾ was used:

$$N_{ReG,t} = 1.837 \times 10^5 N_{ReL}^{-0.184} \quad (15)$$

After solving Eq. (6) for the implicit variable, h_L/d , the boundary between the stratified and the non-stratified flow was determined by the Taitel-Dukler criterion^(3),5) as given in **Appendix A**. If stratified flow was identified, we applied another criterion proposed by Bendiksen and Espedal^(4),5).

$$v_{sF} \geq v_{sT} \quad (16)$$

This condition considers that if the front velocity of a slug, v_{sF} , becomes greater than the translational (tail) velocity, v_{sT} , stable slug flow is initiated to form a large roll-wave that is developed into the slug body. The translational velocity is given by Bendiksen⁽¹²⁾ as

$$v_{sT} = C_o v_M + (0.35 \sin \theta + 0.54 \cos \theta) \sqrt{\frac{g(\rho_L - \rho_G)d}{\rho_L}} \quad (17)$$

The distribution parameter, C_o , was calculated by the correlation of Petalas and Aziz⁽¹³⁾.

$$C_o = (1.64 + 0.12 \sin \theta) N_{ReL}^{-0.031} \quad (18)$$

The Reynolds number, N_{ReL} , in Eq. (18) is based on the mixture velocity, liquid viscosity and density. The front velocity, v_{sF} , can be obtained from the mass balance equation for a slug unit, as given in **Appendix B**.

3.3. Intermittent Flow

Since the transitional region between intermittent flow and dispersed-bubble flow was wider than in the case of low-pressure conditions as shown in **Figs. 8** through **10**, the following condition was implemented for calculations of the intermittent flow: if $L_s/L_U > 1$, then the dispersed bubble flow model was used. The slug body length, L_s , was calculated by an empirical correlation⁽¹⁾, and the slug unit length, L_U , is calculated by the mass balance equation as

$$\ln(L_s) = -26.6 + 28.5[\ln(d) + 3.67]^{0.1} \quad (19)$$

$$L_U = L_s \frac{v_L H_s - v_F H_F}{v_{SL} - v_F H_F} \quad (20)$$

The intermittent flow model used in this study included the elongated-bubble, slug and froth flow patterns. For the transition boundary between the elongated-bubble and slug flow, based on the quantity of the gas bubbles contained in a slug body, assuming that the liquid holdup of slug body is equal to 0.98 on the boundary provides good results⁽⁵⁾. However, this criterion is not appropriate for the present experiments at high-pressure conditions. A possible reason is the presence of two types of elongated-bubble flow as shown in **Fig. 5**. The first type, which is often observed at low-pressure conditions, occurred at low gas and liquid velocities in the inclined pipe as shown in **Figs. 5 (b) and (c)**. For this type, the transition boundary between the elongated-bubble and slug flow

can be determined by the quantity of the gas bubbles included in slug body. The second type occurred when the flow is horizontal, or when the flow was inclined with higher liquid flow rate, and the flow changed to a form as shown in **Fig. 5 (a)** in the horizontal case, or **Figs. 5 (d) and (e)** in the inclined case. These three figures demonstrate that gas bubbles tend to flow at the upper part of the pipe. In the second type, the transition boundary cannot be determined by the quantity of the gas bubbles in slug body. Instead, the minimum value of the average liquid level, h_L/d , was evaluated for the experimental results of the both flow types above, and the value of 0.85 was found to generate good correlation with the results from the liquid holdup data. Therefore, the transition boundary between the elongated-bubble and slug flow was determined as below.

$$\frac{h_L}{d} = 0.85 \quad (21)$$

After determining the boundary between the elongated-bubble and slug flow, the transition to annular flow was determined using the following empirical criterion⁽⁵⁾

$$\frac{h_L}{d} \leq 0.25 \quad (22)$$

Then, the transition from slug to froth flow pattern was calculated by empirical correlations as follows:

$$H_s < 0.48 \text{ or } \frac{h_L}{d} \leq 0.45 \quad (23)$$

4. Calculations of Liquid Holdup and Pressure Drop

Calculations of the liquid holdup and pressure drop were performed on the basis of the flow patterns identified in the previous section.

4.1. Dispersed-bubble Flow

Dispersed-bubble flow occurs at high liquid flow rates. Applying the pseudo homogeneous model, the liquid holdup was evaluated by the slip model⁽¹³⁾ derived from the liquid mass balance.

$$H_L = 1 - \frac{v_{SG}}{1.2v_M + v_b} \quad (24)$$

The bubble rise velocity, v_b , depends on the angle of inclination and is given by⁽¹³⁾

$$v_b = 1.53 \left[\frac{g(\rho_L - \rho_G)\sigma}{\rho_L^2} \right]^{0.25} \sin \theta \quad (25)$$

After calculating the liquid holdup, the mixture density, ρ_M , and velocity, v_M , the pressure drop is calculated by

$$-\frac{dp}{dL} = \frac{2f_M \rho_M v_M^2}{d} + \rho_M g \sin \theta \quad (26)$$

f_M is a mixture friction factor based on the mixture Reynolds number.

4.2. Stratified Flow

In this study, the liquid holdup was first determined applying a simple relationship as

$$H_L = 1 - \alpha \quad (27)$$

The void fraction, α , can be calculated by the correlation proposed by Lockhart and Martinelli¹⁴⁾ as follows:

$$\alpha = f(x_{LM}) = (1 + x_{LM}^{0.8})^{-0.378} \quad (28)$$

x_{LM} is the Lockhart and Martinelli parameter defined as

$$x_{LM} = \sqrt{\frac{(dp/dL)_{SL}}{(dp/dL)_{SG}}} = \sqrt{\frac{4\tau_{SL}/d}{4\tau_{SG}/d}} \quad (29)$$

where shear stresses τ_{SL} and τ_{SG} are based on the superficial liquid and gas velocities, respectively.

Once the liquid holdup was determined, the pressure drop could be calculated by eliminating the interfacial shear stress using both Eqs. (4) and (5).

4.3. Intermittent Flow

Calculations for the intermittent flow were performed for slug, elongated-bubble and froth flow patterns.

Slug flow occurs over a wide range of the gas and liquid flow rates. The slug unit consists of the slug body and film zones. The slug body zone was treated as a homogeneous flow and the film zone is treated as stratified flow. The average liquid holdup of a slug unit was calculated applying the liquid mass balance as

$$H_L = \frac{H_F L_F + H_S L_S}{L_U} \quad (30)$$

where H_F and H_S are liquid holdup in the slug film zone and slug body zone, respectively, L_U is slug unit length, L_F is liquid film length, and L_S is slug body length.

Using Eqs. (C-1), (C-2), and (C-3) in **Appendix C**, the expression for the liquid holdup becomes

$$H_L = \frac{H_S v_{sT} + v_{Gdb}(1 - H_S) - v_{SG}}{v_{sT}} \quad (31)$$

where the velocity of dispersed-bubbles in slug body, v_{Gdb} , is given by¹³⁾

$$v_{Gdb} = C_o v_M + 1.53 \left[\frac{g(\rho_L - \rho_G)\sigma}{\rho_L^2} \right]^{0.25} \sin \theta \quad (32)$$

The translational velocity, v_{sT} , is determined by Eq. (17). C_o is referred to as the distribution parameter or profile parameter that is very important to correctly predict liquid holdup. Usually it is taken as $C_o = 1.2$ for turbulent flow, and 2.0 for laminar flow, although different values are also used depending on experimental conditions¹²⁾. In this study, it was determined so that the calculated values of liquid holdup matched the measured values. Therefore, $C_o = 1.0$ for the bubbles velocity in the slug body, and the empirical correlation Eq. (18) was applied for the translational velocity. The liquid holdup of the slug body, H_S , is given in **Appendix B**.

Once the average liquid holdup was determined, the average pressure drop over the slug unit was calculated as follows:

Fig. 9 Flow Pattern Map for Nitrogen-Water System in 106.4 mm ID Inclined Pipe (1°) at 2060 kPa

$$-\frac{dp}{dL} = \rho_M g \sin \theta + \frac{1}{L_U} \left[L_S \left(\frac{\tau_S \pi d}{A} \right) + L_F \left(\frac{\tau_F S_F + \tau_G S_G}{A} \right) \right] \quad (33)$$

The slug body shear stress, τ_S , gas-wall shear stress, τ_G , and liquid-wall shear stress, τ_F , were calculated based on the friction factors. The gas-wall friction factor, f_G , and liquid-wall friction factor, f_F , are calculated by Eqs. (10) and (11), respectively, and the friction factor in the slug body can be calculated by Eq. (D-1) in **Appendix D**.

For elongated-bubble and froth flow, the slug model is usually used without any modification. In this study, the slug flow model was applied to the elongated-bubble flow. For the froth flow, the distribution parameter, C_o , in Eq. (17) of the slug translational velocity was assumed as 1.05 and 1.2 for horizontal and inclined flow, respectively. These values provided the optimum correlation of the calculated liquid holdup with the measured data.

5. Results

5.1. Experiments and Flow Pattern Predictions

Experimental runs were performed using 81 flow conditions at each of the inclination angles, 0°, 1°, and 3° as shown in **Figs. 8, 9, and 10**, respectively. For each inclination angle, the flow rates of gas and liquid were varied in the range of the superficial velocities $v_{SG} = 0.10$ –3.14 m/s and $v_{SL} = 0.04$ –5.64 m/s. Each data point symbol also indicates the flow pattern identified by interpreting the holdup signals. The results of these experiments were compared with the model predictions of the flow pattern transitions shown by the lines in the figures.

Fig. 10 Flow Pattern Map for Nitrogen-Water System in 106.4 mm ID Inclined Pipe (3°) at 2060 kPa

The model prediction for the transition from intermittent to dispersed-bubble flow is based on the concept of bubble size proposed by Barea *et al.*²⁾ The applicabilities of two critical bubble sizes, d_{CB} and d_{CD} , were first evaluated. Under experimental conditions with high pressures, d_{CB} is smaller than d_{CD} , as shown in Figs. 8 through 10, where the broken lines showing the boundary reflect the criterion $d_{max} = d_{CB}$, and the solid lines reflect $d_{max} = d_{CD}$. The criterion $d_{max} = d_{CD}$ was usually effective in vertical or high inclination flow, whereas the criterion $d_{max} = d_{CB}$ was effective in horizontal or slightly inclined flow. The figures show that the broken lines obtained by the criterion $d_{max} = d_{CB}$ are located within the dispersed-bubble flow region, and that the solid lines based on $d_{max} = d_{CD}$, nearly pass through the intermittent flow region. Therefore, the single criterion $d_{max} = d_{CD}$ was adopted in model calculations for all the inclination angles.

The stratified and non-stratified flow transition estimated by the method proposed by Abdurvayt *et al.*⁵⁾ using the Bendiksen and Espedal⁴⁾ criterion in addition to the Taitel and Dukler³⁾ criterion demonstrated good agreements with the experimental data. Only the stratified roll-wave (SR) region appeared at high gas flow rates for the inclined cases (Figs. 9 and 10).

The intermittent flow pattern was subdivided into elongate-bubble, slug, and froth flow based on the experimental data. Model predictions of these three flow patterns were based on the liquid level. Most intermittent flows were identified as elongated-bubble or slug flow, and well matched the model predictions. The most notable difference from the low-pressure case⁵⁾ was the elongated-bubble region transition directly to the dispersed-bubble region. Froth flow as classified here is not a flow type like churn or froth flow in

vertical or high inclination pipes, but a type of slug flow with a higher gas rate that causes liquid slugs to be split in gas plugs. The experimental results identified as froth flow were also reasonably matched with model predictions.

Three data points showed transitional behavior between the stratified and annular flow as marked by the unfilled squares in Figs. 8 and 9. These points were located on or close to the boundary between the stratified roll-wave and annular regions predicted by the model.

5.2. Liquid Holdup and Pressure Drop Calculations

Based on the flow patterns predicted, liquid holdup and pressure drop were calculated for the individual flow conditions of the experiments. Figures 11 through 23 compare between the predictions by the proposed model and the experimental data for both the average liquid holdup and the pressure drop of all flow patterns.

Three statistical parameters, *i.e.* average percentage error, absolute average percentage error, and standard deviation were used to evaluate the predictions of the proposed model for liquid holdup and pressure drops.

$$\varepsilon_1 = \left(\sum_{i=1}^n e_i \right) / n \quad (\text{average percentage error}) \quad (34)$$

$$\varepsilon_2 = \left(\sum_{i=1}^n |e_i| \right) / n \quad (\text{absolute average percentage error}) \quad (35)$$

$$\varepsilon_3 = \left(\sum_{i=1}^n (e_i - \varepsilon_1)^2 \right)^{1/2} / (n-1)^{1/2} \quad (\text{standard deviation}) \quad (36)$$

where n is the number of data points, and the relative error, e_i , is given by

$$e_i = ((\Delta p_i)_C - (\Delta p_i)_M) \times 100 / (\Delta p_i)_M \quad (37)$$

The results of liquid holdup and pressure drops are listed for the individual flow pattern models in Table 1.

Figure 11 compares the predicted and measured liquid holdups of 54 data points for the dispersed-bubble flow at all inclination angles. The predictions of the proposed model show excellent agreements with this data set, with -1.13% average error, 1.77% absolute average error, and 2.40 standard deviation. Figures 12, 13, and 14 compare the slip and non-slip models with the measured liquid holdups against the gas flow rate for inclination angles 0°, 1°, and 3°, respectively. The liquid velocity was set constant as $v_{SL} = 3.44$ m/s for each inclination angle. Using the slip model for these experimental conditions provided better accuracy.

Figures 15 compares the predicted and measured liquid holdups of 30 data points for the stratified flow. The predictions of the proposed model show good agreements with this data set, with 2.33% average error, 3.03% absolute average error, and 19.57 standard deviation. The circle symbols display the results of

Table 1 Accuracy Evaluation of Liquid Holdup and Pressure Drop Calculations

Flow patterns	Average error ε_1 [%]	Absolute average error, ε_2 [%]	Standard deviation, ε_3	Number of data
Liquid holdup				
Dispersed bubble	-1.13	1.77	2.40	54
Stratified	-2.33	3.03	19.59	30
Intermittent	-1.20	5.96	9.33	159
All patterns	-2.56	6.28	10.25	243
Pressure drop				
Dispersed bubble	3.15	3.73	4.01	162
Intermittent	1.04	16.02	38.40	467
All patterns	1.58	12.85	33.15	629

Fig. 11 Accuracy of Liquid Holdup Calculations for Dispersed Bubble Flow, Horizontal and Inclined Pipes

Fig. 13 Comparison of Slip and Non-slip Models for Dispersed Bubble Flow, Inclined Pipe (1°)

Fig. 12 Comparison of Slip and Non-slip Models for Dispersed Bubble Flow, Horizontal Pipe

Fig. 14 Comparison of Slip and Non-slip Models for Dispersed Bubble Flow, Inclined Pipe (3°)

Fig. 15 Accuracy of Liquid Holdup Calculations for Stratified Flow, Horizontal Pipe

Fig. 17 Accuracy of Liquid Holdup Calculations for All Flow Patterns, Horizontal and Inclined Pipes

Fig. 16 Accuracy of Liquid Holdup Calculations for Intermittent Flow, Horizontal and Inclined Pipes

Fig. 18 Accuracy of Pressure Drop Calculations for Dispersed Bubble Flow, Horizontal and Inclined Pipes

the proposed model, and the triangular symbols show the results calculated by the following equation that can be derived from geometrical relationships assuming a flat surface of the liquid layer:

$$H_L = \frac{\pi - \cos^{-1}\left(2\frac{h_L}{d} - 1\right) + \left(2\frac{h_L}{d} - 1\right)\sqrt{1 - \left(2\frac{h_L}{d} - 1\right)^2}}{\pi} \quad (38)$$

Figure 16 compares the predicted and measured liquid holdups of 159 data points for the intermittent flow at all inclination angles. The predictions of the proposed model provide excellent agreements with this

data set, with -1.20% average error, 5.96% absolute average error, and 9.33 standard deviation. The elongated-bubble flow was treated as the same as the slug flow in the model. For the froth flow, the slug model was applied assuming a constant value of the distribution parameter, C_o , for the translational velocity.

The overall performance of the proposed model for evaluating the liquid holdup is depicted in **Fig. 17**, with -2.56% average error, 6.28% absolute average error, and 10.25 standard deviation.

Figure 18 compares the predicted and measured pressure drops of 162 data points for the dispersed-bub-

Fig. 19 Pressure Drop vs. Gas Velocity for Stratified Flow, $v_{SL} = 0.044$ m/s

Fig. 21 Pressure Drop vs. Gas Velocity for Stratified Flow, $v_{SL} = 0.156$ m/s

Fig. 20 Pressure Drop vs. Gas Velocity for Stratified Flow, $v_{SL} = 0.078$ m/s

ble flow at all inclination angles. Each experimental run provided 3 pressure-drop data at the 10 m and 7.7 m sections and the overall length. The model had 3.15% average error, 3.73% absolute average error, and 4.01 standard deviation.

Figures 19, 20, and 21 compare the model calculations and measured pressure drops for the stratified flow with increased the gas flow rate and constant liquid velocities, $v_{SL} = 0.044$, 0.078 , and 0.156 m/s, respectively. As seen in the figures, the experimental runs of stratified flow generated much lower pressure drops than those of other flow patterns. Under these

experimental conditions, it was difficult to obtain correct data for low gas stratified flow due to the resolution limit of the pressure transducers, since pressure drops were close to zero when the superficial gas velocity was lower than 0.5 m/s. As shown in the figures, the predicted results show good agreements with the experimental data as the superficial gas velocity increases. Since there were only a few reliable data in this case, error analysis was not performed.

For the intermittent flow, the same model based on the momentum balance was applied to the three flow patterns of elongated-bubble, slug, and froth flow. Calculations for the froth flow are sensitive to the translational velocity used for liquid holdup. Predictions of the intermittent model for pressure drop were plotted against the experimental measurements of 467 data points for all inclination angles as shown in Fig. 22 showing 1.04% average error, 16.02% absolute average error, and 38.40 standard deviation.

The overall performance of the model was evaluated against the total 629 experimental data for dispersed-bubble and intermittent flow as shown in Fig. 23 with error analysis showing average error of 1.58%, absolute error of 12.85%, and standard deviation of 33.15.

6. Conclusions

- (1) Experiments of two-phase flow in 106.4 mm ID pipes at high-pressure conditions were conducted at a large-scale test loop, and liquid holdup and pressure drop data were successfully acquired at 81 flow conditions for each of the inclination angles 0° , 1° , and 3° .
- (2) The flow pattern of each flow condition was determined by interpreting liquid holdup signals, and a flow

Fig. 22 Accuracy of Pressure Drop Calculations for Intermittent Flow, Horizontal and Inclined Pipes

Fig. 23 Accuracy of Pressure Drop Calculations for Dispersed-bubble and Intermittent Flow, Horizontal and Inclined Pipes

pattern transition map was developed. The flow pattern boundaries were also predicted using a mechanistic model.

(3) The critical bubble size mechanisms at low pressures proposed by Barnea *et al.* were also applicable at high pressures to predict the boundary between dispersed-bubble and intermittent flow. The results confirmed that only the criterion for bubble distortion $d_{\max} = d_{CD}$ can effectively discriminate between them. The liquid holdup calculated by the slip model matched the experimental data better than the non-slip model.

(4) Stratified flow at high pressures remained stable at higher liquid flow rates than at low pressures. The liquid level in high-pressure flow can be higher than the centerline of the pipe. Sequential application of the Taitel-Dukler and Bendiksen-Espedal criteria could correctly identify the stratified and non-stratified flow transition in large-diameter pipes under high-pressure conditions. For the liquid holdup evaluation, the Lockhart and Martinelli correlation based on the shear stresses was much better than the common correlation based on the material balance.

(5) Elongated-bubble flow changed directly into dispersed-bubble flow as the liquid flow rate increased, which is a notable difference from the low-pressure case. The intermittent flow region can be sub-classified on the basis of liquid level. The slug flow model of liquid holdup can be effectively applied to elongated-bubble and froth flow.

(6) The friction factors, distribution parameter, and fluid properties, which directly influence model calculations for flow pattern, liquid holdup and pressure drop, could be accurately evaluated according to flow conditions.

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Nomenclatures

A	: cross-sectional area
C_o	: velocity distribution coefficient
d	: pipe internal diameter
d_{CB}	: critical bubble size for bubble migration by buoyant force
d_{CD}	: critical bubble size for bubble agglomeration
f	: friction factor
g	: acceleration due to gravity
H	: liquid holdup
h_L	: height of liquid
L	: length
N_{Re}	: Reynolds number
p	: pressure
S	: contact perimeter
v	: velocity
v_{SG}	: superficial gas velocity
v_{SL}	: superficial liquid velocity
<Greeks>	
α	: void fraction
ε	: pipe roughness or error parameter
μ	: viscosity
θ	: angle of inclination
ρ	: density
σ	: interfacial tension
τ	: shear stress
<Subscripts>	
F	: liquid film
G	: gas phase
I	: gas/liquid interface
L	: liquid phase
M	: mixture or measured
S	: slug body

U : slug unit

Appendix A: Criterion of Transition from Stratified to Non-stratified Flow

Taitel and Dukler³⁾ proposed that if the pressure suction force is greater than the gravity force in stratified flow with wave on the gas-liquid interface, stratified flow cannot be preserved. They obtained the following criterion for this transition.

$$v_G > \left(1 - \frac{h_L}{d}\right) \frac{\sqrt{(\rho_L - \rho_G)g \cos \theta A_G}}{\sqrt{\rho_G \frac{dA_L}{dh_L}}} \quad (\text{A-1})$$

where the gradient of the cross-sectional area with respect to the level of the liquid stratum is given by

$$\frac{dA_L}{dh_L} = S_1 = d \sqrt{1 - \left(2 \frac{h_L}{d} - 1\right)^2} \quad (\text{A-2})$$

Appendix B: Slug Front Velocity

The front velocity of a slug, v_{sF} , can be obtained from the mass balance over a slug unit as

$$v_{sF} = v_G \frac{\alpha - \alpha_S \frac{v_{GS}}{v_G}}{\alpha - \alpha_S} \quad (\text{B-1})$$

where α and α_S are gas void fractions for equilibrium stratified flow and in the slug body, respectively. v_G and v_{GS} are gas velocity in equilibrium stratified flow and in the liquid slug body, respectively. α and v_G can be obtained from the momentum balance over the equilibrium stratified flow. α_S can be obtained by letting $v_{GS} = v_M$, because the slug body can be treated as a non-slip flow. The gas void fraction in the slug body, α_S , can be obtained by

$$\alpha = 1 - H_S \quad (\text{B-2})$$

where H_S is liquid holdup in the slug body proposed by Gregory *et al.*¹⁵⁾

$$H_S = \frac{1}{1 + \left(\frac{v_M}{8.66}\right)^{1.39}} \quad (\text{B-3})$$

Appendix C: Mass Balance for Intermittent Flow

An overall liquid mass balance over a slug unit gives

$$v_{SL} L_U = v_L H_S L_S + v_F H_F L_F \quad (\text{C-1})$$

where, H_S and H_F are the liquid holdups in the slug body and film zone, respectively, and L_S and L_F are the lengths of the slug body and film zone, respectively.

A mass balance can also be applied at two cross-sectional areas relative to a coordinate system moving at the translational velocity, v_{sT} , yielding

$$(v_{sT} - v_L) H_S = (v_{sT} - v_F) H_F \quad (\text{C-2})$$

As the total volumetric flow rate is constant through any cross section of the slug unit, the mixture velocities are given by slug body and the film zone cross sections, respectively, as

$$v_M = v_{SL} + v_{SG} = v_L H_S + v_{Gdb} (1 - H_S) \quad (\text{C-3})$$

$$v_M = v_F H_F + v_G (1 - H_F) \quad (\text{C-4})$$

Appendix D: Friction Factor in Slug Body

In this study the frictional factor in slug body, f_s , was calculated by the correlation proposed by Zigrang and Sylvester¹⁶⁾:

$$\frac{1}{\sqrt{f_s}} = -2 \log \left[\frac{2\varepsilon/d}{3.7} - \frac{5.02}{N_{ReM}} \log \left(\frac{2\varepsilon/d}{3.7} + \frac{13}{N_{ReM}} \right) \right] \quad (\text{D-1})$$

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要 旨

高圧条件における気液 2 相流実験およびモデル化に関する研究

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高圧条件における気液 2 相流挙動の研究として, 実験およびモデル予測を行った。実験は, 管内径 106.4 mm のテストチューブに窒素ガスと水を使用し, テスト区間の傾斜角度を 0, 1, および 3 度として実施した。各傾斜角度における 81 回の流動実験で得られた液体ホールドアップの解析によって流動様式を確定した。

低圧条件で構築したメカニスティックモデルを高圧条件に対応して修正した。モデル計算では, まず流動条件における流動様式を予測し, それに基づいて液体ホールドアップおよび圧力損失を計算する。分散気泡流については, 流動様式の予測基準として臨界気泡サイズのマニピュレーションが適用できること, 液体ホ

ールドアップの予測にはスリップモデルが実験データによく適合することが確認された。高圧条件における層状流の領域は, 低圧における場合よりも高い液体流量域に広がっている。層状流と非層状流の遷移判別には, Taitel-Dakler および Bendiksen-Espedal の基準式の逐次的適用が有効であり, Lockhart-Martinelli の相関式が層状流の液体ホールドアップ予測に有効である。高圧における伸長気泡流の流動領域は, 液体流量の増加に伴い, 分散気泡領域に直接遷移する。液体ホールドアップおよび圧力損失の計算結果に関する誤差分析によって, モデルの計算精度が高いことが確認された。

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