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Original Paper

Evaluation of frictional pressure drop correlations for air-water and air-oil two-phase flow in pipeline-riser system

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ABSTRACT

Accurate prediction of the frictional pressure drop is important for the design and operation of subsea oil and gas transporting system considering the length of the pipeline. The applicability of the correlations to pipeline-riser flow needs evaluation since the flow condition in pipeline-riser is quite different from the original data where they were derived from. In the present study, a comprehensive evaluation of 24 prevailing correlation in predicting frictional pressure drop is carried out based on experimentally measured data of air-water and air-oil two-phase flows in pipeline-riser. Experiments are performed in a system having different configuration of pipeline-riser with the inclination of the downcomer varied from -2° to -5° to investigated the effect of the elbow on the frictional pressure drop in the riser. The inlet gas velocity ranges from 0.03 to 6.2 m/s, and liquid velocity varies from 0.02 to 1.3 m/s. A total of 885 experimental data points including 782 on air-water flows and 103 on air-oil flows are obtained and used to access the prediction ability of the correlations. Comparison of the predicted results with the measured data indicate that a majority of the investigated correlations under-predict the pressure drop on severe slugging. The result of this study highlights the requirement of new method considering the effect of pipe layout on the frictional pressure drop.

1. Introduction

Gas and liquid two-phase flows are frequently encountered in a lot of applications such as oil and gas exploitation, air condition systems, thermal energy plants as well as others. Pressure drop is critical for gas-liquid flows and the determination of pressure drop for two-phase flow in tubes is needed by many design practices. In the process of offshore petroleum production and transportation, the oil and gas from production wells is usually transported through a long pipeline placed on the seabed, and then is lifted upward via a riser to the floating platforms (Chen, 2011; Pedersen et al., 2017; Zhang J.X. et al., 2022). Considering the length of the pipeline, power consumption costs of pumps usually constitute a substantial portion of operational costs for the overall pipeline transportation. Thus, a method precisely predicting pressure gradient is highly required for proper design and assessments of practical performance.

It is generally agreed that the total pressure gradient essentially consists of three components, i.e. the frictional pressure drop, the momentum pressure drop, and the static pressure drop. Among these three terms, the frictional pressure drop is the most complicated one (Lockhart and Martinelli, 1949) and has attracted much attention in the past decades. A great number of investigations on two-phase flow frictional pressure drop has been reported and many predicting correlations have been recommended. According to the method adopted, the existing correlations can be generally regarded as two types, i.e. the homogeneous based model and the separated based model. The former model ignores the interfacial slip and regards the two phases as well mixed flows with averaged properties of each phase (Cicchitti et al., 1960; Dukler et al., 1964), while the latter assumes the two phases to share common interfaces between them and move separately with different velocities (Chisholm, 1967; Sun and Mishima, 2009; Xu and Fang, 2012).

Since the actual application of the frictional pressure drop models may beyond the ranges in which they were originally developed, it is necessary to assess their applicability before

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practical application. Choi et al. (2008) compared 15 correlations with experimental data of refrigerant R-410A in 1.5 and 3 mm diameter mini-channels having smooth surface. Equations based on homogeneous based model were found to provide the frictional pressure drop with reasonable accuracy. Li and Wu (2010) tested 11 methods against 769 experimentally measured data points for various working fluids. The correlations developed by Cicchitti et al. (1960) and Muller-Steinhagen and Heck (1986) was recommended. The performance of 29 correlations were evaluated by Xu and Fang (2012) against a database containing 3480 experimental points collected from 26 open sources. The hydraulic diameters of the experimental data points range from 0.0695 to 14 mm. They concluded that the Muller-Steinhagen and Heck (1986) based on separated model, and Beattie and Whalley (1982) based on homogeneous model can predict acceptable results for experimental data of adiabatic liquid-gas. Recently, Liu et al. (2022) used 579 measured data of frictional pressure drop with oscillation to access 24 wildly used models. Large deviation was found.

For these correlations mentioned above, the properties of working fluids, pipe orientations and diameters are considered. However, most of these correlations were developed on the basis of experimental data in pipes with small diameters, ranging from 0.5 to 30 mm. Ghajar et al. (see Cheng and Ghajar, 2019) evaluated 35 correlations with 3147 data points of air-water two-phase flow frictional pressure drop in channels with diameters ranging from 9.5 to 152 mm. Cavallini et al. (2002) was recommended to calculate frictional pressure drop for gas-liquid flow in vertical downward tube. Yao et al. (2018) conducted experimental investigation on frictional pressure drop of downward air-water two-phase flow in pipes with diameters ranging from 15 to 65 mm. A total of 19 correlations were assessed against their 978 experimental data points. However, experimental data were from flows in straight pipe without a flow direction change. As the behavior of two-phase flow in the riser is greatly affected by the preceding flow in the pipeline, the characteristic of flows in pipeline-riser system is different from those in straight tube. Guo et al. (2001) studied frictional pressure drop of single-phase and steam-water two-phase in helical coiled tubes. It was found that the two-phase frictional pressure drop in helical tubes is greater than that of a straight tube due to the effect of secondary flow. The curvature effects on two-phase pressure drop in helical coil flows was further confirmed by Cioncolini and Santini (2016). They tested 25 correlations against 980 data points of two-phase water-steam flows in helically coiled tubes and found that most of the methods notably underpredicted the pressure gradient. With the rapid development of deep learning methods in nowadays, methods based on complex network was proved to be a promising solution for characterizing parameters of gas and liquid two-phase flow. An example can be seen in the literature (Gao et al., 2021).

It is obviously that many of the published correlations provide reasonable accuracy when applied to straight tubes. Some of the correlations providing acceptable predictions include that of Chisholm (1973), Sun and Mishima (2009) and Xu and Fang (2012) based on separated flow model, and that of Cavallini et al. (2002) and Shannak (2008) based on homogeneous model. Nevertheless, the accuracy needs to be evaluated when the existing frictional pressure drop correlations are applied to flows in pipeline-riser system, as the flow patterns in pipeline-riser are quite different from those occurring in pipe with one orientation. However, to the best of the authors' knowledge, the predicting performance of the present existing correlations for gas-liquid two-phase flow in pipeline-riser system has not been reported yet. The overall objective of this work is to provide fundamental understandings of frictional pressure drop of air-water two-phase flow in a pipeline-riser system. In what follows, a comprehensive survey of the existing frictional pressure drop models, which were generated to predict the two-phase frictional pressure drop in straight pipes, is conducted first. Then the experimental setup of pipeline-riser system is presented. Finally, the prediction performances of 24 existing equations are tested against the 885 experimental data points to evaluate the applicability of these correlations to gas-liquid two-phase flow in pipeline-riser system.

2. Review of existing frictional pressure drop models

2.1. Models based on homogeneous flow

In homogeneous flow, the mixture of gas and liquid is regarded as well-mixed single-phase flow. Hence, the properties of the mixture can be determined according to the proportion of each phase. For a steady state flow in a channel with a constant cross-section area, the frictional pressure drop is calculated using correlation proposed for single-phase flow, as defined by Eq. (1).

$$\left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{TP}}=\frac{G_{\mathrm{TP}}^2}{2D\rho_{\mathrm{TP}}}f_{\mathrm{TP}} \tag{1}$$

where f_{tp} represents the friction factor, ρ_{TP} is the density averaged by the proportion of gas and liquid, defined as Eq. (2):

$$\frac{1}{\rho_{\mathrm{TP}}}=\frac{x}{\rho_g}+\frac{1-x}{\rho_l} \tag{2}$$

The two-phase friction factor f_{TP} is a function of the Reynolds number, defined as:

$$Re_{\mathrm{TP}}=\frac{G_{\mathrm{TP}}D}{\mu_{\mathrm{TP}}} \tag{3}$$

As is indicated in Eq. (3), Re_{TP} mainly depends on the mixture average viscosity μ_{TP} of the two-phase. Therefore, the main difference among the homogeneous flow based frictional pressure drop correlations is the methods they used to calculate μ_{TP}. Table 1 shows 8 correlations for different working fluids, pipe orientations and diameters collected in this paper.

2.2. Models based on separated flow

It is a kind of more complicated model compared to the homogeneous flow-based model. The separated flow models considering the effects of two phases separately were generally developed based on the concept of two-phase friction multipliers which was firstly proposed by Lockhart and Martinelli (1949).

2.2.1. $\Phi^2{}_L$ and $\Phi^2{}_G$ based method

Assuming that a single phase passes through the pipe, the two-phase frictional pressure drop gradient can be expressed as Eq. (4) by introducing the two-phase friction multiplier $\Phi^2{}_{\mathrm{L}}$.

$$\left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{TP}}=\Phi_{\mathrm{L}}^2\left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{L}} \tag{4}$$

where $[\mathrm{d}P/\mathrm{d}L]_{\mathrm{L}}$ in Eq. (4) represents the frictional pressure gradient assuming that the liquid phase of a two-phase flow mixture flows alone in the pipe.

$$\left(\frac{\mathrm{d}P}{\mathrm{d}L}\right)_{\mathrm{L}}=\frac{[(1-x)G_{\mathrm{TP}}]^2}{2D\rho_{\mathrm{L}}}f_{\mathrm{L}} \tag{5}$$

where f_{L} is the single-phase frictional factors determined by the mass flux and properties of liquid phase.

Table 1
Correlations for the two-phase mixture viscosity.

Author	Correlation	Source
Akers et al. (1959)	$\mu_{TP}=\mu_L/(1-x)+x(\rho_L/\rho_G)^{0.5}$	Based on experimental data in horizontal pipes
Cicchitti et al. (1960)	$\mu_{TP}=\mu_L(1-x)+\mu_G x$	Based on experimental data of two-phase flow
Dukler et al. (1964)	$\mu_{TP}=[x(\mu_G/\rho_G)+(1-x)(\mu_L/\rho_L)]$	Based on experimental data in large pipes
Beattie and Whalley (1982)	$\mu_{TP}=\mu_L(1-\beta)(1+2.5\beta)+\mu_G\beta$ $\beta=x/[x+(1-x)\rho_G/\rho_L]$	Flow regime dependent
Lin et al. (1991)	$\mu_{TP}=\mu_G\mu_L/[\mu_G+x^{1.4}(\mu_L-\mu_G)]$	Based on refrigerants in capillary channels
Fourar and Boris (1995)	$\mu_{TP}=(1-\beta)\mu_L+\beta\mu_G+2[\beta(1-\beta)\mu_L\mu_G]^{0.5}$	0.18–1 mm horizontal narrow channels
Awad and Muzychka (2008)	$\mu_{TP}=\mu_L[2\mu_L+\mu_G-2(\mu_L-\mu_G)x]/[2\mu_L+\mu_G+(\mu_L-\mu_G)x]$	New definition of two-phase viscosity
Shannak (2008)	$\mu_{TP}=\mu_G[2\mu_G+\mu_L-2(\mu_G-\mu_L)(1-x)]/[2\mu_G+\mu_L+(\mu_G-\mu_L)(1-x)]$ $Re_{TP}=\{G_{TP}D[x^2+(1-x)^2(\rho_G/\rho_L)]\}/[\mu_G x+\mu_L(\rho_L\rho_G)(1-x)]$	Based on air-water mixture in horizontal and vertical pipes

As suggested by Chisholm (1967), Φ^2_L is treated as the function of X and C.

$$\Phi_L^2=1+\frac{1}{X^2}+\frac{C}{X} \tag{6}$$

The coefficient C of the third term in the right hand of Eq. (6) is an adjustable parameter accounting for the interactions between the two phases. X^2 is known as the Lockhart-Martinelli parameter defined as:

$$X^2=\frac{\left[\frac{dP}{dL}\right]_L}{\left[\frac{dP}{dL}\right]_G} \tag{7}$$

where $[dP/dL]_G$ in Eq. (7) is the frictional pressure gradient assuming that that the gas phase of a two-phase flow mixture flows alone in the pipe.

This concept has been used wildly because of its simplicity and sufficient accuracy for field applications. The main complexity of this method is how to calculate the pressure drop multiplier. 8 Φ^2_L and Φ^2 based correlations are collected, as is listed in Table 2. The correlation by Yao et al. (2018) was not considered in the present study because it was developed for downward vertical flow where the buoyancy acts on gas bubbles is opposite to the direction of the flow.

2.2.2. Φ^2_{LO} and Φ^2_{GO} based method

If a single-phase of a two-phase mixture flows in the pipe with the total mass flux of the mixture, the frictional pressure drop of two-phase flow can be defined as:

$$\left[\frac{dP}{dL}\right]_{TP}=\Phi_{LO}^2\left[\frac{dP}{dL}\right]_{LO} \tag{8}$$

or

$$\left[\frac{dP}{dL}\right]_{TP}=\Phi_{GO}^2\left[\frac{dP}{dL}\right]_{GO} \tag{9}$$

where $[dP/dL]_{LO}$ in Eq. (8) and $[dP/dL]_{GO}$ in Eq. (9) is the frictional pressure drop when the liquid or gas phase of a two-phase flow mixture flows in the pipe with the total mass flow rate of the mixture, respectively. Φ^2_{LO} in Eq. (8) and Φ^2_{GO} in Eq. (9) denote the two-phase friction multiplier for liquid and gas phase whose mass flow rate is assumed to be equivalent to the entire two phase mixture flow rate, respectively.

$$\left(\frac{dP}{dL}\right)_{LO}=\frac{G_{TP}^2}{2D\rho_L}f_{LO} \tag{10}$$

$$\left(\frac{dP}{dL}\right)_{GO}=\frac{G_{TP}^2}{2D\rho_G}f_{GO} \tag{11}$$

where f_{LO} and f_{GO} are the single-phase frictional factors for liquid and gas phase, respectively. f_{LO} and f_{GO} can be determined using the properties of single-phase together with the total mass flux of the mixture.

Table 3 lists 8 Φ^2_{LO} and Φ^2_{GO} based empirical equations considered in this work.

3. Experiment

3.1. Test loop

A pipeline-riser system was used for the experiment, see Fig. 1. The test loop was made of stainless steel, consisting of a horizontal pipe with 114 m in length, a downward inclined section with 16 m in length and a vertical riser. The riser has a height of 15.3 m when the inclination angle of the downcomer is 2-degree. The test section has a constant internal diameter of 0.046 m.

Both air-water and air-oil was used as working fluids in the present study. In air-water two-phase flow experiment, tap water in tank was supplied by a centrifugal pump having a capacity of 30 m^3/h. In air-water two-phase flow experiment, light oil was supplied by a gear pump with a capacity of 28 m^3/h. Air was first filtered and then compressed by an equipment with a maximum flow rate of 360 m^3/h. The two phases were transported into the pipeline via a mixing tee in order to obtain a stratified flow at the inlet. The velocity of each phase is controlled by ball valves and a bypass pipe. The pressure at the entrance of the test section is controlled at approximately 700 kPa to make sure that the pressure in the test loop is bearable for the test tubes. After being tested, fluids from the outlet of the riser were discharged into a cyclone separator working at atmospheric condition. Air separated from the mixture was vented out to open air, while water or oil was pumped to the storage tank.

3.2. Instrumentation

Electromagnetic flow meter having an accuracy of ±0.5% was used to measure the velocity of each phase. The gas and liquid flow meters were calibrated to provide superficial gas and liquid velocities (U_{SG} and U_{SL}) directly as output instead of volumetric flow rates. To determine the flow regions for different flow patterns, the ranges of superficial liquid velocity and superficial gas velocity under standard conditions are set as 0.02–1.3 m/s and 0.03–6.1 m/

Table 2
Correlations for $\Phi^2{}_L$ and $\Phi^2{}_G$ based method.

<table>
<tr><th>Author</th><th colspan="6">Method</th><th>Source</th></tr>
<tr><td>Chisholm (1967)</td><td colspan="6">X=1+1/X^2+C/X, C is a constant for different flow conditions</td><td>1.5–26 mm, horizontal pipes</td></tr>
<tr><td>Mishima and Hibiki (1996)</td><td colspan="6">C=21[1−exp(−0.319D)]</td><td>1–4 mm horizontal pipes</td></tr>
<tr><td rowspan="6">Lee and Lee (2001)</td><td colspan="6">$C=A\lambda^q\Psi^R Re^S{}_{LO}$, where $\lambda=\mu_L/(\rho_L\sigma D)$, $\Psi=\mu_L j/\sigma$</td><td rowspan="6">0.4–4 mm rectangular channels</td></tr>
<tr><td>Re_L</td><td>Re_G</td><td>A</td><td>q</td><td>R</td><td>S</td></tr>
<tr><td>≤2000</td><td>≤2000</td><td>6.833×10^{-8}</td><td>−1.317</td><td>0.719</td><td>0.557</td></tr>
<tr><td>≤2000</td><td>>2000</td><td>6.185×10^{-2}</td><td>0</td><td>0</td><td>0.726</td></tr>
<tr><td>>2000</td><td>≤2000</td><td>3.627</td><td>0</td><td>0</td><td>0.174</td></tr>
<tr><td>>2000</td><td>>2000</td><td>0.408</td><td>0</td><td>0</td><td>0.451</td></tr>
<tr><td>Lee and Mudawar (2005)</td><td colspan="6">$C=2.16Re_{LO}^{0.047}We_{LO}^{0.6}$ for vv
$C=1.45Re_{LO}^{0.25}We_{LO}^{0.23}$ for vt
where $Re_{LO}=G_{TP}D/\mu_L$, $We_{LO}=G_{TP}^2D/(\sigma\rho_L)$</td><td>R134a in a micro-channel</td></tr>
<tr><td>Hwang and Kim (2006)</td><td colspan="6">$C=0.227Re_{LO}^{0.452}X^{-0.32}La^{-0.82}$
where $La=[\sigma/g(\rho_L-\rho_G)]^{0.5}/D$</td><td>R134a in 0.244−0.79 mm tubes</td></tr>
<tr><td>Sun and Mishima (2009)</td><td colspan="6">For viscous flow:
$C=26(1+Re_L/1000)[1-\exp(-0.153/(0.8+0.27La))]$
For turbulent flow:
$\Phi_L^2=1+C/X^{1.19}+1/X^2$, $C=1.79\times(Re_G/Re_L)^{0.4}[(1-X)/X]^{0.5}$
where $Re_G=G_{TP}D/\mu_G$, $Re_L=G_{TP}(1-\chi)D/\mu_L$</td><td>Various refrigerants in 0.5–12 mm pipes</td></tr>
<tr><td>Zhang et al. (2010)</td><td colspan="6">C=21[1−exp(−0.674/La)] for adiabatic gas-liquid
C=21[1−exp(−0.142/La)] for adiabatic vapor-liquid
C=21[1−exp(−0.358/La)] for flow boiling</td><td>An extension of Mishima and Hibiki's correlation</td></tr>
<tr><td>Pamitran et al. (2010)</td><td colspan="6">$C=3\times10^{-3}We_{TP}^{-0.433}Re_{TP}^{1.23}$
where $We_{TP}=G_{TP}^2D/(\sigma\rho_{TP})$, $Re_{TP}=G_{TP}D/\mu_{TP}$</td><td>Various refrigerants in 0.5–3 mm tubes</td></tr>
</table>

Table 3
Correlations based on $\Phi^2{}_{LO}$ and $\Phi^2{}_{GO}$ method.

Author	Correlation	Source
Chisholm (1973)	$\Phi^2{}_{LO}=1+(Y^2-1)\{B[x(1-x)]^{0.875}+x^{1.75}\}$ $Y^2=(\Delta P/\Delta L)_{GO}/(\Delta P/\Delta L)_{LO}$ $0<Y<9.5$, $B=55/G^{0.5}{}_{TP}$, for $G_{TP}\geq1900$ kg/(m^2 s) B=2400, for 1900 kg/(m^2 s)$\geq G_{TP}\geq$500 kg/(m^2 s) B=4.8, for $G_{TP}\leq$500 kg/(m^2 s) $9.5<Y<28$, $B=520/(YG^{0.5}{}_{TP})$, for $G_{TP}\leq$ 600 kg/(m^2 s) $B=21/Y$, for $G_{TP}>$600 kg/(m^2 s) $Y>28$, $B=15000/(Y^2G^{0.5}{}_{TP})$	Mathematical expression of Barcozy graph
Friedel (1979)	$\Phi^2{}_{LO}=(1-x)^2+x^2(\rho_L f_{GO})/(\rho_G f_{LO})+[5.7x^{0.78}(1-x)^{0.224}H]/(Fr^{0.045}{}_{TP}We^{0.035}{}_{TP})$ $H=(\rho_L/\rho_G)^{0.91}(\mu_G/\mu_L)^{0.19}(1-\mu_G/\mu_L)^{0.7}$ $Fr_{TP}=G^2{}_{TP}/(gD\rho^2{}_{TP})$	Based on experimental data in large pipes
Friedel (1985)	$\Phi^2{}_{LO}=(1-x)^2+x^2(\rho_L f_{GO})/(\rho_G f_{LO})+[5.7x^{0.7}(1-x)^{0.14}H]/(Fr^{0.09}{}_{TP}We^{0.007}{}_{TP})$ $H=(\rho_L/\rho_G)^{0.85}(\mu_G/\mu_L)^{0.36}(1-\mu_G/\mu_L)^{0.2}$ $Fr_{TP}=G^2{}_{TP}/(gD\rho^2{}_{TP})$	On the basis of more than 25,000 measured data points
Muller-Steinhagen and Heck (1986)	$\Phi^2{}_{LO}=Y^2x^3+(1-x)^{1/3}[1+2x(Y^2-1)]$	On the basis of more than 9000 measured data points
Souza and Pimenta (1995)	$\Phi^2{}_{LO}=1+(\Gamma^2-1)x^{1.75}(1+0.9524\Gamma X^{0.4216}{}_{tt})$ $\Gamma=(\rho_L/\rho_G)^{0.5}(\mu_G/\mu_L)^{0.125}$, $X_{tt}=1/[(1-x)/x]^{0.875}$	Based on experimental data in large pipes
Tran et al. (2000)	$\Phi^2{}_{LO}=1+(4.3Y^2-1)\{[x(1-x)]^{0.875}La+x^{1.75}\}$	Based on B-coefficient method for small tubes
Cavallini et al. (2002)	$\Phi^2{}_{LO}=(1-x)^2+x^2(\rho_L f_{GO})/(\rho_G f_{LO})+(1.62x^{0.6978}H)/We^{0.1458}{}_{GO}$ $H=(\rho_L/\rho_G)^{0.32}(\mu_L/\mu_G)^{1.181}(1-\mu_G/\mu_L)^{3.47}$ $We_{GO}=G^2{}_{TP}D/(\rho_G\sigma)$	Based on experimental data in small horizontal tube
Xu and Fang (2012)	$\Phi^2{}_{LO}=\{Y^2X^3+[1+2x(Y^2-1)]\times(1-x)^{0.33}\}[1+1.54\times La\times(1-x)^{0.5}]$ $La=(\sigma/[g(\rho_L-\rho_G)])^{0.5}/D$	Various refrigerants in 0.5−10 mm channels

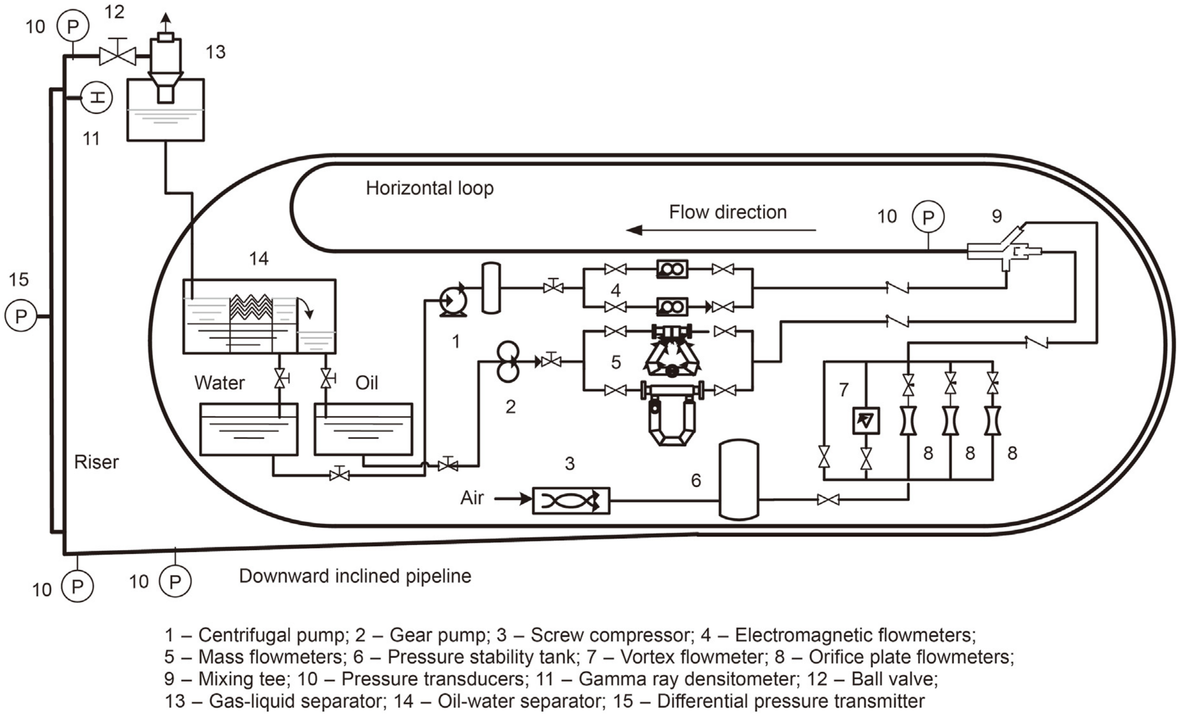

Fig. 1. Schematic diagram of the test loop.

s, respectively. Each experiment was run at a constant gas and liquid velocity. The test rig was equipped with 1 differential pressure transmitter (FS±0.1%) over the riser to monitor the development of the flow structure. To obtain frictional pressure drop, local instantaneous void fraction was also measured by Gamma ray density monitor located at the top of the riser. The void fraction for each test run is the time-averaged value of the measured void fraction profile. All the concerned experimental data was monitored and recorded by PCI-6259 from National Instrument. All the signals were sampled at a frequency of 200 Hz, which is high enough for flow in the riser according to Shannon theorem. The recording time of the concerned signals generally ranged from 5 to 20 min, depending on the actual flow regimes. In order to avoid random error, the experiment is repeated for 5 times at least in each test run, and the arithmetic average value of all the measurements for the test run is taken as the measured result of experimental pressure drop. The pressure drops over the riser and the local void fractions at the top of the riser were measured simultaneously in each run. A data acquisition and analysis program developed based on LabVIEW was used for data processing.

3.3. Experiment condition

Experiments were conducted at room temperature with the inclined angle of the downcomer varied from −2° to −5°. The superficial gas velocity changes from 0.03 to 6.1 m/s, and the superficial liquid velocity changes from 0.02 to 1.3 m/s. The ranges of the test is wild in order to cover all the possible flow regimes. The experimental range for the test is listed in Table 4.

The experiments were carried out at ambient temperature of 26 °C. Table 5 gives the physical properties of working fluids at 26 °C. It was found by carefully comparing pressure drop profiles at different ambient temperature that slightly change in ambient temperature has neglectable effect on the behavior of the flow patterns.

The uncertainty of all the concerned parameters in the experiment is given in Table 6. The temperatures are measured with thermocouple, having a maximum error of 1 °C. The uncertainties of superficial velocity are determined by the accuracy of the electromagnetic flow meter. The uncertainties of void fraction are based on the accuracy of the Gamma ray density monitor, having a maximum error of 1%.

A total of 885 experimental data points, including 782 air-water flows and 103 air-oil flows, are obtained. The distribution of flow condition is presented in Fig. 2. It is obviously that a majority of experiments conducted here are located in *tt* (turbulent liquid and turbulent gas) region and *tv* (turbulent liquid and viscous gas) region.

Table 4
Experiment conditions.

β	Working fluids	U_{SG}, m/s	U_{SL}, m/s	Data points
−2°	Air-water	0.07–6.1	0.09–1.3	193
−3°	Air-water	0.04–6.0	0.03–1.0	190
−4°	Air-water	0.04–6.2	0.03–1.1	198
−5°	Air-water	0.03–6.1	0.02–1.2	201
−2°	Air-oil	0.05–3.19	0.04–1.05	103
Total		0.03–6.2	0.02–1.3	885

Table 5
Physical properties of test fluids at 26 °C.

Properties	Water	Oil	Air
Density, kg/m^3	993	829	1.19
Surface tension, N/m	0.071	0.0301	0.071
Viscosity, mPa/s	1.001	28.6	0.191

Table 6
Uncertainties of the measured parameters.

Parameters	Maximum uncertainty
Superficial gas velocity	0.5%
Superficial liquid velocity	0.5%
Temperature	1 °C
Local void fraction	1%
Pressure drop	0.1%

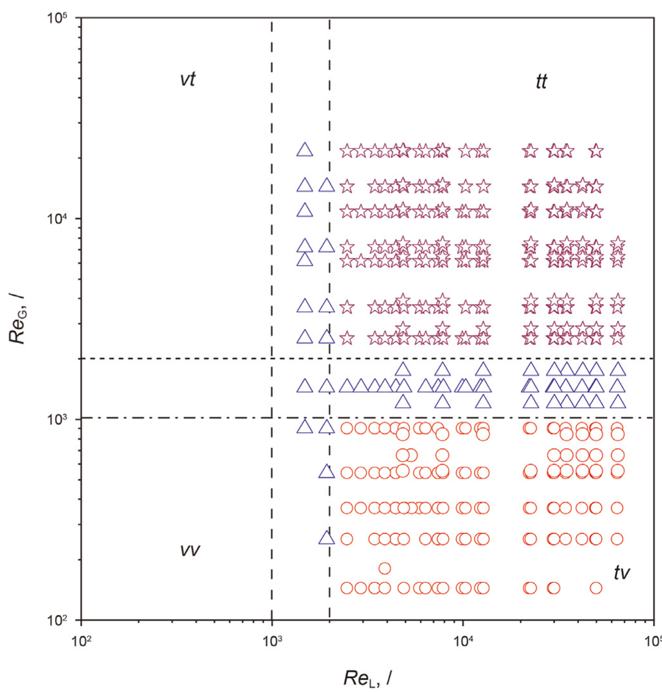

Fig. 2. Distribution of the experimental data by flow conditions.

4. Results and discussion

4.1. Flow patterns

For both air-water and air-oil experiments, severe slugging of type 1 (SS1), severe slugging of type 2 (SS2), transition flow (TRS), bubbly flow (BF), slug flow (SF) and annular flow (AF) were observed over the entire experimental range of gas and liquid flow rates. Flow pattern was identified based on the method developed by our previous works (Li et al., 2013; Ye and Guo, 2013). Figs. 3 and 4 shows profiles of total pressure drop for each flow pattern of air-water and air-oil two-phase flow, respectively. It is obviously that severe slugging of types 1 and 2 induce large amplitude of fluctuations in pressure drop over the riser. Flow pattern of severe slugging and transition flow are characterized by one surge in pressure drop profile in each individual period. The sharp increase and decrease in pressure drop profiles imply gas penetration into the riser. This might cause violent vibration of the riser. Besides, the flow rate at riser outlet is also associated with the surges in pressure drop, meaning an unstable production in liquid. Violent fluctuations in liquid production usually cause serious flooding of the separator (Luo et al., 2014). Zhang R.Y. et al. (2022) pointed out that slug flow might pose challenges to the reliability and performance of facilities such as the helicon-axial multiphase pump which is used to deliver hydrocarbons in subsea pipelines. The violent fluctuations in pressure drop might induce deformation and torque of the shaft. Therefore, these three flow patterns are unstable and undesired in practical process. However, in bubbly and annular flow, the fluctuations in pressure drop is relatively smooth. Detailed information of gas-liquid two-phase flow patterns in pipeline-riser can be seen in our previous work (Li et al., 2013).

The experimental results show that the flow patterns observed here are not exactly the same as the traditional flow patterns in upward vertical tubes, as a result of the pipeline upstream the riser. As the existing pressure drop correlations were mainly proposed on the basis of gas-liquid in straight pipes, their applicability for gas-liquid two-phase flow in complex pipelines where flow patterns differ from those in straight pipes needs to be assessed.

4.2. Applicability of the existing correlations to pipeline-riser flow

It is agreed that the total pressure drop of two-phase flow contains three components, as follows:

$$\left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{TP}} = \left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{F}} + \left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{A}} + \left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{S}} \tag{12}$$

where subscript TP, F, A and S denote the total, the frictional, the acceleration, and the static pressure drop, respectively. In view of the fact that there is no phase transition in the experiment, the acceleration part can be neglected. Thus, the term of frictional pressure gradient can be estimated through Eq. (13):

$$\left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{F}} = \left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{TP}} - \left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{S}} \tag{13}$$

The static pressure gradient $[\mathrm{d}P/\mathrm{d}X]_{\mathrm{H}}$ can be determined by Eq. (14):

$$\left[\frac{\mathrm{d}P}{\mathrm{d}L}\right]_{\mathrm{S}} = g\left[\alpha\rho_{\mathrm{g}} - (1-\alpha)\rho_{\mathrm{l}}\right]L \tag{14}$$

where L represents the length of the measuring section, and α denotes the gas void fraction of the two-phase flow.

The identified 24 two-phase frictional pressure drop correlations are tested against the entire database established in this study, as is shown in Table 7. All the correlations considered here are derived originally for straight pipes without flow orientation changing. The models considering the curvature effects including that of Guo et al. (2001) and Cioncolini and Santini (2016) were not evaluated in this study, because they were specifically developed for flows in helical coils where centrifugal force dominates the flow.

The comparisons of the predictions of the methods with the measured data are shown in Figs. 5–7. The abscissa represents measured pressure drop, while the ordinate denotes the predicted one. The error band of ±30% is also shown by the solid lines in these figures.

The performance of the correlations is evaluated by MARD (mean absolute relative deviation) and MRD (mean relative deviation), defined as Eqs. (15) and (16), respectively.

$$\mathrm{MARD} = \frac{1}{N}\sum_{i=1}^{N}\left|\frac{y(i)_{\mathrm{pred}} - y(i)_{\mathrm{exp}}}{y(i)_{\mathrm{exp}}}\right| \tag{15}$$

$$\mathrm{MRD} = \frac{1}{N}\sum_{i=1}^{N}\frac{y(i)_{\mathrm{pred}} - y(i)_{\mathrm{exp}}}{y(i)_{\mathrm{exp}}} \tag{16}$$

where pred and exp represent results predicted by the correlations and measured experimentally, respectively.

As is seen from Table 7, there are no correlations satisfactorily predict the experimental results of gas-liquid two-phase flow in

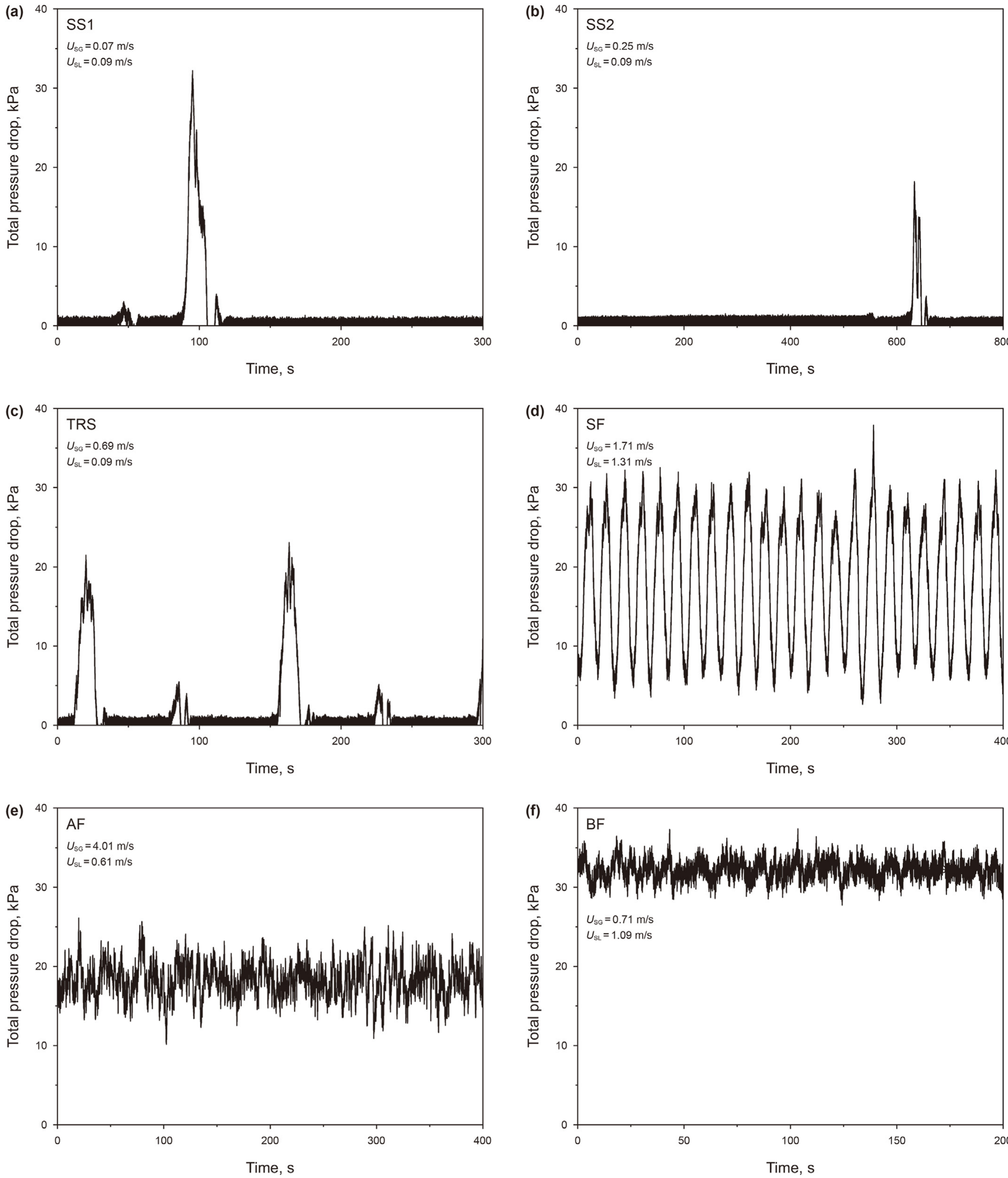

Fig. 3. Profiles of total pressure drop for each flow pattern of air-water two-phase flow in pipeline-riser. (**a**) SS1, (**b**) SS2, (**c**) TRS, (**d**) SF, (**e**) AF, (**f**) BF.

pipeline-riser. The top five methods with MARD<40% for both air-water and air-oil flows are Dukler et al. (1964), Sun and Mishima (2009), Zhang et al. (2010), Muller-Steinhagen and Heck (1986), Xu and Fang (2012). Note that the correlation by Xu and Fang (2012) is also recommended by Yao et al. (2018) for vertical downward flow. Among the four correlations of two-phase viscosity proposed

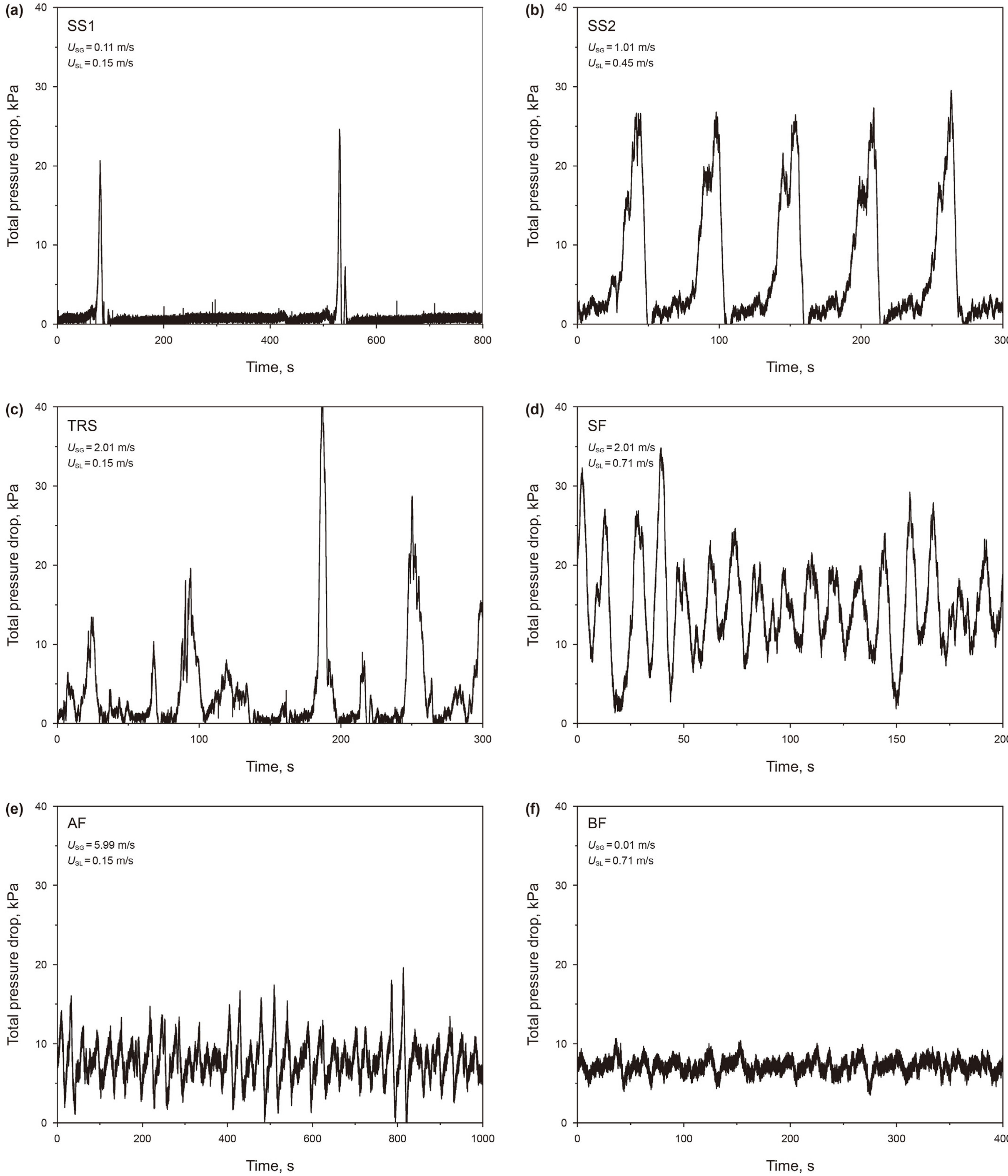

Fig. 4. Profiles of total pressure drop for each flow pattern of air-oil two-phase flow in pipeline-riser. (**a**) SS1, (**b**) SS2, (**c**) TRS, (**d**) SF, (**e**) AF, (**f**) BF.

by Awad and Muzychka (2008), only definition 3 providing the best predictions is given in Table 1. In the preferred five correlations, Muller-Steinhagen and Heck (1986), Xu and Fang (2012) under-predict pressure drop for both air-water and air-oil flows, while Dukler et al. (1964), Sun and Mishima (2009), Zhang et al. (2010) over-predict the pressure drop for air-water flow and under-

Table 7
Comparison of predicted and measured two-phase frictional pressure drop.

Model	Air-water				Air-oil			
	MARD, %	MRD, %	Ψ, ±30%	Ψ, ±50 %	MARD, %	MRD, %	Ψ, ±30%	Ψ, ±50 %
Akers et al. (1959)	127.3	119.2	29.2	46.6	51.8	−24.5	24.5	50.9
Cicchitti et al. (1960)	77.9	68.1	35.9	54.6	90.3	73.1	37.7	55.6
Dukler et al. (1964)	24.3	12.3	72.2	86.6	34.5	−10.2	58.5	81.1
Beattie and Whalley (1982)	51.5	48.5	26.4	48.0	43.9	27.1	55.6	70.8
Lin et al. (1991)	101.6	97.8	28.6	41.4	80.3	62.9	39.6	57.5
Fourar and Boris (1995)	44.5	32.7	47.6	57.4	50.1	−37.1	19.8	45.3
Awad and Muzychka (2008)	72.3	68.2	39.2	56.4	49.8	−30.8	22.6	47.1
Shannak (2008)	51.8	45.6	41.8	62.4	88.8	72.1	36.8	52.8
Chisholm (1967)	40.1	20.2	37.8	73.3	38.6	9.38	57.5	79.2
Mishima and Hibiki (1996)	41.9	32.8	45.4	68.6	40.3	12.9	57.5	77.3
Sun and Mishima (2009)	24.1	10.6	65.4	82.6	38.8	−25.1	35.8	75.5
Zhang et al. (2010)	31.4	13.1	62.1	81.0	39.9	−8.7	51.8	77.3
Hwang and Kim (2006)	231.8	225.1	19.4	30.2	139.1	118.5	25.5	40.6
Lee and Mudawar (2005)	104.7	92.4	27.1	45.6	76.3	58.9	32.1	50.0
Pamitran et al. (2010)	189.7	187.2	11.4	22.0	136.9	130.0	17.9	28.3
Lee and Lee (2001)	183.5	177.1	18.6	28.8	131.2	121.1	21.6	31.1
Chisholm (1973)	107.1	103.2	35.1	51.4	101.1	84.1	34.9	54.7
Friedel (1979)	47.1	32.6	36.8	54.6	44.2	14.1	50.0	73.6
Friedel (1985)	41.3	29.8	39.2	68.6	37.8	13.1	55.6	74.5
Muller-Steinhagen and Heck (1986)	33.1	−19.3	48.2	72.8	38.6	−23.7	37.7	75.5
Cavallini et al. (2002)	39.3	21.5	50.2	70.2	45.8	11.7	46.2	72.6
Xu and Fang (2012)	36.1	−14.2	45.2	71.2	35.4	−17.2	51.9	79.2
Tran et al. (2000)	35.5	−18.8	44.6	81.0	49.1	−40.3	9.4	53.7
Souza and Pimenta (1995)	59.1	28.2	54.8	74.8	50.3	−16.3	22.6	66.9

Ψ denotes percentage (%) of data points in specified error ranges.

predict for air-oil flow. For further illustration, comparisons of the predicted frictional pressure drop by correlations of Dukler et al. (1964) based on homogeneous flow model, and Muller-Steinhagen and Heck (1986) based on separated flow model with the measured data are given in Figs. 8–11. Figs. 8 and 9 compare predictions by correlations of Dukler et al. (1964) with experimental data of air-water and air-oil two-phase, respectively. Figs. 10 and 11 compare predictions by correlations of Muller-Steinhagen and Heck (1986) with experimental data of air-water and air-oil two-phase, respectively. As can be seen from Figs. 8–10, the frictional pressure drop increases with superficial liquid velocity. Besides, the frictional pressure drop increases with superficial gas velocity for a constant superficial liquid velocity. The correlations of Dukler et al. (1964) and Muller-Steinhagen and Heck (1986) show reasonably good performance.

It is also evidently from Table 7 that parameter C in $\Phi^2{}_L$ and $\Phi^2{}_G$ based method is not a constant value but rather a variable depending on flow condition, as the MARD of Chisholm (1967) correlation is greater than the correlation with C specified by flow patterns. A comprehensive description of the correlations can be found in Xu et al. (2012). The correlations of Pamitran et al. (2010), Hwang and Kim (2006), and Lee and Lee (2001) predict pressure drop with MARD greater than 100% in both air-water and air-oil flows. In particular, the correlations of Pamitran et al. (2010) and Hwang and Kim (2006) were derived from experimentally measured data of many kinds of refrigerant in horizontal placed channels with small diameters ranging from 0.2 to 3.0 mm, while the method of Lee and Lee (2001) was proposed using data points of air and water in channels with small gaps between 0.4 and 4 mm, operating at atmospheric pressure condition. It is evident from the study of Xu et al. (2012) that the correlations developed for micro-tubes might be difficult to provide predictions of the two-phase frictional pressure gradient in macro-pipes, mainly due to the effect of gravity might act as a more important role while the influence of surface tension is less pronounced as the diameter of the pipe varying from micro-scale to macro-scale. Besides, the expression of these correlations basically involve restrictive parameters such as diameter D, surface tension σ, and Laplace number La. This suggests that the large deviation can be attributed to the fact that these predictive methods were generated by measured data in small diameter channels where surface tension is more important than gravity. Note that the correlation proposed by Zhang et al. (2010) fits our experimental data well with 62.1% within the error band of ±30% for air-water, and 51.8% within ±30% for air-oil. It is mainly because that both the influence caused by surface tension and the effect due to gravity are considered by introducing La.

Generally, the comparison in the present study indicates that the correlations based on separated flow model performs better than the correlations based on homogeneous flow model. Among the methods considered in this work, Dukler et al. (1964) based on HFM and Sun and Mishima (2009) based on SFM are the best two correlations having performance of MARD<30% and MRD within ±15%. It is also worth noting that all the methods based on HFM show large deviations except the Dukler et al. (1964) correlation, probably because the effect of two-phase density on pressure drop calculation was considered by Dukler et al. (1964). The reason for the large deviations can be given as follows. As stated previously, a homogeneous mixture is an ideal assumption for correlating pressure drop of gas-liquid two-phase flow. In homogeneous flow model, the mixture of gas-liquid flow is usually regarded as a pseudo-fluid. Consequently, the mixture is characterized by a homogeneous nature with the same properties suitably averaged by each phase. It can be concluded from the literature that almost all the homogeneous methods are developed for micro-channels, where surface tension is dominant. However, for gas-liquid two-phase flow in pipeline-riser studied in this work, gravity is dominant in the process of flow structure development. As is shown in Figs. 3 and 4, flow patterns including severe slugging, transition flow and slug flow are characterized by intermittent passage of gas and liquid slug. In contrast with homogeneous flow models, the separated flow models assume that the liquid and gas phases flow separately in a pipe with their own properties. Therefore, separated flow models are generally more comprehensive and more suitable

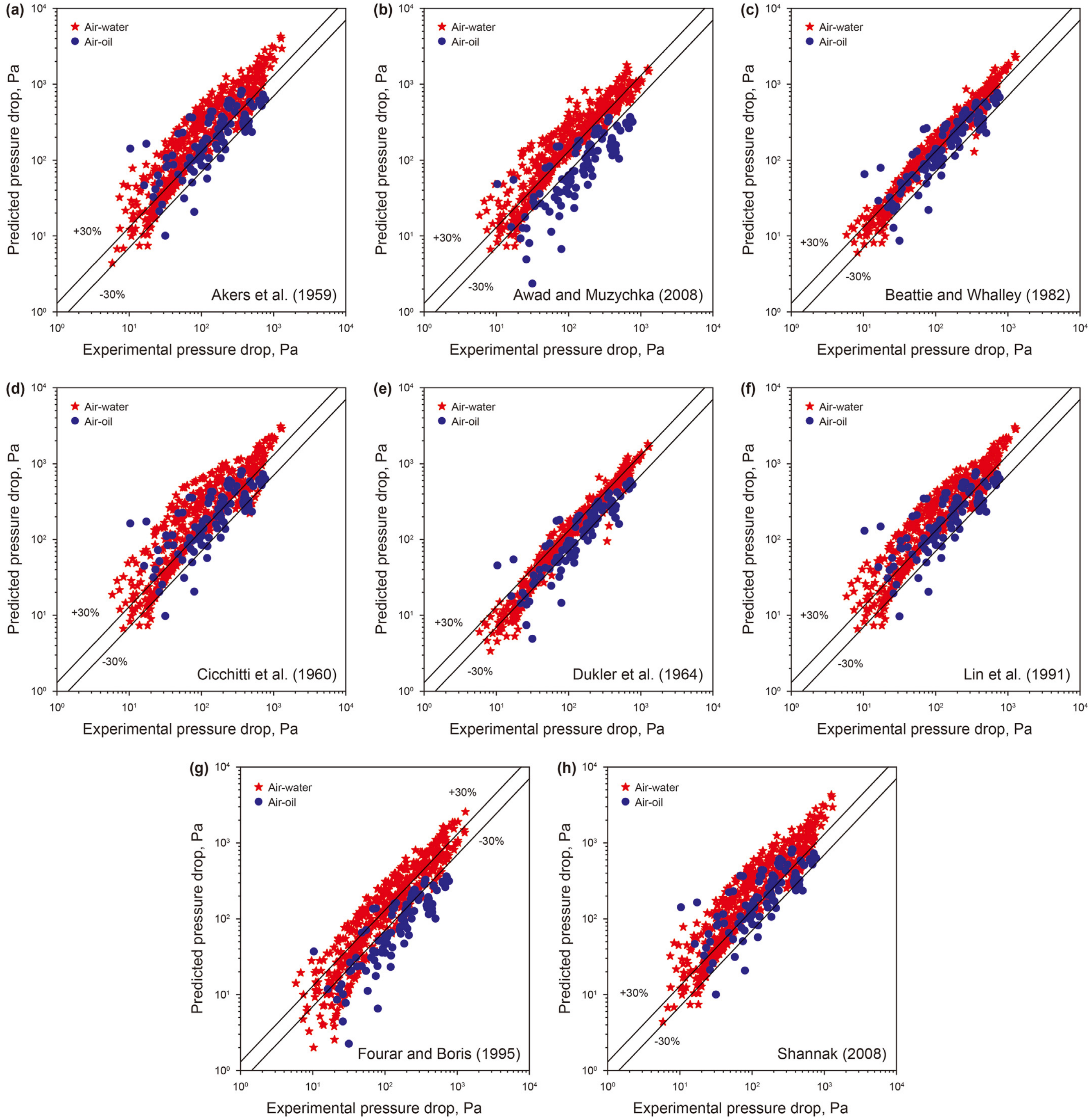

Fig. 5. Comparison of the results predicted by correlations based on homogeneous flow model with the experimental data.

in correlating pressure drop of two-phase flow in conventional pipe than homogeneous flow models. We tested the homogeneous correlations against our experimental data because some of these correlations, i.e. Shannak (2008) correlation, present reasonable accuracy for large pipes according to the published results (Yao et al., 2018).

Although the correlation of Zhang et al. (2010) was originally established for viscous flow, it provides generally good predictions with the MARD equals to 31.4% and 39.9% for air-water flow and air-oil flow, respectively. The same results can be found for Mishima and Hibiki (1996) which was developed for viscous flow and shows good agreements with our experimental data. It should be noted that there are no data points fall into *vt* and *vv* regions according to Fig. 2. In contrast to the correlations of Mishima and Hibiki (1996) and Zhang et al. (2010), the correlations proposed by Souza and Pimenta (1995) is only suitable for viscous flow. However, the disagreement is slightly larger with MARD equals to 59.1% and 50.3% for air-water flow and air-oil flow, respectively.

Special attention is then paid to the Sun and Mishima (2009) correlation which was generated by using 2092 flow conditions

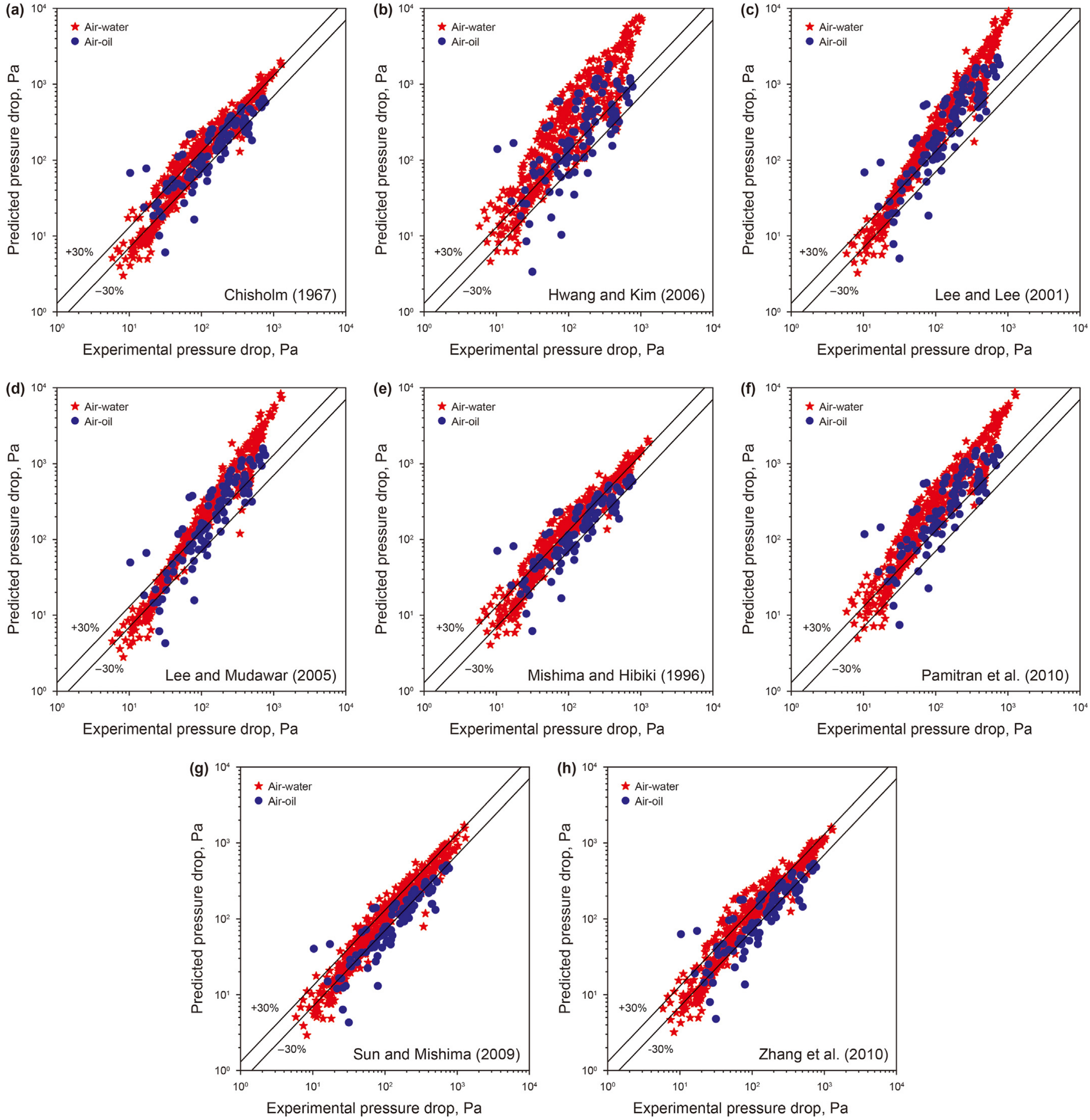

Fig. 6. Comparison of the results predicted by correlations based on $\Phi^2{}_L$ and $\Phi^2{}_G$ method with the measured data.

covering various working fluids including refrigerant, CO_2, air-water in tubes with small diameters ranging from 0.5 to 12 mm. In spite of the fact that the flow conditions it derived from are quite different from this work, comparison shows reasonable agreement for air and water two-phase flow, with 65.4% data points within ±30%. However, large discrepancy is observed when apply the Sun and Mishima (2009) correlation to air-oil flows. The predicted data of air-oil in Figs. 5–7 appears more disperse from the diagonal than the data points of air-water flows, suggesting that the properties of the working fluid might be one of the main reasons for the discrepancy. Even though the deduction of fluid properties on the frictional pressure drop in pipeline-riser might be merely preliminary and more measured data of air-oil flows are required, this result is still important because it highlights the need of method with sufficient accuracy for practical use.

Table 8 indicates the effect of gas void fraction on frictional pressure gradient prediction. As is observed that most methods result in larger deviations in low void fraction area ($\alpha < 0.4$) than in high void fraction area ($\alpha > 0.6$). Low values of the gas void fraction in pipeline-riser usually indicates the formation of severe slug flow,

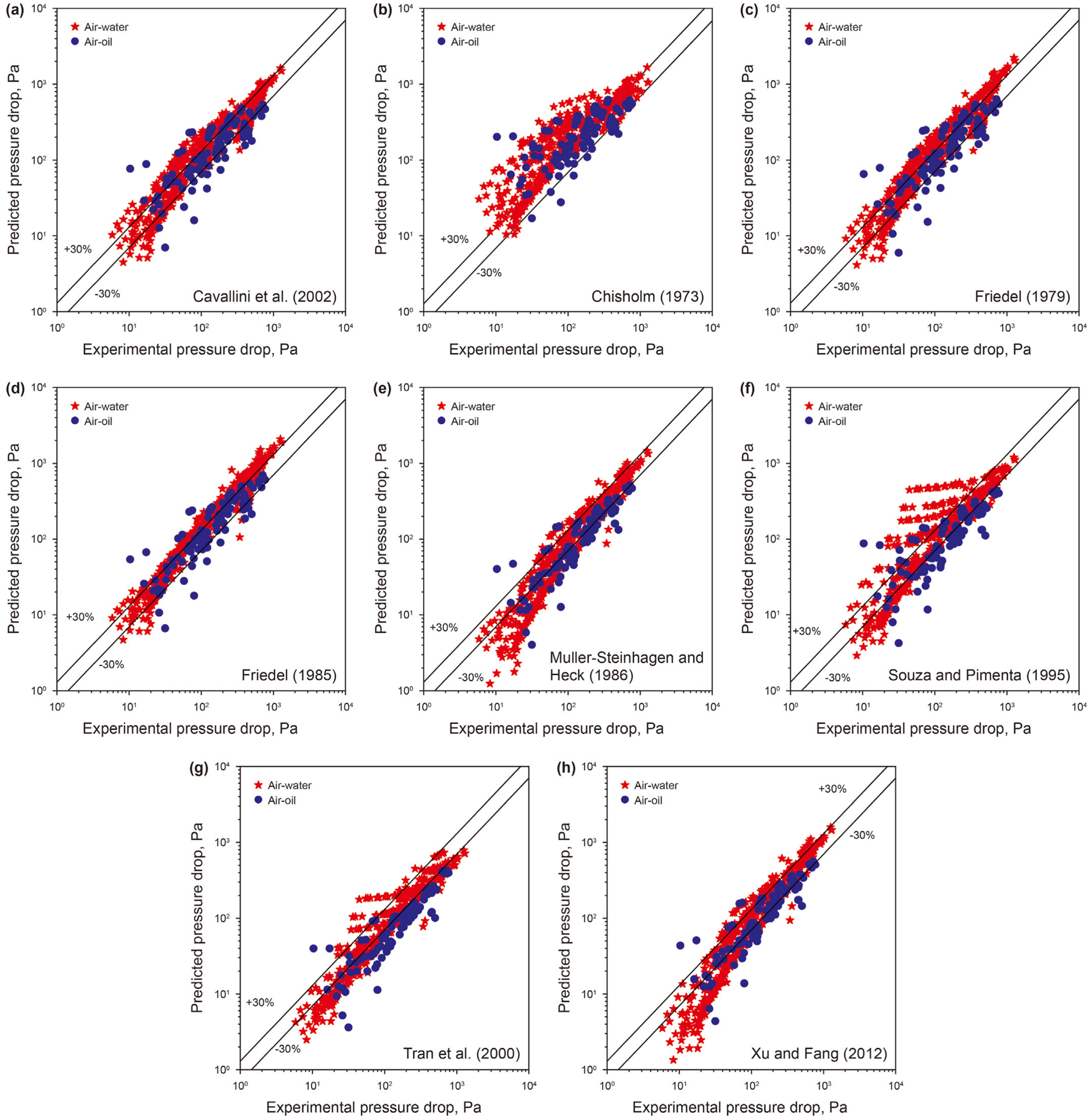

Fig. 7. Comparison of the results predicted by correlations based on Φ^2_{LO} and Φ^2_{GO} method with the measured data.

which is the dominant flow regime located in region of relatively low gas and liquid velocity in a flow pattern map.

As a result of the pipe layout, flow regimes in pipeline-riser are more complicated compared to flows in vertical tubes where bubbly flow, slug flow, churn flow and annular flow are the most common flow patterns. Following the identification method developed by previous works (Li et al., 2013, 2022; Ye and Guo, 2013; Xu et al., 2022), flow regimes investigated in the present work can be recognized as severe slugging (SSG), transitional flow (SST) and stable flow, which includes normal slug flow (SF), bubbly flow (BF) and annular flow (AF). Table 9 shows the predictions of the preferred five methods on different flow patterns to find out the flow conditions under which predicted methods have low accuracy. It is obviously that all correlations show the maximum MARD on SSG and SST. In addition, these models under-predict the two-phase frictional pressure gradient on SSG. This result indicates that frictional pressure drop highly depends on the actual flow patterns. The effect of pipeline layout on the pressure drop in pipeline-riser two-phase flows is also pronounced, since the pressure gradient in pipeline-riser is higher than straight tubes. The

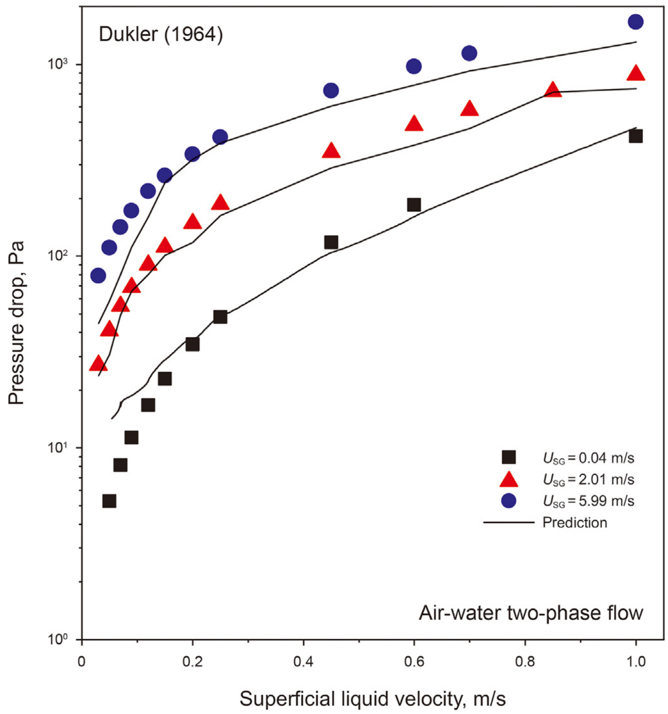

Fig. 8. Comparison of the results predicted by Dukler et al. (1964) with the experimental data of air-water two-phase flow.

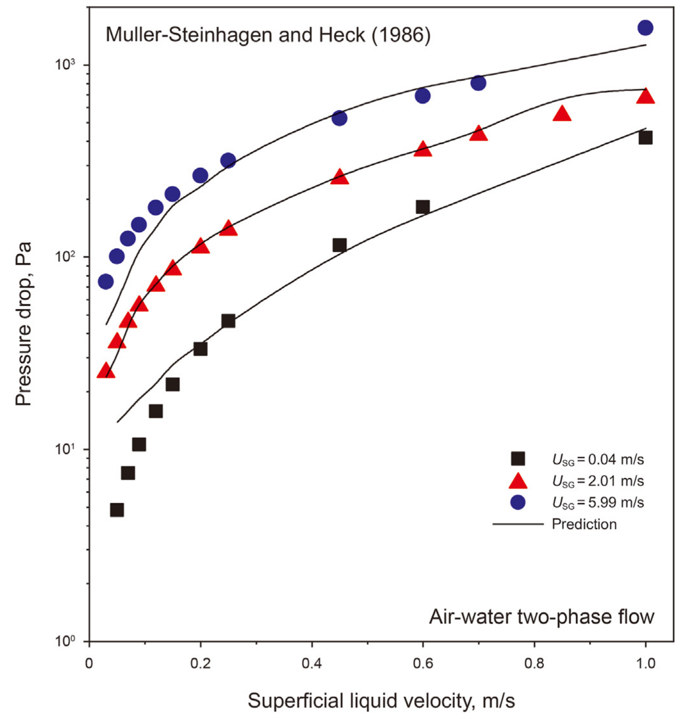

Fig. 10. Comparison of the results predicted by Muller-Steinhagen and Heck (1986) with the experimental data of air-water two-phase flow.

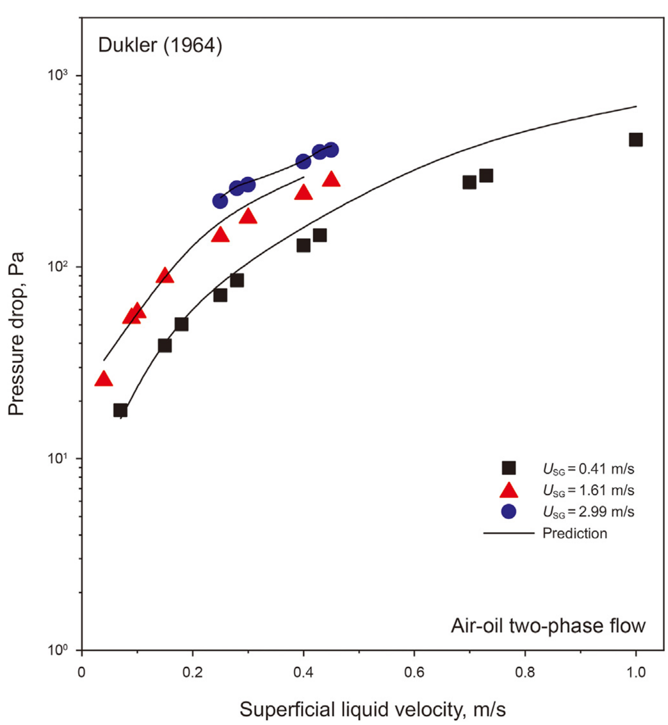

Fig. 9. Comparison of the results predicted by Dukler et al. (1964) with the experimental data of air-oil two-phase flow.

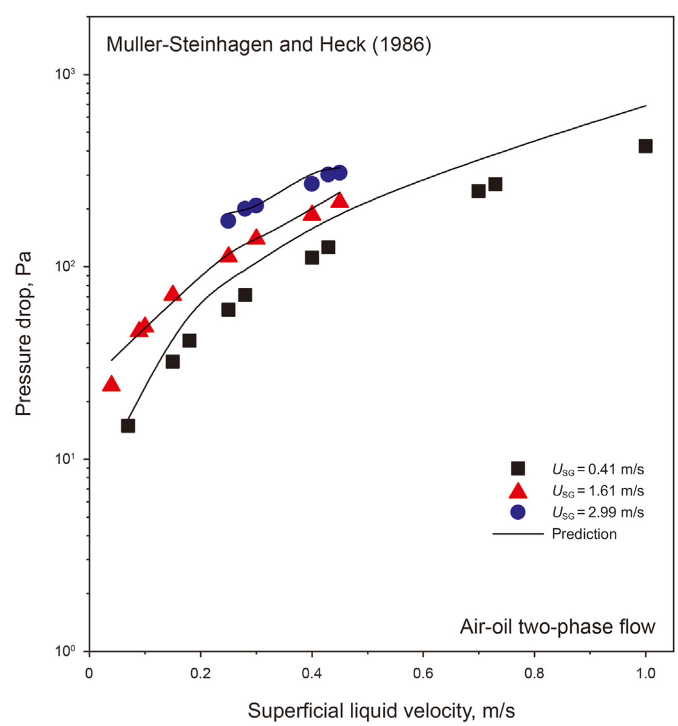

Fig. 11. Comparison of the results predicted by Muller-Steinhagen and Heck (1986) with the experimental data of air-oil two-phase flow.

straight pipe correlations can not be used directly to predict the frictional pressure gradient with sufficient accuracy and the comparison highlights the requirement of new method considering the effect of pipe layout on the frictional pressure drop.

Table 10 depicts the top five prediction methods under different downcommer inclination. The inclination changing from $-2°$ to $-5°$ leads to increase in curvature radius of the elbow connecting the pipeline and the riser. As a result, the effects of curvature on the pressure drop in the riser become more important. It is evidently from Table 10 that MARDs of the correlations increase with increase

Table 8
Predictions of the preferred five correlations against different void fraction.

Correlations	$0.0 \le \alpha \le 0.2$		$0.2 < \alpha \le 0.4$		$0.4 < \alpha \le 0.6$		$0.6 < \alpha \le 0.8$		$0.8 < \alpha \le 1.0$	
	MRD	MARD	MRD	MARD	MRD	MARD	MRD	MARD	MRD	MARD
Dukler et al. (1964)	35.5	38.1	−6.79	23.2	0.49	27.2	11.1	21.7	35.5	28.1
Sun and Mishima (2009)	28.2	32.9	−8.01	24.3	−10.8	27.3	−5.29	20.3	28.2	22.9
Zhang et al. (2010)	89.5	91.3	89.5	91.3	−9.99	22.4	10.9	23.7	−14.8	22.8
Muller-Steinhagen and Heck (1986)	25.1	43.1	−44.6	49.6	−34.3	31.1	−17.5	22.7	25.1	31.1
Xu and Fang (2012)	36.3	46.4	−43.1	50.6	−34.5	45.2	−12.1	25.8	36.3	40.4

Table 9
Predictions of preferred five correlations against different flow regimes.

Correlations	SSG		SST		BF		SF		AF	
	MRD	MARD	MRD	MARD	MRD	MARD	MRD	MARD	MRD	MARD
Dukler et al. (1964)	−6.76	44.7	15.3	20.7	18.6	23.7	25.9	31.0	44.7	20.1
Sun and Mishima (2009)	−10.4	35.1	1.73	20.2	14.3	22.9	7.01	21.3	29.7	24.8
Zhang et al. (2010)	−13.3	86.4	31.8	35.2	10.8	19.0	38.3	44.5	86.4	20.1
Muller-Steinhagen and Heck (1986)	−46.0	38.9	−13.7	17.8	10.2	18.7	8.77	16.8	33.7	33.1
Xu and Fang (2012)	−48.2	48.3	1.01	18.6	19.4	23.4	17.5	29.9	44.3	43.2

Table 10
Predictions of the top five correlations against different downcommer inclination.

Model	$-2°$		$-3°$		$-4°$		$-5°$	
	MRD	MARD	MRD	MARD	MRD	MARD	MRD	MARD
Dukler et al. (1964)	15.7	21.1	15.1	22.7	17.4	32.1	11.1	44.5
Sun and Mishima (2009)	−1.7	19.7	−11.6	26.3	−10.7	28.9	25.4	32.6
Zhang et al. (2010)	10.6	30.5	13.1	30.9	12.4	31.4	15.7	32.7
Muller-Steinhagen and Heck (1986)	−25.7	29.5	−22.3	33.6	−25.6	34.7	−17.1	35.1
Xu and Fang (2012)	−20.1	35.6	−18.9	35.9	−18.3	37.6	−15.2	38.5

in inclination. This result confirms the curvature effect on the frictional pressure gradient.

5. Conclusions

In order to evaluate the performance of the wildly used correlations in predicting frictional pressure gradient, experimental investigation on frictional pressure gradient in the riser was performed in pipeline-riser with downcomer inclination varied from $-2°$ to $-5°$. The experiment was performed with air-water and air-oil two-phase flow. Totally 885 experimental data points are obtained in this experiment and 24 frequently used models are tested against the measured data. The results for this study are concluded below.

(1) There is no method among those collected in this study precisely fits the experimental results. The leading five correlations, whose MARDs are smaller than 40% for both air-water and air-oil flows, are those, Dukler et al. (1964), Sun and Mishima (2009), Zhang et al. (2010), Muller-Steinhagen and Heck (1986), Xu and Fang (2012), in the order of the accuracy.
(2) Generally, existing models based on separated flow method predict frictional pressure drop with higher accuracy than those based on homogeneous flow model. This is mainly because separated flow models are more comprehensive and more suitable than homogeneous flow model in correlating pressure drop of gas-liquid two-phase flow in pipeline-riser system.
(3) Gas-liquid two-flow frictional pressure drop in pipeline-riser is greater than flows in straight pipes. Most of the existing correlations considered here under-predict the pressure drop on SSG.
(4) New method with satisfactory accuracy is needed. The effect of pipeline layout on the frictional pressure gradient in the riser should be taken into consideration in developing new correlation.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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