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Brief communication

Unsteady flow phenomena: implications on the design of experimental facilities

R. van Hout *

Department of Fluid Mechanics and Heat Transfer, Faculty of Engineering, Tel-Aviv University, 69978 Tel-Aviv, Israel

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1. Introduction

Unsteady flow phenomena are an intrinsic feature of many multiphase flows. A common example of unsteady nature in gas–liquid two-phase flow is slug flow which is considered to be "quasi-steady" when consecutive elongated bubbles propagate at almost the same velocities and are separated by relatively large liquid slugs (developed slug flow). In this case, liquid and gas flow rates averaged over a number of passing slug units, consisting of an elongated bubble and a liquid slug, are approximately constant. However, situations may occur where on a reasonable time scale, the average flow rates do fluctuate. One such phenomenon is severe slugging that has been described by Schmidt et al. (1980) and Taitel and Barnea (1990) for a vertical riser and terrain induced slugging (De Henau and Raithby, 1995). However less extreme examples of severe slugging have been observed (Taitel et al., 1990). Taitel et al. (1990) stated that when gas penetrates into a liquid filled riser, the void fraction in the riser tends to oscillate. These oscillations might become dampened and a steady state can be reached. It can however, also result in a quasi-steady cyclic process that resembles severe slugging but lacks the strong blowout that characterizes the severe slugging cycle. In this case the liquid flow rate is not completely cut-off while the gas flow rate into the riser might be stopped.

Usually in the design of an experimental facility one is interested in obtaining steady flow over the whole range of flow rates and experimental setups. This is however not always the case. The purpose of this brief communication is to relate the occurrence of unsteady flow regions to the design of the experimental facility and provide an answer on how to avoid unsteady flow phenomena that lead to fluctuating flow rates at the pipe outlet.

* Address: Department of Mechanical Engineering, Whiting School of Engineering, The Johns Hopkins University, 200 Latrobe Hall/3400 N. Charles Street, Baltimore, MD 21218-2686, USA. Tel.: +1-410-516-5427; fax: +1-410-516-7254.

E-mail address: vanhout@pegasus.me.jhu.edu (R. van Hout).

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2. Description of experiments

The experimental facility (Fig. 1) was designed to study gas–liquid two-phase flow in two parallel, 10 m long, transparent Perspex pipes with internal diameters of $D = 0.024$ and 0.054 m. The pipes could be inclined to any position from the horizontal to the vertical by two hydraulic pistons. Air was supplied from a central high pressure line and the pressure upon entering the rotameters was maintained at 1 barg. The air flow rate was measured by a set of three rotameters that were calibrated to measure the volumetric gas flow rates according to the atmospheric outlet pressure. Tap water was circulated in a closed loop through the system by a centrifugal pump. The liquid flow rate was controlled by a system of five rotameters. The two fluids were introduced through a "mixer"-type annular inlet device, 30 cm long with an inner pipe of the same diameter as the test pipe and an annulus of width 2.5 cm. The inner pipe was perforated with uniformly distributed 1 mm holes, 10 mm apart in the axial direction and 5 mm apart on the circumference in a staggered configuration. Water entered in the axial direction. Air was introduced into the annulus and entered the test section through the holes in the inner pipe.

The air supply pipeline volume, V_p, measured from the gas rotameter exit to the inlet device is an important parameter in the system design and the piping taken into account for the calculation of V_p is shown in Fig. 1 as a dashed line. The equivalent pipe length, L_e, of the air supply pipeline is

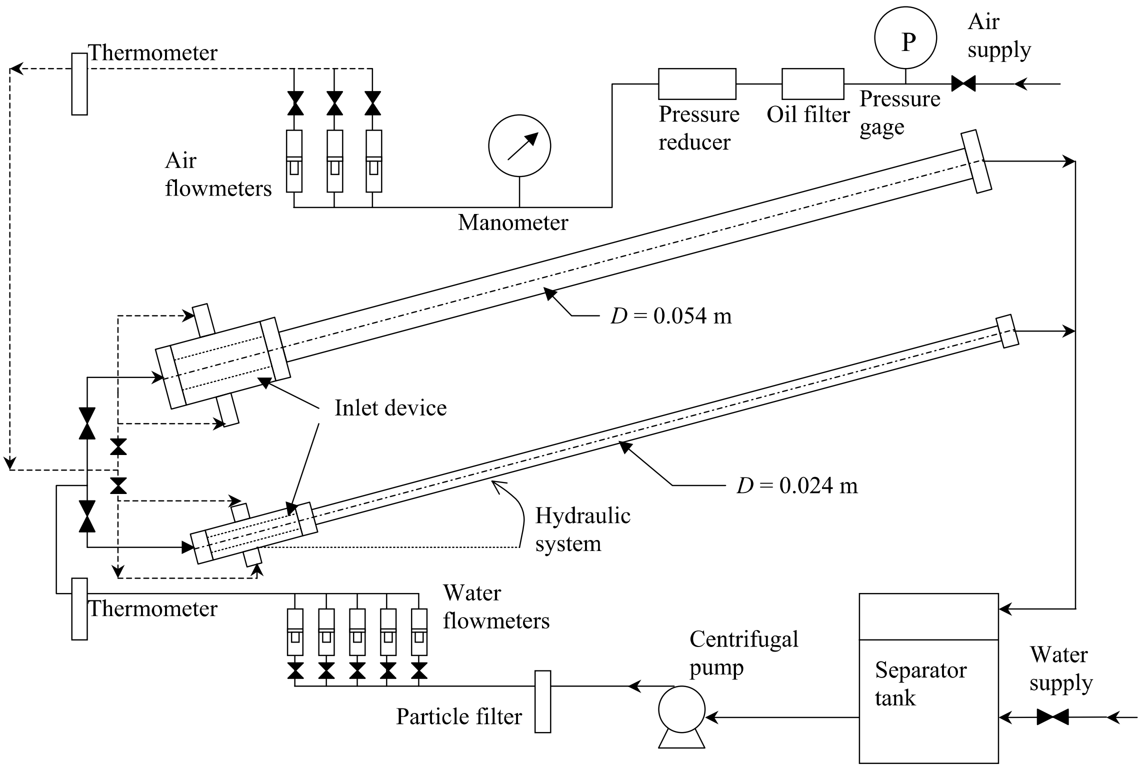

Fig. 1. Schematic layout of the experimental facility. (- - -) Piping taken into account for the calculation of V_p.

defined as the ratio between the pipeline volume, V_p, and the cross-sectional area of each test pipe separately. In the initial design of the facility, the equivalent pipe length was equal to $L_e = 23.4$ m for $D = 0.024$ m and $L_e = 5.6$ m for $D = 0.054$ m. In the course of the here described experiments on unsteady flow phenomena, L_e ranged between approximately $5.6 < L_e < 23.4$ m by changing the pipeline volume V_p.

The experimental facility was designed to study slug flow characteristics in upward flowing co-current flow at different inclination angles (from 0° to 90°) and to compare between the two pipe diameters at the same flow conditions. To this purpose, a set of flow rates was chosen such that slug flow existed in the whole range of inclination angles. Experiments were first carried out for the large pipe diameter and no unstable flow phenomena were observed. However, for the same experimental conditions, steady flow proved to be impossible to obtain for the small pipe diameter. In this case, unstable oscillating flow occurred for inclinations angles ranging between $30° < \beta < 70°$ and steady flow data acquisition was impossible at these inclination angles.

In the following a theoretical analysis is presented that can explain the unstable phenomena observed in the system. Guidelines for the proper design of a facility are proposed.

3. Analysis and results

The flow pattern maps for unstable two-phase flow in a system with an inclined riser can be determined by modifying the analysis presented by Taitel et al. (1990) for a vertical riser. A schematic layout of a system with an inclined riser filled with water is shown in Fig. 2. At some instance, gas penetrates the riser and the hydrostatic pressure at the bottom of the riser decreases.

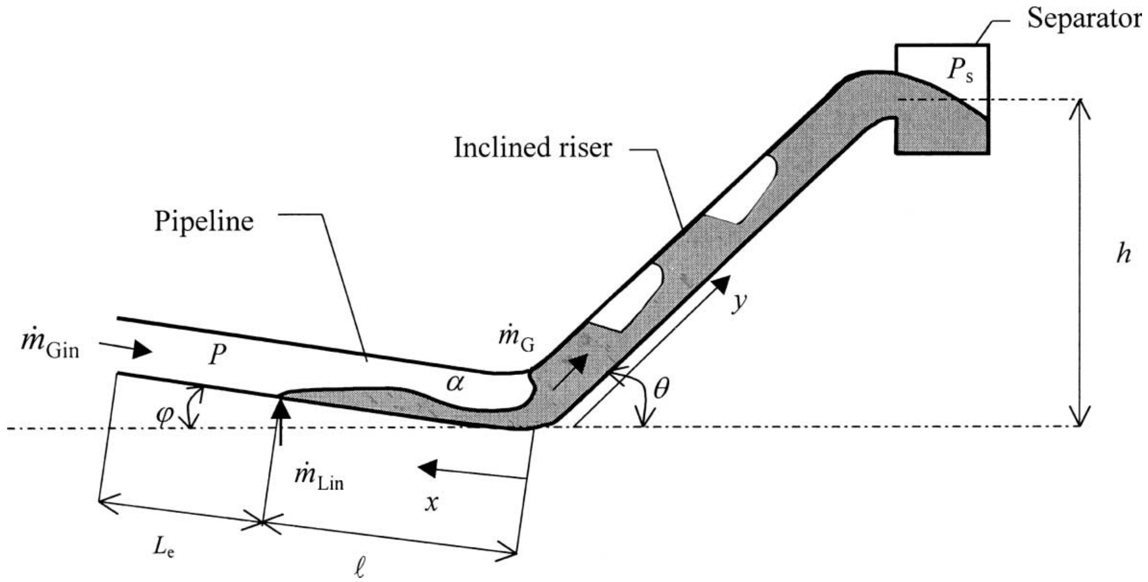

Fig. 2. Schematic layout of unsteady flow phenomena in the case of an inclined riser.

As a result, the mass flow rate of the gas into the riser further increases. Assuming ideal gas behavior, the instantaneous gas mass flow rate, $\dot{m}_G$, that penetrates the riser, is given by:

$$\dot{m}_G = \dot{m}_{Gin} - \frac{(\alpha\ell + L_e)A}{RT}\frac{dP}{dt}, \tag{1}$$

where $\dot{m}_{Gin}$ is the gas mass flow rate at the inlet, ℓ is the length of the corresponding stratified region (assumed zero here), α is the void fraction in the stratified region, L_e is the equivalent length of the pipeline, R is the gas constant, T the temperature and A the cross-sectional area of the pipeline. The pressure in the pipeline (at the bottom of the riser), P, consists of the separator pressure P_s and the hydrostatic pressure exerted by the weight of the liquid column in the riser:

$$P = P_s + \int_0^h \phi\rho_L g\sin\theta\, dy, \tag{2}$$

where ϕ is the local liquid holdup in the riser, ρ_L the liquid density, g the gravitational constant, θ the inclination angle of the riser measured from the horizontal and h the height of the inclined riser. Using Eqs. (1) and (2) together with the procedure formulated by Taitel et al. (1990), the variation of the pipeline pressure, $P(t)$, the gas mass flow rate into the riser as a function of time, $\dot{m}_G(t)$, and the local liquid holdup in the riser, $\phi(y,t)$, can be calculated.

The previous analysis can be implemented as long as gas enters the riser. When the gas flow rate into the riser becomes zero, the pipeline pressure, P, at hydrostatic equilibrium and for any time is given by:

$$P = \rho_L g\phi h\sin\theta - \rho_L gx\sin\varphi + P_s, \tag{3}$$

where $x = x(t)$ is the distance of the liquid interface penetrating into the pipeline and φ is the pipeline inclination. A mass balance on the gas in the pipeline requires that:

$$\frac{\rho_L g\overline{\phi}h\sin\theta - \rho_L gx\sin\varphi + P_s}{RT}[(\ell - x)\alpha + L_e]A = \frac{\rho_L g\overline{\phi}_i h\sin\theta + P_s}{RT}(\ell\alpha + L_e)A + \int_{t_i}^{t}\dot{m}_{Gin}\, dt, \tag{4}$$

where the $\overline{\phi}_i$ stands for the initial average liquid holdup at time t_i. Eq. (4) can be solved for $x(t)$ when the average liquid holdup as a function of time, $\overline{\phi}(t)$, is known (Taitel et al., 1990).

Using the above set of slightly modified equations together with the calculation procedure described by Taitel et al. (1990), the flow maps for unstable flow in vertical and inclined pipes can be calculated (Figs. 4 and 5). Different modes of operation exist. First of all, the regions of cyclic operation (regions I and V) in which at a certain point the gas flow rate into the pipe becomes zero, the liquid enters the gas inlet pipe until the gas pressure increases and the gas again enters the pipe. This cycle is repeated over and over. Two types of cyclic conditions can be distinguished. One is "with fallback" (region I), schematically depicted in Fig. 3a, and the other is "without fallback" (region V) as sketched in Fig. 3b. "Cyclic operation without fallback" is defined as a situation where the gas flow rate at the riser exit, $\dot{m}_G^{exit}$, is cyclically cut-off while at all times a liquid film remains present at the exit of the riser. "Cyclic operation with fallback" is defined when the

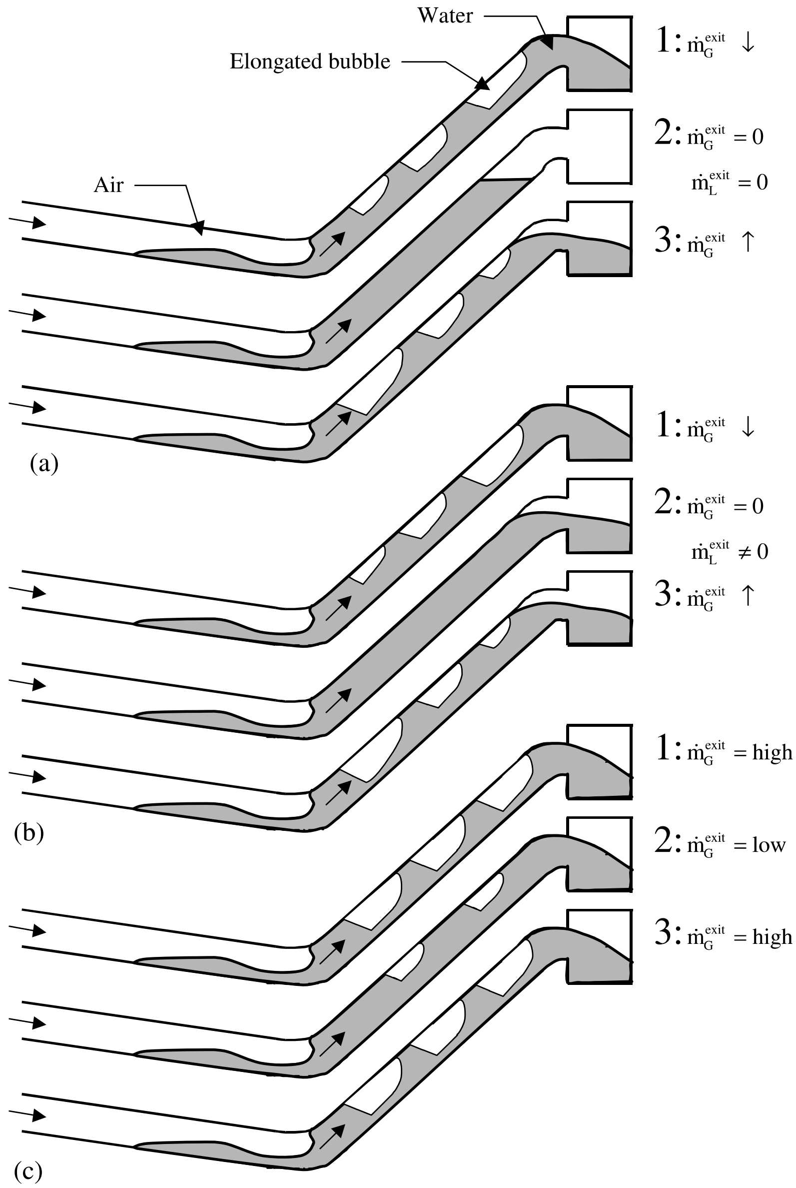

Fig. 3. Sketch of different stages (1–3) in periodic unstable flow patterns: (a) region I: cyclic operation with fallback, (b) region V: cyclic operation without fallback, (c) region II: unstable oscillations.

top of the riser becomes clear of liquid ($\dot{m}_L^{exit} = 0$) and a liquid interface propagates towards the top of the riser. When this situation occurs, the liquid height in the riser, z, is given by:

$$z = \overline{\phi} h \sin\theta, \tag{5}$$

and the calculation procedure is carried out as described by Taitel et al. (1990). In both cyclic modes, once $x(t)$ recedes back to zero, gas penetrates into the riser and the cycle is repeated.

Another type of cyclic unstable flow, termed unstable oscillations (region II), is schematically depicted in Fig. 3c. It occurs when the gas flow rate oscillates due to pressure fluctuations that are not dampened. However, in contrast with cyclic operation (I and V), the gas flow rate does not become zero. The transition from cyclic operation "with fallback" to "without fallback" (region I–V) is designated as region IV while steady flow as region III.

The effect of the inclination angle on the unstable flow regions is depicted in Fig. 4 for the small pipe diameter $D = 0.024$ m with equivalent pipeline length $L_e = 23.4$ m. A decrease in inclination angle strongly decreases region I, cyclic operation with fallback. The theoretical boundary between the unstable oscillations region II and steady flow region III is only slightly influenced by

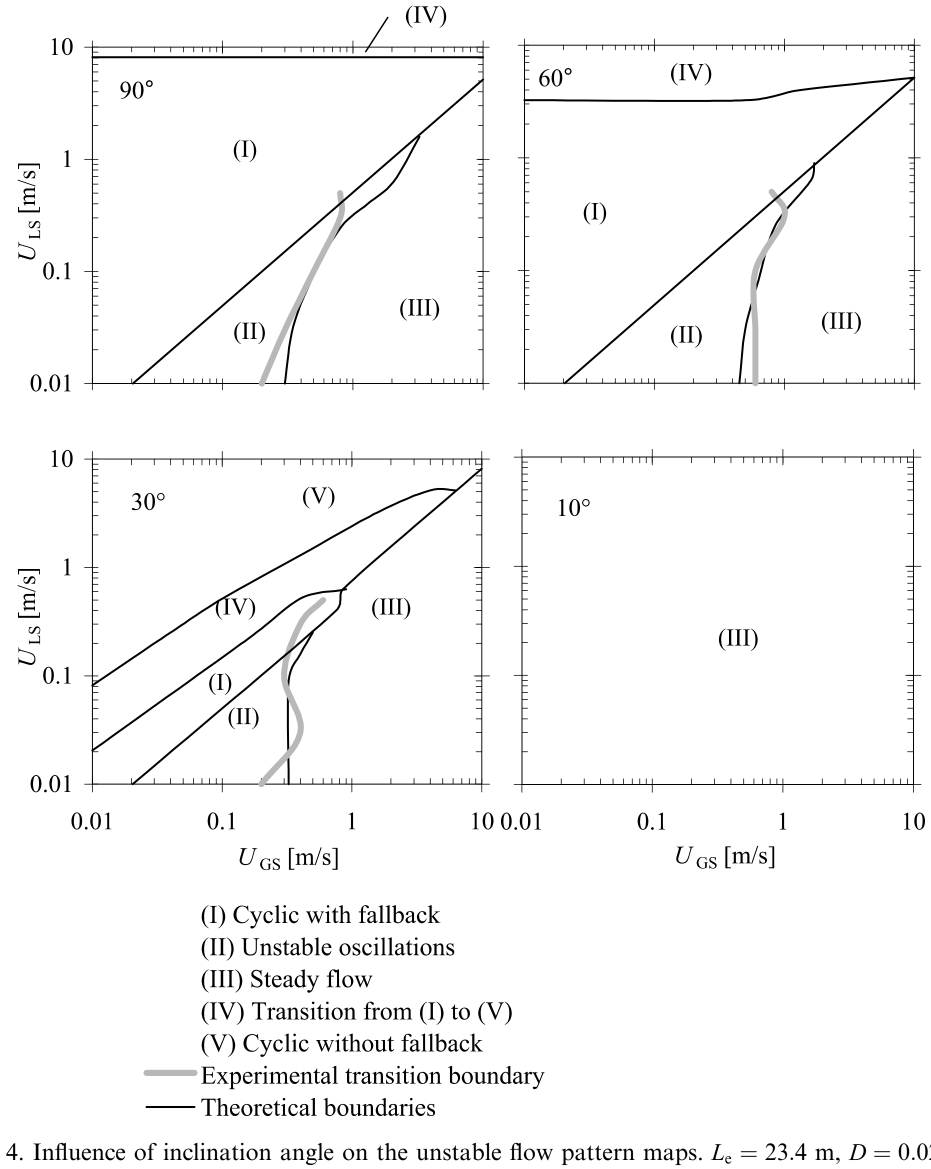

Fig. 4. Influence of inclination angle on the unstable flow pattern maps. $L_e = 23.4$ m, $D = 0.024$ m.

decreasing the inclination angle from 90° to 30°. At 10° inclination the whole flow map consists of steady flow. The theoretical boundary between region II and region III was verified experimentally and the results are presented in Fig. 4 as a solid gray curve. The experimental transition boundaries compare quite well with the theoretical boundaries for all inclination angles. Typical cycle periods were of the order of minutes depending on flow rates and inclination angle. For the large pipe, $D = 0.054$ m, and corresponding equivalent pipe length $L_e = 4.6$ m, stable flow was obtained at all inclination angles and for all flow rates.

The theoretical analysis together with visual observation of the flow have confirmed the existence of unstable flow regions for the small pipe diameter, $D = 0.024$ m. The question is now being asked on how to avoid these unstable fluctuations. Apparently the value of the equivalent pipeline length L_e strongly influences the occurrence of unstable flow regions and its effect on the unstable

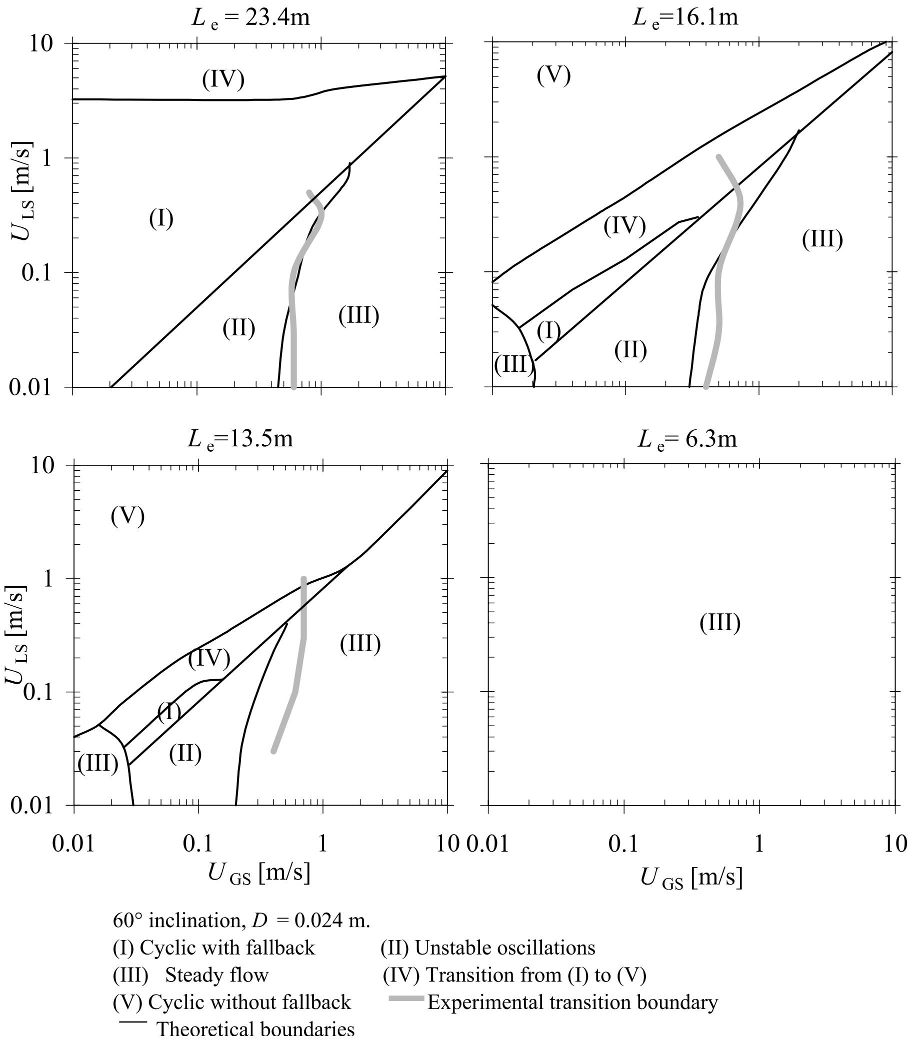

Fig. 5. Influence of the equivalent pipeline length L_e on the unstable flow pattern maps. 60° inclination, $D = 0.024$ m.

flow pattern maps was therefore studied. The results are presented in Fig. 5 for a 60° inclined riser, $D = 0.024$ m. The theoretical analysis shows that decreasing L_e results in an increase of the steady flow region III. The predicted boundary between the unstable oscillating region II and the steady flow region III was validated by visual observation of the flow and the boundaries are depicted in Fig. 5 as a solid gray curve. It can be seen that there is quite a good agreement between the theoretical and the experimental boundaries.

4. Concluding remarks

Unsteady flow phenomena that occur in experimental facilities are often dispatched of being intrinsic features of the complex two-phase flows that are the subject of investigation. This communication, however, has made it clear that this is not necessarily the case and that unstable fluctuations of flow rates can be avoided by a proper design of the experimental facility. In the present case, two risers with internal diameters of $D = 0.024$ and 0.054 m were fed by a common gas pipeline of relatively large diameter. For $D = 0.054$ m, compressibility effects in the pipeline were negligible and the resulting two-phase flow in the riser was steady at all flow rates and riser inclinations. However, for $D = 0.024$ m, compressibility of the gas in the pipeline strongly affected the resulting two-phase flow and unstable flow was obtained in large parts of the flow pattern maps at most inclination angles. Theoretical considerations showed that the unstable flow regions would shrink and finally disappear when the pipeline volume would be decreased. The theoretical boundary between the steady flow region and the unstable oscillations region was validated by visual observation of the flow behavior.

In general when designing a facility for two-phase flow where multiple pipes of different diameter are fed from a common gas inlet pipeline, care should be taken so that the gas pipeline volume is not too large. As a rule of thumb, the equivalent pipeline length L_e should be smaller than the length of the riser.

Acknowledgements

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