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SEVERE SLUGGING IN A RISER SYSTEM: EXPERIMENTS AND MODELING

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Abstract—Previous analysis showed that "severe" slugging would exist in a pipeline–riser system when the liquid in the riser is unstable and gas penetrates into the riser. This would then result in an unstable blowout and a cyclic process. When the liquid column is stable, a steady state is assumed to exist. However, observations made on a small-scale test facility have demonstrated that when the liquid column is stable there is still a tendency for a cyclic process to occur. This cyclic process can be damped and become a steady flow or it can continue indefinitely. From these observations a new theory is developed for predicting the behavior in the stable region and is verified with experimental results. It is also shown that in the region predicted by the Boe criterion to be a steady flow, the flow can be unstable and lead to a severe slugging type of behavior.

Key Words: severe slugging, pipeline–riser system, Boe criterion

INTRODUCTION

The process of severe slugging in a riser system was considered to consist of four steps (Schmidt *et al.* 1980; Taitel 1986): (1) slug formation; (2) slug movement into the separator; (3) blowout; and (4) liquid fallback. Taitel (1986) showed that a necessary condition for blowout is that the liquid column in the riser is unstable. Then, when the gas starts to penetrate into the riser, this quickly develops into an unstable blowout process. It was also assumed that a steady-state operation will result when the riser is stable.

Based on analysis of new experimental results we see that this is not necessarily the case. In fact, when the system is stable and gas penetrates into the liquid-filled riser, there is a tendency for the void fraction in the riser to oscillate. This oscillatory process can become damped and result in a steady-state two-phase flow (as was assumed before); but, it can also continue indefinitely in a quasi-steady cyclic process. The latter process does resemble the severe slugging cyclic process but lacks the spontaneous vigorous blowout which is characteristic of severe slugging.

This quasi-steady process can be described as follows. We start with the step when the riser is full of liquid and the gas first penetrates into the riser. As a result of gas penetration into the riser, the void fraction increases and its hydrostatic pressure decreases. Because of the pressure decrease and the corresponding expansion of the gas in the pipeline, the flow rate of gas into the riser increases. Once the riser is completely aerated, the pressure in the pipeline ceases to decrease and the gas flow rate into the riser decreases. This, in turn, causes a further increase of the liquid holdup in the riser resulting in an increase of the pipeline pressure. As a result, the mass flow rate into the riser decreases. As the rate of the mass of the gas entering the riser decreases, the rate of increase of the pressure with time decreases too. This causes the rate of gas into the riser to increase again. Thus, the end result is a cyclic process. This cyclic process becomes a steady state when the rate of penetration of the gas into the riser is always positive. However, it is also possible that the penetration of the gas into the riser becomes zero. In this case, liquid blocks the bottom of the riser. This is followed by a movement of the liquid interface into the pipeline and blocking of the gas passage into the riser. The gas pressure then increases and pushes the liquid back into the riser until the liquid interface reaches the bottom of the riser. At this point, penetration of gas into the riser starts and then a new cycle begins again.

When liquid penetrates into the pipeline the gas in the riser propagates to the top until all of the gas in the riser disappears. When the liquid input is very low, the propagation of the gas towards the top of the riser causes accumulation of all the gas at the top as the liquid falls back. This process is termed "cyclic process with fallback", while the former case is termed "cyclic process without fallback".

In summary, we identify three different possibilities than can occur as a result of penetration of gas into a liquid column in a quasi-steady "severe" slugging process:

1. Penetration of the gas that leads to oscillation, ending in a stable steady-state two-phase flow.
2. Penetration of the gas that leads to a cyclic operation without fallback of liquid.
3. Penetration of the gas that leads to a cyclic operation with fallback of liquid.

In this work, experimental verification of the aforementioned phenomenon is provided and a theoretical model capable of predicting this complex phenomenon is proposed.

ANALYSIS

Consider a riser system consisting of a pipeline of length l, an additional air line of length L and a riser of height h, as shown in figure 1. Inlet mass flow rates of the liquid and gas $\dot{m}_{\mathrm{Lin}}$ and $\dot{m}_{\mathrm{Gin}}$ are constant. Analysis begins at the point when the riser is full of liquid and gas is just entering the bottom of the riser under equilibrium conditions. We assume that the condition is stable (Taitel 1986) so that no blowout occurs as a result of the penetration of the gas into the riser. Nevertheless, when gas enters the riser the hydrostatic pressure at the bottom of the riser decreases. This causes an expansion of the gas in the pipeline. As a result, the mass flow rate of gas into the riser $\dot{m}_{\mathrm{G}}$ increases. Assuming ideal gas behavior, the instantaneous mass flow rate into the riser can be calculated by

$$\dot{m}_{\mathrm{G}} = \dot{m}_{\mathrm{Gin}} - \frac{(\epsilon l + L)A}{\mathrm{R}T} \frac{\mathrm{d}P}{\mathrm{d}t}, \qquad [1]$$

where P is the pipeline pressure, R is the ideal gas constant, T is the absolute temperature, t is time and ϵ is the void fraction in the pipeline, which is calculated using the stratified equilibrium flow model (Taitel 1986).

The pressure in the pipeline (and at the bottom of the riser) is the separator pressure P_{s} and the hydrostatic pressure exerted by the weight of the liquid column in the riser (the gas weight is neglected). Designating the local liquid holdup in the pipe as Φ, one obtains that

$$P = P_{\mathrm{s}} + \int_0^h \Phi \rho_{\mathrm{L}} g \, \mathrm{d}y, \qquad [2]$$

where ρ_{L} is the liquid density.

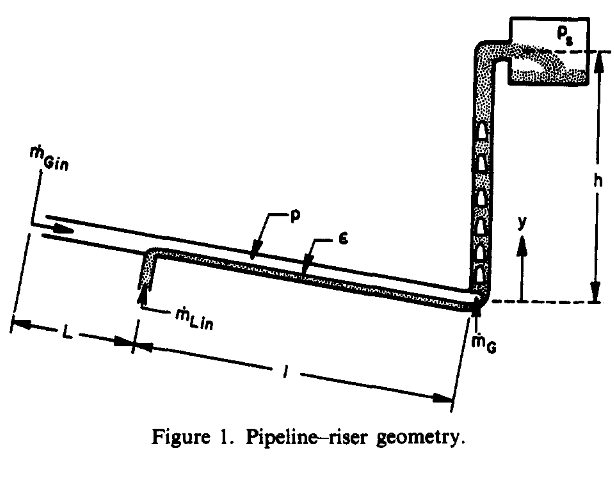

Figure 1. Pipeline–riser geometry.

The gas that penetrates the bottom of the riser is in the form of either small bubbles or larger Taylor bubbles. In either case the translational velocity of the bubbles, u_t, can be expressed in the form

$$u_t = Cu_S + u_d, \tag{3}$$

where u_S is the superficial (mixture) velocity and u_d is the drift velocity.

For slug flow $C = 1.2$ and $u_d = 0.35\sqrt{gD}$ (Nicklin *et al.* 1962) (D is the pipe diameter and g is the acceleration of gravity). For bubble flow we can assume $C = 1$ and u_d to be as given by Harmathy (1960):

$$u_d = 1.53\left[\frac{(\rho_L - \rho_G)\sigma}{\rho_L^2}\right]^{1/4}, \tag{4}$$

where σ is the surface tension.

In order to simplify the problem we consider the gas density within the riser to be constant. Therefore, the mixture velocity in the riser does not vary along the riser (although it is a function of time). For this purpose we calculate the average gas density as follows:

$$\bar{\rho}_G = \frac{\int_0^h (1-\Phi)\frac{P}{RT}\,dy}{\int_0^h (1-\Phi)\,dy}. \tag{5}$$

As can be seen in [5], the average gas density is calculated based on the local pressure in the riser weighted by the local gas void fraction $(1 - \Phi)$. The local pressure is given by

$$P(y) = P_s + \int_y^h \Phi\rho_L g\,dy. \tag{6}$$

Using [5], the superficial gas velocity in the riser is

$$u_{GS} = \frac{\dot{m}_G}{\bar{\rho}_G A}, \tag{7}$$

where A is the pipe cross-sectional area.

The liquid holdup at the bottom of the riser is given by

$$\Phi_b = 1 - \frac{u_{GS}}{u_t}. \tag{8}$$

The local liquid holdup in the riser is determined by simple propagation of the liquid holdup at the bottom of the riser with a velocity u_t. Thus, the local liquid holdup is calculated by

$$\Phi(y) = \Phi_b \quad \text{on} \quad y = \int_0^t u_t\,dt. \tag{9}$$

This mathematical formulation allows one to calculate the variation of the pipeline pressure, gas mass flow rate into the riser as a function of time and the local instantaneous liquid holdup in the riser, $\Phi(y, t)$. Although the formulation is somewhat complex, it is very simple to program using the explicit Lagrangian numerical scheme described below.

At time $t = 0$ the riser is full of liquid, $\Phi = 1$ and $\dot{m}_G = \dot{m}_{Gin}$. The average density of the gas at this time is the inlet density. The gas superficial velocity is given by [7] and the translational velocity is calculated by [3]. The riser is subdivided into small segments of length Δh and the time step Δt is calculated using $\Delta t = \Delta h/u_t$.

After time Δt, Φ_1 (at the bottom of the riser, $= \Phi_b$) is given by [8]; the new pressure is given by [2]; the new average gas density in the riser is given by [5]; and the new gas mass flow rate into the riser is given by [1]. Note that dP/dt in [1] is approximated numerically by the difference between the "new" and "old" pressure divided by Δt. Once the new $\dot{m}_G$ is known, the new gas superficial velocity u_{GS} is calculated from [7] along with the new translational velocity, u_t from [3] and the new time step $\Delta t(\Delta t = \Delta h/u_t)$.

At the next time step, the Φ_{j+1}s are set equal to Φ_j and this takes care of the propagation of the bubbles in the riser. Φ_1 is calculated as before.

This analysis can be used provided the penetration of the gas into the riser $\dot{m}_G$ is positive (which leads finally to a steady-state flow). Under certain conditions $\dot{m}_G$ becomes zero, for which case penetration of liquid into the pipeline occurs. Let $x(t)$ be the distance of the liquid interface penetrating into the pipeline. Under hydrostatic equilibrium the pipeline pressure at any time is

$$P = \rho_L g(\bar{\Phi} h - x \sin \beta) + P_s, \qquad [10]$$

where β is the pipeline inclination from the horizontal and $\bar{\Phi}$ is the average liquid holdup in the riser. A mass balance on the gas in the pipeline requires that

$$\frac{\rho_L g(\bar{\Phi} h - x \sin \beta) + P_s}{RT}[(l - x)\epsilon + L]A = \frac{\rho_L g \bar{\Phi}_i h + P_s}{RT}(l\epsilon + L)A + \int_{t_i}^{t} \dot{m}_{Gin}\, dt. \qquad [11]$$

Equation [11] can be solved for x as a function of time. For this purpose the average liquid holdup $\bar{\Phi}$ should be known as a function of time. The variation of $\bar{\Phi}$ with time can be calculated as before on the basis of the translational velocity u_t from [3]. The mixture velocity u_S is then calculated on the basis of the liquid mass balance to yield

$$u_S = u_{LS} - \epsilon \frac{dx}{dt}. \qquad [12]$$

At time t_i, $x = 0$ and $u_S = u_{LS}$ ($\dot{m}_G = u_{GS} = 0$). For time step Δt, we then calculate the new Φ distribution in the riser and $\bar{\Phi}$, the new x, the new u_S (approximating dx/dt numerically), the new u_t, the new time step Δt etc. As in the case of severe slugging, x increases to a maximum and then recedes back to zero. When $x = 0$ the cyclic process is repeated.

This calculation is valid provided no fallback occurs. A condition of fallback is defined when the top of the riser becomes clear of liquid (or liquid mixture) and a visible liquid interface is propagating towards the top of the riser. The condition of fallback is related to the net liquid velocity at the top of the riser. Once the liquid velocity is less than zero no liquid exits the riser, therefore resulting in fallback of the liquid in the riser. Thus, the point at which fallback occurs is when u_L is negative, where u_L is given by (simple mass balance)

$$u_L = \frac{u_S - u_t(1 - \Phi_{top})}{\Phi_{top}}. \qquad [13]$$

Once this situation occurs we calculate the liquid height in the riser by $z = \bar{\Phi} h$ and the calculation proceeds in the exact manner described by Taitel (1986; appendix A). In this calculation $x(t)$ as well as $z(t)$ are calculated on the basis of two equations: [14], which is a mass balance on the gas (similar to [11]),

$$\frac{\rho_L g(z - x \sin \beta) + P_s}{RT}[(l - x)\epsilon + L]A = \frac{\rho_L g(\bar{\Phi}_i h - x_i \sin \beta) + P_s}{RT}[(l - x_i)\epsilon + L]A + \int_{t_i}^{t} \dot{m}_{Gin}\, dt; \qquad [14]$$

and [15], which is a mass balance on the liquid,

$$z = z_i - \epsilon(x - x_i) + \int_{t_i}^{t} u_{LS}\, dt. \qquad [15]$$

Equations [14] and [15] are used to calculate $x(t)$ and $z(t)$. Once the slug reaches the top of the riser, then $z = h$ and $x(t)$ is calculated by [14] only. The values of x_i and z_i are the values of x and z at the time of fallback, namely when u_L becomes negative. As in the previous case, once x recedes to zero, the gas penetrates the riser and the cycle is repeated.

EXPERIMENTAL APPARATUS

The experimental data that were used to verify the theory were collected using a test facility that consisted of a pipeline 9.1 m long, connected to a riser of height 3 m (see figure 2). The pipeline and the riser pipe were constructed of 2.54 cm dia R-4000 clear PVC pipe and were mounted on aluminum I-beams. The pipeline is connected to the riser by a short flexible hose and is capable of being inclined at angles from $-5°$ to $+5°$ from the horizontal.

The riser flows into a 4.6 m high 20.3 cm dia PVC pipe that serves as a separator. This eliminates any syphon effects and allows easy back-pressure control.

Additional pipeline lengths, L, are simulated by two variable-volume tanks, as shown in figure 2. The tanks can each be used alone or they can be connected in parallel. The gas volume in the tanks can easily be adjusted by changing the amount of water in the tanks.

Fluid handling system

City tap water is used for the liquid phase and flows directly into the pipeline. A pressure regulator was placed at the entrance of the pipeline to eliminate any pressure fluctuations from the city supply line.

Air is supplied by a compressor and flows through two filters to eliminate any impurities. The air pressure is controlled by a pressure regulator. After being separated, the air is vented to the atmosphere and the water is dumped into the drain.

Instrumentation

The water flow rate is measured with a turbine flow meter and controlled by a metering valve upstream of the turbine meter. Orifice meters measure air flow rates. The flow rates are controlled by metering valves.

Important pressure in the system are measured with two pressure transducers, one located at the pipeline inlet and the other at the bottom of the riser. The separator pressure is controlled by a back-pressure regulator placed on the air vent line from the separator.

An IBM PC-AT with a LabMaster data acquisition package was used to gather the data for each run. The computer gathers the flow rate from the turbine meter, the differential pressures and static pressures from the air supply lines and the pressures from the transducers on the pipeline. Computer programs that control the calibration of the equipment and the actual data collection take the data and convert the computer binary numbers into actual flow rates and pressures.

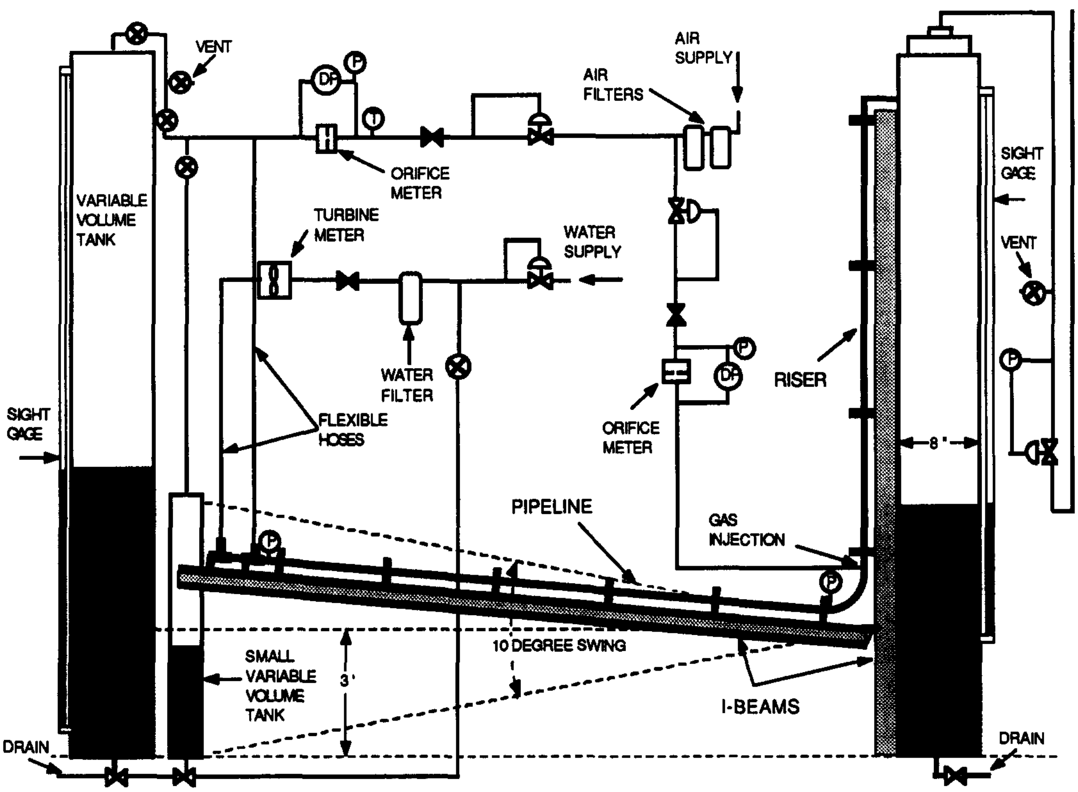

Figure 2. Schematic of the test facility.

A detailed description of the experimental facility and the data are given by Vierkandt (1988).

RESULTS AND DISCUSSION

The experimental runs reported here were collected for one pipeline inclination of $-5°$ and three additional pipeline volumes giving equivalent pipeline extensions L of 1.69, 5.1 and 10 m. For each volume the flow rate of liquid and gas was varied in the range $u_{LS} = 0.05$ to 0.5 and $u_{GSo} = 0.05$ to 1.0 m/s.

We first consider the theoretical results. Figures 3–5 present an example of the theoretical results for the pressure, gas mass flow rate into the riser and liquid penetration into the pipeline as a function of time. In all cases the additional pipeline volume is $L = 1.69$ m, the gas flow rate is $u_{GSo} = 0.1$ m/s and the liquid flow rates are: 1, 0.63, 0.4, 0.2, 0.13 and 0.07 m/s. Decreasing the liquid flow rate causes a change in the characteristic operation of the system. The analysis begins with equilibrium conditions where the riser is full of liquid and the gas is just ready to penetrate into the riser. For a high liquid flow rate the pressure will oscillate, but the oscillation is damped and a steady state finally results. As shown in figure 3, the damping rate is high for a high liquid flow rate, $u_{LS} = 1$ m/s, and decreases as the flow rate decreases. For $u_{LS} = 0.63$ m/s the oscillations are damped very slowly. This oscillation was previously described to be caused by the effect of the pressure variation on the flow rate of gas into the riser.

Figure 4 shows the relative gas mass flow rate into the riser ($\dot{m}_G/\dot{m}_{Gin}$). As can be seen, for the high liquid flow rate, the gas input into the riser fluctuates considerably before a steady state is reached, where $\dot{m}_G = \dot{m}_{Gin}$. This is the first type of operation and leads to a stable steady-state flow. The second and third possibilities described earlier occur for lower liquid rates, when the gas mass flow rate into the riser becomes negative. The pipeline is blocked, liquid penetrates into the pipeline and the gas is compressed above it.

Figure 5 shows the variation of the penetration distance with time for different flow rates $u_{LS} = 0.4$, 0.2 and 0.13 m/s. As can be seen, the maximum penetration increases and then decreases as the liquid flow rate is decreased. When the liquid penetrates into the pipeline the pressure increases, since at the same time the gas in the riser propagates towards the top, which clears the gas from the riser. As a result, the pressure increases and reaches a maximum value when x approaches zero. At this point the riser is clear of gas and the gas in the pipeline is just penetrating into the riser causing a new cycle to take place. The end result is a cyclic operation. This cyclic process can be subdivided into two types: type 2 which is cyclic without fallback; and, type 3 which is cyclic with fallback. The case of fallback occurs when the top of the riser is clear of liquid and a visible interface of liquid is propagating upward in the riser. In our example the cases of $u_{LS} = 0.4$ and 0.2 m/s are of type 2 (no fallback), whereas the case of $u_{LS} = 0.13$ m/s is type 3 (with fallback). As mentioned, in the calculations the criterion for changing from type 2 to 3 is a negative net liquid velocity (and flow rate) at the top of the riser [12].

The effect of the flow rates on the type of operation can be represented on a flow map, as shown in figure 6. In this figure the two broken lines separate the aforementioned three regions. The region

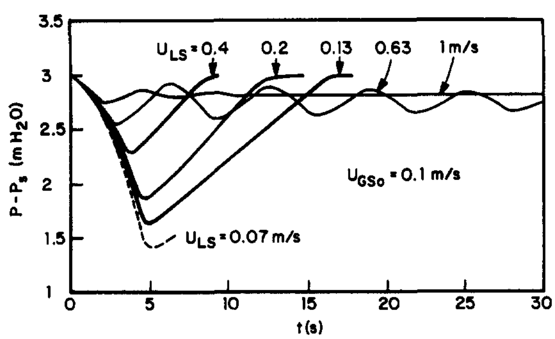

Figure 3. Pressure variation in the pipeline with time, $L = 1.69$ m.

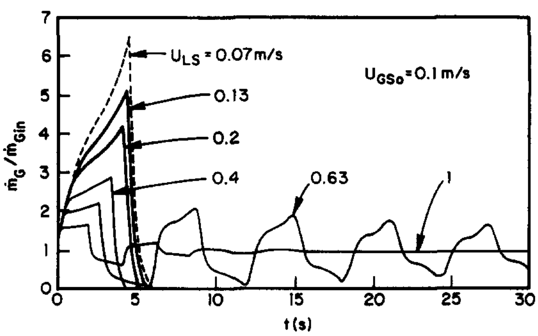

Figure 4. Gas flow rate into the riser, $L = 1.69$ m.

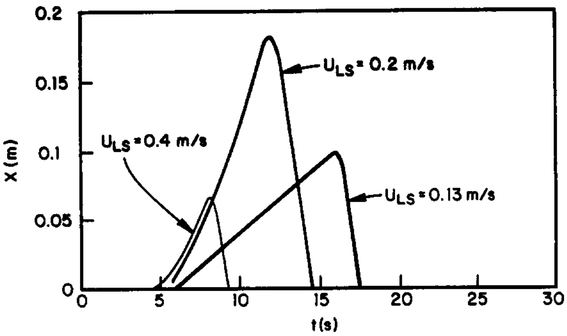

Figure 5. Penetration of liquid into the pipeline, $L = 1.69$ m.

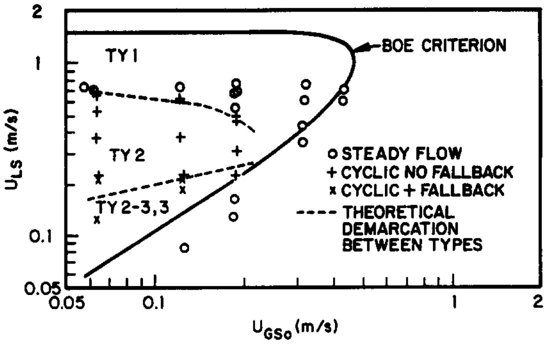

Figure 6. Flow map, $L = 1.69$ m.

above the upper broken line is where type 1 occurs, namely the system oscillates but reaches a steady state. The region between the broken lines is type 2 where cyclic operation without fallback occurs. The region below the the lower broken line is the region where a cyclic operation without fallback takes place (type 3 and type 2–3). Type 2–3 is a cyclic type that starts with no fallback but fallback occurred in the middle of the cycle, whereas in type 3 the fallback occurs immediately when the gas reaches the top of the riser. Thus, type 2–3 can be considered as a transition region from type 2 to type 3.

The heavy solid line is the Boe (1981) criterion for severe slugging. According to this criterion, steady-state flow results when the gas input is sufficiently high and the liquid flow rate is sufficiently low that liquid is not permitted into the pipeline. This line is given by the relation (see Taitel 1986)

$$u_{LS} = \frac{\rho_{Go} RT}{\rho_L g(l\epsilon + L)} u_{GSo}, \qquad [16]$$

where the subscript o refers to standard atmospheric conditions.

The original claim by this criterion was that outside the region of the line (see figure 6) the flow will be of steady-state nature, while inside severe slugging will prevail.

Equation [16] shows that at low liquid flow rates, where $\epsilon \simeq 1$, u_{LS} is a monotonic linear function of the gas inlet flow rate, u_{GSo}. For high liquid flow rates, ϵ approaches zero, and the curve is bent to the left. Note, however, that ϵ here is calculated while neglecting the gas shear (Taitel 1986). Thus, this upper limit is beyond the applicability of the present calculations. The validity of the Boe criterion is consistent with this analysis. Below the Boe criterion liquid does not penetrate into the pipeline and a steady state is reached. We termed this steady state as type 4 (the case of $u_{LS} = 0.07$ m/s in figures 3 and 4).

Figure 7 is a flow map for the case where the additional gas volume is $L = 5.1$ m. There is considerable difference between this case and the previous one where $L = 1.69$ m. The broken lines that separate the regions of the different types move considerably upward. The horizontal solid line seen in this figure is the Taitel (1986) criterion for stability:

$$\frac{P_s}{P_o} > \frac{\dfrac{\epsilon l + L}{\epsilon'} - h}{\dfrac{P_o}{\rho_L g \bar{\Phi}}}, \qquad [17]$$

where ϵ' is the void fraction for a Taylor bubble that penetrates into the riser. Taitel (1986) showed that ϵ' is of the order of 0.9.

Equation [17] for $\bar{\Phi} = 1$ is applied to the blowout phase of the severe slugging process. The region above the line is where [17] is satisfied and the gas penetrates into the riser in a steady, quasi-equilibrium fashion. Below this line a spontaneous blowout will occur when gas penetrates into the riser. It was assumed before (Taitel 1986) that for a steady penetration (above the stability line), a steady-state flow will result. Based on this study we see that this is not necessarily the case, and that in the stable region one may also obtain a cyclic process similar to the severe slugging process although lacking the vigor of the unstable blowout.

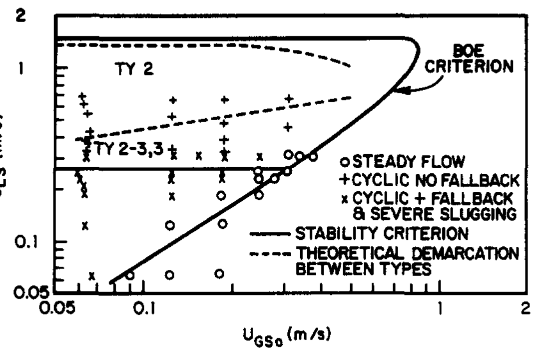

Figure 7. Flow map, $L = 5.1$ m.

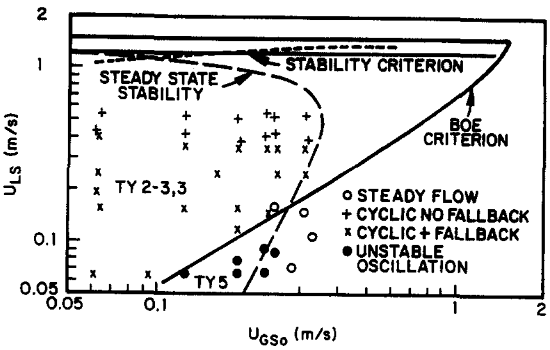

Figure 8. Flow map, $L = 10$ m.

We also found that as one moves closer to the stability line, the numerical procedure does not converge. The gas mass flow rate increases to infinity as we decrease the discretization, Δh. In spite of this theoretical shortcoming it appears that the mathematical procedure also does a reasonable job below the stability line because only the blowout phase of the solution is not converging. The other part of the solution related to the severe slugging process, namely the propagation of the liquid into the pipeline after fallback, is essentially valid as it is a strictly analytical solution of the slug formation process. Thus, although the stability line is not truly a sharp dividing boundary between severe slugging and steady operation, it is a dividing line between a quasi-equilibrium operation and spontaneous blowout expansion. Also, it separates the regions where the mathematical formulation converges and is well-posed and the region where the formulation is not converging and the results should be viewed with some reservation.

A different consideration of the stability criterion is the steady-state stability (Taitel 1986). We may consider a steady state, outside the Boe criterion, and discover that this steady state is unstable. In this case a severe slugging process will take place. Gas in the pipeline will spontaneously expand into the riser and a blowout will occur, followed by liquid fallback. Thereafter, gas will continue to penetrate into the riser and bubble through it while the liquid (mixture) level in the riser, z, rises towards the top of the riser. At the time the liquid level reaches the top of the riser, a steady state is expected to ensue. However, because of the inherent lack of stability, blowout will re-occur. This gives rise to a cyclic severe slugging process except that the slugs produced into the separator are aerated and shorter than the "classic" severe slugging slugs which are longer than the riser length.

The criterion for the existence of severe slugging under such steady-state conditions is obtained using a combination of [17] and the steady-state condition which relates the average liquid holdup in the riser to the flow rates.

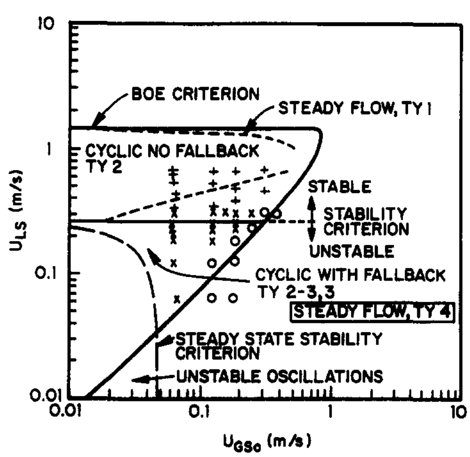

Figure 9. Overall flow map, $L = 5.1$ m.

Consider slug or bubble flow, where the translational velocity of the gas can be given by [3]. The void fraction of the gas in the riser is

$$1 - \Phi = \frac{u_{GS}}{C(u_{GS} + u_{LS}) + u_d}. \quad [18]$$

The gas superficial velocity in the riser, adjusting for the average pressure in the riser, is given by

$$u_{GS} = u_{GSo} \frac{P_o}{P_s + \dfrac{\rho_L \Phi g h}{2}} \quad [19]$$

Equations [17]–[19], in addition to the relation between the gas void fraction in the pipeline, ϵ, and the liquid superficial velocity (as before), are used to yield a steady-state stability line. This line is shown in figures 8 and 9 (for figures 6 and 7 the line is outside the map range). As can be seen, there is a definite region in which one can obtain unstable steady-state flow. As a result the flow will be cyclic, similar to the severe slugging cycle. We term this cyclic behavior "unstable oscillations".

Comparison with experimental results

The experiments and the theoretical results are summarized in tables 1–3 and in figures 6–8. Tables 1–3 present the experimental results for the type of flow obtained, the maximum penetration of the liquid into the pipeline, x_{max}, and the cycle time, t. Theoretical results for these variables are also given in tables 1–3.

In the theoretical results type 1 is a steady state that results due to the dampening of the oscillations; type 2 is cyclic with no fallback; type 3 and type 2–3 are cyclic with fallback; type 4

Table 1. Comparison between experimental and theoretical results for L = 1.69 m

		Experimental			Theoretical			
u_{GSo} (m/s)	u_{LS} (m/s)	Type	x_{max} (m)	t (s)	Type	x_{max} (m)	t (s)	Error t (%)
0.063	0.124	Cyc + fall	—[a]	24	2–3	0.28	22	−8
0.064	0.209	Cyc + fall	0.61	20	2	0.23	17	−17
0.123	0.183	Cyc + fall	—[a]	15	2–3	0.15	14	−9
0.124	0.212	Cyc + fall	—[a]	14	2–3	0.16	13	−9
0.062	0.679	Cyc no fall	—[a]	6	2	0	6	−3
0.063	0.367	Cyc no fall	0.53	13	2	0.1	11	−17
0.063	0.679	Cyc no fall	—[a]	9	2	0	6	−36
0.064	0.535	Cyc no fall	—[a]	10	2	0.03	7	−26
0.065	0.226	Cyc no fall	0.63	19	2	0.21	16	−17
0.122	0.374	Cyc no fall	—[a]	11	2	0.07	9	−19
0.123	0.621	Cyc no fall	—[a]	8	1	0	0	
0.126	0.228	Cyc no fall	—[a]	13	2	0.15	12	−7
0.187	0.226	Cyc no fall	—[a]	11	2–3	0.01	10	−8
0.188	0.466	Cyc no fall	—[a]	8	2	0	7	−19
0.188	0.502	Cyc no fall	—[a]	7	1	0	0	
0.19	0.312	Cyc no fall	—[a]	10	2	0.02	8	−16
0.058	0.705	Steady flow	0	0	1	0	0	
0.063	0.698	Steady flow	0	0	1	0	0	
0.122	0.730	Steady flow	0	0	1	0	0	
0.126	0.673	Steady flow	0	0	1	0	0	
0.126	0.085	Steady flow	0	0	4	0	0	
0.184	0.127	Steady flow	0	0	4	0	0	
0.185	0.161	Steady flow	0	0	4	0	0	
0.187	0.551	Steady flow	0	0	1	0	0	
0.188	0.755	Steady flow	0	0	1	0	0	
0.19	0.685	Steady flow	0	0	1	0	0	
0.313	0.433	Steady flow	0	0	1	0	0	
0.314	0.347	Steady flow	0	0	1	0	0	
0.319	0.614	Steady flow	0	0	1	0	0	
0.321	0.744	Steady flow	0	0	1	0	0	
0.43	0.604	Steady flow	0	0	1	0	0	
0.433	0.701	Steady flow	0	0	1	0	0	

[a]Not observed.

Table 2. Comparison between experimental and theoretical results for $L = 5.1$ m

		Experimental			Theoretical			Error (%)	
u_{GSo} (m/s)	u_{LS} (m/s)	Type	x_{max} (m)	t (s)	Type	x_{max} (m)	t (s)	x_{max}	t
0.060	0.252	Cyc + fall	1.42	33	2–3	1.25	36	−12	9
0.061	0.230	Cyc + fall	1.35	32	2–3	1.26	38	−6	18
0.063	0.206	Cyc + fall	1.30	34	2–3	1.24	39	−4	14
0.064	0.121	Cyc + fall	0.94	39	3	0.96	45	2	16
0.064	0.187	Cyc + fall	1.24	34	3	1.22	40	−2	18
0.125	0.231	Cyc + fall	0.77	19	3	0.87	21	13	12
0.126	0.184	Cyc + fall	0.58	20	3	0.68	23	17	14
0.126	0.253	Cyc + fall	0.84	19	2–3	0.93	21	11	8
0.187	0.254	Cyc + fall	0.00	14	3	0.56	15		5
0.187	0.250	Cyc + fall	0.00	14	3	0.55	15		6
0.066	0.063	Cyc + fall	0.00	40	3	0.26	48		19
0.063	0.320	Cyc + fall	1.24	27	2–3	1.12	28	−10	5
0.064	0.301	Cyc + fall	1.27	28	2–3	1.16	30	−8	6
0.065	0.307	Cyc + fall	1.24	27	2–3	1.15	29	−7	8
0.127	0.314	Cyc + fall	0.91	16	2–3	1.01	18	11	14
0.155	0.309	Cyc + fall	0.76	14	2–3	0.88	16	15	14
0.186	0.229	Cyc + fall	0.00	14	3	0.46	15		9
0.188	0.303	Cyc + fall	0.65	14	2–3	0.7	14	8	−1
0.250	0.311	Cyc + fall	0.00	11	3	0.43	11		−2
0.062	0.688	Cyc no fall	0.81	11	2	0.39	10	−53	−2
0.063	0.624	Cyc no fall	0.84	12	2	0.47	12	−44	1
0.064	0.378	Cyc no fall	1.22	24	2–3	0.98	24	−20	0
0.064	0.333	Cyc no fall	1.23	26	2–3	1.1	27	−11	4
0.065	0.546	Cyc no fall	0.89	13	2	0.59	15	−34	9
0.065	0.369	Cyc no fall	1.22	24	2–3	1	24	−18	0
0.066	0.433	Cyc no fall	0.85	21	2	0.83	20	−2	−5
0.126	0.342	Cyc no fall	0.93	15	2–3	1.03	18	11	20
0.126	0.525	Cyc no fall	0.99	11	2	0.79	12	−20	5
0.126	0.662	Cyc no fall	0.86	10	2	0.59	9	−32	−9
0.188	0.321	Cyc no fall	0.66	13	2–3	0.74	13	12	7
0.189	0.482	Cyc no fall	0.97	11	2–3	0.84	10	−13	−3
0.189	0.391	Cyc no fall	0.86	12	2–3	0.85	12	−2	5
0.189	0.324	Cyc no fall	0.69	13	2–3	0.75	13	9	5
0.190	0.660	Cyc no fall	0.84	9	2	0.62	8	−26	−15
0.309	0.469	Cyc no fall	0.57	9	2–3	0.53	8	−9	−17
0.309	0.673	Cyc no fall	0.53	8	2	0.48	6	−9	−28
0.090	0.064	Steady flow	0.00	0	4	0	0		
0.124	0.064	Steady flow	0.00	0	4	0	0		
0.124	0.123	Steady flow	0.00	0	3	0.27	25		
0.182	0.065	Steady flow	0.00	0	4	0	0		
0.185	0.184	Steady flow	0.00	0	3	0.24	16		
0.186	0.125	Steady flow	0.00	0	4	0	0		
0.247	0.255	Steady flow	0.00	0	3	0.24	12		
0.248	0.230	Steady flow	0.00	0	3	0.12	12		
0.250	0.186	Steady flow	0.00	0	4	0	0		
0.280	0.230	Steady flow	0.00	0	4	0	0		
0.307	0.257	Steady flow	0.00	0	4	0	0		
0.310	0.316	Steady flow	0.00	0	3	0.18	9		
0.338	0.309	Steady flow	0.00	0	3	0.05	8		
0.377	0.308	Steady flow	0.00	0	4	0	0		

is a steady state that is below the Boe criterion; and type 5 is unstable oscillation flow outside the Boe criterion. Note that types 2 and 3 can be of either a severe slugging or quasi-equilibrium nature, depending on whether they are above or below the stability line [17]. In the experimental results there is no distinction between types 1 and 4 since both result in a final steady state.

All results are reported for calculations in which the number of subdivisions in the riser is 50. As mentioned before, results for liquid flow rates below the stability criterion are nonconverging. In spite of this theoretical difficulty the results seem to be quite reasonable, even when used somewhat below the stability line. Table 3 shows that even in the worse case, when the calculations are well below the stability line, the error for both x_{max} and the cycle time is usually $<50\%$. The reason for the disagreement is due to the underprediction of the fallback in the blowout simulation. Note also, that the experimental results for $x_{max} \lesssim 0.5$ m were not observed, since the connection of the pipeline to the riser is opaque and therefore small values of x could not be observed. Also,

Table 3. Comparison between experimental and theoretical results for $L = 10$ m

		Experimental			Theoretical			Error (%)	
u_{Gso} (m/s)	u_{LS} (m/s)	Type	x_{max} (m)	t (s)	Type	x_{max} (m)	t (s)	x_{max}	t
0.061	0.064	Cyc + fall	0.85	56	3	0.94	73	10	31
0.062	0.191	Cyc + fall	1.79	47	3	2.37	68	32	45
0.063	0.247	Cyc + fall	1.98	47	3	2.64	66	33	39
0.063	0.405	Cyc + fall	1.66	36	3	3.22	62	94	72
0.064	0.157	Cyc + fall	1.80	50	3	2.09	67	16	34
0.094	0.064	Cyc + fall	0.00	39	3	0.29	48		22
0.123	0.357	Cyc + fall	1.42	23	3	2.40	32	69	39
0.124	0.157	Cyc + fall	0.97	27	3	1.22	35	26	29
0.157	0.249	Cyc + fall	1.17	21	3	1.58	27	35	26
0.185	0.118	Cyc + fall	0.00	21	3	0.15	24		13
0.185	0.155	Cyc + fall	0.00	20	3	0.57	23		17
0.186	0.351	Cyc + fall	1.07	16	3	1.88	22	76	35
0.232	0.351	Cyc + fall	0.97	14	3	1.52	17	58	23
0.233	0.147	Cyc + fall	0.00	16	3	0.12	19		17
0.247	0.349	Cyc + fall	0.94	14	3	1.41	16	50	15
0.249	0.246	Cyc + fall	0.61	15	3	0.83	17	36	12
0.304	0.339	Cyc + fall	0.77	12	3	1.03	13	32	11
0.311	0.247	Cyc + fall	0.00	13	3	0.45	13		4
0.124	0.065	Unst. osc.	0.00	32	5	0.00			
0.185	0.078	Unst. osc.	0.00	24	5	0.00			
0.185	0.066	Unst. osc.	0.00	24	5	0.00			
0.229	0.067	Unst. osc.	0.00	20	5	0.00			
0.230	0.091	Unst. osc.	0.00	19	5	0.00			
0.246	0.087	Unst. osc.	0.00	18	5	0.00			
0.062	0.433	Cyc no fall	1.65	36	3	3.32	62	101	73
0.064	0.538	Cyc no fall	1.65	29	3	3.58	58	117	99
0.124	0.414	Cyc no fall	1.47	20	3	2.64	31	79	57
0.124	0.523	Cyc no fall	1.22	14	3	3.03	30	148	116
0.184	0.513	Cyc no fall	1.22	12	3	2.52	21	107	71
0.187	0.375	Cyc no fall	1.09	15	3	1.98	21	81	42
0.228	0.405	Cyc no fall	1.14	13	3	1.80	17	58	33
0.230	0.543	Cyc no fall	1.24	12	3	2.33	16	87	36
0.245	0.527	Cyc no fall	1.22	11	3	2.18	15	79	40
0.247	0.416	Cyc no fall	1.16	12	3	1.72	16	49	32
0.307	0.532	Cyc no fall	1.09	11	3	1.85	12	69	12
0.313	0.385	Cyc no fall	0.89	12	3	1.20	13	35	6
0.247	0.158	Steady flow	0.00		3	0.14	17		
0.280	0.071	Steady flow	0.00		4	0.00	0		
0.308	0.149	Steady flow	0.00		4	0.00	0		
0.327	0.108	Steady flow	0.00		4	0.00	0		

the comparison of x_{max} with the theory when x_{max} is of this order (up to about 0.6 m) should be considered with caution. The test facility used a connecting section with a flexible rounded geometry, whereas the theory considers a sharp connecting section and interface that is perpendicular to the pipeline (in practice this interface is horizontal).

Comparison of the theory with the experimental data for the various types of flow are also summarized in figures 6–8. It is clearly seen that the theoretical results differentiate quite well among the different types of "severe" slugging above the stability line. Also, the theory can predict reasonably well when the flow will be steady state vs a cyclic nature. In figure 6 the stability criterion is below the map range and the whole map is within the stable region. Indeed, the flow is also of a quasi-equilibrium nature for all the cyclic operations. In figure 7 the stability criterion is in the "middle" of the region. The cyclic nature below this line is of a severe nature in which the blowout is spontaneous and vigorous, whereas above this line it is of much calmer nature. Admittedly, the change of this "severity" nature with decreasing flow rate is somewhat gradual and in this respect the stability line cannot be considered as a sharp demarcation line.

In figure 8 the whole experimental region is unstable and the entire range inside the Boe region is cyclic of severe nature. Note that in this case, the transition boundary between type 1 and 2 is missing, since for the case it occurs above the Boe criterion. Of special interest is the small region of unstable oscillation (type 5). This is the first time "severe" slugging behavior has been reported outside the Boe criterion with an almost perfect match with the considered stability theory.

In general, comparison of the experimental data with the theory is good. In particular, we stress the considerable success of the theory to describe qualitatively and quantitatively the different phenomena taking place for this complex system.

SUMMARY AND CONCLUSIONS

Experimental results of severe slugging in a pipeline–riser system show that one can obtain four different flow characteristics: steady flow; cyclic flow with fallback; cyclic flow without fallback; and unstable oscillations. The term severe slugging is used for either of the cyclic processes when the blowout process is "severe" or occurs as a spontaneous unstable expansion.

A new model is presented that is capable of predicting the type of flow and the flow parameters, such as local liquid holdup in the riser as a function of time, pressure fluctuations, liquid penetration into the pipeline, cycle time, blowout time etc. The theoretical results of this model differentiate between the four types of severe slugging. Type 1 is characterized by damped oscillations leading to steady flow. Type 2 is cyclic flow without fallback. Type 3 is cyclic flow with fallback. Type 4 is outside the Boe criterion, leading to steady flow or unstable osillations. Type 2–3 is a transition between types 2 and 3. It starts with cyclic flow without fallback, but fallback occurs while the liquid penetrates into the pipeline.

The Boe criterion [16] differentiates quite well between steady and cyclic operations with two exceptions. At high liquid flow rates a steady flow can also exist within the "severe slugging" region predicted by the Boe criterion. Also, there is a region outside the Boe criterion which is in an unstable steady state and leads to unstable oscillations.

The stability criterion [17] is applied to the case of severe slugging (inside the Boe region) where the riser contains only liquid, and to the case of steady flow of liquid and gas in the riser. The former is an approximate boundary dividing severe and "nonsevere" cyclic operations. The latter indicates when steady flow outside the Boe criterion is not possible and one obtains unstable oscillations. The stability criterion also serves as a guide when the aforementioned quasi-equilibrium theoretical model is valid.

Figure 9 is an enlarged scale map for the case of $L = 5.1$ m that show the different types and criteria used in this study.

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REFERENCES

BOE, A. 1981 Severe slugging characteristics, selected topics in two-phase flow. NTH, Trondheim, Norway.

HARMATHY, T. Z. 1960 Velocity of large drops and bubbles in media of infinite or restricted extent. *AIChE Jl* **6**, 281–288.

NICKLIN, D. J., WILKES, J. O. & DAVIDSON, J. F. 1962 Two-phase flow in vertical tubes. *Trans. Instn chem. Engrs* **40**, 61–68.

SCHMIDT, Z., BRILL, J. P. & BEGGS, H. D. 1980 Experimental study of severe slugging in a two-phase flow pipeline–riser pipe system. *Soc. Petrol. Engrs J.* 407–414.

TAITEL, Y. 1986 Stability of severe slugging. *Int. J. Multiphase Flow* **12**, 203–217.

VIERKANDT, S. 1988 Severe slugging in a pipeline–riser system, experiments and modeling. M.Sc. Thesis, Univ. of Tulsa, Okla.