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Slug flow control using topside measurements: A review

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ABSTRACT

Slugging flow is a condition caused by a liquid obstruction at the riser base. It exhibits cyclic behaviour. The cycle consists of a protracted time of no gas production at the riser's top, followed by the arrival of a liquid slug with a length greater than the riser height, and ultimately the breakthrough of a significant gas surge. The cycle time might range from a few minutes to a few hours, depending on the system size and flow conditions. In offshore oil production, feedback control is a practical and cost-effective way to prevent slug flow. To control the flow rate or the pressure in the pipeline, adjusting the choke valve opening on the topside facility is generally utilised as the control input. From a practical standpoint, designing a control system based on topside data rather than seabed measurements is preferable. Controlling the topside pressure alone is difficult and ineffective in reality, but combining it with the flow rate results in a more reliable control solution. Measuring the flow rate of a multiphase flow, on the other hand, is difficult and expensive. All the topside measurements-based slug control techniques was critically reviewed and necessary recommendations for enhanced control performance provided. In conclusion, this review acknowledged that slugging is a well-defined flow pattern, yet despite having been studied for several decades, current slug control methods still have robustness issues. Slug flow problems are expected to become even more intense in the future as a result of longer vertical risers driven by deep-water Exploration and Production (E&P).

1. Introduction

Flow assurance refers to the "guarantee of flow" that involves combining technology and science to assure safe and cost-effective hydrocarbon fluid production from reservoir to endpoint during the life of a well in any deep-water environment [1]. The most significant problem in deepwater oil and gas production wells is flow assurance in flowlines, which attracts enormous financial losses due to production downtime and pipeline damage due to high pressure, long lengths, and low temperatures that are common in deep-water environments [2]. With the recent rise of deep-water and even ultra-deep-water wells, it is critical to limit the hazards of slugging formation on flowlines due to the inherent high financial expenses and restricted accessibility. As a result, early detection of potential hazards and the implementation of suitable remediation techniques are required for successful well completion and smooth operation of a well. This might be accomplished by implementing an effective sampling and analysis programme that offers insight into tentative flow assurance concerns. Flow assurance is based on production and fluid characteristics theories, spatial distribution of phases (gas, oil, solids, and water), and multiphase flowing conditions (temperature, pressure, velocity) for a deeper understanding of flow conditions, which supports the design and operation of multiphase production systems during start-up, shutdown, metering, pigging, leak detection, and so on [3].

Subsea pipelines and risers transfer multiphase fluids from offshore oil wells to topside facilities. Slugging flow patterns occur in pipeline-riser systems when specific input circumstances exist, such as low pressure and low inflow rates. Significant flow and pressure fluctuations define such flow regimes [4]. Overflow of inlet separators, poor separation, and undesired gas flaring are all difficulties caused by these flow conditions in oil production [5,6]. The pressure variations cause strains on some parts of the system such as bends and valves. In most cases, the burden in the compressors and separators at the topside becomes so large that it causes a lot of damage and shutdown of the plant, this is a

Abbreviations: E&P, Exploration and Production; GLR, Gas Liquid Ratio; ISC, Inferential Slug Controller; OPC, Open Platform Communications; OLGA, Oil and Gas simulator software; PCA, Principal Component Analysis; PDO, Petroleum Development Oman; PT, Pressure at the Top; PI, Proportional Integral; PID, Proportional–Integral–Derivative; RHP, Right Half-Plane; S^3, Shell Slug Suppression System.

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huge disadvantage to oil producers.

The traditional approach for controlling slugging flow is to limit the opening of the topside choke valve (choking), however, this increases back-pressure in the producing oil wells and reduces production rate. As a result, a solution that ensures steady flow, as well as the highest feasible production rate, is preferable [6]. Slugging flow prevention or control offers significant economic benefits, which is why so much time and money has been spent looking for a reliable solution.

The benefit of applying active control is substantial as it relieves the system from strain, reducing unplanned maintenance. This is economically efficient and effective in comparison to the deployment of new equipment. A high production rate is more feasible with active control than the conventional choking of the topside valve.

Subsea multiphase measurements, for instance pressures at the riser base or further upstream, are normally included during control design. Subsea measurements have been proven to be effective in slugging mitigation/control [7]. Hence, in the absence of these measurements, slug mitigation/control gets tougher.

However, subsea measurements are often not readily available, expensive to implement, and less reliable than topside measurements. It is imperative to ascertain the possibility of slug control using only topside measurements. However, questions remain such as the possibility of improving the system performance by using a single topside measurement or combinations of topside measurements, and whether these results are the same or comparable to those obtained with controller based on subsea measurements.

In practise, measuring a multiphase flow is dependent on the density, velocity, and phase fractions of specific phases. The density and phase fraction measuring instruments are known to be either costly or imprecise. For example, gamma-ray sensors for detecting phase-fraction are expensive, and they must be calibrated for flow conditions and serviced regularly [11].

Simon et al. [12] thoroughly reviewed the state-of-the-art in the analysis, identification, dynamical modelling, and removal of slugs in offshore oil and gas Exploration and Production (E&P) operations. Talimi et al. [13] presented a review of numerical research on slug flow hydrodynamics and heat transfer in non-boiling two-phase flows at small (mini and micro) sizes. Simon et al. [14] reviewed some of the most recent findings and significant problems in slug identification, dynamical modelling, and slug flow control. They conclude that cost-effective and optimum slug modelling and control are still unresolved issues with several limitations. From previous research, slug flow control using only topside data appears to have received no or limited attention. Due to the difficulties encountered while employing subsea measurements, it is critical to undertake a comprehensive review focusing only on slug control utilising just topsides data. As a consequence of assessing existing slug flow control techniques, the current work creates a complete understanding of slug flow control. The findings of several slug techniques were gathered from various published papers in order to highlight the most significant measurement that may facilitate the control/mitigation of slug flow in oil and gas pipeline systems.

This review paper's contribution summary is as follows: (1) to analyse various multiphase flow control measurements techniques for slug flow control; (2) to the best of the authors' knowledge, this is the first review paper on slugging control based solely on topside measurements; (3) to report recent key findings in the literature on topside measurements for slug flow control; and (4) to present research gaps that need to be considered for future research work in the area.

This article is organised as follows. Section 2 describes slugging flow. Section 3 covers selected slug types, and Section 4 shows slug detection and criteria. Section 5 examines the slug flow measurements utilised in control. Section 6 provides an outlook for the future and necessary recommendations. Section 7 summarises the main points and conclusions.

2. Slug flow

The slugging phenomenon is a key flow assurance problem in multiphase flow. The flow regimes frequently observed with the liquid and gaseous phases of hydrocarbon in transit results in the production of slugs. Slugging is defined by Shotbolt [15] as an intermittent flow that results in alternating delivery of liquid and gas phases. This delivery is produced by a discrepancy in the phases' superficial velocities, which can generate liquid surges within the pipes. Slugging may be seen in a variety of system designs or offshore topologies, including the vertical, inclined flexible riser and the horizontal portion of piping resting on the seabed.

2.1. Properties of slugging flow

Multiphase flow in a pipeline can assume a variety of forms depending on the flow characteristics. However, these shapes may be categorised based on distinct sorts of fluid distributions, which are generally referred to as flow patterns. The study [16] revealed the most common multiphase gas-liquid flow patterns in vertical pipelines: *bubble flow* occurs when there is a dispersion of gas within the liquid, slug flow occurs when larger bubbles fill the diameter of the tube, causing gas pockets, *churn flow* occurs when the slug bubbles have broken down, resulting in a churn pattern, and *annular flow* occurs when the liquid flows on the pipe wall while the gas flows in the middle. A variety of elements influence the flow pattern, including the flow rates, gas-to-liquid ratio, diameters, and forms of the pipes, liquid, and gas material properties, and materials. Multiphase fluids may contain solid components such as sand, although the main phases are usually liquids and gases. Desanding occurs early in the process, for example, at the wellhead, at installations where the solid components have a substantial negative influence on the daily output.

In the oil and gas industry, a slug flow is a typical flow pattern. Under certain operational conditions, gas and liquid may not be dispersed uniformly across wells, transportation pipes, and risers. Gas can sometimes move through the pipeline in huge plugs. Slugs are a term for this occurrence [17].

Slugs can be induced by physical construction in both vertical and horizontal pipelines, as well as transient scenarios such as shut-down, start-up, pigging, changes in flow rate, or pressure set-points [7]. An example of terrain-induced slugs is a steep seafloor topography. Multiphase flow passing via a vertical well or riser can cause severe slug. In practice, process engineers describe a severe slug as a rise in the amplitude of pressure and flow oscillations; hence, reducing the amplitude of the pressure or flow oscillations would lower the negative impact of the slug [18].

2.2. Negative effects of slugging flow

The adverse effects of heavy slugging flow on production platforms are widely researched in [14,19,20]. According to the research, the following are the major negative effects of the slug:

Fatigue as a result of repeated impact: Due to the additional fatigue stress, fluctuating pressure might reduce the lifespan of a pipeline [20].

Reduced throughput: The daily average production rate is diminished, both because of increased friction in the liquid blowout phase and because the production rate is temporarily negative in the liquid fallback phase. Several published works such as [21–23], have demonstrated that employing feedback control to eliminate slug flow has the potential to enhance throughput. Unexpected shut-off of production can potentially cause production loss.

Overloading of gas compressors: The operating capability of the equipment can be exceeded because the slugs frequently create significantly higher pressures and flow rates than the system is intended for. This can be particularly difficult for compressors that are not built to manage such high-pressure rates [12,24,25].

Erosion: Erosion is a widespread problem in the oil and gas sectors, and repairing pipeline fittings damaged by erosion is highly expensive [26]. Slugs of liquid separated by areas of the gaseous phase can occupy the intact pipe cross-sectional area. The slug body is the major source of erosion because erodents are carried primarily by the liquid phase and the slug body has the largest liquid holdup. The bulk of the entrained sand moves in the slug body's fluids, causing erosion in pipes [27–29].

Speedup corrosion rate: It is generally known that slugs may speed up corrosion in pipes [30,31]. The higher flow fluctuations generate higher friction in the pipes. The large frictional pressure drop causes greater shear straining to the wall, which raises the corrosion rate [20].

Overflowing liquid and excessive pressure in the separators: The liquid blowout causes variable flow inputs to the separator. Although the separator serves as a buffer tank, significant slug blowouts might result in larger liquid volumes than the buffer tank can manage. Yang et al. [32] discovered that changing input flow can influence a controller's ability to minimise separator output flow, lowering separation efficiency. Additionally, Husveg et al. [33] demonstrated that poor separation in the separator has an impact on the performance of the entirety of the separation process. Thus, managing the slugs upstream of the separator would be preferred since the slug affects the effectiveness of the oil-water-gas separation, leading to a lower production quality, and difficulties for the produced water treatment.

Slop in production: To maintain a safe pressure, the slugs can force the generated gas to be released as waste via a gas combustion unit. This issue develops when gas surges cause pressures to increase over the safe threshold. As a result, gas flaring is used to manage the increasing volumes of gas in the separator. Avoiding the slug is essential for both safety and economic reasons. Slugging flow may be prevented by altering operational conditions, thus it is crucial to understand which flow regime occurs under which particular conditions. Flow maps have traditionally been utilised for this purpose. Flow maps have been utilised in research for decades [34,35] and they are still used today [36]. The maps usually show the connection between gas and liquid surface velocities and the associated flow regime. Experiments may be used to generate maps for each particular system. The flow maps demonstrated that significant slugs develop when both the surface fluid velocities are low. This was similarly concluded by [37] who also found experimentally that raising the gas or liquid superficial velocity causes the varying amplitude of the riser bottom pressure to initially increase owing to the accompanying higher friction. Consequently, in the end, this pressure will have decreased, resulting in a constant, non-slugging flow. Although flow maps can predict where slugs would develop, acquiring these maps may need extensive testing.

3. Types of slug flow

3.1. Severe slugging

Severe slugging, also known as riser-induced slugging, is a liquid obstruction phenomenon that starts at the riser base. It has a strong cyclic behaviour. The cycle consists of a protracted time of no gas production at the riser's top, followed by the arrival of a liquid slug with a length usually greater than the riser's height, and ultimately the breakthrough of a gas surge. The cycle time can range from a few minutes to several hours, depending on the system size and flow conditions. Fig. 1 depicts the five cyclical stages of severe slugging.

There are five characteristics of cyclic stages of severe slugging and they are as follows [38]:

- **Riser base blockage:** When production is low, the riser will function in the hydrodynamic slugging condition. The liquid will fall back and obstruct the riser base, causing a severe slugging cycle to occur.
- **Slug growth:** Liquid accumulates at the riser base, forcing a slug to develop upstream into the flowline and in the direction of the riser. During this period, liquid production is halted. To compensate for the increased hydrostatic head, gas entering the pipeline will build pressure upstream of the slug.
- **Production of liquid:** The terrible slug develops until the riser is filled with liquid. Production of the liquid slug into the production separator begins when the slug reaches the top of the riser.
- **Fast liquid production:** As the tail of the slug in the pipeline approaches the riser base, the hydrostatic head in the riser begins to drop, causing faster liquid production at the riser's top.
- **Gas blow-down:** This occurs when all of the liquid has been produced and all of the excessive gas has been discharged.

The riser-induced (Severe) slugging can take place if these four conditions are fulfilled:

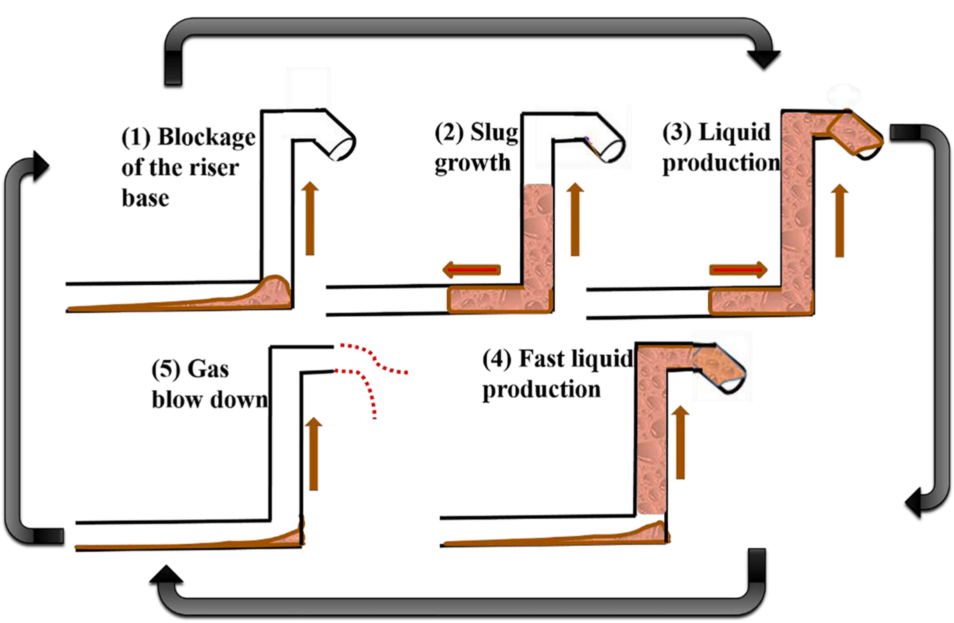

Fig. 1. Severe slugging cyclic stages[38].

1 The severe slug number (π_{ss}), which is a function of the flow condition of gas/liquid and the length of the flowline, needs to be less than one. This is because blockage of the liquid at the riser base can only happen if $\left(\frac{dp}{dt}\right)flowline < \left(\frac{dp}{dt}\right)riser$. The former pressure gradient $\left(\left(\frac{dp}{dt}\right)flowline\right)$ is the rate of increase in pressure at the pipeline upstream the riser-base because of gas compression while the latter gradient $\left(\frac{dp}{dt}\right)riser$ is the rate of increase in pressure at the riser-base because of the liquid head increase caused by liquid flowing into the riser [38]. As defined in Appendix A, the severe slugging number can be written as:

$$\pi_{ss} = \frac{\left(\frac{dp}{dt}\right)flowline}{\left(\frac{dp}{dt}\right)riser} = \frac{P_s}{(1-\alpha_F)\rho_L g L_F} GLR \tag{1}$$

where:

$P_s s$ denotes the pressure at standard conditions in Pascal units ($1.01325 \times 10^5\ N/m^2$).
α_F denotes the liquid holdup fraction in the pipeline upstream of the riser base.
ρ_L denotes the liquid density
L_F denotes the length of the pipeline upstream of the riser base
g denotes the gravitational acceleration
GLR is the Gas-to-Liquid volume ratio at standard conditions

In deriving this equation, we assume that (a) the riser is a vertical riser, (b) the riser and pipeline have the same diameter, and (c)$\rho_G \ll \rho_L$.

Thus, for a severe slugging to occur, $\pi_{ss} < 1$ is known as the Boe criterion, which expresses that a decreasing gas-liquid ratio (GLR) will reinforce severe slugging [39]. This happens in the field's late life, where the breakthrough of water will decrease the GLR.

1 If there is a low point at the riser base of the pipeline topography which can cause liquid blockage.
2 If the flowline is operated in an annular or stratified flow regime but not in a slugging flow. This condition is not strong enough, as some cases exist where a severe slug-like cycle happens with hydrodynamic slug flow in the upstream pipeline of the riser base.
3 Uneven riser flow. This indicates that a decline in production will induce a higher instead of a lower pressure drop over the riser. This happens below a certain flow condition where pressure drop becomes gravity dominated instead of friction dominated. In this case, low production yields an increased liquid holdup in the riser, which gives a hydrostatic head increase. A rough indication for the onset of an uneven flow is when the densimetric gas Froude number

$$Fr_G = \sqrt{\frac{\rho_G}{(\rho_{L-}\ \rho_G\ )}\frac{V^2{}_{SG}}{gD}} \text{ is lower than one.} \tag{2}$$

The summary of the criteria for severe slug flow can be seen clearly in the flow regime map for an upstream pipeline of the riser base as depicted in Fig. 7. Severe slug flow usually happens at the fields' late-life or production turndown when the riser system is in the hydrodynamic slugging regime and at low gas production.

As presented in Appendix A, the determined severe liquid length is L_R / Π_{ss} and the time is given as

$$\Delta t = \frac{L_R}{V_{SLs}} \frac{1}{\Pi_{ss}} \tag{3}$$

where V_{SLs}, is the pipeline superficial velocity at standard conditions while L_R is the height of the riser.

During severe slugging, a lot of energy is exhausted because of the gas and liquid velocities' regular variation in the line.

3.2. Transient slug flow

The transient slugs' behaviour being produced via the riser is as shown in Fig. 2. In phase 1, the slug is getting close to the riser. The slugs have already been initiated at the earlier phase. In phase 2, the slug velocity reduces because the slug pressure is not high enough to push the slug into the riser and the hydrostatic pressure has to be atoned for [40]. The pressure, therefore, builds up in the pipeline. In phase 3, the maximum pressure is attained and the slug moves into the riser. In phase 4, the slug gets to the riser top. The hydrostatic pressure does not have to be atoned for anymore, inducing the slug acceleration. In phase 5, a gas surge induced by high pressure behind the slug follows. Examples of transient slugging are growing slugs, start-up slugs, ramp-up slugs, and surge. Fig. 2 depicts the stages of five transient slugs entering the riser.

4. Slug detection and criteria

Slug criteria have typically been utilised in pipeline and equipment project designs to address slug issues early in the design stage. The major benefit of utilizing slug criteria is that they can be obtained quickly.

The study reported in [35] provided general criteria for stratified flow in horizontal pipes based on the maximum allowable surface gas velocity for stratified flow. This approach was commonly employed in conjunction with the research of [41] who identified the presence of stratified flow in a horizontal pipeline as a prerequisite before severe slugging could occur. The Bøe criteria, introduced in [42] expanded the [35] criterion further with assertions based on ideal gas law [41]. For a severe slug to form in the riser, the rate of gas accumulation at the riser base must be higher than the rate of pipeline gas pressure increase. The pressure rise is generated by the liquid column build-up in the riser during slug, which raises the hydrostatic pressure as well. [43] investigated similar criteria, which were similarly based on the accumulation of gas pressure in the riser and the generated hydrostatic head in the riser. This study developed a quantitative measure to indicate the degree of slugging to assess the severity of a specific slug.

The study presented by Taitel et al. [44] examined the definition of slug since the Bøe criteria states that if the liquid column is constant with no variation in the operating parameters, a constant steady-state exists. This assumption, however, was refuted by the study of Taitel et al. [44], who discovered that a cyclic process may still occur even when the liquid column is constant. Another research [45] finds that the Bøe Criterion is only applicable when no elimination procedures are used, which limits its practical application. Taitel [6] presented additional criteria [44]. This detection process relies on the gas holdup at the riser base, which is the void fraction of a Taylor bubble that enters the riser during slugs. A Taylor bubble, also known as a gas slug, is a huge uneven bullet-shaped bubble in a gas-liquid multiphase flow that occupies the whole cross-section of the pipe and has a length multiple times greater than the pipeline diameter [46]. This approach solely detects severe slugging from the superficial liquid velocity and nevertheless worthless when the gas velocity is taken into account. Fuchs [47] devised a criterion based on the severe slug's release time. The release time is the point at which the slug tail transitions from the horizontal pipeline to the riser, initiating the gas surge phase of the slug cycle. However, this

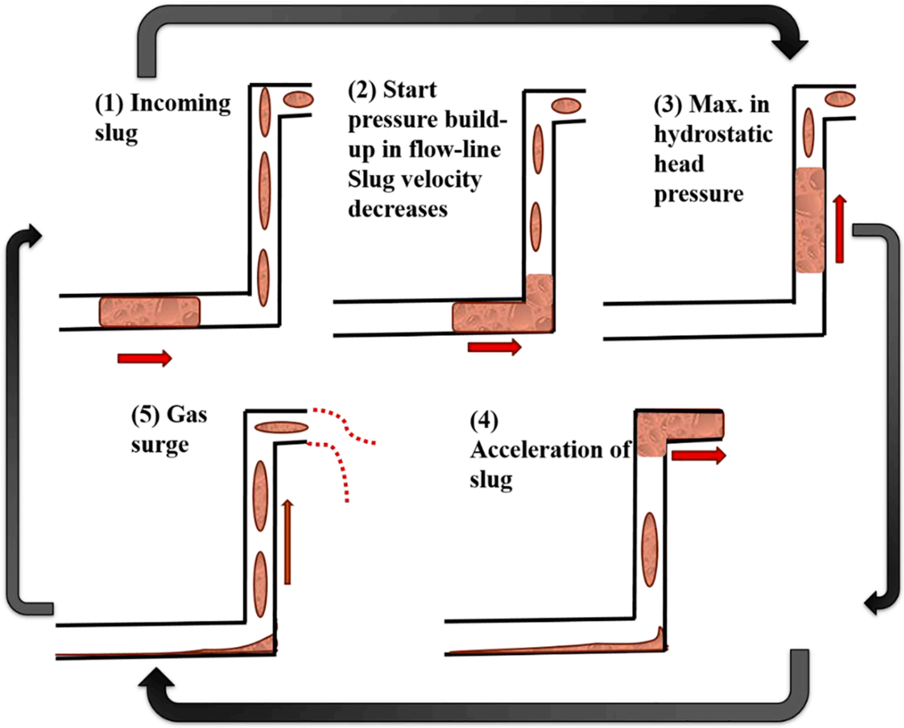

Fig. 2. The five transient slugs' phases entering the riser [38].

physical behaviour is not specific to severe slugging and is not usually distinguishable from a hydrodynamic slug or bubble movement. The work by Jansen et al. [45] proposed a modified version of the criteria from Taitel et al. [44], which includes valve choking and gas lifting but only with constant values. The work [48] proposed two variables for detecting gas-lift instabilities. If at least one of the variables is greater than one, the flow will be stable. Other methods, such as vertical well casing-heading instability [49] and slugs induced by S-shaped risers, have been used in recent years for specific subsections of the well-pipeline-riser constructions [50,51]. Online detection of the slug regime for controller decision-making has recently been researched. The study by Durdevic et al. [25] proposed two simple distinct criteria for controlling online detection as part of a monitoring anti-slug control scheme. The researchers presented two new slug criteria for detecting the slug regime in real-time:

i The first criteria depends on the changing rate of the bottom pressure over time, which is rising proportionally to the frequency of pressure fluctuations. A criterion for calculating the minimum frequency for severe slugging to occur was implemented.
ii The second criteria were based on the pressure drop across the riser because the hydrostatic pressure in the riser varies rapidly when slug exists. Both methods need a riser low-point pressure measurement, however, the first criteria may also apply to topside pressure measurements. The techniques can only forecast steady-state flow regimes under numerous assumptions, such as constant inlet flow rates and no closed-loop conditions with sensor data and feedback controllers because the slug criteria are often used in the early design stage. While there are ways for detecting slugs online, slug prediction for controller design is still in its development phase.

5. Slug control/mitigation

The goal of these slug management systems is to guarantee that the peak liquid and gas production levels at slug production do not exceed a predetermined maximum rate. This maximum is selected to ensure that the processing facilities' capacity is not exceeded. As a result, slug control will:

- Increase field life since it can withstand regular end-of-field life slugging.
- Reduce separator flooring and compressor trips.
- Expand the processing facilities' production capacity.
- Reduce deferment during transient scenarios such as ramp-up, pigging, and startup operations.
- Improve the processing facility's stability.

5.1. Slug control using only topside measurements

Some of the researchers' strategies to slug flow mitigation/control needed at least one or two subsea measurements. As a result, there are several drawbacks. In most cases, these subsea measurements are not commonly accessible, and their implementation to an existing pipeline/flowline incurs extra costs. Subsea measurement equipment must be inspected on a regular basis, calibrated, maintained, and perhaps replaced due to wear and tear. As a result, these are not cost-effective.

5.1.1. Single topside measurements

Topside pressure measurement

Diverging viewpoints have been expressed about the use of riser topside pressure measurement as a parameter for severe slugging control. According to Storkaas et al. [52] controllability study, riser top pressure alone is not a viable measurement for pipeline-riser system control. This is due to the fact that the associated transfer function's zeros are in the complex plane's right half-plane (RHP). From the research, it was revealed that to achieve an efficient control performance that the use of upstream pressure is the best and only measurement that can be used in a single loop controller for slug attenuation [53].

A controller-based only on the use of single topside pressure measurement was designed by Sivertsen et al. [10]. During the experiment conducted in the medium-scale rig riser, it was observed that achieving stability with this type of controller is not possible [10].

Topside density measurement control

Using a model established by Storkaas et al. [54], Sivertsen et al. [7] performed controllability analysis to screen the various measurement possibilities. According to the results of the research, it should be feasible to manage the flow using solely topside data. The findings of this study were then utilised as a backdrop for the tests conducted in the lab. As predicted by the study, the density measurement was inadequate. This measurement might be used to regulate flow, but only at an average valve opening greater than 17–18 percent, which is just inside the unstable zone. As a result, the advantages of employing control are insignificant [53].

Riser flow rate Measurement Control

The riser flow rate has also been explored for stabilising slugging flow. The volumetric flow rate can be utilised as a controlled variable for stability. Storkaas' model-based research revealed that utilising volumetric flow rate as a controllable variable results in poor performance at low frequency. The author also stated that employing volumetric flow rate in the inner loop of a cascade control system to manage the flowrate out of the riser might improve performance.

Topside differential pressure control

Somtochukwu et al. [5] conducted an experimental study into the practicality of active riser slug suppression by measuring topside differential pressure from the venturi flow meter's inflow to the throat. When compared to manual choking, experimental findings show that with active slug flow control, the system was able to remove slug flow at a greater valve opening.

Estimated pressure drop over the riser

Di Meglio et al. [55] conducted an experiment to examine innovative control solutions, relying on a reduced-order slugging model presented in [56]. It is made up of a three-state system of nonlinear ordinary differential equations (ODEs) that reflect mass conservation rules in two volumes of gas and one of liquid, and it can replicate pressure and flow rate fluctuations that correlate to slugging behaviour.

The study [55] gives the four control methods that were subjected to comparison testing.

i Controller 1: PI controller on the bottom pressure: The most advanced control technique employs a PI feedback control law with the bottom pressure as the output. Several effective implementations of this technique have been described in the literature [57,58], and precise tuning criteria for the proportional and integral gains of the PI may be found in [9]. The main disadvantage of this approach is that it is dependent on the presence of a bottom pressure sensor: when the sensor fails, as stated in [59] the slugging restarts.

ii Controller 2: PI controller based on pressure drop over the riser: The authors of [60] use the pressure drop across the riser as the output of a PI feedback law to stabilise an experimental slugging riser with a low point. Surprisingly, there have been relatively few additional attempts to construct such a controller described in the literature. It should be noted that it, similar to Controller 1, requires a bottom pressure sensor.

iii Controller 3: PI controller based on the expected pressure drop through the riser (topside): This controller is quite similar to Controller 2, except that it only employs the observer's estimations, which are derived using the topside pressure sensor. Even in the lack of a bottom pressure sensor, this approach allows one to stabilise slugging.

iv Controller 4: A nonlinear stabilising controller (topside): For the model, the authors suggested a nonlinear control rule. The primary impact of this feedback law is to linearize the dynamics of the riser's liquid mass.

The experimental findings show the potential of the suggested control methods, which overcome two significant weaknesses of the reference PI controller: they can control the flow for higher valve opening set points and give a viable solution when no bottom sensor is present. The PI controller, in particular, utilising the pressure drop above the riser as the output, is a simply practical control law that does not need any extra sensor to the reference control solution (PI on the riserbase pressure) or the usage of the model. Furthermore, the suggested model-based nonlinear feedback rule compensates for the lack of a bottom pressure sensor nearly entirely. The results are presented in Table 1

Positive findings from model-based control methods would most likely be reduced during real-world deployments. The model, in particularly, may not be realistic over a wide variety of operation conditions on a real well, because inflows often vary with bottom pressure, but they are presumed constant in the model. These issues might be addressed by using adaptive methods or including pressure-driven inflows into the model. On-line adaptation might also be utilised to reduce part of the off-line tuning work, which is still time-consuming. Furthermore, the constraints imposed by unstable zeros when no bottom pressure sensor is present may be limiting for a practical system [55].

Additionally, the list of potential controllers explored in [55] is far from complete. New contributions could compare systems incorporating various alternative positions for pressure sensors, particularly at the entrance of flowlines, as well as other different sensors. It is demonstrated in [7] that cascade controllers may be utilised to accomplish stability by combining topside density, flow, and pressure data. Surprisingly, these sorts of measures have the potential to improve the model-based approach's output (Tables 2 and 3).

5.2. Combination of topside measurements

The idea of controlling slugging flow using cascade configuration arose due to the inability of using topside measurement as a single variable in achieving an effective control performance. Considering the controllability challenges observed using topside measurement as mentioned above, the cascade method was implemented with the hope that it will proffer a better solution [53].

Storkaas and Skogestad [8] were the first to suggest cascade control for slug flow stabilisation. Godhavn et al. [9] and Sivertsen et al. [10] then tested it experimentally.

Godhavn et al. [9] carried out experiments at the SINTEF Petroleum Research Multiphase Flow Laboratory in Tiller. The goal of the experiment was to validate the results reported offshore at Heidrun, test novel methodologies, and establish easy tuning guidelines for future offshore installations. The flow was controlled using pressure and density data. This research, however, did not compare the data obtained with what may be obtained utilising subsea measurements in the control system. The author also presented a cascade control method in which the top pressure (upstream of the control choke) was used in a slow outer loop and the flow through the choke was used in an inner loop. To the best of the author's knowledge, this control system has never been published. The flow line pressure at the riserbase reacts differently from the top pressure. The sum of the pressure downstream the choke (separator pressure) and the differential pressure above the choke produces the top

Table 1
The experimental results of the 4 control method investigated [55].

	Pressure sensor(s) used	Control variable	Maximum valve opening
1	Bottom	Bottom pressure	27%
2	Bottom &topside	Pressure drop over the riser	36%
3	Topside	Estimated pressure drop over the riser	30%
4	Topside	Estimated states of the model	35%

Table 2
Slug control using single topside measurements.

Measurements	Results	Drawbacks	Reference
inlet pressure	Based on the controllability study, the inlet pressure was found to be the sole parameter that could be utilised as a single measurement for anti-slug control.	Offshore, subsea measurements may not be accessible. Subsea measurements are capital intensive	[61]
Topside density or pressure	It was unsuccessful	Encountered problems due to RHP pole/zero location when using a topside density measurement or pressure as single measurements for control. In addition, problems due to drift (low steady-state stationary gain) are possible.	[10]
Volumetric flow rate	The use of volumetric flow rate as a controlled variable leads in poor performance at low frequencies.	Practically impossible.	[73]
Density	As predicted by the study, the density measurement was inadequate. This measurement might be used to regulate flow, but only at an average valve opening greater than 17–18 percent, which is just inside the unstable zone. As a result, the advantages of employing control are insignificant.	Very noisy measurements.	[7]
Differential pressure from Venturi inlet to the throat	It significantly boosted hydrocarbon output by controlling the slugging flow.	Capital intensive.	[5]

pressure. The pressure at the riserbase is given by the top pressure plus the hydrostatic pressure produced by the weight of the contents in the riser, friction loss, and pressure decrease due to riser acceleration. In the following, friction and acceleration will be disregarded. Observations revealed that the top pressure exhibits non-minimum phase behaviour, as it has a short reaction in the opposite direction to the stationary response. The bandwidth of non-minimum phase systems is upper limited, which is a well-known fact in control theory. This implies that the PID controller has a long integration period and a low gain. The differential pressure over the choke, defined as the difference between the top and separator pressures, is a direct measure of controllability and durability. A high differential pressure gives controllability and durability, whereas a low differential pressure gives good performance with maximum output.

Sivertsen et al. [61] eliminated the requirement for subsea measurements by proposing a cascade design that used topside pressure in the outer loop and mass flow in the inner loop. It was initially evaluated in simulations using a simpler riser-slug model. The findings revealed that the cascade controller was successful in stabilising the flow. The cascade controller performed poorly in testing on a small two-phase loop. The cause for this was that the volume flow was calculated using hold up data that had to be severely filtered. If superior measurements for pipeline hold-up were available, the cascade design might still be used. If stronger measurements for pipeline hold-up were available, it is still feasible that the cascade design might produce superior results. Capacitance-based instruments, rather than the fibre optic slug sensors employed in the research, would have been a preferable option. Other, more robust control systems may also be able to solve the problems better than the cascade model. One such example is the H∞ controller, which they want to test in future studies.

Table 3
Slug flow control using measurements combination.

Measurements	Results	Drawbacks	References
Pressure and density	Proven to have a very strong slug suppression performance, allowing the operating pressures to be reduced and the production capacity to be significantly increased.	The results were not compared to what may be obtained in the control structure utilising subsea measurements.	[65]
Topside pressure and estimated mass flow	The cascade controller was successful in stabilising the flow.	When the cascade controller was tested on a small two-phase system, the findings were not promising. This was due to the volume flow being calculated using hold up data that had to be severely filtered.	[61]
Inlet pressure, pressure upstream production choke, density, mass flow rate, and volumetric flow rate through the topside choke	Using the density and mass flow controller, it was feasible to control the flow in the unstable region.	The controllers were slow and did not manage to stabilize the flow very far into the unstable region.	
The topside density results were not as efficient as the inlet pressure and topside volumetric flow rate measurements, but the flow could still be controlled.	[10]		
Virtual flow (pressure drop across the valve and valve opening)	This virtual flow is measured without taking into account phase fraction or density, yet it yields adequate results for stabilising control.	A further frequency-domain study is necessary to thoroughly examine these problems.	[66]
Combination of any available measurements	Controlled the severe slugging and increased throughput.	The use of gamma densitometer.	[67]
Liquid density, total liquid mass flowrate, riser top pressure, and gas mass flowrate	The controller was able to control the slugging flow far into the unstable region.	No known setback was recorded	[64]

Sivertsen et al. [7], performed further experiment to evaluate/analyse the effectiveness of using cascade control configuration to attenuate or eliminate the severe slug flow in a multiphase medium-scale rig riser. The following variables were tested in cascade configuration:

- The topside volumetric flowrate in the inner loop and topside choke valve opening in the outer loop.

- The topside volumetric flowrate in the inner loop and pressure in the outer loop.
- The topside density was used in the inner loop while the topside choke valve opening was used in the outer loop.
- The topside density in the inner loop and the topside pressure in the outer loop.

The measurement combinations used in cascade control were able to suppress the slugging flow but after some time, the system becomes unstable again. Hence, it was also deduced that the controller can suppress the slugging flow only when they are activated on a particular flow condition order wise stability will not be achieved. More so, there was no structured approach (systematic approach) when testing the performance of the controller in that some of the results obtained were when the controller was already activated (switched on). Because of that, the time required to stabilize the flow was not estimated. Due to the lack of an effective systematic approach during the experiment, repeatable flow behaviour was not obtained because of the inability to start the analysis during a severe slugging flow regime. This type of chaotic (nonlinear) system will affect the results significantly due to flow hysteresis. With regards to the above-mentioned issues, it is clear that the control system described is not good enough to be used in the industry.

In the work of Esmaeilpour [62], cascade control deploying topside pressure with density was investigated. One of the objectives in the work was to run the closed-loop deploying a cascade control system with topside density and pressure as the control variables. Thus, the idea was to tune this control system by trial and error and thereby examine the controllability features in comparison with the single-loop system. The control variable of the inner loop was the outflow density and the control variable for the outer loop was top pressure. The conductance probe was the device deployed to measure the outflow density. Hence, it was observed that the conductance probe used to measure density is not a good measuring device for it. From the analysis, it was deduced that the density signal could not be the appropriate measurement for control because the density signal is very noisy and does not indicate a good response to step changes. Hence, to achieve effective cascade control using density as a control variable in the inner loop, more clear and accurate density signals are essential.

Cao et al. [63] invented a novel method to suppress slugging flow called Inferential Slug Control (ISC). This method was known for its effectiveness to detect and suppress hydrodynamic slugging flow in real-time, subsea production facilities. This novel technique implements multiple topside measurements signals to suppress slugging flow in the offshore pipeline-riser system. The topside measurements that were fed into the controller are the pressure at the outlet of the riser, the separator pressure at the topside, the separator pressure of the 3-phase, the topside separator liquid level, the topside separator gas outlet flow rate, the topside separator liquid outlet flow rate, the mass flow rate riser outlet measurement from Coriolis metre, the density measurement of the riser outlet from Coriolis metre, the riser outlet soft count from Gamma metre, the riser outlet hard count from Gamma metre.

This idea of applying a linear combination of all available topside measurements in designing a novel slug control system has been investigated both in OLGA simulations and experiments. The benefits of this approach are numerous such as:

- Increase in oil production: The PC1 which denotes the combinations of measurements most sensitive to slug flow, when used as the controlled variables, the control choke valve will be maintained at a larger valve opening to suppress the flow. Hence, increasing the production rate.
- Sensitive to slugging flow: The more the available measurements deployed, the more sensitive the system will be to slugging flow.
- Simple and easy to implement: The PC1 variable can be designed by deploying any available topside measurements in as much as the measurements will react to slugging flow.

However, inferential slug controller (ISC) controller gain is determined by trial and error although, its measurement combination coefficients can be estimated systematically.

Ehinmowo et al. [23] proposed a systematic approach to determine the minimum control gain for a slug flow controller. In the study on flow stability at a large valve opening using active feedback control, the gas flow rate was used as the input variable for a purely catenary riser. The input variable (gas flow rate) was used successfully to stabilise the flow but determining this variable practically is not easy.

Tandoh et al. [64] extended the work proposed by [23] to design a controller gain of the inferential slug controller on a U-shaped riser to make the method more feasible for a real system. In the study, the liquid density, total liquid mass flowrate, riser top pressure, and gas mass flowrate were the topside measurements used due to their ready availability. One of the benefits of the approach is that it can be used to determine the control gain of the inferential slug controller (ISC).

5.3. Selected field and lab proven active slug control methods using topsides measurements

5.3.1. Intelligent choke

The Intelligent Choke is made consisting of a dynamic choke that is installed at the top of the riser, directly upstream of the stage separator (see Fig. 3). A PID controller, which regulates the liquid level in the downstream first stage separator, determines the opening of the choke. This liquid level is largely regulated by the separator's liquid outlet valve, but if it reaches a predetermined amount, the intelligent Choke activates and begins to close. Peak levels of liquid production caused by slug arrival can therefore be minimised. The Intelligent Choke, on the other hand, cannot prevent gas surge (due to pressure build-up upstream of the choke) after a liquid surge. The Intelligent Choke's PID settings are likewise significantly reliant on the PID settings of the first stage separator's liquid and gas output valves. If several pipes are linked to the same first stage separator, the Intelligent Choke's operation will be endangered.

Shell Expo implemented the Intelligent Choke at the Tern, a platform in the North Sea, in 1999 to handle the flow via the Hudson pipeline. Expro's Gannet platform was also used for testing to manage production across two 6″ flowlines of the Gannet D field [1]. Shell Expro installed the Intelligent Choke for otter at the Eider platform in 2003, and for Kestrel at the Tern platform in 2004. Petroleum Development Oman (PDO) also used the Intelligent Choke for two on-shore multiphase flow pipelines, Burhaan and Harweel-to-Birba.

ABB Optimize^{IT}

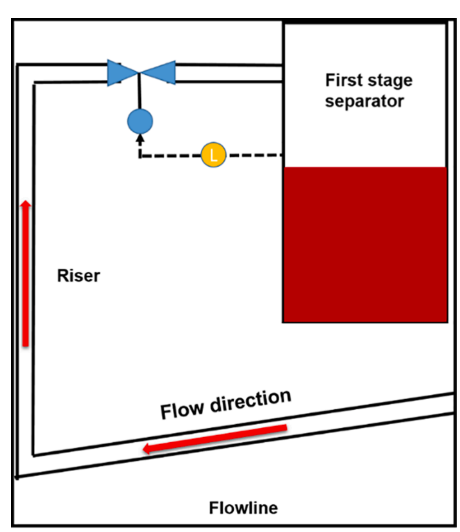

Fig. 3. Diagram of an intelligent choke [1].

Optimize^{IT} Active flowline control, which was originally branded as the ''SlugCon," is a slug mitigation system created by ABB. A flow control valve is installed at the top of the riser, upstream of the production separator, using this technique. The fundamental control mode was to maintain a constant pressure at the pipeline's intake. This technique is intended to avoid or reduce terrain and extreme slugs [68].

The subsea pipeline is subjected to pressure, temperature, and flow monitoring. These measurements are sent into a slug controller, which performs appropriate control action depending on them. A typical OptimizeIT active flow controller requires pressure, temperature, and flow data from two points, one upstream of the pipeline to detect the onset of slugs and the other downstream of the pipeline to detect the onset of slugs [68]. The control system configuration of the OptimizeIT installed for BP's Hod-Valhall pipeline in Norway is shown in Fig. 4.

Fig. 4 depicts the slug controller arrangement, which demonstrates how pressure readings at the pipeline's intake and output are utilised to modify the pipeline choke valve. If flow data are available, they are also employed for feed-forward control of the controller nominal operating point and tuning parameters.

The slug control system has the following features [68]:

- Slug control with dynamic feedback to guarantee pipeline stability.
- Choking was used to reduce the slug's impact on the separation and compression units.
- Feed forward control was used to fine-tune the dynamic feedback controller nominal operating point and settings.
- Slug signature for identifying slugs and monitoring the dynamic feedback controller effectiveness.

Separator train control interface:

- Feed forward control of the separator
- Slug control can be overridden (in case of a fault or error in the separator train).

The following facilities have either tested or implemented the Optimize^{IT} active flowline control:

- The Hod-Valhall Platform of BP, January 2000
- Visund North Development by Norsk Hydro, August 2001
- Brage Development by Norsk Hydro, July 2002
- Tor platform, Phillips Petroleum, May 2001
- Statfjord North Flank, Statoil, October 2002
- Statoil's Sigyn in December 2002
- The Hudson field of Shell is producing on the Tern Alpha platform.

Statoil's choke

Statoil created a cascading slug controller [68]. Fig. 5 depicts the control structure. The system is made up of two traditional PID controllers. The pressure controller and flow controller are combined in the cascade configuration. By adjusting the set point of the inner flow control loop, the outer control loop maintains a constant pressure at the inlet. The inner control loop, which is quicker, manipulates the choke position to deliver the required flow. By deactivating the inner flow control loop, the controller may be simply converted into a pure pressure controller. By deactivating the outside pressure control loop, it may be converted into a pure flow controller [68].

The flow controller function is to control the volumetric flow of the multiphase fluid through the choke. In this control mode, the choke position V_{pos} is given by a flow controller FIC with feedback from the volumetric flow F via the choke and set point F_{SP} set by the operator. A basic valve equation for a liquid/gas combination gives us the volumetric flow:

$$F = g(V_{pos})\sqrt{\frac{\Delta P}{\rho}} \tag{4}$$

F is the predicted volumetric flow, g is a function of the choke position V_{pos} supplied by the valve characteristics scaled as 0–100 percent, ΔP is the observed differential choke pressure, and ρ is the density of the multiphase fluid passing through the choke as measured by a gamma densitometer.

StatoilHydro has built two devices on the Heidrum TLP to stabilise the flow at Heidrun Norther Flank in Norway.

Shell Slug Suppression System (S^3)

Shell Slug Suppression (S^3) was intended to stop the severe slugging cycle and minimise peak liquid and gas production rates caused by transient slugs [69]. Fig. 6 depicts a schematic drawing of S^3, which is installed between the pipeline outlet and the production separator. Because the gas and liquid fractions in a multiphase flow can change dramatically over time, measuring and controlling the overall volumetric flow or mixture velocity with a single control valve is not simple. The S^3 functions as a control valve and is implemented as a "mini--separator", with separate control valves for each of the two phases present in the system and the use of typical measurement equipment for level, pressure and mass flow. A time-average pressure of 1 to 3 bara over the gas and liquid control valves is generally required for control [69]. The S^3 can also be used in conjunction with allocation or fiscal metering. The Penguins S^3, for example, includes a wet-gas metre at the gas output, and a multiphase metre in the North Sea may be recombined

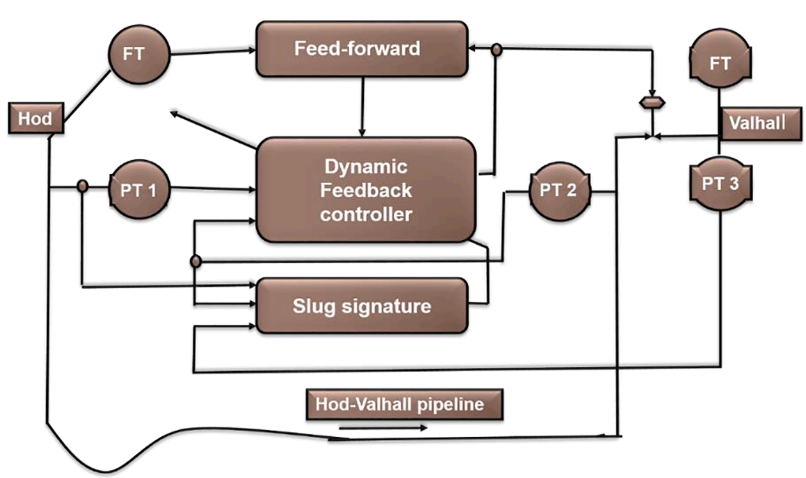

Fig. 4. Schematic diagram of the ABB Optimize^{IT} [68].

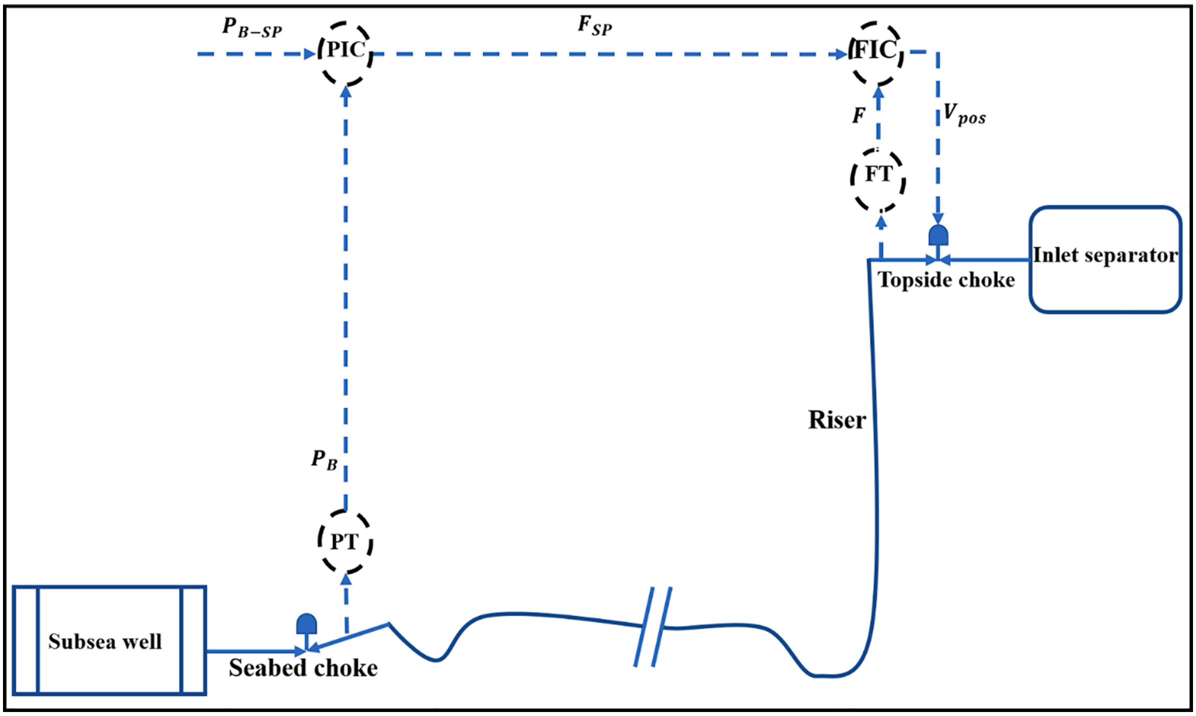

Fig. 5. A schematic diagram of Statoil controller [68].

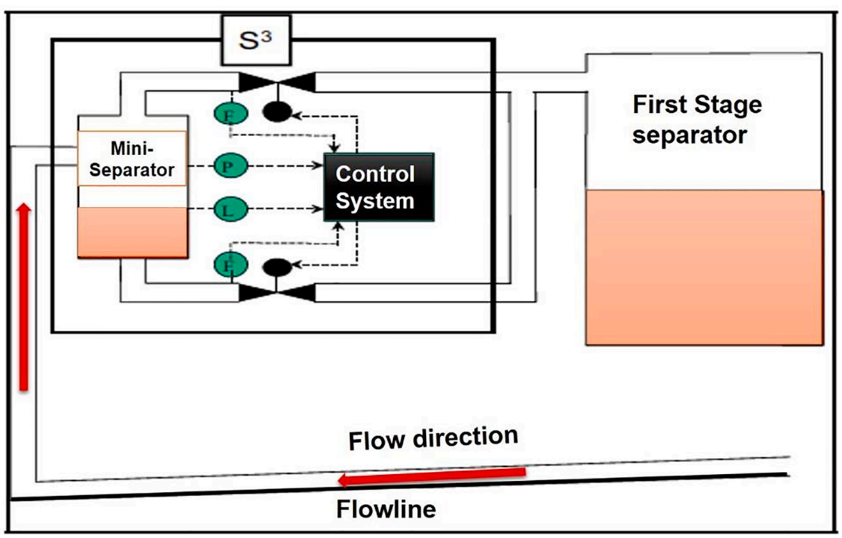

Fig. 6. Shell's slug suppression schematic diagram [69].

to enter the production separator. Instead, the flow following the S^3 can be split and enter straight into a liquid or gas train. During the S^3 commissioning, the controller settings are optimised and proven to be highly reliable.

The S3 control technique is focused on overall volumetric flow control as well as liquid flow control. The liquid valve is regulated in complete volumetric control mode to maintain a level fixed point. Furthermore, the gas valve is programmed to maintain a total volumetric flow set point. Flow metres in the gas and liquid outlets monitor the real flow rates. The variable to be regulated is thus the sum of the flow metre outputs. A pressure controller, in conjunction with other algorithms, changes the pressure set point in the mini-separator and factors based on the size of the pipeline [69]. The set point of the total volumetric flow is restricted by high and low limit values in order to function safely at startup.

If the riserbase becomes clogged with liquid and a severe slug occurs, the volumetric flow controller senses a reduction in output and opens the gas valve. The pressure in the S^3 mini-separator is reduced, and the liquid blockage is pulled into the riser, thereby ending the severe slugging cycle. However, the whole volumetric flow control technique will not function under transient slugging situations because it does not bind the level controller at high liquid flow rates [69]. To retain the liquid at its fixed point, the entire volumetric flow control mode would completely open the liquid valve. As a result, a liquid flow control mode of operation is established, which is activated when: (a) the liquid level gets to a certain threshold value or (b) the liquid flow rate exceeds a threshold value. The liquid flow rate threshold value may be affected by external factors such as the liquid level or liquid drainage capacity of downstream facilities. This threshold value can also be utilised as the liquid flow controller set point. Liquid flow management will slow down the slug velocity and prevent the slug from blowing out [69]. Because gas pockets and released associated gas in the mini-separator are constantly present during liquid slug generation, it is possible to provide for a regulated gas flowrate to reduce gas production famine. The gas flowrate is constrained by the total volumetric flow set point.

Once the slug is formed (indicated by a lowering liquid level in the mini-separator), the control strategy is switched from liquid flow control to total volumetric flow, and the gas in the line after the slug is decompressed (the gas flowrate is limited by the set-point of the total volumetric flow). The S^3 also controls the gas surge; otherwise, transitory slugs will occur due to the increased gas flow rates. As a result of these regulating operations, riser-induced slugging is reduced, and transitory slugs and accompanying gas surges are produced at a predetermined pace. This will result in a more reliable pipeline behaviour [69]. Thus, the economic loss from the additional equipment must be noted as a significant drawback, because this also introduces more equipment maintenance.

The S^3 has been extensively tested in lab flow loops and in the field, and it has been employed on two Shell Expro North Sea platforms, North Cormorant and Brent Charlie. A third S^3 was in operation at the Piper Bravo site in the North Sea to reduce slugging in Talisman's Tweedsmuir pipeline.

Shell's Vessel-less S^3

Since the two S^3 systems were deployed in the field, a version of the S^3 without the mini-separator has been produced [70]. The major reason for the vessel-less S^3 is to minimise system costs, as removing the min-separator has a considerable cost advantage. In addition to cost advantages, the vessel-less S^3 is an effective substitute for platforms with limited space and weight. Fig. 7 depicts a schematic of the vessel-less S^3. The mini-function separator's is essentially achieved by a stratifier, downcomer, and two T-junctions [70].

The stratifier aids in liquid/gas separation at the T-junction sections, and the downcomer handles incoming liquid slugs, allowing the control valves to respond with enough time. A level transmitter measures the level of liquid in the downcomer (DP cell). The rest of the vessel-less system, including the control algorithms for slug mitigation, is identical to the standard S^3 [70].

The air-water test rig in Amsterdam was used to test the vessel-less S^3.

Shell's Smart Choke

While the traditional S^3 [69] and Vessel-less S^3 [70] are effective in mitigating liquid slug and controlling gas surges, they necessitate pre-separation of the gas and liquid phases in each pipeline where slugging may occur. Furthermore, the method necessitates the use of two control valves, which, when combined with the separator, may be prohibitively expensive for some slug issues that emerge only at the end of field life. Furthermore, the S^3 is not a feasible solution when several risers must be on boarded to a single production plant [71]. The recommended choice is a slug mitigation method based on a single valve that can sufficiently minimise liquid slugging while also controlling gas surges. The Shell Smart Choke, which implements the S^3 control algorithm on a single control valve, was created for this purpose.

Fig. 8 is a schematic of the Shell Smart Choke. The valve is controlled by three inputs: upstream pressure, pressure drop *dP* over the choke, and valve position. The flow through the valve is calculated using a model correlation that takes the three input factors into account. A PID controller is then applied to the valve to maintain a consistent flow through it. Constant flow can be either volumetric or mas flow. The control mode is changed based on whether a liquid slug or a gas surge is created [71].

The Shell Smart Choke, as contrasted to a fixed choke, attempts to maintain a constant overall volumetric flowrate. When the valve is in automated mode, it will open to the point where a flow rate set point is achieved; this set point is defined by a pressure set point for the system, which is required to manage the flow. The flowrate is determined as follows:

$$Q_v = X(t)\sqrt{\frac{dP}{\rho_x}} \tag{5}$$

In the preceding equation, Q_v represents the estimated two-phase flowrate (m^3/h), dP represents the pressure drop over the valve (bar), and ρ_x represents the fluid density (kg/m^3). The variable $X(t)$ is affected by the valve size and the valve opening. The time-average pressure decrease across the valve is typically 1 to 3 bar.

This estimated flowrate is utilised to regulate the flow and is not always equal to the real flowrate. The system operates in three modes: default mixing mode, liquid mode, and gas mode. A pressure drop across the valve is required to regulate the flow. The value of dP/dt is used to transition between modes, each of which corresponds to a particular density to be utilised in the Q_v computation. These densities are- ρ_{mix},ρ_{liquid}, and ρ_{gas}, respectively. ρ_{mix} is typically computed using the gas/liquid flowrates in the pipeline. The gas and liquid densities are not the real densities, but they are carefully adjusted to dampen the system. To account for the compressibility effects of the gas, the density for the gas and mixed modes is multiplied by the upstream pressure [71].

When dP quickly increases and dP/dt reaches a particular threshold value, a liquid slug is expected to arrive at the valve. As a result of this

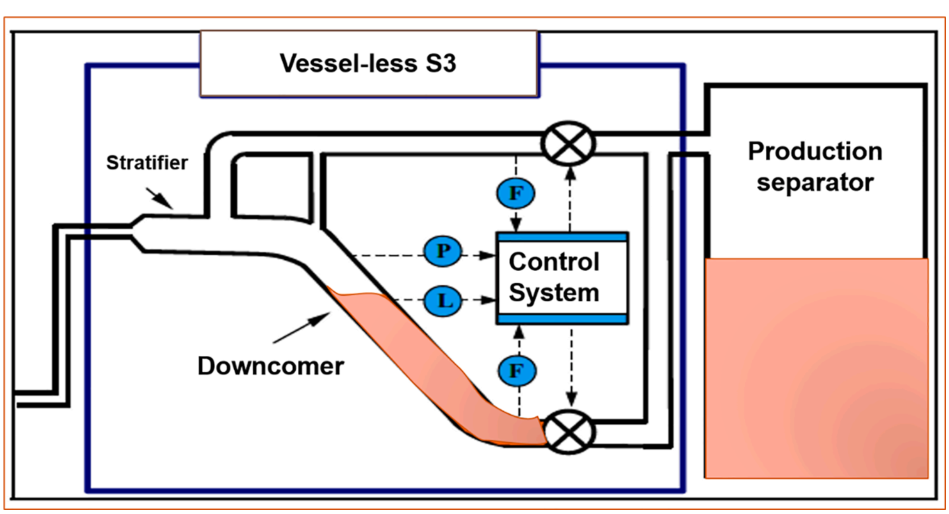

Fig. 7. Shell's vessel-less schematic diagram [70].

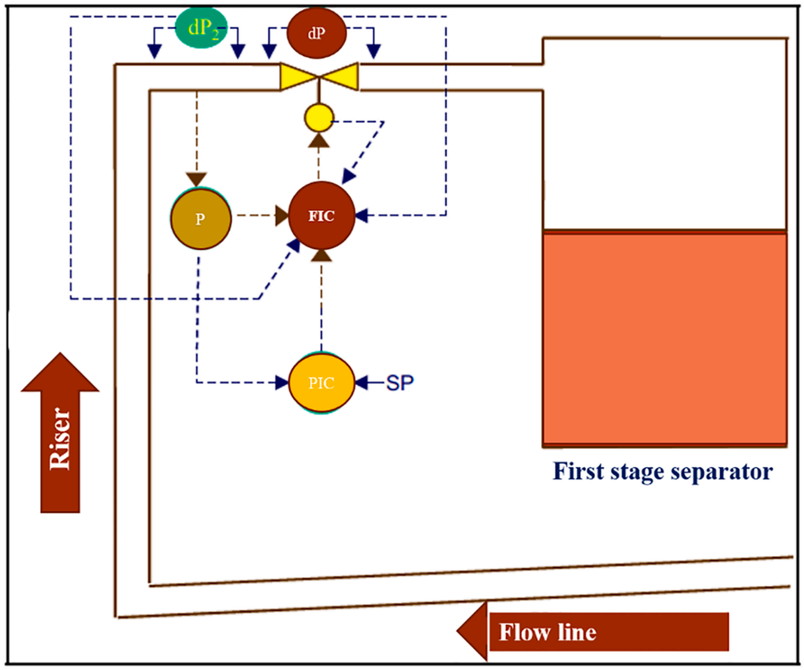

Fig. 8. Shell Smart Choke schematic diagram [71].

arrival, the default mixing mode is switched to liquid flow mode. As ρ_{liquid} exceeds ρ_{mix}, the estimated value of Q_v falls, and the system opens the valve to return to the Q_v set point. By opening the valve, the liquid outflow is promoted, and additional liquid build-up in the riser is prevented. When the flowrate reaches the Q_v specified point, the valve will automatically close.

When the dP/dt value goes below the threshold, the system returns to the mixed flow mode. After all liquid has been generated, the dP/dtwill soon fall below another (negative) threshold value. In this scenario, the system expects a gas rush and responds by altering the gas density and closing the valve. This mode switching activity can be dampened by utilising an average dP/dtvalue, where the number of values used to generate the average defines the degree of damping. Another approach to affect the valve opening in gas-liquid modes is to carefully determine

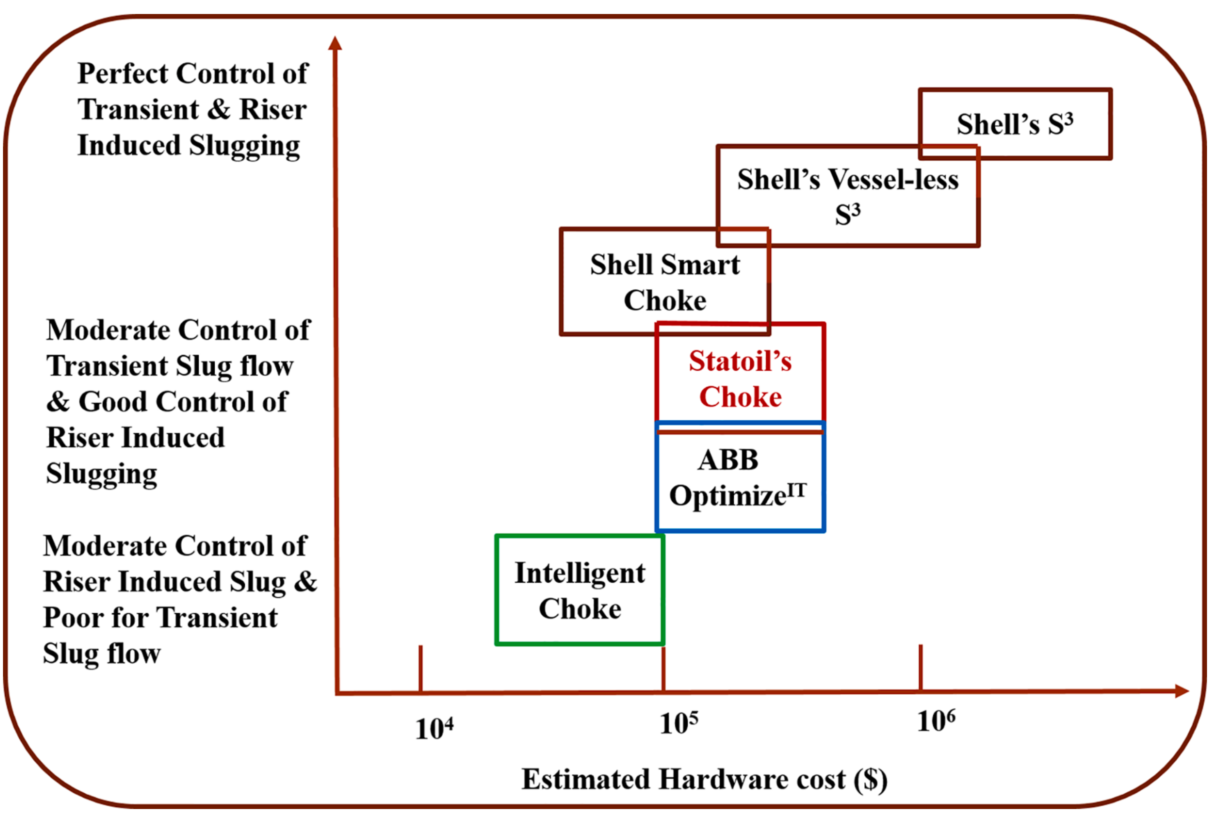

Fig. 9. Active slug control benefits Vs costs graphical analysis for different control methods [72].

the densities involved. While topsides choking can help reduce slugs in a pipeline-riser, it usually causes an increase in backpressure on the wells [71].

Fig. 9 depicts a cost-benefit analysis of several control methods.

6. Future outlook and recommendation

Beyond the fundamental aspects of slug flow control systems that are being researched, there are numerous opportunities to integrate a range of control technologies and systems to take slug flow control improvement to the next level. Amongst these are:

I Research is required in the deployment of non-intrusive Doppler ultrasonic sensor signals for the online detection and control of slug flow.

II A study is required to investigate the feasibility of using information from neurofuzzy for online flow monitoring and control. The idea is shown in Fig. 10.

7. Conclusion

Slug control is a challenging phenomenon. The major problem with severe slug flow control/mitigation is to address the robustness.

Several major slugging elimination techniques have been examined and they are categorised into active and passive slug control methods. The most common active slugging control methods are the feedback control of a riser/well artificial gas-lifting and topside choke valve feedback control of a riser/well. The major aim of every slug control is severe slugging flow elimination and production optimization.

Although a lot of effort and money has been spent for the past decades in slug control investigations by both industrial and academic sectors, getting the most accurate, economical, optimal, and robust remedies are still problems that remain. As the potential hydrocarbon resources reduce with time, the motivation for exploring deeper hydrocarbons is on the increase. Hence, slugging problems will get more profound in the future due to the use of longer risers and wells. The following observations are relevant based on the information given in this review:

a The use of simply the topside pressure measurement for stabilising control is not robust; this has been explored using a linear controllability analysis. The reason for this is that the right half-plane (RHP) zeros seen in the inverse response behaviour are quite close to the system's unstable poles, resulting in an unavoidably high peak in the sensitivity transfer function. Subsea pressure measurements, on the other hand, do not contain RHP zeros, and a basic PI controller is utilised in practice.

b In a condition where the only accessible measurement is topside pressure, a typical control method is to develop an observer to estimate the system states, including subsea pressure, and then utilise these estimations for control. We do know, however, that the observer mechanism and state feedback cannot be utilized in all control applications. The observer, in particular, cannot overcome the basic controllability constraint associated with RHP zeros.

c The subsea solution is better in terms of robustness. The deployment of riser-base pressure as the controlled variable for flow stabilization is robust and easy. Hence, the main drawback to this idea is its requirement of subsea equipment which in most cases is difficult to achieve and very expensive.

Furthermore, slug control/mitigation research is still ongoing and will become more important in the future.

Appendix A. The characteristics of severe slug flow

Severe Slug flow Number derivation

The upstream pipeline of the riser dip has cross-section (*A)* and length (L_F). Assuming that the flow is stratified flow in this part of the pipeline with liquid holdup fraction (α_F). Hence, considering the time the liquid blocked the riser dip, the increase in pressure at the back of the liquid blockage can be stated mathematically as:

$$\left(\frac{dP}{dt}\right)_{flowline} = RT\,\frac{d\rho_G}{dt} \tag{6}$$

Ideal gas law is believed to hold, with gas density(ρ_G), temperature (*T*) and gas constant (*R*). The gas density is equal to the ratio of the gas volume and gas mass (M_G) in the upstream pipeline:

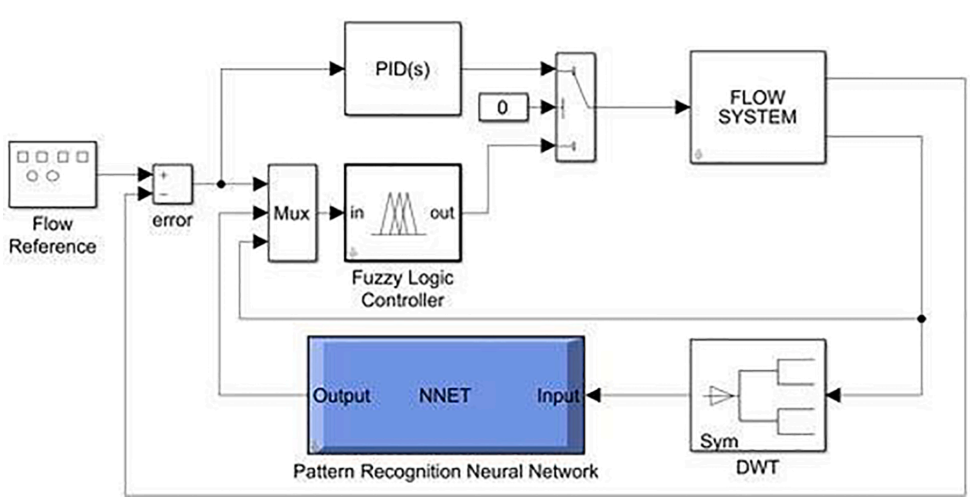

Fig. 10. Flow control using information from Neurofuzzy.

$$\rho_G = \frac{M_G}{(1-\alpha_F)L_F A} \tag{7}$$

Substituting Eqs. (6) into (7) gives

$$\left(\frac{dP}{dt}\right)_{flowline} = \frac{RT}{(1-\alpha_F)L_F A}\frac{dM_G}{dt} \tag{8}$$

When the liquid blocked the riser, the gas mass change is simply because of the gas inflow at the inlet of the pipeline, which gives:

$$\left(\frac{dP}{dt}\right)_{flowline} = \frac{P_s}{(1-\alpha_F)L_F}V_{SGs} \tag{9}$$

where $P_s$isthepressureatstandardcondition,V_{SGs} is the superficial gas velocity at standard condition.

The pressure downstream of the blockage of liquid is because of the hydrostatic head increase and it can be shown mathematically as:

$$\left(\frac{dP}{dt}\right)_{riser} = (\rho_L - \rho_G)V_{SL,\ R}\ g\ \sin\varphi \tag{10}$$

where $V_{SL,R}$is the riser's superficial liquid velocity while ϕ is the riser angle with respect to the horizontal (hints: for vertical riser, $\phi = 90°$). The riser superficial velocity follows from the pipeline superficial velocity as stated below:

$$V_{SL,\ R} = \left(\frac{D_F}{D_R}\right)^2 V_{SL} \tag{11}$$

Here, D_R is the riser diameter while D_F is the pipeline diameter.

The severe slugging number (Π_{ss}) can be said to be the ratio of upstream pressure increase and downstream of the liquid blockage as shown mathematically below:

$$\Pi_{ss} = \frac{\left(\frac{dp}{dt}\right) flowline}{\left(\frac{dp}{dt}\right) riser} = \frac{Ps}{(1-\alpha_F)(\rho_L - \rho_G)\left(\frac{D_F}{D_R}\right)^2 g\ \sin\varphi L_F}GLR \tag{12}$$

where the gas-liquid ratio at standard condition is GLR (in the unit, m^3/m^3) (hints: $GLR = V_{SGs}/V_{SL}$). This simplifies to give

$$\Pi_{ss} = \frac{\left(\frac{dp}{dt}\right) flowline}{\left(\frac{dp}{dt}\right) riser} = \frac{Ps}{(1-\alpha_F)\rho_L g L_F}GLR \tag{13}$$

Assuming (a) $\rho_G \ll \rho_L$ (b) the riser is vertical, (c) the riser and the pipeline have an equal diameter.

Slug length and period

The severe slug flow cycle period can be deduced from the time (Δt) it takes the gas in the pipeline to accumulate enough pressure to push the liquid out of the riser. Starting at the time the slug was just blown out and at minimum rise-base pressure ($\Delta tlevel1$) and ends at the time the liquid-filled the riser, and at the maximum riser-base pressure ($\Delta tlevel2$). Hence, the mass balance of the gas upstream the riser base of the pipeline can be stated mathematically as:

$$\mathfrak{m}_G$$

where

$$\mathfrak{m}_G \Delta t = (\rho_2 - \rho_1)(1-\alpha_F)L_F A$$

is the inlet gas mass flowrate which is equal to ($\rho_{Gs}V_{SGs}A$) with the gas superficial velocity and gas density taken at the standard conditions. Applying the ideal gas law, we can rewrite the mass balance to give:

$$\Delta t = \frac{\Delta P}{Ps}\frac{(1-\alpha_F)L_F}{V_{SGs}} \tag{15}$$

with $P_2 - P_1 = \Delta P$. The term ΔP can be taken to be equal to the hydrostatic head of the riser when filled with liquid ($i.e\Delta P = L_R\rho_L g$).

$$\Delta t = \frac{L_R}{V_{SLs}}\frac{1}{\Pi_{ss}} \tag{16}$$

where $\frac{L_R}{\Pi_{ss}}$ is the liquid slug body length.

The pipeline superficial liquid velocity at standard condition is denoted as V_{SLs}. The relation for Δt solely denotes the severe slugging cycle part where there is production starvation. The liquid blowout time is excluded. Although, the liquid blowout time is lower than that of production starvation time.

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Slug flow control using topside measurements: a review

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