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Abstract

The present thesis is focused on the problem of severe slugging and ways to mitigate it. Severe slugging is an oscillatory multiphase flow regime characterized by high variations in production rates occurring in offshore pipeline-riser systems.

Chapter 1 provides basic notions related to multiphase flow, which are essential for understanding of the rest of the thesis.

Chapter 2 gives a thorough description of the severe slugging occurrence mechanism and preconditions as well as introduces different types of the phenomenon. Special attention is given to the effect of mass transfer and how it alters the flow regime's behavior. Detrimental effects of severe slugging are discussed and some examples are provided.

Chapter 3, making a significant part of the thesis, provides its reader with carefully gathered data concerning severe slugging alleviation and mitigation methods published from 1973 to 2015, both conventional and purely speculative methods are discussed. Examples, where possible, are given.

Chapter 4 considers modeling of severe slugging in a vertical riser with aids of the multiphase simulation program OLGA. A constructed study case is considered and described with some of the mitigation techniques implemented and tested.

Chapter 5 evaluates a novel severe slugging mitigation method proposed by Caltec Ltd. UK. The method assumes pipeline system depressurization by installation of a Surface Jet Pump on the production platform. The chapter gives the method description and verifies its feasibility using a simulation model within OLGA.

The thesis ends with Conclusions and Recommendations for further work and self-evaluation.

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Nomenclature

A – cross sectional area, [m^2];

C – choke coefficient [$Pa \cdot s^2 \cdot m^{-2}$];

D – diameter, [m];

F – force per area, [$N \cdot m^{-2}$, Pa];

M – molar mass [$g \cdot mol^{-1}$];

U – phase velocity, [$m \cdot s^{-1}$];

P – pressure, [Pa];

Q – volumetric flowrate, [$m^3 \cdot s^{-1}$];

R – gas constant [=8314 $J \cdot K^{-1} \cdot kmol$];

T – temperature [K];

f – friction coefficient [-];

h – height, [m];

l – length, [m];

$\dot{m}$ – mass flow rate [$kg \cdot s^{-1}$];

z – compressibility factor, [-];

α – phase area fraction, [-];

β – pipeline inclination angle to horizontal line, [rad];

γ – specific gravity, [-];

ρ – density, [$kg \cdot m^{-3}$];

ϕ – average holdup, [-];

Subscripts

B – backpressure;

G – relating to gas;

HP – high pressure;

L – relating to liquid;

LP – low pressure;

M – relating to mixture;

MP – medium pressure;

O – relating to oil;

R – relating to riser;

P – relating to pipeline;

b – relating to bubble-point;

bub – relating to bubble;

res – relating to reservoir conditions;

sep – relating to separator;

wf – relating to flowing bottom hole pressure;

0 – at standard conditions;

Abbreviations

API – American Petroleum Institute;

$C-ESP$ – Caisson – Electrical Submersible Pump;

EPI – elimination performance index;

ESP – electrical submersible pump;

FVF – formation volume factor;

GOR – gas-oil ratio;

HP – high pressure;

ID – internal diameter;

IPR – inflow performance relationship;

$I-SEP$ – compact cyclonic gas/liquid separator by Caltec Ltd.;

LP – low pressure;

MP – medium pressure;

$OLGA$ – (stands for OiL and GAs) multiphase flow simulator;

PI – productivity index;

PVT – pressure-volume-temperature;

$RSGO$ – solution gas-oil ratio (the same as R_S);

SJP – surface jet pump;

SS – severe slugging;

STB – barrel at standard conditions;

S^3 – Slug Suppression System;

$VASPS$ – Vertical Annular Separation and Pumping System;

Contents

Introduction

Nowadays, a significant amount of hydrocarbons is being produced from marginal fields, which development assumes subsea production followed by a flowline terminated into a platform or onshore production facilities via a riser. Because of the challenges related to subsea processing, production fluids are often transported in a multiphase flow, which imposes additional requirements on flow assurance, such as oscillatory flow regimes.

The problem of severe slugging became evident in early 70s due to both an increased number of marginal fields' developments and increased water depths. Severe slugging may arise when a number of preconditions are present, such as multiphase flow at relatively low production rates, and the specifics of the pipeline-riser geometry.

Severe slugging can be characterized by low-frequency oscillatory flow with significant pressure and flowrate fluctuations. It has proved to be hazardous for the production system in general, potentially causing high stress loads on riser pipework and topside piping and affecting efficiency of process equipment, such as separators, pumps and compressors. If untreated, severe slugging may result in complete system shut-down. In addition to that, the phenomenon increases backpressure imposed on the wells, decreasing production by a substantial amount or killing the wells completely in some extreme examples.

A number of mitigation techniques have been proposed over the last 40 years. The thesis gives a thorough description of all the methods found in literature ranging from 1973 to present and evaluates on a novel mitigation technique proposed by Caltec Ltd., UK.

1. Background

The purpose of this chapter is to introduce some of the basic concepts important for the thesis topic. In case the reader is familiar with the definitions, the chapter may be skipped without loss of understanding.

1.1. Multiphase flow

Multiphase flow in pipeline simply refers to flow regimes, consisting of more than one phase. For petroleum multiphase pipeline transportation the most commonly encountered fluids are different fractions of hydrocarbons (gas, oil, gas condensate), produced water, and solid particles. Therefore, a typical case when gas, oil, and water are present in the conduit refers to three-phase flow.

Unlike single phase flow, multiphase flow may appear in different forms depending on the conduit geometry and the fluids properties and quantities. These forms are called ***flow regimes***. Figure 1.1 gives an example of a flow regimes map of two immiscible fluids plotted with respect to gas and liquid ***superficial velocities***. The superficial velocity is a multiphase flow specific definition and for a given flow phase may be defined as a ratio between the phase volumetric flowrate and the pipe cross section area:

$$U_{Si} = \frac{Q_i}{A}. \tag{1.1}$$

The superficial velocity definition above provides an intuitive understanding of the term: *superficial velocity of a phase in a multiphase flow is the average instantaneous velocity of the given phase if it was a single phase flow with the given volumetric flowrate*. On the other hand, the volumetric flow rate of a given phase may be expressed in terms of the absolute phase velocity (U_i) and the area occupied by the phase ($\alpha_i A$):

$$Q_i = \alpha_i A U_i, \tag{1.2}$$

where $\alpha_i = A_i/A$ is the ***area (volume) fraction*** of phase i.

When referred to liquids, area fraction is commonly called ***(liquid) holdup***. For petroleum related topics, water phase fraction is often referred to as ***water cut***.

Thus, we can describe superficial velocity of phase i in terms of its absolute velocity in the flow and area fraction:

$$U_{Si} = \alpha_i U_i. \tag{1.3}$$

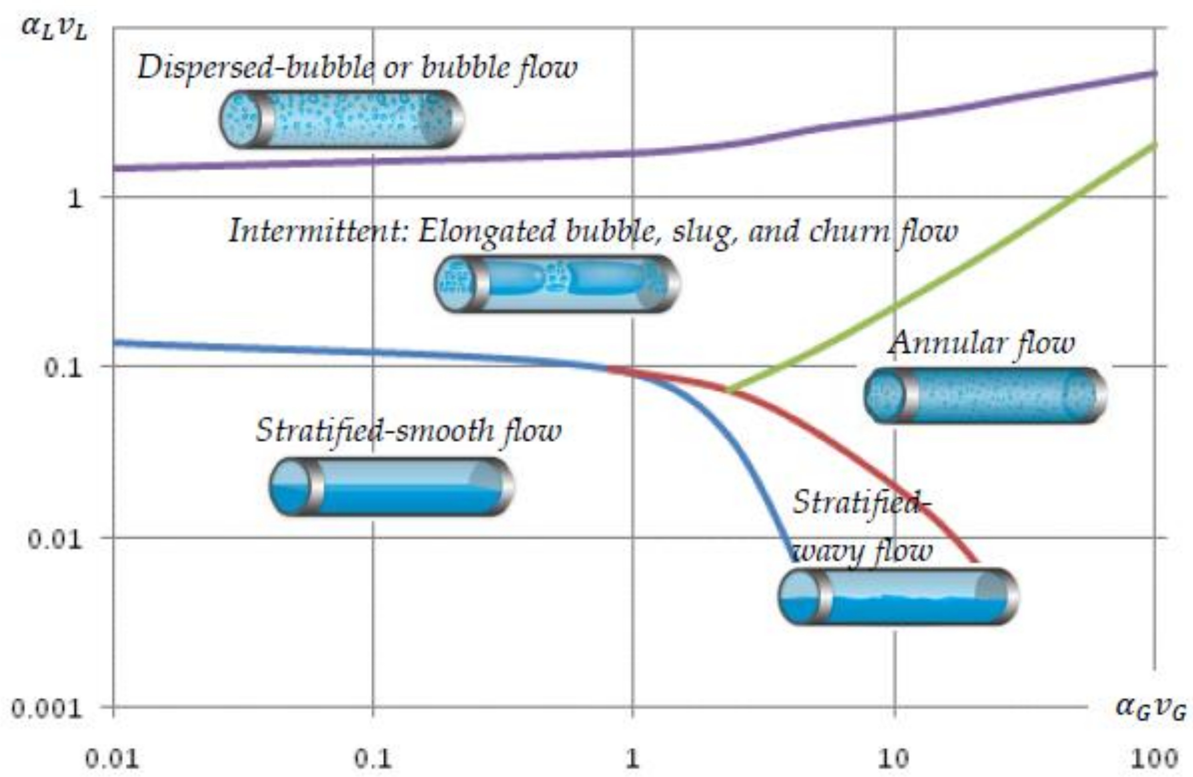


Figure 1.1. An example of flow regime map for two-phase gas-liquid flow in a horizontal pipe [1].

The figure uses nomenclature different from the rest of the thesis.

Some of the flow regimes observed in horizontal pipes essential for the thesis are to be explained here. Starting with modest amount of liquid and gas in the pipe, the ***stratified flow*** may be observed, refer to Figure 1.1. Substantial increase of gas quantities flowing through the pipe induce waves on the surface of the liquid phase forming ***stratified wavy flow*** up to the point where the wave crests reach the top of the pipe and block the whole cross section of the conduit thus forming relatively short liquid slugs. Such phenomenon is called ***hydrodynamic slugging***. Further increase of gas velocities creates sufficient turbulence to tear the liquid droplets of the liquid phase, forming a layer of liquid around the inner circumference of the pipe and dispersed liquid droplets traveling with the gas. This flow regime is called ***annular flow*** and frequently encountered in gas with gas-condensate flows. On the other extreme, when liquid superficial velocity is high enough, the gas phase is dispersed in liquid and ***dispersed-bubble flow*** occurs. Between the phases mentioned above, a number of intermittent flow regimes lay, characterized by less even distribution of liquid and gaseous phases. To those ***slug*** and ***churn flows*** may be attributed. The overall flow regimes classification is not limited to those mentioned in this paragraph and not universal across the field. It may be encountered that individual researchers as well as publications of different years tend to specify flow regimes in a slightly different way.

However, flow maps are restricted in their usefulness due to predefined fluids properties and pipeline geometry. They are also of limited use for more-than-two-phase flows. The situation is further worsened if the fluids are able to have ***mass transfer*** between them, which is often the case for petroleum pipelines. Thus, depending on PVT conditions and oil properties, gas is dissolvable in oil phase; water may be present in form of vapor in the gas phase, and so on.

1.2. Slug flow

Slug flow is characterized by varying liquid and gas flowrates and may impose significant threat to production facilities, inducing oscillations, increasing wear and tear of equipment, hindering separation efficiency. Severity of slug flow mainly depends on its origin and may be described in terms of liquid slug length and pressure fluctuations.

Three main types of slugging by its origin related to petroleum multiphase production are:

1. ***Hydrodynamic slugging***. As was described before, this type of slugging is formed due to waves generated by gas blowing with sufficient speeds over the liquid phase layer. When the wave crests reach the top of the pipe, they block the whole cross section of the conduit and form slugs. As a rule of thumb, slugs generated by this mechanism are relatively short.
2. ***Operationally induced slugging***. Refers to transient flow regimes in multiphase pipeline, such as production shut-down/ramp-up/restart and pigging operations.
3. ***Terrain induced slugging***. Refers to slug flow originated at the dips of pipelines following the profile of the terrain (therefore the name). Liquid accumulated at a dip may block the pipeline cross section and form a slug. If a number of criteria met, the

slug may grow up to considerable lengths until the upstream gas pressure build-up pushes it out of the dip.

An extreme case of terrain induced slugging often occurs in pipeline-riser systems and called ***riser (induced) slugging***, or ***severe slugging***. Even though severe slugging is attributed to a separate type of slugging by some authors, both have the same mechanism and only differ in the location of the origin.

2. Severe slugging phenomenon

2.1. Mechanism

Severe slugging as a separate phenomenon received much attention in early 80s as part of the Tulsa University Fluid Flow Projects run by Schmidt [2-4] and was probably linked to higher number of subsea developments and increased depths of offshore developments, therefore taller riser and severer riser induced slugging conditions. In a series of studies conducted for the project, two main flow regimes associated with slug flow in risers were distinguished. First, the hydrodynamic slugging, originated in upstream pipe and arriving to the riser base and moving up the riser conduit. The second one, severe slugging, was classified as terrain induced slugging due to its nature, was governed by different mechanism and had severer consequences.

Severe slugging was defined as such a terrain induced slug flow regime resulting in slugs of length equal or greater than that of the riser. It was also found to be a cause of high magnitude pressure fluctuations and varying production characteristics, both in terms of production rates and phases delivered to the platform (see Figure 2.1). First precondition for severe slugging occurrence was found to be low gas and liquid flowrates [3].

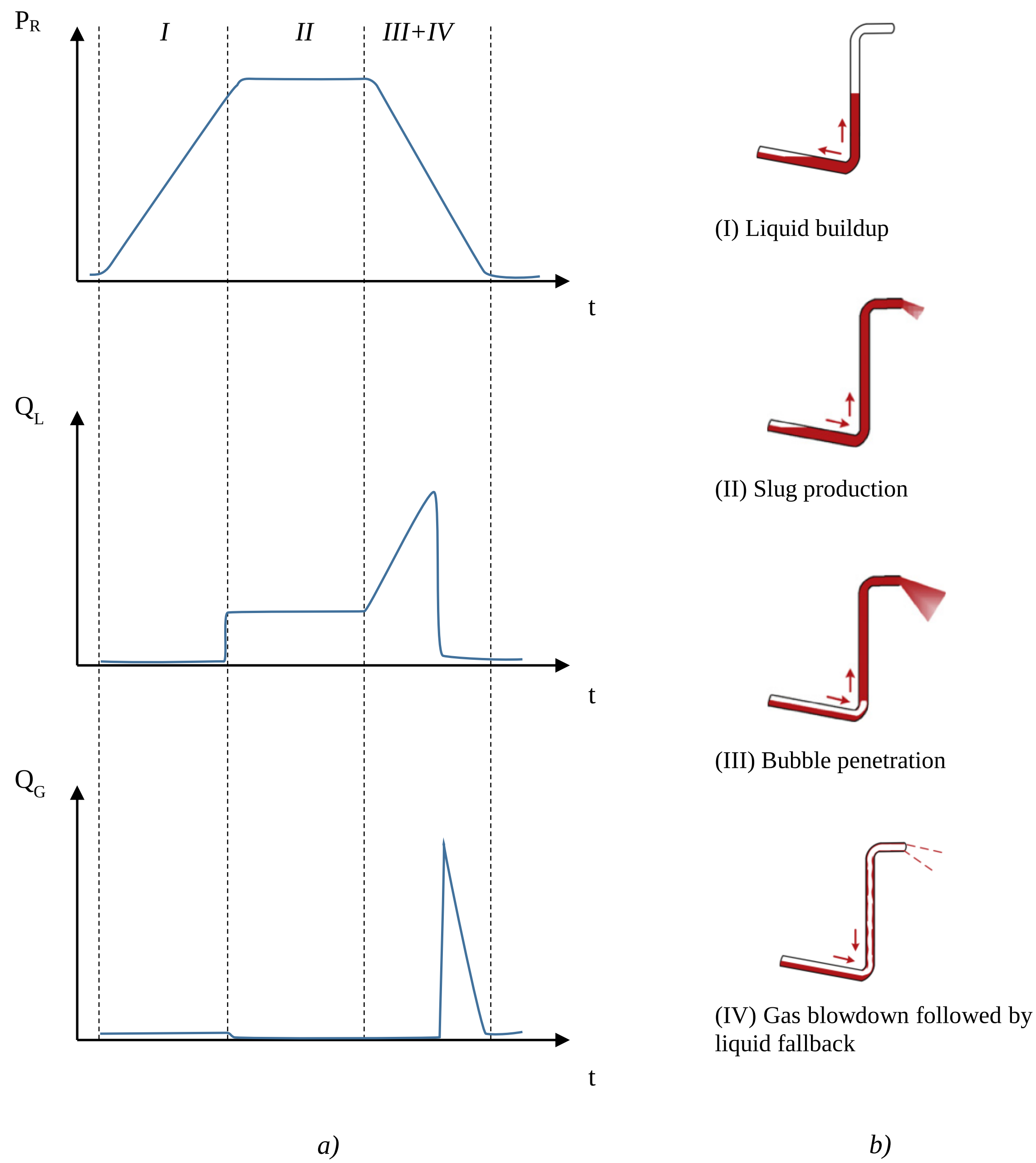


Figure 2.1. Qualitative depiction of one severe slugging cycle.

Figure a) represents variations of riser base pressure, liquid and gas flowrates with time. Figure b) shows liquid holdup distribution along the system (adopted from [5]).

Severe slugging mechanism is initiated by formation of liquid blockage at the riser base due to a low point in the pipeline, which prevents the gas phase from penetrating inside of the riser, and followed by force balance between the increasing hydrostatic pressure of the liquid column in the riser and compressed gas pressure inside of the pipeline. The whole process may be broken down on four major stages (see Figure 2.1):

I. Liquid buildup

Liquid accumulated at the riser base blocks the conduit cross section and creates pressure difference between upstream and downstream of its origin. Often initiated by the preceding stage of liquid fall back, when some part of the slug produced during the blowout phase loses momentum and falls to the riser base. The liquid phase, continuing flowing to the riser base in stratified flow, initially increases the liquid level in the riser and liquid propagation into the pipeline. Gradually, as the pressure in the pipeline increases due to the gas compressibility, the gas phase pushes the liquid slug further into the riser. The stage continues until the slug reaches the riser top (for slugs of length greater than riser length) or until the bubble penetration stage occurs.

II. Slug production

At this stage the slug occupies whole riser length. The phase is characterized by liquid production at relatively slow rate equal to liquid inflow from the pipe and volume displaced by the compressed gas. Pressure distribution along the riser height remains constant and follows approximately hydrostatic pressure distribution. The phase ends when the liquid-gas level inside of the pipe reaches the base and the gas enters the riser.

III. Bubble penetration or gas blowout

With liquid-gas interface reaching the riser bend, the gas phase enters the riser, forming a bubble front similar to a large Taylor bubble. The bubble, continuously expanding due to reduced static pressure, propagates up the riser decreasing hydrostatic pressure on the riser base, therefore involving more gas into the riser. With the gas front propagating further into the pipe, the pipeline pressure continues declining, letting the gas expand. Eventually, the force balance between the hydrostatic and compressed gas in broken and the riser blowout occurs characterized by ever-increasing flowrate of the liquid into the separator due to expanding gas flowing from the pipeline.

IV. Gas blowdown and fallback

The fast liquid production phase is followed by violent production of pressurized gas at high rates, called blowdown. After the gas has expanded and the pipeline pressure drops to its minimum, the gas velocity in the riser drops low enough to let the liquid fallback from the riser inner surface. The liquid fallen back from the riser accumulates at the riser bend and starts the cycle again. At certain conditions liquid fallback may not occur. In this case the new cycle is initiated from stratified flow blocking the riser bend cross section.

The description somehow varies between the researchers, attributing the bubble penetration to the fourth stage or distinguishing a fifth one. However, the principle remains the same.

Based on the description given above, one can infer some of the necessary preconditions for severe slugging. First, severe slugging is sensitive to pipeline-riser geometry, with the pipeline declination being an important precondition. Otherwise, the gas would be redistributed and moved further down the stream due to the lower density compared to the liquid phase; therefore, the condition under which the gas balances out and allows the liquid build up in the riser would not be met. In addition to that, stratified flow regime in the pipeline section preceding the riser must be present. As was shown by Schmidt [6], normal slug flow in the pipeline and riser generated slug flow are mutually exclusive since pipeline slugs and bubbles would pass the riser base almost unchanged, eliminating possibility for liquid accumulation at the riser base.

Second, for the riser-generated slugging to be strictly classified as severe slugging, the rate of increase of hydrostatic pressure must be greater than the rate of increase in pipeline pressure to the point when the liquid column reaches the riser outlet. If the condition is not met, the flow must be considered as being a transitional to severe slugging [3]. However, transitional slugging may still present enough hazard even with lower fluctuations in pressures and production rates compared to pure severe slugging.

Third, the accumulated liquid column in the riser must be stable. It means that initiation of the bubble penetration phase won't cause rapid expansion of the gas phase resulting in blowout. The stability criteria is met when increase of gas flowrate at a given liquid flowrate through the vertical pipe causes increase of the pressure drop over the pipe. In other words, the gas does not produce gas-lift effect.

Linga [7] pointed out at an alternative mechanism for severe slugging generation related to slugs formed upstream of the pipeline-riser junction. Thus, a train of terrain-induced slug of sufficient length decelerated at the riser base was observed to be a possible reason for severe slugging initiation. The slug stopped near the riser blocks the riser base cross section and initiates the process. Slugs arriving afterwards contribute to build up of the liquid column inside the riser. Eventually the liquid column, being in near unstable state, is destabilized by an additional slug. The gas entrapped between the column and the fresh slug forms a buffer, pushing the column into the riser and initiating blowout, and decelerating the slug at the junction point. Thus, the process repeats itself.

Possibility of severe slugging without stratified flow being present in the pipeline could seriously affect the phenomenon prediction since slug flow occurs for different combination of superficial gas and liquid velocities. Moreover, some of the elimination methods discussed further in the thesis could prove to be not feasible or less effective, e.g. mixing and self-lift.

2.2. Severe slugging classification

The early works on the topic defined severe slugging as a riser induced slug flow resulting in slugs of length equal or greater than the riser length. Those flow regimes with gas blowdown before the liquid level had reached the riser top, therefore had slug length less than that of the riser, were classified as transitional to severe slugging flows.

However, even transitional to severe slugging flow can be “severe” enough to call for actions. A riser induced slugging classification by Tin and Sarshar [8], introduced in 1993, is often used as a standard. It is presented in the following paragraphs.

Severe slugging of type 1 (SS1) is the regime we referred to before as severe slugging. During SS1 the slug reaches the riser top prior to blowout. This type of severe slugging has a distinctive feature of constant liquid production phase with approximately constant rate, during which the riser base pressure remains roughly equal to hydrostatic head of the riser.

Severe slugging of type 2 (SS2) is characterized by blowout occurring prior to the liquid slug reaching the riser top and was referred to before as transitional to severe slugging.

Severe slugging of type 3 (SS3) is initiated by transitional slugs generated in the upstream pipeline and traveling along the riser, partially falling down and forming highly aerated liquid slug at the riser base. The slug then pushed into the riser by increasing backpressure from the pipeline side. Eventually, the main gas cap (in contrast to smaller gas bubbles, which keep passing to the riser throughout the whole process) penetrates into the riser, starts rapid expansion, and pushes the aerated liquid slug up the riser. Blowdown follows, generating new portion of transient hydrodynamic slugs inside of the pipeline and the cycle repeats. In the previous section we described this type of slugging as Linga's mechanism.

2.3. Mass transfer effect on severe slugging

So far, the discussion on the topic only considered systems containing fluids which do not allow mass transfer between them. The majority of studies published to date restrict their effort exclusively on that type of system. It may be explained by two factors. First, the phenomenon may be easier described and modeled by neglecting the mass transfer term, however at the cost of the model accuracy. Second, such models are easier to test and verify in laboratory conditions due to the fact that mass transfer primarily occurs due to pressure changes. Therefore, a test setup would have to be pressurized that greatly increases the cost of experimental campaign. Thus, the majority of the publications on the topic use air-water (and some other liquid-gas combinations) flow loops to verify their models.

However, there is a number of difficulties related to extrapolation of such results onto real production system. Those primarily arise due to the fact that the real production system have pipeline lengths of the order of kilometers, compared to a fraction of that in test loops [9]. Even more important is the effect imposed by significant pressure drop due to friction losses and especially hydrostatic pressure drop. Pressure in turn affects physical properties of the fluids and solvability of the gas fraction in oil.

Severe slugging in real production systems, being primarily governed by gravitational forces and pipeline gas compressibility, exhibits high pressure variations in the riser with potential release of the solved gas; therefore, mass transfer effect must not be ignored in modeling of the phenomenon in subsea production systems [9].

A recent paper by Nemoto and Balino [9] provides a mathematical model for severe slugging description with mass transfer between oil and gas based on Blackoil approximation for relatively light oils ($API < 45$) at isothermal conditions. Simulations based on the model showed the same evolution of the phenomenon through the four severe slugging stages; however, significant changes in the behavior compared to air-water system were observed [9].

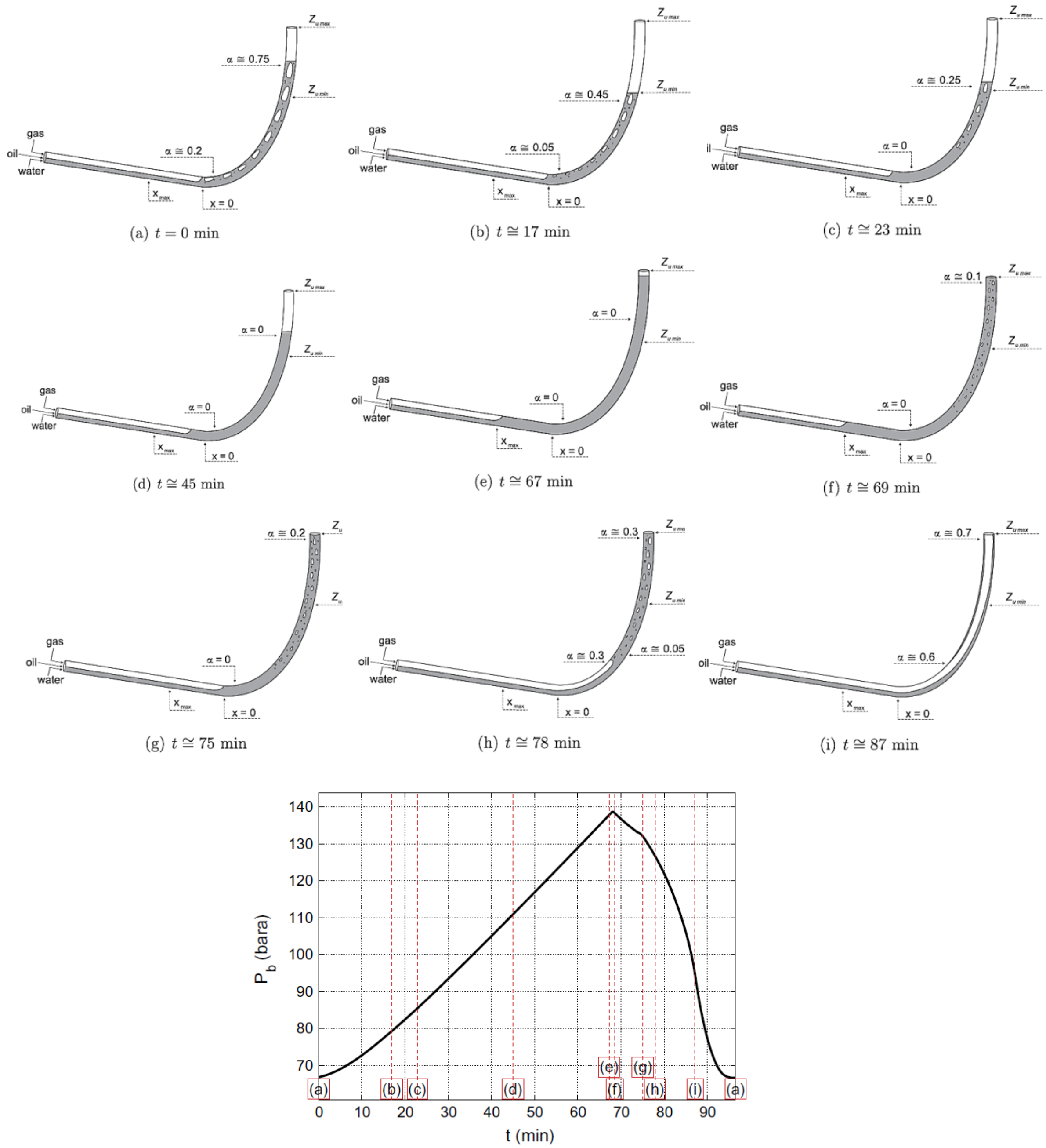


Figure 2.2. Severe slugging of type 1 cycle with effect of mass transfer[9].

A thorough description is provided in text.

The model assumed three-phase flow consisting of oil, water and released gas, where oil and water have the same absolute velocity and exist in homogenized flow. Mass transfer between oil of API equal to 19 and gas of specific gravity equal to 0.6602 was allowed and calculated as an equilibrium state at given PVT conditions according to Blackoil model with initial GOR of 145 [9]. The production system was modelled as 1 km pipeline with 2 degree declination followed by catenary riser of total height of 1300 m; both had internal diameter of 4 inches.

Some qualitative differences were observed over the course of one severe slugging cycle. The following description of one cycle of SS1 is based on [9] and refers to Figure 2.2.

At the initial conditions, chosen to be when the minimum riser base pressure is observed, the riser is partly filled with highly aerated liquid due to partial gas penetration over the riser bend followed by expansion and accompanied by vaporization of some of the gas from the oil, see Figure 2.2.(a). The upper level of the liquid accumulation starts to move down due to relative motion of the gas bubbles and decrease of gas volumes penetrating into the riser, whereas the lower level remains at the riser bend point (b). Then the liquid slug starts moving into the pipeline, completely blocking gas passage (c). At some point all the gas phase has moved to the top of the liquid slug or has condensate into the liquid; no gas observed over the slug length (d). The holdup continues to remain equal to unity until the slug reaches the riser top. As the slug reaches the riser outlet, some gas evaporates from the liquid (e-f). Liquid production phase continues until the point when the pipeline gas reaches the riser bend and initiates blowout (g).The gas front moves up the riser, depressurizing conduit (h). Blowdown occurs when the gas is at the riser top, characterized by high gas flowrate over the stratified flow in the pipe and annular flow in the riser, inducing new portion of transient slugs in the pipeline and initiating the cycle again (i).

An important consequence of the described process is that the liquid production is always associated with some of that of gas due to mass transfer from liquid to gas, opposite to no-mass-transfer case when pure liquid production is possible [9]. It also affects the time profile of riser base pressure, see Figure 2.2.(j). Despite the fact that the process may be seen as SS1, there is no plateau of constant pressure during liquid production phase. That in turn is important for understanding the process as well as flow regime identification in the following parts of the thesis.

2.4. Severe slugging occurrence criteria

Steady state models consider the balance of forces acting on the fluids in pipeline-riser system in conditions close to severe slugging. The main purpose of the steady state models is to find the criteria necessary for severe slugging occurrence. They usually neglect friction forces due to stationary conditions, and assume constant inflow rates.

It is worth mentioning prior to discussion that due to steady state nature, the models consider only severe slugging resulting from the classical severe slugging mechanism and incapable of describing the system behavior with fast variations pipeline holdups, such as that stated by Linga [7].

Under the stationary assumption, three main conditions must exist for severe slugging to form [3]:

1. stratified flow in the negatively inclined pipeline section preceding the riser base;
2. rate of hydrostatic pressure increase in the riser due to liquid build up must be higher than the rate of pipeline gas pressure increase;
3. the liquid column instability inside the riser.

Each of the criteria below address to one of the three preconditions.

2.4.1. Stratified flow criterion

Classical mechanism for severe slugging occurrence requires stratified flow to be present in the pipeline. The criterion for transition to stratified flow in relation to severe slugging was described by Taitel and Dukler [10]. They considered conditions necessary for wave generation in two phase flow between parallel plates and extended the model to the flow in circular pipe, taking into account interaction between the generated waves and gas flow accelerated over the wave crests. The criterion, expressed in terms of superficial gas velocity, can be written as [10]:

$$U_{SG} < K_2 \left[\frac{(\rho_L - \rho_G) g \cos\beta \, \alpha_G}{\rho_G \, d\alpha_L / dh_L}\right]^{0.5}, \tag{2.1}$$

where $K_2 \approx \alpha_G / \alpha_L$,

$d\alpha_L / dh_L$ – change of liquid flow area with liquid height.

When the criterion is followed, stratified flow is expected to be observed in the pipeline.

The relation expressing the liquid holdup with respect to liquid height may be derived based on geometrical considerations [10]:

$$\frac{d\alpha_L}{dh_L} = D\sqrt{1-\left(2\frac{h_L}{D}-1\right)^{0.5}}. \tag{2.2}$$

2.4.2. Bøe criterion

The Bøe criterion [11] focuses on the second precondition for severe slugging, namely that the rate of increase of hydrostatic pressure in the riser must be greater than the rate of increase of the pipeline pressure. The criterion, assuming constant gas and liquid flowrates, can be written as [11]:

$$U_{SL} \geq \frac{P_P}{\rho_L g(1-\alpha_L)L} U_{SG}. \tag{2.3}$$

To resolve the inequality, the liquid holdup must be evaluated from the superficial gas and liquid velocities. The first approximation by Bøe [11] assumed no-slip condition between the phases, resulting in identical absolute speeds and equation for the liquid holdup in the form of:

$$\alpha_L = \frac{U_{SL}}{U_{SL}+U_{SG}}. \tag{2.4}$$

This condition results in a straight line on the superficial velocities graph, distinguishing severe slugging and no severe slugging regions.

More rigorous correlations for liquid holdup, such as that presented in Taitel [12], result in an operational envelope outlining the severe slugging region. Complete derivation of the iterative steps for finding the correlation may be found in the Appendix of Montgomery [13].

2.4.3. Pots criterion

Pots et al. [14] formulated a criterion for liquid build up in the riser resulting in severe slugging. Similarly to the Bøe's criterion, it considered the pressure balance between the hydrostatic pressure of the liquid in the riser and accumulated in the pipeline gas pressure, under assumption that all of the liquid entering the flow line goes to the riser. The resulting condition may be written as:

$$\Pi_{SS} = \frac{zRT/M}{g\alpha_L L}\frac{\dot{m}_G}{\dot{m}_L}. \quad (2.5)$$

Severe slugging of type 1 occurs when $\Pi_{SS} < 1$. Additionally, Pots et al. [14] proposed to use the measure of Π_{SS} to determine the degree of slugging severity, where smaller values correspond to more harsh slugging conditions.

Since both Bøe and Pots criteria are based on the same principle, it may be shown that the Pots criterion takes form of the first one when expressed in terms of pressure and superficial velocities [13].

2.4.4. Taitel criterion

Taitel [12] speculated on the stability of operations under conditions that are close to blowout (precondition 3 discussed above). He considered stability of the liquid column in the riser at the moment of first bubble penetrating the riser bend. If the column is not stable, the bubble will cause the full cycle of blowout, sweeping almost all the liquid from the riser. Otherwise, a bubble or slug flow will occur.

To derive the criterion, the difference of the forces acting on the column before and after the penetrating bubble front was considered. The force difference may be written as [12]:

$$\Delta F = \left[\left(P_{sep} + \rho_L g h_R\right)\frac{\alpha_{GP}L}{\alpha_{GP}L + \alpha'_G y}\right] - \left[P_S + \rho_L g(h_R - y)\right], \quad (2.6)$$

where α'_G stands for the gas holdup in the gas cap penetrating the liquid column.

Then the criterion for stability becomes [12]:

$$\frac{\partial(\Delta F)}{\partial y} < 0 \quad (2.7)$$

for $y = 0$ (at the riser base).

Combining the two equations above and referencing to atmospheric conditions (to express gas flow rate in terms of superficial gas velocity at standard conditions) the criterion takes form [12]:

$$\frac{P_{sep}}{P_0} > \frac{(\alpha_{GP}/\alpha'_G)L - h_R}{P_0/\rho_L g}. \quad (2.8)$$

The resulting criterion is mainly dependent on the pipeline-riser system geometrical parameters and the separator backpressure. The gas holdup values vary within 0.8 to 1.0 interval for stratified flow [12]. The model is relatively insensitive to the value of α'_G, and Taitel [12] advocates that 0.89 is a good assumption. The pipeline gas holdup, α_{GP}, may be expressed in terms of superficial velocities under the stratified flow model, allowing to plot the criterion on a flow regime map.

2.5. Effect on production system

Slug flow in general may impose significant problems on the production system. As a rule of thumb, the longer the slugs the greater negative effects imposed. Therefore, severe slugging may be considered as the most violent form of slug flow. The three parameters of main concern associated with riser induces slug flow are pressure fluctuation, high superficial velocities, and variability in production rate.

High variability of pipeline pressure affects the field productivity, imposing additional backpressure on the producing wells. In general, resulting loss in production is system specific and may vary greatly. In some extreme cases severe slugging may kill a well or cause putting it on hold [15]. Additionally, it causes high stress loads on riser pipework and topside piping [16].

Variability of the production rates on the platform is, by far, the main challenge of severe slugging. As it was demonstrated above, severe slugging is characterized by periods of no production, slow liquid production, and increasing liquid production, reaching extreme flow rates and followed by violent gas blowdown. Pots et al. [14] states that the liquid production at its peak may reach velocities as high as 70 times its average value. As the result, increased vibration, erosion, and general wear and tear rates are experienced by the system. It is especially true for erosion-corrosion processes since they are strongly dependent on the media flow speed. Moreover, droplet induced erosion may be caused by high gas flow rate even without solids being present in the flow [17].

At the same time, variability in the production rate imposes large disturbances in the separator train, decreasing the separator effectiveness, resulting in poor separation of the phases, water carryover to the export system, and improper water treatment, leading to oil content in discharged water and potential environmental requirements violations [15]. Extreme cases may result in separator flooding, emergency shutdowns, and liquid carryover to the process flare system [16]. Moreover, gas flow rate instability causes inefficient compressor operation, resulting in increased maintenance costs and higher flare gas volumes [15].

3. Severe slugging elimination

3.1. Conventional methods

3.1.1. Backpressure increase

As it may be inferred from Bøe and Pots criteria, severe slugging of type 1 may be alleviated by increasing the backpressure imposed on the system, thus increasing the "stiffness" of the gas phase resulting in earlier initiation of the blowout phase of severe slugging cycle. However, the method cannot stabilize the flow completely, merely reducing the length of the slugs produced.

In addition to that, the method is connected to sufficient losses in production volumes. Yocum [18] reports flow capacity reduction up to 70% in some cases. For that reason the technique was deemed to be inefficient even for low depth offshore developments [19].

3.1.2. Choking

Choking as a way to control slug flow in vertical conduits was considered relatively early in the history of multiphase flow production. Thus, Yocum [18] in 1973 elaborates on choking efficiency for slug flow mitigation and states that at that time the method was used inefficiently, inducing excessive pressure losses and decreased production rates.

Topside choking addresses to the flow stability criterion described in Section 2.4.4. Combination of choke and flowline characteristic curves allows reduction of the minimum required gas flowrate to stabilize the flow at the cost of higher pressure drop, see Figure 3.1.

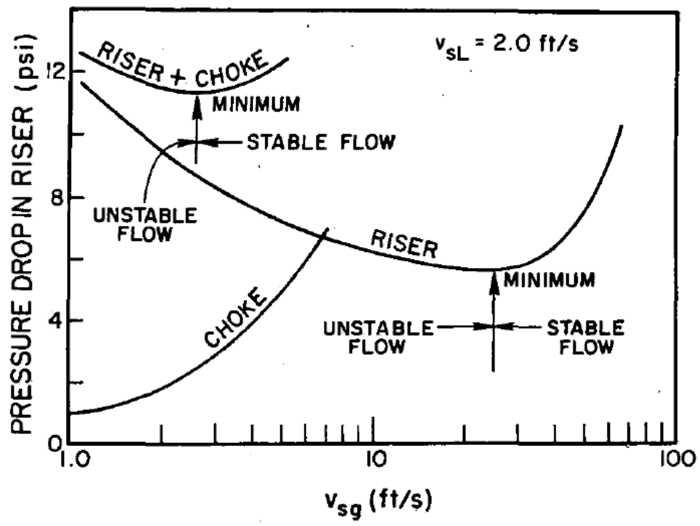


Figure 3.1. Flow stabilization by topside choke introduction [6].

The graph shows the result of characteristic curves combination plotted at constant superficial liquid velocity. It may be seen that superficial gas velocity guaranteeing stable flow is reduced by a factor of 10.

Schmidt [20] recognized topside choking as a means for severe slugging alleviation and later [6] provided a description of the control mechanism. Proportional to the liquid velocity increase of backpressure due to a choke installed topside creates an additional retarding force on the gas phase during the bubble penetration phase. Suppressed bubble could not accelerate to the level of blowdown initiating. Therefore, normal slug flow regime was forced instead of severe slugging.

A more thorough description of the choking mechanism with theoretical and experimental investigation was done by Jansen [21-24]. The work was based on a modified Taitel's stability criterion [12], taking into account additional effect of the choke. Thus, backpressure upstream from the choke right before the bubble penetration occurs may be written as:

$$P_B = P_{sep} + CU_{SL}^2, \tag{3.1}$$

where C – choke coefficient, [Pa·s^2·m^{-2}].

An important assumption of the model for two-phase flow is that time averaged pressure drop over the choke is dependent only on liquid phase and superficial liquid velocity. Therefore, the pressure drop as a function of time is assumed to be a function of both liquid holdup and the mixture velocity [21]. These statements was reported to be verified by experiments of Jansen [21]; however, no specific data about the experiments were provided.

When the gas phase is entering the riser, it creates an additional response from the choke due to acceleration of the liquid column, which can be proved to be linearly dependent on the gas penetration height [21]:

$$P_B = P_{sep} + CU_{SL}^2 + Ky, \tag{3.2}$$

where K – proportionality constant, [Pa·m^{-1}].

Similar to the stability criterion discussed in Section 2.4.4, the force per area difference applied to the interface between the penetrating gas phase and the bottom end of the liquid column may be written as [21]:

$$\Delta F = \left[\left(P_{sep} + CU_{SL}^2 + \rho_L g h_R\right)\frac{\alpha_{GP}L}{\alpha_{GP}L + \alpha'_G y}\right] - \left[P_{sep} + CU_{SL}^2 + Ky + \rho_L g(h_R - y)\right]. \tag{3.3}$$

Following the same logic as it was presented in Section 2.4.4, stability criterion may be found by differentiating the equation above, bearing [21]:

$$\frac{P_{sep} + CU_{SL}^2}{P_0} > \frac{\frac{\alpha_{GP}}{\alpha'_G}L(1 - \frac{K}{\rho_L g}) - h_R}{P_0/\rho_L g}. \tag{3.4}$$

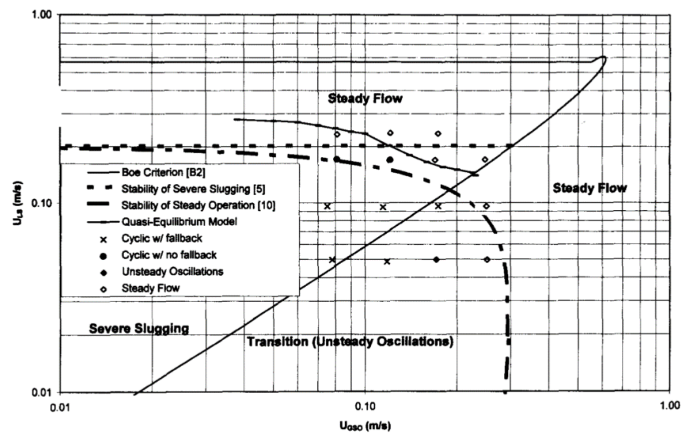
Relation between the choke coefficient and the proportionality coefficient may be found from the equation of motion written for the liquid column in the riser with respect to applied force ΔF [21]. The solution brings [21]:

$$K = \frac{2CU_{SL}^2}{h_R}. \tag{3.5}$$

The model, however, only describes flow stability at the initial point when the riser is fully occupied by the liquid phase. When the steady state operations is established, some fraction of the gas will be present in the riser. Therefore, the average density over the riser is reduced. Neglecting the gas density compared to that of liquid, the average density may be approximated to $\phi_R\rho_L$. The stability criterion then modified as follows [21]:

$$\frac{P_{sep} + CU_{SL}^2}{P_0} > \frac{\frac{\alpha_{GP}}{\alpha'_G} L(\phi_R - \frac{K}{\rho_L g}) - \phi_R h_R}{P_0/\rho_L g}. \tag{3.6}$$

Introduction of a choke into the system increases the flow stability, contracting the unstable flow operational envelope along the superficial liquid velocity axis: the greater the choke coefficient C, the smaller the value of the transitional velocity from unstable to stable flow. Figure 3.2 shows how different choke valve settings affect the flow regime.



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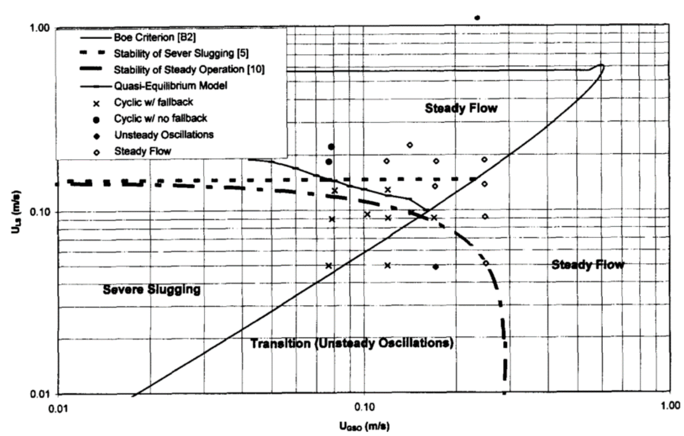


b)

Figure 3.2. Choking effect on the severe slugging envelope [21].

Fig. a) shows stability region for choke set to C=120000 Pa·s^2·m^{-2} and b) for C=245000 Pa·s^2·m^{-2}. It may be seem how with increased choke resistance, severe slugging envelope contracts along liquid velocity axis.

3.1.3. Riser-base gas injection

The method of gas injection for severe slugging remediation is based on the artificial gas lift principle and, according to Mokhatab and Towler [25], was one of the most frequently used. Gas injected in the riser base is deemed to lower the static pressure of the liquid column and shift the flow regime to annular or dispersed flow, thus solving the problem of slugging. However, Mokhatab and Towler [25] highlight that gas injection may be relatively useless for transient slugging when an already formed liquid plug arrives to the riser.

Despite the fact that the method was considered not economically feasible by some researchers at the beginning of 80s [6], some investigation was conducted. Pots et al. [14] experimentally studied effect of gas injection on severe slugging and concluded that for injection rates of about 50% the slug arrival velocities are considerably lower than in the case without treatment. However, about four times the amount of produced gas must have been injected to completely eliminate slugging, approaching annular flow regime [14]. Hill [26] reported real case implementation of gas injection on the S.E. Forties field, indicating reduction in severe slugging extension. The method also succeeded in reviving some of the wells previously killed-off by the back pressure from the riser slugging [27].

The method feasibility and effectiveness reduces with the water depth owing to increased pressure loss on friction and gas compressibility. Thus, gas cooling due to Joule-Thomson effect rising from high water depth makes the flow conditions more susceptible to wax and hydrate formation inside of the riser [25]. In addition to that, higher water depth requires the gas injection line to be longer, making it more costly and the friction losses more prominent.

Jansen et al. [21] described the performance of the gas lift method based on the Taitel's stability criterion. The modified stability criterion may be written as [21]:

$$\frac{P_{sep}}{P_0} > \frac{\frac{\alpha_{GP}}{\alpha'_G} L - h_R}{P_0/\phi_R \rho_L g}, \tag{3.7}$$

where ϕ_R is the average liquid holdup over the riser length, that can be expressed in terms of gas and bubble velocities:

$$\phi_R = 1 - \frac{U_{SG,R}}{X_0 U_M + Y_0}. \tag{3.8}$$

Parameters X_0 and Y_0 are the drift parameter and drift velocity, respectively; both vary depending on the flow regime inside of the riser.

Gas injection affects the flow map diagram in a way that the severe slugging is contracted similar to the case of choking with the difference that contraction occurs along the superficial gas velocity axis with increasing rate of gas injection [21].

3.1.4. Active control methods

The basic principle behind all of the active control methods is to make use of online measurement of the process and/or pipeline information to control the system by adjusting available degrees of freedom such as choke opening level, process pressure, and levels inside of the process facilities [15].

Feed-forward control relies on prediction of slug occurrence and preparation of the separator train to accommodate them [15].

The method of slug flow suppression based on the active control of the choking system got a lot of attention in 90s and was successfully introduced into practice later. The method relies on the pressure readings either from subsea pipeline or topside facilities and based on prediction of the slug slow occurrence. As the result, choke valve, either subsea or topside, is adjusted accordingly and the flow is maintained within required regime. Solutions of that kind were developed by many of the oil and gas producers and service companies, such as ABB, Statoil, and Total [15, 28-30].

The active choking technique has a significant advantage over the classical manually operated choking for slugging suppression since it may operate with average valve opening greater than that of regular choking, e.g. [15]. The stability achieved due to online tuning of the valve opening, thus maintaining desired multiphase flow regime.

Slug Suppression System (S^3), belonging to Shell, was first published in 1995 [31] and assumed installation of an additional “mini-separator” (in later modifications substituted with a piping section [32]) prior separation of the phases with sufficient measuring of gas and liquid flowrates and pressure, see Figure 3.3. Based on the readings and processing, separate chokes for gas and liquid are actuated to facilitate flow stability.

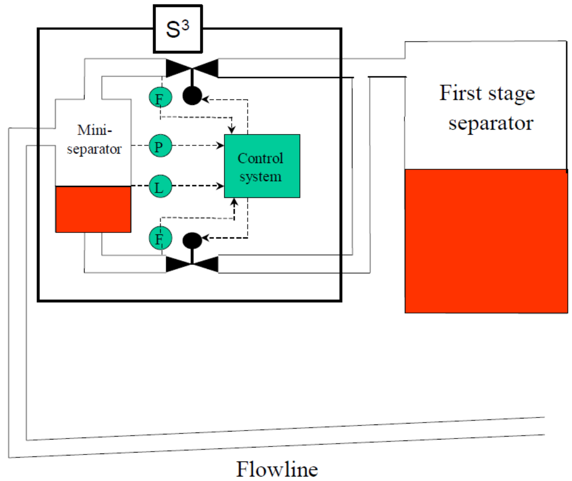


Figure 3.3. S^3 principle scheme [33].

Based on the readings of the pressure (P) and liquid level (L) from the mini-separator and from the gas and liquid flowmeters (F), the control system actuated the one phase chokes.

The manufacturer states that the technology creates less backpressure on the pipeline system compared to regular topside choking (only 1 to 3 bar pressure drop over the system in control mode claimed [32]) and occupies considerably less space compared to a slug catcher [33].

3.2. Unconventional methods

3.2.1. Self-lift

The method of severe slugging elimination based on injection of the produced gas from the production pipeline into the riser base, in contrast to that when the gas is treated and compressed on the platform and then transported through a separate conduit to the riser inlet, was first proposed by Barbuto in 1995 [34].

The statement consisted of a bypass, connecting the pipeline and the riser, allowing the flow of gas and its injection at the predetermined position of the riser. It was said to be located at approximately a third of the total riser height [34]. In addition, the patent describes some of the ways to control throughput of the bypass line to alter the pipeline pressure. However, no further information, tests or theoretical proof followed.

An investigation of the method feasibility was conducted at the University of Tulsa and published in the thesis of Tengesdal [35]. The experimental campaign of the thesis was run on a test facility of 65 ft. [19.8 m] pipeline followed by 49 ft. [14.9 m] riser, both with ID of 3 in. [76.2 mm]. Effective pipeline length for the system was equal to 280 ft. [85.3 m] due to tank at the pipeline inlet. A conduit of 1 in. [25.4 mm] with installed choke valve was used as a bypass allowing connection with any combination between 4 take-off points on the pipeline and 4 injection points on the riser, see Figure 3.4 [35]. Therefore, the facility allowed investigating effectiveness of the system with variation of all the parameters of interest: gas/liquid superficial velocities, inclination, take-off/injection point location, pressure drop over the bypass.

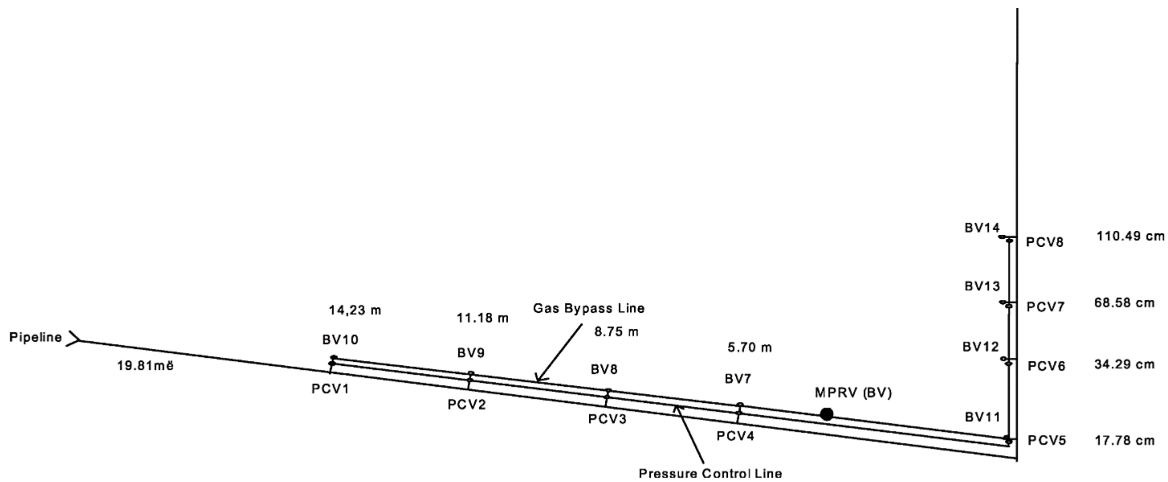


Figure 3.4. Pipeline-riser system with bypass used in experiments of [36].

The drawing shows take-off and injection points locations for bypass length adjustments:
PCV 1-8 – pressure control valves;
BV 7-10 – ball valves to operate the take-off point location;
BV 11-14 – ball valves to operate the injection point location;
MPRV – ball valve used for choking.

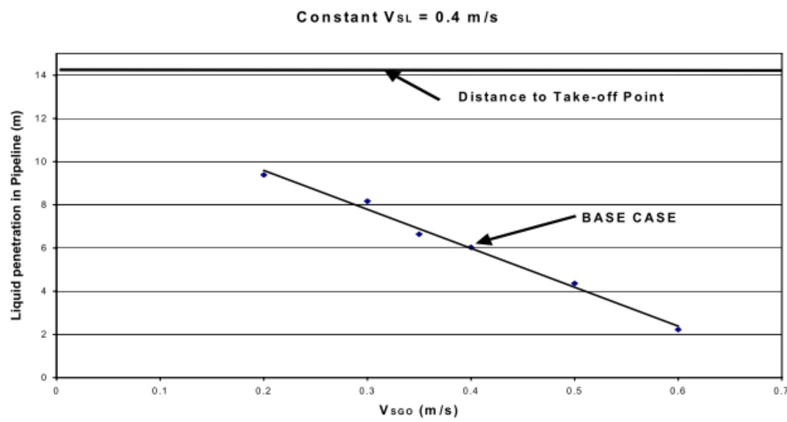
Overall, a stable production point was found for any combination of superficial gas and liquid velocities and pipeline inclination with the use of choke valve installed in the bypass line. However, a stable flow could be achieved for fully open choke valve only with relatively high superficial gas velocities, due to increased pressure drop over the bypass; otherwise, the system should have been tuned with the aid of the choke valve.

It was found that situations when the injection and take-off point levels were close, stabilized flow conditions were obtained more easily due to pressure balance between them [35]. Was the injection point moved upwards, the take-off should has been moved upstream away from the increased liquid penetration level into the pipeline, or, alternatively, additional pressure loss over the bypass could be provided by the choke. Otherwise, the bypass inlet is blocked with liquid, resulting in instable flows.

An extreme situation of another sort occurs when the injection point is too close to the riser base. In that case, dual gas penetration may happen, one from the bypass injection point and one from the riser base, resulting in instable flow [35].

Ironically enough, one of the necessary features for severe slugging occurrence – the pipeline declination – has a positive effect on the self-lift severe slugging elimination technique. With increasing level of the pipeline inclination, it is easier to achieve a stable flow. As experiments by Tengesdal [35] show, at -1° downward angle, the stable flow was easily interrupted by perturbations between the phases. The gas phase fingers towards the riser base and, if reaches the riser bend, partial blowdown occurs, causing instabilities in the pressure and flow rate levels. At higher declination angles, a better separation between the phases occurs and gas is less likely to reach the riser via the riser base.

The self-lift technique is relatively insensitive to the liquid and gas flow rates. One of the main limiting factors when the designing the system for different flow rates, is the distance to the take-off point, see Figure 3.5. Presumably, it is possible to find an optimal choke and distance to the take-off point combination satisfying all encountered flow rates during the field life.



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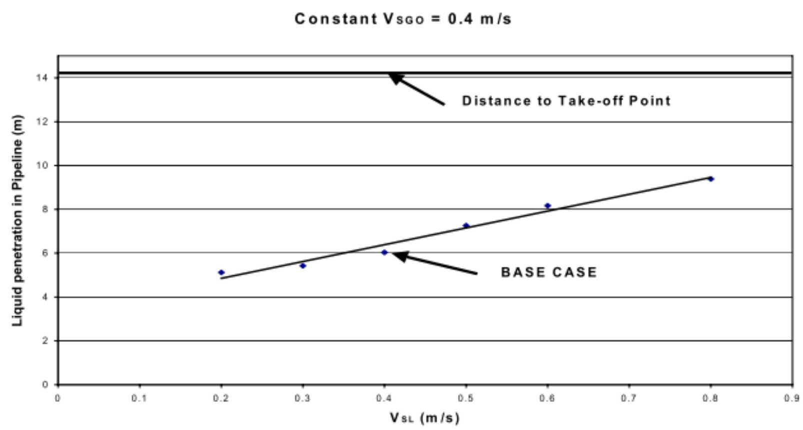


b)

Figure 3.5. Flow rate sensitivity with constant superficial liquid (a) and gas (b) velocities [35].

Declination level -3°. Injection point: BV13; take-off point: BV10.

3.2.1.1. Self-lift steady-state model

In his work Tengesdal [35] derived two steady-state models of the technique: a simplified and a rigorous one. Both of them are based on the analysis of the hydrostatic balance between cross sections A and B, see Figure 3.6.

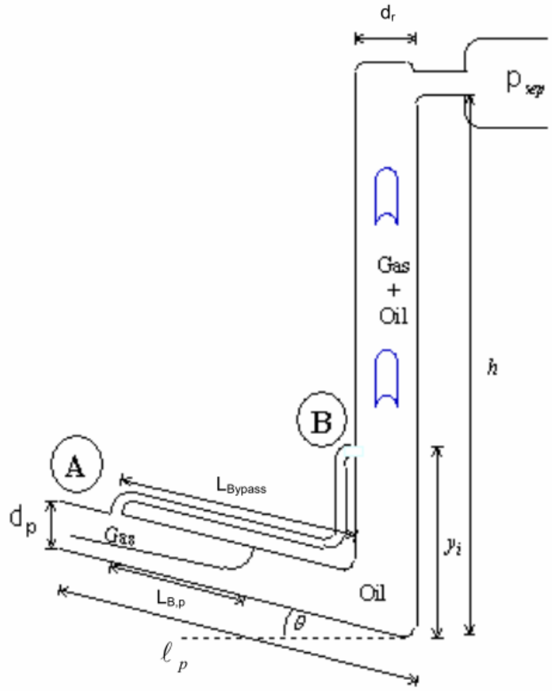


Figure 3.6. Schematic depiction of the self-lift system [36].

The simplified model assumes no pressure loss on friction and, constant PVT and flow properties of the fluids due to small variations in pressure and temperature within area of interest, and that density of the gas, which is much less then density of liquid, may be assumed to be zero [35].

The resulting criterion for stable flow with installed bypass may be written as [35]:

$$\frac{dP_{sep}}{dt} + \frac{g}{A_R}\dot{m}_L \leq \frac{zRT}{M\left(l_P - l_{Bypass}\right)A_P\alpha_{G,P}}\dot{m}_G. \tag{3.9}$$

The criterion is similar to that of Bøe for severe slugging occurrence in the riser system.

The more strict criteria takes into account the flow pressure loss over the distance and the riser base elbow. The new limiting criterion becomes the bubble penetration length, L_{Bub}, which must lay within $0 \leq L_{Bub} \leq L_{Bypass}$. If the bubble length is less than 0, meaning that the bubble recedes to the point above the take-off point, and the liquid enters the bypass. The bubble length of greater than L_{Bypass} would mean that the gas front reaches the riser base elbow [35]. Both of the scenarios would result in unsteady flow with slugging in the riser of sufficient length.

Therefore, to satisfy the steady flow criterion, the following system of equations must be resolved with respect to the parameters of interest for design process [35]:

$$\begin{cases} \Delta P_{L,min} \leq \Delta P_{bypass} \leq \Delta P_{L,max} \\ \Delta P_{L,max} = \rho_L g(y - D_P) + \dfrac{2f_{L,R}\rho_L}{D_R}\left(\dfrac{Q_0}{A_R}\right)^2 (y - D_P) + \Delta p_{elbow} \\ \Delta P_{L,min} = \rho_L g\left(y - D_P - L_{bypass} sin\theta\right) + \dfrac{2f_{L,P}\rho_L}{D_P}\left(\dfrac{Q_0}{A_P}\right)^2 L_{Bub,P} + \\ \quad + \dfrac{2f_{L,R}\rho_L}{D_R}\left(\dfrac{Q_0}{A_R}\right)^2 (y - D_P) + \Delta P_{elbow} \\ \Delta P_{bypass} = \Delta P_{bypass,fric} + \Delta P_{bypass,choke} \end{cases} \tag{3.10}$$

According to Tengesdal [35], the model predicted accurately occurrence of severe slugging. The accuracy of the model was low in cases when the liquid-gas interface was not stable enough. When liquid penetration level is close to either the take-off point or the riser base, oscillations of the interface could cause entering of the liquid into the bypass or partial blowdown, respectively. The effect was less prominent with increased pipeline declination, as it was mentioned earlier. Thus, as Tengesdal [35] demonstrates, for declination of -1°, 25 out of 92 fell outside of the prediction model; for declination -3°, 5 out of 151 tests were predicted erroneously by the model; for declination of -5°, all 162 experimental points were predicted correctly. The outliners corresponded to the cases when the liquid-gas interface instability could affect the stable operation of the system.

3.2.2. A conduit of reduced diameter

The method relies on the principle that the gas and liquid superficial velocities may be increased by reduction of the conduit diameter, moving the operational point out of severe slugging envelope. As early as 1973, Yocum [18] proposed to make use of this fact by using a riser with smaller diameter or substituting it with a number of such risers.

Wyllie and Brackenridge [37] proposed to use an insertion of a smaller diameter pipe inside of the riser for severe slugging mitigation. The annulus (number 2 on Figure 3.7) was presumably used as a gas injection point further enhancing effectiveness of the concept. However, the insertion could impose restriction on pigging, which is an important flow assurance measure for wax deposits cleaning, especially for deep waters [19]. Later a patent was filled by Wyllie [16] for an apparatus of such a design. The patented solution allowed the insertion to be retrievable, however a workover operation was needed. The conceptual drawing is presented on Figure 3.7.

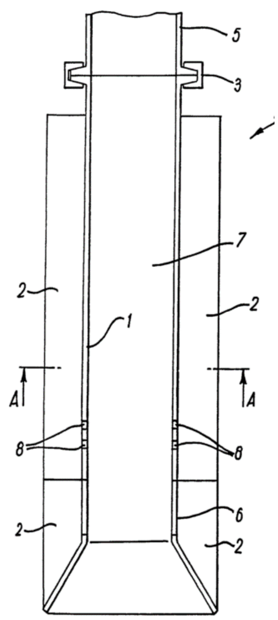


Figure 3.7. Smaller diameter conduit insertion [16].

The riser insertion device. Perforations denoted on the figure by number 8 allow gas injection into the flow.

3.2.3. Flow conditioning, phase agitation, mixing

The concept of remixing of the multiphase flow right before the riser entrance for slug flow alleviation was first mentioned by Yocum [18]. He proposed to use either helices with properly selected pitch, turbulence inducing flow restrictors (bafflers or impact mixing barrels), or agitation by mixers rotated from the platform [18].

The purpose of a mixing tool being present in the conduit is to break up the stratified flow inside of the pipeline, which is one of the preconditions for severe slugging. Entering the riser in form of droplets, liquid, assuming that the gas velocity inside of the riser exceeds the critical velocity, cannot accumulate at the riser base, creating the conduit cross section blockage, therefore, the classical mechanism for severe slugging initiation is eliminated.

The fact that the technique is passive makes it attractive for implementation in extremely remote conditions, such as for terrain induced slugging mitigation in horizontally drilled wells, see [38]. Recently, a series of studies on that topic was conducted by Brasjen et al. [39, 40]. Number of passive devices, such as mixers, swirls, perforated liners and choke, was investigated. The researchers highlight quick restructuration of the mixed flow downstream the mixing devices. Therefore, it was concluded that the optimal placement of such a device is near to flowline undulation dip. Reduction of slugging frequency up to 16% in large scale test facility was observed, however, at the price of significant pressure drop over the system [39].

3.2.3.1. Phase agitators

Some of the proposed designs assumed possibility of the multiphase flow homogenization by introducing some kind of obstacle to the flow stream, by that facilitating the mixing processes in the stream, as it was mentioned referring to Yocum [18].

To that category the patent after Arnaudeau and Corteville [41] may be attributed. The authors proposed a train of mixing members of a defined shape to be placed inside of the multiphase conduit. The members were to be jointed on a rod and, supposedly, lowered down by means of a wireline, see Figure 3.8.

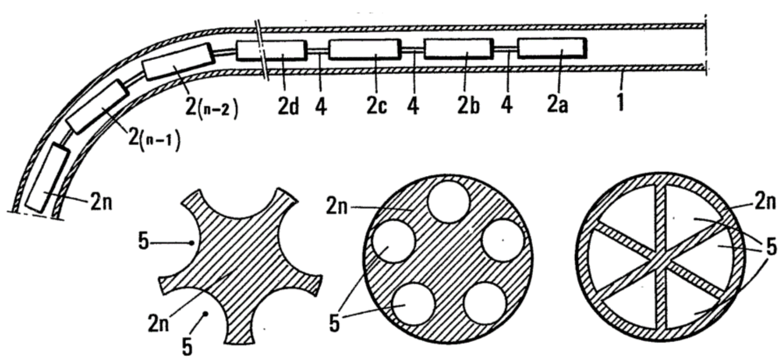


Figure 3.8. Mixing members inside of a conduit [41].

The three circular cross sections give an example of mixing members geometry.

However, no further information on the topic may be found in literature. In 1992 the patent was ceased due to non-payment of the annual fee, probably indicating abandonment of the invention.

3.2.3.2. Undulation

Despite the fact that undulating shape of the pipeline may be one of the preconditions to severe slugging [21, 38], similar to the effect of downwards inclination of the pipeline, some research interest was generated around the altering of the pipeline shape as a passive method for severe slugging alleviation.

The patent by Makogon and Brook [42] proposes installation of at least one section of pipeline configuration, consisting of a positive inclined pipe followed by a horizontal and declining section, see examples on Figure 3.9. It was claimed that the setup of proper configuration and placement allows breaking up of severe slugging flow into smaller slugs, at the same time not restricting pigging operations due to appropriate bending radiuses [42].

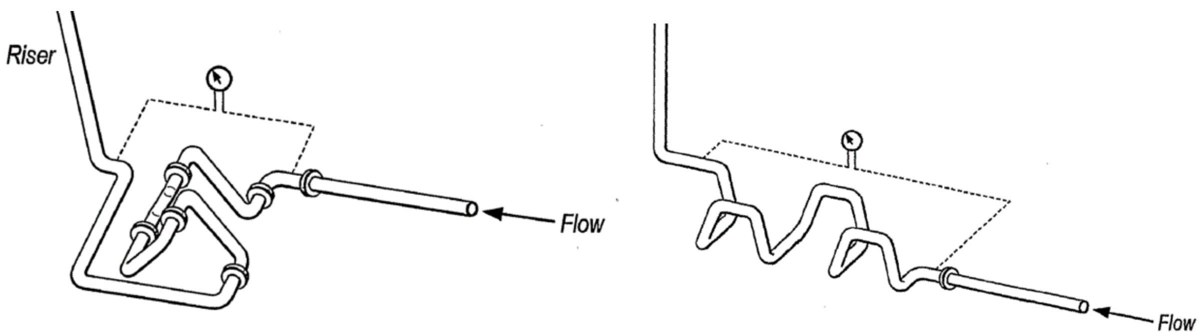


Figure 3.9. Examples of flow conditioning by Makogon and Brook [42].

The device was tested by Makogon et al. [43] over the commercial transient multiphase flow simulator OLGA and on scaled down model at the University of Tulsa. The concept proved to be viable and provided severe slugging alleviation, reducing the pressure fluctuations over the riser due to generation of smaller slugs.

Another undulating pipeline design is mentioned in the patent belonging to Shen and Yeung [44], which describes a pipeline configurations to serve the purpose of stratified flow mixing, assembled from consecutively jointed tubular segments with predetermined angles, forming the shape of a spiral or a wave, see Figure 3.10.

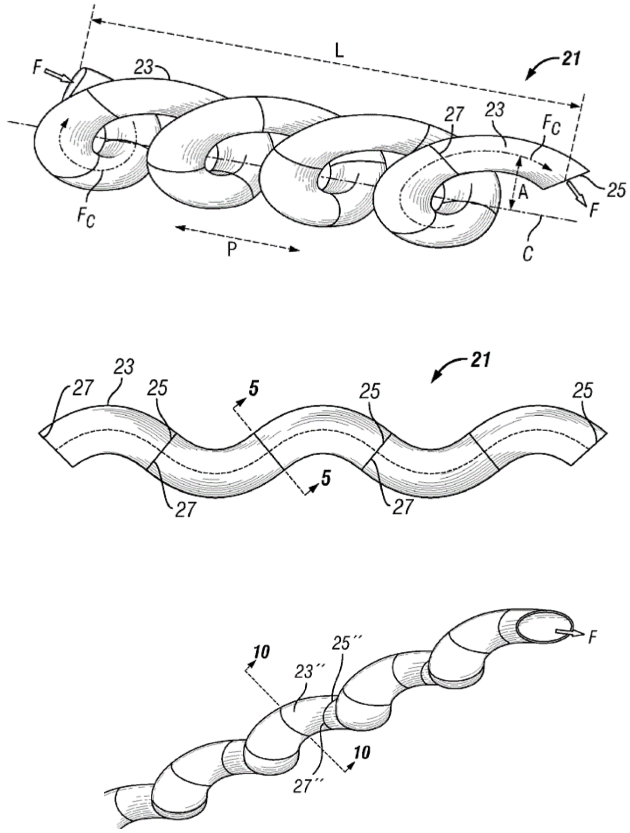


Figure 3.10. Undulating pipes design proposed by Shen and Yeung [44].

Each section of the pipe is a standard piping bend of 90 degree.

It may be inferred from the patent [44] that the shape depicted in the middle of Figure 3.10 (later called “wavy pipes”) showed the best results over others, providing the greatest reduction in the operational region attributed to severe slugging.

Additional research was conducted by Xing et al. [45-47] on the effectiveness of the wavy pipes for slugging alleviation. The wavy section was assumed to be made of standard piping sections, therefore allowing pigging operations and having lower manufacturing costs.

Experimental study on the topic published in 2013 by Xing et al. [47] proved the concept to be relatively efficient in the small scale model. The authors claim that the presence of the wavy pipe section before the riser base consistently reduce the severity of slugging by reducing the slug length and the severe slugging occurrence region on the flow regime map. For a number of test trials, transition from severe slugging to oscillatory flow was achieved. It was also found that the pressure drop over the system with wavy pipes installed in it was consistently lower compared to the case without those. The number of the up-down section was showed to have correlation with the method efficiency, provided some distance between the wavy section outlet and the riser base.

3.2.3.3. Venturi tube

An interesting approach for severe slugging elimination was proposed by Almeida and Goncalez in 1999 [48]. The patent assumed installation of a venturi tube near the riser base for temporary rearrangement of stratified flow into non-stratified flow over the distance sufficient to overpass the riser bend.

The device represents a venturi tube consisting of convergent-divergent nozzle installed inside of the conduit, as it is shown on Figure 3.11.

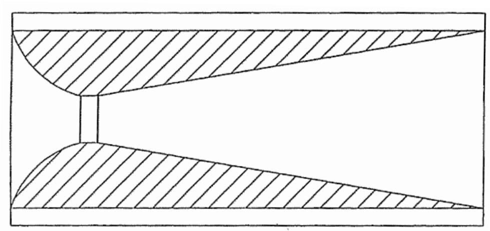


Figure 3.11. Venturi tube for stratified flow [48].

In addition to stratified flow rearrangement the patent claims that the device, due to sudden drop in pressure in the throat part of the nozzle, is capable to release significant quantities of the gas dissolved in the liquid phase [48]. According to the claim, it would further contribute to establishment of a non-stratified flow regime.

The consecutive paper [49] on the experimental investigation of severe slugging elimination effectiveness with venturi demonstrated that pressure fluctuations may be suppressed with the device, provided that it has sufficiently small throat diameter. The test trials comparing the venturi to the choking showed almost identical results.

Criticism.

The paper claims that venturi would impose "the smallest pressure drop through the device", despite the fact that only experiments on water and air were conducted [49]. For the real case scenario this statement is rather doubtful. Such effects as mass transfer between the phases and possible presence of water in oil-gas flow, forming highly viscous dispersion with oil, must be taken into account to make a sound statement about the pressure loss over a venturi tube in multiphase flow.

Second, the paper [49] compares effectiveness of choking to venturi in severe slugging mitigation. However, no information is given considering the choke setup; therefore, no judgment may be given about how efficiently the choke was set.

Third, the paper [49] considers devices with two different throat diameters: 6 and 8 mm. The former one eliminated severe slugging, whereas the latter failed to do so. To illustrate the effect, authors plotted three points on the flow pattern map, representing regimes with the two devices (6 and 8 mm throats) and without it. The points plotted correspond to the superficial velocities inside of the throat (it may be inferred from the fact that all three points are plotted along a line, having the angle of 45 degrees and from the statement given in the paper and from the wording of the paper). It is highly doubtful that such an assumption is valid for description of the overall regime in the pipe. Superficial velocities will return to their initial values after they exit diffusor, especially taking into account "the smallest pressure drop through the device" (no sufficient gas compression happens; the statement is always true for incompressible fluid, given constant boundary conditions of the conduit system).

Based on the arguments presented above, the approach chosen by the authors to describe the proposed may be questioned.

3.2.4. Separation

Presence of at least two phases with different densities is necessary for severe slugging occurrence. Therefore, terrain slugging, in principle, must be completely eliminated by separation of the liquid and gaseous phases. However, some researchers point out at observed intermittent flow regimes in pipelines transporting only oil and water, due to the fact that oil is more viscous and less dense and the two liquids are immiscible.

The concept of subsea separation may eliminate slugging flow in the riser and increase production; however, it is associated with a number of problems. First, after the gas phase is separated in must be transported via an additional conduit. Provided that there is no pipeline or/and riser available for this purpose, the concept may turn out to be economically not feasible. Benetti and Villa [50] propose to utilize an existing pipeline of two parallel pipelines from a field, one of which was abandoned due to flow assurance considerations, when a subsea separation system is installed near to wellhead.

In addition to that, the liquid flow separated from gas will impose additional backpressure due to increased averaged density in the riser, therefore increased hydrostatic pressure. For that reason a typical subsea separation system must include a pump for the liquid phase boosting. Operations of a pump added to that of a separator may be considered troublesome.

There are a number of benefits associated with subsea separation and boosting apart from slugging elimination and production increase. Thus, implementation of the technique may be economically efficient due to reduced investments in flow assurance management. Two-phase separation, boosting, and further transportation via separate flowlines depressurizes the system, considerably reduces amount of water in gas-phase flowline, and significantly decreases Joule-Thomson cooling of the entrained liquid, all of which positively affect hydrate curve, lessening the required amount of glycol injected [27]. Potential benefits from subsea separation rise with increase of water depth and proved to be important decision drivers for such projects as Perdido [51], Parque das Conchas [52] and Pazflor [53].

Three-phase separation (with additional separation between oil and water) brings additional benefits such as potential reduction of topside capacity and backpressure and possibility to re-inject the separated water in situ. However, it goes beyond the scope of the thesis.

The remaining portion of the section gives a brief overview of two-phase subsea separation and boosting solutions that have already been tested and proved their effectiveness. For information on other subsea separation technologies not involving the boosting component consult Hannisdal et al. [27].

3.2.4.1. Vertical annular separation and pumping system (VASPS)

Vertical Annular Separation and Pumping System (VASPS) concept was proposed at the beginning of 90s as a subsea solution combining two phase separation and liquid phase boosting. The system had significant height and was supposed to be installed and secured in vertical position inside of a 30 or 36 in. conductor pipe in a dummy well of approximate depth of 30-70 meters, drilled and completed by conventional techniques [50].

The VASPS body unit consisted of three concentric casings [50]:

1) Outer casing with nominal diameter of the conductor pipe provided pressure integrity of the system;
2) Helix Separator Joint (intermediate casing) consisted of a 16-20 in. pipe with perforations and welded to the outer surface helix plates, spiraling to the bottom of the assembly and being in sufficient sealing contact with the inner surface of the outer casing;
3) Liquid Discharge Tubing (inner conduit) was a 8-10 in. pressurized conduit for the liquid phase pumped from the bottom of the assembly by an ESP, accommodated inside the tubing.

Thus, a liquid-gas mixture enters the outer-intermediate casing annulus and forced to flow in spiral due to gravity drainage and initial momentum. Under the action of gravity and induced centrifugal forces the gaseous phase is separated and escapes the helix through the perforations in the intermediate casing, entering the annulus between the intermediate casing and inner tubing from which it is directed to the gas pipeline. The liquid phase is drained to the bottom of the assembly where it completes its degassing in a sump and then pumped by the ESP into the inner tubing and directed into the liquid phase pipeline to the process facilities. Figure 3.12 provides schematic illustration of the VASPS.

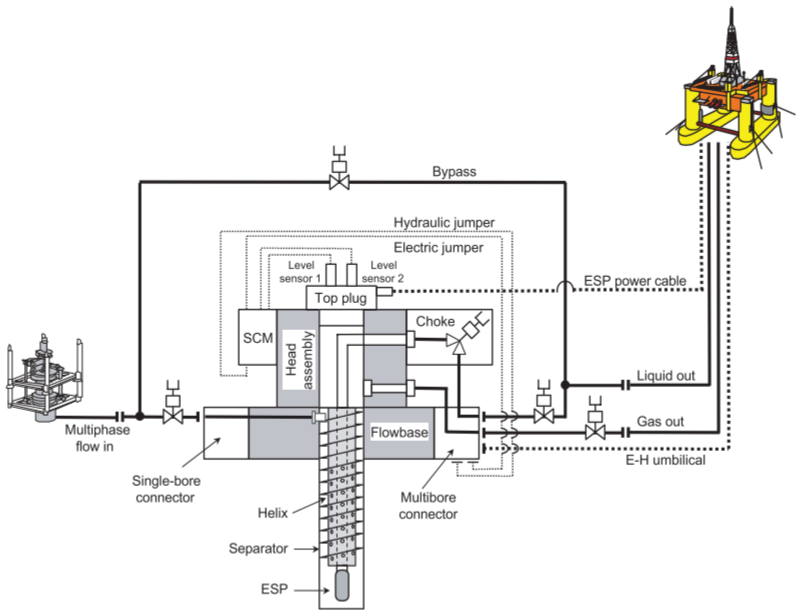
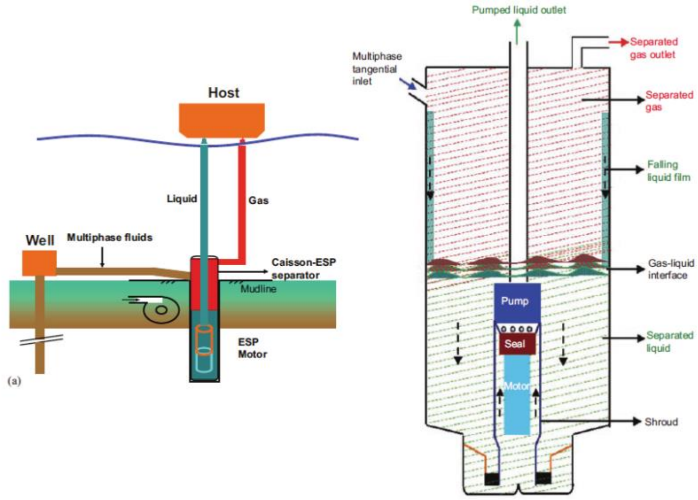


Figure 3.12. Schematic depiction of the VASPS principle and main components [54].

Initial phase of the field test of the system in 2001 at the Marimba field, Brazil, proved VASPS to be a feasible solution. Installation of the system on the well MA-01 reduced the wellhead pressure from 36 to 11 bars, increasing production from 750 m^3/d to 1000 m^3/d and eliminating the need for injection of 100,000 m^3/d gas-lift gas [55].

3.2.4.2. Caisson-ESP (C-ESP)

Caisson-ESP concept is similar to that of VASPS. It consists of a vertical caisson of approximately 100 meters of height secured on the seabed inside of a dummy well drilled and completed for that purpose. A multiphase flow from producing wells enters the assembly at the top through a specifically designed angled and tangential inlet. Thus the fluid enters cylindrical cyclonic separator where the gas is separated from the liquid phase under the action of gravitational and centrifugal forces. The liquid falls to the bottom of the caisson where it is pumped by an ESP into a dedicated liquid phase outlet, whereas the gaseous phase expands into the gas outlet [52], see Figure 3.13.



a) *b)*

Figure 3.13. Caisson-ESP concept [52].

a) Schematic depiction of field development with use of C-ESP
b) Simplified drawing of a C-ESP

The concept was implemented on Perdido [51] and Parque das Conchas [52] fields. Iyer et al. [52] report that implementation of C-ESP on Parque das Conchas allowed efficient severe slugging management, which was predicted to occur at intermediate GLR values, among other benefits.

3.2.4.3. Topside compact separator I-SEP by Caltec Ltd.

A different approach to severe slugging mitigation with phase separation was undertaken by Caltec Ltd. They propose to use an in-line compact uni-axial cyclonic separator and after a choke valve installed topside.

A paper by Jones et al. [56] from 2014 reports a series of experimental trials aimed at investigation of severe slugging mitigation capabilities of I-SEP® (compact cyclonic gas/liquid separator by Caltec Ltd.). The experimental campaign conducted at the Cranfield University, utilized its three-phase flow test facility using water and air.

The comparative analysis between the topside choking and combined choke/I-SEP® severe slugging mitigation techniques was conducted. The choking/I-SEP® combination proved to be a more efficient flow stabilizer compared to standard topside choking, reducing the pressure oscillations by up to 75% at the same pressure drop over the test facility [56]. Moreover, in passive regime I-SEP® reduces the pressure fluctuations at the riser base by 60%, whereas the flow is stabilized completely at higher relative choke opening and lower riser base pressure, see Figure 3.14.

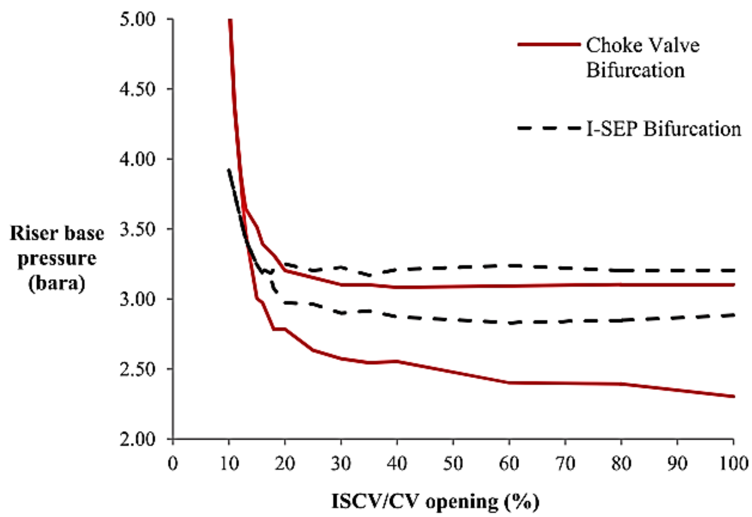


Figure 3.14. Standard choking compared to choking/I-SEP® bifurcation diagrams [56].

The authors point out that there is no rigor quantitative explanation to the effect of the I-SEP® on severe slugging [56]. It well may be that the separator applies additional retardation force during the blowout phase further stabilizing the flow in a manner comparable to choking, however utilizing different mechanism. I-SEP® is more sensitive to a perturbation in flow rate, resulting in greater stability benefit, thus there is active control involved in the use of I-SEP®.

3.2.5. Foaming

Hassanein and Fairhurst [57], discussing challenges related to mechanical and hydraulic design of risers in deep water, pointed out that with increased water depth occurrence of bigger slugs was expected owing to increase in flow line diameter. They proposed the use of foams as a means to control severe slugging [57]. However, no further information regarding this technique followed, neither experimental nor theoretical.

In 2014, Sarica et al. [58] published the results of the first experimental investigation of the severe slugging mitigation using foaming. The foam was generated inside of the pipeline near to the riser base by injecting surfactant into the flow.

The experiment was conducted on the experimental facility at the University of Tulsa aimed at modeling severe slugging, which consisted of 20 m long pipeline section with negative inclination of 3° followed by a 15 m riser. Inner diameter of the pipeline was equal to 3 in. Air and water were used as testing fluids. Heavy Duty Foaming Agent SI-403 was used as foaming agent [58]. It is a blend of surfactants that is used for unloading water from depleted gas wells and is capable of forming stable foams under various conditions and phase composition of the transported fluid. The injection point was located at the end of the pipeline, near to the riser base.

A number of tests were run without surfactant injection to contour steady and transitional slugging (SS2 and SS2) envelopes, followed by test with the foaming agent added to the flow at riser base. Figure 3.15 maps all the trials.

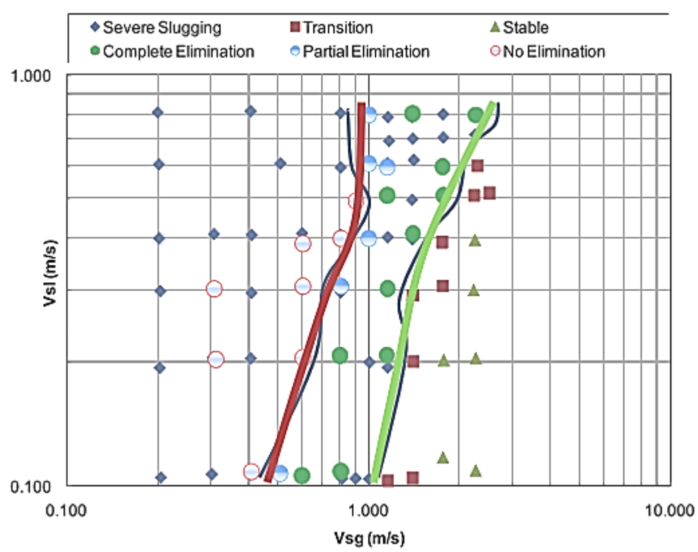


Figure 3.15. Severe slugging map comparison [58].

Diamonds, squares and triangles represent results of the base case, whereas circles the trials with foaming agent injected. The thick lines contour the efficient foaming region.

Addition of surfactants shifted the severe slugging curve to the left, assuring stable or near to stable flow at lower superficial gas velocities, approximately halving the severe slugging region [58].

Figure 3.16 gives a closer look at one of trials completely eliminating severe slugging. At some point of the experiment, the pressure fluctuations in the riser drops almost to zero as it is shown by the blue dots, indicating establishment of stable flow. In addition to that, the total pressure loss over the riser length was significantly reduced compared to the average value between the maximum and minimum of the pressure fluctuations, which is typical for other anti-slugging techniques [58]. Therefore, implementation of the surfactants, in addition to its effect of slugging reduction, brings along reduction in backpressure on the pipeline.

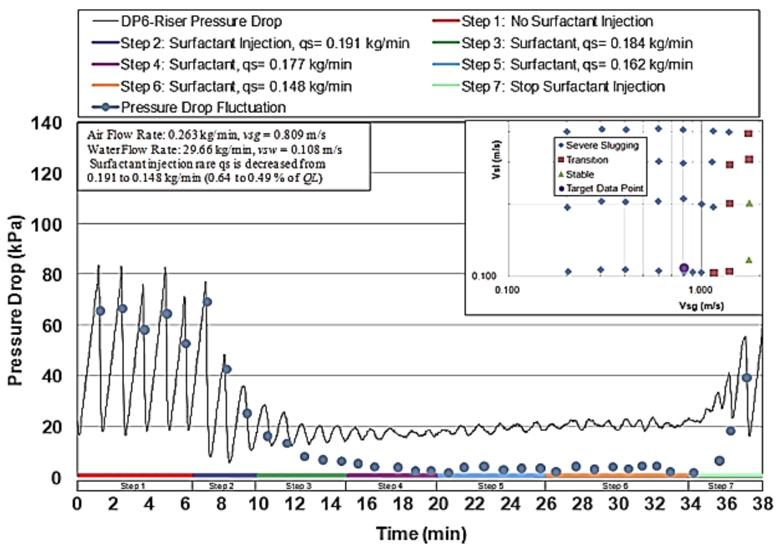


Figure 3.16. Severe slugging elimination case [58].

The purple point on the right part of the figure indicates the case position on the slugging map. The blue dots depict the value of the pressure fluctuation over one cycle. Notice how the mean value of the pressure drop is lower with surfactant added compared to no injection case (Step 1).

Independent of the amount of the foaming agent injected into the flow, surfactants cannot eliminate slugging for some combinations of liquid-gas flowrate, as it may be seen on Figure 3.15 (blue dots with red rim). Moreover, surfactants may cause detrimental effects on the pressure fluctuations in some cases, slightly increasing its value. That is related to improved sweeping of the liquid during the blowdown phase caused by the presence of the surfactant [58].

An overview of all the trials with respect to their effect on slugging is mapped by Figure 3.17. The measure of reduction in pressure fluctuations is expressed by dimensionless elimination performance index (EPI), which is expressed as [58]:

$$EPI = \left[1 - \frac{(\Delta P_{max} - \Delta P_{min})_{after\ inj.}}{(\Delta P_{max} - \Delta P_{min})_{before\ inj.}}\right]. \quad (3.11)$$

It worth mentioning here, that in Sarica et al. [58] all the experiments were conducted with constant pressure at the riser outlet. For that reason pressures are expressed as pressure drop over the riser (Δp). In general case, one would see the same pattern of pressure fluctuations and the formula above also applies to pressure at its absolute value.

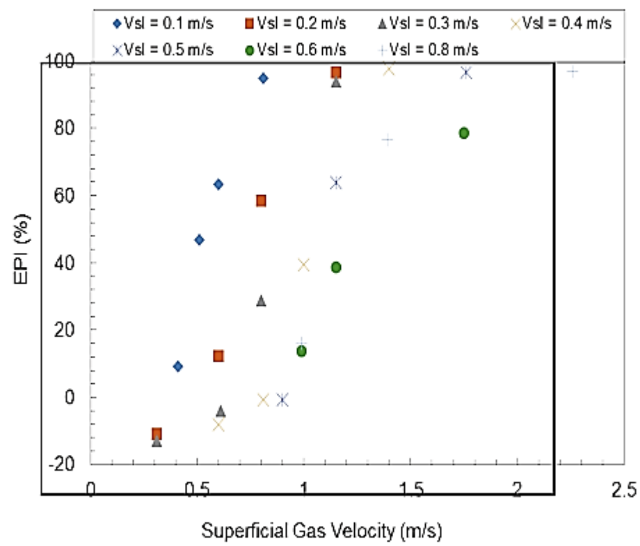


Figure 3.17. EPI of gas and liquid superficial velocities [58].

The graph above shows that EPI has strong correlation with both gas and liquid superficial velocities. It may be inferred that decrease in superficial liquid velocity causes increase of slug elimination efficiency, whereas reduction of superficial gas velocity has an opposite effect. Moreover, Figure 3.15 shows that superficial gas velocity is the crucial factor for the efficiency of the technique since the contraction of severe slugging region occurs along the gas velocity axis. Injection of additional amounts of gas would move the operational point to the right, see Figure 3.15, when the resulting amount of injected gas would be sufficiently lower compared to conventional gas lift.

Based on the arguments above we can come to a conclusion that surfactant injection efficiency could be enhanced by injection of a relatively small amount of gas. To address to that point, a series of following trials on feasibility of combination of foaming and gas injection was run by Sarica et al. [36]. The experiment was conducted on the same facility, using air and water with rates close to the previous investigation and the same surfactant. In addition, possibility to inject gas at the riser base was allowed. In total, 30 tests (combinations of gas and liquid superficial velocities before injection) were run, each consisting of several steps with different gas and surfactant injection rates.

The results of the experiment are plotted on Figure 3.18. Surfactant injection rate was always above critical concentration.

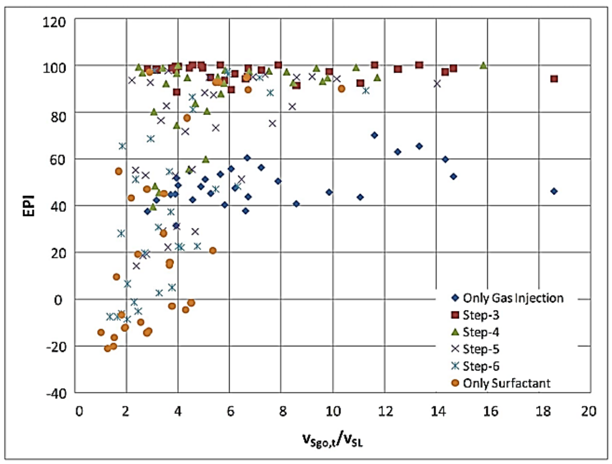


Figure 3.18. Severe slugging elimination with foam and gas injection [36].

Each test consisted of 7 steps:

Step 1 – no injection (used to determine the base line);

Step 2 (Only Gas Injection) – maximum gas injection applied;

Step 3 – gas injection with surfactant;

Step 4-6 – gradual reduction of gas injection.

Step 7 (Only Surfactant) – gas injection stopped.

Table 3.1. Severe slugging elimination with foam and gas injection [36].

Test	$V_{Sg(m/s)}$	$V_{SL\ (m/s)}$	$V_{Sgo,inj}$ (m/s)							Surfactant rate (kg/min)	EPI (%)					
			Step 1	Step 2	Step 3	Step 4	Step 5	Step 6	Step 7		Step#2	Step#3	Step#4	Step#5	Step#6	Step#7
1	1.16	0.11	0	0.92	0.92	0.62	0.42	0.1	0	0.16	46	94	100	92	89	90
2	1.40	0.21	0	0.92	0.92	0.62	0.40	0.1	0	0.16	44	92	93	95	95	95
3	1.76	0.32	0	0.95	0.95	0.61	0.42	0.1	0	0.17	41	91	98	95	98	93
4	1.73	0.4	0	0.94	0.94	0.6	0.43	0.11	0	0.17	38	94	92	87	86	-4
5	2.24	0.51	0	0.95	0.95	0.6	0.42	0.11	0	0.17	48	96	95	95	81	78
6	0.32	0.11	0	0.96	0.96	0.62	0.40	0.1	0	0.84	70	100	93	51	5	-13
7	0.42	0.11	0	0.96	0.96	0.62	0.43	0.11	0	0.84	63	98	99	75	23	-3
8	0.51	0.11	0	1	1	0.60	0.44	0.11	0	0.84	66	100	95	82	47	-2
9	0.60	0.11	0	1.0	1.0	0.62	0.43	0.11	0	0.23	60	97	99	95	48	21
10	0.81	0.12	0	0.96	0.96	0.6	0.41	0.1	0	0.17	52	99	95	94	88	90
11	0.32	0.21	0	0.96	0.96	0.62	0.44	0.11	0	0.84	56	89	56	22	-8	-20
12	0.41	0.21	0	0.99	0.99	0.65	0.41	0.1	0	0.84	61	95	60	31	-5	-12
13	0.60	0.22	0	0.95	0.95	0.62	0.4	0.1	0	0.84	56	98	88	29	3	-14
14	0.81	0.22	0	0.92	0.92	0.68	0.39	0.1	0	0.24	50	100	97	73	22	15
15	1.15	0.21	0	0.92	0.92	0.57	0.4	0.1	0	0.16	46	97	97	96	96	93
16	0.32	0.31	0	0.90	0.90	0.62	0.42	0.1	0	0.84	52	89	40	14	-7	-14
17	0.40	0.31	0	0.98	0.98	0.62	0.46	0.12	0	0.84	55	99	46	19	-7	-21
18	0.61	0.31	0	0.98	0.98	0.63	0.46	0.12	0	0.84	51	99	74	29	-1	-12
19	0.81	0.32	0	0.97	0.97	0.57	0.42	0.21	0	0.81	53	100	95	53	31	-10
20	1.15	0.31	0	0.94	0.94	0.66	0.42	0.1	0	0.23	44	99	98	88	23	16
21	0.61	0.39	0	0.97	0.97	0.62	0.43	0.11	0	0.81	49	99	49	19	-6	-16
22	1	0.41	0	1	1	0.62	0.45	0.11	0	0.84	48	100	97	83	20	19
23	1.4	0.4	0	0.96	0.96	0.67	0.43	0.11	0	0.17	40	93	80	55	38	28
24	0.91	0.5	0	0.94	0.94	0.62	0.45	0.11	0	0.84	45	99	80	53	7	-7
25	1.16	0.53	0	0.91	0.91	0.65	0.38	0.1	0	0.48	45	99	99	93	51	43
26	1.77	0.51	0	0.92	0.92	0.62	0.42	0.1	0	0.17	45	95	84	72	55	45
27	1.0	0.61	0	0.95	0.95	0.62	0.45	0.11	0	0.84	42	98	97	55	28	10
28	1.75	0.6	0	1	1	0.66	0.43	0.11	0	0.52	42	100	100	98	98	97
29	1.39	0.81	0	0.90	0.90	0.62	0.4	0.1	0	0.52	38	98	99	94	66	55
30	2.25	0.8	0	0.90	0.90	0.58	0.42	0.11	0	0.16	31	98	92	76	69	47

Note: color coding of the EPI columns:

Blue – severe slugging eliminated (EPI>0.5);
Yellow – partly severe slugging elimination (0<EPI<0.5);
Red – detrimental effect of the pressure fluctuations (EPI<0).

As it shown in Table 3.1 and figures above, addition of the surfactant to the flow characterized by slugging with pressure fluctuations of about a half of the base case reduced pressure fluctuations almost to zero ($EPI = 1$). Further reduction of the gas injection rate showed results similar to those in Sarica et al. [58].

Along with the flow conditioning effect, injection of the surfactant decreased backpressure on the pipeline. However, the measure for backpressure reduction (BPR) provided by Sarica et al. [36] takes into account the maximum value of the pressure fluctuations instead of its mean. This choice seems questionable.

To the downsides of the technique undesired foaming inside of the separator may be attributed. To mitigate foaming an additional chemical must be added. There is also some evidence that addition of surfactants and consequent addition of antifoaming agents caused flow assurance problems related to increased proneness to wax deposition [59].

Criticism.

Some of the tests provided by Sarica [36] give opposite results despite their similarity. Let us take a closer look at Tests 3 and 4 in Table 3.1. Both have comparable initial gas and liquid superficial velocities and almost identical gas and surfactant injection rates. Both show comparable results in Steps 1-6. However, in the case of slugging mitigation with foaming only, the two tests show completely different results. Test 3 has almost no pressure fluctuations, meaning that no severe slugging is present, whereas the situation in Test 4 got worse compared to the base case. Figure 3.15 can be used to determine the predicted flow regime. Both of the cases lay in transitional severe slugging region and slugging should have been completely eliminated by foaming. This could mean either a misprint, a mistake in the experiment, or incomplete understanding of the process. The authors provide no comments on the result under discussion.

The choice of the criteria to measure the effect on the backpressure reduction also seems to be questionable. It takes into account the change between the maximum exerted pressures during fluctuations instead of its mean values. The mean value of the pressure seems to be a better way to describe the backpressure created by slugging.

3.2.6. Multiphase riser base lift

The multiphase riser base lift concept proposed by Johal et al. [60] was aimed to overcome the main drawbacks of the standard gas lift method, especially for deep water installations. As it was described above, to those one can attribute the necessity for high pressure gas supply and Joule-Thompson cooling effect, resulting in increased possibility of hydrate and wax deposition. The last is especially prominent for deep waters due to higher hydrostatic pressure difference, therefore higher expansion of the rising gas inside of the riser.

The authors proposed to use a cross over path between the existing flowlines to reroute some of the produced multiphase flow near the riser base. Presumably, it would allow to establish a desired flow regime without supplying external energy to the system. It was highlighted that the system is especially effective for start-up of operations, which may be considered to be the worst situation, and could be more attractive in economic terms compared to the gas lift system [60].

The system requires a high capacity multiphase line available at the site and therefore is limited only for specific cases [19].

4. Mitigation methods modelling

4.1. Case study construction

For the purpose of comparison between different mitigation methods and further evaluation of the proposed method, a case study needs to be chosen. However, the author was unable to find an example from the industry, which would fit all of the criteria. For that reason, an artificial case study was modelled. This part describes the model and the main principles that were used to construct it.

4.1.1. The pipeline-riser system

The overall pipeline geometry is presented in Figure 4.1. and summarized in Table 4.1. It is comprised of 5.9 km pipeline section followed by 500 meters high vertical riser. At the point of 1 km downstream of the wellhead a 250 meters long pipeline section with negative inclination of 2° is situated. It is followed by an identical section with positive inclination, bringing the pipeline back to its zero level. The two sections are thought to simulate some of the hilly-terrain effects. The riser outlet is followed by 50 meters horizontal section according to the manual recommendations [61]. Moreover, some evidence in literature was found that absence of the horizontal section prior to separator may result in unphysical results, primarily due to excess of fallback liquid [62].

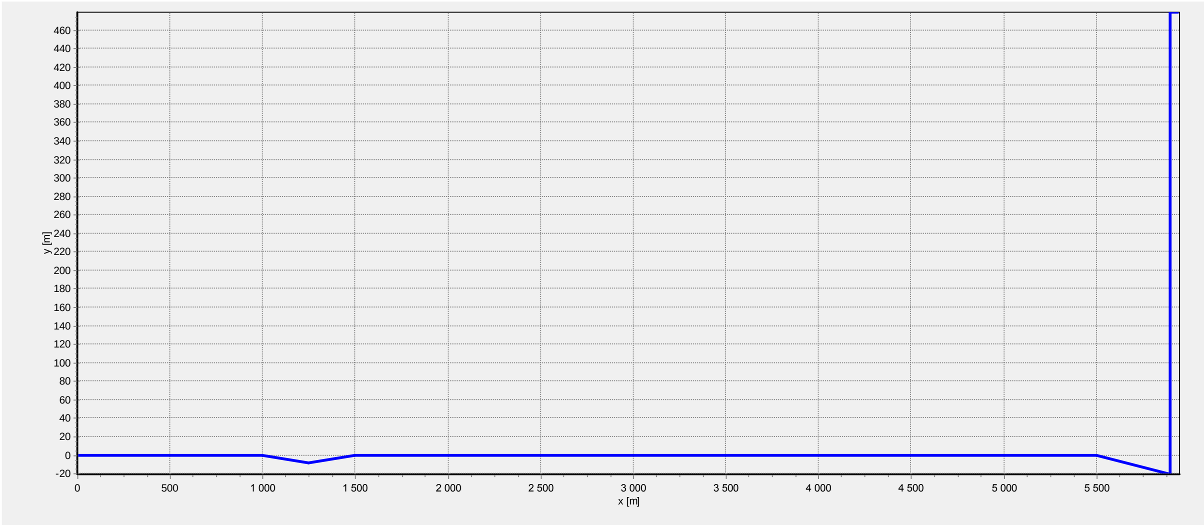


Figure 4.1. The study case pipeline geometry generated in OLGA.

NB! The axises have different scale.

Table 4.1. Pipeline geometry

Pipeline section	Length, [m]	Inclination, [grad]	Inner diameter, [inch]	Roughness, [mm]
1	*1000*	*0*		
2	*250*	*-2*		
3	*250*	*2*		
4	*4000*	*0*	*6*	*0,03*
5	*400*	*-3*		
Riser	*500*	*90*		
Horizontal pipe	*50*	*0*		

4.1.2. The fluid model

As a descriptive model for the fluid properties the Blackoil module within OLGA was chosen. The Blackoil model is appropriate for computations when little information is available about the production fluid. Oil, gas and water specific gravities, and gas-oil ration (GOR) are the only parameters required for the model [61], making it a perfect choice for the limitations imposed by the scope of the thesis.

A PVT table is not required when Blackoil model is used. Instead, at each instance of time and at every position along the pipe the module uses predetermined correlations to calculate equilibrium between oil and gas, assuming that gas is dissolvable in oil, but no oil can exist in the gas as gas phase. Based on that, physical properties of fluids are calculated.

Glasø correlations were chosen to model gas-oil dissolvability. It is advised to use for oils of North Sea region of API greater than 15. The bubble-point pressure may be acquired implicitly from the equations [61]:

$$\begin{cases} \log\left(\frac{P_b}{C_{Pb}}\right) = 1.7669 + 1.7447\log(A) - 0.30218[\log(A)]^2, \\ \qquad A = (GOR/\gamma_G)^{0.816}(T^{0.172}/API^{0.989}). \end{cases} \tag{4.1}$$

For given pressure and temperature combination the equilibrium $RSGO$ may be calculated as follows [61]:

$$RSGO = C_{RSGO}\gamma_G\left[API^{0.989}/T^{0.172} \cdot 10^{\frac{0.5\log(P)-1.329}{2.6256}}\right]^{-0.816}. \tag{4.2}$$

It is important to notice that the model units are psia and °F.

The oil density is primarily a function of pressure, temperature and $RSGO$. At pressures higher than bubble-point pressure, $RSGO = GOR$, and the density follows regular isothermal compressibility rule:

$$\rho_O = \rho_{Ob}\exp\big(c_O(P - P_b)\big), \tag{4.3}$$

where c_O is the compressibility coefficient and may be calculated as [61]:

$$c_O = 10^{-5}(-1433 + 5RSGO + 17.2T - 1180\gamma_G + 12.6API)/P. \tag{4.4}$$

For pressures below the bubble-point, the oil density is computed in the following way [61]:

$$\rho_O = \frac{62.4\gamma_0 + 0.0136\gamma_G RSGO}{0.9759 + 1.2 \cdot 10^{-4}[RSGO(\gamma_G/\gamma_0)^{0.5} + 1.25T]^{1.2}}. \tag{4.5}$$

The correlation above may be also used to compute oil density at the bubble-point by substitution $RSGO$ with GOR.

The model units for pressure, temperature and FVF are psia, °F, and bbl/STB, respectively.

The gas density is calculated with assumption of the real gas model, where compressibility factor z is precalculated and stored within the software. Phase viscosities and interphase surface tensions are calculated in the same fashion by using the model correlations.

The oil is assumed to have API of 30 and the produced gas specific gravity of 0.64.

Due to variety of flowrates modeled, the flow is assumed to be adiabatic. This was done to get results independent of heat exchange with the environment, which in turn is highly dependent on the flow regime, flow composition, and flowrate. Thus, by assuming adiabatic flow, heat exchange was excluded from the model in hope to get more consistent results. To compensate for that fact, the reservoir temperature was intentionally underestimated. A similar approach was used in [63] where isothermal approximation was used. The authors advocate that severe slugging effect is highly dominated by gravity forces and temperature variations typical for subsea pipeline systems do not affect the results significantly.

Using the Blackoil model described above, the relation between bubble-point pressure and GOR may be calculated and used to choose reasonable estimations of reservoir conditions, see Figure 4.2.

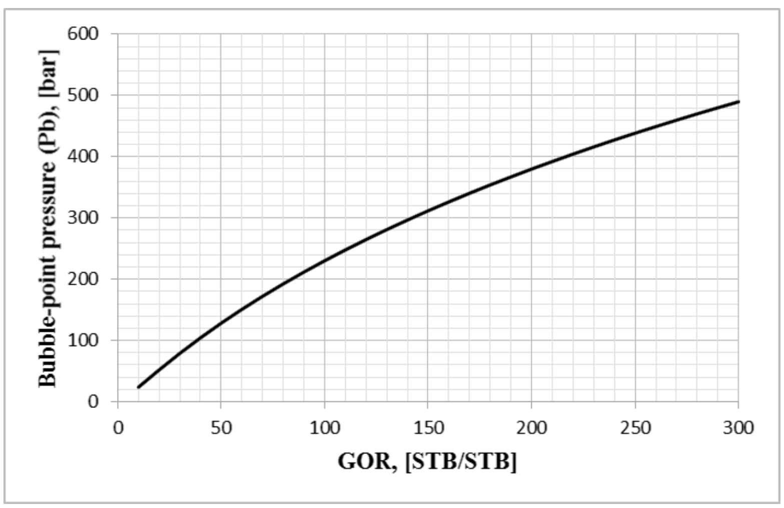


Figure 4.2. Bubble-point pressure dependence on initial GOR at 50°C (assumed reservoir temperature).

4.1.3. The well and wellbore

Production volumes are assumed to be conducted through the 4½ in. production tubing, with inner diameter of 3.958 in. [64]. Wall absolute roughness is set to 0.06 mm. The wellbore of total length of 3 kilometers deviates from vertical line at 2 kilometers depth. The last kilometer is drilled with 45° inclination.

The reservoir is assumed to be a saturated oil reservoir; therefore, Vogel's IPR equation is used to model the oil-well performance. Vogel's equation reads as follows [61]:

$$Q_0 = Q_{0,max}\left[1 - 0.2\left(\frac{P_{wf}}{P_{res}}\right) - 0.8\left(\frac{P_{wf}}{P_{res}}\right)^2\right], \tag{4.6}$$

where $Q_{0,max}$ – maximum oil flow at flowing bottom hole pressure equal to zero.

Given the combination of true vertical depth from the mean sea level and desired GOR of 150, a combination of reasonable reservoir pressure, flowing bottom hole pressure, and bubble-point pressure were chosen. Thus, the reservoir exists at 315 bar of absolute pressure at 50 °C in saturated condition. At flowrate of about 1500 Sm^3/d the flowing bottom hole pressure fluctuates around 270 bar. Maximum oil flowrate was assumed to be 6800 Sm^3/d.

On importance of wellbore inclusion.

The system volume is one of the most important factors for severe slugging process due to gas compressibility and its direct effect of the force balance described above. Meng and Zhang [65] report erroneous flow assurance evaluation due to absence of a wellbore in the model studied. Other researchers report severe slugging induced after connection of an additional well to a flowline. Thus, despite increased flowrates, additional volume created conditions favorable for terrain slugging. The model studied in this project has total volumes of pipeline and wellbore of the same magnitude. Therefore, inclusion of a wellbore was considered to be necessary.

4.1.4. The topside facilities

As was briefly mentioned before, a riser model should include a horizontal section at the outlet. Thus, the vertical riser part translates into 50 meters horizontal part and terminates into a constant pressure node, which simulates operations of a separator. The pressure node is assumed to operate at constant pressure and temperature. Exclusion of a separator simplifies the model. Moreover, evidence was found that inclusion of a separator in OLGA models does not affect severe slugging significantly. Thus, Ioannou et al. [66] reports that addition of a separator into the model just slightly changed severe slugging behavior, affecting slugging frequency, but leaving pressure fluctuations value unchanged, see Figure 4.3.

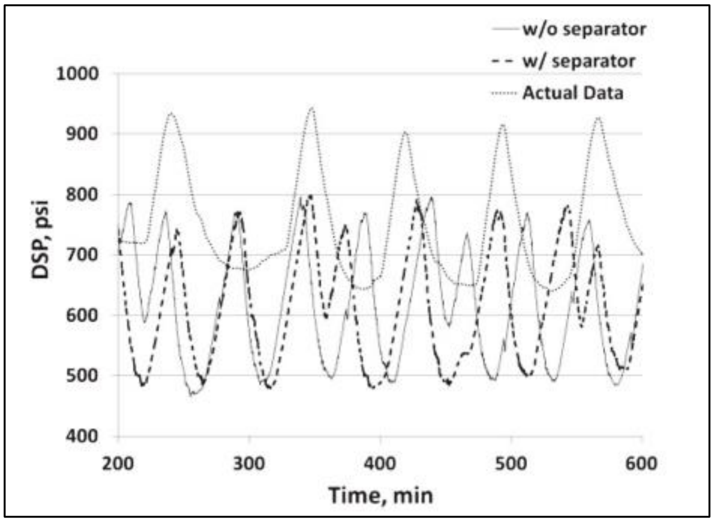


Figure 4.3. Pressure readings after the production choke for cases with and without separator [66].

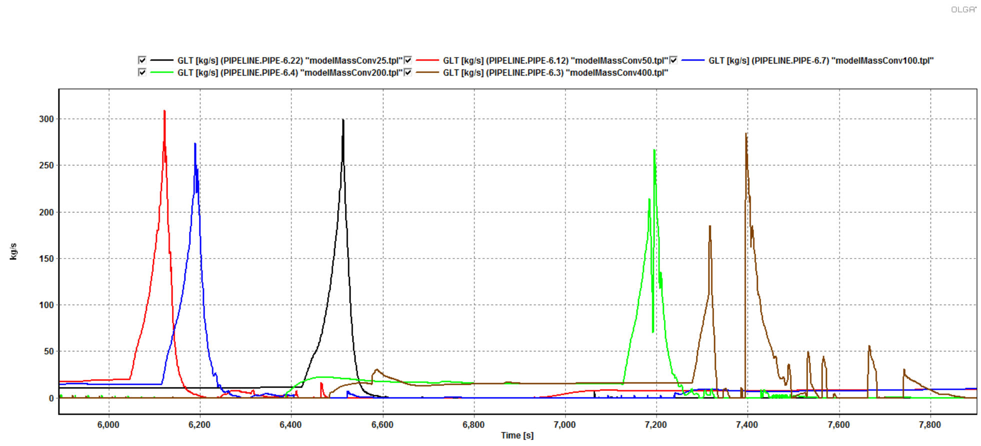
Simulations were done in OLGA7 with HD module. Delay constant was kept at 150. Introduction of a separator affects slug frequency; pressure fluctuation remains the same.

The constant pressure node serves as a boundary condition, allowing inflow of all phase and outflow of only gas phase. It is made possible by setting `GASFRACEQ` node key (gas fraction equilibrium) to 1, whereas `WATERFRACEQ` and `OILFRACEQ` to 0.

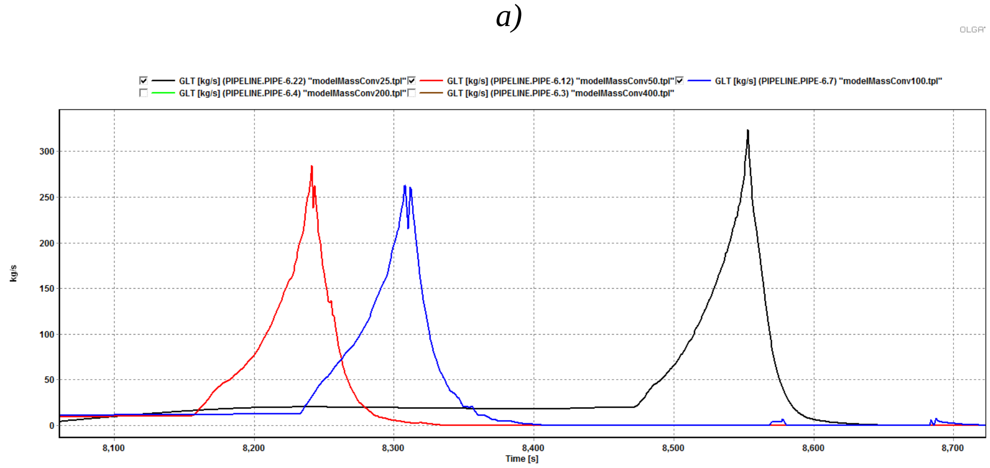
The separator pressure was set to 50 bar at normal operating conditions.

4.1.5. Solution convergence analysis

The base case assumed discretization method over the pipeline length with the rule of at least 2 sections per pipe and pipeline sections less than 100 meters. To investigate on the solution convergence with respect to pipeline section length, sensitivity analysis was conducted with discretization limits of 25, 50, 100, 200, and 400 meters. For all of the cases the well model was substituted with constant mass flowrate source with gas flowrate of 0.84 kg/s and liquid flowrate of 10.24 kg/s. At given mass flowrates transitional to severe slugging regime with stable slug occurrence frequency was observed during preliminary studies. Readings of the total liquid mass flowrate at the riser outlet were used to characterize if riser-induced slugging was properly simulated. Figure 4.4. (a) plots third slugs occurred in the system for each of the cases.



a)



b)

Figure 4.4. Riser outlet liquid mass flowrate for different discretization resolutions.

Note: color coding of discretization rules (in addition to minimum number of sections per pipe equal 2):

Black – MAX pipeline section length 25m;
Red – MAX pipeline section length 50m;
Blue – MAX pipeline section length 100m;
Green – MAX pipeline section length 200m;
Brown – MAX pipeline section length 400m.

It may be inferred from the figure that resolutions of 200 and 400 meters provide numerical solutions inconsistent with physical nature of the phenomenon. In addition to that, slug frequency for the latter cases is consistently greater than for those with finer resolution. Closer look at the fine resolution cases (Figure 4.4.(b)) shows that 100 and 50 discretizations have flaws in solutions similar to those of rough discretization. The effect disappears if pipeline section equal or less than 25 meters.

The drawback of having a fine meshing resolution is increased computational time. As a rule of thumb, reduction of each section length by 2 translates into 4 times longer simulation time [61]. To maintain computational time within acceptable margins and benefit from a finer discretization, progressive discretization was used. Thus, pipeline section lengths vary between 400 meters for long horizontal sections and 25 meters for the dip, pipeline-riser junction, and riser. The ratio between adjacent pipeline sections was kept at about 1.5.

Resulting pipeline discretization encounters 71 pipeline sections, compared to 261 for the meshing rule limiting sections length to 25 meters.

To test how well the refined discretization stands to that of 25 meters limiting factor generated by OLGA standard tool, comparative simulation was run over 3 hours with the same flowrates as described before. The result is presented on Figure 4.5. As one can see, the results are comparable to a high degree. At the same time, significant amount of computational time was saved due to reduced number of sections.

Investigation on the influence of the wellbore pipeline section length was conducted where discretization length was reduced from basic 200 meters to 50 meters. Simulation showed almost identical results, increasing computational time. Therefore, the wellbore section length left to be 200 meters.

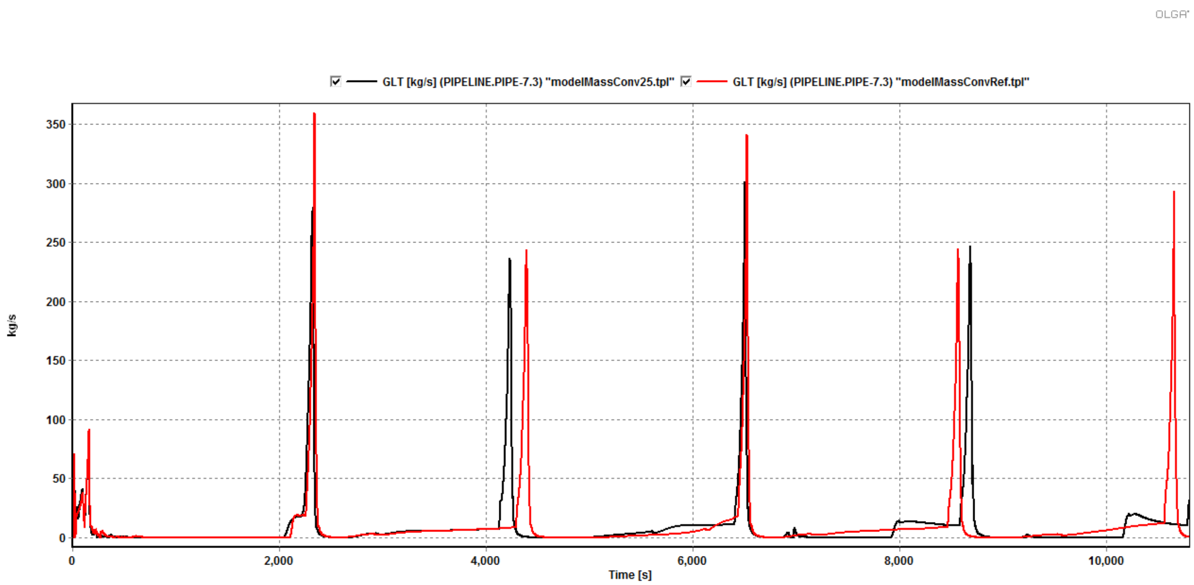


Figure 4.5. Riser outlet liquid mass flowrate for progressive and fixed discretization.

Note: color coding of discretization rules (in addition to minimum number of sections per pipe equal 2):

Black – MAX pipeline section length 25m;
Red – progressive discretization (section length=[25;400] meters);

4.1.6. Flow regime map identification

Despite the fact that severe slugging occurrence criteria are beneficial for understanding of the phenomenon mechanics, it is troublesome to use the same tool for real pipeline riser systems. Thus, the common metrics used for severe slugging identification are superficial gas and liquid velocities. In the laboratory conditions they may be accurately measured and controlled. Water and air at PVT conditions close to standard are used as test fluids. Despite the fact that superficial velocities may be the best choice in severe slugging effect description in laboratory conditions, they are less suited for the task for a number of reasons:

1. The system of concern experiences great variety of pressure conditions, primarily affecting the gaseous phase compressibility, hence its superficial velocity.
2. Unlike the behavior of gaseous and liquid phases of typical test fluids, gas is highly dissolvable in oil meaning possibility of mass transfer between the phases.

An in-depth description of mass transfer effect on severe slugging was covered in Section 2.3.

For that reasons GOR and liquid production volume flowrate and standard conditions were chosen to describe severe slugging occurrence in the system. To make the results comparable with those from other sources, flow regime maps were also expressed in terms of volumetric liquid and gas flowrates at standard conditions.

For the purpose of severe slugging operational conditions identification the model of producing well was substituted with a constant mass flow source. The flowrate was tuned by changing `FEEDSTDFLOW` key, set to change volumetric flowrate at standard conditions of the oil phase (`PHASE=OIL` key) of the Blackoil modelled fluid.

A series of parametric studies was conducted with each of the simulation lasting for 8 hours, with steady state pre-processor turned on (`STEADYSTATE=ON` key). First 2 hours of each simulation were rejected due to oscillatory processes resulted from transition from steady state pre-processor state to transient simulation. Remaining 6 hours were analyzed and flow regimes identified and sorted into three categories:

1. Severe slugging of type 1;
2. Severe slugging of type 2;
3. Severe slugging of type 3;
4. Stable flow (a flow of relatively high frequency, no liquid production starvation phase).

The classification is consistent with the aim of the thesis, namely – severe slugging identification and elimination. Transition from severe slugging state to slug flow was considered to be sufficient. Therefore, no distinction between bubble and slug flow of relatively short slugs were made and those two were combined on the flow map.

The flow regimes were primarily identified by readings of two graphs: mass liquid flowrate (`GLT` trend variable) at the pressure (`PT` trend variable) at the riser base. Additionally, surge liquid volume (`LSLEXP` trend variable specified at the riser outlet) and superficial gas velocity at standard conditions at the riser base (`USGST` trend variable) were used for some flows.

`GLT` readings allowed verifying severe slugging presence and SS1 identification due to characteristic feature of slow liquid production phase. SS2 and SS3 were differed by superficial gas velocity readings at the riser base; for SS3 there is gas penetration at the riser base almost at all times.

The resulting flow map outlining severe slugging occurrence region is presented on Figure 4.6. The simulation summary is presented in form of tables in Appendix A.

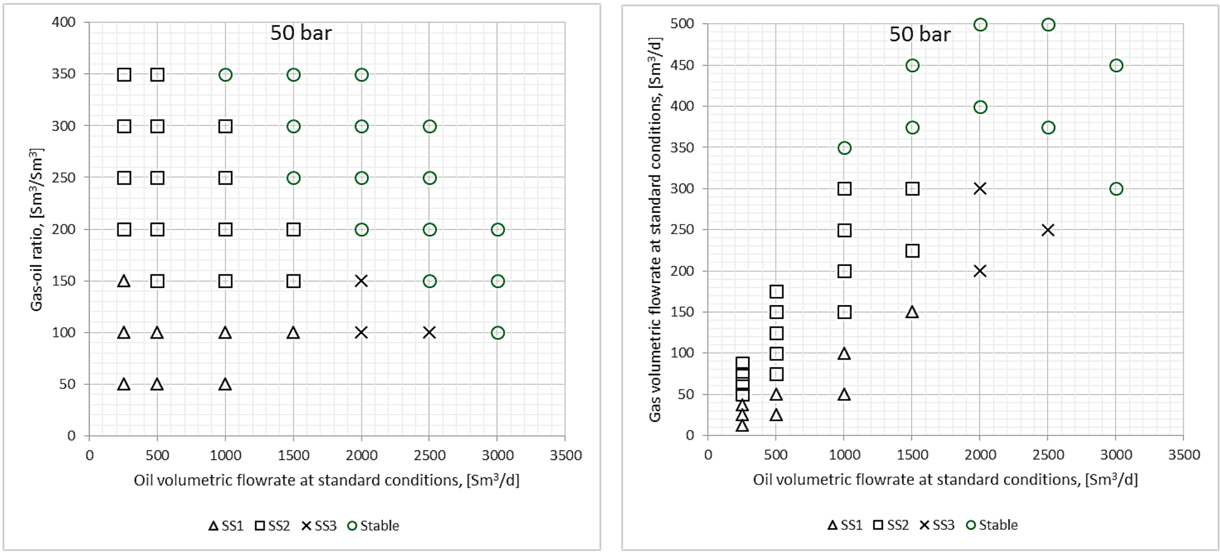


Figure 4.6. Severe slugging occurrence flow regime map for the separator pressure of 50 bara.

The green circles represent non-severe-slugging flow. The triangles, squares and crests stand for severe slugging of type 1, 2 and 3, respectively.

Some peculiar results were observed in cases of low GOR values. Instead of severe slugging with increasing period, starting from oil volumetric flowrate of 1500 Sm^3/d at GOR equal to 50 steady flow regime with sudden release of fast liquid surges was present. This flow regimes were not included on the flow regime map. The same phenomenon was observed for 40 bara pressure in Section 5.4.

4.2. Topside choking

To model the effect of topside choking, the basic study case model was complimented with a choke valve installed in the middle of the horizontal section before the separator. The following settings were chosen based on the modelling purpose and the user manual recommendations:

```
MODEL = HYDROVALVE;
EQUILIBRIUMMODEL = HENRYFAUSKE;
THERMALPHASEEQ = NO;
SLIPMODEL = NOSLIP;
```

```
RECOVERY = YES;
DIAMETER = 0.1524 [m];
CD = 0.84;
CR = 1.
```

The stepwise closing of the choke valve from fully opened position to 0.05 opening was performed similar to that in [15]. Each step lasted for 3 hours, resulting in 21 hour simulation. Time required to shift between the choke positions was set to 0.1 hour and relative opening was changed linearly. The result is presented on Figure 4.7.

From here the relative choke opening is defined as a fraction between open cross sectional area of partially actuated choke to its fully open inner cross sectional area, which for our case is equal to inner cross sectional area of the pipeline.

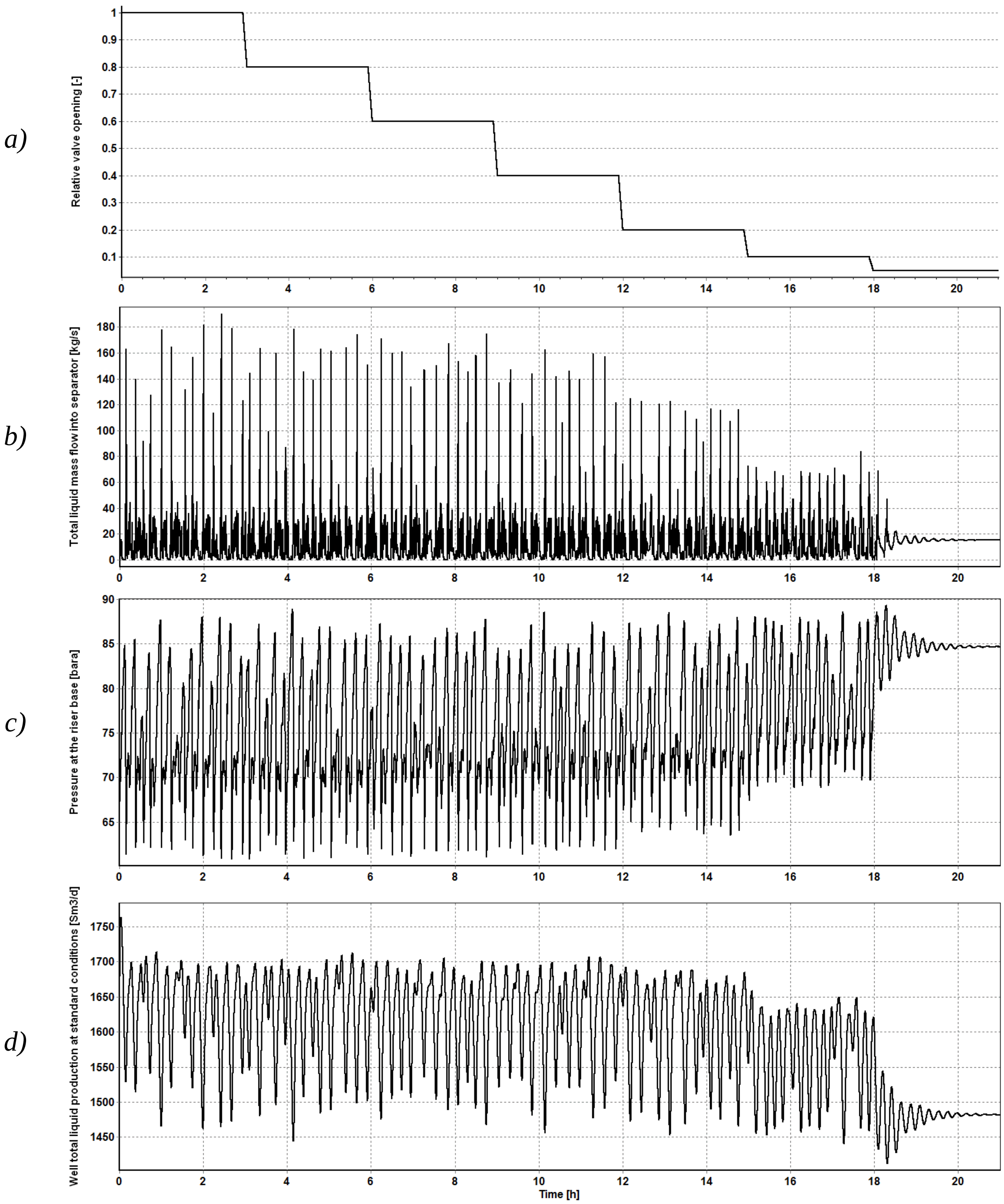


Figure 4.7. Severe slugging mitigation with topside choking of relative valve opening ranging from 1 to 0.05.

The graphs shows change in liquid mass flowrate into separator (b), pressure fluctuations at the riser base (c), and total liquid production (d) with respect to change in relative choke opening (a). Graphs are generated in OLGA.

The graphs show that the severe slugging effect is alleviated with increase of choking. Thus, at relative opening of 0.1 a relatively mild slug flow observed, whereas at 0.05 opening the flow is fully stabilized. However, the effect comes at cost of total production rate. The initial production rate, fluctuating between approximately 1460 and 1700 Sm^3/d and having arithmetic average of 1627 Sm^3/d, is reduced to 1470 Sm^3/d in the case of fully stabilized flow due to increased backpressure on the producing well.

A series of consecutive studies with shorter step of closing ratio was conducted to give a closer look at the flow behavior at low relative choke openings, which summarized on Figure 4.8. As before, the more aggressive the choking effect was, the better flowrate was stabilized at the cost of lowered production and increased back pressures on the production wells.

The study intentionally does not provide conclusion on which closing ratio must be preferred. The choice is case sensitive and highly depends on the capabilities of the process facilities available. Thus, if the first stage separator has the maximum liquid throughput of 3400 Sm^3/d (approximately double the maximum well production in the initial case) and the system may tolerate surges of up to 3 Sm^3, the optimal choice of relative choke opening would be 0.12 as it is shown on Figure 4.9.

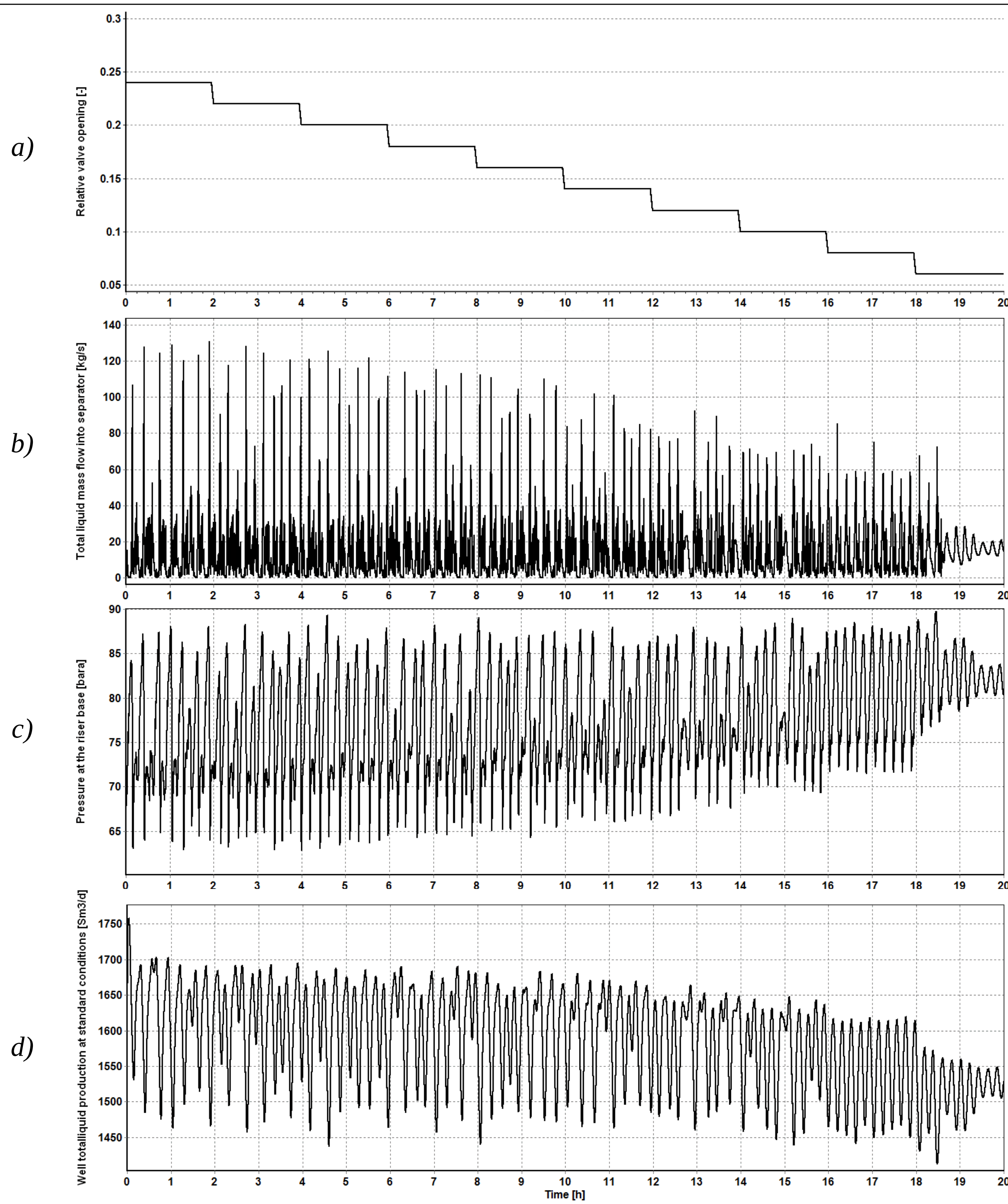


Figure 4.8. Severe slugging mitigation with topside choking of relative valve opening ranging from 0.24 to 0.06.

The graphs shows change in liquid mass flowrate into separator (b), pressure fluctuations at the riser base (c), and total liquid production (d) with respect to change in relative choke opening (a). Graphs are generated in OLGA.

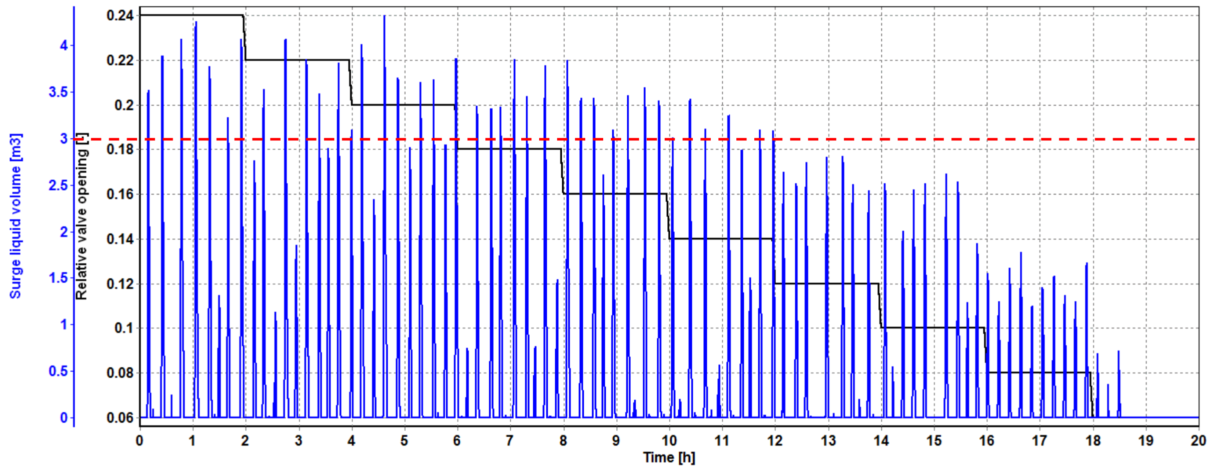


Figure 4.9. Surge liquid volume of the system with maximum drainage rate of 3400 Sm^3/d with respect to relative valve opening.

The graph shows approximate solution to the problem of optimal choking choice provided surge liquid capability of 3 m^3. Relative valve opening of 0.12 satisfies conditions of no carry-over. Graph is generated in OLGA.

The effect of topside choking with respect to pressure fluctuations at the riser base may be neatly summarized under a common bifurcation diagram depicted on Figure 4.10. The bifurcated part represent maximum and minimum values of pressure gauge reading with the average value between plotted as a dash line. At relative opening between 0.06 and 0.07 the flow is fully stabilized and the two curves merge into one.

Simulation results that allowed to plot the bifurcation diagram may be found in Appendix E.

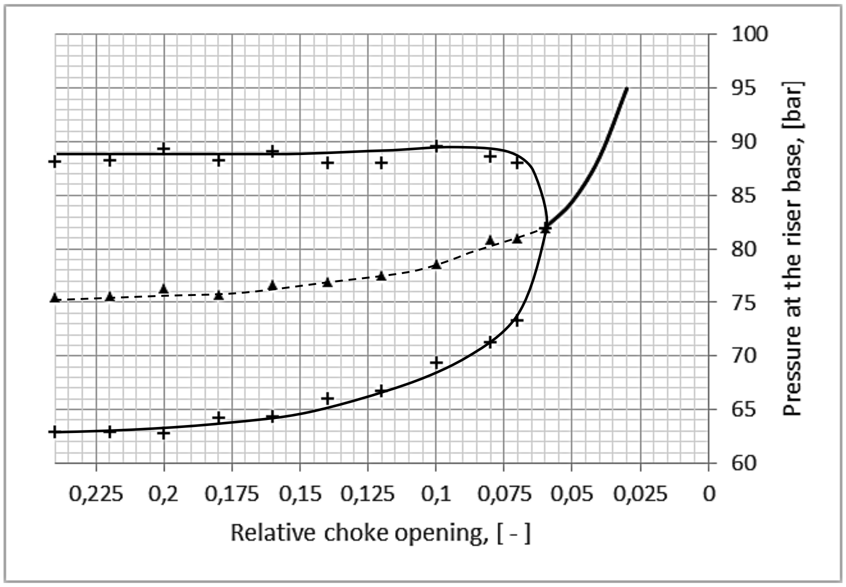


Figure 4.10. Topside choking bifurcation diagram

The graph shows absolute pressure fluctuations value dependence on the relative choke opening. The upper limb represents pressure maximum, whereas the lower – minimum. At relative opening of ~0.065 the flow is fully stabilized and the two lines merge into one.

4.3. Riser-base gas injection

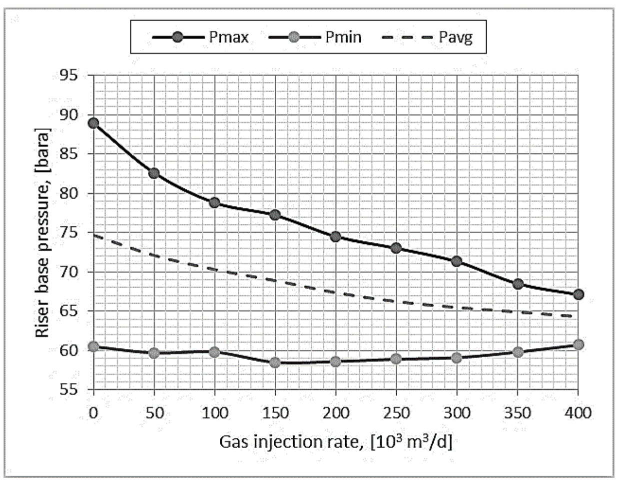
To simulate severe slugging elimination with gas injection at the riser base, the base model was altered by adding constant mass flowrate source at the riser base. The gas source was specified to have constant injection temperature of 40 °C, using the Blackoil feed with `GASFRACEQ` node key (gas fraction equilibrium) set to 1, whereas `WATERFRACEQ` and `OILFRACEQ` to 0, allowing only gas injection with specified flowrate.

Based on the work of other authors, flowrates of the same order as the amount of gas produced were expected to be necessary to stabilize the flow [6, 65]. The average gas flowrate at standard conditions was estimated at approximately 243,000 Sm^3/d. A gas injection rates of 50, 100, 150, 200, 250, 300, 350 and 400 thousand Sm^3/d were investigated and compared to the base case.

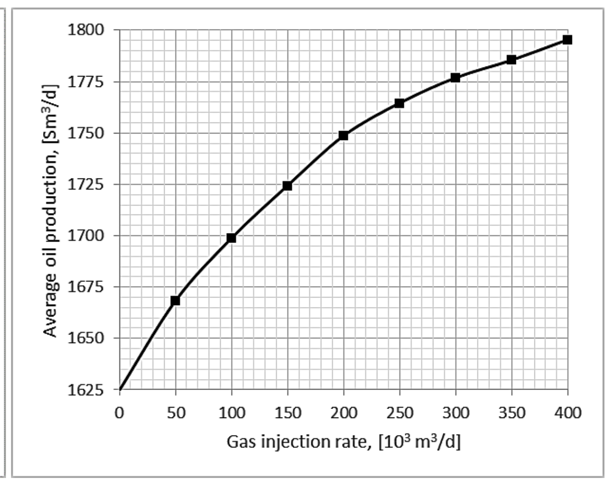
The simulations results are summarized in Table 4.2. and Figure 4.11.

Table 4.2. Gas injection effect on severe slugging

Gas injection rate, [10^3·Sm^3/d]	Maximum liquid flowrate observed, [kg/s]	Slugging cycle average period, [h]	Maximum pressure at riser base, [bara]	Minimum pressure at riser base, [bara]	Pressure amplitude, [bar]	Average oil production, [Sm^3/d]	Average pressure at riser base, [bara]	Maximum liquid surge volume, [m^3]
0	*187*	*0,26*	*88,9*	*60,5*	*28,4*	*1624,8*	*74,6*	*5,1*
50	*128*	*0,19*	*82,6*	*59,7*	*22,9*	*1668,4*	*72,1*	*3,3*
100	*90*	*0,14*	*78,8*	*59,8*	*19*	*1698,9*	*70,3*	*2,2*
150	*77*	*0,12*	*77,2*	*58,5*	*18,7*	*1724,5*	*68,9*	*2*
200	*66*	*0,12*	*74,5*	*58,6*	*15,9*	*1748,7*	*67,3*	*1,2*
250	*58*	*0,1*	*73*	*58,9*	*14,1*	*1764,6*	*66,2*	*1*
300	*52*	*0,1*	*71,3*	*59,1*	*12,2*	*1776,9*	*65,5*	*0,5*
350	*37*	*0,08*	*68,5*	*59,8*	*8,7*	*1785,5*	*64,9*	*0*
400	*30*	*0,07*	*67,1*	*60,7*	*6,4*	*1795,5*	*64,3*	*0*



a)



b)

Figure 4.11. Pressure fluctuations (a) and average production rate (b) dependence on the gas injection rate.

The left plot shows pressure fluctuations decrease with increased amount of injected gas. As the result of gas lift the average production increases (right plot).

It may be inferred from Figure 4.11.(a) that increase of the gas injection flowrate significantly reduces pressure fluctuations at the riser base, at the same time reducing the average pressure. As the result, the backpressure imposed on the producing well also decreases, boosting oil production rate as it is shown on Figure 4.11.(b).

4.4. Diameter reduction

Riser diameter reduction as a severe slugging alleviation technique, proposed by Yocum [18] and calculated as a mitigation method in [65], was investigated on its effectiveness. To do so, the case study model geometry was altered so that cross sectional area was equal to 0.9, 0.8 and 0.7 of the base case as it is shown in Table 4.3. These values were assumed to be reasonable with respect to other operational considerations such as pigging.

Table 4.3. Riser diameter reduction effect on severe slugging

	Area ratio to the base case	Riser inner diameter, [m]	Average cycle period, [h]	Period standard deviation, [h]	Maximum observed liquid mass flowrate, [kg/s]
Base case	*1*	*0.1524*	*0.227*	*0.024*	*187*
Case 1	*0.9*	*0.1446*	*0.259*	*0.044*	*175*
Case 2	*0.8*	*0.1363*	*0.303*	*0.038*	*184*
Case 3	*0.7*	*0.1275*	*0.372*	*0.071*	*177*

Diameter reduction had almost no effect on the system behavior. Liquid surge volumes, maximum liquid flowrates and total production remained at approximately the same level. One notable change occurred to slugging period. In contrast to results in [65], riser diameter reduction reduced severe slugging frequency. A possible explanation to that effect could be increased superficial gas velocities resulting in improved liquid sweep out during blowdown phase. However, no direct evidence favoring the explanation was found.

5. Severe slugging mitigation by backpressure reduction with a Surface jet pump

5.1. Proposal description

The majority of the severe slugging elimination techniques are based on the principle of increase of backpressure on the pipeline-riser system, mainly by choking of the flow stream from topside either in an automated or manual mode. The approach was thoroughly described in the previous chapters. However, it may seem unreasonable to induce additional backpressure load on already ageing production wells. Even though it may ultimately eliminate the problem of slugging, it also decreases production rate and the lifetime of the field.

The method proposed here belongs to Caltec Ltd. and involves reduction of the backpressure caused by processing facilities on a platform. It is achieved by outlet pressure reduction from the first stage separator by means of an optional pump (-s) for the liquid phase and a surface jet pump for the gaseous phase as it is shown on Figure 5.1.

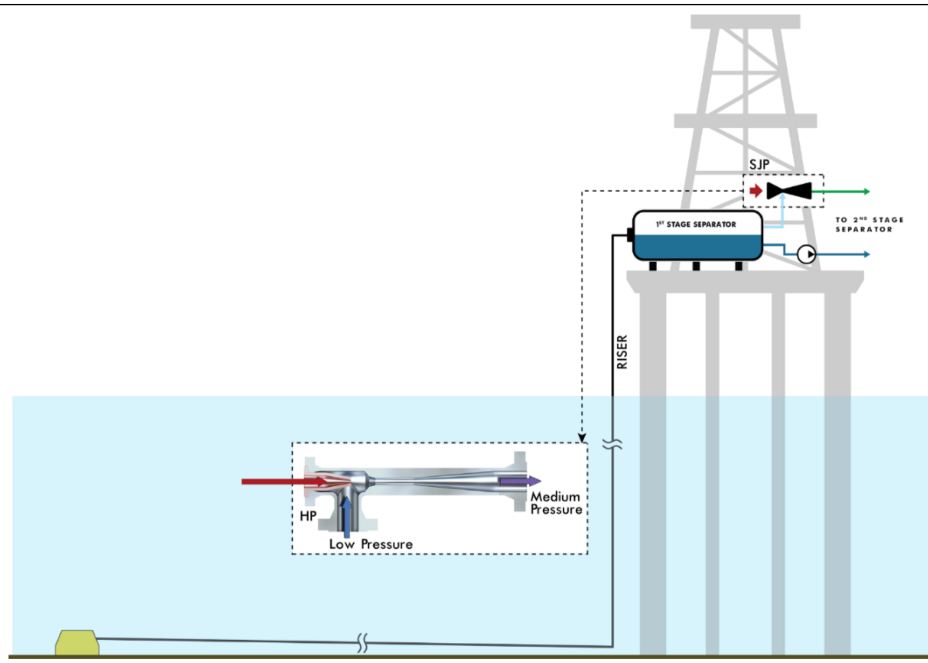


Figure 5.1. The layout of proposed method.

The surface jet pump installed on the gas line after the first stage separator is deemed create a pressure difference, hence lowering the separator outlet pressure while discharging gas at the required downstream pressure. An optional pump for each of the liquid phase leaving the separator may be also installed only if the reduced operating pressure is lower than downstream pressures. As the result of the system alteration the first stage separator may be lowered.

5.2. Hypothesis

The setup is deemed to lower the overall backpressure load created by the topside facilities on the subsea conduit for the cases where it may not be done in an alternative way, consequently lowering the backpressure imposed on the producing well. Depressurized pipeline conditions will allow the gas phase to expand, increasing its volumetric flowrate, thus increasing superficial gas velocity, resulting in change of the flow regime inside of the pipeline from stratified flow to stratified wavy flow and sweep-out of the accumulated liquids, dropping the pressure loss over the transportation system even further. In addition to that, the well is supposed to respond on decreased pressure with increased production rates, provided sufficient PI of the well, which will further increase superficial liquid and gas velocities. Increased superficial velocities push the operational point of the riser out of the severe slugging boundary, solving the problem of severe slugging.

The primary effect from the method implementation should come from increased superficial gas velocities and as a consequence possible transition from stratified flow to slug flow induced by hydrodynamic mechanism. This in turn will effectively restrict severe slugging occurrence as was previously discussed. The secondary effect comes from increased production rates due to lowered backpressure on the system. There is another advantage in Brownfields, by lowering the imposed backpressure some of the marginal backed-out wells which are tied-into some common manifold would have chance to flow too. Thus bring additional production gains.

As an additional benefit, lowered operational pressure inside of the separator may have positive effect, increasing the effectiveness of phase separation.

One of the major advantages for the proposed method is its relatively low CAPEX and OPEX and possibility of making use of already existing on platform facilities.

5.3. Surface Jet pump technology

Ejectors, also known as jet pumps, eductors, injectors, or thermocompressors, are a type of hydraulic pumps, which is operated based on the Venturi effect. The difference in the naming primarily stems from the particular application, whether it is injection, pumping, compression, creation of vacuum, etc. However, the underlying physical principle stays the same. Thus, ejectors, operated with incompressible fluids are normally referred to as jet pumps or eductors, whereas the terms ejector and injector used for compressible fluids [67].

Further reference to the technology as "surface jet pump" is deemed to eliminate ambiguity when the term "jet pump" is referred to a downhole artificial lift technology.

The device has no moving parts and allows mixing of two fluids, one of high total pressure (P_{HP}) and of low total pressure (P_{LP}), discharging a fluid at medium pressure (P_{MP}), so that $P_{HP} > P_{MP} > P_{LP}$. Due to the fact that the intermediate pressure is always greater than the pressure at inlet, the device can be used for raising pressure.

5.3.1. History and applications

The ejector was invented by Henry Giffard in 1858 and found its first application in steam locomotives, where the steam was used as a motive fluid to pump the boiler feedwater into the boiler [68]. A part of high pressure steam from the boiler was used as motive fluid. In the combining cone or mixer chamber, steam would condense, interacting with cold water, see Figure 5.2. Surplus of steam is flushed via overflow outlet. Slightly heated and pressurized water after delivery cone had enough pressure to overcome control valve and enter the boiler.

At the beginning of the 20th century, the principle was also used for removing air from a steam engine condenser. Later, ejectors were put into use to create vacuum and for cooling systems. Thus, in the early 1930s the technology was widely spread as a way for air conditioning of large buildings [69].

Steam jet pumping is used in desalination systems to eliminate mechanical vapor compression cycle, which uses high grade electrical energy. Jet ejectors proved to be the simplest, cheapest, and most reliable systems for vapor compression compared to other methods of compression making use of low grade heat energy [70].

Nowadays, injectors are found in power, chemical and refinery plants as well as in aerospace, refrigeration and vacuum applications, reliably fulfilling their main target – injection or sucking out.

At the same time, eductors remain the only alternative for pumping of high abrasive mixtures, due to high resistance to erosion. It allows eductors to be used for bulk handling of grains or other granular or powdered materials, flushing and pumping of slurries.

Ejectors are also widely used in wet scrubbers, which are an effective way of gas cleaning. The same principle is in place; the motive fluid, however, accomplishes two purposes at the same time: it creates suction for the gas to be cleaned and as the way to entrain particles or gases.

Apart from that, ejectors are used for small bubbles creation for industrial needs [71].

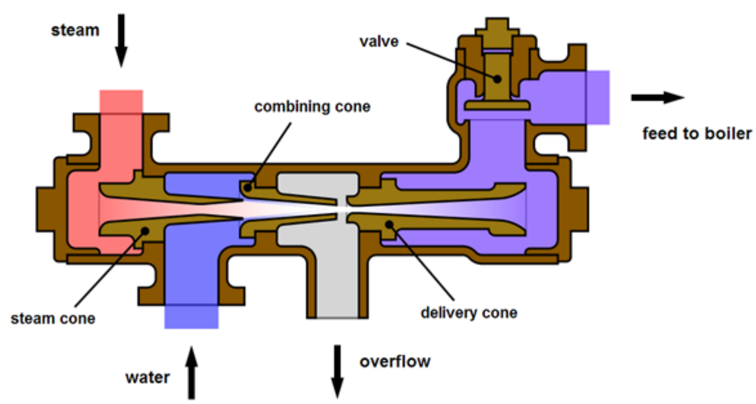


Figure 5.2. Steam locomotive boiler injector [72].

One of the first implementations of the jet pump and a beautiful example of smart engineering. Steam was used to drive feed water into the boiler.

5.3.2. Principle of operation

As it was mentioned above, injectors are based on the Bernoulli principle. They are used to pump or compress the secondary fluid by transfer of momentum and energy from the primary fluid.

Motive fluid, having high pressure and low velocity, is accelerated to high velocity jet through a converging or converging-diverging nozzle, where the initial potential energy of motive fluid is converted into the jet momentum at the nozzle exit according to Bernoulli equation. Thus, the escaping from the nozzle jet has high velocity and low static pressure and accelerates the secondary fluid in the direction of the stream. The two fluids are mixed in the mixing tube. Finally, the diffuser converts kinetic energy of the mixed fluids back to static pressure [67].

Let us have a look at an example of energy conversion according to Bernoulli equation in the nozzle. For the sake of simplicity, we consider a horizontal injector, operated with incompressible motive and driven fluids. No energy losses, such as losses in the nozzle or due to friction, are present in the system. Then, for the flow of HP fluid before and after the nozzle, Bernoulli equation may be written as:

$$\left(\frac{P}{\rho}+\frac{1}{2}V^2\right)_1=\left(\frac{P}{\rho}+\frac{1}{2}V^2\right)_4, \tag{5.1}$$

where indexes 1 and 4 correspond to cross sections 1 and 4 on Figure 5.3, respectively. Due to smaller diameter at the nozzle exit the flow speed at point 4 should be greater. Consequently, the pressure at the same point should drop to keep the equation in balance. Thus, on the right side of the nozzle the sucking pressure is created.

Due to reduced pressure and the action of friction forces, the driven fluid is entrained in the flow. Momentum transfer occurs and the two fluids are mixed in the mixing tube. In the diffuser the kinetic energy of the mixed fluids is converted back to potential, i.e. pressure. Thus, in the outlet of the ejector we have pressure greater than initial pressure of LP fluid.

Compressible fluid ejector based on the same principle as described above. However, the calculations become more difficult and effects, such as choking phenomena, compressibility, and energy balance equation should be taken into account.

5.3.3. Typical construction

A typical injector consists of inlets for motive (high pressure) and driven (low pressure) fluids, motive fluid nozzle, mixing tube and diffuser, see Figure 5.3. The mixing tube may be represented by a constant-area mixing shape, as it is depicted on Figure 5.3, or a constant-pressure mixing shape, where the mixing tube has conical shape with the wider side pointing towards the nozzle, see Figure 5.4 [69].

In case of the motive fluid being a gas, the nozzle may utilize de Laval design, so called converging-diverging nozzle, allowing the gas achieve speeds greater that the speed of sound in given conditions. Injection of the motive fluid with supersonic speed allows a greater conversion of primary fluid energy to secondary pressure increase [67]. For incompressible fluids, simple converging nozzles are used.

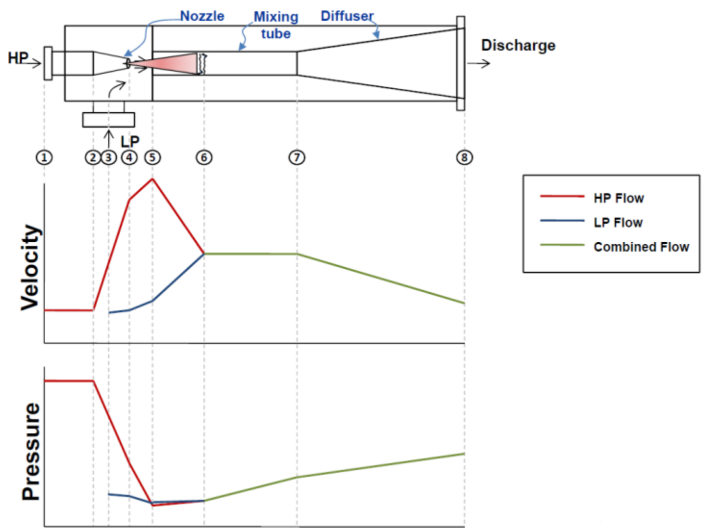


Figure 5.3. Velocity and pressure distribution along the stream axis of a jet pump [73].

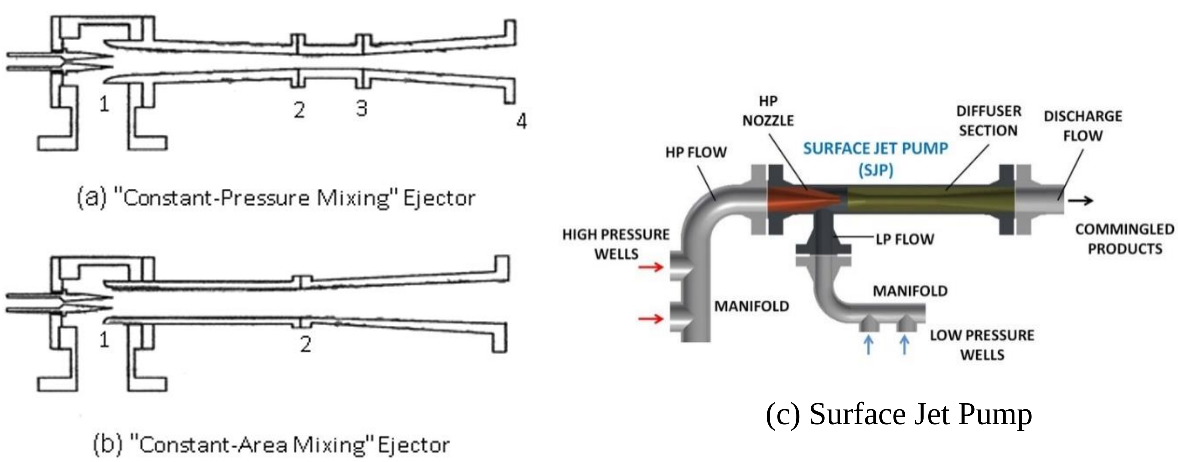


Figure 5.4. Different principles of mixing tubes *(a,b)* and design of an SJP *(c)* [69].

5.3.4. Applications in oil and gas industry

Surface jet pumps have a number of significant advantages making them an important option for compression or boosting purposes. To those one may attribute [74]:

1. No moving parts;
2. Little or no maintenance;
3. Reliable operation;
4. Easy in installation and controlling;
5. Low associated costs;
6. Low weight and size;
7. High safety level;
8. High tolerance to presence of multiphase flow, including solids;
9. Possibility to utilize otherwise wasted energy from HP sources;
10. Possibility to use liquid or gas as power fluid.

The main drawbacks of the technology are:

1. Relatively inefficient methods, with mechanical efficiency peaking at approx. 30%;
2. Significant volumes of power fluid required;
3. Performance may drop significantly with presence of multiphase flow;
4. Power fluid mixes with driven fluids (could be an advantage for mixing duties)

It is worth noticing that jet pump’s lower mechanical efficiency should not be overemphasized, since the technology may harness already available sources of HP fluids or utilize existing facilities.

The area where jet pumps are applied the most within the petroleum industry is their use for artificial lift of oil. The jet pump is insensitive to its orientation in space and therefore can be successfully applied in deep, directionally drilled, and crooked wells [75]. Absence of moving parts, and relatively small dimensions of the pump make maintenance infrequent and inexpensive. In addition, casing type installation allows a jet pump retrieval without pulling out of the tubing. Instead, the pump may be pulled by reverse flow or by introducing a wireline. The jet pump has high tolerance to corrosive and abrasive produced fluids as well as low requirements for pureness of motive fluid (in terms of solid particles). Moreover, water can be potentially used as the motive fluid. For that reasons, jet pumps are one of the best solutions for highly deviated deep wells producing heavy oils [75].

The jet pump found some of its implications in petroleum processing. Thus, it is the only possible solution to create vacuum for dearation purposes, start-up of operations, or sea water handling [76]. Implementation of a jet pump may also effectively prevent gas flaring. After many stages of separation, resulting gas pressure is often too low to be efficiently gathered. In that cases, the gas is usually burned using gas flares. As an alternative, low pressure gas may be compressed by means of a jet pump and rerouted directly to the export line or to processing facilities. Either gas or liquid may be used as a motive fluid for flare gases due to small rates [73].

Especial interest present the perspective of oil and gas production boosting with use of the jet pump either installed on a platform or subsea. Various sources of motive fluid may be utilized for that purpose.

Thus, Caltec [77] reports two implementation of Surface jet Pump on Hewett field between 1987 and 1991. As a result of utilization of available gas from HP producing wells, production from LP gas wells was increased from 100 MMScf/d to 125 MMScf/d for Lower Bunter wells, and from 36 to 51 MMScf/d for Upper Bunter wells.

Caltec [73, 78] refer to Gemini gas field, where a surface jet pump installed on the platform used a compressor recycle line as a source of motive gas. It resulted in reduced backpressure from the platform by 200 psi, which increased gas production from three subsea satellite wells 27 miles away. Increased flow rate swept away liquids accumulated in the pipeline, causing further drop of 140 psi. As the result, production increased by 24%; 2.5 BScf of otherwise lost reserves recovered because of the lowered backpressure.

Oil wells production boosting is somewhat more complicated and often involves implementation of the inline separator (I-SEP by Caltec).

Caltec [78] reports that separated liquid from a high pressure oil well with additional boost from an oil pump was used as a motive fluid in the WELLCOM system to increase production from some oil wells. As the result, backpressure on the wells dropped by 11 bar, increasing oil production 350 bbl/d and gas production by 1.5 MMScf/d.

Caltec [78] reports implementation of the WELLCOM system to bring some closed low pressure oil wells back to production making use of the energy from a HP well. It resulted in increased oil production by more than 35% (about 150 bbl/d).

5.4. Effect on a constant mass flowrate system

The main mechanism allowing severe slugging mitigation by pipeline internal pressure reduction is based on the effect of gas phase compressibility. Thus, when the pressure is reduced the gas phase expands, resulting in increased superficial gas velocity. If the velocity exceeds the limit set by stratified flow criterion (see Section 2.4.1), the flow regime inside the pipeline shifts from stratified flow to stratified wavy flow and further to the point where hydrodynamic slugs are induced, restricting liquid accumulation at the riser base and severe slugging initiation. The secondary effect comes from the fact that fluids that are typical for oil and gas pipelines allow mass transfer between themselves. Therefore, pipeline pressure reduction shifts the equilibrium point between the liquid and gaseous phases of petroleum to the favor of gas, further increasing superficial gas velocity inside the conduit.

To investigate on the effect of reduced pipeline pressure on severe slugging, a series of simulations was run similar to that of Section 4.1.6 to get the flow properties and plot flow regime maps. The total of 122 simulations were documented, all of which may be found in Appendices A to D. Resulting flow regime maps are presented on Figure 5.5 expressed in terms of oil production rate/GOR and oil/gas production rate at standard conditions.

Figure 5.5 and Figure 4.6 from Section 4.1.6 show how severe slugging occurrence region diminishes with reduced backpressure imposed on the system due to induced hydrodynamic slugging inside the pipeline. The effect is more noticeable in the region of medium liquid flowrates (>500 Sm^3/d) and gas-oil ratios (>50 Sm^3/Sm^3).

One of the most important factors influencing the efficiency of this approach is the pipeline declination angle, which positively correlates with tendency of stratified flow regime being present in the conduit. Therefore, one must expect to see lessening effect of the proposed mitigation technique with increasing declination angle value since stratified flow will be forced before the riser bend.

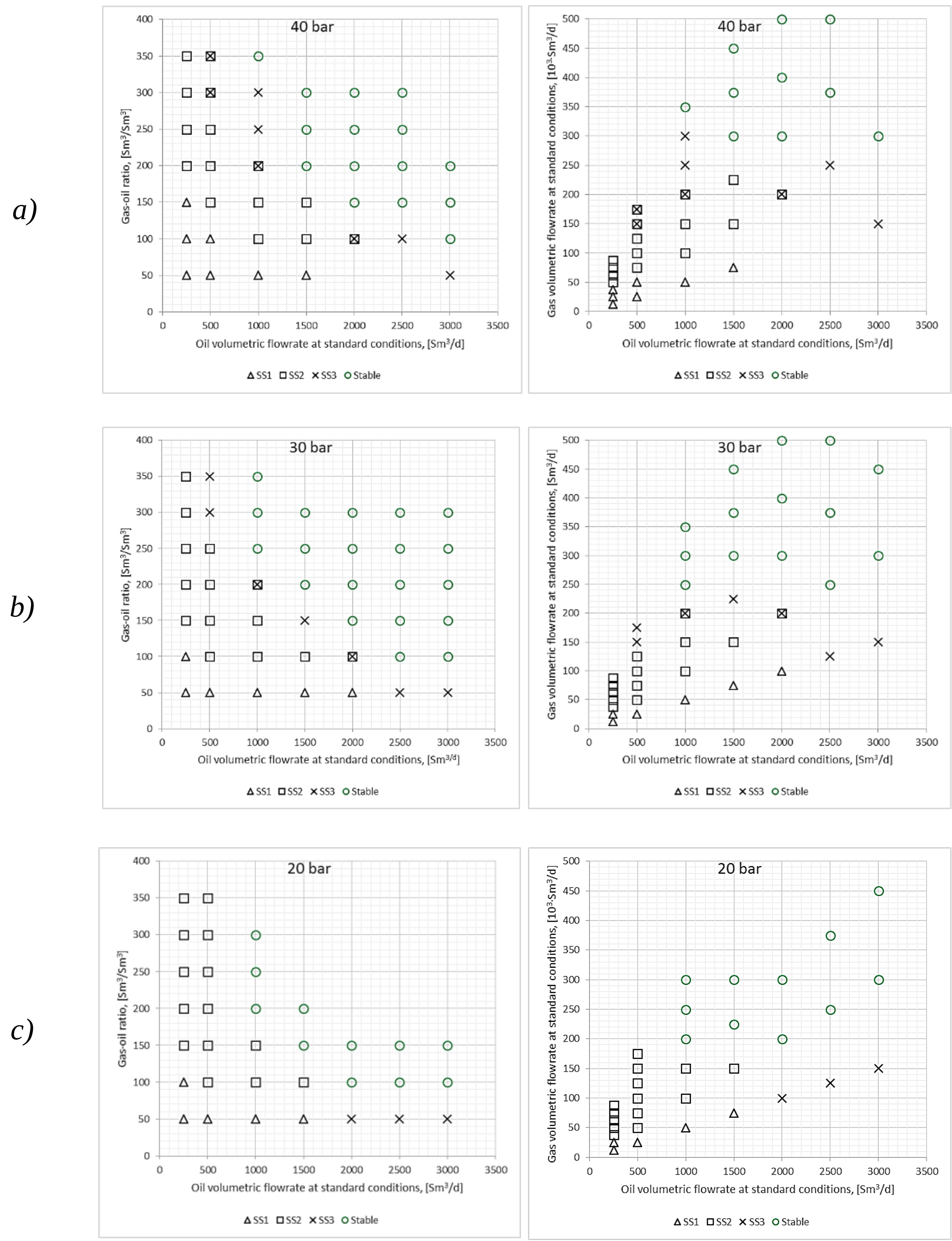


Figure 5.5. Severe slugging occurrence flow regime map for the separator pressure of 40 (*a*), 30 (*b*) and 20 (*c*) bara.

The green circles represent non-severe-slugging flow. The triangles, squares and crests stand for severe slugging of type 1, 2 and 3, respectively. The squares with crosses inside represent the cases when it was hard to distinguish SS2 from SS3.
The plots show how with decreased separator pressure the SS region shrinks.

5.5. Effect on a system with producing well

An additional benefit from the pipeline depressurization is increased production due to lower backpressure on the producing wells. Increased volumes of produced fluid will increase superficial velocities, further affecting severe slugging occurrence criteria in a positive way. The effect is highly dependent on the reservoir characteristic, especially PI value.

To illustrate the mitigation technique, the study case was altered in the way described in Table 5.1.

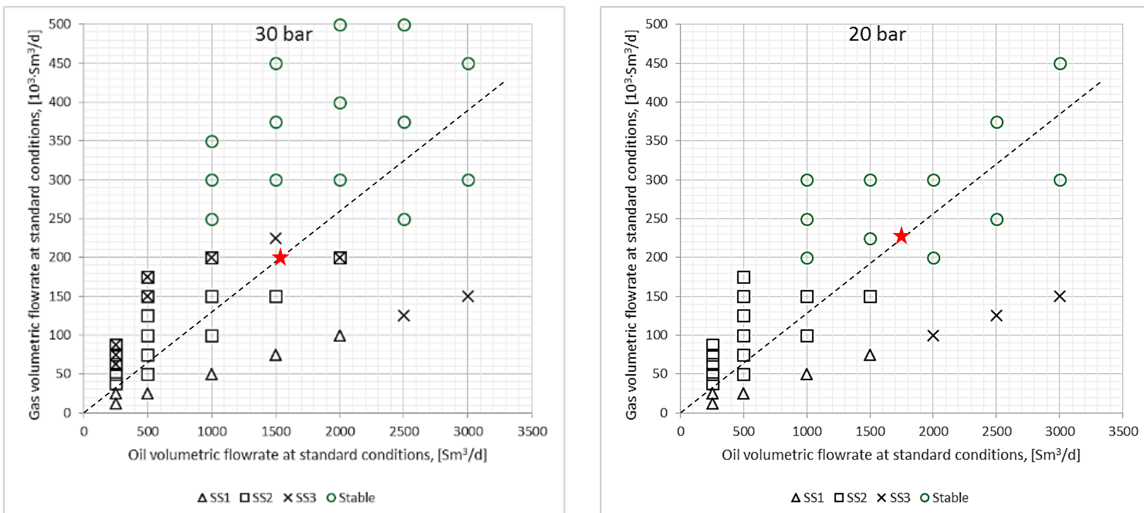
Table 5.1. Reservoir characteristics

Reservoir pressure, [bara]	*280*
Maximum production rate, [Sm^3/d]	*6800*
Separator inlet pressure, [bara]	*30*
Gas-oil ratio, [Sm^3/Sm^3]	*130*

As before, the saturated oil reservoir exists at 50 °C. At the specified conditions the flowing bottom hole pressure is at 242 bara with average oil production rate of 1540 Sm^3/d, resulting in PI of approximately 18 STB/d/psi. With given parameters flow regime inside of the riser may be characterized as severe slugging of type 3 with pressure at the riser base fluctuations greater than 20 bar. It is further assumed that implementation of a jet pump at the first stage separator gas outlet allowed its inlet pressure to be reduced by 10 bar.

The resulting flow regime transition is showed on Figure 5.6. Depressurization of the pipeline system with implementation of a jet pump caused operational point, which is denoted as a red star, to move from severe slugging region (Figure 5.6.(a)) to stable flow region (Figure 5.6.(b)).

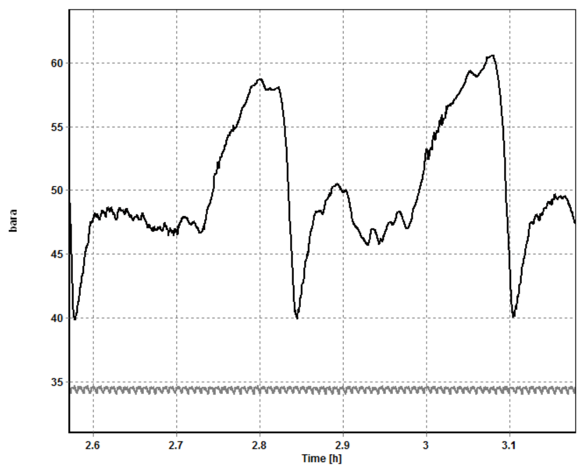
Figure 5.7 compares the time histories of the key parameters for severe slugging, where the black curves represent the case before and the grey curves after pipeline depressurization. Figure 5.7.(a) shows that the riser base pressure after jet pump implementation became close to constant value, indicating a stable flow regime, whereas the initial case exhibits pressure fluctuation of more than 20 bar. Liquid mass flowrate graph (Figure 5.7.(b)) with the mitigation technique being implemented shows arrival of slugs with consistent frequency, compared to severe slugging liquid surges peaks depicted on the same plot. Finally, Figure 5.7.(c) shows that the backpressure imposed on the producing well in the case of depressurization is close to constant, translating into a constant production rate, compared to that of untreated case, which displays high production flowrate variations due to effect of severe slugging. It is important to notice that production rate for the case of stable pipeline-riser flow is higher than the average production rate in case of severe slugging, resulting in the operational point shift.



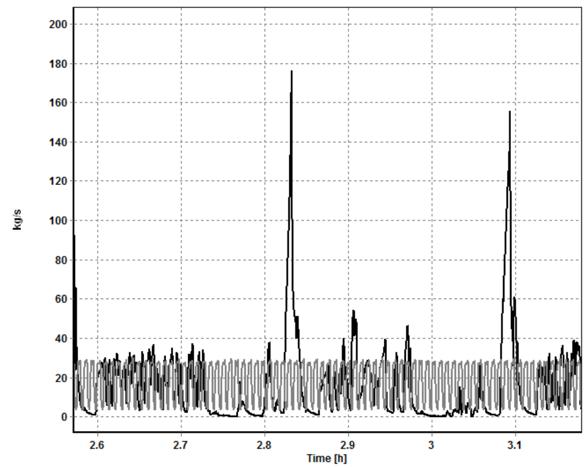
a) *b)*

Figure 5.6. Flow regime transition from the untreated case (*a*) to the depressurized system (*b*).

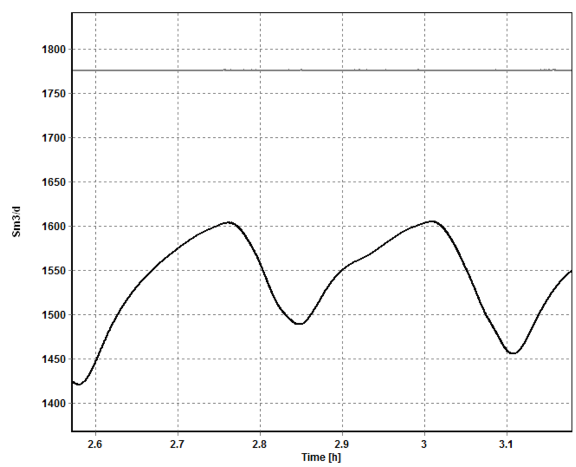
The dashed line represent all possible flow combinations with given gas-oil ratio of 130. Red stars correspond to gas-oil flowrates combination for 30 and 20 bara separator inlet pressure.



a)



b)



c)

Figure 5.7. Surface jet pump implementation effect on a system with a producing well expressed as time history of pressure at the riser base (*a*), mass liquid flowrate into the separator (*b*), and the well production rate (*c*).

Black curves stand for the case before the jet pump implementation and the grey curves for after. Graphs are generated in OLGA.

5.6. Surface Jet pump calculation example

This chapter provides a simplified ejector calculation procedure for choice of HP fluid quantities, as can be found in "Surface Jet Pumps (SJPs) for Enhanced Oil & Gas production" by Beg and Sarshar [74].

A number of assumptions are to be made:

- Pressure drop over the first stage separator is equal and constant to 5 bar;
- Temperature inside of the separator is equal and constant at 25 °C;
- Gas from the first stage separator must meet the pressure requirement of 25 bara to be fed into the last stage compressor;
- The compressor performance is not sensitive to gas flowrate;
- Oil and gas inside the first stage separator exist at equilibrium specified by inlet pressure and separator temperature and described by Blackoil model (see Section 4.1.2);
- Motive and driven gas have the same properties.

The case of Section 5.5 is considered. The surface jet pump is installed on the gas line between the first stage separator and the last stage compressor to lower the outlet pressure from 25 to 15 bara, reducing the inlet pressure to 20 bara from initial 30 bara. Therefore, pressure difference of 10 bar must be created by the surface jet pump to achieve the same pressure as in the base case.

The production rate with the mitigation technique being installed was calculated before and equal to 1780 Sm^3/d, see Section 5.5. To estimate the amount of gas routed to the LP inlet of the surface jet pump, solution gas oil ratio at the first stage separator conditions have to be calculated. Assuming that the fluids exist at equilibrium specified by inlet pressure of 20 bar and temperature equal to 25 °C, *RSGO* may be calculated according to Equation 4.2:

$$R_S|_{P=20bara;T=25°C} = 4.4\ [Sm^3/Sm^3]. \tag{5.2}$$

Therefore, the amount of gas fed into the LP line of the jet pump is equal to:

$$Q_{G0} = Q_{O0}(GOR - R_S) = 1780 \cdot (130 - 4.4) \approx 224 \cdot 10^3 \ \left[\frac{Sm^3}{d}\right]. \tag{5.3}$$

Engineers' Handbook in Surface Jet Pump [74] provides with a graphical way of Surface Jet Pump performance estimation, see Figure 5.8. Assuming that the source of HP gas at 100 bara is available (e.g. export line, compressor recycle line), the amount of HP required may be estimated. As it follows from the figure, discharge pressure to LP and HP to LP ratios are necessary to do so:

$$\frac{P_{discharge}}{P_{LP}} = \frac{25\ bara}{15\ bara} \approx 1{,}67, \tag{5.4}$$

$$\frac{P_{HP}}{P_{LP}} = \frac{100\ bara}{15\ bara} \approx 6{,}67. \tag{5.5}$$

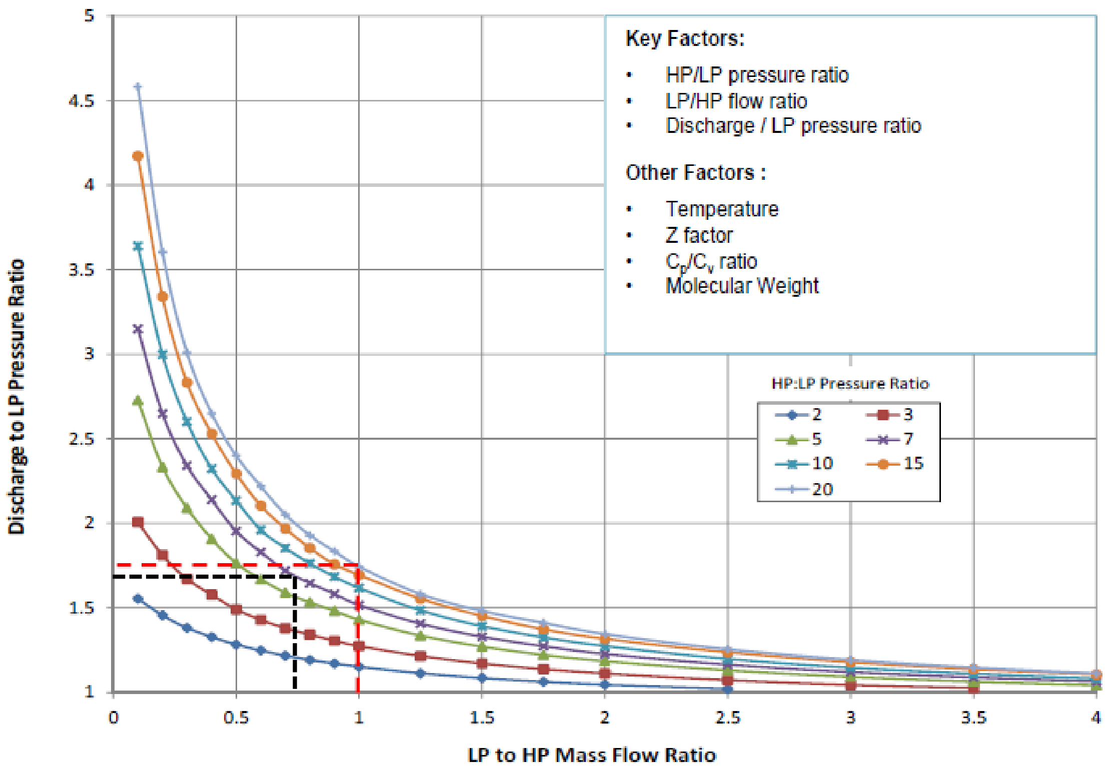


Figure 5.8. Ejector performance curves [74].

The curves show how the key ratios relate to each other. The black dashed line represents the example of this part.

Applying the values to Figure 5.8, LP to HP gas mass flow ratio may be estimated to be around 0.7. Thus, taking into account that the motive and driven gases have the same properties, required volumetric flowrate of HP gas is equal to:

$$Q_{G0,HP} = \frac{1}{0.7} \cdot Q_{G0} = 320 \cdot 10^3 \left[\frac{Sm^3}{d}\right]. \quad (5.6)$$

Therefore, about 320,000 Sm^3/d of motive gas at 100 bara is required to implement the method and eliminate severe slug flow at a given study case. The most probable gas sources would be partly diverted export line gas, HP gas for injection wells, or gas from the last compressor recycle line.

Conclusions

This thesis had two main tasks to accomplish:

1) Provide a thorough theoretical overview of severe slugging phenomenon and techniques for its mitigation (Chapters 2-4);

2) Evaluate a novel mitigation method proposed by Caltec Ltd., UK, based on application of the Surface Jet Pump technology (Chapter 5).

The first task is of importance due to the scarcity of the literature on the topic. The main body of knowledge is scattered in various papers over the last four decades, making its understanding somewhat troublesome. The author hopes that this thesis will become a reference work for those interested in the topic. The present work explains in detail the phenomenon mechanism and classification, points out at some possible confusions that may arise, and highlights the importance of recent publications, e.g. contributing mass transfer effect to the understanding of severe slugging cycle.

It may be even more difficult to find information regarding severe slugging mitigation techniques. The thesis provides a state of the art, an overview of all alleviation methods that were possible to find, including patents in the field. Chapter 4 provides some guidance and examples to simulation of some of those in the multiphase simulation software OLGA.

Chapter 5 describes a new severe slug mitigation method based on pipeline system depressurization. Even though its basic mechanism is not novel and well understood, its application may have been overlooked due to restrictions imposed by process facilities and the impossibility to reduce the pipeline outlet pressure. The thesis provides a new option, showing how a surface jet pump implementation may eliminate severe slugging in a cost-efficient manner and further increase production rates, in contrast to the majority of mitigation methods used up to date, which to some degree restricts production flow.

The chapter demonstrates that the method induces non-detrimental hydrodynamic slugging inside the pipe by increased superficial gas velocities and avoids severe slugging initiation process at the riser base.

As the result of simulations conducted, positive results were achieved showing that the proposed mitigation method is feasible for flow stabilization at some flowrates, see Section 5.4. However, the thesis has a limited scope, which mostly considers pipeline-riser system's behavior under reduced pressure. To get a thorough picture, restrictions imposed by the process facilities and the reservoir must be included. Thus, undesirable effects such as increased water cut and excessive GLR due to higher flowrate and migration in the reservoir should be taken into consideration.

The second part of the thesis achieved its main goal – to show that severe slugging may be avoided by pipeline depressurization with a surface jet pump – by providing a study case where the mitigation technique proposed by Caltec Ltd. eliminated severe slugging flow, making transition to a stable slug flow.

Recommendations for further work, self-evaluation

Due to time constrains for this project and delays related to getting a version of simulation software, it was not possible to model the effect of active controlled choking on the example of a topside choke controlled by a PID controller; as it was planned initially. It would be also beneficial, but very time demanding, to include frequency spectrum analysis of flow oscillations based on Fourier transformation.

Some inconsistency with other authors was found when identifying different types of severe slugging. Based on general description of the types, SS3 regimes were observed for high gas flowrates for the simulations of the separator pressure of 30 and 40 bara, see Figure 5.5.(a-b).

Another improvement would include gas-lift mandrel node within OLGA model for riser-base injection method instead of constant mass flowrate as it was done in the thesis.

The dip at the beginning of the pipeline was not necessary. Even though it was interesting to observe, no actual documentation of observations is made in the thesis. Also it may have interfered with the results, though in a very subtle way.

Recommendations for further work:

1) In-depth investigation on the effect of mass transfer. Comparative analysis between two models one of which would include Blackoil modeled oil-gas solution, the other would consist of a PVT table with two fluids of identical properties, but not allowing for mass transfer to occur (refer to [63, 79]). Development of such a model if not too complicated would help in active control methods design and tuning prior to installation.
2) Those mathematical models describing mass transfer effect on severe slugging assume oil and gas existing at equilibrium at a given pipeline cross section [63]. They may be further improved by developing a severe slugging model which takes into account discrepancy between boiling and condensation of the gaseous phase out of oil. However, it may be that the resulting model will improve results just marginally, whereas increasing complexity significantly.
3) It may be possible to implement the method proposed by Caltec Ltd. and evaluate its effect on a gas condensate pipeline.

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Appendix A.

Table A.1. Flow regime identification for separator pressure of 50 bara.

Gas-oil-ratio, [Sm^3/ Sm^3]	Absolute pressure reading at the riser base, [bara]		Pressure fluctuation value, [bar]	Severe slugging cycle period, [h]	Maximum observed liquid mass flowrate, [kg/s]	Flow regime
	Maximum	Minimum				
$Q_{o0} = 250\ \mathrm{Sm^3/d}$						
50	*91,2*	*60,4*	*30,8*	***	*130*	*SS1*
100	*90,6*	*57,4*	*33,2*	*1,25*	*168*	*SS1*
150	*91,3*	*57,3*	*34*	*1,05*	*193*	*SS1*
200	*90,2*	*57,1*	*33,1*	*0,97*	*188*	*SS2*
250	*85*	*56,9*	*28,1*	*0,68*	*153*	*SS2*
300	*77,7*	*56,2*	*21,5*	*0,56*	*110*	*SS2*
350	*79,7*	*56,4*	*23,3*	*0,55*	*122*	*SS2*
$Q_{o0} = 500\ Sm^3/d$						
50	*91,3*	*63,9*	*27,4*	*1,63*	*120*	*SS1*
100	*91,2*	*60,7*	*30,5*	*0,81*	*163*	*SS1*
150	*91,2*	*58*	*33,2*	*0,52*	*197*	*SS2*
200	*87,4*	*58,2*	*29,2*	*0,37*	*173*	*SS2*
250	*82*	*57,3*	*24,7*	*0,33*	*226*	*SS2*
300	*77,7*	*56,9*	*20,8*	*0,3*	*123*	*SS2*
350	*75*	*57,1*	*17,9*	*0,25*	*102*	*SS2*
$Q_{o0} = 1000\ Sm^3/d$						
50	*91*	*64*	*27*	*0,9*	*119*	*SS1*
100	*90*	*63*	*27*	*0,3*	*165*	*SS1*
150	*90,5*	*60,1*	*30,4*	*0,29*	*195*	*SS2*
200	*85*	*60*	*25*	*0,21*	*163*	*SS2*
250	*81,2*	*58,5*	*22,7*	*0,26**	*152*	*SS2-SS3*
300	*78*	*59*	*19*	*0.37**	*130*	*SS2-SS3*
350	*71*	*59,1*	*11,9*	*-*	*-*	*stable*
$Q_{o0} = 1500\ Sm^3/d$						
50	*89,5*	*75,3*	*14,2*	*0,27*	*53*	****
100	*89,4*	*63,1*	*26,3*	*0,18*	*165*	*SS1*
150	*88,8*	*60,6*	*28,2*	*0,25*	*191*	*SS2*

200	*83,1*	*59,6*	*23,5*	*0,29**	*84,7*	*SS2*
250	*71,3*	*61,4*	*9,9*	-	*54*	*stable*
300	-	-	-	-	-	*stable*
350	-	-	-	-	-	*stable*
$Q_{o0} = 2000\, Sm^3/d$						
50	*91*	*64*	*27*	*0,9*	*119*	****
100	*86*	*65*	*21*	*0,15*	*140*	*SS3*
150	*87,6*	*62,3*	*25,3*	*0,22**	*168*	*SS3*
200	*67,6*	*67,2*	*0,4*	-	-	*stable*
250	-	-	-	-	-	*stable*
300	*66,2*	*64,8*	*1,4*	-	-	*stable*
$Q_{o0} = 2500\, Sm^3/d$						
50	*91*	*64*	*27*	*0,9*	*119*	****
100	*86*	*68*	*18*	*0,15*	*115*	*SS3*
200	*68,8*	*68,2*	*0,6*	-	-	*stable*
300	*68,1*	*67,3*	*0,8*	-	-	*stable*
$Q_{o0} = 3000\, Sm^3/d$						
50	*91*	*64*	*27*	*0,9*	*119*	****
100	*78,3*	*74,5*	*3,8*	-	-	*stable*
200	*70,6*	*69,5*	*1,1*	-	-	*stable*

* - flow exhibit irregular behavior; the value of period is either approximate or omitted.
** - seemingly stable flow with sudden release of hast liquid surges (mentioned in Section 4.1.6).

Appendix B.

Table B.1. Flow regime identification for separator pressure of 40 bara.

Gas-oil-ratio, [Sm³/ Sm³]	Absolute pressure reading at the riser base, [bara]		Pressure fluctuation value, [bar]	Severe slugging cycle period, [h]	Maximum observed liquid mass flowrate, [kg/s]	Flow regime
	Maximum	Minimum				
$Q_{o0} = 250\ \mathrm{Sm^3/d}$						
50	*81,3*	*49,9*	*31,4*	*1,94*	*144*	*SS1*
100	*81,6*	*47,1*	*34,5*	*1,25*	*186*	*SS1*
150	*81,8*	*47,4*	*34,4*	*1,14*	*226*	*SS1*
200	*73,9*	*46,8*	*27,1*	*0,73*	*166*	*SS2*
250	*72*	*46,7*	*25,3*	*0,74*	*118*	*SS2*
300	*65,2*	*45,8*	*19,4*	*0,5*	*102*	*SS2*
350	*64,7*	*45*	*19,7*	*0,5*	*148*	*SS2*
$Q_{o0} = 500\ Sm^3/d$						
50	*81,4*	*53*	*28,4*	*1,15*	*146*	*SS1*
100	*81,2*	*49*	*32,2*	*0,58*	*193*	*SS1*
150	*81,6*	*47,9*	*33,7*	*0,47*	*223*	*SS2*
200	*71,6*	*47,1*	*24,5*	*0,37*	*169*	*SS2*
250	*69,7*	*46,9*	*22,8*	*0,26*	*130*	*SS2*
300	*64,3*	*46,1*	*18,2*	*0,25*	*132*	*SS2-SS3*
350	*62,3*	*46,2*	*16,1*	*0,24*	*115*	*SS2-SS3*
$Q_{o0} = 1000\ Sm^3/d$						
50	*80,9*	*57,6*	*23,3*	*0,5*	*114*	*SS1*
100	*80,1*	*50,4*	*29,7*	*0,29*	*196*	*SS2*
150	*80,2*	*48,6*	*31,6*	*0,33*	*204*	*SS2*
200	*73,7*	*47,8*	*25,9*	*0,28*	*185*	*SS2-SS3*
250	*69,5*	*48,2*	*21,3*	*0,39*	*161*	*SS3*
300	*58,4*	*49,3*	*9,1*	*0,37*	*55*	*SS3*
350	*-*	*-*	*-*	*-*	*-*	*stable*
$Q_{o0} = 1500\ Sm^3/d$						
50	*79,6*	*60,6*	*19*	*0,31*	*96*	*SS1*
100	*80*	*51*	*29*	*0,19*	*197,4*	*SS2*
150	*75,5*	*49*	*26,5*	*0,24*	*174*	*SS2*

200	-	-	-	-	-	stable
250	-	-	-	-	-	stable
300	-	-	-	-	-	stable

$Q_{o0} = 2000\ Sm^3/d$

50	78,7	65,1	13,6	0,21	87,2	**
100	79	52,1	26,9	0,25	174	SS2-SS3
150	-	-	-	-	-	stable
200	-	-	-	-	-	stable
250	-	-	-	-	-	stable
300	-	-	-	-	-	stable

$Q_{o0} = 2500\ Sm^3/d$

50	78,9	64,6	14,3	0,17	80	**
100	76,4	56	20,4	*	143	SS3
150	-	-	-	-	-	stable
200	-	-	-	-	-	stable
300	-	-	-	-	-	stable

$Q_{o0} = 3000\ Sm^3/d$

50	79,6	66,3	13,3	0,14*	77	SS3
100	-	-	-	-	-	Stable
150	-	-	-	-	-	Stable

* - flow exhibit irregular behavior; the value of period is either approximate or omitted.

** - seemingly stable flow with sudden release of hast liquid surges (mentioned in Section 4.1.6).

Appendix C.

Table C.1. Flow regime identification for separator pressure of 30 bara.

Gas-oil-ratio, [Sm^3/ Sm^3]	Absolute pressure reading at the riser base, [bara]		Pressure fluctuation value, [bar]	Severe slugging cycle period, [h]	Maximum observed liquid mass flowrate, [kg/s]	Flow regime
	Maximum	Minimum				
$Q_{o0} = 250\ \mathrm{Sm^3/d}$						
50	*71,3*	*39,7*	*31,6*	*1,98*	*171*	*SS1*
100	*71,9*	*37,1*	*34,8*	*1,21*	*221*	*SS1*
150	*71,8*	*36,5*	*35,3*	*1,09*	*248*	*SS2*
200	*59,8*	*35,4*	*24,4*	*0,65*	*133*	*SS2*
250	*54,6*	*35,5*	*19,1*	*0,56*	*146*	*SS2*
300	*50,4*	*34,8*	*15,6*	*0,4*	*107*	*SS2*
350	*49,3*	*34,4*	*14,9*	*0,34*	*105*	*SS2*
$Q_{o0} = 500\ Sm^3/d$						
50	*71,6*	*42,5*	*29,1*	*0,89*	*151*	*SS1*
100	*71,7*	*38,2*	*33,5*	*0,61*	*228*	*SS2*
150	*66,3*	*37,5*	*28,8*	*0,52*	*213*	*SS2*
200	*58,2*	*36,6*	*21,6*	*0,38*	*130*	*SS2*
250	*54,6*	*36,6*	*18*	*	*149*	*SS2*
300	*52,7*	*35,6*	*17,1*	*	*120*	*SS3*
350	*51,4*	*35,1*	*16,3*		*105*	*SS3*
$Q_{o0} = 1000\ Sm^3/d$						
50	*70,9*	*49*	*21,9*	*0,45*	*155*	*SS1*
100	*70,9*	*39,9*	*31*	*0,37*	*228*	*SS2*
150	*65,6*	*38,6*	*27*	*0,36*	*218*	*SS2*
200	*58,9*	*38,2*	*20,7*	-	*138*	*SS2-SS3*
250	*48,1*	*38,7*	*9,4*	-	*45*	*stable*
300	-	-	-	-	-	*stable*
$Q_{o0} = 1500\ Sm^3/d$						
50	*69,9*	*49*	*20,9*	*0,27*	*138*	*SS1*
100	*69,5*	*39,8*	*29,7*	*0,25**	*217*	*SS2*
150	*60,4*	*39,9*	*20,5*	*0,36**	*141*	*SS3*
200	-	-	-	-	-	*stable*

300	-	-	-	-	-	*stable*

$Q_{o0} = 2000\ Sm^3/d$

50	*67,7*	*49,9*	*17,8*	*0,18*	*136*	*SS1*
100	*68,4*	*41,6*	*26,8*	*0,37* *	*207*	*SS2-SS3*
150	-	-	-	-	-	*stable*
200	-	-	-	-	-	*stable*
300	-	-	-	-	-	*stable*

$Q_{o0} = 2500\ Sm^3/d$

50	*67,5*	*51,6*	*15,9*	*0,15*	*118*	*SS3*
100	-	-	-	-	-	*stable*
150	-	-	-	-	-	*stable*
200	-	-	-	-	-	*stable*
300	-	-	-	-	-	*stable*

$Q_{o0} = 3000\ Sm^3/d$

50	*66*	*52,7*	*13,3*	*0,13*	*103*	*SS3*
100	-	-	-	-	-	*stable*
150	-	-	-	-	-	*stable*
200	-	-	-	-	-	*stable*
300	-	-	-	-	-	*stable*

* - flow exhibit irregular behavior; the value of period is either approximate or omitted.

Appendix D.

Table D.1. Flow regime identification for separator pressure of 20 bara.

Gas-oil-ratio, [Sm^3/ Sm^3]	Absolute pressure reading at the riser base, [bara]		Pressure fluctuation value, [bar]	Severe slugging cycle period, [h]	Maximum observed liquid mass flowrate, [kg/s]	Flow regime
	Maximum	Minimum				
$Q_{o0} = 250\ \mathrm{Sm}^3/\mathrm{d}$						
50	*61,9*	*27,2*	*34,7*	*1,96*	*210*	*SS1*
100	*62,1*	*28,5*	*33,6*	*1,35*	*245*	*SS1*
150	*50,2*	*25,7*	*24,5*	*0,87*	*175*	*SS2*
200	*47*	*24,6*	*22,4*	*0,67*	*140*	*SS2*
250	*38,3*	*24,4*	*13,9*	*0,44*	*105*	*SS2*
300	*37,7*	*23,4*	*14,3*	*0,43*	*120*	*SS2*
350	*36,2*	*23,7*	*12,5*	***	*65*	*SS2*
$Q_{o0} = 500\ Sm^3/d$						
50	*61,7*	*31,2*	*30,5*	*0,71*	*214*	*SS1*
100	*59,5*	*28*	*31,5*	*0,71*	*228*	*SS2*
150	*49,2*	*26,1*	*23,1*	*0,52*	*153*	*SS2*
200	*43,8*	*25*	*18,8*	*0,34*	*175*	*SS2*
250	*42,8*	*25*	*17,8*	*0,37*	*107*	*SS2*
300	*39,5*	*24,4*	*15,1*	*0,44*	*146*	*SS2*
350	*36,3*	*24,3*	*12*	*0,5*	*84*	*SS2*
$Q_{o0} = 1000\ Sm^3/d$						
50	*61,3*	*33,7*	*27,6*	*0,38*	*203*	*SS1*
100	*61,5*	*30*	*31,5*	*0,5*	*234*	*SS2*
150	*47,7*	*26,9*	*20,8*	***	*148*	*SS2*
200	-	-	-	-	-	*Stable*
250	-	-	-	-	-	*Stable*
$Q_{o0} = 1500\ Sm^3/d$						
50	*60,1*	*35,7*	*24,4*	*0,3*	*185*	*SS1*
100	*55,3*	*29,8*	*25,5*	*0,37*	*210*	*SS2*
150	-	-	-	-	-	*Stable*
200	-	-	-	-	-	*Stable*

$Q_{o0} = 2000\ Sm^3/d$						
50	*57,1*	*36,2*	*20,9*	*0,14*	*186*	*SS3*
100	-	-	-	-	-	*Stable*
150	-	-	-	-	-	*Stable*
$Q_{o0} = 2500\ Sm^3/d$						
50	*57,6*	*37,8*	*19,8*	*0,12*	*163*	*SS3*
100	-	-	-	-	-	*Stable*
150	-	-	-	-	-	*Stable*
$Q_{o0} = 3000\ Sm^3/d$						
50	*56,9*	*41,6*	*15,3*	-	*120*	*SS3*
100	-	-	-	-	-	*Stable*
150	-	-	-	-	-	*Stable*

* - flow exhibit irregular behavior; the value of period is either approximate or omitted.

Appendix E.

Table E.1. Riser base pressure readings with topside choking

Relative choke opening, [-]	Maximum observed pressure, [bara]	Minimum observed pressure, [bara]	Average pressure, [bara]
0,24	*88,1*	*62,9*	*75,4*
0,22	*88,3*	*62,9*	*75,6*
0,2	*89,3*	*62,8*	*76,3*
0,18	*88,2*	*64,3*	*75,7*
0,16	*89,1*	*64,3*	*76,7*
0,14	*88,0*	*66,1*	*76,9*
0,12	*88,0*	*66,7*	*77,5*
0,1	*89,6*	*69,4*	*78,5*
0,08	*88,6*	*71,3*	*80,9*
0,07	*88,0*	*73,3*	*81,0*
0,06	*82,4*	*80,8*	*82,0*
0,05	*84,5*	*84,3*	*84,4*
0,04	*88,6*	*88,5*	*88,5*
0,03	*95,0*	*94,9*	*95,0*