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New slug control strategies, tuning rules and experimental results

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Abstract

The challenge of handling intermittent flow with liquid slugs followed by gas pockets in multiphase flow lines becomes more important when the number of satellite fields increases. Oil, gas and water are transported from the wells in some km long flow lines along the seabed and up through a riser to the oil rig. Slugging may cause several problems for topside processing. This paper concerns suppression of slug flow by active use of the topside choke. Process measurements such as pressure and density are used in a PID controller. Slugging has been reduced significantly with such a system in operation offshore since April 2001. This paper contains results within simplified modelling of flow dynamics. New experimental results verify the dynamic model. A control scheme in operation offshore has been tested and a new control scheme independent of subsea measurements has been developed and tested in experiments.

Keywords: Slug control; Multiphase flow

1. Introduction

A slug is a lump of liquid in a multiphase well stream (see Fig. 1). Slugging refers to varying or irregular flows and surges of gas and liquid through any cross-section of a flow line [1].

There are many kinds of slugging. Hydrodynamic slugs are built in near horizontal parts of the flow line, but may also occur in wells and risers. These slugs are usually short and appear frequently. The inlet separator will in most cases handle these slugs well, since the amount of liquid in each slug is little compared to the free volume in the separator. Gravity forces can generate riser slugs (see Fig. 2), when the flow line has a low point in front of the riser. A riser slug contains a lot of liquid and can represent a major challenge to the downstream processing system. Separator levels and compressor flow will oscillate, when the production rates vary. Variations will also take place at later stages in the separator and compressor trains. Many factors are important for the degree of slugging. The most important are flow line pressure, gas and liquid production rates and flow line topography. Riser slugging is most likely to occur at low rates and a low flow line pressure towards the end of the field lifetime.

Slugging may have undesirable effects on the oil and gas production process. Severe slugging will affect the inlet separator liquid level, may give poor separation, and in some cases lead to separator flooding. The oscillating pressure may cause wear on processing equipment and will reduce the lifetime and increase the maintenance costs compared to production with even flow. Well pressure will oscillate during severe slugging, and this might reduce well performance. A varying gas flow will result in varying separator pressure and poor separation and

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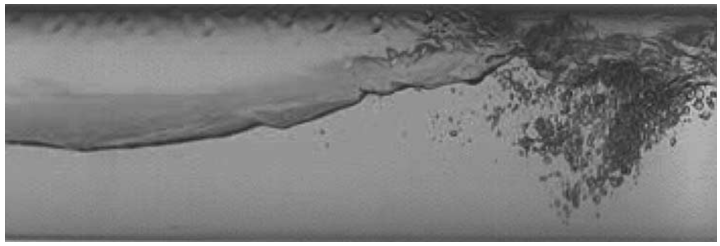

Fig. 1. Lab photo of multiphase slug flow. The picture is by courtesy of Sintef Petroleum Research.

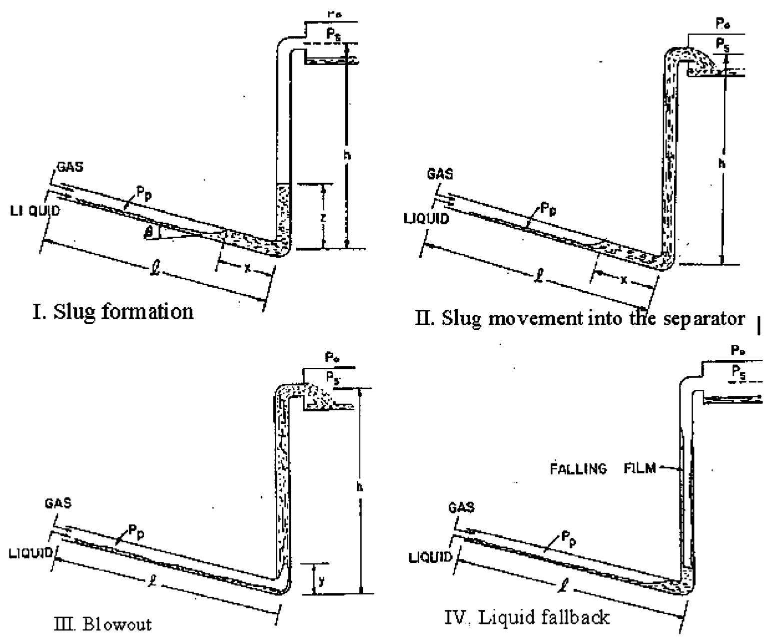

Fig. 2. Schematic view of riser slugging cycle.

some liquid may follow the gas into the compressors. A varying gas rate may also result in flaring.

There is a great economic potential if slugging can be reduced or removed completely. The regularity will be improved with fewer shutdowns if the flow into the separator is stabilized. The most important economic factor is, however, the possibility of an enhanced oil recovery. Possible increased and accelerated production is the main motivation for the installation of slug control, assuming that there is available downstream production capacity. The conventional solution to prevent severe slugging is by static topside choking. The well (and reservoir) pressure is often reduced towards the end of the well lifetime. Topside choking will increase the flow line pressure. The well lifetime can be extended and the production can be increased if topside choking can be minimized.

The motivation for this work was to gain more knowledge on slug control. Slugging is an increasing phenomena in Statoil operated fields, and a slug control system was installed at Heidrun in 2001. A series of experiments were run to verify results from offshore and to test new methods. In this paper it will be shown that a feedback control strategy makes it possible to operate at stable production within a region that is unstable with manual choking, for example, with a lower flow line pressure. A lower flow line pressure makes it possible to produce more. The paper is organized as follows. In Section 2 slug suppression methods are reviewed, and in particular those using active feedback control. The effect of topside choking is also explained. The control methods applied in the experiments are discussed in Section 3, while the experimental results are given in Section 4.

2. Slug suppression

Slugging can be reduced in many ways. If slug flow is anticipated, then a possible solution is to redesign the process equipment to reduce the probability of getting slug flow. If slug flow was not expected during design, then a solution is to make devices that can handle the slugs well. These approaches are based on stationary conditions and will in general be expensive and not optimal. They can also be sensitive to changes in the rate of production and varying flow line pressure.

One expensive and space demanding solution to handle slugs offshore is to build a sufficiently large inlet separator that will not trip or flood during severe slugging. A similar solution onshore, which is also very expensive, is to have a big tank upstream the separator, a Slug Catcher. The flow velocity can be increased if the diameter of the flow line is reduced. Increased velocity means increased flow line pressure drop. However, this will also reduce the flow line capacity. A gas lift at the foot of the riser will increase the gas to oil ratio (GOR) and may reduce slugging. The gas lift is also an expensive solution, since it will require compression of gas and an extra pipe to transport the gas back from the platform to the riser foot or the well. The easiest and most applied solution is simply to reduce the topside choke opening until slugging is reduced to an acceptable level and produce with this static choke opening.

Experiments with slug control were run in small scale already in 1979 [2] and in a larger scale in the late 1980s [3] using a topside choke actively to control the pressure at the riser foot in a two-phase flow line with a simple feedback PI control scheme. Several companies have developed similar solutions for slug control. Shell has a patented slug control system [4] where a mini separator with flow control is used for stabilization. ABB [5] and Total [6] have similar slug control systems. The first Statoil slug control installation [7] was completed in April 2001 at the Heidrun oil platform in the Norwegian Sea. Here two satellite fields from the Heidrun Northern Flank are connected to the Heidrun oil platform with 4 km and 7 km long multiphase flow lines. Severe slugging in the riser was experienced and a slug control system was installed to suppress this. The slug control suppressed the slugging and stabilized the flow, and the flow line pressure was reduced significantly. This slug control system has been operating successfully since the beginning.

A lot of work has been spent in the development of the multiphase flow simulator OLGA. A simplified nonlinear model more useful for control has been implemented in MatLab [8]. The slugging observed in the experiments described in this paper, has been reproduced in OLGA simulations [9], both with an open choke and with active slug control.

The physical mechanism for riser slugging is relatively simple and may be described by simple terms. However, slug suppression is more complex to describe. Let it suffice to say that topside choking helps eliminate riser slugging by increasing the pressure drop across slugs. This pressure drop is given by the pressure difference between the foot of the riser and downstream the choke, typically the inlet separator. An increased pressure drop over the choke counteracts slug acceleration and keeps the slug in the riser until a more stable bubbly gas and liquid flow is established. Static choking reduces the production rate. Dynamic choking by active slug control is therefore a better solution. It is then possible to avoid slugging with a choke opening that is higher on the average than what is possible with static choking. With dynamic choking it is possible to attain a relatively constant flow through the choke. The active slug control system will open the valve to increase the flow if the flow is reduced. In the opposite case, when the flow is increased, the choke opening will be reduced. The control system will either seek a specific flow rate with flow control or a desired pressure with pressure control.

3. Slug control methods, modelling and tuning

A successful slug control system will suppress riser slugging and can, at the same time, be able to operate a stable production with a minimum flow line pressure. It is also important to handle start-up and shutdown of wells in a robust way, and to be robust towards varying rates.

In slug control, the production choke is used to control the flow and stabilize the flow line pressure. Several slug control structures have been proposed and tested experimentally. Fig. 3 shows the control structure that has been implemented offshore at Heidrun [7]. This controller was the background for our experiments. It can run in four different modes as shown in Table 1.

3.1. Volumetric flow PI control

The idea here is to stabilize the volumetric flow through the choke. A multiphase flow meter will provide the necessary measurement, if available. However, such measurements are in many cases not available. It is then possible to use density measurements from a densitometer together with differential pressure measurements for the choke and the choke position to estimate the volumetric flow through the choke. It is possible to have

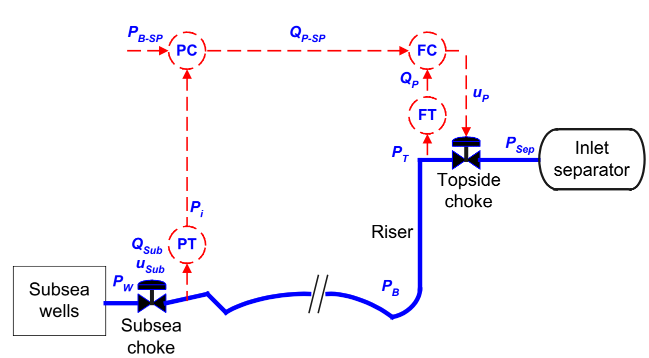

Fig. 3. Cascade control of flow line pressure (P_i) and volumetric flow (Q_P).

Table 1
Controller modes for Heidrun slug control system

	Set point (SP)	Controlled variable (CV)	Manipulated variable (MV)
Manual control	–	–	Choke position SP
Flow control	Operator	Volumetric flow	Choke position SP
Pressure control	Operator	Flow line pressure	Choke position SP
Cascade control			
Slave	Master MV	Volumetric flow	Choke position SP
Master	Operator	Flow line pressure	Volumetric flow SP

tighter control with faster response to variations, since there is no time delay on the measurement. Faster control makes it possible to suppress shorter slugs. The difficulty with this approach is to find a set point for the flow controller. The valve will open 100% and the flow line will induce slugging, if the set point is set too high. A low set point will increase the flow line pressure and reduce the production more than necessary. This set point can, however, be used to choose the desired production rate directly. If e.g. a new well is opened for production, then the flow set point must be increased to allow increased production. A standard PI volumetric flow controller was applied:

$$\frac{u_P}{Q_P - Q_{P-SP}} = K_C\left(1 + \frac{1}{T_i s}\right) \tag{1}$$

The purpose of the flow controller is to even out the flow through the choke by linearising the valve equation with choke dynamics. A good choice for controller parameters is to let the integral time match the choke time constant ($T_i = T_v$) and a controller gain $K_C = \frac{a}{K_P}\sqrt{\frac{\rho_P}{\Delta P_C}}$, where a is a tuning parameter typically between 0.3 and 2.0. The resulting closed loop volumetric flow dynamics is then

$$\frac{Q_P}{Q_{P-SP}} = \frac{1}{1 + \frac{T_v}{a}s} \tag{2}$$

Perfect volumetric flow control ($Q_P = Q_{P-SP}$) will be assumed in the rest of this section.

An asymmetric effect was found during the experiments. A small increase in the choke opening during stable flow results in a much faster pressure reduction in the foot of the riser than the pressure increase observed with a similar small decrease in the choke opening. A decreased choke opening increases the amount of liquid in the riser, due to increased slip between the gas and the liquid. The pressure build-up is limited by the liquid rate into the riser, and therefore slow (see the dashed line in Fig. 4). An increased choke opening reduces the amount of liquid in the riser. The liquid is replaced by compressed gas from the flow line that will expand into the riser. This event is therefore fast (see the solid line in Fig. 4). A fast flow controller is the conventional solution to linearize a choke. A flow controller will reduce asymmetric responses.

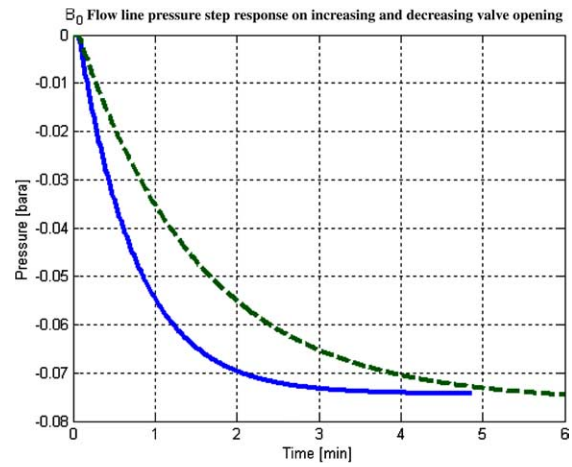

Fig. 4. Asymmetric flow line pressure response with 1% increased (solid) and 1% reduced (dashed) choke. Note that the response for the reduced valve opening is inverted in the figure.

3.2. Flow line pressure PI control

The idea with this control structure is to keep a stable pressure in the flow line at the seabed. This solution requires a subsea pressure sensor and on-line communication with the topside process control system. The controller must be relatively slow, when the distance between the subsea pressure measurement and the control input is long, and the update rate is low. Experience, e.g. from Heidrun [7], ABB [5], and simulations [8] have shown that this control mechanism can be used to suppress riser slugs effectively.

3.3. Flow line pressure and volumetric flow cascade PI control

Here a slow outer control loop maintains a stable pressure at the inlet of the flow line by using a PI pressure controller to provide a set point to the inner flow control loop (Fig. 3). A pressure measurement at the seabed is necessary for this solution. A fast inner control loop will provide the desired flow through the choke using a PI flow controller to provide a set point for the choke.

3.3.1. Simple physical model of flow line pressure dynamics

A local linearized first order mathematical model valid when the pressure at the foot of the riser is close to its set point P_{B-SP} is developed in Appendix A

$$\frac{P_B - P_{B-SP}}{Q_P} = \frac{-K_B}{1 + T_B s}. \tag{3}$$

Inlet choke characteristics are required to compute the K_B and T_B.

3.3.2. Model of flow line pressure dynamics with active slug control

Here perfect flow control is assumed and the simplified dynamic model (3) is applied. The desired volumetric flow is given by the output of the PI pressure controller

$$\frac{Q_{P-SP}}{P_B - P_{B-SP}} = K_C\left(1 + \frac{1}{T_i s}\right) \tag{4}$$

and a first order open loop model (3) gives the following closed loop transfer function

$$\frac{P_B}{P_{B-SP}} = \frac{1 + T_i s}{1 + \left(1 + \frac{1}{K_C K_B}\right)T_i s + \frac{T_i T_B}{K_C K_B} s^2} \tag{5}$$

The time constant T_C and damping ξ_C for the second order closed loop transfer function (5) are

$$T_C = \sqrt{\frac{T_i T_B}{K_C K_B}}$$

and

$$\xi_C = \frac{1}{2}\left(1 + \frac{1}{K_C K_B}\right)\sqrt{\frac{K_C K_B T_i}{T_B}} \tag{6}$$

Solving (6) for the controller parameters T_i and K_C gives

$$K_C = \frac{2\xi_C T_B - T_C}{K_B T_C} \quad \text{and} \quad T_i = T_C\left(2\xi_C - \frac{T_C}{T_B}\right) \tag{7}$$

A proposed set of tuning parameters is, for example,

$$K_C = \frac{3}{K_B} \quad \text{and} \quad T_i = T_B \text{ giving}$$

$$T_C = \frac{T_B}{\sqrt{3}} \quad \text{and} \quad \xi_C = \frac{2}{\sqrt{3}} \tag{8}$$

It must be noted that the models developed in this paper are only valid locally, i.e. for small pressure variations at steady flow. No attempt has been made to model multiphase flow, and the models developed in this paper cannot be used to predict slugging. The purpose of the mathematical modelling was to provide tools for controller tuning, i.e. to find relations between controller parameters, stability and performance. Experience gained during the experiments has shown that the models are sufficiently accurate for controller tuning.

3.4. Top pressure and volumetric flow cascade PI control

Subsea pressure measurements are expensive to install and maintain and sometimes less available and less reliable than topside measurements. Topside measurements are usually updated more often and they are in many cases more accurate than subsea measurements. One goal with the experiments was to develop and test a control structure depending on topside measurements only. Such a solution will be a good backup alternative for slug control, for example, if the flow line pressure transmitter is not working or is unavailable for some reason for a period of time. Slug control with topside measurements only should also be considered for installations where the distance from the platform to the flow line pressure measurement is very long, or, if there is no subsea flow line pressure measurement available online for control. Pressure variations at the inlet of a very long flow line will be out of phase with the pressure at the foot of the riser.

Fig. 5 shows a new cascade control solution using the top pressure (upstream the control choke) in a slow outer loop and the flow through the choke in an inner loop. This control structure has, to the knowledge of the authors, not been published before. It has not been implemented at Heidrun, and is therefore not mentioned in Table 1. The top pressure behaves differently from the flow line pressure at the riser foot. The top pressure is given by the sum of the pressure downstream the choke (separator pressure) and the differential pressure over the choke. The pressure at the riser foot is given by the top pressure plus hydrostatic pressure given by the weight of the contents in the riser, friction loss and pressure drop due to acceleration in the riser. Friction and acceleration will be neglected in the following. From observations it is clear that the top pressure shows some sort of non-minimum phase behaviour, since it has a temporary response in the opposite direction of the stationary response. It is a well-known fact in control theory that the bandwidth of non-minimum phase systems is upper bounded. This suggests a large integration time and low gain for the PID controller.

The differential pressure over the choke, that is the difference between the top pressure and the separator pressure, is a direct measure for both controllability and robustness. Controllability and robustness are achieved with a high differential pressure, while good

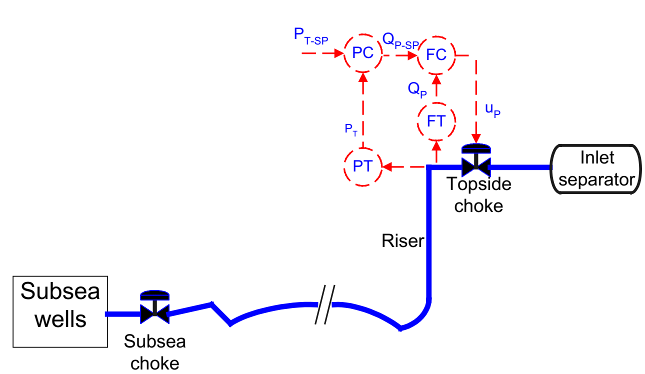

Fig. 5. Top pressure and volumetric flow cascade control.

performance with maximum production is achieved with a low differential pressure.

4. Experimental results

The objectives for the experiments were to verify results obtained offshore at Heidrun, test new methods and to develop simple tuning rules for later offshore installations.

4.1. Description of experimental test loop

An experimental set-up at the SINTEF Petroleum Research Multiphase Flow Laboratory [10] at Tiller outside Trondheim was used to develop and test different control strategies. The tests took place in June and August 2002. The heavy gas SF6 and the liquid Exxsol D80 were mixed in a 231 m long closed loop with 2.5″ inner diameter pipes equipped with a vertical riser (see Figs. 6 and 7) for circulation of oil and gas. The first 100 m have a −0.1° declination. After a horizontal U-turn, the pipe is declined −0.7° for about 100 m before the 15 m vertical riser with a control valve on top. A vertical 8″ drop leg ends in a gas–liquid separator, where the gas is drawn into a de-mister to remove droplets and then into the compressor. The oil is drained to a horizontal separator, and recycled through an oil pump. The gas is mixed with the oil through a 45° downward inclined pipe at the inlet section. The oil and gas then pass through a 7 m long flexible (rubber) pipe section, and into the initial −0.1° section. Temperature sensors, pressure cells and gamma densitometers are distributed along the loop.

Fig. 6. The 15 m high riser at Tiller with the control choke at the top.

The following set of input signals is available for control: choke differential pressure, densities at the top, middle and foot of the riser, pressures at the top and the foot of the riser and the measured choke position. The volumetric flow rate Q_P through the choke is not directly measured, but in this paper it will be assumed that it can be estimated with sufficient accuracy from available measurements according to this simplified valve equation:

$$Q_P = K_P u_P \sqrt{\frac{\Delta P_C}{\rho_P}}, \tag{9}$$

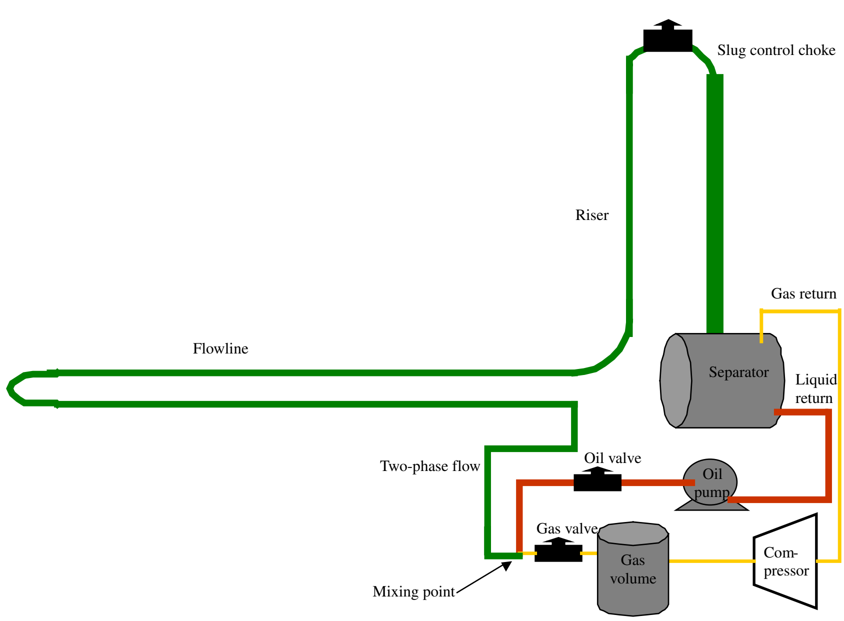

Fig. 7. Schematic overview of Tiller medium scale multiphase experimental loop.

where u_P is the actual choke position, K_P is a choke constant, ΔP_C is the pressure drop over the choke and ρ_P the density of the fluid flowing through the choke. The control output signal is the desired choke position u_C given by the choke dynamics

$$u_P = \frac{1}{1 + T_v s} u_C, \tag{10}$$

where T_v is the time constant and s is the Laplace operator. A nomenclature list is given in Table 2. A linear valve characteristic has been assumed.

Table 2
Nomenclature

A	Pipe cross-section area [m^2]
CV	Controlled variable
FC	Flow controller
FT	Flow transmitter
GOR	Gas–Oil ratio [Sm^3/Sm^3]
g	Gravity [9.82 m/s^2]
H	Hold-up (liquid volumetric fraction)
K_B	Model gain [bar/h m^3]
K_C	Gain [% s m^3] or [m^3/h bar]
K_P	Choke constant [m^2/%]
K_T	Model gain top pressure [bar/h m^3]
K_{Sub}	Subsea choke constant [m^2/%]
MV	Measured variable
m_{riser}	Mass of riser fluid in riser [kg]
P_B	Flow line pressure at riser foot [bara]
P_{B-SP}	Flow line pressure set point [bara]
PC	Pressure controller
P_i	Flow line pressure at inlet [bara]
PID	Proportional, integral and derivative controller
P_{Sep}	Separator pressure [bara]
P_T	Top pressure [bara]
PT	Pressure transmitter
P_W	Well pressure [bara]
Q_B	Riser foot volumetric flow [m^3/h]
Q_P	Choke volumetric flow [m^3/h]
Q_{P-SP}	Volumetric flow set point [m^3/h]
Q_{Sub}	Subsea volumetric flow [m^3/h]
s	Laplace operator [rad/s]
SP	Set point [bara] or [m^3/h]
T_B	Time constant without control [s]
T_C	Time constant with control [s]
T_i	Controller integral time [s]
T_n	Model zero time constant [s]
T_p	Model pole time constant [s]
T_v	Time constant for choke [s]
u_C	Control output [%]
u_P	Measured choke position [%]
u_{Sub}	Subsea choke position [%]
U_{sg}	Superficial gas velocity [m/s]
U_{so}	Superficial oil velocity [m/s]
w_P	Choke mass flow [kg/h]
ξ_C	Damping coefficient [ ]
κ	Subsea choke gain [m^3/h bar]
ρ_B	Fluid density at riser foot [kg/m^3]
ρ_P	Density upstream choke [kg/m^3]
ρ_{Sub}	Subsea choke fluid density [kg/m^3]
ΔP_C	Choke differential pressure [bar]
ΔQ_P	Volumetric flow step [m^3/h]

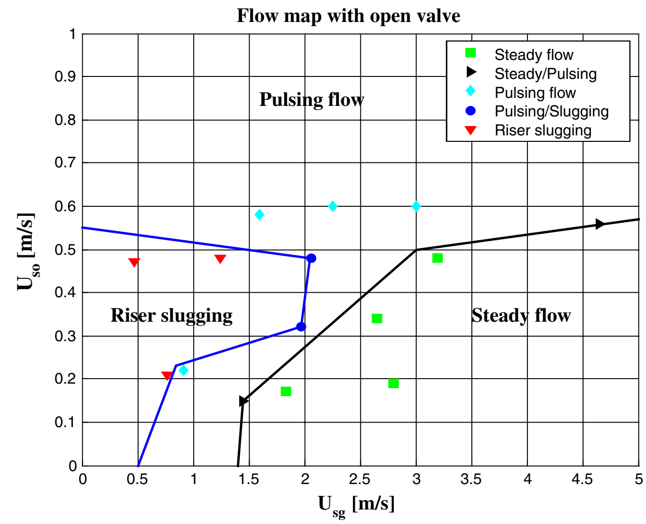

Fig. 8. Flow map with 100% open choke. The superficial gas velocity U_{sg} is plotted along the x-axis and the superficial oil velocity U_{so} is plotted along the y-axis.

4.2. Multiphase flow map

A flow map for the vertical riser is presented in Fig. 8. The different states are defined by the variation in a derived quantity called riser hold-up H (liquid volumetric fraction) defined by

$$H = \frac{P_B - P_T}{(P_B - P_T)_{liquid}} \tag{11}$$

The riser hold-up is 1 if the riser is filled with liquid and (close to) 0 if it is filled with gas. In this paper riser slugging is defined as a state where the riser is filled with liquid ($H_{max} = 1$) and less than 25% liquid ($H_{min} < 0.25$) is left in the riser after blow-out. Pulsating flow is defined as a state where $H_{max} - H_{min} > 0.25$ and steady flow as the state where the hold-up variations are less than 25%. The lines that separate the three flow regimes are drawn through points with flow conditions at the boundaries. This flow map shows for what rates riser slugging can be expected, when producing with an open choke. The gas superficial velocity U_{sg} must, for example, exceed 0.5 m/s to avoid slugging and 1.4 m/s to get the desired steady flow even for the lowest liquid rates. Note that this flow map applies only to the actual experimental conditions.

4.3. Cascade control of flow line pressure and volumetric flow

An experimental transfer function from volumetric flow Q_P to flow line pressure P_B can be modelled well as a first order linear filter for the Tiller experimental setup

$$\frac{P_B - P_{B-SP}}{Q_P} = \frac{-0.12 \text{ bar/m}^3\text{ h}}{1 + 65 \text{ s}} \tag{12}$$

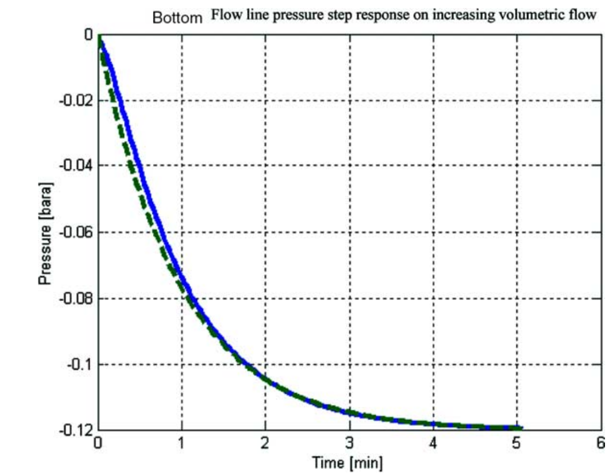

Fig. 9. Flow line pressure step response (solid) and linear approximation (dashed).

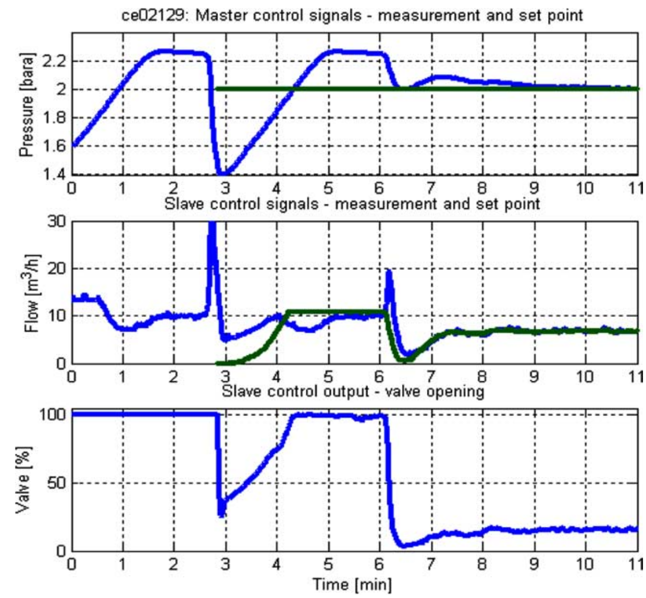

Fig. 10. Cascade PI control of flow line pressure and volumetric flow: slug suppression.

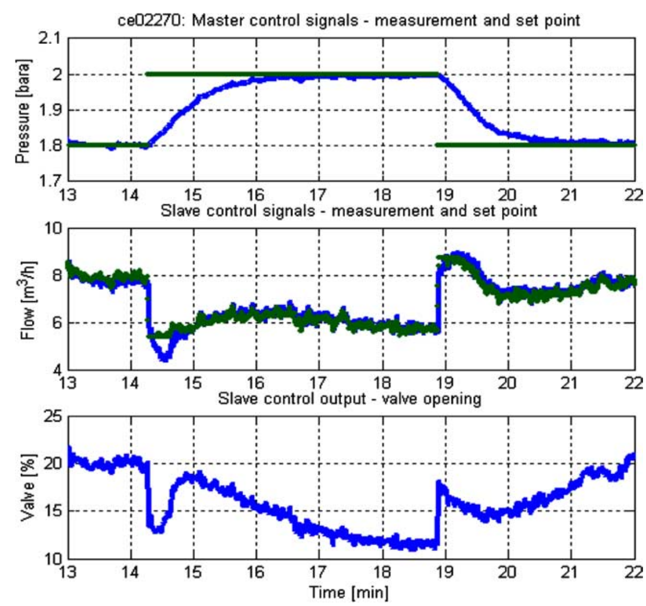

Fig. 11. Cascade PI control of flow line pressure and volumetric flow: step responses in closed loop.

(see dashed line in Fig. 9). The main difference between the physical and the experimental model is that the latter includes the response of the differential pressure across the choke. The parameters (T_C, ξ_C) in Table 3 for the simplified physical model and the experimental model using controller parameters (K_C, T_i) used in the experiments are computed using (8). The two approaches give similar results. Both can therefore be used as a starting point for controller tuning.

The experiment shown in Fig. 10 starts with a 100% open choke and slugging with a slug period of 185 s. The controller is activated after 3 min. Slugging is suppressed immediately and the flow line pressure set point is reached 2 min after blow-out. The controller performance during another test is shown in Fig. 11. Here it is shown how the controller makes the process follow the given set points. The flow line pressure set point is changed from 1.8 bara to 2.0 bara after 14.2 min. The flow line pressure reaches its set point after 2 min. The volumetric flow set point provided by the pressure controller jumps from $8\,m^3/h$ to $6\,m^3/h$ when the pressure set point is increased from 1.8 bara to 2.0 bara, and the calculated volumetric flow finds its set point rapidly.

4.4. Cascade PI control of top pressure and volumetric flow

An inverse response on the top pressure was observed compared to the flow line pressure response on steps in the volumetric flow. A small opening of the choke during stable flow gives the expected pressure fall in the foot of the riser. A temporary pressure increase was observed

Table 3
Model and controller parameters

	K_B/T_B	T_B	K_C	T_i	T_C	ξ_C
Physical model	$10.8\,bar/m^3$	Unknown	$47.1\,m^3/h\,bar$	69 s	22 s	1.5
Experimental model	$6.5\,bar/m^3$	65 s	$47.1\,m^3/h\,bar$	69 s	28 s	1.4

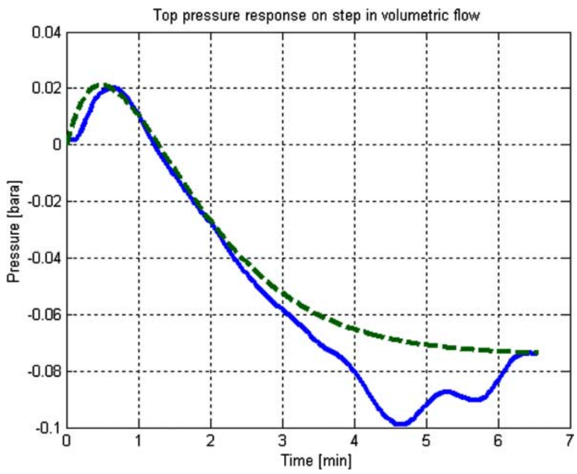

Fig. 12. Top pressure step response (solid) and linear approximation (dashed).

at the top of the riser, before the top pressure was reduced to a new lower stationary value. Similarly, a small closure of the choke gives an increased flow line pressure. There is a temporary pressure reduction at the top before the expected stationary increase. Fig. 12 shows the response and an approximate linear model. The approximate linear model is given by

$$P_{\mathrm{T}} = -K_{\mathrm{T}}\frac{1 - T_n s}{(1 + T_{\mathrm{p}}s)^2}Q_P \tag{13}$$

This transfer function has one zero in the right-half plane and has therefore non-minimum phase. The parameters are $K_{\mathrm{T}} = 0.075\,\mathrm{bar/m^3h}$, $T_n = 65\,\mathrm{s}$ and $T_{\mathrm{p}} = 55\,\mathrm{s}$.

A physical explanation of the inverse response is as follows. A decreased choke opening gives a reduced mass rate through the choke and out of the riser. Liquid will then be accumulated in the riser and give an increased flow line pressure, since there is a higher rate that flows into than out of the riser. The relative velocity between the gas and liquid phases (slip velocity) in the top of the riser increases, resulting in reduced liquid (mass) rate through the choke. A reduced mass rate ($w_P = \rho_P\, Q_P$) through the choke results in a reduced differential pressure (ΔP_{C}) across the choke. The same reduction will also be observed for the top pressure (P_{T}), since the separator pressure is controlled ($P_{\mathrm{T}} = P_{\mathrm{Sep}} + \Delta P_{\mathrm{C}}$). The first part of a pressure oscillation towards a new equilibrium consists of a liquid build-up and a flow line pressure increase. This can result in a riser slug if the change of the choke opening change is sufficiently large. Vice versa, an increased choke opening gives an increased mass rate out of the riser, reduced relative (slip) velocity between gas and liquid, initially increased differential pressure across the choke and increased top pressure, reduced amount of liquid in the riser and a reduced flow line pressure.

The top pressure controller is also a standard PI controller

$$Q_{P-SP} = K_{\mathrm{C}}\left(1 + \frac{1}{T_{\mathrm{i}}s}\right)(P_{\mathrm{T}} - P_{T-SP}) \tag{14}$$

Perfect flow control is assumed in the analysis here, and the volumetric flow is then given directly by the output of the controller ($Q_P = Q_{P-SP}$). Combining the experimental model with the controller gives the following closed loop transfer function

$$P_{\mathrm{T}} = \frac{(1 + T_{\mathrm{i}}s)(1 - T_n s)}{1 + \left(1 - \frac{T_n}{T_{\mathrm{i}}} + \frac{1}{K_{\mathrm{T}}K_{\mathrm{C}}}\right)T_{\mathrm{i}}s + \left(\frac{2T_{\mathrm{p}}}{K_{\mathrm{T}}K_{\mathrm{C}}} - T_n\right)T_{\mathrm{i}}s^2 + \frac{T_{\mathrm{i}}T_{\mathrm{p}}^2}{K_{\mathrm{T}}K_{\mathrm{C}}}s^3} \times P_{T-SP} \tag{15}$$

The closed loop transfer function is asymptotically stable if all poles are in the left half plane. This puts an upper limit on the controller gain and a lower limit on the controller integral time

$$K_{\mathrm{C}} < K_{\mathrm{C,max}} = \frac{2T_{\mathrm{p}}}{K_{\mathrm{T}}T_n}$$

and

$$T_{\mathrm{i}} > T_{\mathrm{i,min}} = \frac{T_n}{1 + \frac{1}{K_{\mathrm{T}}K_{\mathrm{C}}}} \tag{16}$$

The experiment shown in Fig. 13 starts with a 100% open choke and slugging with a slug period of 185 s. Fig. 13 shows the top pressure with set point (upper plot), and flow rate with set point (middle plot) and the choke opening (lower plot). The controller is activated after 26 min. Slugging is suppressed immediately and the set point is reached 6 min after blow-out. Master output constraints ($4\,\mathrm{m^3/h} < Q_{SP} < 9\,\mathrm{m^3/h}$) were imposed.

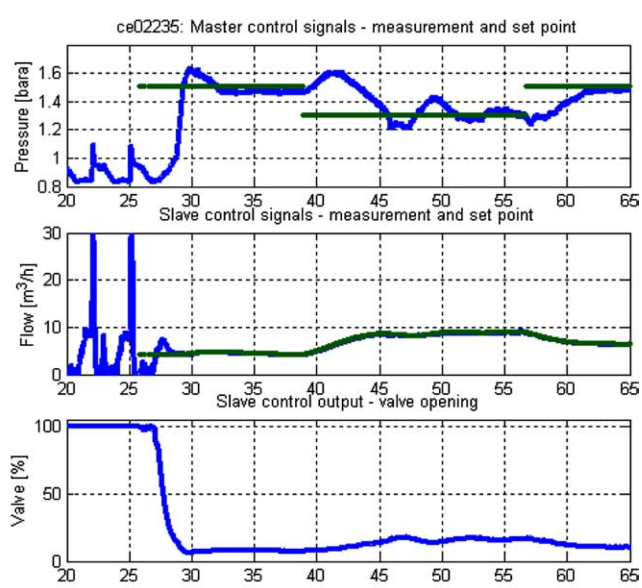

Fig. 13. Cascade PI control of top pressure and volumetric flow: pressure at the top with set point, volumetric flow with set point and choke opening.

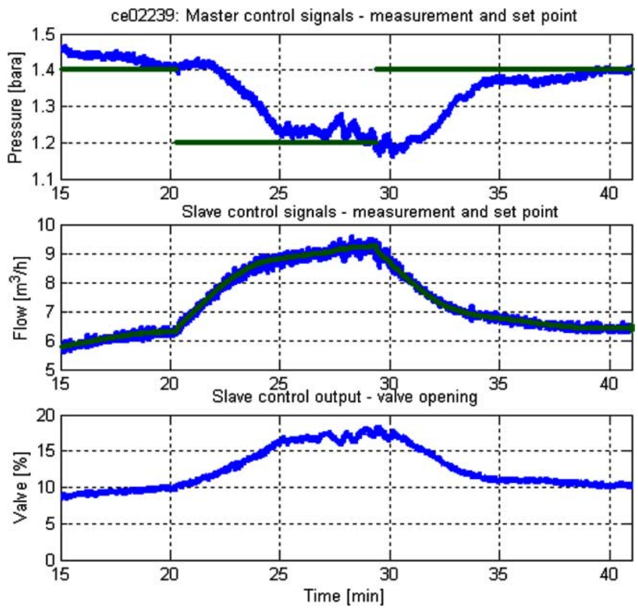

Fig. 14. Cascade PI control of top pressure and volumetric flow: step responses in closed loop.

Fig. 14 shows the step response in closed loop with this controller. It takes about 10 min to step 0.2 bar. This is much slower than the 2 min it takes with feedback from the flow line pressure (see Fig. 11). The inverse response in the top pressure restricts the bandwidth.

5. Conclusion

This paper contains simple models for slug controller tuning, simple tunings rules, a new slug control scheme independent of subsea measurements and experimental results. It has been shown by experiments how riser slugging can be suppressed by conventional feedback control. Feedback control of the flow line pressure has been verified to suppress riser slugging and stabilizes both the flow line pressure and the flow through the riser. Arguments have been given for how a volumetric flow controller can be used to improve performance by reducing slugging with higher frequency and shorter lengths. The response on flow line pressure to choke usage has been shown to be asymmetric. The response is faster when the choke opening is increased, than it is when the choke opening is reduced. These effects can effectively be reduced with a cascade controller, where a flow controller is slave for a master pressure controller. A pure inlet pressure controller is recommended because of its simplicity to avoid riser slugging for pipelines with limited length. The more advanced cascade controller is recommended if the pipeline is long, or if one wants to suppress also shorter terrain induced slugs.

A linearized dynamic physical model has been shown to be similar to an experimental step response model. Both these models can be used to tune the pressure controller to achieve the desired closed loop properties such as damping. A new controller based on topside measurements only has been proposed and tested experimentally. Here, an experimental step response model has been computed. This model is linear with a right-half plane zero and two left half plane poles. The right-half plane zero reproduces the inverse response on the upstream platform choke pressure. The model has been used to compute stability constraints for the controller parameters. The mathematical models can be used to tune slug controllers offshore based on process knowledge or measured step responses.

Appendix A. Simple model of flow line pressure dynamics at steady flow

The simplified dynamic model used in section 3.3.1 is developed in this section. A simple dynamic relation between changes in riser foot pressure and mass changes in riser is given by

$$A\dot{P}_B \approx g\dot{m}_{\text{riser}} \tag{A.1}$$

This approximation is good if the pressure at the top of the riser is almost constant and both the flow and the acceleration are low. Friction loss in pipes and choke can then be neglected, and the mass balance for the riser gives

$$\dot{m}_{\text{riser}}(t) = \rho_B Q_B - \rho_P Q_P \tag{A.2}$$

$$\dot{P}_B(t) \approx \frac{g\rho_B}{A} Q_B - \frac{g\rho_P}{A} Q_P \tag{A.3}$$

The flow through the subsea choke can be linearized around a reference pressure

$$\begin{aligned} Q_{\text{Sub}} &= K_{\text{Sub}} u_{\text{Sub}} \sqrt{\frac{P_W - P_{\text{i}}}{\rho_{\text{Sub}}}} \\ &\approx Q_{\text{Sub},0} - \kappa (P_{\text{i}} - P_{\text{i}0}), \end{aligned} \tag{A.4}$$

where P_{i} is the pressure at the inlet of the flow line downstream the subsea choke, $P_{\text{i}0}$ is a reference pressure and

$$Q_{\text{Sub},0} = K_{\text{Sub}} u_{\text{Sub}} \sqrt{\frac{P_W - P_{\text{i}0}}{\rho_{\text{Sub}}}},$$

$$\kappa = \frac{K_{\text{Sub}} u_{\text{Sub}}}{2\sqrt{(P_W - P_{\text{i}0})\rho_{\text{Sub}}}} \tag{A.5}$$

The pressure, density and rate at the inlet can be approximated to the corresponding at the foot of the riser, if the flow line is short. A short flow line is a strict requirement for the following assumptions to be valid.

$$P_i \approx P_B, \quad P_{i0} \approx P_{B-SP}$$
$$Q_{Sub} \approx Q_B, \quad Q_{Sub,0} \approx Q_{B0} \tag{A.6}$$
$$\rho_{Sub} \approx \rho_B$$

The friction loss in a long flow line is comparable to the friction loss in the subsea choke. Current experience with flow lines up to 10 km is good. If the flow line is considerably longer, then pressure and flow dynamics in the flow line must be considered in the control design. A long flow line means, for example, that there is a time delay between pressure measurements at the riser foot and at the inlet given by the length of the flow line and the speed of sound in the multiphase fluid. Note that pressure dynamics in long flow lines and the speed of sound are not accurately modelled in the multiphase dynamic simulator OLGA. The speed of sound depends on several factors like pressure, fluid composition and flow regime.

A linearized first order model valid when the flow line pressure is close to its set point P_{B-SP} is

$$\frac{P_B - P_{B-SP}}{Q_P} = \frac{-K_B}{1 + T_B s}$$
$$K_B = \frac{\rho_P}{\kappa \rho_B} \quad \text{and} \quad T_B = \frac{A}{g \kappa \rho_B} \tag{A.7}$$

T_B is called the open loop time constant. Numerical values for the model parameters can be computed if the subsea choke characteristics and the well pressure are known. Here the subsea choke parameters are not known. A simplified model of the time derivative of the pressure initially after a flow increase ΔQ_P is given by

$$\dot{P}_B \approx -\frac{K_B}{T_B} \Delta Q_P = -\frac{g \rho_P}{A} \Delta Q_P \tag{A.8}$$

For example, the calculated initial pressure reduction rate for the experiments at Tiller is $\dot{P}_B \approx -10.8$ bar/h for an increased flow rate $\Delta Q_P = 1.0\,\mathrm{m^3/h}$, with an assumed topside density $\rho_P = 400\,\mathrm{kg/m^3}$ and a pipe cross-section area $A = 0.0037\,\mathrm{m^2}$.

Comment: From (A.8) it is seen that an increased density and pressure gives faster dynamics. A changed flow rate will give a faster change in pressure with increased density.

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