Investigation of Slug Suppression System in Deepwater Scenario

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ABSTRACT

In pipeline-riser systems, pressure fluctuations which result from the formation of large liquid slugs and gas surges due to operational changes or low mass flow rate from production wells and the profile of pipeline-riser systems often lead to trips at the inlet of the separator; and thereby, the problem causes a a loss of the production.

In this study, on a sample deep-water oil field off the coast of West Africa is focused. The field lies in water depths greater than 1000 m. Moreover, the wells are connected via a pipeline-riser system to the topside. The slug suppression system (S^3) was changed as a control structure on the field case study.

S^3 comprises of a mini separator coupled with dynamically controlled valves at the liquid and gas outlets. This control structure was modeled on OLGA, a one-dimensional, and two-fluid equations based commercial multiphase flow simulation tool. In implementing the S^3, it was transformed into a parallel configuration of two proportional-integral (PI) controllers (the separator level and pressure controllers) which controls the total volumetric flow and liquid flow respectively by subsequent opening of the valves at the outlets while stabilizing the riser base pressure. In addition, separator sizing was based on the volume of multiphase fluid at the riser-top. Also, controller-tuning parameters were obtained from parametric studies with pressure and liquid level set point at 20.5 bar and 0.5 m.

Finally, it is found out that S^3 is able to stabilize the riser base pressure and flow rate at the outlet of the mini-separator. Moreover, the comparison of production rates before and after the implementation of the control structure indicated an increase of 12.5% in the production rate.

Keywords: Severe Slugging, OLGA, Pipeline-Riser, Controller, Proportional-Integral.

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INTRODUCTION

The current projected global oil demand for the 2nd quarter of 2017 is over 96 million barrels per day (mbbl/day), based on IEA 2017 Chart shown in Figure 1. Global population is also rapidly increasing, hence, energy demand will continue to increase. Although energy sources are evolving, fossil fuel (oil and gas) remains a viable means to meet transportation needs. Therefore, there is the need for optimization of production from deep-water reserves. Slugging is a major flow assurance issue with the capacity of disrupting production by as much as 50% as emphasized by Yocum in 1973 [1]. This paper is focused on a numerical investigation of the application of active slug suppression system (S^3) to a deep-water slugging scenario, in order to understand and assess the behavior of the S^3 slug mitigation technique in mitigating slugging in a typical pipeline-riser system in the deep-water scenario.

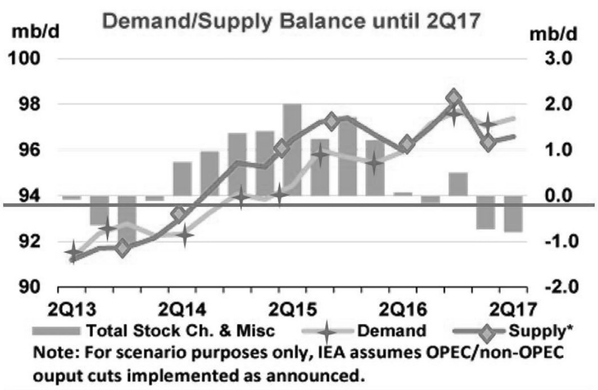

Figure 1: Oil Demand/Supply until 2Q2017 (IEA, 2017 Report).

Background on Slugging Problems in Pipeline-Riser Systems

Slugging basically involves flow rate and pressure fluctuations, and it can be classified mainly into hydrodynamic slugging and terrain induced slugging. Citing [2], severe slugging in a pipeline-riser system is an undesirable flow regime because of its potential to initiate and maintain system instability. By considering the huge variation in pressure and flow rate associated with severe slugging, its consequences in oil and gas production are a serious concern as severe slugging can lead to a drop in reservoir productivity, poor separation, overloading of compressors, platform trips and production loss [2]. Considering existing literature and field experience, severe slugging can be controlled or mitigated via mainly topsides choking and gas-lift [3,4]. Other methods have been proposed in several literature such as suppression of slug flow by active use of topside choke; relying on the process measurement of pressure, and density parameters as PID control parameters which may be difficult to control, especially density [5], deployment of wavy pipe upstream of the riser base to prevent stratified flow condition from occurring at the riser-base, which is a precursor to severe slug flow [6]; however, wavy pipe may have applicability challenge of pigging during maintenance, and recently a combination of self-lift and gas-lift techniques was also proposed by Okereke et al in 2018 [7]; however, this technique is yet to be tested in a field operation. This paper focussed on evaluating the performance of S^3 (Slug Suppression System) proposed by Kovalev et al in 2003 [8], in a deep-water scenario of over 1000 meters water depth. In October, 2002, S^3 was deployed or used in a typical shallow water field scenario in Otter field development (TotalFinaElf), in the North Sea, where S^3 was installed in the North Cormorant platform (13km flow line, 12” riser diameter and 18 bara separator pressure) with a 189 m export riser and the results indicated a flexible and robust control of the liquid and gas outflow based on the S^3 control philosophy (Kovalev, Cruickshank and Purvis, 2003) [8]. In January, 2003, the S^3 was also deployed on the Brent Charlie platform in the North Sea and was used for the Penguins development (64 km flow line, 14" riser diameter and 35 bara separator pressure and 140m water depth) and it also showed good results of flexible and robust control of pressure, liquid and gas volumetric flow [8,10]. As highlighted earlier, this paper is therefore focused on investigating the pressure, liquid and gas volumetric flow behavior when S^3 is deployed in a typical deep-water scenario of a field operating at over 1000 m water depth.

EXPERIMENTAL PROCEDURES

Description of Slug Suppression System and Basis for S^3

In a typical pipeline-riser system, operators typically want to achieve a scenario where the gas-liquid phases arrive at the inlets of the separator in a stable manner. However, in most cases, the gas-liquid phases flow in an unstable manner predominantly as a result of the low mass flow rate of the gas-liquid phases from the inlets of wells connected to the pipeline-riser systems coupled with the configuration of the pipeline-riser systems with a change in elevation between the pipeline and riser section at the riser-base. In order to restore stability in the gas-liquid phase, S^3 (Slug suppression system) has been recommended by Kovalev et al in 2003 [8].

Slug suppression system (S^3) involves volumetric flow control of the gas and liquid flow arriving in a mini-separator positioned upstream of the inlet of the production separator to control the flow of liquid and gas at certain set points, in order to prevent chaotic/irregular flow of liquid and gas phases into the inlets of the separator.

In Figure 2, an illustration of the S^3 deployed between the outlet of the pipeline-riser and the inlet of the separator is shown. During the occurrence of slugging, the liquid and gas holdup of a two-phase flow often fluctuated at high frequency especially over time from the onset of the slugging. Following on, the need for a system that can effectively control the total volumetric flow of the gas-liquid phase with the support of a single control valve is difficult. The idea of the S^3 is a system that acts like a control valve, providing control for the liquid level and gas pressure via feedback mechanism set to regulate the liquid level and gas pressure at a certain set-point to avoid a chaotic fluctuation of the liquid and gas phase within the inlet of the separator.

The S^3 works with a PID (Proportional-Integral-Derivative) kind of system in a mini-separator structure, either to provide control for total volumetric flow or liquid flow control. In a scenario where the S^3 is operating in the total volumetric flow mode, the liquid is controlled to maintain a certain level of set point at the liquid outlet, while the gas pressure feedback at the outlet of the gas is used to control the gas at certain set-point.

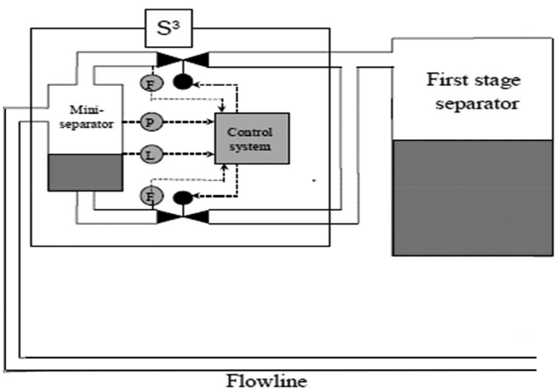

Figure 2: An illustration of the S^3 deployed between the pipeline-riser outlet and a first stage separator [8].

It is important to note that the flow rates are measured by the flow meters in the gas and liquid outlet regions. In the total volumetric mode scenario, the sum total of the feedback from the gas and liquid phase flow meters is the variable to be controlled. The set-point of the total volumetric flow is adjusted by a pressure controller in combination with other algorithms. These adjustments depend on the actual pressure and the set-point of the pressure in the mini-separator and other factors such as the diameter of the pipeline riser system.

However, in the liquid flow control mode, the liquid level is the variable to be controlled, based on the feedback from the liquid phase flow meter to a certain level of set-point.

Recent studies on the application of active slug control (smart choke system) on a sample offshore field operating at an average water depth of 1000 m is shown in Figure 3. In the case-study field, smart choke system was deployed to mitigate the occurrence of slugging at flow rate range of less than 20,000 to 30,000 BLPD (Barrel of Liquid Per Day) as highlighted in Lacy et al 2014 [9].

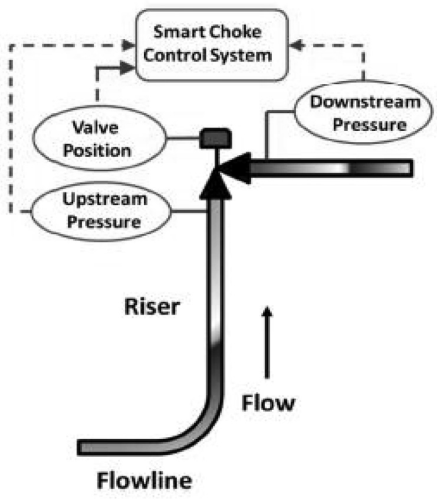

Figure 3: Smart Choke System for Slug Suppression [9].

The smart choke system as highlighted in Figure 3 is based mainly on controlling the upstream pressure relative to the riser top section as well as the downstream pressure relative to the riser-top section. Based on the results shown in Figure 4 (a), the smart choke system combined with gas-lift performed better than the gas-lift combined with fixed choke, considering the increased valve opening highlighted in the black line and the smooth decrease in pressure upstream of the valve highlighted in the grey trend when the smart choke was combined with gas lift. As shown in Figure 4 (b), the gas-lift combined with a fixed choke slug mitigation strategy showed fluctuation in the upstream pressure highlighted in grey pressure trend and fluctuations in valve opening highlighted in black spikes, which could still lead to trips on the inlets of the separator.

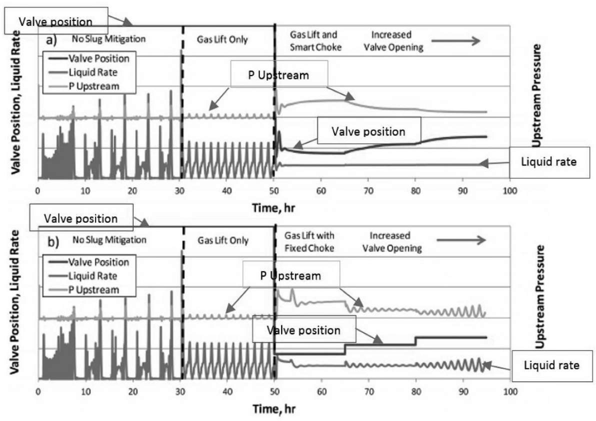

Figure 4: Smart Choke System Combined with Gas lift.

Based on [9], the application of a smart choke system succeeded in lowering the minimum liquid volume at which slugging was suppressed, reduced pressure fluctuation within the compression train, and finally reduced the gas-lift requirement for slug control/suppression.

The operational strategy of the smart choke system is based mainly on the management of pressure upstream and downstream of the riser-top as compared to the strategy of S^3 which is based both on the liquid volume control and pressure control in the form of a mini-separator, preventing pressure build-up within the inlets of the separator.

In summary, the recent study of deploying a smart choke system in slug mitigation performed better when combined with gas-lift. However, considering the extra cost associated with gas-lift, S^3 appears a better approach, hence this current work focused on modeling the deployment of S^3 in a deep-water scenario.

Field Case Study-Description

In this study, the focus was on a sample deep-water oil field off the coast of West-Africa. The field lies in a water depth of about 1447.8 m [10]. It consists of twenty production wells centered on six drilling center manifolds. The production wells are tied to an FPSO (Floating Production Storage and Offloading) vessel by eight production risers [10]. Currently, sixteen of the production wells have been drilled and are in production. The field currently produces over 200,000 BoPD [10]. Table 1 shows a highlight of the total vertical depth, pressure and temperature at the core points associated with Pipeline-Riser X1. In Pipeline-Riser X1, wells X1 and X2 are connected through the pipeline-riser system to the topsides.

Table 1: Total Vertical Depth, Pressure and Temperature at Core Points of Pipeline-Riser X1.

Station	X1		
	TVD (m)	**Pressure (Pa)**	**Temperature (°C)**
Separator	49.99 (164 ft)	1.99×10^{6} (290 psi)	65.56
Manifold	-1463.04 (-4800 ft)	8.96×10^{6} (1300 psi)	75.56
Wellhead	-1447.8 (-4750 ft)	1.16×10^{7} (1678 psi)	82.22
Sandface	-3916.68 (-12850 ft)	2.37×10^{7} (3444 psi)	100.56

Wells X1 and X2 are connected via MFX1 (Manifold X1) and identified as Pipeline-Riser X1 as highlighted in Figure 5. Pipeline-Riser X1 from the field report obtained experienced hydrodynamic slugging when it was operating at 3000 BoPD in the early life of the field.

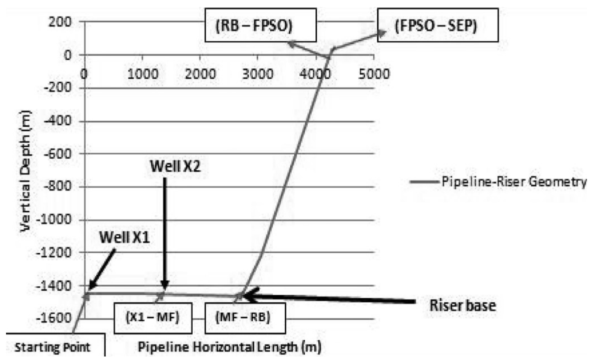

Figure 5: Geometry of Pipeline-Riser X1 System Showing the Profile from Seabed to Topside.

As a key part of this study, the initial hydrodynamic slugging scenario of Pipeline-Riser X1 was firstly modeled on OLGA (version 7.3). Subsequently, three other case scenarios were considered by moderating the flow velocity of well X1 and well X2 at different rates and in slug tracking mode, in order to generate severe slugging scenarios before S^3 was then changed to Pipeline-Riser X1 in order to investigate the viability of S^3 mitigating slugging in deep-water scenario. Key parameters considered in this study include pressure trend behavior, ID (Flow Regime Indicator) profile behavior and QLT (Total Volumetric Flow) behavior. In reaching conclusion, the QLT (Total Volumetric Flow) before and after implementation of S^3 on Pipeline-Riser X1 was considered in order to evaluate the impact of S^3 on production.

Pipeline-Riser X1 Geometry Description

In Pipeline-Riser X1, Well X1 (Source 1) is at the inlet and lies about 2712 m upstream of the riser-base. Well X2 (Source 2) lies at about 1067 m downstream of well X1 connected via the manifold. The riser height is about 1512 m, connecting to the separator. The description of Pipeline-Riser X1 is captured in Figure 5 and the OLGA model GUI (Graphical User Interface), showing wellhead representing well X1 (Source 1) and Manifold representing well X2 (Source 2) as well as Topsides representing the topsides is captured in Figure 6.

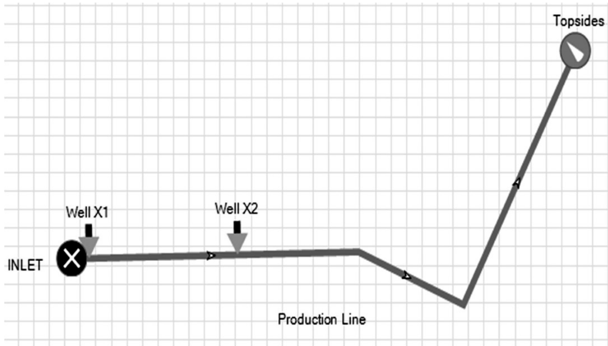

Figure 6: OLGA Model of Pipeline-Riser X1 (Not Geometrically Accurate).

The detailed sectional length analysis of Pipeline-Riser X1 system is also clearly shown in Table 2 as defined in the OLGA model of the pipeline-riser section.

Table 2: Pipeline-riser co-ordinates and section lengths for pipeline-riser X1.

Pipeline-Riser	x [m]	y [m]	Length [m]	Elevation [m]	No. of Sections	Length of Sections (m)	Diameter (m)	Roughness (m)
Starting Point	0.00	-1447.80						
Pipe-1 (X1-MF)	1066.80	-1447.80	1066.8	0.00	35	35:30.48	0.1524	0.002
Pipe-2 (MF-RB)	2712.72	-1463.04	1645.92	-15.24	54	54:30.48	0.3048	0.002
Pipe-3 (RB-FPSO)	4236.72	0.000	1524	1463.04	50	50:42.25	0.3048	0.002
Pipe-4 (FPSO-Sep)	4319.02	49.987	82.296	49.99	3	3:32.10	0.3048	0.002

Fluid Description

The fluid composition of well X1 and well X2 are as detailed in Table 3. The fluid composition is defined in PVTSim20 based on the mole percentage of each constituent that made up the well X1 and well X2 fluid.

Table 3: Fluid Properties of Field Data.

Component	Composition (Mol. %)
	Well X1 and X2
Carbon Dioxide (CO_2)	0.81
Nitrogen (N)	0.13
Methane (CH_4)	43.30
Ethane (C_2H_6)	7.49
Propane (C_3H_8)	7.29
Iso-Butane (iC_4)	2.61
N-Butane (nC_4)	3.28
Iso-Pentane (iC_5)	1.98
N-Pentane (nC_5)	1.56
Hexanes (C_6H_{14})	2.72
Heptane Plus (C_{7+})	28.83

The GOR (gas/oil ratio) was verified as 385.91 sm^3/sm^3 from the PT flash at a pressure range of minimum 1 bar and maximum 300 bar. The temperature range for the PT flash was of minimum 20 °C and maximum 120 °C as defined on PVTSim20. Fluid API was defined as API 47 degree. The fluid API suggests that the fluid is a relatively light fluid with an API 47 degree and a moderate GOR of 385.91 sm^3/sm^3. The long pipeline-riser section of over 4000 m horizontal length and the change in configuration at the riser-base has the tendency to cause multiphase fluids to experience a drop in the superficial gas velocity leading to liquid accumulation at the riser-base and possible slugging along Pipeline-Riser X1. Hence, this became a basis for the focusing of this study on researching the ability of S^3 to mitigate slugging on a typical deep-water pipeline-riser system.

Preliminary Simulation Results

In Figure 7, the field pressure profile was compared with the simulation pressure profile at 6722 BoPD for well X1 and 22,157 BoPD for well X2 condition for validation purpose. The detailed conversion of the 6722 BoPD and 22,157 BoPD volumetric flow rates to mass flow rates for well X1 and X2 are clearly highlighted in Appendix 1 and Appendix 2 respectively. The comparison between them in Figure 7 shows a variation within $\pm 20\%$ which falls within a similar range as the comparison of Leda flow (a transient slug capturing software) and OLGA with experimental results and with each other [11]. It is also important to note that the over-prediction of pressure by OLGA is similar to the trend obtained in the literature [11].

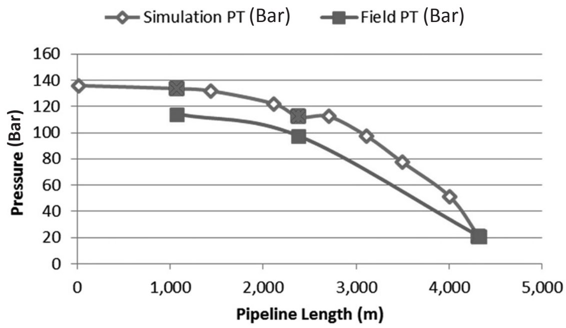

Figure 7: Field Data Vs Simulation Result Comparison (Pressure).

The similarity in pressure trend shown in Figure 7 is a basis for confidence in further simulation results of the field case study.

The initial hydrodynamic slugging scenario obtained at 3000 BoPD was firstly modeled on OLGA 7.3 by converting the volumetric flow rate to mass flow rate at both well X1 and well X2. The corresponding mass flow rates for well X1 and X2 are modeled as (8.745 kg/s and 25.13 kg/s) and run to an end time of 24 hours. The hydrodynamic slugging behavior is confirmed with cyclic pressure fluctuation between 58.70 bar and 59.25 bar as shown in Figure 8.

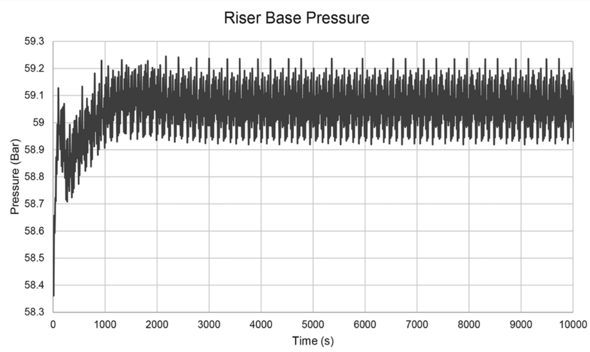

Figure 8: Hydrodynamic slug condition on Pipeline-Riser X1 at 3000 BoPD.

In subsequent simulation studies, the initial condition was modified by lowering the mass flow rate until the model exhibited severe slugging behavior with an increased pressure fluctuation and liquid accumulation at the riser-base. The S^3

technique was then changed to the Pipeline-Riser X1 case to assess its ability to suppress slug in a typical deep-water scenario.

Slugging Scenario Conditions (Case Scenarios 1, 2, and 3)

Case scenario 1 (Source 1 reducing with source 2 shut off)

Initially the flow rate at source 1 (Well X1) was varied at 8.745 kg/s, 7 kg/s, 6 kg/s and 5 kg/s respectively to tune the Pipeline-Riser X1 model to severe slugging scenario. Pressure fluctuation is one of the most critical parameter for assessing typical pipeline-riser system flow instability behavior. The pressure trend (PT) was assessed for case scenario 1 and the results show a high level of pressure fluctuation at the riser-base, with pressure fluctuating between 39 bar to over 120 bar as highlighted in Figure 9. This range of pressure fluctuation is capable of causing trips on the inlets of the separator as the multiphase fluid arrives at the inlets of the separator at a potential high-high level.

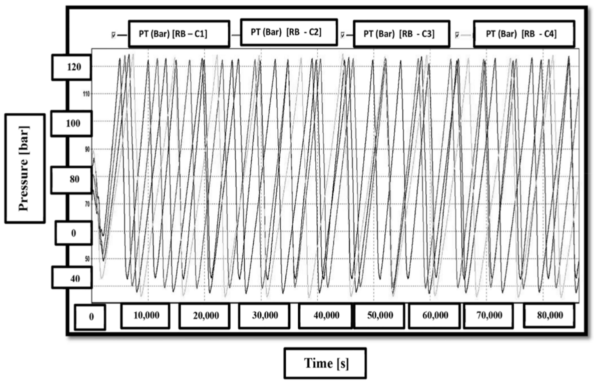

Figure 9: Plot of riser-base pressure at Case Scenario 1 (all curves fluctuate the same as each other).

Case Scenario 2 (Source 1 decreasing with Source 2 constant)

Further simulation was carried out to study the slugging behavior when source 1 is reducing while source 2 is kept constant, the flow rate at source_1 (Well X1) was varied at 8.745kg/s, 7kg/s, 6kg/s, and 5kg/s while source 2 (Well X1) was kept constant at 56.128kg/s to represent the 22157 BoPD production rate of Well X2 commingled at the manifold. Total Pressure (PT), Flow regime ID, Liquid Holdup (HOL), and Total liquid flow rate (QLT) are taken into consideration and observed. For the case scenario 2 considered, Figure 10 also shows a high level of pressure fluctuation between 84 bar and 106 bar. However, it is important to note that for the scenario with Well X1 at 6 kg/s (condition 3-C3) highlighted in blue, the fluctuation stabilized from about 6000 s. Also, for the scenario with Well X1 at 7 kg/s (condition 2-C2) reflected in red and 5 kg/s (condition 4-C4) reflected in green both cases stabilized at about 68,000 s and 75,000 s respectively.

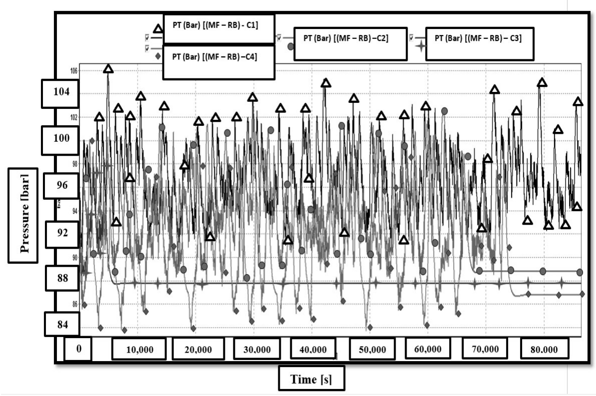

Figure 10: Plot of pressure trend between the manifold to riser base at Case Scenario 2 (The line which is straight is related to the "PT (Bar) [(MF-RB) – C3]".

Figure 11 suggests a fluctuation in flow regime from the stratified flow, through slugging regime to bubble flow regime. The major region of the change in flow regime was around the riser-base, which is attributed to the sharp change in elevation around the riser-base.

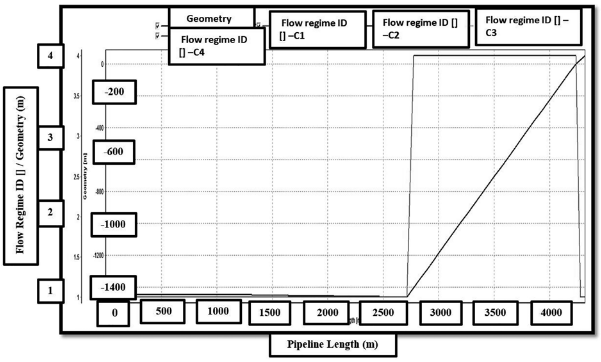

Figure 11: Plot of the flow regime at Case Scenario 2 (The four (4) conditions considered in case scenario 2 were within a close range in terms of mass flow rate of well X1 (8.745 kg/s, 7 kg/s, 6 kg/s and 5kg/s). Hence, the similarity in flow regime behavior has been observed, and there is considerable overlap between the results.

Figure 12 also shows fluctuation in QLT trend between 7,500 m^3/day to about 11,500 m^3/day, this is attributed to the extra flow coming from well X2 which was initially shut-off in case scenario 1. Terrain slugging is basically not witnessed as a result of an increase in flow rate from the commingled well X2 as shown in Figure 10 with flow stabilizing at 7 kg/s, 6 kg/s, and 5 kg/s scenarios respectively. Also, from Figure 11, the pipeline-riser section was not stable at flow regime ID-3, indicating that Pipeline-Riser X1 was not under severe slugging condition at case scenario 2.

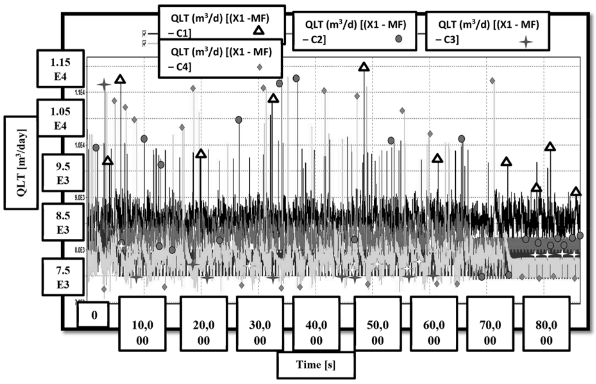

Figure 12: Plot of the volumetric flow rate at Case Scenario 2.

Case Scenario 3 (Both Source 1 and Source 2 Reducing)

Finally, a scenario with a reduction of the mass flowrates of both sources 1 and 2 was simulated. Source_1 was gradually reduced from 8.745 kg/s through 5 kg/s, and source_2 was reduced from 25 kg/s through 10 kg/s. With reference to [12], the worst kind of terrain-induced slugging is severe slugging caused by an abrupt change from the horizontal to vertical flow directions. Severe slugging is frequently seen in the risers. This typically occurs when both gas and liquid flow rates are relatively low. Severe slugging was observed in the case 3 scenario, with worst fluctuation being the scenario with 5 kg/s as captured in the green fluctuation in Figure 13. Subsequently, S^3 was adapted to the Pipeline-Riser X1 with the worst case of severe slugging recorded in case scenario 3 with source 1 at 5 kg/s and source 2 at 10 kg/s.

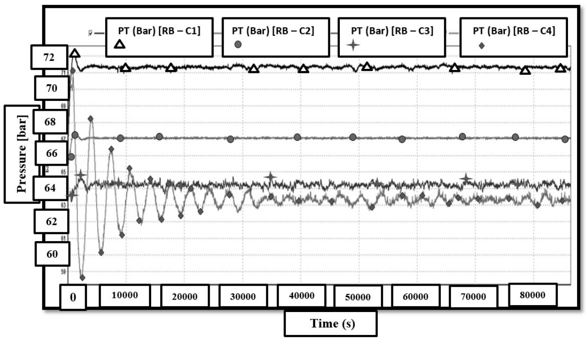

Figure 13: Plot of riser-base pressure trend at Case Scenario 3.

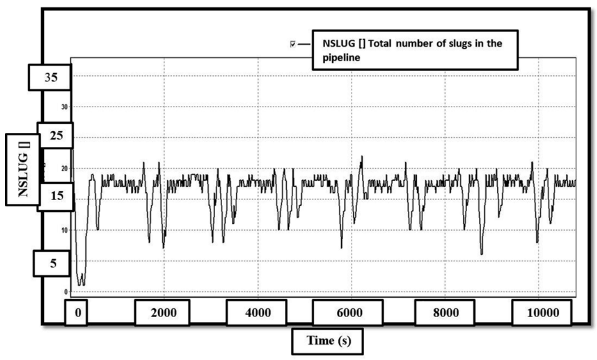

Figure 14: Slug frequency of the flow across the pipeline-riser system at Case Scenario 3.

In the slugging scenario captured in case scenario 3, it can be observed from Figure 14 that the pipeline-riser system operated in a slugging regime has an average slug cycle of 15 slugs/s. A slug suppression system using the mini-separator was deployed to mitigate the slugging scenario in this case 3 scenario with source 1 at 5kg/s and source 2 at 10kg/s.

According to [8], the control strategy of the S^3 is based on total volumetric flow control and liquid flow control. A major advantage of the S^3 is that implementation of the S^3 results in a stabilized production of gas and liquid as an approach to the ideal production system. Also one of the simplest solutions for active slug control is to use the classic PI/PID controller to stabilize the riser base pressure. The main advantages of S^3 are the ease of application and its well-proven effectiveness in the case of severe slugging mitigation in shallow water scenario. Figure 15 captured the coupling of S^3 to Pipeline-Riser X1 on OLGA.

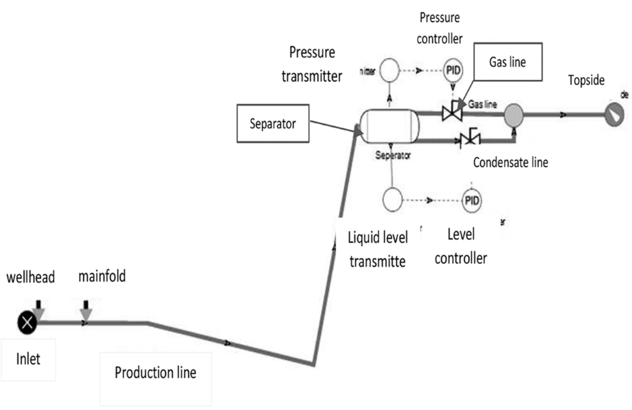

Figure 15: OLGA model of the S^3.

Separator Design

As a major step in modeling the S^3 in OLGA, the existing model was equipped with a horizontal two-phase gas-liquid separator. In order to avoid the complexity of modeling on OLGA, a simplified dimension of the separator was changed as stated below:

Diameter of the separator (d_{sep}) = 1.5m

The height of the separator (H_{sep}) = 3m

The volume of the separator can be calculated as follows (Equation 1):

$$V_{sep} = \pi \times \left(\frac{d_{sep}}{2}\right)^2 \times h_{sep} \quad (1)$$

$$V_{sep} = \pi \times \left(\frac{1.5m}{2}\right)^2 \times 3.0m \approx 5.30m^3$$

Table 4: Sizing and weight calculation of the S^3 unit for Pipeline-Riser X1 in comparison to the Otter and Penguins project [13].

	Otter	Penguins	X1
Pipeline Diameter	30 cm	40 cm	20 cm
Gas production	0.30 m^3/s	1.30 m^3/s	0.348 m^3/s
Oil/water production	0.08 m^3/s	0.10 m^3/s	0.026 m^3/s
Conventional S^3			
Vessel Height	3.00 m	3.50 m	3.00 m
Vessel diameter	1.30 m	2.00 m	1.50 m
Vessel volume	3.98 m^3	11.00 m^3	5.30 m^3
System weight	16.65 t	26.02 t	20.42 t

The initial liquid level in the separator was set at 0.5 m, and the efficiency of separation was to 100%. Liquid and gas outlets were attached to the separator, and controller valves were installed on both outlets to control the flow (Table 5). For the preliminary study, two pressure nodes were added for both gas and liquid outlets of the separator at the desired conditions, configuration of the gas and liquid outlets as shown in Table 6.

Table 5: Mini-separator vessel construction information.

Chemical Engineering Index	252.00
Material Type	Carbon Steel
Mass Density (kg/m^3)	7861.08
FMC	1.00
Allowable Stress (kg/(m.s^2))	94458.20
Shell Thickness (mm)	100.01
Corrosion Allowance (mm)	3.18
Efficiency of Joints	1.00

Gas and liquid outlet were later commingled so as to reach the first stage separator at the desired temperature and pressure. Moreover, in order to model the control scheme, two controller systems were added: a level controller system (Level Transmitter (LT) and PID controller) to prevent the separator from being ran empty or flooded and a pressure controller system (Pressure Transmitter (PT) and PID controller) installed, the transmitters were configured to output the absolute liquid level and pressure signals from the separator with the use of a bar unit scale. The PID block was added once the control variable was configured and was connected to the choke valves located at the separator outlets. Figure 15 shows an overview of the control system.

Table 6: Configuration of the S^3 liquid and gas outlets.

Parameter (keyword)	Gas Outlet	Liquid Outlet
TYPE	MASS	MASS
GASFRACEQ	1.0	0.0
OILFRACEQ	0.0	1.0
LIQUIDFRACEQ	0.0	1.0
PRESSURE [bar]	20.5	20.7
TEMPERATURE [C]	65.5	65.5

Controller Tuning

Controller tuning is the process of selecting the controller parameters to achieve given performance specification.

The PID controller used in the OLGA model is described by the following equation (Equation 2):

$$u = k_c\left(e + \frac{1}{T_i}\int_{t0}^{t} e\,dt + Td\,\frac{de}{dt}\right) + bias \qquad (2)$$

where u is the output of the controller, e is the calculated error of the controller, and t is the initial time at which the controller starts, and bias is the controller initial output.

The parameters of the PI controllers used were tuned based on trial and error methodology. By performing a series of parametric studies under slug flow conditions, it was possible to adjust the gain and integral time in such a way that variations and disturbances in separator liquid level was kept as low as possible. Optimum values achieved as a result of this simulation are:

K_{LC} = 0.006 level controller gain,

Ti_{LC} = 5sec level controller integration time step, and the set point of the controller was set to maintain a separator liquid level of 0.5 m.

Similarly, the pressure controller system was tuned in order to stabilize the pressure in the separator which was kept at 20.5 bar, and the same method used above was applied to achieve optimum controller tuning. Values achieved are:

K_{PC} = 0.7 Pressure controller gain

Ti_{PC} = 10 sec Pressure controller integration time step.

RESULTS AND DISCUSSION

Control Results

The separator liquid level and pressure controller performed considerably well at the optimum tuning conditions at simulation runs of 6 hours which are shown in Figures 16 and 17. Large variation in liquid level characterized by high peaks was initially observed in the separator, but the controller swiftly responded by bringing the level to the desired value which was achieved at 3 hours, but unlike the pressure which had very little variation as a result of marginal pressure difference between the upstream pressure at the separator entry and the final desired topside pressure.

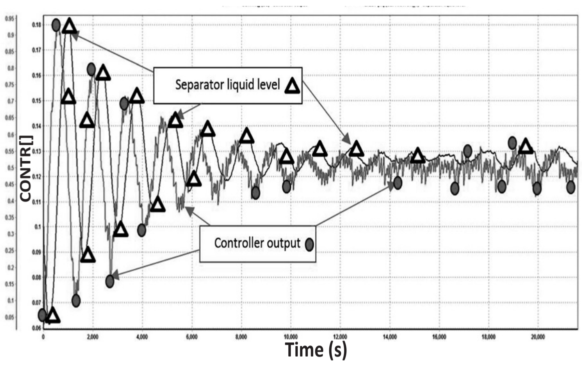

Figure 16: Controller response to liquid level variation.

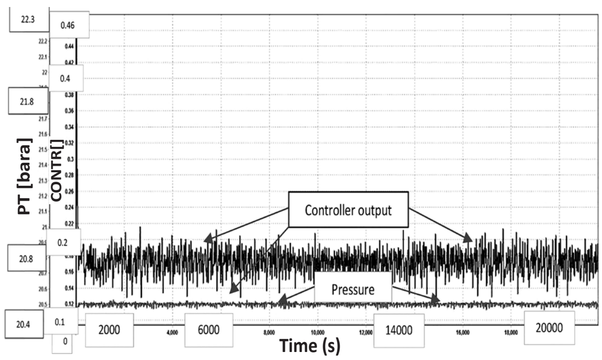

Figure 17: Controller response to pressure variation.

The controller system had a positive impact on the cyclic undulations of the liquid volumetric flow rate flowing out of the outlet of the separator as shown in Figure 18; moreover, the liquid volume flow in red spikes fluctuated between 38.74 m^3/hr and 321 m^3/hr. The controller was able to stabilize the liquid volume flow oscillations highlighted in red resulting in fluctuations between 131.54 m^3/hr and 146.89 m^3/hr as shown in Figure 19. Also shown in Figures 18 and 19 is the gas volumetric flow rate at the separator outlet obtained from simulations before and after the implementation of the S^3 control, it can also be observed from the plots that the amplitude of oscillation of the parameters reduced after the controller has been turned on. The most important benefit of introducing the S^3 was the increase of the daily production rate, in order to determine the improvement, the production rates before and after the implementation of the S^3 was calculated from the volumetric flow rate as shown in Figure 20. The result shows an increase from 131 m^3/hr to 143 m^3/hr which indicates that, the introduction of this slug control scheme; production can be increased by about 12.5%.

Based on this current research, one of the major benefits of the S^3 is the increase in production as a result of the reduction of the gas and liquid volume flow instability as observed in Figures 20 and 21.

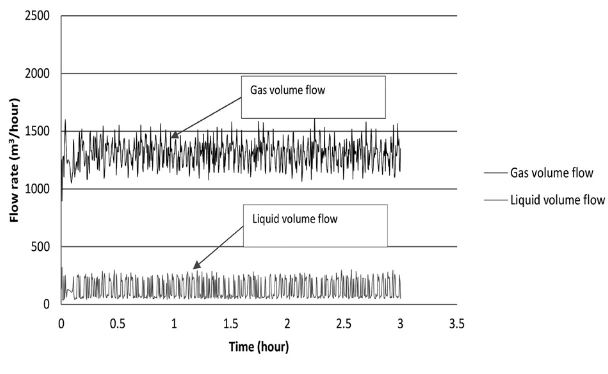

Figure 18: Outlet gas and liquid production rate before the implementation of S^3.

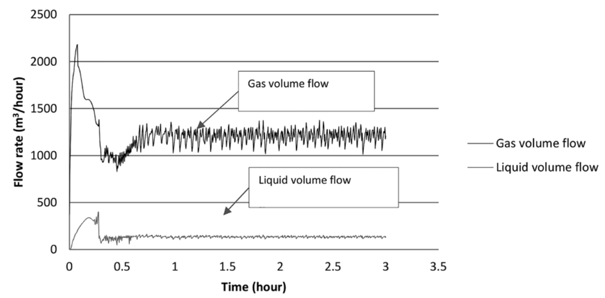

Figure 19: Outlet gas and liquid Production rate after the implementation of S^3.

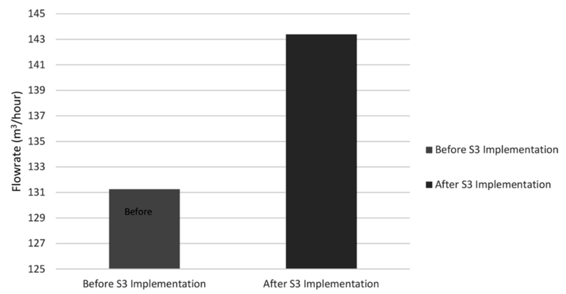

Figure 20: Difference in production rate after the implementation of S^3.

CONCLUSIONS

Considering field experience as highlighted by the work of Kovalev et al in 2003 [8] (Kovalev, and in view of this current research), and the series of numerical simulation results presented in this work, the following conclusions can be drawn:

• The presence of slugs can cause severe effects on the overall production to the topside, with potentials of causing trips on the inlet of the separator because of high-pressure fluctuations.

• The ability of OLGA is able to model the slug suppression system including the controllers was demonstrated.

• Implementation of the S^3 in mitigating severe slugging is effective. This is achieved via total volumetric flow control and liquid flow control. Implementation of the system can result in the following benefit:

• Ensuring the stability of the gas and liquid volumetric flow into the first stage separator,

• Reduction in production loss because of slugs by using a robust control strategy.

• Reduction in platform trips,

• Increase in production which is estimated in the order of 12%, and

• Increase in production because of the reduced margin needed in the facilities to accommodate slugs.

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Appendix 1

Well X1 Volumetric flow rate conversion

Volumetric flow conversion to mass flow:

Well X1:

Q_{oil} = 6722 bopd

Q_{gas} = 4 MMScf/d

Q_{water} = 0 STB/day

Q_{oil} = 6722 = 0.012369 m³/s; $U_{sl} = \frac{Q_l}{A_{pipe}} = \frac{0.012369}{0.0324}$ = 0.3818 m/s

Q_g = 4 MMScf/day

$$\frac{P_{s*V_s}}{T_s} = \frac{P_{o*V_o}}{T_o} = \frac{1*4}{15} = \frac{19.994*V_o}{65.55}$$

V_o=0.8743 MMCf/day

Q_{gas} = 24757.42 m³/day

$$Q_{gas} = \frac{24757.42}{86400} = 0.2865 \text{ m}^3/\text{s}$$

$$U_{sg} = \frac{Q_g}{A_{pipe}} = \frac{0.2865}{0.0324} = 8.8426 \text{ m/s}$$

$U_m = U_{sl} + U_{sg}$ = 0.3818+8.8426=9.2244 m/s

$$\lambda_l = \frac{U_{sl}}{U_m} = \frac{0.3818}{9.2244} = 0.0414 \text{ [-]}$$

$\rho_{mix} = \lambda_l \rho_l + (1-\lambda_l)\rho_g = 0.0414 * 641 + (1 - 0.041\ 4) * 18.2 = 43.9874\ kg/m^3$

$m\cdot_{mix} = \rho_{mix}(Q_{oil} + Q_{gas}) = 13.15\ kg/s$

Appendix 2

Well X2 Volumetric flow rate conversion

Well X2 (Volumetric flow conversion to mass flow) resulted in;

Q_{oil} = 22,157 bopd

Q_{gas} = 23 MMScf/day

Q_{water} = 6 STB/day

Q_oil = 22,157 = 0.04077 m^3/s ; $U_{sl} = \frac{Q_l}{A_{pipe}} = \frac{0.04077}{0.0324} = 1.2583$ m/s

Q_g=23 MMScf/d

Appendix 3

$$\frac{P_{s*}V_s}{T_s} = \frac{P_{o*}V_o}{T_o} = \frac{1*23}{15} = \frac{19.994*V_o}{65.55}$$

V_o=5.027 MMCf/d

Q_{gas} = 142,348.79 m^3/d

$$Q_{gas} = \frac{142,348.79}{86400} = 1.64756\ m^3/s$$

$$U_{sg} = \frac{Q_g}{A_{pipe}} = \frac{1.64756}{0.0324} = 50.85\ m/s$$

$U_m = U_{sl} + U_{sg} = 1.2583 + 50.85 = 52.1083$ m/s

$$\lambda_l = \frac{U_{sl}}{U_m} = \frac{1.2583}{52.1083} = 0.02415\ [-]$$

$\rho_{mix} = \lambda_l \rho_l + (1-\lambda_l)\rho_g = 0.02415 * 641 + (1 - 0.02415) * 18.2 = 33.24\ kg/m^3$

$m\cdot_{mix} = \rho_{mix}(Q_o + Q_{gas}) = 33.24\ (0.04077 + 1.64756)$ kg/s = 56.12 kg/s

Q_W=6 STB/d = (0.000008280) m/s

$m = \rho_W * Q_W$ =980*0.000008280

$m\cdot_W = 0.0081144$ kg/s

$m\cdot_{mix\ (owg)} = 56.12 + 0.008114 = 56.128$ kg/s

Appendix 4
OLGA Simulation Input Parameters

	Mass flow rate (Kg/s)	Mass flow rate (Kg/s)	Temperature (°C)		Pressure (Pa) (Separator)
Scenario	Well X1 (Source 1)	Well X2 (Source 2)	Well X1	Well X2	
Scenario 1(Case 1)	8.745	Nil	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 1 (Case 2)	7	Nil	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 1 (Case 3)	6	Nil	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 1 (Case 4)	5	Nil	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 2 (Case 1)	8.745	56.128	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 2 (Case 2)	7	56.128	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 2 (Case 3)	6	56.128	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 2 (Case 4)	5	56.128	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 3 (Case 1)	8.745	25	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 3 (Case 2)	7	20	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 3 (Case 3)	6	15	82.22	75.56	1.99 X10^6 (290 psi)
Scenario 3 (Case 4)	5	10	82.22	75.56	1.99 X10^6 (290 psi)

NOMENCLATURES

FPSO	: Floating Production Storage and Offloading
FMC	: Field Maintenance Condition
GOR	: Gas/Oil Ratio
GASFRACEQ	: Gas Fraction
GUI	: Graphical User Interface
LT	: Level Transmitter
MDC	: Field Maintenance Condition
MF	: Gas/Oil Ratio
NSLUG	: Gas Fraction
OILFRACEQ	: Graphical User Interface
PVT	: Pressure Volume Temperature
PT	: Pressure reading
PID	: Proportion Integral Derivative
PI	: Proportional Integral
TM	: Temperature reading
RB	: Riser-base
SEP	: Separator
IEA	: International Energy Agency
LIQUID FRACEQ	: Liquid Fraction

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