

## Modeling and Control Solutions for Riser Slugging in Offshore Oil Fields

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### Abstract

A highly transient flow regime called *riser slugging* can develop in pipelines from wellheads to separation tanks which introduces severe oscillations of pressure and flow rates in off-shore processes. Mitigation of the intensity of *riser slugging* may be achieved via anticipative control actions on valves positioned at the entrance and at gas/oil exits of the separators. In *Phase 1*, this work is concerned with the development of a simplified phenomenological model for the producing system, including a set of wells performing gas-lift, their corresponding risers, entrance manifold and a gas-oil separator. The input variables to this integrated model are the percentage of opening of choke and exit valves on the separator. The corresponding output variables are its internal pressure and liquid level in the separator. The oscillating slugging in the pipelines arises naturally as a consequence of the values chosen for specific model parameters like down-hole pressures, gas injection flow rates, fluid properties and geometry of wells and lines. In *Phase 2*, this work is involved with the mitigation of slugging consequences via determination of optimum adaptive decentralized control strategies actuating on the choke and exit valves on the gas-liquid separator. Finally, in *Phase 3*, we approach the identification of a stochastic predictor for the entire system (risers + tanks) adopting an ARX MIMO structure, periodically tuned to the real process by using recorded time series of inputs and outputs. With a tuned ARX predictor, optimum anti-slugging control strategies can be periodically established for the real process..

Keywords: riser slugging, gas-lift, slugging control, ARX predictor

### 1. Introduction

In offshore oil fields, risers are used to transport a multiphase mixture (composed by oil, gas, water and sand) from the wellheads to separation tanks on producing platforms. In case of inexistence of separation and pumping facilities near the wellheads, on the sea bottom, this multiphase mixture must be propelled to the sea surface at expenses of the reservoir pressure. For deep water, a common situation is

that the reservoir is not sufficiently pressurized to promote the flow of oil at satisfactory rates, requiring artificial ascension of oil via gas-lift techniques, where injection of compressed natural gas is maintained at some points in the well column. In consequence, high flow rates of gas have to be accommodated in the system of risers for further recovery by gas-liquid separators at the platform, before re-compression and re-injection. In this scenario, and depending on the flow conditions, a flow regime called *riser slugging* can develop in the pipelines. This regime, naturally highly transient, introduces severe oscillations of pressure and flow rates into the system.

Aamo et al. (2005) reported that flow in wells is highly oscillatory showing instability phenomena aggravated for the case of mature wells. Space and load constraints for off-shore units favor compact equipments which increases sensitivity of downstream processes, mostly in the early stage of phase separation. Multiphase separators are hence a key-step in off-shore facilities, with a two-fold process objective: (a) phase separation; and (b) dampening *riser slugging*, as in surge tanks. Faanes and Skogestad (2003) defined a *buffer tank* as a unit at which the holdup is explored to promote smooth operation. While in some level applications it is desirable to maintain tight regulation, such as in a reactor to achieving maximum capacity, this is seldom the case for level control in surge tanks, as it betrays the anti-surge requirement.

Despite the importance of level control, loops operate poorly due to inadequate tuning of control parameters. In practice, to meet the conflicting objectives of flow filtering and keeping the level within constraints is not a simple task. Most levels of surge tanks are currently controlled by detuned PI controllers. However, simply detuning the control loop, by decreasing the proportional gain while retaining the integral action is not recommended as the damping factor is reduced, leading plant to cyclic swing. Mitigation of the intensity of *riser slugging* requires an *averaging level* control strategy and anticipative control actions on valves at the entrance and at gas/oil exits of the separators. In *averaging level control*, fluctuations of discharge flow are minimized under constraints of level remaining between limits, typically 30% to 70% of level scale (Horton et al., 2003).

In this work, a two-phase separator was modelled and an adaptive control law of PI controller parameters was employed and tuned to optimize a control performance criteria subject to minimum and maximum limits of level, with penalty imposed on control movements in order to reduce downstream disturbances.

## **1. Simplified Modeling of Oscillating Producing Wells and Reception System**

We firstly addressed the development of a simplified phenomenological model for the producing system, including a set of wells performing gas-lift, their corresponding risers, entrance manifold and a gas-oil separator.

1.1. Gas-Lift Mass Balances

The dynamic behavior of the gas lift derives from the interaction between the annular and tubular volumes above the gas injection point, described by mass balances in Equations 1 to 3.

$$\frac{dH_A}{dt} = W_G - W_I \quad (1)$$

$$\frac{dH_G}{dt} = W_I - W_C \quad (2)$$

$$\frac{dH_L}{dt} = W_R - W_P \quad (3)$$

The valve responsible for injecting gas into the annular region is taken as a *swing-check-valve*:

$$W_I = S_I \sqrt{2\rho_I \frac{\max\{0, P_I - P_V\} \cdot 10^5}{K_{CV}}} \quad (4)$$

The choke valve for feeding gas to the annular region and the production choke valve are modeled as gate valves, according to Equations 5 and 6, assuming that the oil reservoir and the off-shore separator are at constant pressure.

$$W_C = S_P \cdot Z_G \cdot X_P \sqrt{2\rho_M \frac{\max\{0, P_T - P_{OUT}\} \cdot 10^5}{K_{GV}}} \quad (5)$$

$$W_P = S_P \cdot Z_L \cdot X_P \sqrt{2\rho_M \frac{\max\{0, P_T - P_{OUT}\} \cdot 10^5}{K_{GV}}} \quad (6)$$

$$W_G = 0.6 \frac{\max\{0, P_G - P_A\}}{|P_G - P_A| + \delta}, \quad (\delta = 10^{-6}) \quad (7)$$

The oil mono-phase flow section between the reservoir and the gas injection point is solved in pseudo-stationary mode – Equation 8 – with the reservoir and injection point pressures, assuming incompressible flow with friction factor given in Equations 9 and 10 (Chilton's Equations), which comprise all hydraulic regimes.

$$\frac{P_R - P_V}{\rho_R} = g \cdot L_R + \left( \frac{f_R}{2} \right) \left( \frac{L_R}{D_R} \right) \left( \frac{W_R}{A_R \rho_R} \right)^2 \quad (8)$$

$$f_R = f(\varepsilon_R / D_R, \text{Re}_R), \quad (\varepsilon_R / D_R = 10^{-5}) \quad (9)$$

$$\text{Re}_R = \frac{4W_R}{\pi D_R \mu_R} \quad (10)$$

The remaining gas-lift equations are:

$$\frac{P_R - P_V}{\rho_R} = g.L_R + \left(\frac{f_R}{2}\right) \left(\frac{L_R}{D_R}\right) \left(\frac{W_R}{A_R \rho_R}\right)^2 \quad (8)$$

$$P_I = \left(\frac{RT_A}{M}\right) \frac{H_A}{A_A L_A} + \left(\frac{gH_A}{A_A}\right) \cdot 10^{-5} \quad (11)$$

$$P_A = P_I - \left(\frac{gH_A}{A_A}\right) \cdot 10^{-5} \quad (12)$$

$$\rho_I = \frac{P_I M}{RT_A} \quad (13)$$

$$\rho_M = \frac{H_G + H_L}{A_T L_T} \quad (14)$$

$$Z_G = \frac{H_G}{H_G + H_L}, \quad Z_L = \frac{H_L}{H_G + H_L} \quad (15)$$

$$P_T = \left(\frac{RT_T}{M}\right) \frac{H_G}{A_T L_T - H_L / \rho_R} \quad (16)$$

$$P_V = P_T + g \left(\frac{H_L + H_G}{A_T}\right) \cdot 10^{-5} \quad (17)$$

$H$ ,  $W$ ,  $S$ ,  $X$ ,  $K$ ,  $P$ ,  $g$ ,  $Z$ ,  $\rho$  are, respectively, hold-up (kg), mass flow rate (kg/s), cross-sectional area (m<sup>2</sup>), valve opening, valve coefficient when 100% open, pressure (bar), gravity acceleration (m/s<sup>2</sup>), mass fraction and density (kg/m<sup>3</sup>). Subscripts  $A$ ,  $V$ ,  $G$ ,  $L$ ,  $M$ ,  $I$ ,  $T$ ,  $C$ ,  $R$ ,  $P$ ,  $OUT$ , stand respectively for annular region, production column base, gas in the production tube, liquid in the production tube, bi-phase mixture, in the production tube, injected gas, top of the production tube, gas in the production choke, reservoir, production choke, product tank. S.I. Units are used and the gas is fed to the annular region at 0.6 kg/s.

The gas phase behavior is simplified to an isothermal ideal gas. Furthermore, temperatures in the gas-lift subsystems are simplified to annular temperature equal to fed gas temperature, and production tube temperature identical to the oil temperature in the reservoir.

Equations 1 to 3 are integrated numerically (*ode15s*, MATLAB, *The Mathworks Inc.*). The simulated Base Case is summarized in Table 1 and the control structure of the separator is pictured in Figure 1.

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The unstable behavior presented at reduced gas flow rate is presented in Figure 3, with a dynamic behavior that defies the performance of controller loops in the offshore facility, and requiring a control strategy in the multiphase separator to mitigate downstream disturbances.

OIL	P (bar)	160
	T (°C)	108
	Density (kg/m <sup>3</sup> )	850
	Viscosity (Kg/m.s)	0.012
	Production Valve (X <sub>P</sub> )	0.8
GAS	P (bar)	120
	T (°C)	60
	Flow Rate (kg/s)	0.6
	Gas Valve (X <sub>G</sub> )	0.5

Table 1. Base Case for Gas-Lift Simulation

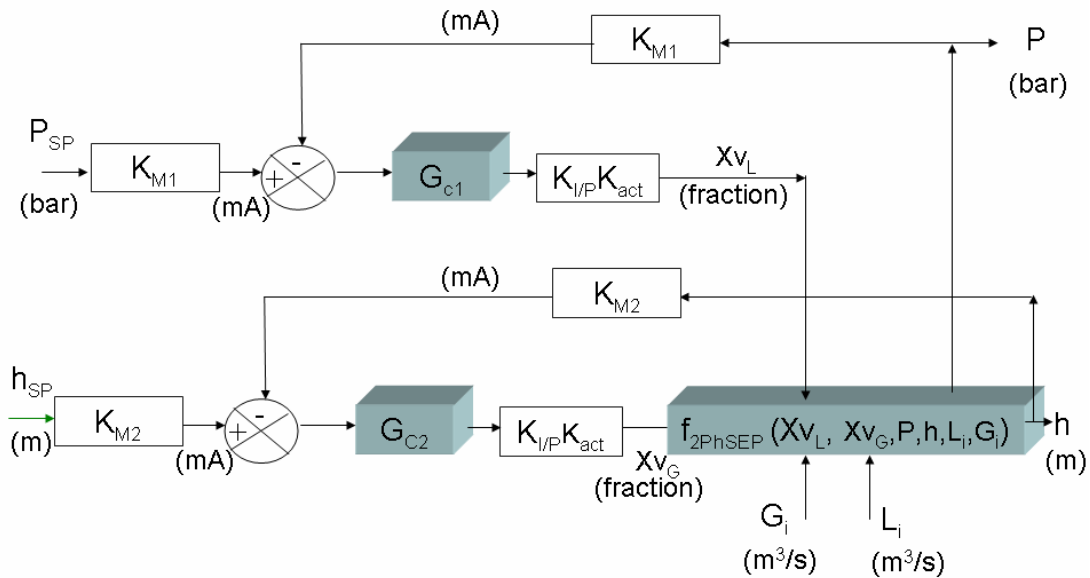


Figure 1. Control Structure of 2-phase separator.  $K_{M1}$ ,  $K_{M2}$  = sensor gains,  $G_{C1}$ ,  $G_{C2}$  = pressure controller and adaptive level controller,  $K_{I/P}K_{act}$  = factor (converts mA to valve opening),  $f_{2PhSEP}$  = non-linear 2-Phase Separator,  $X_{VL}$  = opening of liquid valve,  $X_{VG}$  = opening of gas valve,  $L_i$  = inlet liquid flow rate,  $G_i$  = outlet liquid flow rate.

Profiles obtained for the Base Case are illustrated in Figure 2.

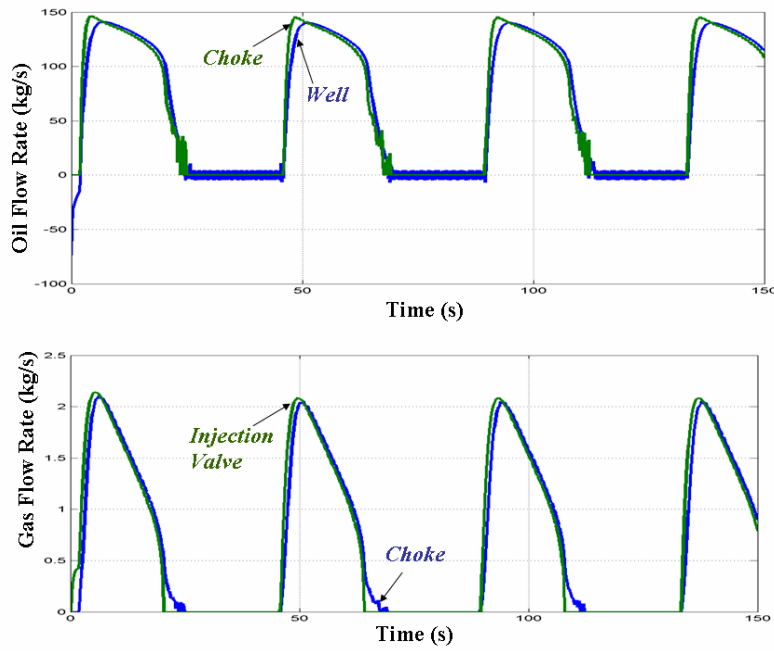


Figure 2. Flow Profiles for Gas-Lift Simulated at Base Case

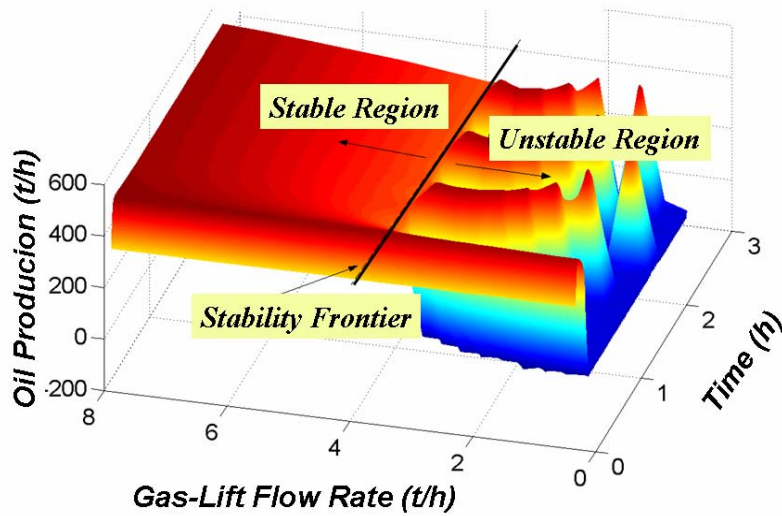


Figure 3. Flow instability in Gas-Lift Process According to Equations 1 to 17: Sensitivity of Oil Production to Gas-Lift Flow Rate.

### 1.2. Liquid Mass Balance

Let  $V_L$  be the volume of the liquid phase in the separator ( $m^3$ ),  $L_i$  and  $L_o$  respectively be the feed and discharge flows ( $m^3/s$ ), and  $h_L$  the liquid level in the tank (m), hence:

$$\frac{dV_L}{dt} = L_i - L_o \quad (18)$$

$$\frac{dh_L}{dt} = \frac{L_i - L_o}{2C\sqrt{h_L(D-h_L)}} \quad (19)$$

### 1.3. Gas Balance in the Tank

Equation 20 derives from the gas balance in the tank, where  $P$  is the vessel operating pressure (bar),  $V_G$  is the volume of the gas and  $V_T$  total volume of the cylindrical vessel ( $V_G, V_T$  in  $m^3$ ),  $G_i$  and  $G_o$  are, respectively, gas feed and gas discharge flow rates ( $m^3/s$ ).  $C$  and  $D$  are the length and diameter of the separator (m).

$$\frac{dP}{dt} = \frac{P(L_i + G_i - L_o - G_o)}{V_T - V_L}; V_T = C \frac{\pi D^2}{4} \quad (20)$$

### 1.4. Liquid Volume

The volume occupied by the liquid fraction is described by Equation 21.

$$V_L = C \left\{ \left( \frac{D^2}{4} \right) a \cos\left(1 - \frac{2h_L}{D}\right) - \frac{1}{2} (D - 2h_L) \sqrt{h_L(D - h_L)} \right\} \quad (21)$$

### 1.5. Valve Equations

The gas flow ( $m^3/s$ ) at operating pressure and temperature, for a given valve opening fraction ( $x_G$ ), is given in Equation 22, where  $P$  and  $P_G$  are respectively upstream and downstream pressures (bar),  $T$  and  $C_V^{MAXG}$  are the temperature (K) and the valve coefficient for the gas, and  $MM$  the molecular mass (air –  $AIR$ , and gas –  $G$ ) in  $kg/kgmol$ . The liquid flow rate is presented in Equation 23 as a function of the opening fraction ( $x_L$ ), where  $P_L$  is the upstream pressure (bar),  $\rho$  is the density  $kg/m^3$  (subscripts  $L$  and  $H_2O$  for liquid and water, respectively),  $C_V^{MAXL}$  is the valve coefficient for the liquid, and the factor  $10^{-5}$  is in  $bar/Pa$ .

$$G_o = 2.881 * 10^{-4} * C_V^{MAXG} * x_G \sqrt{\frac{(P - P_G)(P + P_G).T. \frac{MM_{AIR}}{MM_G}}{P^2}} \quad (22)$$

$$L_o = 2.4028 * 10^{-4} * C_V^{MAXL} * x_L \sqrt{\frac{P - P_L + \rho_L * g * h_L * 10^{-5}}{\frac{\rho_L}{\rho_{H_2O,15.5^\circ C}}}} \quad (23)$$

The input variables to this integrated model are the percentage of opening of choke and exit valves on the separators. The corresponding output variables are the pressure

and liquid level in the separator. The oscillating slugging in the pipelines arises naturally as a consequence of the values chosen for specific model parameters like down-hole pressures, gas injection flow rates, fluid properties and geometry of wells and lines. The well model uses a non-distributed approach based on the division of the well into annular and tube compartments. The risers adopt a pseudo-stationary two-phase (slug) flow model. The separators are described via conservation equations for gas and liquid compartments coupled to valve flow rate models.

## 2. Control Structure for Reception of Slugging Feeds

We now consider the mitigation of slugging consequences in the separation system via optimum adaptive decentralized control strategies actuating on the choke and exit valves on the gas-liquid separators. In offshore production plants, gravity separators, compressors, hydrocyclones (de-oilers) and electrostatic treaters are used to specify oil, gas and water for exportation. A simple structure of vessels in series results, which indicates that substantial attenuation of feed disturbances can be obtained if proper control is done. Merely fast acting control loops, designed for disturbance rejection, facilitate propagation of downstream disturbances in the flow rates.

Cheung and Luyben (1980) explore different level control strategies, such as the strategy proposed by Shunta and Fhervari (1976), the wide-range controller, with adaptation of PI controller tuning parameters ( $K_C$  and  $\tau_I$ ) according to Equation 24:

$$K_C = f K_{CO}; \quad \tau_I = \frac{\tau_{I0}}{f} \quad (24)$$

where  $K_{CO}$  and  $\tau_{CO}$  are the proportional gain and the reset time for zero error ( $e$ ) and

$$f = (1 + |eK \ln(K_1)|) (25^{|e|K_1}) \quad (25a)$$

with  $K$  and  $K_1$  as additional fixed controller parameters.

In this work, an alternative adaptation law was proposed, in substitution for Equation (25a)

$$f = (1 + \frac{1}{1 + e^{(\lambda_1(E_1 - \text{abs}(e))})}} + \alpha(1 + \frac{1}{1 + e^{(\lambda_2(E_2 - \text{abs}(e))})}}) + \beta(1 + \frac{1}{1 + e^{(\lambda_3(E_3 - \text{abs}(e))})}}). \quad (25b)$$

The parameters in Equation 25a allow for smooth transition of controller behavior along three plateaus of intensity defined by values  $I$ ,  $\alpha$  and  $\beta$ . Parameters  $\lambda_1$ ,  $\lambda_2$  and  $\lambda_3$  regulate the sharpness of transition between neighboring plateaus, while  $E_1$ ,  $E_2$  and  $E_3$  represent the respective error thresholds.

Figure 4 shows the distinct shape of Cheung and Luyben (1980) adaptation factor (Equation 25a) and the function proposed in the present work (Equation 25b).



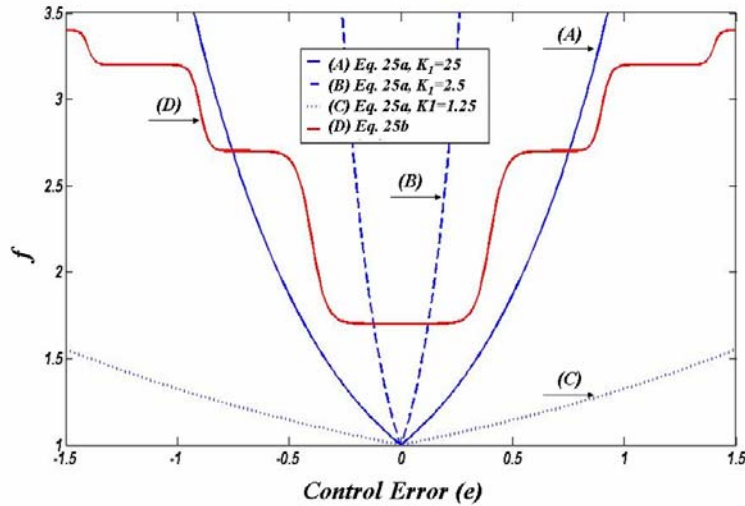


Figure 4. Adaptation factor  $f$ : (A) Equation 25a,  $K_I=25$ , (B) Equation 25a,  $K_I=2.5$ , (C) Equation 25a,  $K_I=1.25$ , (D) Equation 25b, with  $\alpha=0.5$ ,  $\beta=0.2$ ,  $\lambda_1=30$ ,  $\lambda_2=50$ ,  $\lambda_3=70$ ,  $E_1=0.4$ ,  $E_2=0.9$  and  $E_3=1.4$  mA

$K_{CO}$  and  $\tau_{CO}$  (in Equation 24) and the additional parameters used in Equation 25b along with PI pressure controller parameters ( $K_C$  and  $\tau_C$ ) were tuned in a procedure posed as a constrained optimization problem, with objective function defined as the weighted sum of level controller error, pressure controller error and control effort, subject to minimum and maximum limits on separator level.

For minimizing CPU time at each optimization step, prior Gas-Lift process simulation, producing a time series for  $L_i$  and  $G_i$  was used, assuming  $P_{SEPARATOR}$  constant.

### 2.1. Tuning Procedure for Level and Pressure Controllers

The controllers were tuned through optimization of an objective function, described in Equation 26a, accounting simultaneously for performance of level and pressure control loops, subject to constraints of minimum ( $h_{min}$ ) and maximum ( $h_{max}$ ) levels (Equation 26b).

$$F_{OBJ} = R \text{var}(\text{abs}(\delta x_G)) + R \text{var}(\text{abs}(\delta x_O)) + Q \text{var}(\text{abs}(\delta P)) \quad (26a)$$

$$h_{min} \leq h \leq h_{max} \quad (26b)$$

where  $\{\delta x_G\} = (1 - q^{-1})\{x_G\}$ ,  $\{\delta x_O\} = (1 - q^{-1})\{x_O\}$  and  $\{\delta P\} = (1 - q^{-1})\{P\}$ ,

with  $\{\cdot\}$  standing for time series in the time window from 0 to 9000s, and  $q^{-1}$  represents the backwards shift operator.  $R$  is a weighting factor which establishes the impact in the objective function of penalties on the control effort, and  $Q$  the contribution of pressure controller error. Note that the controller errors are not

considered in the performance metrics of the control structure, as long as it is kept in a preset range (e.g.,  $h_{min} = 0.1 D$ ,  $h_{max} = 0.9D$ ). Hence the surge capacity of the separator vessel is used, averaging disturbances. The decision variables in the optimization problem ( $\underline{y}$ ) were the pressure controller parameters ( $K_C$  e  $\tau_I$ ), and level controller parameters  $K_{Co}$ ,  $\tau_{I0}$ ,  $\alpha$ ,  $\beta$ . The remaining parameters were fixed at  $\lambda_1=30$ ,  $\lambda_2=50$ ,  $\lambda_3=70$ ,  $E_1=0.4$ ,  $E_2=0.9$  and  $E_3=1.4$ . Nelder & Mead optimization algorithm, available in MATLAB Optimization Toolbox (The Mathworks Inc.) was used. In order to impose limits in the search region, decision variables ( $\underline{y}$ ) were transformed according to Equation (27).

$$\underline{\theta} = \sqrt{\frac{\underline{y} - \underline{y}_{min}}{\underline{y}_{max} - \underline{y}}} \quad (27)$$

Limits on the decision variables are displayed in Table 2.

Tuning Parameters ( $\underline{y}$ )	Lower Limit ( $\underline{y}_{min}$ )	Upper Limit ( $\underline{y}_{max}$ )
$K_{Co}$	0.01	1
$\tau_{I0}$	0.05	1000
$\alpha$	0.01	5
$\beta$	0.05	1000
$K_C$	1	1000
$\tau_I$	1	1000

Table 2. Limits in Decision Variables

Constraint of minimum and maximum level were imposed as a penalty function ( $SATL$ ) added to the objective function according to Equation 28a

$$F_{OBJ}^{MOD} = F_{OBJ} + W \int_{t=0}^{t_{final}} (SATL) dt \quad (28a)$$

where

$$\int_{t=0}^{t_{final}} (SATL) dt = \int_{t=0}^{t_{final}} \left( 1 + \frac{1}{1 + \exp(200(h_{max} - h))} - \frac{1}{1 + \exp(200(h_{min} - h))} \right) dt \quad (28b)$$

For a separator of diameter equal to 3m,  $SATL$  is presented in Figure 5.

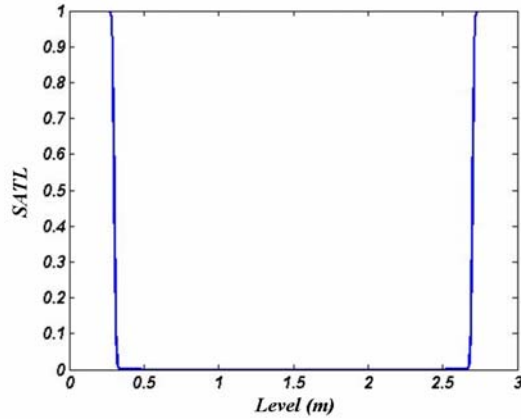


Figure 5. Penalty of Minimum and Maximum Levels

### 2.2. Tuning Results

Simulation conditions were: level set-point at 1m, pressure set-point at 8 bar, separator geometry as  $D=3\text{m}$  and  $C=8\text{m}$  (length),  $PL = 6$  bar,  $PG = 6$  bar,  $C_V^{MAXL}=1025$ ,  $C_V^{MAXG}=120$ . Sensor ranges for level and pressure were assumed, respectively, as 0-2.2m and 1-100bar. Tuning procedure was applied in 5 weighting scenarios (Base Case, and Cases a through d). All cases corresponded to  $W=1000$ . Table 3 presents the resulting tuning. Case b presented large deviations with respect to set-points, as can be observed in Figure 6.

	Base Case R=1000 Q=10	Case a R=100 Q=10	Case b R=10 Q=10	Case c R=10 Q=100	Case d R=10 Q=1000
<u>Tuning</u>					
<u>Level</u>	$K_{C0} = 0.62$	$K_{C0} = 0.92$	$K_{C0} = 0.54$	$K_{C0} = 0.51$	$K_{C0} = 0.49$
	$\tau_{I0} = 126.3$	$\tau_{I0} = 80.8$	$\tau_{I0} = 119.6$	$\tau_{I0} = 102.9$	$\tau_{I0} = 204.7$
	$\alpha = 0.53$	$\alpha = 0.02$	$\alpha = 0.44$	$\alpha = 0.05$	$\alpha = 0.11$
	$\beta = 0.25$	$\beta = 0.08$	$\beta = 0.21$	$\beta = 0.05$	$\beta = 0.06$
<u>Pressure</u>	$K_C = 10.33$	$K_C = 20.4$	$K_C = 55.1$	$K_C = 212.5$	$K_C = 286.0$
	$\tau_I = 123.2$	$\tau_I = 891.7$	$\tau_I = 85.3$	$\tau_I = 290.9$	$\tau_I = 170.6$

Table 3. Tuning Parameters as function of  $R$  and  $Q$

Results show that increasing weight on pressure control loop although favoring tracking of pressure set-point introduces severe disturbance in level control and flow rates. Balancing both weights (Case b:  $R=10$ ,  $Q=10$ ) favors smoothing of flow rates, with level oscillating within preset limits.

To test the robustness of the adaptation mechanism, tuning obtained for Base Case was tested for the process under unsteady feed condition (riser slugging) and steady condition (with  $L_i=0.1\text{ m}^3/\text{s}$ ,  $G_i=0.165\text{ m}^3/\text{s}$ ). Additionally, at  $t=7000\text{s}$ , a step in the

level controller set-point was applied, with amplitude of 1m. The achieved performance is pictured in Figure 7. It is worth noting that a conventional PI controller was tested (but not herein shown) and a same tuning set could not stand both flow regimes.

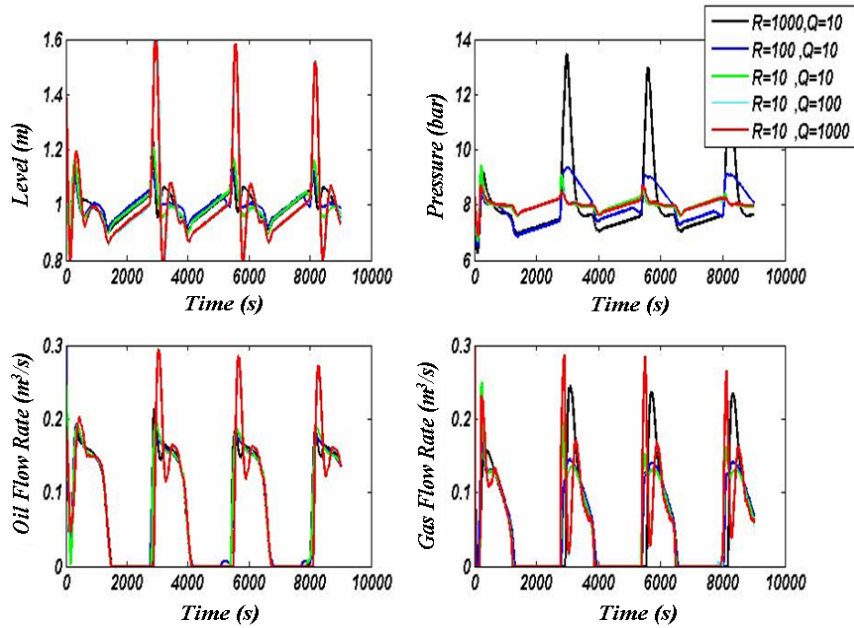


Figure 6. Control Performance: Level (m), Pressure (bar), Oil Flow Rate (L, m3/s), and Gas Flow Rate (G, m3/s).

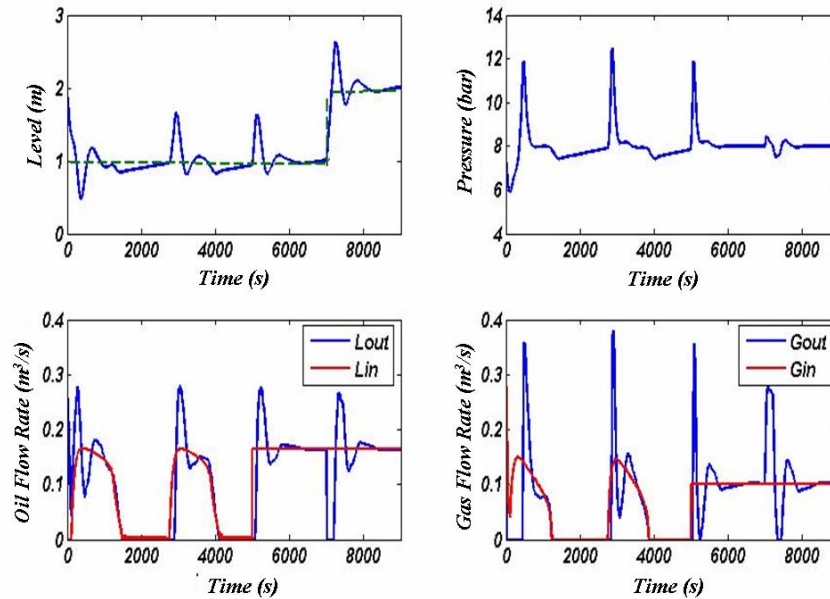


Figure 7. Controller performance (Base Case) considering two flow regimes and set-point change: Level (m), Pressure (bar), Oil Flow Rate (L, m3/s), and Gas Flow Rate (G, m3/s).

### 3. Stochastic Prediction of the Dynamic Behavior of the Producing System

Finally, the identification of a stochastic predictor, capable of representing the entire process (gas lift + separator), was accomplished via ARX MIMO structures. The ARX predictor was chosen because it is simple, accurate and can be periodically tuned to the real process through recorded time series of inputs and outputs. The utilization of the ARX predictor was demonstrated with simulated data generated with the model described in Section 1, in closed loop according to control structure proposed in Section 2. In the real process, optimum anti-slugging control strategies can be periodically established by using permanently tuned ARX predictors. A proposed strategy for slugging effects mitigation through ARX estimators is shown in Figure 8, where:

- (i) Separator and Gas-Lift, herein simulated with models in Section 1 (block *S-function*, SIMULINK / MATLAB). Pressure in the Separator affects GAS-LIFT performance.
- (ii) Averaging level control, according to adaptation law and optimization procedure described in Section 2;
- (iii) ARX predictors (SEPARATOR and GAS-LIFT estimators) in serial arrangement to produce an estimated value of  $X_{CHOKE}$ . ARX Predictors are posed as invers process model, i.e., given the output, estimates the associated process input; and
- (iv) Estimated  $X_{CHOKE}$  is used to correct choke valve opening.

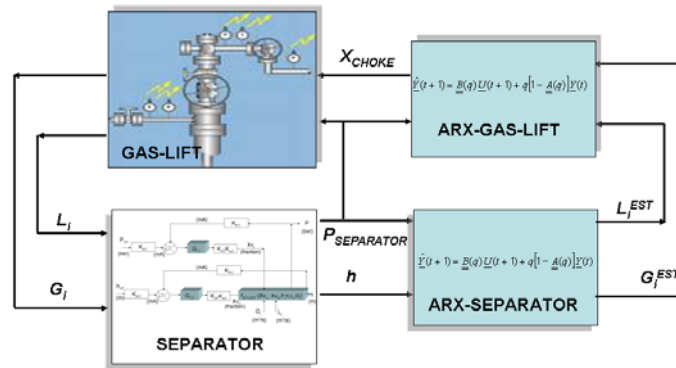


Figure 8. Control Strategy: ARX Predictors for Establishing  $X_{CHOKE}$ .

#### 3.1. ARX-MIMO Predictors

In a real application, recorded measured data covering process inputs and selected outputs are used for training ARX-MIMO predictors (Box et al., 1994). In the present case, the process time series were generated according to models presented in Section 1, with closed control loops, according to Section 2 (with Base Case tuning, Table 3). A general ARX-MIMO predictor has the following form:

$$\hat{Y}(t+1) = \underline{\underline{B}}(q)U(t+1) + q[1 - \underline{\underline{A}}(q)]Y(t) \quad (29)$$

with  $\underline{\underline{A}}(q) = 1 + \sum_{k=1}^n \underline{\underline{A}}_k q^{-k}(\cdot)$  and  $\underline{\underline{B}}(q) = \sum_{k=1}^m \underline{\underline{B}}_k q^{-k}(\cdot)$

where  $\hat{Y}(t+1)$  is the predicted vector of outputs for instant  $t+1$ ;  $q$  represents the forward shift operator;  $\underline{\underline{A}}(q)$  and  $\underline{\underline{B}}(q)$  are matrix filters – respectively the AutoRegressive filter with order  $n$  and the eXogenous input filter with order  $m$ ,  $\underline{\underline{A}}_k$  and  $\underline{\underline{B}}_k$  are constant matrices with sizes  $(ny \times ny)$  and  $(ny \times nu)$ , respectively. The predictor identification consists in estimating matrices  $\underline{\underline{A}}_k$  and  $\underline{\underline{B}}_k$  from recorded values of  $\underline{Y}$  and  $\underline{U}$  during a training window with  $NT$  time instants (the training phase). This is a linear estimation problem that can be conducted via standard techniques. The predictor is used subsequently for the next  $NI$  instants estimating process outputs (the predictive phase). Training and predictive phases can then be repeated. The quality of the training and the degree of deterioration of the predictor, as it departs from the training window, can be measured through the semi-width of 99% confidence intervals for correct responses in the training and predictive phases: narrow confidence intervals mean high adherence to the process outputs, and vice-versa.

### 3.2. Input Generation

Generation of time-series followed the pattern proposed by Pottmann and Pearson (1998), consisting of evenly distributed random inputs. In each sampling instant  $k$ , the input  $u_i(k)$  has a transition probability  $P$  of assuming a new random value in the interval  $[u_j^{MIN}, u_j^{MAX}]$  and  $1-P$  of maintaining the antecessor value  $u_i(k-1)$ .

Two predictors were identified: (i) Gas-Lift predictor – estimates  $X_{CHOKE}$  one-step ahead given present and past values of  $P_{SEPARATOR}$  (bar),  $L_i$  ( $m^3/s$ ) and  $G_i$  ( $m^3/s$ ); (ii) Separator predictor – estimates  $L_i$  ( $m^3/s$ ) and  $G_i$  ( $m^3/s$ ) given present and past values of  $h$  (m) and  $P_{SEPARATOR}$  (bar). All predictors were ARX filters with  $m=3$  and  $n=3$ . Figures 9 and 10 present the process inputs ( $\underline{U}$ ) and outputs ( $\underline{Y}$ ) for simultaneous disturbances in the inputs ( $\underline{U}$ ), in for  $P=99\%$ . For the separator, time series were produced with level and pressure controller in closed-loop, with tuning parameters obtained with Base Case (Table 3).

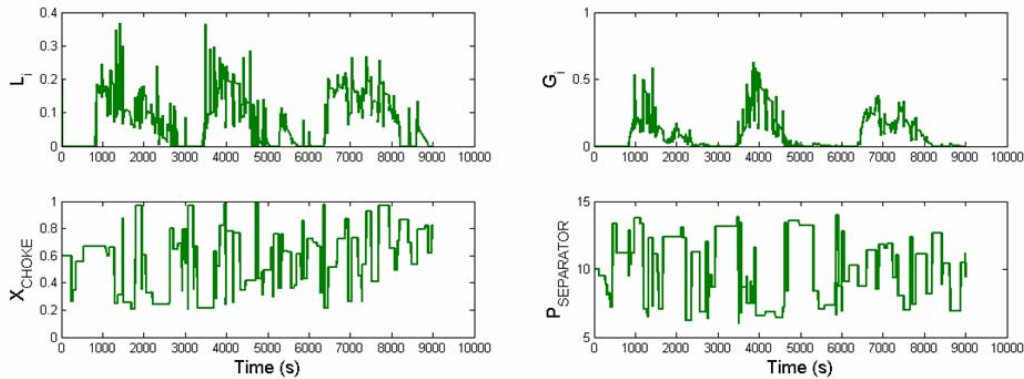


Figure 9. Time Series for GAS-LIFT ARX Predictor Identification: Pressure (bar), Opening of Choke Valve ( $X_{CHOKE}$ ) (inputs to process) and Oil Inlet Flow Rate ( $L_i$ ,  $m^3/s$ ) and Gas Flow Rate ( $G_i$ ,  $m^3/s$ ) (outputs to process).

## Modeling and Control Solutions for Riser Slugging in Offshore Oil Fields

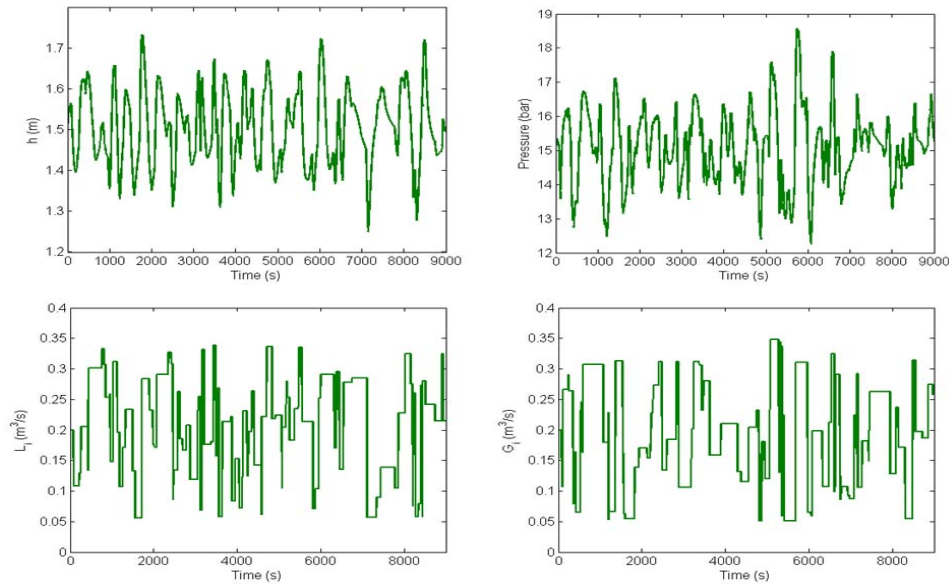


Figure 10. Time Series for SEPARATOR ARX Predictor Identification: Oil Inlet Flow Rate ( $L_i$ , m<sup>3</sup>/s), Gas Flow Rate ( $G_i$ , m<sup>3</sup>/s) (Process Inputs) and Opening of Choke Valve ( $X_{CHOKE}$ ) and Level (m), Pressure (bar) (Outputs).

Inverse predictors were identified in order to estimate process inputs for given process outputs. Predictors responses are displayed in Figures 11 and 12, with confidence intervals.

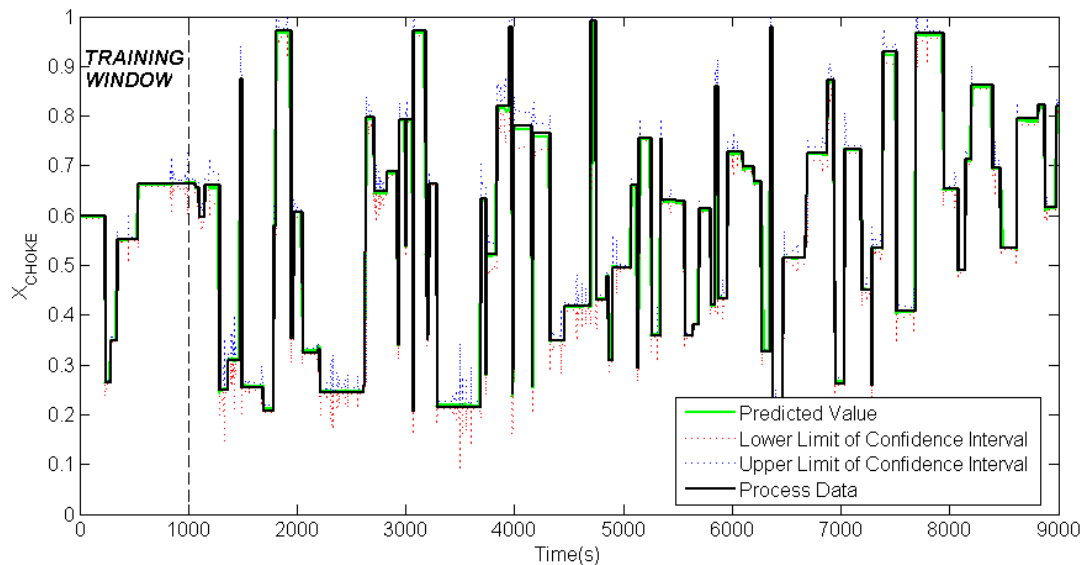


Figure 10. GAS-LIFT ARX Prediction: Given Pressure (bar), Oil Inlet Flow Rate ( $L_i$ , m<sup>3</sup>/s), Gas Flow Rate ( $G_i$ , m<sup>3</sup>/s) estimates Opening of Choke Valve ( $X_{CHOKE}$ ).

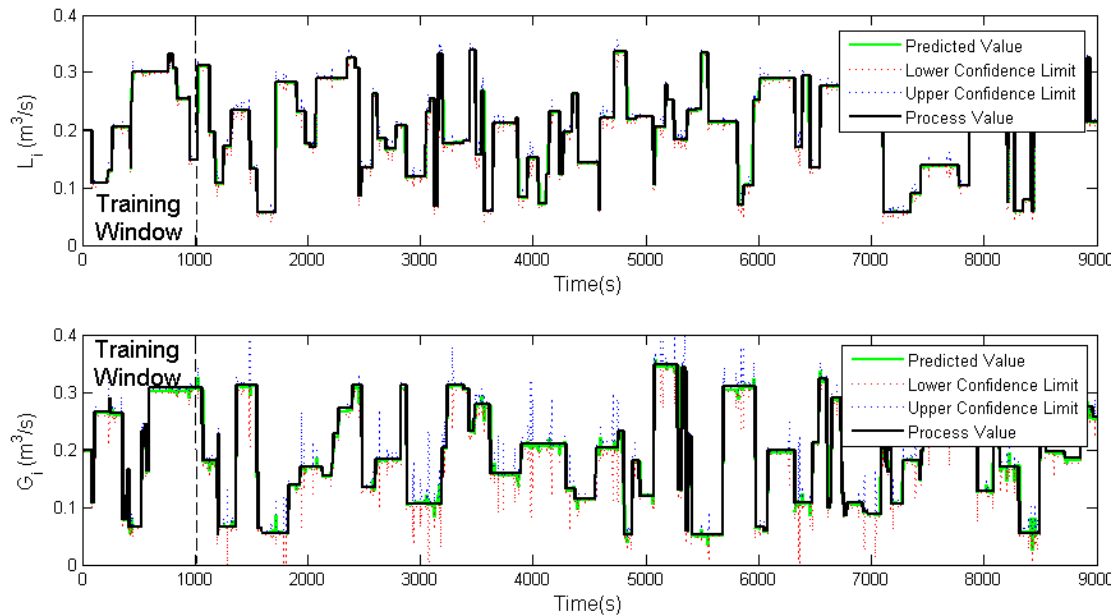


Figure 11. SEPARATOR ARX Prediction: Given Pressure ( $P_{SEPARATOR}$ , bar) and Separator Level ( $h$ , m) estimates Oil Inlet Flow Rate ( $L_i$ , m<sup>3</sup>/s), Gas Flow Rate ( $G_i$ , m<sup>3</sup>/s).

#### 4. Concluding Remarks

In this work, a two-phase separator was modeled and an adaptive control law of PI controller parameters was employed and tuned to optimize a control performance criteria subject to minimum and maximum limits of level, with penalty imposed on control movements in order to reduce downstream disturbances.

The resulting averaging level control stood both a quiescent feed condition and a slugging feed situation.

Furthermore, with a proposed Gas-Lift model, along with the separator model, time series were generated to identify ARX estimators, in a serial arrangement, for predicting the necessary choke valve opening ( $X_{CHOKe}$ ) for mitigating effects of riser *slugging*. The prediction remains within confidence intervals for 8000 s following training window of 1000 s. A control strategy including the ARX estimators are proposed.

Testing the strategy is the object of further study.

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