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Simplified Dynamical Models for Control of Severe Slugging in Multiphase Risers

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Abstract: In order to prevent the severe slugging flow regime in multiphase transport pipelines, active feedback control is the recommended solution. Instead of elaborated models such as CFD and OLGA® models, a simple dynamical model with few state variables is required in a model-based control system. We propose a new simplified dynamical model for severe slugging flow in pipeline-riser systems. The proposed model, together with five other simplified models existing in the literature, is compared to the results from the OLGA® simulator. The proposed model could be able to maintain the main dynamics of the severe slugging flow regime.

1. INTRODUCTION

Slugging has been recognised as a serious problem in the North Sea oilfields and many efforts have been made in order to prevent this problem (Courbot 1996). Severe slugging flow regime usually occurs in the pipeline-riser systems which transport oil and gas mixture form the seabed to the surface facilities. This problem, also referred to as “riser slugging”, is characterised by severe flow and pressure oscillations. Slugging problem has also been observed in the gas-lifted oil wells and two types of instabilities, casing heading and density wave instability, which both can result in production loss have been reported (Hu and Golan 2003).

The irregular flow caused by slugging can cause large operational problems for the downstream receiving facilities, and an effective way to handle or remove riser slugging is needed. Reducing opening of a top-side choke valve (choking) is the conventional solution, but this reduces the production rate. Therefore, a solution that guarantees stable flow together with the maximum possible production rate is desirable. Finally, automatic control was shown to be an effective strategy to eliminate the slugging problem (Havre and Dalsmo 2002) and different slug control strategies were tested in experiments (Godhavn, Fard et al. 2005). Anti-slug control is an automatic control system that stabilises the flow in the pipeline at the same operating conditions that uncontrolled would yield riser slugging. This control system usually uses top-side choke as the manipulated variable and some measurements (i.e. pipeline and riser pressures) as control variables.

There have been some researches involving simulation of riser slugging phenomenon using the OLGA® simulator to demonstrate the system behavior and also slug control (Fard, Godhavn et al. 2006), but for controllability analysis and controller design a simple dynamical model of the system is required. The purpose of this research work is to find or develop a simple dynamical model which has a good fit with the real system. In this modeling approach the shape and length of the slugs are not the matter of interest, because the aim is to change slug flow regime to a stable regime. Five simplified dynamical models were found in the literature. The “Storkaas model” is a three dimensional state-space model presented in (Storkaas and Skogestad 2003) and used for controllability analysis in (Storkaas and Skogestad 2007). Also, a four dimensional state-space model (Tuvnes 2008) was found, we refer to it as the “Eikrem model” (Eikrem 2008). Another simplified model referred to as the “Kaasa model” (Kaasa, Alstad et al. 2008), only predicts pressure at the bottom of the riser. Another model is the “Nydal model” (Da-Silva and Nydal 2010), which is the only one that includes friction in the pipes. The most recently published simplified model is the “Di Meglio model” (Di-Meglio, Kaasa et al. 2009), (Di-Meglio, Kaasa et al. 2010). In addition, a new dynamical model is developed in this research and it is compared with the existing models.

Different simplified models are simulated in time domain and they are compared to an OLGA reference model in the following five aspects listed in order of importance:

- Stability margin (critical valve opening)
- Frequency of the oscillations in the critical point
- Response to the step change in the valve opening
- Prediction of the steady-state values
- Maximum and minimum of the oscillations

Different simplified models are also analysed in frequency domain. Complex conjugate poles and important unstable RHP zeros in the critical operating point are considered. This paper is organised as the following. The OLGA reference model is introduced in Section 2. In Section 3, the new low-order model is proposed. The simulation results of the different models are compared in Section 4. Finally, the results are concluded in Section 5.

2. OLGA REFERENCE MODEL

In order to study the dominant dynamic behavior of a typical, yet simple riser slugging problem, the test case for severe slugging in OLGA is used. OLGA is a commercial multiphase simulator widely used in the oil industry (SPT-Group 2006). The geometry of the system is given in Fig. 1. In the OLGA test case the pipeline diameter is 0.12 m and its length is 4300 m. starting from the inlet, the first 2000 m of the pipeline is horizontal and the remaining 2300 m is inclined with a 1° angle. It causes 40.14 m descent and creates a low pint at the end of pipeline. Riser is a vertical 300 m pipe with the diameter of 0.1 m. The 100 m horizontal section with the same diameter as that of the riser connects the riser to the outlet choke valve. The feed into the system is nominally constant at 9 kg/s, with $w_L = 8.64$ kg/s (oil) and $w_G = 0.36$ kg/s (gas). The pressure after the choke valve (P_0) is nominally constant at 50.1 bar. This leaves the choke valve opening Z as the only degree of freedom in the system.

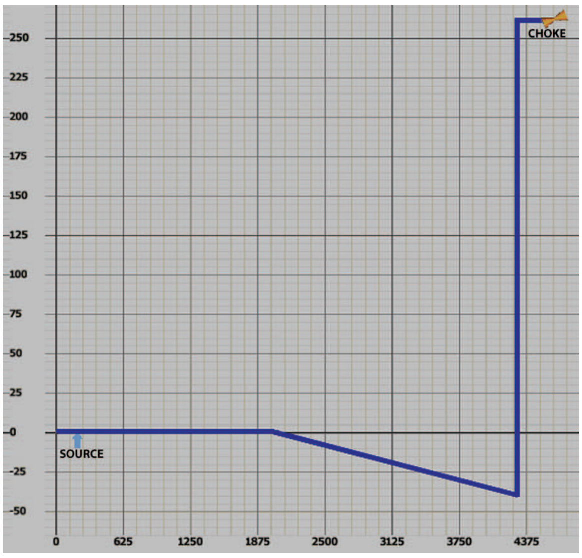


Fig. 1: Geometry of the system in OLGA

For the present case study, the critical value of the relative valve opening for the transition between a stable non-oscillatory flow regime and riser slugging is Z = 5%. This is illustrated by the OLGA simulation of inlet pressure, topside pressure and outlet flow rate in Fig. 2, with the valve openings of 4% (no slug) 5% (transient) and 6% (riser slugging).

Simulations, such as those in Fig. 2, were used to generate the bifurcation diagrams in Fig. 5, which illustrate the behavior of the system over the whole working range of the choke valve. The line in between represents the steady-state solution which is unstable without control for valve opening larger than 5%. OLGA includes a steady state preprocessor which calculates the initial values to the transient simulations. For unstable operating points, initial values provided by OLGA are considered as the steady-state solution.

For valve openings more than 5%, in addition to the steady-state solution, there are two other lines giving the maximum and minimum of the oscillations shown in Fig. 2.

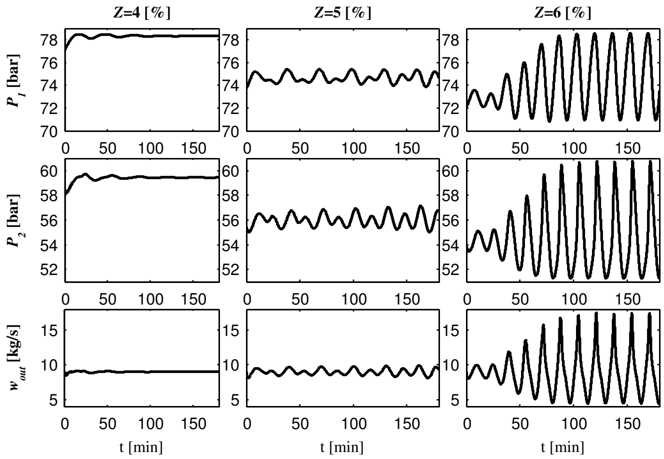


Fig. 2: Simulation results of OLGA case for different valve openings.

3. NEW LOW-ORDER MODEL

The four state equations of the proposed model are simply the mass conservation law for individual phases in the pipeline and the riser sections:

$$\dot{m}_{G1} = w_{G,in} - w_{G,lp} \tag{1}$$

$$\dot{m}_{L1} = w_{L,in} - w_{L,lp} \tag{2}$$

$$\dot{m}_{G2} = w_{G,lp} - w_{G,out} \tag{3}$$

$$\dot{m}_{L2} = w_{L,lp} - w_{L,out} \tag{4}$$

The state variables of the model are as the following.

m_{G1} : mass of gas in the pipeline

m_{L1} : mass of liquid in the pipeline

m_{G2} : mass of gas in the riser

m_{L2} : mass of liquid in the riser

Four tuning parameters are used to fit the model to the desired pipeline-riser system.

K_h : correction factor for level of liquid in pipeline

K_{pc} : choke valve constant

K_G : orifice coefficient for gas flow through low point

K_L : orifice coefficient for liquid flow through low point

3.1 Boundary Conditions

a) Inflow conditions:

In the equations (1) and (2), $w_{G,in}$ and $w_{L,in}$ are inlet gas and liquid mass flow rates. They are assumed to be constant, but the inlet boundary conditions can easily be changed. The liquid volume fraction in pipeline section can be written based on the liquid mass fraction and densities of the two phases (Brill and Beggs 1991):

$$\alpha_L = \frac{\alpha_{Lm} / \rho_L}{\alpha_{Lm} / \rho_L + (1-\alpha_{Lm}) / \rho_G}$$

The average liquid mass fraction in the pipeline section can be fairly well approximated using the inflow boundary condition:

$$\bar{\alpha}_{Lm1} \cong \frac{w_{L,in}}{w_{G,in} + w_{L,in}}$$

Combining the two above equations gives the average liquid volume fraction in the pipeline:

$$\bar{\alpha}_{L1} \cong \frac{\bar{\rho}_{G1} w_{L,in}}{\bar{\rho}_{G1} w_{L,in} + \rho_L w_{G,in}} \tag{5}$$

In equation (5), the gas density $\bar{\rho}_{G1}$ can be calculated based on the nominal pressure (steady-state) of the pipeline:

$$\bar{\rho}_{G1} = \frac{P_{1,nom} M_G}{R T_1} \tag{6}$$

b) Outflow conditions:

A constant pressure (separator pressure) condition and a choke valve model for outflow of two-phase mixture are assumed as the boundary conditions at the outlet of the riser.

$$w_{mix,out} = K_{pc} f(z) \sqrt{\rho_t (P_2 - P_0)} \tag{7}$$

$0 < z < 1$ is the relative valve opening and $f(z)$ is the characteristic equation of the valve. For sake of the simplicity a linear valve is assumed in the simulations (i.e. $f(z) = z$). $w_{L,out}$ and $w_{G,out}$, the outlet mass flow rates of liquid and gas, are calculated as follows.

$$w_{L,out} = \alpha_{Lm,t} w_{mix,out} \tag{8}$$

$$w_{G,out} = (1 - \alpha_{Lm,t}) w_{mix,out} \tag{9}$$

Liquid mass fraction at top of the riser, $\alpha_{Lm,t}$, is given by equation (38).

3.2 Pipeline model

Consider the steady-state condition in which gas and liquid are distributed homogeneously along the pipeline. In this situation the mass of liquid in the pipeline is given by $\bar{m}_{L1} = \rho_L V_1 \bar{\alpha}_{L1}$ and the level of liquid in pipeline at the low-point is approximately $\bar{h}_1 \cong h_c \bar{\alpha}_{L1}$. If liquid content of pipeline increases by Δm_{L1}, it starts to fill up the pipeline from the low-point. A length of pipeline equal to ΔL will be occupied by only liquid and level of liquid increases by $\Delta h_1 = \Delta L \sin(\theta)$.

$$\Delta m_{L1} = \Delta L \pi r_1^2 (1 - \bar{\alpha}_{L1}) \rho_L$$

$$h_1 = \bar{h}_1 + \Delta L \sin(\theta)$$

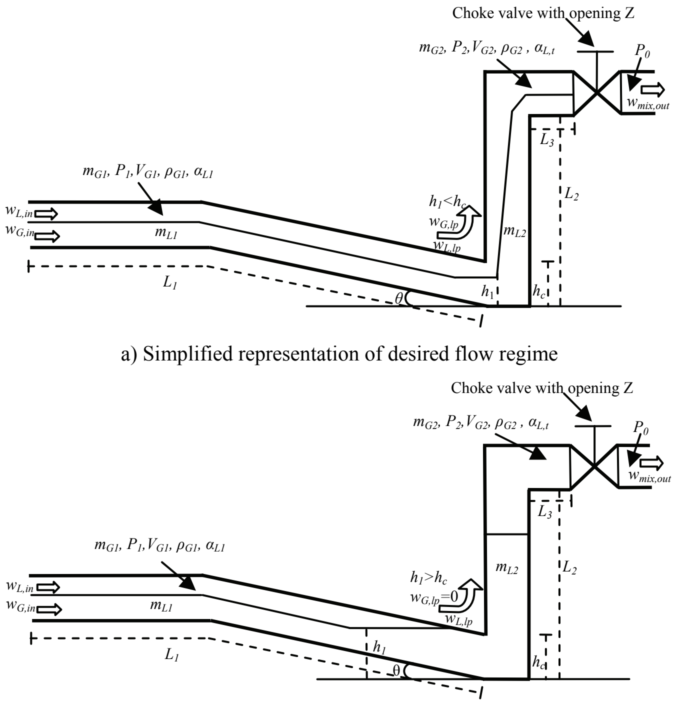
$$h_1 = \bar{h}_1 + \frac{\Delta m_{L1}}{\pi r_1^2 (1 - \bar{\alpha}_{L1}) \rho_L} \sin(\theta)$$

$$\bar{h}_1 = K_h h_c \bar{\alpha}_{L1}, \tag{10}$$

where K_h is a correction factor around unity which can be used for fine-tuning of the model.

$$h_1 = \bar{h}_1 + \left(\frac{m_{L1} - \rho_L V_1 \bar{\alpha}_{L1}}{\pi r_1^2 (1 - \bar{\alpha}_{L1}) \rho_L} \right) \sin(\theta) \tag{11}$$

Therefore, level of liquid in the pipeline h_1 can be written as a function of liquid mass in pipeline m_{L1} which is a state variable of the model. All of the other parameters are constant.



b) Simplified representation of liquid blocking leading to riser slugging

Fig. 3: Schematic representation of model parameters

Volume occupied by gas in pipeline:

$$V_{G1} = V_1 - m_{L1} / \rho_L \tag{12}$$

Gas density in pipeline:

$$\rho_{G1} = \frac{m_{G1}}{V_{G1}} \tag{13}$$

Pressure in pipeline assuming ideal gas:

$$P_1 = \frac{\rho_{G1} R T_1}{M_G} \tag{14}$$

For pressure loss due to friction in the pipeline only friction of liquid is considered.

$$\Delta P_{fp} = \frac{\bar{\alpha}_{L1} \lambda_p \rho_L \bar{U}_{sl,in}^{\ 2} L_1}{4 r_1} \tag{15}$$

The correlation developed by Drew, Koo and McAdams (Brill and Beggs 1991) for turbulent flow in smooth wall pipes is used as the friction factor in the pipeline.

$$\lambda_p = 0.0056 + 0.5 \,\mathrm{Re}_p^{\ -0.32} \tag{16}$$

Reynolds number in pipeline:

$$\mathrm{Re}_p = \frac{2 \rho_L \bar{U}_{sl,in} r_1}{\mu}, \tag{17}$$

where μ is viscosity of liquid and $U_{sl,in}$ is superficial velocity of inlet liquid:

$$\bar{U}_{sl,in} = \frac{w_{L,in}}{\pi r_1^2 \rho_L} \tag{18}$$

3.3 Riser model

Total volume of riser:
$$V_2 = \pi r_2^2 (L_2 + L_3) \tag{19}$$
Volume occupied by gas in riser:
$$V_{G2} = V_2 - m_{L2} / \rho_L \tag{20}$$
Density of gas at top of the riser:
$$\rho_{G2} = \frac{m_{G2}}{V_{G2}} \tag{21}$$
Pressure at top of riser form ideal gas law:
$$P_2 = \frac{\rho_{G2} R T_2}{M_G} \tag{22}$$
Average liquid volume fraction in riser:
$$\bar{\alpha}_{L2} = \frac{m_{L2}}{V_2 \rho_L} \tag{23}$$
Average density of mixture inside riser:
$$\bar{\rho}_m = \frac{m_{G2} + m_{L2}}{V_2} \tag{24}$$
Friction loss in riser:
$$\Delta P_{fr} = \frac{\bar{\alpha}_{L2} \lambda_r \bar{\rho}_m \bar{U}_m^{\,2} (L_2 + L_3)}{4 r_2} \tag{25}$$
Friction factor of riser using same correlation as pipeline:
$$\lambda_r = 0.0056 + 0.5 \,\mathrm{Re}_r^{-0.32} \tag{26}$$
Reynolds number of flow in riser:
$$\mathrm{Re}_r = \frac{2 \bar{\rho}_m \bar{U}_m r_2}{\mu} \tag{27}$$
Average mixture velocity in riser:
$$\bar{U}_m = \bar{U}_{sl2} + \bar{U}_{sg2} \tag{28}$$
$$\bar{U}_{sl2} = \frac{w_{L,in}}{\rho_L \pi r_2^2} \tag{29}$$
$$\bar{U}_{sg2} = \frac{w_{G,in}}{\rho_{G2} \pi r_2^2} \tag{30}$$

3.4 Gas flow model at the low-point

As shown in Fig. 3 (b), when the liquid level in the pipeline section is above the critical level ($h_1 > h_c$), liquid blocks the low-point and the gas flow rate $w_{G,lp}$ at the low-point is zero.
$$w_{G,lp} = 0, \quad h_1 \geq h_c \tag{31}$$
When the liquid is not blocking at the low-point ($h_1 < h_c$ in Fig. 3 (a)), the gas will flow from volume V_{G1} to V_{G2} with a mass rate $w_{G,lp}$ [*kg/s*]. From physical insight, the two most important parameters determining the gas rate are the pressure drop over the low-point and the opening area. This suggests that the gas transport could be described by a "orifice equation", where the pressure drop is driving the gas through a "orifice" with opening area of A_G (Skogestad 2009):
$$w_{G,lp} = K_G A_G \sqrt{\rho_{G1} \Delta P_G}, \quad h_1 < h_c \tag{32}$$
where
$$\Delta P_G = P_1 - \Delta P_{fp} - P_2 - \bar{\rho}_m g L_2 - \Delta P_{fr} \,. \tag{33}$$

4.8 Liquid flow model at the low-point

The liquid mass flow rate at the low-point can also be described by an orifice equation:
$$w_{L,lp} = K_L A_L \sqrt{\rho_L \Delta P_L} \,, \tag{34}$$
in which
$$\Delta P_L = P_1 - \Delta P_{fp} + \rho_L g h_1 - P_2 - \bar{\rho}_m g L_2 - \Delta P_{fr} \,. \tag{35}$$
In (Storkaas and Skogestad 2003) the free area for gas flow is calculated precisely using some trigonometric functions. For the sake of simplicity, a quadratic approximation is used in the proposed model.
$$A_G \cong \pi r_1^2 \left(\frac{h_c - h_1}{h_c} \right)^2, \quad h_1 < h_c \tag{36}$$
$$A_L = \pi r_1^2 - A_G, \tag{37}$$

3.5 Phase distribution model at outlet choke valve

In order to calculate the mass flow rate of the individual phases in the outlet model in equations (8) and (9), also mixture density at top of riser ρ_t, used in equation (7), the phase distribution at top of the riser must be known.

Liquid mass fraction at the outlet choke valve:
$$\alpha_{Lm,t} = \frac{\alpha_{L,t} \rho_L}{\alpha_{L,t} \rho_L + (1 - \alpha_{L,t}) \rho_{G2}} \tag{38}$$
Density of two-phase mixture at top of riser:
$$\rho_t = \alpha_{L,t} \rho_L + (1 - \alpha_{L,t}) \rho_{G2} \tag{39}$$
The liquid volume fraction at top of the riser, $\alpha_{L,t}$, can be calculated by the entrainment model proposed by Storkaas (Storkaas and Skogestad 2003), but the entrainment equations are complicated and make the model very stiff. Instead of entrainment, we propose a very simple relationship by using some physical assumptions. We use the fact that in a vertical gravity dominant two-phase flow pipe there is approximately a linear relationship between the pressure and the liquid volume fraction. The gradient of pressure along the riser can be supposed to be constant for the desired smooth flow regimes. Because of the assumed linear relationship, the liquid volume fraction also maintains approximately a constant gradient along the riser for the stable flow regimes.
$$\frac{\partial \alpha_{L2}}{\partial y} = \text{constant} \tag{40}$$
This assumption suggests that the liquid volume fraction at middle of the riser is the average of the liquid volume fractions at the two ends of the riser. On the other hand, the liquid volume fraction at middle of the riser is approximately equal to the average liquid volume fraction in the riser given by equation (23). Therefore,
$$\bar{\alpha}_{L2} = \frac{\alpha_{L,lp} + \alpha_{L,t}}{2} \,. \tag{41}$$
The liquid volume fraction at the bottom of riser, $\alpha_{L,lp}$, is determined by the flow area of the liquid phase at low-point:

$$\alpha_{L,lp} = \frac{A_L}{\pi r_1^2} \tag{42}$$

Therefore, liquid volume fraction at the top of the riser, $\alpha_{L,t}$, can be written as

$$\alpha_{L,t} = 2\bar{\alpha}_{L2} - \alpha_{L,lp} = \frac{2m_{L2}}{V_2 \rho_L} - \frac{A_L}{\pi r_1^2} \tag{43}$$

4. SIMULATION RESULTS

Different simplified models were simulated in Matlab and their tuning parameters were used to fit them to the OLGA reference model. Actually the only way to find the best values for the tuning parameters is trial and error and we tried to tune all of the models as good as possible. One approach has been proposed in (Di-Meglio, Kaasa et al. 2010) for tuning the Di Meglio model, but the best parameter values for the present case study are far from values calculted by this tuning method. Our most important criteria for the model fitting are critical value of the valve opening and oscillation frequency at this point. As shown in Table 1, at first we try to get the correct critical valve opening (5%) together with period time of oscillations at this point (15.6 min), and then we look at the other criteria.

4.1 Frequency of oscillations

The models were linearised at Z=5%. They demonstrated a pair of complex conjugate poles close to the imaginary axis, $\omega = \pm 0.0067i$ which are corresponding to the period time of the oscillations in this operating point:

$$\omega = \frac{2\pi}{T}, \; 0.0067 = \frac{2\pi}{15.6 \times 60}$$

The Eikrem model and the Nydal model are not able to get the right period time.

4.2 Step response

Fig. 4 demonstrates the response of the pressure at top of the riser to the step change is the valve opening form Z=4% to Z=4.2% for the OLGA reference model and the new model. Step responses of the OLGA model has one undershoot and one overshoot. The amplitudes of the overshoot and undershoot for different simplified models are given in Table 1 in the form of errors form those of the OLGA model. The inverse response (overshoot) is corresponding to the RHP zeros near the imaginary axis which are also given for Z=5% in Table 1.

4.3 Bifurcation diagrams

Fig. 5 shows the steady-sate behaviour of the new model (central solid line) and also the minimum and maximum of the oscillations compared to those of the OLGA model. In order to have a quantitative comparison, deviations of the different simplified models from the OLGA reference model for fully open valve (Z=100%) are summarised in Table 1.

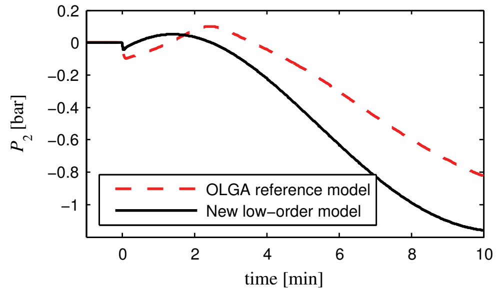


Fig. 4: Step response of the pressure at top of the riser

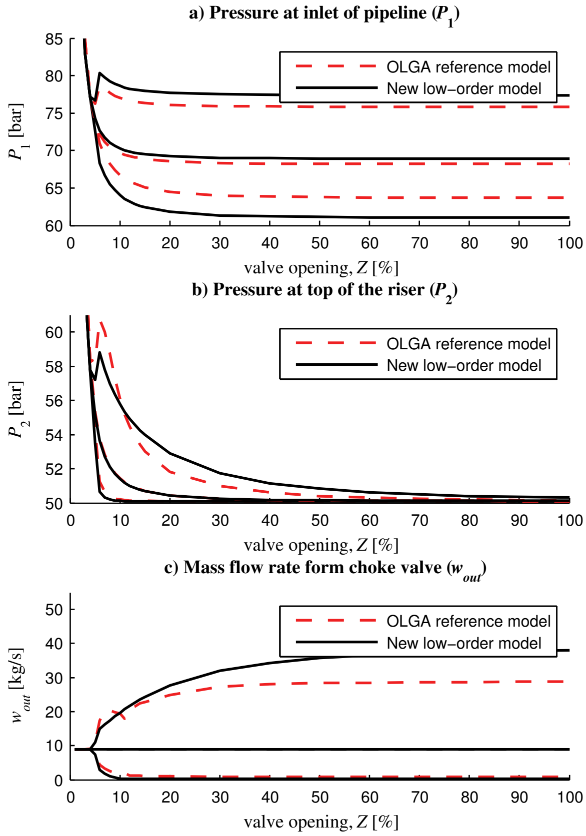


Fig. 5: Bifurcation diagrams of the new low-order model compared to those of OLGA reference model.

5. CONCLUSIONS

There is a trade-off between complexity of the models and the number of tuning parameters they use to match properly to a desired process. The simple models like the Kaasa model (with seven parameters) and the Di Meglio model (with five parameters) are able to demonstrate a good fit in expense of using more tuning parameters. However, finding parameter values is difficult. The Nydal model and the Eikrem model are simple and both use only three tuning parameters, but they are not able to match to the OLGA model. We included more physics with some simplifying physical assumptions into the new model in order to fit the model properly. Despite the other simplified models, the new model does not use any physical property of the system, such as volume of gas in the pipeline, as tuning parameter. Comparing quality of the different simplified models, the new model and the De Meglio model have approximately the same accuracy in prediction of the steady-state and also minimum and maximum values. But, the new model is able to demonstrate a better step response. We conclude that the proposed model maintains a good fit for steady-state and dynamics; therefore it will be used in our future works for controllability analysis and controller design. Also, the De Meglio model is quite simple and easy to use, and it can be considered as an alternative.

Parameters	**OLGA Model**	**Storkaas Model**	**Eikrem Model**	**Kaasa Model (only $P_{2,b}$)**	**Nydal Model**	**Di Meglio Model**	**New Model**
State equations	**Many**	3 diff + 1 alg.	4 diff.	3 diff.	4 diff.	3 diff.	4 diff.
Complexity	**Complicated**	Average	Simple	Very simple	Average	Simple	Average
Tuning Parameters	**Many**	5	3	7	3	5	4
Important unstable zeros of P_2 (Z=5%)	———	0.0146	.006 +.005i .006 -.005i	———	0.0046	.0186 + .034i .0186 - .034i	0.0413 0.0126
Values from OLGA simulator		**Error of the simplified models with respect to OLGA**					
Critical valve opening	**5%**	0 (0%)	0 (0%)	0 (0%)	0 (0%)	0 (0%)	0 (0%)
Period [min] (at Z=5%)	**15.6**	0 (0%)	1569 (168%)	0 (0%)	958 (102%)	0 (0%)	0 (0%)
Step response of P_2 undershoot overshoot at t=10 min	 **-0.098** **0.1987** **-0.824**	 0.097 (99%) 0.16 (80%) 0.43 (53%)	 0.076 (78%) 0.18 (89%) 0.62 (75%)	———	 0.118 (89%) 0.01 (50%) 0.25 (30%)	 0.067 (68%) 0.17 (84%) 0.47 (58%)	 0.053 (54%) 0.10 (51%) 0.33 (41%)
Steady-State P_1 [bar] P_2 [bar] w_{out}[kg/s]	 **68.22** **50.105** **9**	 1.9 (2.7%) .008 (.016%) 0 (0%)	 4.4 (6.46%) .009 (.017%) 0 (0%)	 for $P_{2,b}$ 0.02 (.034%)	 1.77 (2.6%) .011 (.021%) 2.9 (32%)	 0.70 (1%) .007 (.014%) 0 (0%)	 0.69 (1%) .008 (.016%) 0 (0%)
Minimum P_1 [bar] P_2 [bar] w_{out} [kg/s]	 **63.50** **50.09** **0.791**	 2.7 (4.3%) 4e-4 (8e-4%) 0.55 (69%)	 9 (14.2%) 5e-4 (9e-4%) 0.14 (17%)	 for $P_{2,b}$ 2.57 (4.1%)	 8.3 (13%) 0.41 (0.82%) 0.79 (100%)	 2.6 (4.1%) .003 (.006%) 3.3 (405%)	 2.7 (4.2%) 4e-4 (8e-4%) 0.55 (68%)
Maximum P_1 [bar] P_2 [bar] w_{out} [kg/s]	 **75.83** **50.14** **31.18**	 1.4 (1.8%) 1.5 (3%) 80 (278%)	 1.89 (2.5%) 0.95 (1.9%) 57 (200%)	 for $P_{2,b}$ 0.55 (.075%)	 1.00 (1.3%) 1.09 (2.2%) 13.3 (43%)	 1.3 (1.7%) 0.71 (0.35) 22 (77%)	 1.5 (2%) 0.33 (0.16%) 9.1 (32%)

Table 1: Comparison of the different simplified models to the OLGA reference model

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