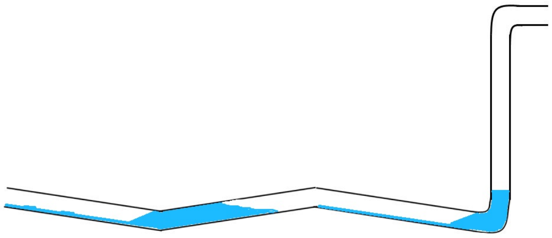

Severe Slugging in pipelines:
Modelling, Simulation and Mitigation

André Mendes Sancho

Thesis to obtain the Master of Science Degree in

Chemical Engineering

Supervisor(s): Prof. Dr. Henrique Aníbal Santos de Matos
Dr. Apostolos Giovanoglou
Dr. Yemi Zaccheus

Examination Committee

Chairperson: Prof. Dr. Sebastião Manuel Tavares da Silva Alves
Supervisor: Prof. Dr. Henrique Aníbal Santos de Matos
Member of the Committee: Prof. Dr. Vítor Manuel Geraldes Fernandes

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Para a rapariga que não ri...

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Abstract

The petroleum industry heavily relies on the simultaneous transport of gas and liquid phases in a single pipeline. Due to the pipeline-riser system configuration, severe slugging might occur. This phenomena is unwanted and it is important to have a multiphase dynamic model capable of accurately represent it.

A drift-flux model was developed with the purpose of predicting severe slugging. This dynamic and isothermal model based on the one dimensional conservation equations of mass and momentum used Shi correlation as the general slip law. The model was implemented in gPROMS, using the software internal implicit temporal discretization. For the spatial discretization it was developed a finite volume scheme with staggered grid, making this model numerically stable.

A comparison was made against experimental data from different literature and the state of the art software, OLGA, showing very good results for the prediction of the cycle time and severe slugging type.

Different mitigation strategies, such as gas-lift, increase in the separator pressure and pipeline design parameters, were studied. The model developed described correctly the behavior of such strategies.

Keywords: Severe Slugging, Pipeline-riser system, Drift-flux model, gPROMS

Resumo

A indústria petrolífera tem uma grande dependência do transporte em simultâneo de fase líquida e gasosa num único oleoduto. Devido à configuração do sistema de tubagens, o fenómeno de severe slugging pode ocorrer. Este fenómeno é indesejado e é importante ter um modelo multifásico dinâmico capaz de representá-lo com precisão.

Foi desenvolvido um modelo drift-flux com o objetivo de prever o severe slugging. Este modelo dinâmico e isotérmico baseado nas equações de conservação da massa e momento usou a correlação de Shi como a regra geral de deslizamento. O modelo foi implementado em gPROMS, usando a discretização temporal implicítica interna do software. Para a discretização espacial foi desenvolvido um esquema de volumes finitos com malha deslocada, a qual tornou o modelo estável numericamente.

Foi feita uma comparação com os dados experimentais provenientes de fontes literárias diferentes e do estado da arte do software, OLGA, mostrando resultados muito bons para a previsão do tempo de ciclo e tipo de severe slugging.

Diferentes estratégias de mitigação, como a injeção de gás, aumento da pressão no separador e design do oleoduto, foram estudados. O modelo desenvolvido conseguiu descrever corretamente o comportamento em todas as situações referidas.

Palavras-chave: Severe Slugging, Pipeline-riser system, Drift-flux model, gPROMS

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Nomenclature

Symbols

α_k	Holdup of phase k.
ρ_k	Density of phase k.
SS1	Severe Slugging Type I.
SS2	Severe Slugging Type II.
SS3	Severe Slugging Type III.
USO	Unstable Oscillations.
θ	Pipe inclination in relation to horizontal.
p	Absolute pressure.
u_k	Velocity of phase k.

Chapter 1

Introduction

The petroleum industry played a primary rule on the recent development in the multiphase transport. Being able to transport oil-gas-water together from the offshore collection site to an onshore separation facility immensely reduces the costs, for example taking the need of an offshore production platform whose cost might have been be prohibitive. This advancement in technology also allowed the opening of deepwater fields that were previously non-profitable or inaccessible. However, the transport of gas and liquid in the same pipeline also as its down sides, because it creates a highly complex and possible unstable flow.

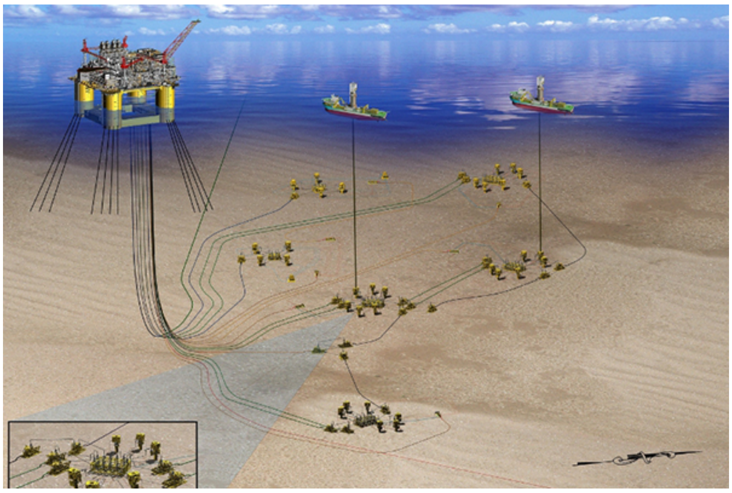


Figure 1.1: Appomattox deepwater field, water depth of 2200 m.

1.1 Motivation

A typical petroleum pipeline follows the terrain topography, having uphill and downhill sections (see Figure 1.1). The way that the different phase are spatially distributed, also know by flow regimes, are dependent not only on the flow-rates but also on the inclination of the pipe, witch adds another layer of complexity.

Under certain conditions, the liquid accumulates at the lowest points of the pipeline, as shown in Figure 1.2, until it's blown out afterwards by the compressed gas, leading to high instantaneous flow rates. This phenomena is know as Severe Slugging or Terrain-induced Slugging and has received enormous amount of attention, because it can lead to whole facilities shutdowns, if they were not design to account for it, or if mitigation strategies were not implemented.

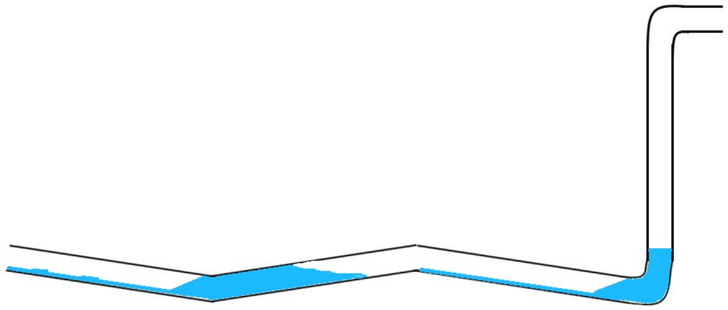


Figure 1.2: Liquid buildup during severe slugging.

Many companies model multiphase systems as a pseudo-homogeneous mixture. While this approach might be valid for a vast range of applications and even has some advantages over the multiphase counterpart like being more CPU efficient and numerically stable, it fails in being able to simulate certain transient phenomena like severe slugging. There is therefore the need for the development of a multiphase dynamic model capable of not only accurately representing the physical phenomena but also simple enough that it's feasible to run on the current computer technology.

Process Systems Enterprise (PSE), the world's leading supplier of Advanced Process Modelling technology, is a company highly recognized in the industry and has showed interest in expanding their knowledge in the multiphase flow area. The current work was developed in PSE's Oil & Gas department with the main objective being the development of a multiphase model suitable for severe slugging studies.

1.2 Scope of Work

The objective of the present work is understand the severe slugging phenomena and numerically study it with the purpose of having a model capable of predicting this phenomena and that can be integrated in the company libraries and solve projects on that field.

In order to do that, a drift-flux model was developed. This multiphase model is a dynamic, isothermal based on the one dimensional conservation equations of mass and momentum. It is numerically stable and capable of handling flow reversal and phase disappearance, two big numerical challenges.

The developed model is validated against experimental data from different authors and configurations. These results are also compared with models developed by other authors, and it's performance is tested against the industry standard multiphase simulator, OLGA.

Finally, some strategies for mitigating severe slugging are studied to demonstrate the capability of the model of applying such strategies.

1.3 Outline

In Chapter 2, an overview is first given of the basic concepts of two-phase flow, followed by a physical description of the severe slugging phenomena, and then review on the current modelling approaches and correlations on gas-liquid flow.

Chapter 3 presents the Schmidt model, explaining the governing equation and the needed closure correlations and ending with its applicability and limitations.

The main model developed is presented in Chapter 4, delineating the governing equations behind it as well as the correlations and submodels used.

In chapter 5 it's explained the numerical scheme that was used to build the model described in Chapter 4 with gPROMs in such a way that can circumvent some challenging numerical problems related to multi-phase flow.

In chapter 6 the model is validated against different experimental data, other researchers models and state of the art competitor software (ex: OLGA).

Strategies to mitigate severe slugging are discussed at Chapter 7 and the capability of the model at simulating those strategies is shown.

A sensibility analysis is done in Chapter 8 to see how some models parameters and discretization affects the predicted results and the influence of design aspects such as pipe diameter, length and inclination on severe slugging.

The main conclusions are drawn in Chapter 9 and some possible areas where future work can be done are suggested.

Chapter 2

Background

2.1 Gas-Liquid Flow Concepts

The multiphase flow, or more specifically gas-liquid flow, can be encountered in a wide number of industries like chemical, refrigeration, nuclear and petroleum.

Understanding the dynamics of gas–liquid multiphase flows is critical from an engineering and scientific point of view and it has received a lot of attention. Knowing the behavior of the gas-liquid flow, void fraction, pressure drop and heat transfer is essential in the design and optimization of industrial processes and equipment, such as the design of a single pipeline capable of transporting oil and gas together from the offshore field to the onshore facility or a a heat exchanger where some currents change phase.

Much of the theory behind gas-liquid flow comes from the nuclear industry, where the coolant boiling in the reactor core leads to a gas-liquid flow [1]. For safety reasons, it's very important to be able to accurately simulate the two-phase flow so that tests like lost of coolant accident (LOCA) and safety codes can be done in order to prevent disastrous accidents like Chernobyl [2] and more recently Fukushima [3].

One of the biggest difficulties in gas-liquid flow is that it can vary in form depending on a wide number of variables, such as velocities of each phase, densities, pipe inclination, etc. Two well known and opposite cases of this type of flow are the transportation of liquid as droplets by the gas media, forming a film around the pipe, and the transport of gas as small bubbles dispersed in the liquid.

2.1.1 Flow Regimes

Taitel & Duckler [4] proposed to categorize the flow regimes into three types: dispersed, separated, and intermittent.

The cases previously described are denominated as annular flow and bubble flow, respectively, and both are dispersed flows, the liquid droplets being the dispersed phase in the first case and in the second the gas bubbles. With small gas and liquid flow rates and horizontal or downhill pipe stratified flow happens, where both phases travel separated. If the flow rates are increased liquid will temporally reach the top wall of the pipe, forming elongated bubbles or slugs. Figure 2.1 represents typical flow regimes that can occur in a horizontal or near-horizontal gas-liquid flow. The slug flow regime is not to be confused with severe slugging, as the first forms hydrodynamics slugs that last seconds while the second forms much more dangerous slugs that happen during severe slugging, which can last for several hours [1]. Figure 2.2 shows some regimes of an air-water mixture in a horizontal pipe.

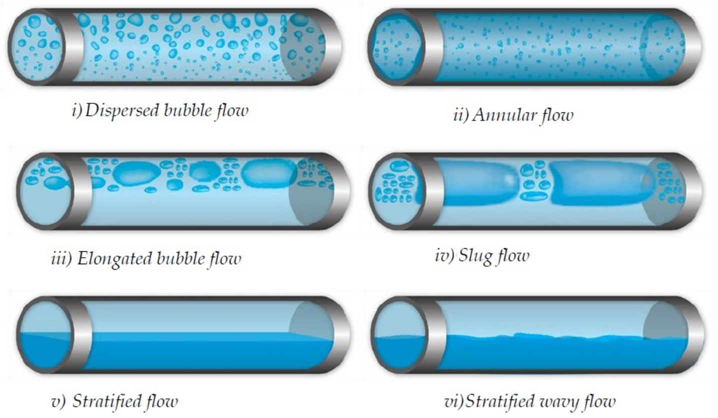


Figure 2.1: Typical flow regimes in a horizontal pipe [1].

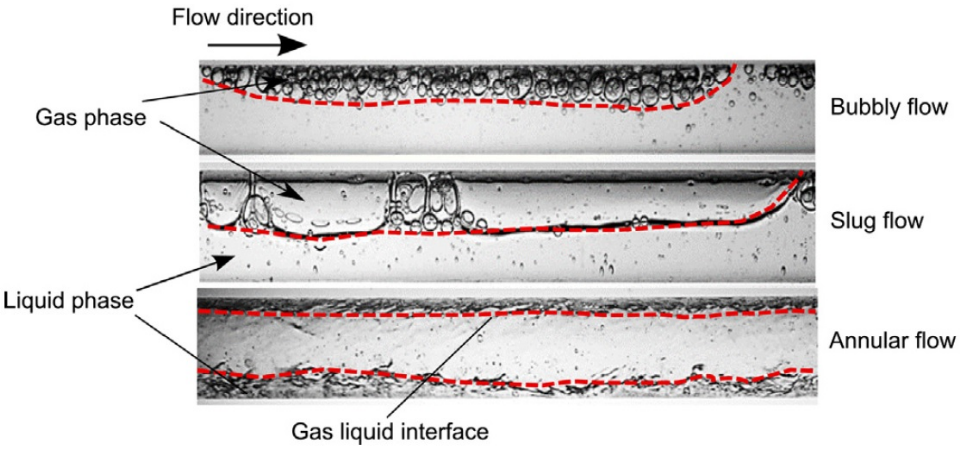


Figure 2.2: Gas-liquid flow in a horizontal pipeline. (image courtesy OSU, Two Phase Flow Lab)

Not all flow regimes are the same when the pipe inclination change. For example, in a vertical pipe, stratified flow can not occur while other flows like Churn flow appear.

It is useful to know in what regime the flow is at, since different regimes have different pressure drops or mass and heat transfer between phases. A practical way to see this is with flow regime maps, where the regimes are plotted depending alone on the superficial velocity of the gas and liquid. Figure 2.3 illustrates a typical flow regime map for a horizontal pipe.

It is important to refer that flow maps are only valid for their data set, as the regime also depends on others factors as pressure, temperature, pipe inclination, pipe diameter, mixture composition and others.

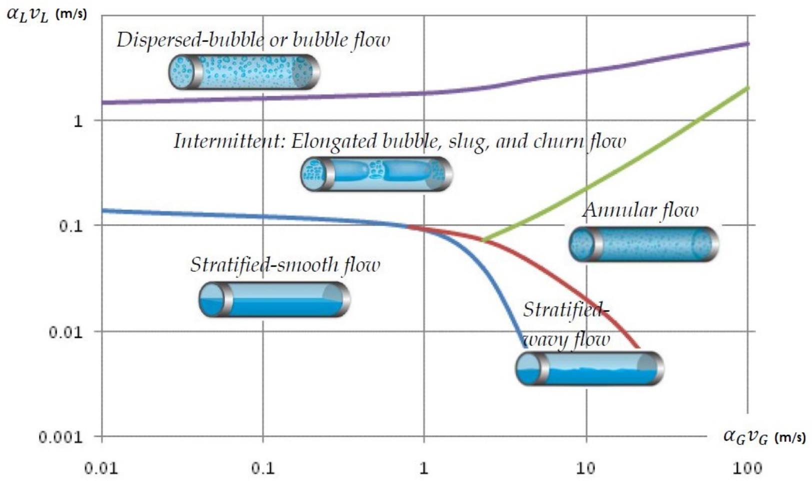


Figure 2.3: Typical flow regime map for a horizontal pipeline [1].

2.2 Severe Slugging

Severe slugging is a cyclical phenomena that might happen in pipelines with sections with different inclination, characterized by the accumulation of liquid at certain areas of the pipe and generation of long liquid slugs that are followed by a fast gas blowdown.

This phenomena was first reported by Yocum [5] and it's prone to happen on petroleum offshore facilities, where usually there are downward sections on the pipe system, denominated for now on as pipelines, and upward sections, denominated as risers. The key phenomena behind severe slugging are the liquid buildup at the bottom of the riser, local flow reversal and local phase disappearance.

The existence of severe slugging is a major issue for the production facilities as it's responsible for the following:

- Increase of pressure at the wellhead, which causes production losses.
- Oscillations on the reservoir flow caused by the unsteady system.
- High instantaneous outflow of liquid and gas to the separator, see Figure 2.4, which leads to large oscillations in the separator control system and might cause separator flooding, compressor trips and even production shutdown.

It is important to have a model that can simulate the severe slugging behavior correctly. Some of the big benefits of modelling severe slugging are:

- Know at which conditions severe slugging is going to occurs.
- Being able to determine the slug length and period, which are important for the design and control of the downstream facility.
- Being able to perform simulations of different scenarios where mitigation strategies are employed.

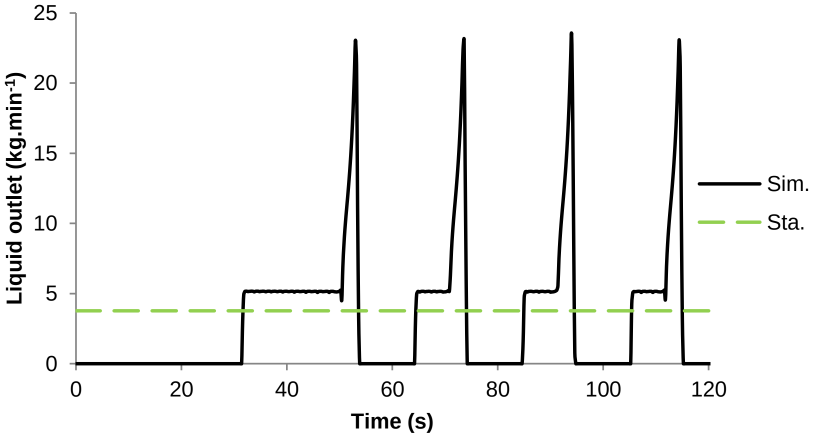


Figure 2.4: Comparison of liquid output during severe slugging and stable flow.

2.2.1 Severe Slugging Classification

Severe slugging has been experimentally studied by several investigators [6, 7, 8, 9] at a laboratory scale to better understand it's characteristics. Schimidt [6] was the first to divide the severe slugging cycle into four main stages, see Figure 2.5:

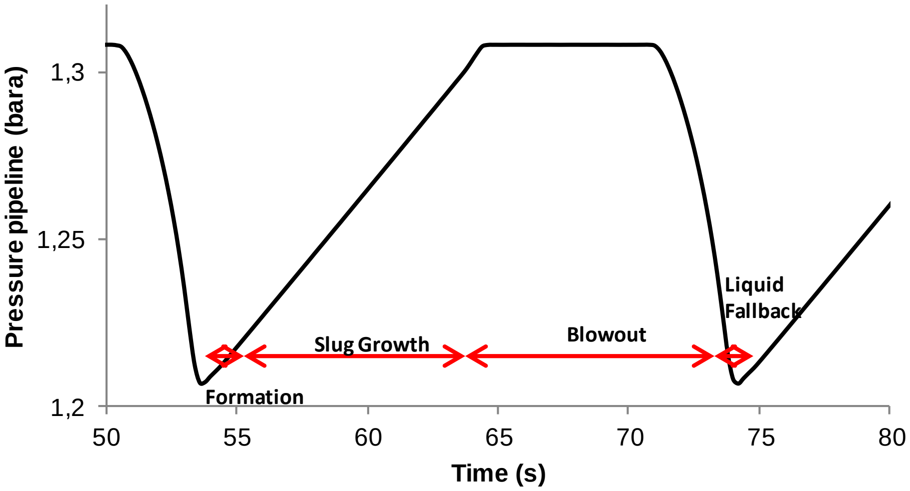


Figure 2.5: Severe slugging stages.

1. Slug formation: If there is accumulation of enough liquid at the bottom of the riser it will block the riser entrance to gas, generating a liquid slug. This initial liquid buildup can also arise as a result of liquid fall-back from the riser and transient hydrodynamic slugs from the pipeline.

2. Slug growth: The liquid level in the riser will increase as the slug grows. The gas is the pipeline will be compressed until its pressure becomes greater than the hydrostatic head of the liquid slug (other pressure drop terms neglected in explanation).

3. Blowout: The compressed gas will expand as it pushes the liquid out of the riser. According to

Malekzadeh [9], this stage should be divided in two to better distinguish between different types of severe slugging:

   (a) Liquid production: If the liquid slug is bigger than the length of the riser then when the slug reaches the top of the riser the liquid will start to flow out with the gas pushing the slug tail from the pipeline to the riser.
   (b) Fast liquid production: When the compressed gas reaches the bottom of the riser, the hydrostatic head in the riser will decrease, making the gas expand and push the liquid out of the riser rapidly.

4. Liquid fall-back: The gas is expelled at a high rate, which will cause a quick system depressurization. When system reaches its minimal pressure the small liquid amounts that still remains in the riser will fall-back to the bottom.

Severe slugging can also be divided according to certain characteristics like slug length or riser blockage. Balino [10] divided severe slugging in the following categories : severe slugging I (SS1), severe slugging II (SS2), severe slugging III (SS3) and unstable oscillations (USO). The different types of severe slugging can be distinguished between each other as follows:

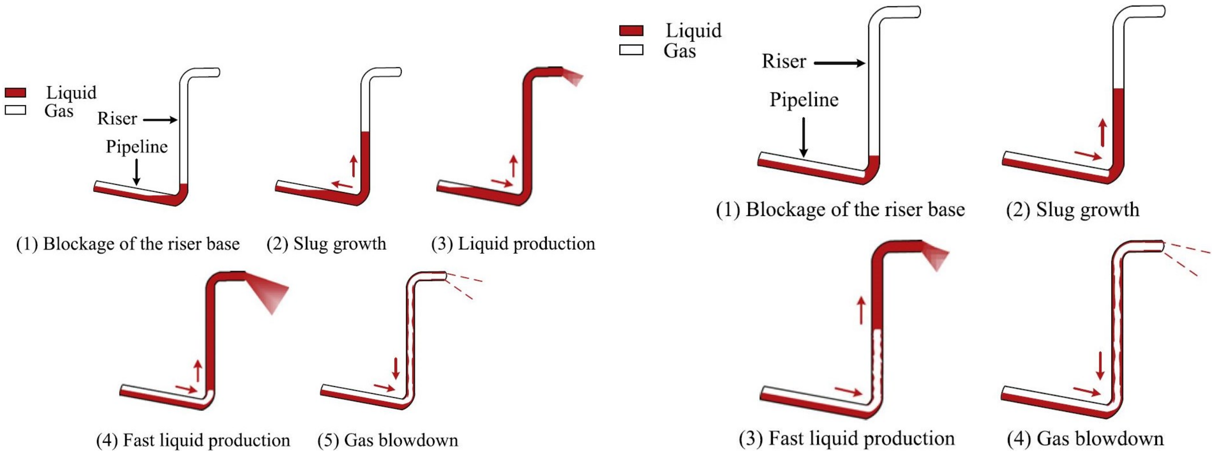
- Severe slugging I (SS1) : The maximum pressure at the bottom of the riser is equal to the hydrostatic head of the riser filled with liquid (neglecting other pressure drop terms) and the liquid slug length is equal or bigger than the riser length (see Figure 2.6(a)).
- Severe slugging II (SS2) : The liquid slug length is smaller than the riser length and there is a full blockage of the bottom of the riser until the blowout phase (see Figure 2.6(b)).
- Severe slugging III (SS3) : The bottom of the riser is never fully blocked so gas can still pass. Pressure and slug length are smaller compared to severe slugging I (see Figure 2.6(c)).
- Unstable oscillations (USO) : In this regime both gas and liquid flow into the riser and there isn't a vigorous blowdown. This type is not even considered severe slugging by some as it usually as very small pressure oscillations compared to the other types.

2.2.2 Severe Slugging Mitigation

Since severe slugging affects the profitability and safety of a facility, much time as been spent on studying ways of eliminating or mitigating it.

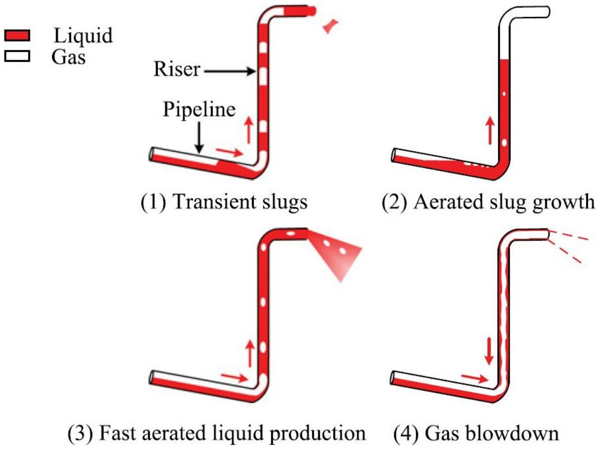
Yocumm [5] demonstrated that by increasing the separator pressure steady-state could be reached, at the cost of lower production flow-rates. Schmidt [6] and Jasem [11] suggested choking has a way to eliminate severe slugging by increasing the back pressure proportionally to the velocity at the choke. This however will also decrease the production rate leading to a premature closing of the field.

The injection of gas in the system in order to decrease the back pressure by decreasing the hydrostatic head of the riser, also known as gas-lift, is capable of reducing or even eliminating severe slugging.



(a) Severe slugging I

(b) Severe slugging II



(c) Severe slugging III

Figure 2.6: Different types of severe slugging.

One of the biggest drawback of this method is that it needs a large volume of gas to completely eliminate severe slugging [11] and some authors say it's too expensive [6, 5].

Tengesdal [12] managed to eliminate severe slugging by using self-lifting method, where the gas in the pipeline is collected and bypassed to the riser, thus reducing the hydrostatic head in the pipeline. Huawei [13] also successfully eliminated severe slugging in a pipeline+riser system using a bypass-pipe method.

2.3 Model Approaches

The first model of severe slugging was done by Schmidt [6]. This model can simulate the growth of a liquid slug for a downward pipeline + vertical riser system. However, the applications of these model, as explained on Chapter 3, are limited. Despite all the shortcomings, the Scmidt model is capable of predicting cycle times with some approximations and it was further improved by Schmidt [14] by accounting with the liquid production and gas blowdown stages.

Malekzadeh and Balino [9, 15] numerically generated stability maps and the stability curves by running simulations to show the gas and liquid flow rates the different severe slugging regime changes.

2.3.1 Conservation Models

A more general approach, also used in the development of this work model is to model, the gas-liquid flow using one dimensional equations of conservation of mass, momentum and energy. Although some researchers have been able to simulate severe slugging using CFD, higher dimensions are currently restricted to research work due to the computational time it needs.

These models can be categorized depending on how you model each phase, and most can be divided in two groups: Two-fluid models and Mixture models. In the two-fluid approach both gas and liquid have their own conservation equations, while in the mixture approach, at least one of the conservation equations are considered to be a mixture equation with mixture variables. Some state of the simulators like LedaFLOW use Three-fluid models, where they model the water and oil as two different liquid phases [16].

Mixture Models

Sometimes the gas-liquid flow can be approximated to a pseudo single-phase flow that has properties determined by the amount of gas and liquid in the mixture. One of the most simple model and widely used is the pseudo-homogeneous mixture model where all the conservation equations are for the pseudo mixture-fluid. This is only valid for very dispersed flow regime, as it assumes that there is no slip between the phases, meaning that the gas and liquid have the same velocity. With the correct correlations the validity of the model can be extended but since it can't track separately the buildup of liquid or gas it is not suitable to predict severe slugging.

Another mixture model that was idealized by Zuber [17] and later developed by Wallis [18] is the drift-flux model. This model is similar to the two-fluid model with the difference being the use of only one momentum equation, for the pseudo mixture-fluid.

A simplified version of this model is to impose no slip condition. Until now, the described models aren't appropriate to simulate severe slugging as the velocities of gas and liquid during this phenomena are very different at certain times and might even be of opposite directions. The use of a slip law, also known as drift-flux correlation, is of high importance to the results of the model.

Masella, Malek and Osiptsov [19, 20] have successfully used a drift-flux model to simulate severe slugging in different conditions. Some commercial multiphase simulator like TACITE and ECLIPSE are also based on this approach.

Balino, Taitel and Fabre [10, 7, 8] have developed specific models for severe slugging where the pipeline is modeled with one set of equations, usually lumped, and the regime is considered to be stratified. The riser is modelled with a drift-flux correlation. The switch between both models usually happens with a liquid slug front tracking scheme.

Two-Fluid Models

With the two-fluid model approach it's possible to have a more rigorous and realistic model. This approach as been for long used in the nuclear safety codes like RELAP, TRAIC and SIMMER [21].

Bediksen [22] developed a two-fluid model, named OLGA, a dynamic multiphase simulator that is now considered the standard in the industry.

DeHenau [23] developed his own model two-fluid model with severe slugging in mind in a pipeline with many downhill and uphill sections in mind.

Two-fluid models are strongly based on regime flow maps in order to calculate the friction factor and the pressure difference between phases. The modelling of the inter-phase momentum transfer is not well established [23] and the velocities when a phase disappears, which happens during severe slugging, become unbounded, creating numerical problems.

Brevik [24] tried to model severe slugging using a two-fluid model on gPROMs, but failed due numericall instability issues. This comes up that numerical problems must be addressed for this type of phenomena in order to have a stable model and solution.

Drift-flux vs. Two-Fluid Models

Much discussion has been heard about whether or not the drift-flux approach can have the same accuracy as the two-fluid approach.

Some authors even advocate that the drift-flux approach for the momentum equation should be a better choice once the correct correlations are developed [25]. Two-fluid models have present in their equations terms that are difficult to define and get correlations for, such as the interfacial shear stress or the interfacial area, which are very hard to define in some regimes, resorting also in the use of correlation to estimate the variables. Another numerical problem that arises in the two-fluid models is the numerical discontinuities when there is a change in flow regime.

In the Drift-flux approach there is no need to model the interaction between phases as those terms cancel each out in the mixture momentum equation. Also, due to being a simpler model than two-fluid model, the drift-flux model should be faster.

In the end, both model approaches require correlations whether directly as the slip law on the drift-flux approach correlation or more indirectly as the interfacial friction factor, in the more mechanistic two-fluid approach. Both Malezadeh and Osiptsov [20] compared a drift-flux model with a two-fluid model and reached the conclusion that both models are capable of predicting severe slugging accurately and there is space to improve the correlation used for both models.

Numerical Schemes

Multiphase models have long been known to have numerical challenges, from phase appearance or disappearance [26] to solving of the transient conservation equations, a problem that also appear in single-phase flow [21, 27].

When velocity and pressure and located in the same node, this often lead to non-physical behavior of the model due to the velocity-pressure coupling in the momentum equation. Harlow and Welch [28] proposed a solution for the problem by using a staggered grid arrangement where scalar variables are defined at the center of the control volume and vectorial variables like velocity at the faces.

Different numerical schemes and solvers have been applied for both drift-flux and two-fluid models model [29, 30, 31], from upwind-scheme type to AUSM-scheme type.

The handling of these numerical problems in gPROMs is further explained in Chapter 5.

2.3.2 Stability Maps

The stability map a plot defined by superficial liquid velocity towards superficial gas velocity, and it is possible to show regions where different types of severe slugging can occur as well as the regions of stable flow. A general Stability Map can be seen in Figure 2.7. and it's major benefit is to easily tell if severe slugging will occur. Similarly to the regime maps, this map is only valid for the experimental setup and corresponding conditions.

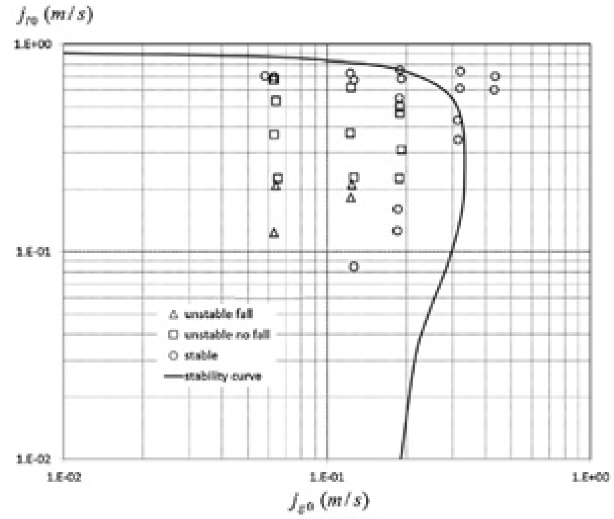


Figure 2.7: Typical stability map [15].

Stability Criterions

Generating a stability map for a specific setup or running simulations for different flow rates can take a long time. Any quick idea if severe slugging will occurs, without more details, could be of interest. The stability critirions are simple expressions that only need input conditions and the holdup in the pipeline in order to determine if severe slugging happens.

Boe [32], based on Schimdt's model, developed a condition that predicts whether severe slugging is going to happen based on inlet flow rates, pipe configuration and liquid holdup. Beniksen [put ref of technical paper] developed an improved version of Boe's critireon that accounts for the riser inclination and gas density. Taitel [7] also proposed another criterion and Jansen [33] extended it to include the effect of a choke valve at the top of the riser.

2.4 Drift-flux Correlations

The drift-flux model only as one momentum equation, so there is a need for an additional algebraic equation in order to be able to solve the model. This closure law also know as slip law relates the velocity of the gas to that of the the liquid, see eq. 4.10.

The slip law started out as a correlation for regimes where there is strong motion coupling between phases. However more recently Ishii and Ghajar [34] have showed that by changing the parameters to account for the flow patterns it is possible to extend the correlation to separated flow regimes.

The parameters for the correlation started as constants [35] and became more and more complex over the last decades, being now modeled as function of densities, Froude number, pipe orientation and others [34]. With the rise of many drift-flux correlations there was need for knowing what were the top performing correlations. Ghajar [34] did an extensive comparison against 8255 data point from more than 60 sources and found out that his own correlation and Choi [36] correlation were the top performing.

Malekzadeh and Osiptsov use Shi's [37] correlation in their model of severe slugging. The parameters for the correlation suggested by the author were already optimized for Oil & Gas industry data and they can be fitted to experimental data.

Another important point about both Ghajar and Shi correlations is that they are regime free and continuous, so there isn't a need for a mechanistic model to determine on which regime the flow is as that is already incorporated in the correlations, making the model less complex.

Chapter 3

Schmidt Model

Schmidt [6] was the first to model the severe slugging phenomena. He created an hydrodynamic model, represented in Figure 3.1, based on fundamental physical principles that represent the buildup of a liquid slug both on the pipeline and riser, a behavior exclusive of severe slugging type I. Although simple, this model provided information on the time that the liquid slug took to reach the top of the riser, given us some information about the cycle time.

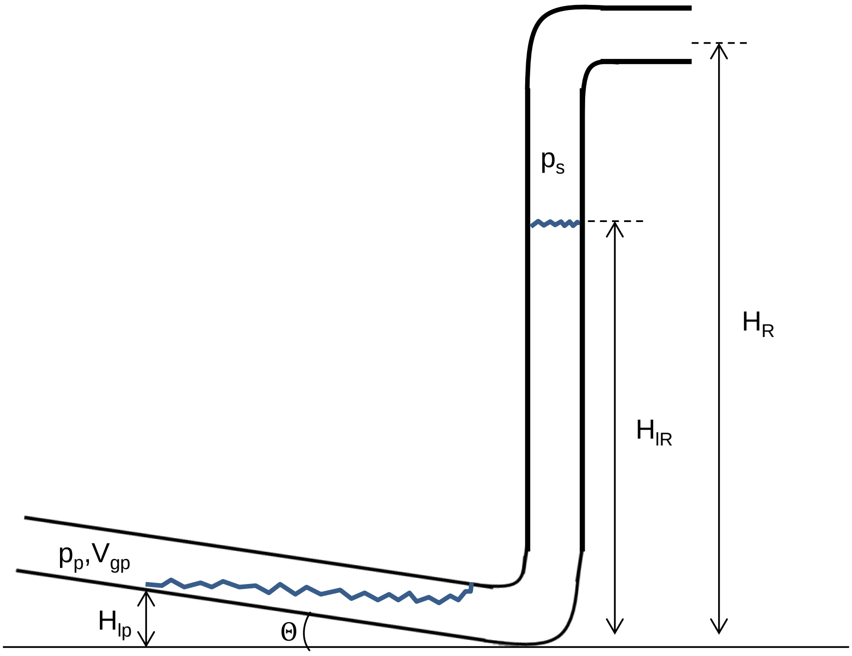


Figure 3.1: Hydrodynamic model of severe slugging.

To develop is model Schmidt had to do many simplifications and assumptions:

- Both liquid and gas mass flow rates remain constant.
- The separator pressure is constant.
- The only pressure drop that exists is due to the hydrostatic head of the slug.

- The generated slug does not contains any entrained gas.
- All the properties in the pipeline don't change with axial position.
- Gas liquid interface in the pipeline is horizontal.
- The pipeline is isothermal.

3.1 Governing Equations

The gas phase can be considered to behave according to the ideal gas expression, with the compressibility factor z.

$$pV = \frac{zmRT}{M} \tag{3.1}$$

By differentiating Eq.3.1 in respect to time we get

$$p_p \frac{\mathrm{d}V_{gp}}{\mathrm{d}t} + V_{gp} \frac{\mathrm{d}p_p}{\mathrm{d}t} = \frac{zRT}{M_g} \frac{\mathrm{d}m_g}{\mathrm{d}t} \tag{3.2}$$

Where the right side of Eq.3.6 is constant, as $\mathrm{d}m_g/\mathrm{d}t$ is the gas inlet mass flow rate that was assumed it doesn't change with time.

By neglecting the frictional and inertial pressure drop the pressure balance can be written as

$$p_p = p_s + \rho_l g \left(H_{lR} - H_{lp}\right) \tag{3.3}$$

Where the separator pressure, p_s, is constant and $H_{lR} - H_{lp}$ is the difference between the liquid slug height on the riser and the liquid slug height on the pipeline, respectively. By differentiating with respect to time eq.3.3 we get 3.4,

$$\frac{\mathrm{d}H_{lR}}{\mathrm{d}t} - \frac{\mathrm{d}H_{lp}}{\mathrm{d}t} = \frac{1}{\rho_l g} \frac{\mathrm{d}p_p}{\mathrm{d}t} \tag{3.4}$$

The change of the liquid volume in the pipeline and riser is equal to the liquid inlet flowrate and the liquid fallback from a previous severe slug cycle. This is represented on 3.5, where $A\frac{1-\alpha_l}{\sin\Theta}$ is the liquid area in the pipeline.

$$A\frac{\mathrm{d}H_{lR}}{\mathrm{d}t} + A\frac{1-\alpha_l}{\sin\Theta} \frac{\mathrm{d}H_{lp}}{\mathrm{d}t} = A\alpha_l u_l + A\frac{\mathrm{d}F_b}{\mathrm{d}t} \tag{3.5}$$

Where A is the cross section area of the pipeline. As the slug level increases in the pipeline, the volume occupied by gas decreases according to

$$\frac{\mathrm{d}V_{gp}}{\mathrm{d}t} = -\frac{A(1-\alpha_l)}{\sin\Theta} \frac{\mathrm{d}H_{lp}}{\mathrm{d}t} \tag{3.6}$$

3.2 Closure Correlations

In order to solve the system of equations shown above it is necessary to have closure correlations. In this case, the governing equations contain six unknown variables, so there is a need for two closure correlations, one for the liquid holdup and other for the liquid fallback.

Liquid holdup

For low gas and liquid flow rates, that are needed for severe slugging to occur, the mixture in the pipeline travels in stratified flow. This is a typical regime in negative inclination pipelines and resembles open-channel flow. Schmidt suggested then to use the open-flow correlations to predict the liquid holdup in the pipeline.

The liquid velocity in a open-channel flow is calculated according to Manning correlation (Eq. 3.7) .

$$u_l = 149 \left(\frac{A_l}{P_l}\right)^{\frac{2}{3}} (\sin\Theta)^{0.5} \quad \text{(ft/s)} \tag{3.7}$$

The liquid area and the liquid wetted perimeter are given by Eqs.3.8 and 3.9, respectively.

$$A_l = \frac{d^2}{4}(\delta - 0.5\sin\delta) \tag{3.8}$$

$$P_l = d\delta \tag{3.9}$$

Where d is the pipeline diameter and δ is half angle subtended by liquid level chord In order to solve this system of equations and calculate the liquid holdup we just need to remind ourselves that the liquid area is the liquid holdup times the pipeline cross section area, $A_l = \alpha_l A$.

Liquid Fallback

After the blowout stage of the severe slug cycle most of the liquid is expelled to the separator. The remaining liquid, however, falls back into the bottom of the riser. The liquid fall-back is the length of a liquid slug that was formed only with the liquid that as fallen back to the riser.

Schmidt found that the amount of liquid fall, F_b, back was only a function of the gas superficial velocity, u_{sg}, and for is experimental setup the following correlation was obtained.

$$F_b = -7.71 + 58\left(u_{sg} + \frac{1}{u_{sg}}\right) \quad \text{(ft)} \tag{3.10}$$

Looking at the correlation it's possible to observe that the liquid fall-back only depends on the gas flow rate. This is due to the fact that in the blowdown stage the riser is receiving only the compressed gas that was in the pipeline, making the liquid flow rate play little to none role on the fall-back amount.

To extend eq.3.10 to other riser heights Schmidt suggested multiplying the fallback by the ratio between the new riser height and the original riser height. The fallback expression is then represented by 3.11.

$$F_b = -7.71 + 58\left(u_{sg} + \frac{1}{u_{sg}}\right)\frac{H_R}{50} \quad \text{(ft)} \tag{3.11}$$

Note that liquid fallback is likely to change according to the experimental configuration, and factors such as pipe diameter will affect the gas superficial velocity and were not studied by Schmidt.

During his experiments, Schmidt found that all the fall-back was occurs within the first 10s of a new cycle. The liquid fallback relation with time is given by eq. 3.12

$$\frac{\mathrm{d}F_b}{\mathrm{d}t} = \frac{F_b}{10}1.26t^{-0.8} \quad \text{(ft/s)} \tag{3.12}$$

Initial Conditions

In order to solve the system of differential equations initial conditions must be provided. In the beginning there is no slug in the system, and it's assumed that the pressure drops are negligible. The initial conditions are summarized in Eq. 3.13.

$$\begin{aligned} H_{lR} &= 0 \\ H_{lp} &= 0 \\ p_p &= p_s \\ V_{gp} &= V_p \end{aligned} \tag{3.13}$$

3.3 Applicability and limitations

Being the first model available on severe slugging, Schmidt model can predicts the cycle time, or at least, the growth of the slug time. The steady state regime could be assumed if the cycle time of the simulation is smaller than 5 s.

This model also served as base for Boe's Critireon [32], the first simple expression to tell whether severe slugging is going to occur.

Moreover, the model is rudimentary and simple, with some of it's biggest limitations listed bellow.

- Simulates the behavior of Severe slugging I (SS1), meaning that it's not possible to distinguish the different severe slugging types.
- Only works for downward inclined pipeline followed by vertical riser configuration. In actual offshore pipelines systems there is always more uphill and downhill sections that follow the topography of the sea bottom.
- The input flow rates need to be constant, which also doesn't happen in actual offshore pipelines systems.

- The liquid fall-back correlation only works for the particular dataset used by Schmidt. It would be needed to do create a new correlation that could account for important factor like pipe diameter and riser length.
- It always simulates the slug growth, meaning that it's not possible to tell whether there is severe slugging happening in reality or not.
- It's not possible to incorporate mass transfer between phase or mitigation strategies like gas lift or choking.

Chapter 4

Drift-flux Model

The mathematical model developed in the current work for simulating severe slugging is presented is this chapter. First, the governing equations are derived and explained. Then the flow properties model is chosen and validated against a third party package. Afterwards, the drift-flux closure relation is described and discussed, with emphasis on it's numerical advantages. Finally, two multiphase homogeneous friction models are presented and compared.

4.1 Governing Equations

The drift-flux model is an hybrid between the one-pseudo fluid model and two-fluid model. It is based on the mass conservation equation for each phase and a momentum equation for the mixture. The energy equation can be for each phase or for the mixture. However, the system is assumed to be isothermal, eliminating the need for including the energy equation. This was done because all the experimental cases cases are isothermal.

This model rose in popularity is mainly due to being more simple than the two-fluid model, more numerically stable, faster, but at the same time being able to deliver results with the same accuracy.

The derivation of these one dimensional conservation equation as been done by many authors, so the author in this papers encourages the reader to see the work of Bratland [1] for that purpose.

The mass equations for the gas and liquid phase are described bellow, respectively :

$$\frac{\partial \alpha_g \rho_g}{\partial t} + \frac{\partial \alpha_g \rho_g u_g}{\partial z} = \Gamma_{gl} + \Gamma_{gw} \,. \tag{4.1}$$

$$\frac{\partial \alpha_l \rho_l}{\partial t} + \frac{\partial \alpha_l \rho_l u_l}{\partial z} = \Gamma_{lg} + \Gamma_{lw} \,. \tag{4.2}$$

Where α_i, ρ_i and u_i are the volume fraction, density and velocity of phase i, respectively. The terms on the right side of Eq. 4.1 and 4.2 represent the mass flow rate transfer per volume from other sources. Γ_{iw} represents the quantity of phase i that entered the system through perforations in the pipe wall.

The other term represents the mass transfer due to phase change. The change of phase does not change the total mass, so the quantity that one phase lose must be the same as the other phase gains. For example, the amount of gas that is lost due to condensation must be the same amount of liquid that is gained. This leads to this simple relation:

$$\Gamma_{lg} + \Gamma_{gl} = 0\,. \tag{4.3}$$

The mass transfer between phases term is given by the simulation to determine how much condensation or boiling takes place at a certain temperature, pressure and composition. This term is usually given by third-party software or another models due to it's complexity in real mixture like petroleum.

The volume fraction of the phase i, α_i, is the fraction of the pipe cross section that phase i occupies. The sum of all fractions must then fill the pipe. Eq.4.4 express this relationship:

$$\alpha_g + \alpha_l = 1\,. \tag{4.4}$$

The momentum conservation equation usually is the source of numerical problems and solver failures, mainly due to the high non-linear coupling between the pressure and the velocity, as it will be further discussed in Chapter 5. A more stable version of the momentum equation is to neglect the dynamic momentum term. Viviand [19] showed that this approximations is valid when the velocities of the phases are smaller than the sound velocity, which is the case for severe slugging.

$$\frac{\partial p}{\partial z} = R_{fric} + R_{grav}\,. \tag{4.5}$$

Where the variables on right side terms represent the pressure drop due friction and gravity, respectively.

$$R_{fric} = -\frac{2f}{d}\rho_m u_m |u_m|\,. \tag{4.6}$$

Eq.4.6 describes the pressure drop based on mixture properties and the mixture fanning factor, f. The last is defined more ahead depending on the friction model used. The mixture density, ρ_m, and the mixture velocity, u_m are defined, respectively, by:

$$u_m = \alpha_g u_g + \alpha_l u_l\,. \tag{4.7}$$

$$\rho_m = \alpha_g \rho_g + \alpha_l \rho_l\,. \tag{4.8}$$

The pressure drop due to gravity effects, R_{grav}, taking in account the pipe inclination, is expressed as:

$$R_{grav} = -\rho_m g \sin\theta\,. \tag{4.9}$$

Since there is only a momentum equation for both phases, there is need for a correlation that relates the velocity of the liquid to that of the gas. In this model this is done using a slip law, whose generic form is given by Eq.4.10.

$$u_g = C_0 u_m + u_{drift}\,. \tag{4.10}$$

Where C_0 is the distribution parameter and u_{drift} the drift velocity. These variables are the ones that affect how the velocity of both phases relate, meaning that the multi-phase flow regimes are imbued on then. A more in depth explanation on the slip law used is given on the next section.

4.1.1 Physical Properties

In the chemical area almost every model takes the physical properties with utmost importance and the final results of the simulations greatly depend on the correct prediction of this properties, such as density, viscosity and enthalpy.

These properties are normally a function of temperature, pressure and composition. Usually, as the petroleum industry, the mixture is complex making this properties hard to predict correctly. In that case, specialized third party software are usually used.

Multiflash

gPROMs has the benefit of being able to use third-party software easily through the Foreign Object interface [foreign object guide]. This enables the model developer to worry more about the main model development and call the physical properties with ease from the external package.

MultiFlash$^{\text{TM}}$is state of the art third-party package developed by KBC and is the primary physical properties package used on gPROMs models and libraries. It is capable of accurately predict the properties of complex mixtures for the petroleum.

Within MultiFlash$^{\text{TM}}$, it is possible to define which model should be used for estimating the physical properties. The model selected was the Peng-Robinson revised 1978 equation of state to model the gas phase physical properties. This equation is a cubic equation of state and is suitable for ideal or near-ideal system, being widely used in oil and gas industry.

However, using an external package will slow down the simulation time. The computational time will be greatly affected since gPROMs, at every timestep, has to pass down information (pressure, temperature, composition) to Multiflash through the Foreign Object interface, and then Multiflash must calculate the physical properties and give it back to gPROMs.

Another solution is define in the model the physical properties without using a external package. A good approximation for the gas density is the use the ideal gas law, with the compressibility factor, z. It can be rewritten as:

$$\rho_g = \frac{p}{R_g T}\,. \tag{4.11}$$

Where R_g is the individual gas constant, give by,

$$R_g = \frac{zR}{M_g} \, . \tag{4.12}$$

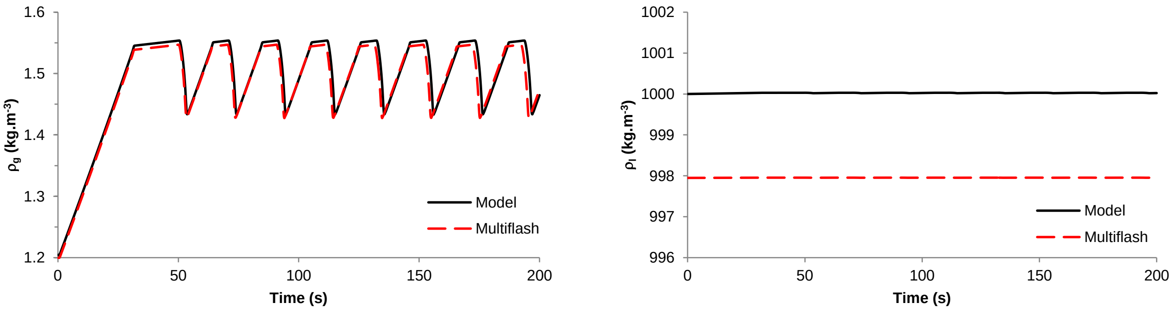
Since the model is isothermal, and it's assumed that there is no mass transfer, the density of the liquid phase will only change due to the pressure. One way to express this relation is:

$$\rho_l = \rho_{l,0} + \frac{p - p_{l,0}}{a_l^2} \, . \tag{4.13}$$

Where $\rho_{l,0}$ is the reference liquid density at the reference pressure, $p_{l,0}$ and a_l is the sound velocity in the liquid phase.

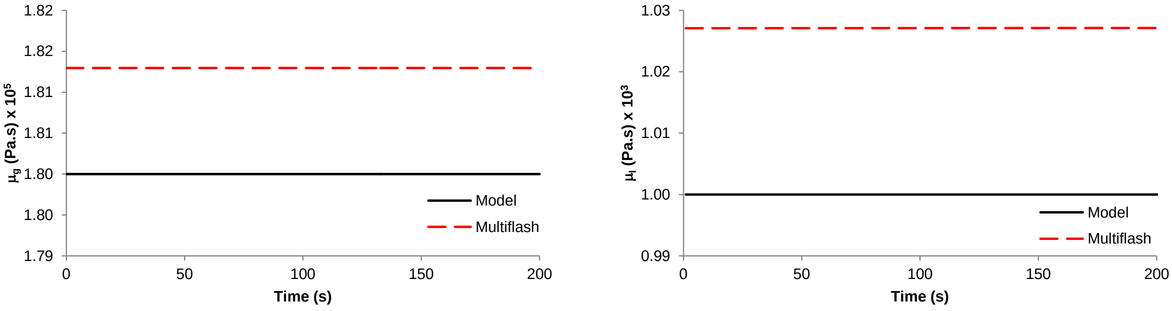
Other physical properties like the viscosity or the surface tension are assumed to be constant parameters.

Although much simpler, these expressions suffice to accurately predict the physical properties when the liquid phase is water and the gas phase is air, as table 4.1 and Fig4.1 show. So, unless otherwise specified, these expressions are used to predict the physical properties instead of MultiFlash$^{\text{TM}}$, as properties predicted using this expressions are similar to that of MultiFlash$^{\text{TM}}$, as shown on table 4.1.



(a) Gas density profile.

(b) Liq density profile.



(c) Gas viscosity profile.

(d) Liq viscosity profile.

Figure 4.1: Properties profiles during severe slugging.(Taitel, Table 4.1)

4.2 Shi Correlation

In order to solve the model there is need for a closure correlation. The slip law, a correlation that relates the velocities of both phases with the general form of eq. 4.10 is used.

Table 4.1: Physical properties of the mixture

		Period (s)	$\bar{\rho_g}$(kg.m^{-3})	$\bar{\rho_l}$(kg.m^{-3})	$\bar{\mu_g}$ (Pa.s) $\times 10^5$	$\bar{\mu_l}$ (Pa.s) $\times 10^3$
Taitel	Present work	20.4	1.49	1000	1.81	1.03
	Multiflash	20.3	1.49	998	1.80	1.00
	Deviation (%)	-0.5	-0.4	-0.2	-0.7	-2.7
Malek	Present work	115.3	3.00	1000.1	1.80	1.00
	Multiflash	115.6	2.99	1000.1	1.79	1.14
	Deviation (%)	0.3	-0.2	0.0	-0.5	12.4

Although there is many correlations available in literature [34], most of then are not suitable as they are developed for specific flow regimes or limited to horizontal and vertical inclinations. Another shortcoming is that the parameters in the correlations were determined experimentally using small diameter pipes and using air-water, which are not suitable for the petroleum industry.

Shi [37] developed a correlation that is regime independent and the parameters were optimized using an oil and gas mixture in industrial size pipelines. Although the model developed in this work was validated against an air-water mixture in small diameter pipeline, the slip correlation developed by Shi was chosen as the model developed will be used with petroleum mixtures in industrial projects. Another advantage of this correlation is that it is possible to fit the parameters to experimental data, being able in this way to tune the model for each project.

The relationship between the velocities can be described as a combination of two mechanism, as shown on eq. 4.10. The distribution parameter, C_0, represents the distribution of gas over the pipe cross section and also acknowledge that the concentration profile and the velocity of the mixture profile can vary independently of each other across the pipe section. The other mechanism represents the tendency of the gas phase to rise vertically due to buoyancy effects.

Distribution Parameter

The distribution parameter peaks on bubbly and slug flow regime, reaching a value of 1.2. As the void fraction increases the distribution parameter approaches unity. For the case that the void fraction is unity, that is, there is only gas, the distribution parameter is 1.0. The distribution parameter is expressed according to:

$$C_0 = \frac{A}{1 + (A-1)\gamma^2}\,. \tag{4.14}$$

Where A is the value of the distribution parameter on bubble and slug flow regimes and γ is a term that makes C_0 reduce to 1.0 at high values of void fraction or mixture velocity, and is defined by:

$$\gamma = \frac{\beta - B}{1 - B}\,. \tag{4.15}$$

where β approaches 1.0 at high values of void fraction or mixture velocity. B is the value of void fraction at which the the distribution parameter drops below A.

$$\beta = MAX\left(\alpha_g, F_v \frac{\alpha_g u_m}{u_{sgf}}\right) . \tag{4.16}$$

Shi choose the transition to the annular regime to eliminate the phase slip velocity. This transition occurs when the gas superficial velocity is higher than the flooding velocity, defined in eq. 4.17, being sufficient drag the liquid film.

$$u_{sgf} = \alpha_g K_u (\frac{\rho_l}{\rho_g})^{0.5} u_c \,. \tag{4.17}$$

Where u_c is the characteristic velocity, defined in eq. 4.18, and K_u is the critical Kutateladze number, which is related to the inverse of the adimensional pipe diameter, La, according to eq. 4.19:

$$u_c = \left[\frac{\sigma_{gl} g \left(\rho_l - \rho_g\right)}{\rho_l}\right]^{\frac{1}{4}} . \tag{4.18}$$

$$K_u = \begin{cases} 3.2 & \text{if } La \leq 0.02 \\ 0 & \text{if } La \geq 0.5 \\ 12.6La^2 - 13.1La + 3.41 & \text{else} \end{cases} \tag{4.19}$$

The inverse of the adimentional pipe diameter is given by

$$La = \left[\frac{\sigma_{gl}}{g\left(\rho_l - \rho_g\right)}\right]^{0.5} \frac{1}{d} \tag{4.20}$$

Where σ_{gl} is the superficial tension the the mixture.

Drift velocity

The vertical rise of the gas bubbles due to buoyancy effects is accounted on the slip law by the drift velocity term. It can be expressed as:

$$u_{drift} = \frac{(1-\alpha_g)C_0 K u_c}{(\frac{\rho_g}{\rho_l})^{0.5}\alpha_g C_0 + 1 - \alpha_g C_0}\Phi(\theta) \,. \tag{4.21}$$

Where K is a term that ramps down the flooding curve at low void fractions in order to account for the bubble rise and is defined by:

$$K = \begin{cases} \frac{1.53}{C_0} & \text{if } \alpha_g \leq a_1 \\ K_u & \text{if } \alpha_g \geq a_2 \\ \frac{1.53}{C_0} + \left(K_u - \frac{1.53}{C_0}\right)\frac{\alpha_g - a_1}{a_2 - a_1} & \text{else} \end{cases} \tag{4.22}$$

The change between curves are done by the ramping parameters a_1 and a_2.

To account for other inclination that are not vertical there was need to add the following correction term to Eq. 4.23.

$$\Phi(\theta) = n_1(\Theta) \cos\Theta^{n_2} \left(1 + \sin\Theta\right)^{n_3} \ . \tag{4.23}$$

The correction term developed by Shi was only valid for inclination ranging from 90° to 30°. The correction factor was modified in order to extend the correlation to account for negative inclination by changing the drift velocity sign to negative [38] as eq. 4.24 shows.

$$\Phi(\theta) = n_1 \text{sgn}(\theta) |\sin\theta|^{\ n_2} \left(1 + |\cos\theta|\right)^{n_3} \ . \tag{4.24}$$

It should be noted that the parameters used in this expression stayed the same. Further studies on the effects of inclination on the drift velocity are advised.

Shi parameters ā

Table 4.2: Shi correlation default parameters.

Parameter	A	B	F_v	a_1	a_2	n_1	n_2	n_3
Default Value	1	0.3	1	0.06	0.21	1.85	0.21	0.95

4.3 Friction Model

The pressure drop due to friction usually plays an important role on the transport over long pipelines. For multiphase flow the friction factor of the mixture depends on what type of flow regime is happening. Wallis [18] simply considered a mixture average Reynolds number and then used the Coole-Brook equation is order to determine the pressure drop. Garcia [39] estimated a friction factor that accounts for the different flow regimes based on a modified Reynolds velocity.

Since for the phenomena of severe slugging the pressure drop due to friction is much smaller compared to the gravitational pressure drop the Wallis [18] model was implemented. The fanning friction factor was calculated according to:

$$f = \begin{cases} \frac{16}{Re} & \text{if } Re \leq 2300 \\ \frac{.0055}{4} \left[1 + \left(2 \times 10^4 \frac{\varepsilon}{d} + \frac{10^6}{Re} \right)^{1/3} \right] & \text{else} \end{cases} \tag{4.25}$$

Where Re is the adimentional Reynolds number of the mixture, defined as:

$$Re = \frac{\rho_m u_m d}{\mu_m} \ . \tag{4.26}$$

With the mixture viscosity being an volume average of the velocity of each phase $\mu_m = (\alpha_g \mu_g + \alpha_l \mu_l)$.

Chapter 5

Numerical Scheme

The one dimensional transport equations described in the previously chapter form a set of partial differential equations. In order to solve it,an adequate numerical method is required.

The modelling of multiphase phase flow as long been known for have numerical issues. Some came from single phase flow like the velocity-pressure coupling in the momentum equations, while others like numerical discontinuities when changing flow regimes are exclusive for multiphase flow. The phenomena of severe slugging brings another layer of complexity and numerical challenges because there is local flow reversal of each phase separately as well local phase disappearance.

In the current chapter is presented the discretizations methods that were used for the implementation of the model in gPROMS language. Later it is given a short summary of how the model was build in gPROMS platform.

5.1 Discretization Methods

The drift-flux model variables are function of both time and space. There is need to discretize both the temporal and spacial domain with suitable methods.

The temporal discretization is done by DAE solver, a internal gPROMS solver that uses an implicit scheme. The use of implicit scheme makes the solver more robust and usually allow it to that larger time steps than the explicit counterpart.

Although gPROMS also has some discretization methods like finite differences, they are not suitable for reversible flow. So unlike the temporal discretization where the method was already implemented, there was need to develop and implement a discretization method capable of handling flow reversibility in gPROMS language.

5.1.1 Finite Volume Method

The set of PDE were discretized spatially by using the finite volume method. In this method, the variables are calculated at the center of a control volume having in account the values that pass the volume face (boundaries).

Staggered Grid

Due to the velocity-pressure coupling in the momentum equation, if both variables are defined at the node a cell it can give rise to non-physical simulations. Harlow [28] used a staggered grid for the velocities as a solution for this problem. In this approach the velocities are defined using a different control volume, and the node where the velocity is defined matches the face of the cell of the normal grid, as shown on figure 5.1.

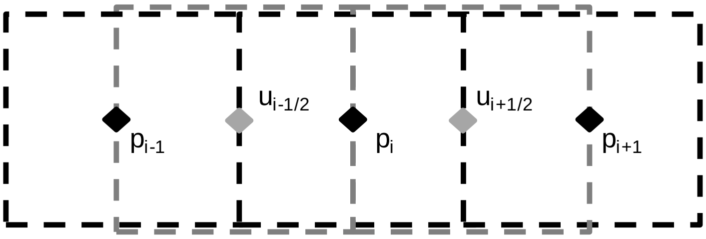


Figure 5.1: Staggered grid.

The momentum equation is discretized over the staggered grid domain while the continuity equations are discretized over the normal domain. Figure 5.2 is and example of a pressure field that illustrates well the problem with the co-allocated approach.

If the momentum equation is discretized using the co-allocated grid approach the pressure gradient is approximated by:

$$\frac{\partial p_i}{\partial z} = \frac{p_{i+1/2} - p_{i-1/2}}{\Delta z} = \frac{\left(\frac{p_{i+1}+p_i}{2}\right) - \left(\frac{p_{i-1}+p_i}{2}\right)}{\Delta z} = \frac{p_{i+1} - p_{i-1}}{2} \,. \tag{5.1}$$

It is possible to see that the pressure gradient doesn't depend on the pressure field at the point. Looking at the pressure values of Fig 5.2, the pressure gradient using Eq. 5.1 would be zero, even if the it's clear that the pressure changes. However, if the momentum equation is discretized using the staggered grid approach, the pressure gradient is defined as:

$$\frac{\partial p_{i+1/2}}{\partial z} = \frac{p_{i+1} - p_i}{\Delta z} \,. \tag{5.2}$$

It is possible to see that the use of staggered grid avoids the unrealistic behaviour of the momentum equation that occurs when the pressure spatial profile is of the checker-board type, like the one shown on 5.2. Another big advantage of using the staggered grid approach is that the velocities are in the correct location for the calculation of the cell properties as it will become evident in the next section.

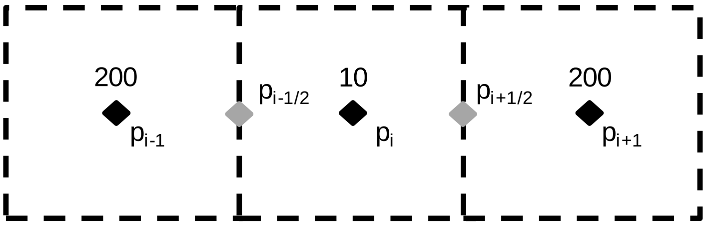


Figure 5.2: Checker-board type pressure field.

Cell-surface quantities

Since properties like the densities and fractions of both phases are at the position z_i, there is need for estimating this properties for the $z_{i+1/2}$. An intuitive away of doing this is by simply average between the two closest points,

$$\psi_{i+1/2} = \frac{\psi_i + \psi_{i+1}}{2}\,. \tag{5.3}$$

This approach leads to oscillations near regions with sharp gradients leading to numerical artifacts that can make the simulation to fail. A solution for this problem is approximating the term using a upwind scheme. This scheme causes strongly diffused solutions, so there is need to use a higher resolution on zones of high gradient. However, this scheme is much more robust and stable due to his diffusional part.

In the upwind scheme, also known as donor-cell scheme, the property to be approximated at $z_{i+1/2}$ is either the value at the node behind or the node ahead, depending on a condition. In this case, the condition is the direction of the specific phase flow. Eq. 5.4 shows an example of how the density of gas is calculated at the cell boundary:

$$\rho_{g\;i+1/2} = \begin{cases} \rho_{g\;i} & \text{if } u_g \geq 0 \\ \rho_{g\;i+1} & \text{if } u_g < 0 \end{cases} \tag{5.4}$$

Chapter 6

Model Validation

Several researchers have conducted experimental investigations on severe slugging [7, 23, 9] with the objective of better understand and the phenomena.

In this chapter the performance of the model developed in the present work is compared with different experimental studies performed, each of them having their own pipeline-riser configuration an all using air and water as work fluids.

The model is also compared with others models and the industry standard OLGA.

Important information like pressure profiles during severe slugging and the stability maps are also discussed.

The validation of the model against simple and well known fluids at isothermal conditions was preferred because it allows the testing of the model performance without having to worry about the physical properties.

6.1 Pipeline-Riser System

Taitel [7] studied severe slugging occurrence in a downward pipeline connected to a vertical riser, a typical setup for a offshore production facility. The experimental setup, extensively explained in is work, consists in a buffer tank where only gas passes and gives an additional length of 1.69m to the pipeline, followed by a 9.1m pipeline with an inclination of -5° and a then by a vertical riser of 3m. Figure 6.1 shows the gPROMS representation of the experimental setup.

Driscretization Strategy

For obtaining the correct cycle time, that is, the time it takes for the slug to generate, grow and being expelled, the system must be correctly discretized, as showed further on chapter 8, since zones with high transient behavior demand a more refined grid.

Although all the system could have an uniform and refined grid, this would take much more time. In this way, the pipeline itself is subdivided in two parts: the stratified region, where there isn't to much change in the void fraction, and a slug region, that is the region where the slug might grow in the pipeline.

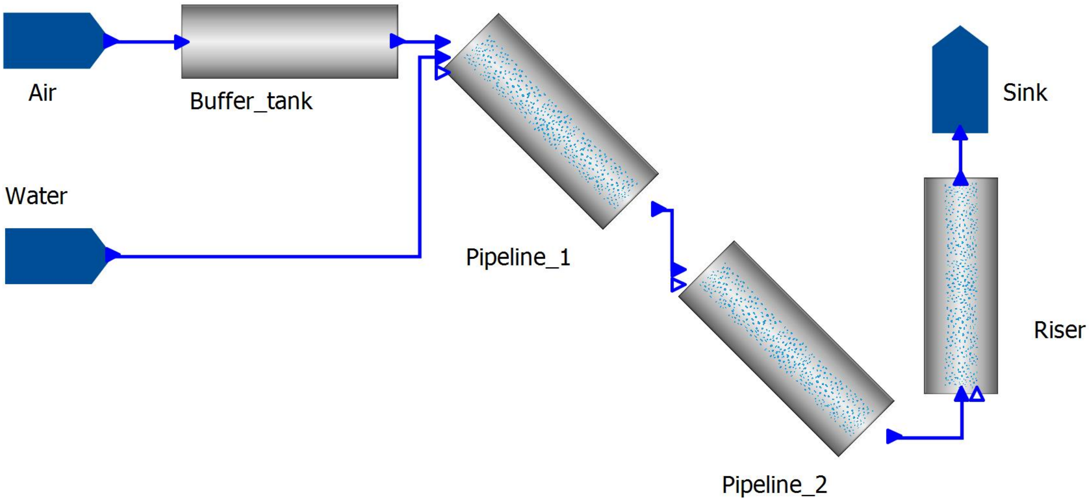


Figure 6.1: Taitel experimental setup.

This regions where chosen by running first a simulation with a smaller resolution just to approximately see how much of the slug could penetrate the pipeline. After that, the high transient zone was discretized in intervals of 5 cm, this latest value needs to be small enough. This value is obtained as the maximum allowed that doesn't change the results (see Figure 8.1 for graphical example). Table 6.1 gives a summary of the configuration used.

Table 6.1: gPROMS configuration of Taitel experimental setup.

	Buffer_tank	Pipeline_1	Pipeline_2	Riser
Length (m)	1.69	8.1	1	3
N^o of cells	4	16	20	60
Inclination	0	-5^o	-5^o	90^o
Diameter (cm)	2.54	2.54	2.54	2.54

The experiments were done for different inlet conditions and the separator pressure was 1 bar. Table 6.2 shows the experimental results of the different cases.

Others researchers including Taitel also developed their own severe slugging models and compared it against Taitel experimental points. Figure 6.2 shows the relation between the measured and simulated period of severe slugging for the drift-flux model while Table 6.3 shows a summary comparison between the different models in order to compare the drift-flux performance.

It is possible to see that the drift-flux model developed in the present work is capable of predicting within a small margin of error the cycle time of severe slugging when it exists an can also predict that there won't be severe slugging, and the discrepancy between the stable cases reported is explained further ahead.

The Schmidt model, as it was expected, didn't predict stable flow as the model always assumes the existence of severe slugging and it also is the worse at predicting the cycle time.

The pressure changes with time in the pipeline allow to better understand and categorize severe slugging. Figure 6.3 shows the pressure at the end of the pipeline for case 1 for both models developed

Table 6.2: Taitel experimental data.

Case	u_{sg0}	u_{sl}	t_{exp}(s)
1	0.063	0.124	24
2	0.064	0.209	20
3	0.123	0.183	15
4	0.124	0.212	14
5	0.062	0.679	6
6	0.063	0.367	13
7	0.063	0.679	9
8	0.064	0.535	10
9	0.065	0.226	19
10	0.122	0.374	11
11	0.123	0.621	8
12	0.126	0.228	13
13	0.187	0.226	11
14	0.188	0.466	8
15	0.188	0.502	7
16	0.19	0.312	10
17	0.058	0.705	Stable
18	0.063	0.698	Stable
19	0.122	0.73	Stable
20	0.126	0.673	Stable
21	0.126	0.085	Stable
22	0.184	0.127	Stable
23	0.185	0.161	Stable
24	0.187	0.551	Stable
25	0.188	0.755	Stable
26	0.19	0.685	Stable
27	0.313	0.433	Stable
28	0.314	0.347	Stable
29	0.319	0.614	Stable
30	0.321	0.744	Stable
31	0.43	0.604	Stable
32	0.433	0.701	Stable

Table 6.3: Summary of results for Taitel data.

	Experimental	Drift-flux	Schmidt	Balino [15]	Taitel [7]
Period Average Error (%)	-	6.3	15.9	4.9	13.8
Stable Cases	16	7	0	6	18

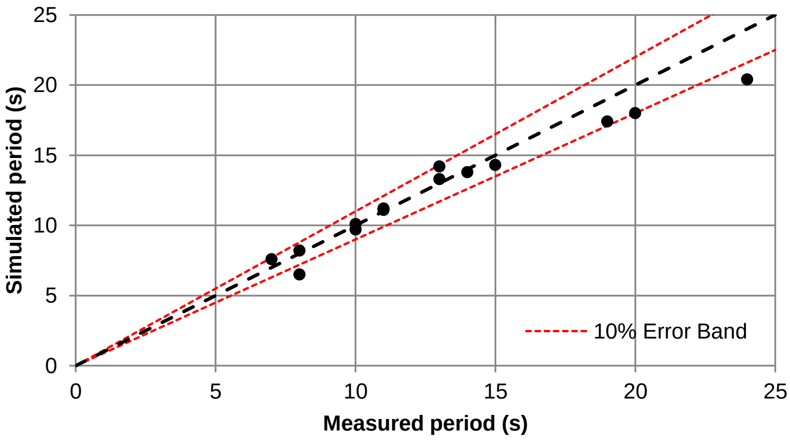


Figure 6.2: Comparison between experimental a simulated period.

in the present work.

As it was stated back in Chapter 3, the Schmidt models can only simulate the growth of a liquid slug, which correspond to the pressure increase in the pipeline. For here on out, only the drift-model is going to be used as it is more accurate and general than the Schmidt model.

After the first few cycle the system enters a periodic state, with shorter periods than the first cycle. This is because in the begging there was only gas in the pipeline, but once the slug being expelled there is some liquid fall back that contributes to generation of the next slug.

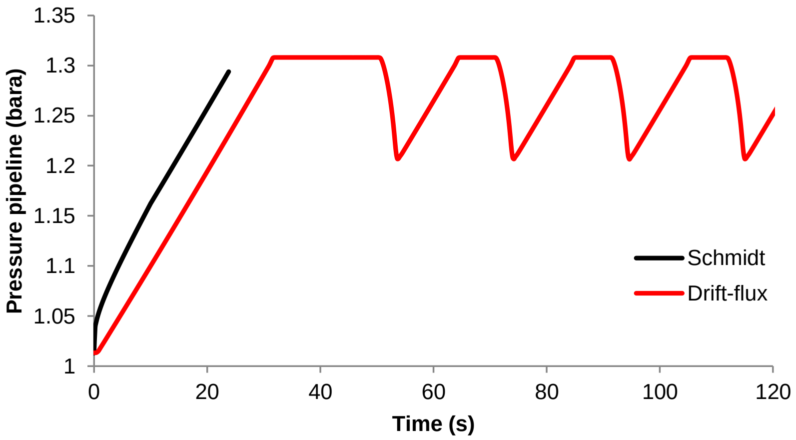


Figure 6.3: Pressure profile. (Taitel case 1)

Besides the higher pressure drop caused by during severe slugging, other maybe even more important consequence is the intermittent inlet that the separator downstream receives. Figure 6.4 shows the simulation results of the liquid outlet with time for case 1.

As expected, it is possible to see in Fig. 6.3 that liquid production starts to happen at the same time the pressure reaches it's maximum, this is when the slug reaches the top of the riser. It is then pushed by the compressed gas at a steady rate until the gas penetrates the bottom of the riser and accelerates due to the pressure drop decrease, spiking the liquid production for a few seconds before the gas slows

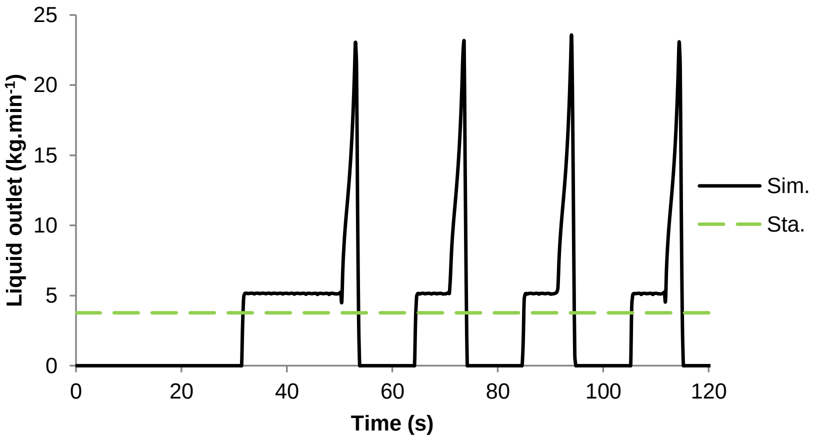


Figure 6.4: Liquid production profile.

down and the remaining liquid fall back to the bottom of the riser again.

As mentioned in Chapter 2, severe slugging can be categorized as type I, II or III based on the slug length and riser blockage. By looking at the pressure plot or the liquid outlet plot it clear that it's severe slugging type I, because the is full blockage of the riser and the slug is bigger than the riser length. If the pressure at the begging of the riser hits a plateau after the slug as reached the top of the riser it means that there is still part of it being pushed out of the pipeline.

The slug length time profile can be used to determine the maximal slug length. Figure 6.5 shows the slug size profile for case 1.

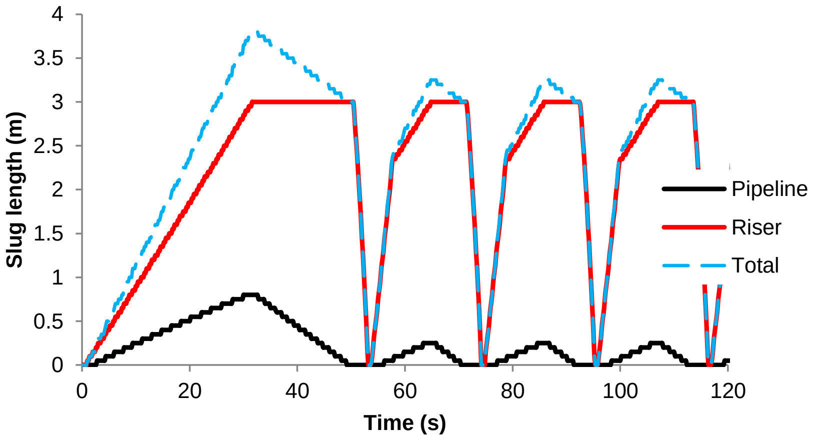


Figure 6.5: Slug length profile. (Taitel case 1)

It is possible to see that maximum slug length is around 3.5 m and. This confirms a severe slugging type I as the riser length is 3 m.

To better understand where the liquid slug starts to form and how it behaves figure 6.6 shows the liquid distribution in function of the axial position at different times, with the vertical blue line representing the end of each section of the system configuration written on gPROMS (Fig. 6.1), and the grid resolution (number of cells) is not the same for every sector as it's possible to see on Table 6.1. As an example, Buffer_tank has 1.69 m and is the first 4 cells, while Pipeline_2 has 1 m and goes from 20 to 40 in the plot.

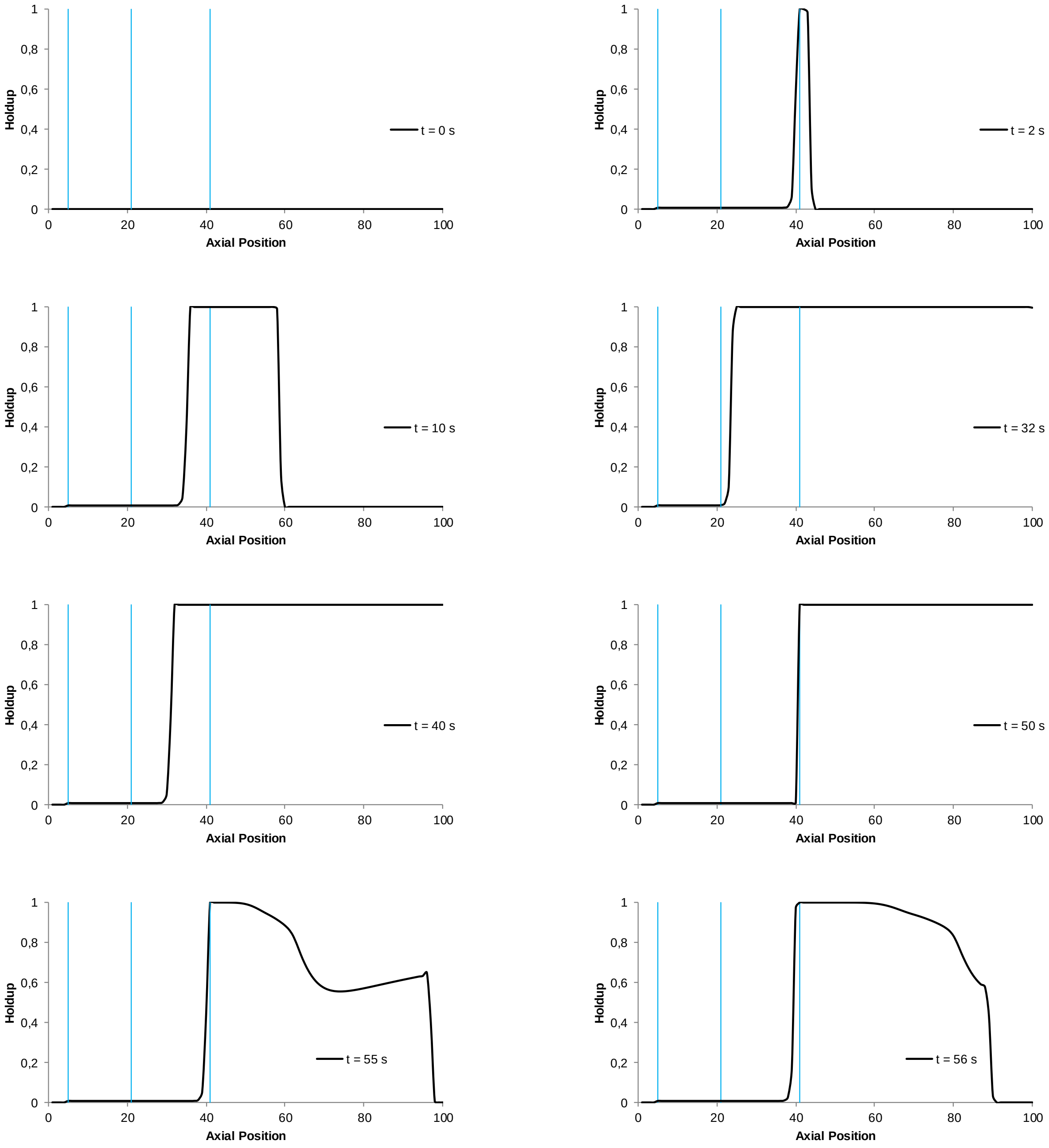


Figure 6.6: Liquid gas axial distribution. (Taitel case 1)

Initially the pressure in all the system is almost the same as the separator pressure as there is only gas($t = 0$s). The liquid then start to accumulate in the bottom of the riser, blocking it($t = 2$s). In this case, the liquid slug grows both into the riser and pipeline($t = 10$s). At $t = 32$s, the slug reaches the top of the riser and it's maximum length. The compressed gas starts pushing the slug out($t = 40$s) of the pipeline. When the gas penetrates the riser($t = 50$s), it greatly accelerates due to the decrease of the hydrostatic liquid head quickly blowing out the liquid. This is followed by a quick depressurization($t = 55$s)and slowing of the gas, making the remaining liquid fall back into the riser($t = 56$s). This last part of liquid production that only lasts about 5 seconds in the case corresponds to the peak in liquid production of Fig. 6.4.

Stability Map

As it is possible to see by in table 6.2, the severe slugging cycle time (period) reduces as the superficial velocities at the inlet increase. When the liquid and gas flow rates are high severe slugging will not occur and steady state is reached. It is then of the utmost importance for flow assurance to correctly predict when severe slugging flow becomes stable flow.

Taitel plotted his experimental points in a stability map in order to try to find the stability curve experimentally, show in Fig. 6.7.

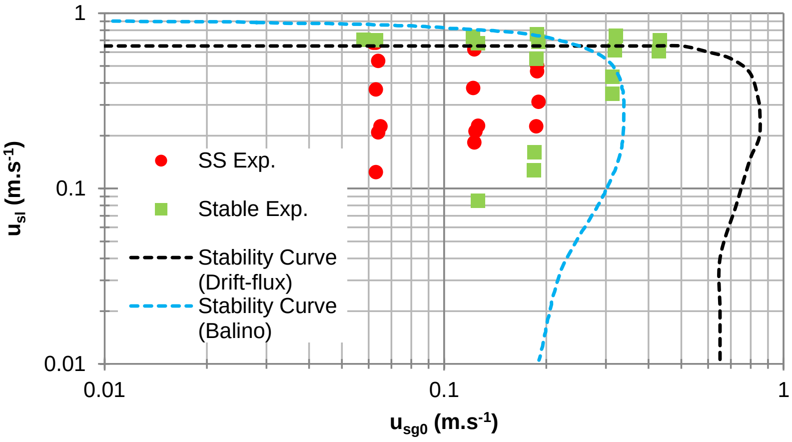


Figure 6.7: Taitel's stability map.

The stability curve in Fig. 6.7 was generated using the present model by running simulations at different inlet velocities and see when the system became stable. It is possible to conclude that the model enclosed all the experimental severe slugging cases.

However, some of the experimental points that were considered stable fell under the unstable region. A possible explanation for this is that the points that Taitel considered stables were in fact in unstable flow regime. In this regime both gas and liquid penetrate the riser at all points, but there is still small fluctuations on the pressure profile over time, never reaching steady state. Some authors like Malekzadeh [9] considered the system stable if the pressure fluctuations are smaller than 0.4 bar but in the present work the system was only called stable if it reached steady state, that is, if the variables converged to a single values independent of time, thus the difference on the right side of the stability curve.

6.2 Hilly-Terrain

A typical offshore pipeline system follows the terrain topography before reaching the riser. With negative and positive inclination depending on the section, it is also possible that severe slugging occurs when certain conditions are met.

De Henau [23] studied both experimentally and numerically severe slugging in hilly-terrain by using a pipeline that consists in two downhill sections and two uphill sections, interchanged and connected with flexible hoses. Each section had a length of 3.84 m and each flexible hose had a length of 0.314 m. A more detailed description of the experimental setup can be found in his work. The experimental setup, defined in Fig. 6.8 and Table 6.4, also contained a tank for the gas to pass before entering the pipeline system in order to gain an additional pipeline length of 53 m. The inclinations of each pipeline section varies with the angles being defined at Table A.1.

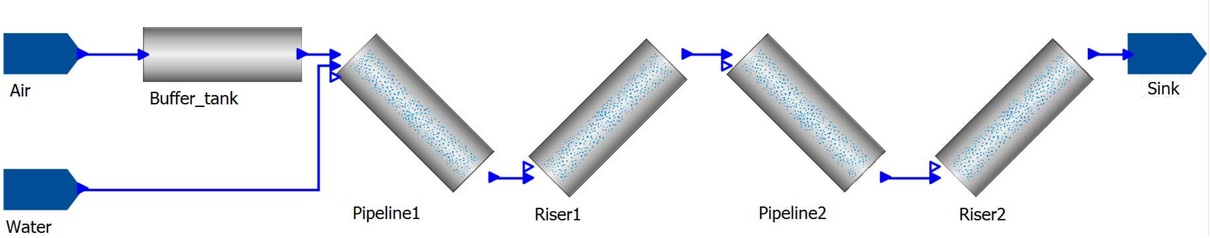


Figure 6.8: DeHenau experimental setup.

Table 6.4: gPROMS configuration of DeHenau experimental setup.

	Buffer_tank	Pipeline1	Riser1	Pipeline2	Riser2
Length (m)	53	4	4.15	4.15	4
N^o of cells	10	80	83	83	80
Inclination	0^o	-15^o - -26^o	15^o - 26^o	-15^o - -26^o	14^o - 24^o
Diameter (cm)	5.18	5.18	5.18	5.18	5.18

The experiments were conducted with the separator at atmospheric pressure with the air entering at the buffer tank and the water after it. As mentioned earlier, DeHenau also developed is own model along his experimental studies. Table 6.5 shows the experimental results of the different cases of severe slugging as well the simulation results predicted by the developed model and the simulations carried out by DeHenau.

Table 6.5: Experimental and numerical results.

Case	Experiment	Drift-flux		DeHenau [23]	
	t_{exp}(s)	t_{sim}(s)	Error (%)	t_{sim}(s)	Error (%)
3	92.4	69.5	-24.8	87.4	-5.4
4	57.9	44.5	-23.1	51.7	-10.7
5	45.2	35.0	-22.6	45.0	-0.4
7	12	21.0	75.0	11.6	-3.3
8	182.5	219.0	20.0	180.9	-0.9
9	95.2	103.5	8.7	93.0	-2.3
10	89.2	93.0	4.3	89.8	0.7
11	75.2	90.5	20.3	50.3	-33.1

The drift-flux model predicted correctly the existence of severe slugging for all the cases above, even if with an higher error than DeHenau model on the cycle time. The relation between the measured and simulated period of severe slugging for the drift-flux model is presented on Figure 6.9 and Table 6.6 shows the average error of both models against the experimental data.

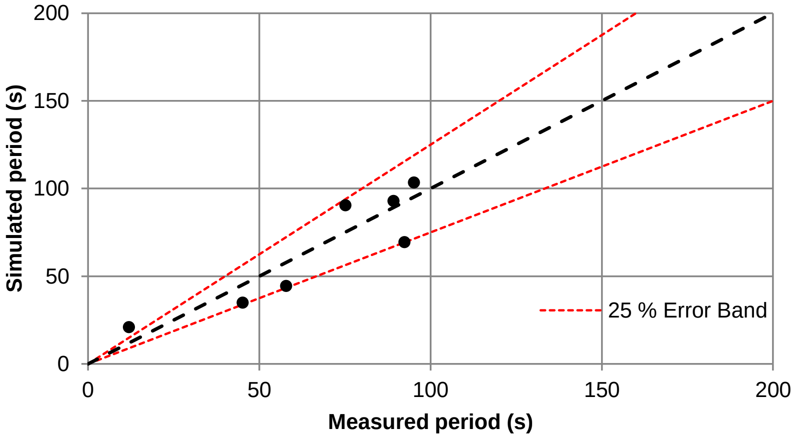


Figure 6.9: Comparison between experimental a simulated period.

Table 6.6: Cycle time average Error.

	Period Average Error (%)
Drift-flux	24.9
DeHenau [40]	7.1

The drift-flux model didn't manage to predict the cycle time of severe slugging with the same accuracy as DeHenau model, getting an average error more than three times higher.

This is due to different factor such as the correlations used and even how the experimental setup was written in gPROMS.

The drift-flux model develop in this work uses Shi's correlation as the general slip law that dictates the relationship between the velocities of the gas and liquid phases. This correlation, however, was created with the focus on it's application on wellbores, which are usually vertical or nearly vertical but never negative. The correlation was extended to account negative inclinations in this work. The parameters were not changed and optimized for these inclinations.

All the experimental setup have negative inclination, however DeHenau experimental setup have negative inclination up to -26^o, which makes the approximation that the author made on using the same values as Shi for the parameters less valid, and explains why the average error of the model for this experimental case is higher than the others.

Another reason is that in the gPROMS configuration of the experimental setup the flexible hoses volume is divided and added to the adjacent pipe models. Therefore, instead of having a junction model that varies with inclination the hoses are considered to have fixed inclination for each half. Although this is a good approximation for the majority of the systems, in this case the junction is around 10 % of the pipeline section size, therefore the error introduce by this approximation might be relevant. Nevertheless, the model develop in the present work shows capability of reproduce this phenomena in a complex pipeline system.

In a hilly-terrain there are different low points in which a single or multiple slugs might form. By measuring the pressure at the bottom of each riser it's possible to see where and when the slugs form. Figure 6.10 compares the pressure profile simulated by the developed model with the experimental for case 9.

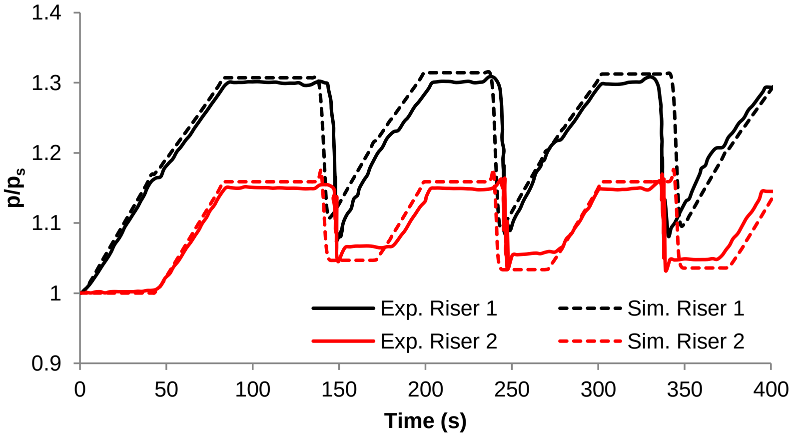


Figure 6.10: Pressure profile.(DeHenau case 9)

There is only gas in the pipe initially. First, the liquid blocks the first riser and forms a slug that starts to grow. At around 40 seconds the slug reaches the top of the first riser and the compressed gas will push it out like in a normal severe slugging cycle for the downward pipeline + riser system. While when there is only a pipeline followed by a riser the slug is pushed to a separator, in this case the slug goes to the other pipeline-riser section, that was initially filled only with gas. When the gas in the second pipeline penetrates the second riser there is the blowout followed by liquid fall back.

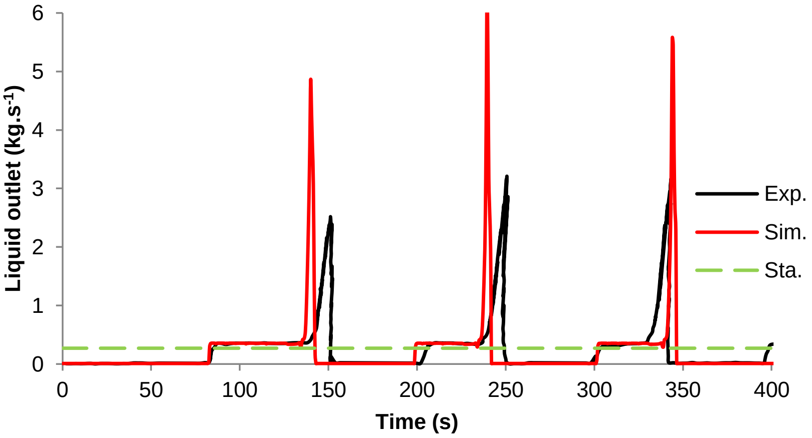


Figure 6.11: Liquid production profile.

Figure 6.11 show the liquid output to the separator. As expected, the liquid only starts to get produced when the slug reaches the top of the second riser.

The peak intensity in the liquid output production is higher than the measured experimentally because the device used to determine the flow rate probably didn't take points with really small time intervals.

By looking at both Figure 6.10 and 6.11 it is possible to conclude that the model correctly represents the severe slugging phenomena.

6.3 Drift-flux model *vs.* OLGA - Performance

In this section the drift-flux model developed in the present work is compared against OLGA. OLGA was started it's development on the 80s by Bendiksen [22]. Now with more than thirty years of development it is considered the petroleum industry standard for dynamic multiphase simulation.

This performance test was done using Malekzadeh [9] experimental and numerical investigation of severe slugging in a long pipeline-riser system. Malekzadeh used OLGA v5.1.2 to simulate his experimental cases and test the performance of the simulator. In this way, it is possible to see how close is the drift-flux model performs for a simple mixture of air and water against the top performer simulator.

Most of the experimental studies done on severe slugging were done on system with a small pipelines. In the industry, the pipeline that connects to the riser is much longer, so Malekzadeh created an experimental setup to study severe slugging with a 65 m horizontal pipeline, connected to a 35 m pipeline which had an inclination of -2.54° from horizontal and followed by a 15.5 m vertical riser. Similar to previous experimental works, it was also included a 250 L gas buffer tank in order to provide additional gas compressibility, which corresponds to an additional pipeline length of 123 m. The gPROMS configuration of the experimental setup is defined on Figure 6.12 and Table 6.7.

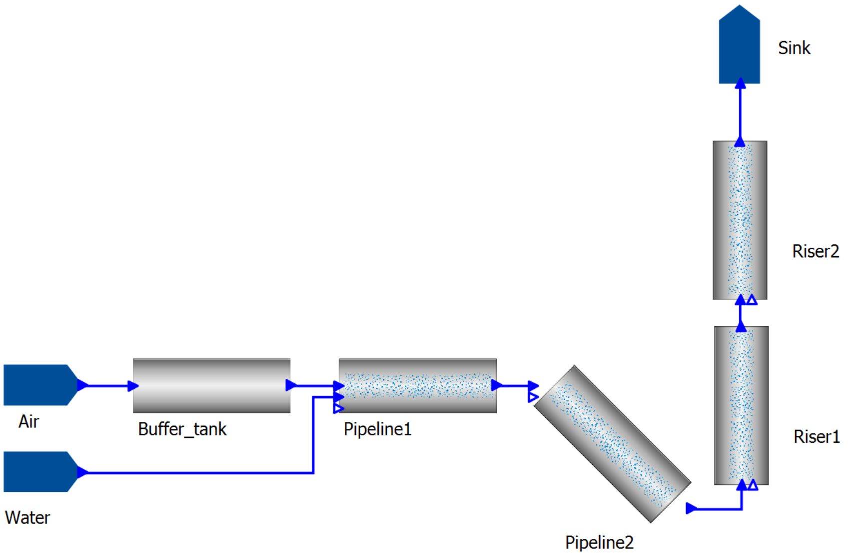


Figure 6.12: Malekzadeh experimental configuration.

The experiments were carried out at atmospheric pressure with the separator at atmospheric pressure with the air entering at the buffer tank and the water after it. It should also be noted that before the separator there is a valve with a flow factor, K_v, of 10 m^{3}h^{-1}bar$^{-1/2}$. A choke valve model used was also developed in the present work an is explained on the next chapter.

Table 6.7: gPROMS configuration of DeHenau experimental setup.

	Buffer_tank	Pipeline1	Pipeline1	Riser1	Riser2
Length (m)	123	65	35	10	13
N^o of cells	20	260	140	2.5	52
Inclination	0^o	0^o	-2.54^o	90^o	90^o
Diameter (cm)	5.08	5.08	5.08	5.08	4.50

Table 6.8 shows the experimental results of the different cases of severe slugging as well the simulation results predicted by the developed model and OLGA.

Malekzadeh reported in is experimental investigation the type of severe slugging for each case. This allows to evaluate the models ability to predict the correct severe slugging regime. The definition of each type was described back in chapter 2 and they were identified with the help of the pressure profile plots, as the one on Fig.6.13.

Another interesting detail is that Malekzadeh reported dual slugging for case 11 and 15, which consists in severe slugging that has two slugs that have different cycle times and length. It is important to refer that the developed model did not predict dual slugging and OLGA predicted correctly both dual slugging cases, but mispredicted six normal cases as dual slugging, with cycle times much different from the experimental.

Table 6.9 shows a summary of the results produced by both models for Malekzadeh dataset.

The unstable flow regime is the one that is less well described by the developed model, but is more important to refer that OLGA simulator mispredicted every unstable flow regime as severe slugging II sometimes with dual slugging occurrence.

Both models had a similar performance in guessing the correct regime, with the drift-flux model coming slightly ahead.

To have a fair comparison between the average error in both models all the unstable flow regime cases were neglected and a new average error was determined. With this change, the drift-flux model average error went down almost 10% while OLGA model was lowered by around 4%. This supports the supposition that the unstable flow regime is the one that gives rise to bigger errors and that the developed model can have similar performance as the industry standard, OLGA.

Figure 6.13 shows the pressure profile for case 1 of Malek (Table 6.8. The first thing to conclude it that both models follow reasonably well the experimental pressure profile, with the developed model getting a very good match with the experimental data. OLGA predicts a slower cycle time, delivering minus one slug for the timespan showed.

Another interesting fact is that both Malekzadeh and OLGA categorized case 1 as being severe slugging type I. When looking to the pressure profile, it is hard to see the typical horizontal pressure line that means that the riser is completely filled with liquid. Similarly, in Fig. 6.14, the liquid output should first reach a small and steady value and only peaking when the slug was pushed out of the pipeline. Both of this profiles resemble more that a severe slugging II (SS2), where the slug is not bigger than the riser.

Table 6.8: Experimental and numerical results.

Case	Experiment		Drift-flux			OLGA [20]		
	t_{exp}(s)	Type	Type	t_{sim}(s)	Error (%)	Type	t_{sim}(s)	Error (%)
1	179	SS1	SS2	171	-4	SS1	211	18
2	135	SS1	SS1	150	11	SS1	154	14
3	109	SS1	SS1	126	16	SS3	105	-4
4	78	SS1	SS1	83	7	SS3	77	-1
5	95	SS1	SS1	118	24	SS3	95	0
6	185	SS1	SS1	182	-1	SS1	211	14
7	101	SS1	SS2	98	-3	SS1	108	7
8	148	SS1	SS1	154	4	SS1	138	-7
9	173	SS1	SS1	224	30	SS3	143	-17
10	90	SS1	SS1	97	8	SS3	85	-6
11	92 and 138	SS2	USO	72	NA	SS2	76 and 191	NA
12	72	SS2	USO	100	38	SS2	108 and 160	NA
13	96	SS2	SS2	87	-10	SS2	98	2
14	68	SS2	SS2	55	-19	SS2	63	-7
15	227 and 159	SS2	USO	837	NA	SS2	138 and 286	NA
16	119	SS2	USO	330	177	SS2	211	77
17	93	SS2	USO	179	92	SS2	143	54
18	82	SS2	SS2	62	-24	SS2	72	-12
19	72	SS3	SS1	76	5	SS3	72	0
20	69	SS3	SS1	73	6	SS3	70	1
21	63	SS3	SS1	70	11	SS3	67	6
22	60	SS3	SS3	68	13	SS3	66	10
23	72	SS3	SS3	80	11	SS3	73	1
24	64	SS3	SS3	70	9	SS3	66	3
25	57	SS3	SS3	65	14	SS3	63	11
26	113	SS3	SS3	158	40	SS3	98	-13
27	86	SS3	SS3	110	28	SS3	89	3
28	64	SS3	SS3	77	20	SS3	68	6
29	57	SS3	SS3	63	11	SS3	59	4
30	64	USO	USO	89	38	SS2	98	53
31	64	USO	USO	88	38	SS2	98	53
32	65	USO	USO	83	28	SS2	95 and 191	NA
33	53	USO	USO	43	-18	SS2	86 and 155	NA
34	60	USO	SS2	47	-22	SS2	60 and 236	NA
35	43	USO	USO	34	-20	SS2	60	40
36	72	USO	USO	99	38	SS2	56 and 111	NA
37	73	USO	USO	63	-13	SS2	95	30
38	67	USO	USO	49	-26	SS2	87 and 174	NA

Table 6.9: Performance of the drift-flux model and OLGA model.

	Drift-flux	OLGA [9]
Period Average Error (%)	23.6	15.9
Period Average Error (without USO) (%)	15.6	12.3
Cases with correct type	27	24

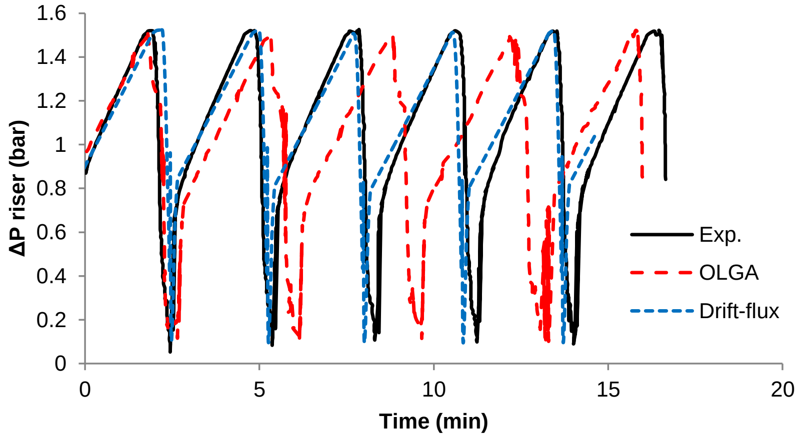


Figure 6.13: Pressure profile Taitel.

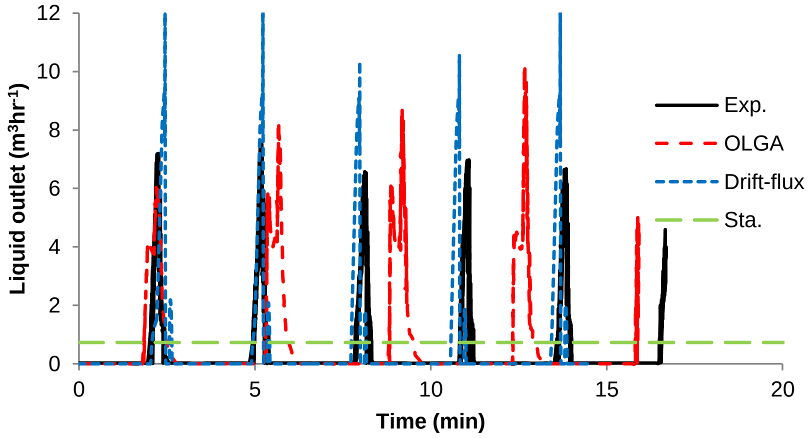
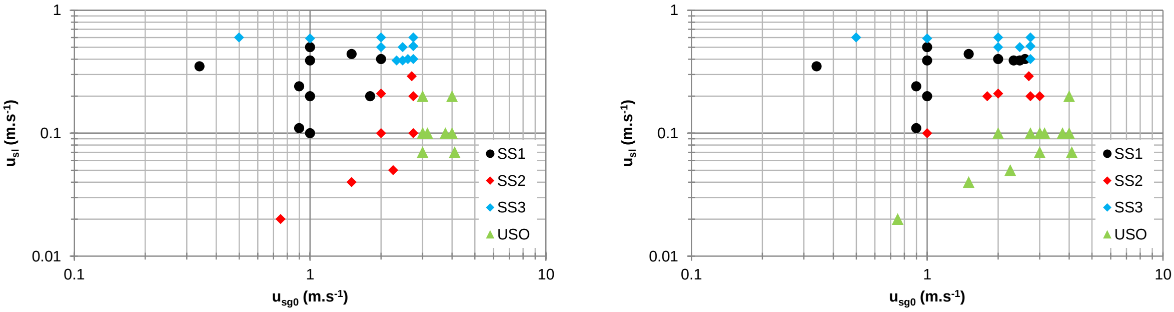


Figure 6.14: Liquid outlet Malek.

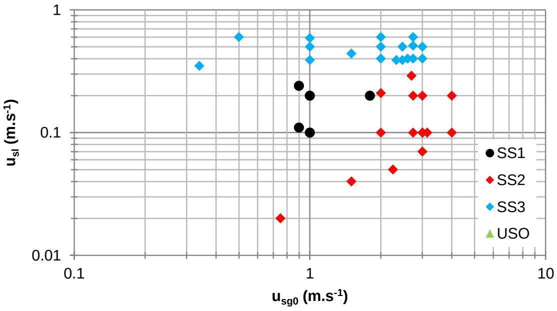
The author of the present work decided to categorized this case as severe slugging II, but it is really near the transition to severe slugging I (SS1).

Stability Map

Figure 6.15 shows three stability maps, one experimental and the other two numerically generated. These stability maps show the regions for each type of severe slugging and it can be seen that both numerical maps predict reasonably well the upper regions but both fail to capture correctly the transition region between severe slugging II and unstable oscillations flow.



(a) Experimental (b) Drift-flux



(c) OLGA

Figure 6.15: Comparisson between stability maps.

Chapter 7

Mitigation Strategies

Severe slugging causes production losses and unstable and intermittent production of liquid to the separator. For getting the highest profit and for safety reasons this phenomena must be eliminated.

Several strategies have been capable of mitigating severe slugging, such as: pressure increase, gas injection or even the improvement of pipeline.

This chapter focus on the discussion of some common mitigation strategies like the use of choke valves, gas-lift and separator pressure changes. Simulations were performed for the last two in order to evaluate the capability of these methods in reducing severe slugging.

7.1 Choking

The use of a choke valve at the end of the riser as a mitigation strategy was suggested by some researchers after they observed that it reduces the cycle time of severe slugging or even reaches steady state. The theory behind it is that the back pressure will increase proportionally to the velocity at the choke, and this increase in pressure will eliminate severe slugging.

Malekzadeh [9] used a choke valve in the experimental setup. In order to test the model performance in the last chapter, there was a need for develop a simple valve model.

7.1.1 Choke Valve Model

The choke valve creates a pressure drop that is usually proportional the velocity square. The pressure drop across the valve can be estimated by the definition of the flow factor.

$$K_v = Q_v \sqrt{\frac{\rho_m}{\rho_{ref} \Delta Pchoke}} \tag{7.1}$$

By rearranging Eq. 7.1 in order to the pressure drop,

$$\Delta Pchoke = \frac{1}{K_v^2} \frac{\rho_m}{\rho_{ref}} Q_v^2 \tag{7.2}$$

Where ρ_{ref} is the water reference density and equal to $1000kgm^{-3}$. Since there is two phases flowing thorough the choke, an expression for the volumetric flow rate must be defined.

$$Q_v = \alpha_l u_l A \tag{7.3}$$

Eq. 7.3 assumes that the only the liquid phase is important for the estimation of the pressure drop across the valve. This is generally a good approximation as the gas density is low.

It is also important to note that the flow factor provided was obtained using only water as single phase. The usage of this flow factor to calculate the pressure the pressure drop of air passing the choke would be wrong.

By decreasing the flow factor, which will be related to the valve stem position by some expression, the back pressure will increase and reduce severe slugging.

7.2 Separator Pressure

Yocum [5] was the first to identify the severe slugging in a offshore facility and he concluded that increasing the separator pressure would mitigate or even eliminate the phenomena.

To study the effect of the separator pressure in severe slugging the pressure at the separator was increased for one scenario. Table 7.1 shows the cycle time for the case 1 of Taitel experiments for the different scenarios.

Table 7.1: Pressure scenario definition

	p_s(bara)	Period (s)
Base case	1.013	20.4
Scenario 1	1.5	10.9

The increase in pressure indeed lessens the effects of severe slugging as suggested. By plotting the pressure in the pipeline for both scenarios it is possible to see that the pressure fluctuations are shorter in amplitude. In Scenario 1 (Fig. 7.1) the initial separator pressure is atmospheric, then, at 50 s, it is increased to 1.5 bara.

The importance of the stability maps comes again to play here on mitigation strategies. By determining the stability curve for the different scenarios, as shown bellow in Fig. 7.2, it is possible to determine area were severe slugging was completely eliminated.

In conclusion, increasing the separator pressure will eliminate severe slugging for a constant inlet flow rate. In real situations, however, the increase in pressure will lower the flow rate coming from the reservoir, so even though the severe slug region in the map is small, it might be possible to still be inside it as the flow rates of liquid and gas will also be smaller.

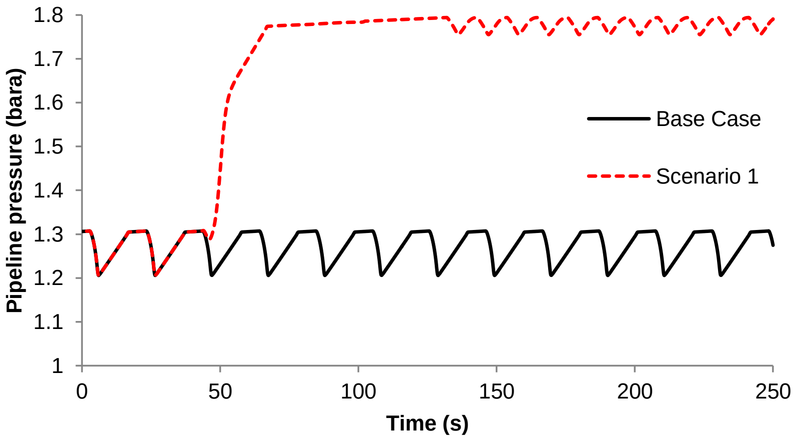


Figure 7.1: Pressure profile pressure.

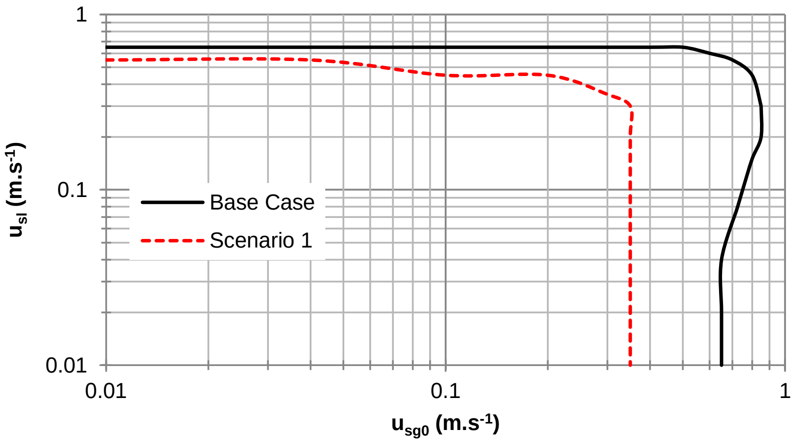


Figure 7.2: Stability map pressure.

7.3 Gas-lift

Gas-lift is a method well known in the petroleum industry. This method is used to decrease the pressure drop in order to increase production. In the severe slugging case, inserting gas at the bottom of the riser will also decrease it's hydrostatic head and mitigate severe slugging.

Some researchers [14] concluded that for the gas-lift method completely eliminate severe slugging there is need for a large volume of gas injected. More recently it was demonstrated the ability of using self-lift for mitigating severe slugging [12, 13]. In this strategy the gas inserted at the bottom of the riser comes from the pipeline section where stratified flow is dominant.

In this work the focus will be only on gas-lift. Table 7.2 shows the cycle time for the case 1 of Taitel experiments for the different scenarios.

Figure 7.3 shows the pressure profile for the different scenarios when air is injected at the bottom of the riser.

As the gas is injected, the hydrostatic head of the riser is smaller and the pressure drops. Even if the amount of gas injected is not enough to make the system stationary, the cycle time and amplitude of the

Table 7.2: Gas-lift scenarios definition

	Gas injected (kg.hr^{-1})	Period (s)
Base case	0	20.4
Scenario 1	0.9	7.1
Scenario 2	1.3	5.6

fluctuations are much smaller.

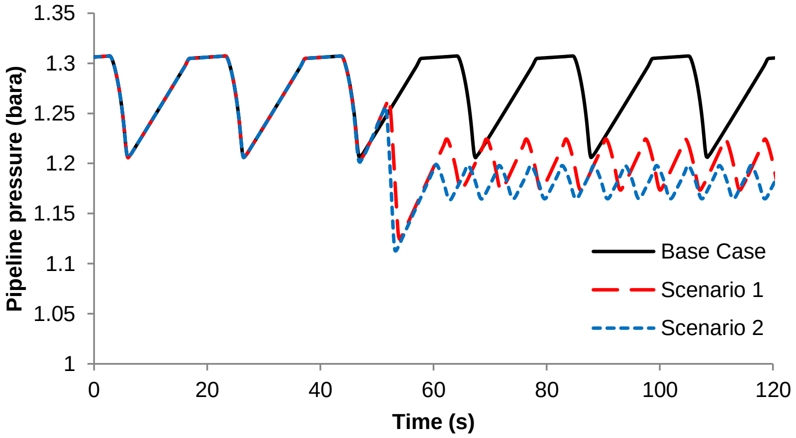


Figure 7.3: Pressure profile gas-lift.

Like in the separator pressure analysis, the elaboration of a stability map and stability curves for each scenario is presented on Fig. 7.4.

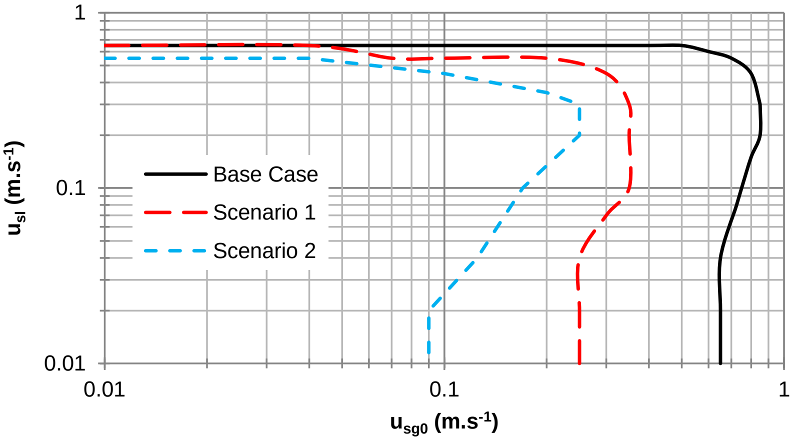


Figure 7.4: Stability map gas-lift.

As expected, the injection of gas at the bottom of the riser moves the stability curve to the left, so even with smaller flow rates of gas at the inlet the system might be steady state.

The gas-lift method is good strategy to mitigate severe slugging and other benefit is that it will increase the production as the pressure drop is related with the inlet flow rates to the pipeline-riser system in the industry. This means that if gas is injected the pressure drop will be reduce, increasing the inlet flow rates and moving the experimental point to the right and up and the stability curve will move to the opposite direction.

7.4 Summary

Choke is a viable option for eliminating severe slugging by increasing the back pressure. Similarly, increasing the separator pressure will eliminate severe slugging for a constant inlet flow rate. In real situations, however, the increase in pressure will lower the flow rate coming from the reservoir, so even though the severe slug region in the map is small, it might be possible to still be inside it as the flow rates of liquid and gas will also be smaller.

The gas-lift method is good strategy to mitigate severe slugging and other benefit is that it will increase the production as the pressure drop is related with the inlet flow rates to the pipeline-riser system in the industry. This means that if gas the experimental point will move to the right and up and the stability curve will move to the opposite direction. Self-lifting might be a better alternative for the same features if the gas-lift cost are too high.

The design of the system also plays an important effect on severe slugging as it will be shown in the next chapter, so this might also be a viable way of preventing severe slugging if the project is still on the early stages.

In the end, the decision of which strategy to apply will depend on a economical analysis that should weight the loss in gains due to production losses (choke valve and separator pressure) with the increase of operating costs (gas-lift) in order to maximize profit. Future studies on the optimization of eliminating severe slugging using one or more strategies are encouraged to be done.

Chapter 8

Sensitivity Analysis

The results of a model might significantly change by changing a parameter like the grid resolution or the correlation parameters. How sensible is the model response to the change of determined parameter is an important aspect to study as it can explain some of the model deviations from experimental data.

The influence of the pipeline design in severe slugging is also of great importance as it can prevent its occurrence and greatly reduce the severe slugging region.

This chapter starts by showing the model sensitivity to the grid resolution and some empirical parameters. After, the influence of some the pipeline diameter, length and inclination is numerically studied and discussed.

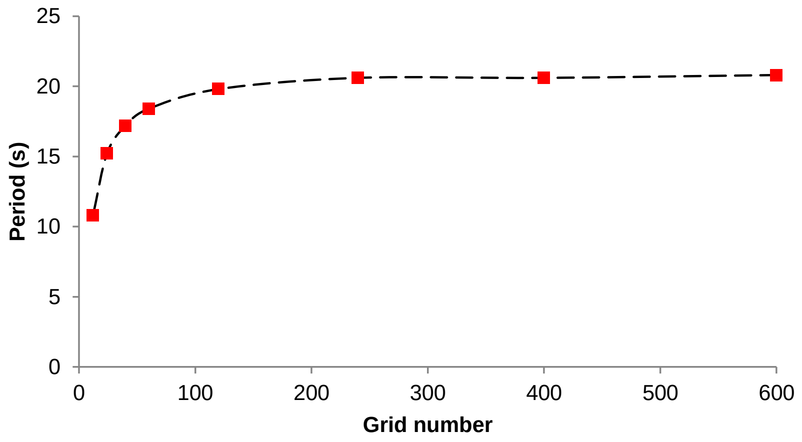
8.1 Grid Resolution

In chapter 5 it was shown that the model is discritized using the finite volume method with cells.

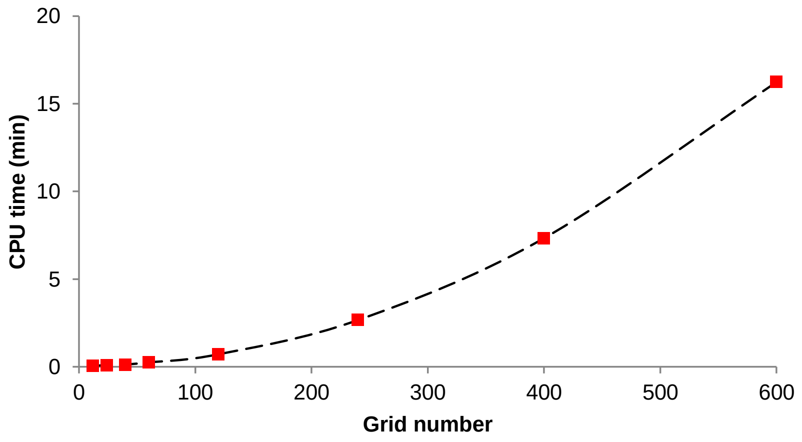
Depending on the grid size, the simulation results might not be the same. That is why in Chapter 6 is was told that for high transient sections a more refined grid needed to be used. However, the use of a grid with a smaller size will increase the number of equations to be solved and thus the CPU time it takes for the simulation to run. Table 8.1 and Fig. 8.1 shows these two issues.

Table 8.1: Grid resolution effects on severe slugging and CPU time.

Grid length (cm)	Grid Number	Period (s)	CPU Time (min)
2	600	20.8	16.2
3	400	20.6	7.3
5	240	20.6	2.7
10	120	19.8	0.7
20	60	18.4	0.3
30	40	17.2	0.1
50	24	15.2	0.1
100	12	10.8	0.1



(a) Influence of grid resolution on the simulated period.



(b) Influence of grid resolution on simulation time.

Figure 8.1: Influence of grid resolution on severe slugging. (Taitel case 1, Table 4.1)

From the sesults shown in Fiq. 8.1 for a lower number of cells the cycle time is completely wrong, but as the number of cells used increase, the results start to converge.

As for the computational time, on the other hand, increases exponentially with the increase of the grid number. As such, an optimal grid resolution must be chosen with the aim of having the converged result with the minimal computational time. For example, in this case, that would be to discritize the domain with cells with 5 cm length.

8.2 Drift-flux Correlation Parameters

One of the reasons for using the Shi correlation in the development of this model was the existence of empirical parameters that latter could be improved and better tuned. It is then of extreme importance to perform a sensitivity analysis to these parameter in order to determine the ones that make the model more sensible to their changes.

This analysis was performed by changing each parameter separately, with some exceptions that are discussed further ahead, within a range of 20% of their original value. Table 8.2 show the sensitivity analysis for a case were the period with the default values is 20.4.

Table 8.2: Sensitivy analysis of Shi parameters.

		Pertubation (%)				Deviation (%)			
Parameters	Default Values	20	10	-10	-20				
A	1	16.4	18.3	23.5	27.5	20	10	15	35
a_1	0.06	20.4	20.5	20.5	20.4	0	0	0	0
a_2	0.21	20.9	20.6	20.3	20.2	2	1	0	1
n_1	1.85	18.3	19.3	21.7	23.2	10	5	6	14
n_2	0.21	20.4	20.4	20.6	20.5	0	0	1	0
n_3	0.95	20.6	20.5	20.4	20.3	1	0	0	0

It is important to note that the sensitivity analysis of the parameters B and F_v was not performed because, as it is possible to see on Eq. 4.25, if the A it the default value the values of B and F_v won't matter as C_0 will always constant and equal to 1.

From this study it is possible to state that the model is more sensible to the changes in the parameters A and n_1.

A is the value that the distribution parameter, C_0, take the bubble and slug flow regimes. Although the physical meaning of change this parameter to values small than unity is not correct, as the distribution parameter should be between 1 and 1.2, changing the parameter shows a change in behavior. Usually A should be 1.2 because that's is the typical distribution parameter value on bubble flow and slug flow. However, due to parameter tuning done by Shy, this parameter reached the lower bound of it's possible values.

The second parameter, n_1, it's related to the correction term of the drift velocity due to the pipe inclination. As already stated before, the term was extended to allow pipe with negative orientation, but values for the parameters in the correction factors remained the same.

Although in the present study the parameters were set to default, the results of this author sensitivity analysis strongly suggest that a new parameter estimation should be done, taking better into account the pipe inclination.

8.3 Pipeline Design

For this numerical study, a typical setup of a downward pipeline + vertical riser system. Figure 8.2 shows the gPROMS representation of the base case presented in Table. 8.3 :

Table 8.3: Base case configuration.

	Pipeline	Riser
Length (m)	30	15
Inclination	-2^{o}	90^{o}
Diameter (cm)	5.08	5.08

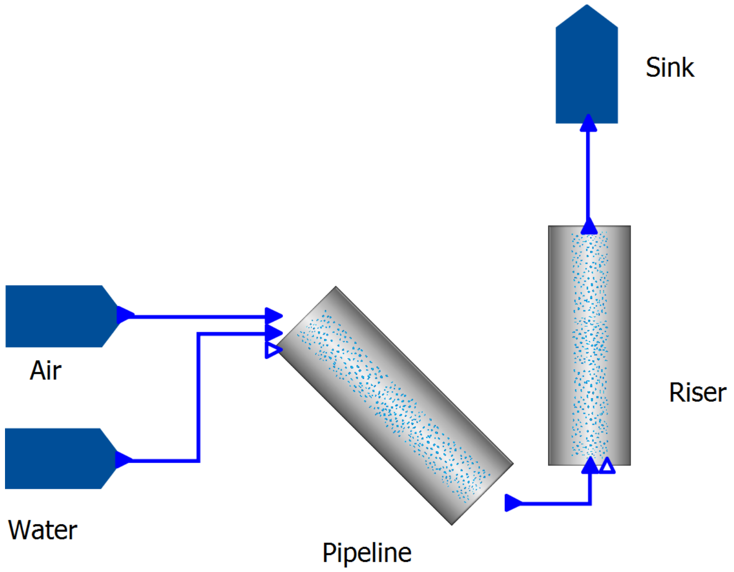


Figure 8.2: Base case experimental configuration.

8.3.1 Pipeline Diameter

During the design phase, the choice between the pipe diameters for the configuration has several alternatives. It is expected that this parameter will have an affect on severe slugging, as it also affects the velocities.

For a constant mass flow rate of both phases, four scenarios were defined with different only by changing the pipeline diameter in order to study the influence of the diameter on severe slugging.

Table 8.4: Diameter sensitivity

	d(cm)	Period (s)
Base case	5.08	133
Scenario 1	6.10	81
Scenario 2	4.06	175
Scenario 3	2.54	Stable

Table 8.4 shows the decrease in pipe diameter also decreases the intensity of severe slugging and scenario 3 shows that the phenomena can even be eliminated. Figures 8.3 and 8.4 show the pressure profile and the liquid production profile, respectively.

As the diameter of the pipeline lowers, for the same mass flow rates, the velocities inside it increase. This increase in velocities make the slugs smaller, decreasing the cycle time of severe slugging. If the velocities continue to increase the liquid will no longer accumulate at the bottom of the riser and steady state will be reached.

The liquid profile shows us that the time without liquid production is also shortened with the decrease in the diameter. This is because the slug reaches the top of the riser faster.

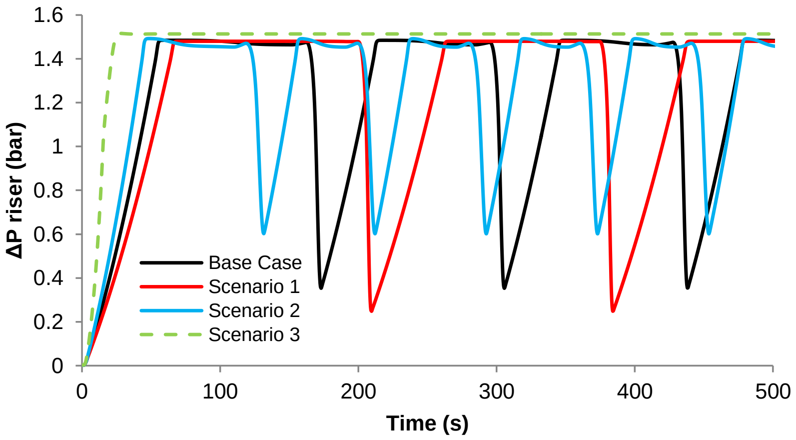


Figure 8.3: Pressure profile diameter.

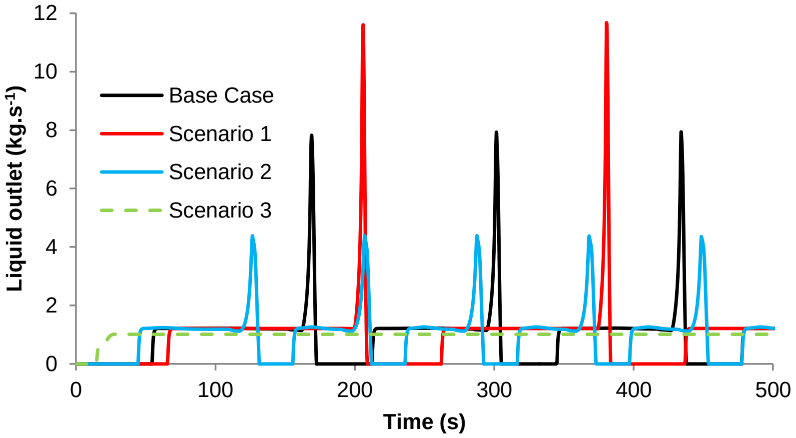


Figure 8.4: Liquid outlet diameter.

8.3.2 Pipeline Length

Other interesting design parameter is the pipeline length. Researchers for long have been using gas buffer tanks in order to add virtual pipeline length to the experimental configuration. The reason is that the amount of gas present in the system upstream of the slug influences it's duration. The scenarios defined used in this numerical study are presented on Table 8.5.

Table 8.5: length sensitivity

	Pipeline length (m)	Period (s)
Base case	30	133
Scenario 1	50	226
Scenario 2	15	37
Scenario 3	10	Stable

The decrease in pipeline length reduces the effect of severe slugging and can even eliminate it.

If the pipeline is longer, the volume of gas inside it will be bigger, so it will take more time to be compressed an win the slug hydrostatic head pressure. If the volume of gas inside the pipeline is small enough it will not let the liquid slug forms and the system will becomes stable, as shown in Scenario 3 in Fig. 8.5.

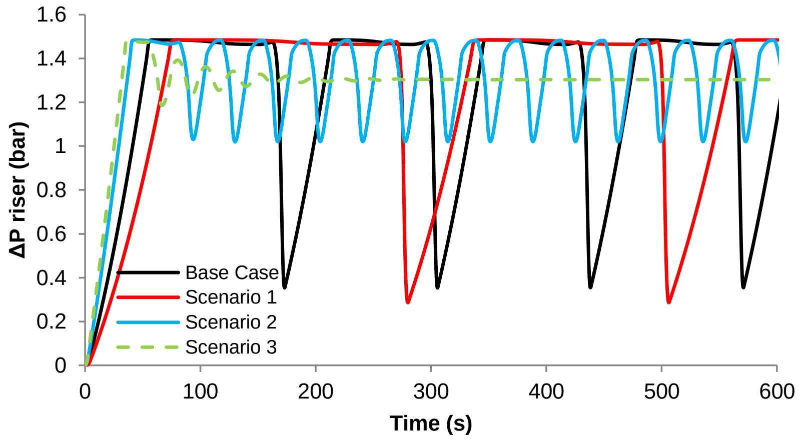


Figure 8.5: Pressure profile Length.

Therefore, it is possible to conclude that if possible the riser should be close to the well, minimizing the length of the pipeline and the amount of gas to be compressed, reducing in this way severe slugging.

8.3.3 Pipeline Inclination

Severe slugging is a phenomena where liquid accumulates at the lowest points of the system and a slug is created. It is of interest to see the influence of pipeline inclination on severe slugging, in particular if the pipeline doesn't have a negative inclination. Table 8.6 show the results of the numerical simulations.

Table 8.6: inclination sensitivity

	θ	Period (s)
Base case	-2°	133
Scenario 1	-5°	141
Scenario 2	-20°	148
Scenario 3	0°	Stable

Two very important conclusions can be taken when analyzing Table 8.6 and Fig. 8.6.

By changing the negative inclination of the pipeline will have almost no effect on severe slugging. The small increase in the period when the pipeline inclination is steeper is due to the slug grow slightly more in the pipeline, because the pressure there is smaller the more downward inclined it is.

The other conclusion is that if the pipeline is no longer negative inclined, stable flow will be achieved. However, this does not invalidate the existence of severe slugging in a horizontal pipeline + riser system, as some researchers already observed it both experimentally [8] and numerically [19]. Rather, correct conclusion to take is that if the pipeline isn't downward inclined it greatly reduces the area of severe

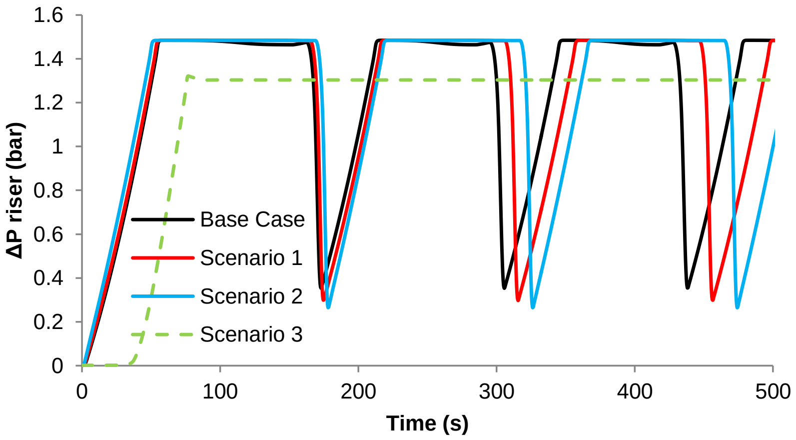


Figure 8.6: Pressure profile Inclination.

slugging, in most of the cases eliminating it for meaningful flow rates.

Chapter 9

Conclusions

The petroleum industry heavily relies on the simultaneous transport of gas and liquid phases in a single pipeline. Due to the pipeline-riser system configuration, severe slugging might occur. This phenomena is unwanted and it is important to have a multiphase dynamic model capable of accurately represent it.

A drift-flux model was developed with the purpose of predicting severe slugging. This dynamic and isothermal model based on the one dimensional conservation equations of mass and momentum used Shi correlation as the general slip law. This correlation was chosen because its parameters were optimized for petroleum industry and its potential for improvement, as its parameters can be fitted to experimental data.

The current model was developed in gPROMS, using the software internal implicit temporal discretization. A finite volume scheme with a staggered grid was developed in gPROMS language in order to be able solve numerical problems such as the ones caused by local flow reversal and phase disappearance.

The model was then validated against experimental data for downward pipeline + vertical riser systems and for hilly terrain systems. It was also compared with models developed by other authors, and it's performance was tested against the industry standard multiphase simulator, OLGA.

Different mitigation strategies that are usually used to eliminate severe slugging were studied. A simple choke valve model was developed in order to estimate the pressure drop caused by the valve. Scenarios with varying gas-lift at the bottom of the riser and different separator pressure were done in order to study the capability of these strategies in reducing severe slugging.

Lastly, a sensitivity analysis was done to the model to show the importance of the grid resolution chosen and of the values of some parameters. The influence of some design parameters in severe slugging was also studied.

9.1 Conclusions

The developed model was validated and can predict accurately the occurrence of severe slugging and characteristic properties like the cycle time and slug length. This showed capability of achieving the similar performance other model used in the present. This model can also be used to simulate other multiphase flow regime for different pipeline configurations, either in steady state or dynamic.

Shi correlation was successfully extended to allow negative inclination. However, the parameter were not change for steeper negative inclination the model results get worse.

Different mitigation strategies typically used to eliminate severe slugging were discussed and the model showed is capability of simulating these methods. Injecting gas at the bottom of the riser or increasing the separator pressure successfully reduce severe slugging. The latest method, however, will cause production losses as an side effect.

The choice of the design parameters will influence severe slugging. Smaller pipes both in diameter and length will help mitigating severe slugging. More importantly, the pipeline section should not have a downward inclination, as that will greatly help the liquid accumulation for the generation to severe slugging.

9.2 Future Work

This work was created to serve as base for a new and better multiphase model capable of predicting severe slugging. Due to the large applications of this model there are many ways the model can be extended/improved. Some suggestions for future work are listed bellow:

- Extend the model to include the energy balance.
- Improve and extend Shi correlation by making some of the parameters inclination dependent.
- Estimate the Shi new parameters with industrial data.
- Study the model performance in typical multiphase flows.
- Compare the different friction models available for multiphase flow.
- Use the model for the design and operation of pipeline systems.
- Study self-lift as an method to eliminate severe slugging.
- Perform economical analysis and optimization in order to find which are the optimal variables for the mitigation methods in order to reduce severe slugging.

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