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An Approach for Slug Catcher Design based on Multiphase Flow Simulation

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Abstract

The purpose of this thesis is the development of two programs in Microsoft Excel which can be used to perform a multiphase flow analysis and to design a slug catcher. From the development of these tools ILF expects to gain knowledge regarding multiphase flow in pipelines.

After explaining multiphase flow phenomena the main features of the mechanistic model according to Xiao *et al*. (1990) are introduced which serves as a basis for the multiphase flow analysis performed in this master thesis. It is furthermore shown how to determine the temperature profile in a multiphase flow pipeline and methods of slug analysis are discussed.

In order to validate the multiphase flow tool simple flow line profiles of different lengths and at different inclination angles are calculated. Furthermore a 15,5 km flow line with a given elevation profile is considered. The calculated exit pressures and temperatures are compared and validated with PipeSim. Regarding the simple profiles deviations at small inclinations (<5°) in terms of pressure prediction and deviations at large inclinations (>30°) in terms of temperature determination are observed. For the given 15,5 km flow line the deviation is 2,4 % in the predicted exit temperature and 1,3 % in case of the calculated exit pressure.

Various multiphase flow correlations are selected to perform a correlation comparison with PipeSim. It is found that the correlations vary in terms of the predicted flow pattern, the pressure drop and especially in the predicted slug length. A decision tree to select appropriate correlations is developed based on selection criteria provided by PipeSim user guide.

Furthermore, a sensitivity study is carried out in PipeSim where the influence of pipe roughness, pipe diameter, oil flow rate, gas flow rate, API gravity and gas gravity on the pressure drop and slug length are investigated. It is shown that the pipe diameter and the gas flow rate have a significant influence on the predicted slug length.

Theoretical concepts and design criteria for slug catcher are discussed which serve as a basis for the development of the slug catcher tool. Finally, the slug catcher tool is validated on the basis of a slug catcher design study. It is found that the sizing with the vessel type tool can be done with sufficient accuracy. Additionally, it is found that apart of other factors the transition criterion between stratified and intermittent flow pattern plays a crucial role in finger-type slug catcher sizing.

Kurzfassung

Ziel dieser Arbeit ist die Entwicklung zweier Programme in MS Excel um einerseits Multiphasenströmungen analysieren und andererseits Slugcatcher dimensionieren zu können. Die ILF verfolgt damit das Ziel, "Knowhow" im Bereich der Multiphasenströmungen in Pipelines aufzubauen.

Anschließend an eine Einführung in die wichtigsten Phänomene der Multiphasenströmung wird das mechanistische Model nach Xiao *et al*. (1990) kurz erläutert, welches als Basis für die Analyse von Multiphasenströmungen in dieser Arbeit dient. Außerdem wird gezeigt, wie das Temperaturprofil in einer Pipeline mit Multiphasenströmung und wie Slug-Charakteristika bestimmt werden können.

Zur Validierung des Multiphasen-Programmes werden einerseits einfache Profile mit unterschiedlichen Längen und unterschiedlichen Neigungen berechnet. Andererseits wird ein konkreter Fall mit gegebenem Höhenprofil und 15,5 km Länge betrachtet. Die Berechnungen der Drücke und Temperaturen werden zur Validierung mit Ergebnissen aus PipeSim verglichen. Dabei zeigt sich, dass es im Fall von den einfachen Profilen zu Abweichungen bei der Temperatur bei großen Neigungen (>30°) bzw. zu Abweichungen beim Druck bei kleinen Neigungen (<5°) kommt. Die Abweichungen für den konkreten Fall mit gegebenen Höhenprofil betragen für den berechneten Austrittsdruck 1,3 % und für die Austrittstemperatur 2,4 %.

Ein Vergleich verschiedener Korrelationen für Multiphasenströmungen wurde mit PipeSim ausgeführt. Dabei wurde festgestellt, dass die Korrelationen zu unterschiedlichen Ergebnissen für das Strömungsmuster, den Druckverlust und speziell für die Sluglängen kommen. Auswahlkriterien für die anzuwendenden Korrelationen sind im Rahmen dieser Arbeit dargestellt worden.

Darüber hinaus wurde eine Sensitivitätsstudie mit PipeSim durchgeführt, um den Einfluss von Rohrrauhigkeit, Durchmesser, Öl- und Gasfluss, API-Schwere und der Gas-Schwere auf Druckverlust und Sluglänge zu untersuchen. Es konnte gezeigt werden, dass der Durchmesser und die Gasflussrate einen signifikanten Einfluss auf die Sluglänge haben.

Theoretische Konzepte und Designkriterien für die Auslegung von Slugcatcher wurden ebenfalls diskutiert, die als Grundlage für die Entwicklung des Slugcatcher-Programmes dienten. Abschließend wurde das Slugcatcher-Programm basierend auf einer Slugcatcher Design Studie validiert. Dabei stellte sich heraus, dass sich das entwickelte Program besonders für die Auslegung eines „vessel type" Slugcatcher eignet. Für das Design eines „finger-type" Slugcatcher kommt dem Übergangskriterium zwischen der Schichten- und Schwallströmung eine besondere Bedeutung zu.

Table of Contents

1. Introduction

Chapter 1 introduces the reader to the topic of this master thesis. Therefore, the reader finds general background information on multiphase flow investigations in section 1.1. The statement of the problem is discussed in section 1.2 and finally the scope of this thesis is presented in section 1.3.

1.1. Background information[1]

An enormous amount of theory on multiphase flow has been developed in the last six decades by numerous investigators. According to Brill and Arirachakaran (1992) this period can be subdivided into three phases.

The Empirical Period (1950-75): Especially in the beginning of this period data was commonly obtained from laboratory tests with only a few investigators using field data. Liquid hold up and flow pattern were still predicted empirically even though pressure gradient equations involved conservation of momentum and mass principles. Fluids were treated as homogenous mixtures resulting in frictional pressure losses based on single-phase flow equations. The attempt to account for basic physical mechanisms - as can be seen in the work of Dukler and Hubbard (1975) and Taitel and Dukler (1976) - gave rise to the development of mechanistic models for slug flow and flow pattern prediction.

The Awakening Years (1970-85): Even though the introduction of the personal computer (PC) and it’s coupling with the available empirical correlations evolved, the accuracy of the predictions could not be improved. It was recognized that empirical flow pattern maps were inadequate since the transitions are not only dependent on flow rates (as assumed) but also on other parameters, especially inclination. Furthermore, an inadequate liquid hold up correlation and the oversimplified assumption of the homogenous mixture made many investigators aware of the need to introduce more basic physical mechanisms.

The Modelling Period (1980-Present): The demand of a better understanding of multiphase flow in pipes was met by combining the experimental and the theoretical approach in the 1980’s. Improved experimental research guaranteed by sophisticated test facilities resulted in improved understanding of multiphase flow mechanisms, leading to further development of mechanistic models. Transient simulators based on the two-fluid modelling approach were developed where simplifications such as the single mixture-energy equation were found to be convenient. In case of steady-state mechanistic models the flow-pattern prediction independent of the inclination was a crucial factor for further improvement resulting

[1] (Brill & Mukherjee, 1999)

in comprehensive mechanistic models such as Xiao et al. (1990), Ansari et al. (1988) and Hasan and Kabir (1988). Evaluation of these models showed that the comprehensive approach is more precise compared to the empirical one.

Currently available multiphase flow simulators are, for example, PipeSim®, UniSim® and OLGA-S. As stated in “Offshore Engineer” (July 2011, p.70) a “next-generation” multiphase flow simulator called “LedaFlow” (Kongsberg Oil & Gas Technologies) is to be announced soon. Both OLGA-S and LedaFlow are developed to model two- and three phase flows.

1.2. Statement of the problem

The purpose of this thesis is to develop tools in MS Excel for multiphase flow analysis and slug catcher design. Based on this development it is desired to gain knowledge for ILF concerning multiphase flow in pipelines.

Therefore, chapter 2 introduces the most common multiphase flow phenomena. Chapter 3 and 4 provide theoretical concepts which serve as basis for the multiphase flow analysis tool. Chapter 5 is dedicated to perform multiphase flow analysis with both the developed Excel tool and the multiphase flow simulator PipeSim. This chapter is also intended to validate the developed multiphase flow tool. Chapter 6 discusses concepts of slug catcher design and the implementation of these concepts in the developed slug catcher tool. A validation of the tool based on a given study is also included.

Finally, chapter 7 and 8 provide a summary and conclusions.

1.3. Scope of this thesis

Multiphase flow analysis are either based on empirical correlations or on mechanistic models. For the development of the multiphase flow tool the mechanistic model according to Xiao et al. (1990) was selected. Therefore, the explanations found in chapter 3 are limited to this model, even though several (empirical) multiphase flow calculation models are available. Detailed descriptions of empirical correlations are given in Brill and Mukherjee (1999).

Furthermore, chapter 3 does not comprise any theoretical abbreviations but states the main equations to predict the pressure gradient depending on the predominant flow pattern. If further details on the pressure gradient or on the flow pattern prediction are required the work of Xiao et al. (1990) and Taitel and Dukler (1976) respectively can be used.

The temperature calculation in section 3.3 is based on the PipeSim user guide (Schlumberger, PipeSim User Guide, 2010) where detailed relationships for determining heat transfer parameters are given. Discussing these relationships in detail is not subject of this master thesis. Therefore, if details on thermal conductivities and Nusselt-Numbers are required it is recommended to refer to the PipeSim user guide.

In analogy, it is resigned to introduce empirical correlations of the Black Oil Model introduced in section 4.2. A detailed description of these correlations can be found in Brill and Mukherjee (1999).

2. Multiphase flow phenomena

Multiphase flow is characterized by specific phenomena which are introduced in this section. Section 2.1 is emphasizing on the different types of flow pattern found in multiphase flow and section 2.2 explains typical parameters required to perform multiphase flow calculations.

2.1. Multiphase flow pattern[2]

What multiphase flow basically distinguishes from single phase flow are “flow patterns” or “flow regimes” which are induced due to different forces such as buoyancy, turbulence, inertia and surface tension.

Since the flow patterns are not completely identical it is a common practice to consider either horizontal or vertical flow. In case of two-phase horizontal flow four main flow patterns are found. As Figure 1 suggests there can be further classifications within these four patterns.

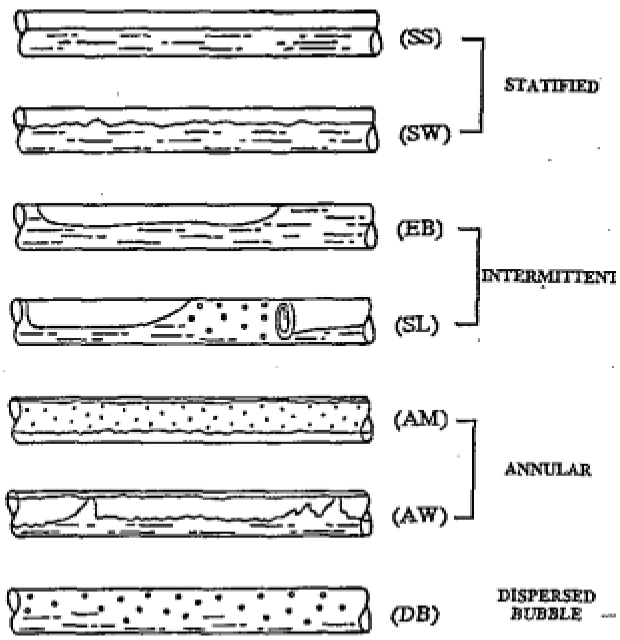


Figure 1: Flow pattern classification in two-phase horizontal flow acc. to Xiao *et al.* (1990). SS = Smooth Stratified; SW = Stratified Wavy; EB = Elongated Bubble; SL = Slug; AM = Annular Mist; AW = Annular Wavy and DB = Dispersed Bubble.

[2] (Brill & Mukherjee, Multiphase Flow in Wells, 1999)

Similarly to horizontal flow in vertical flow four main flow patterns are found as illustrated in Figure 2.

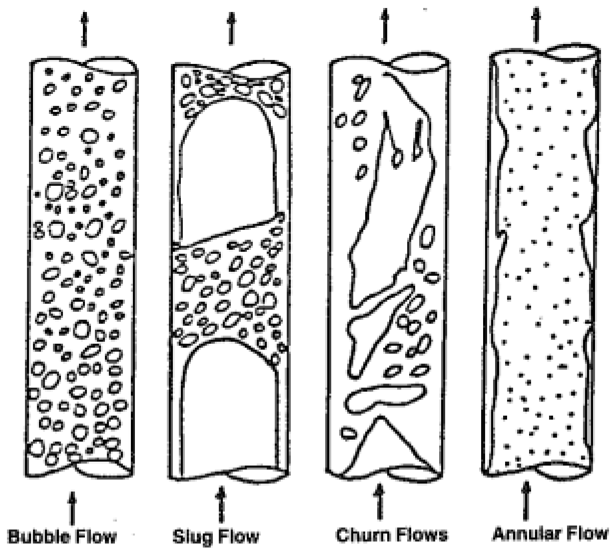


Figure 2: Flow pattern classification in two-phase vertical flow acc. to Brill and Mukherjee (1999).

2.1.1. Stratified flow

The stratified flow pattern is subdivided into “stratified wavy” and “stratified smooth”. Both patterns typically show a clear separation between the phases. While in “stratified smooth flow” the interface between liquid and gas phase is relatively flat in the “stratified wavy” flow pattern waves tend to grow on the liquid surface. In general, stratified flow occurs at relatively low liquid and gas velocities.

2.1.2. Intermittent flow

If gas and/ or liquid velocities are increased the flow pattern changes into intermittent flow where either “elongated bubble flow” or “slug flow” can develop. Higher gas velocities increase the transfer of impulse between gas and liquid phase causing waves on the liquid phase which can completely fill the cross section of the pipe. This phenomenon is called “slug flow”. The gas core between two waves accelerates the liquid phase and draws liquid from the slug which moves ahead and emits liquid to the slug behind the gas phase. The mass transfer of liquid phase takes place in a thin film beyond the gas core which travels at velocities lower than the gas core. In case of “elongated bubble flow” large gas bubbles are travelling along with the liquid phase (Stapelberg & Mewes, 1991).

2.1.3. Dispersed bubble flow

A further increase in liquid velocities results in distributed flow where the gas phase is dispersed in small and spherical gas bubbles.

2.1.4. Annular flow

Annular flow requires high gas velocities. In annular flow the gas phase forms a continuous core in the center of the pipe and the liquid is flowing as a thin film along the pipe wall as well as a dispersed phase in the gas core. Two different forms of annular flow (annular mist and annular wavy) may be encountered.

2.1.5. Churn flow

Churn flow is likely to occur in vertical flow and it can be characterized by chaotic flow behaviour where there is no continuity, neither in the gas nor in the liquid phase. Churn flow requires relatively high gas velocities in order to separate the liquid.

2.2. Multiphase flow variables[3]

Multiphase flow calculations require specific variables such as liquid hold up, mixture velocities and mixture fluid properties in order to describe the behaviour of the involved phases. The main objective of this chapter is to introduce these variables.

2.2.1. Liquid hold up

The liquid hold up H_L is a weighting factor which accounts for gas and liquid phase flowing simultaneously through a pipe section and is calculated as

$$H_L = \frac{volume\ of\ liquid\ in\ a\ pipe\ segment}{volume\ of\ pipe\ segment} \qquad \text{Equation 2-1}$$

[3] (Brill & Mukherjee, Multiphase Flow in Wells, 1999)

The remaining pipe section which is not occupied by the liquid must be occupied by gas and therefore the liquid hold up and the gas hold up can be correlated as

$$H_g = 1 - H_L \qquad \text{Equation 2-2}$$

where H_g is the gas hold up.

If a gas and liquid phase is present in a pipe the gaseous phase is usually travelling at higher velocities than the liquid phase because of the higher mobility of the gas phase (Brill & Mukherjee, 1999). The difference in velocities is termed “slippage”.

2.2.2. Velocities

Multiphase flow calculations involve different velocities such as “superficial velocity”, “mixture velocity” and “slip velocity”. Superficial velocities are important for the determination of liquid hold up and volumetric flow rates. Depending on which phase is considered the superficial velocity is defined as

$$v_{SL} = \frac{q_L}{A_P} \qquad \text{Equation 2-3}$$

and

$$v_{Sg} = \frac{q_g}{A_p} \qquad \text{Equation 2-4}$$

where q_L = liquid in-situ flow rate in m^3/s; q_g = gas in-situ flow rate in m^3/s and A_P = pipe area in m^2. Therefore, the superficial velocity corresponds to the velocity that the liquid or gas phase would reach if it flowed through the total pipe cross section alone.

By combining liquid and gas superficial velocities one can obtain the mixture velocity as

$$v_m = v_{SL} + v_{Sg} = \frac{q_L + q_g}{A_p} \qquad \text{Equation 2-5}$$

As Equation 2-5 shows the mixture velocity is the velocity at which both the gas and liquid phase would flow if there were no slip between the phases. But since slip between the phases is a common phenomenon in multiphase flow both phases flow at divergent velocities from the mixture velocity. While the velocity of the gas phase is typically greater the liquid usually flows at a velocity less than the mixture velocity.

Based on the superficial velocity the actual velocity for each phase can be calculated with the liquid hold up as

$$\mathrm{v_L} = \frac{\mathrm{v_{SL}}}{\mathrm{H_L}} \qquad \text{Equation 2-6}$$

and

$$\mathrm{v_g} = \frac{\mathrm{v_{Sg}}}{1-\mathrm{H_L}} \qquad \text{Equation 2-7}$$

Finally, a slip velocity can be defined as the difference in the actual phase velocities

$$v_s = v_g - v_L \qquad \text{Equation 2-8}$$

3. Multiphase flow calculation models

In order to perform comprehensive multiphase flow analysis specific calculation models are necessary. Typical parameters of interest in a multiphase flow analysis are the flow pattern, the pressure gradient, the temperature profile and slug prediction. For each of these parameters individual models were developed and the aim of this chapter is to introduce these models.

3.1. Flow pattern prediction[4]

The mechanistic model according to Xiao *et al.* (1990) is a comprehensive model developed for gas-liquid two phase flow in horizontal and nearly horizontal pipelines. It is able to detect the existing flow pattern and to predict the pressure drop (section 3.2) of any flow pattern.

For predicting the flow pattern this model is based on the work of Taitel and Dukler (1976) with some modifications. The following flow pattern transitions are identified:

Stratified-Non Stratified

$$v_g > \left(1 - \frac{h_L}{D}\right)\left[\frac{(\rho_L - \rho_g) g \cos\alpha A_g}{\rho_g \left(\frac{dA_L}{dh_L}\right)}\right]^{1/2}$$

where v_g = gas velocity in m/s; h_L = liquid hold up; D = diameter in m; ρ_L = liquid density in kg/m^3; ρ_g = gas density in kg/m^3; g = acceleration due to gravity in m/s^2; α = inclination in rad; A_g = pipe cross section occupied by gas m^2 and A_L = pipe cross section occupied by liquid in m^2.

Intermittent-Annular

$$\frac{h_L}{D} < 0.35$$

where h_L = liquid hold up and D = diameter in m.

[4] (Xiao, Shoham, & Brill, 1990)

Intermittent-Dispersed Bubble

$$v_L > \left[\frac{4A_g}{S_i}\frac{g\cos\alpha}{f_L}\left(1-\frac{\rho_g}{\rho_L}\right)\right]^{1/2}$$

where v_L = liquid velocity in m/s; A_g = pipe cross section occupied by gas m^2; S_i = periphery between gas and liquid in m; ρ_L = liquid density in kg/m^3; ρ_g = gas density in kg/m^3 and f_L = friction factor of the liquid.

Stratified-Smooth and Stratified-Wavy

$$v_g > \left[\frac{4\mu_L(\rho_L-\rho_g)g\cos\alpha}{s\rho_L\rho_g v_L}\right]^{1/2}$$

where v_g = gas velocity in m/s; v_L = liquid velocity in m/s; μ_L = liquid viscosity in Pa*s; ρ_L = liquid density in kg/m^3; ρ_g = gas density in kg/m^3 and s = sheltering coefficient.

For stratified flow in downwardly inclined pipes

$$\frac{v_L}{\sqrt{gh_L}} > 1.5$$

where v_L = liquid velocity in m/s and h_L = liquid hold up.

Further details on the transition criteria are found in the paper of Xiao *et al*. (1990) and in the work of Taitel and Dukler (1976).

The different flow patterns found by the Xiao mechanistic flow pattern are visualized in Figure 3, a so-called “flow pattern map”. Line A represents the transition from stratified-wavy (SW) to intermittent (I) if liquid superficial velocities are increased. In case of high gas superficial velocities line A indicates the transition from stratified-wavy to annular (AN) flow pattern. From intermittent flow pattern either the dispersed bubble regime can be reached by crossing line C or the annular flow pattern by crossing line B. If low gas and liquid velocities are found the smooth-stratified (SS) flow pattern may be encountered. Increasing liquid velocities result in a transition from smooth-stratified to stratified-wavy flow regime as indicated by line E and D respectively.

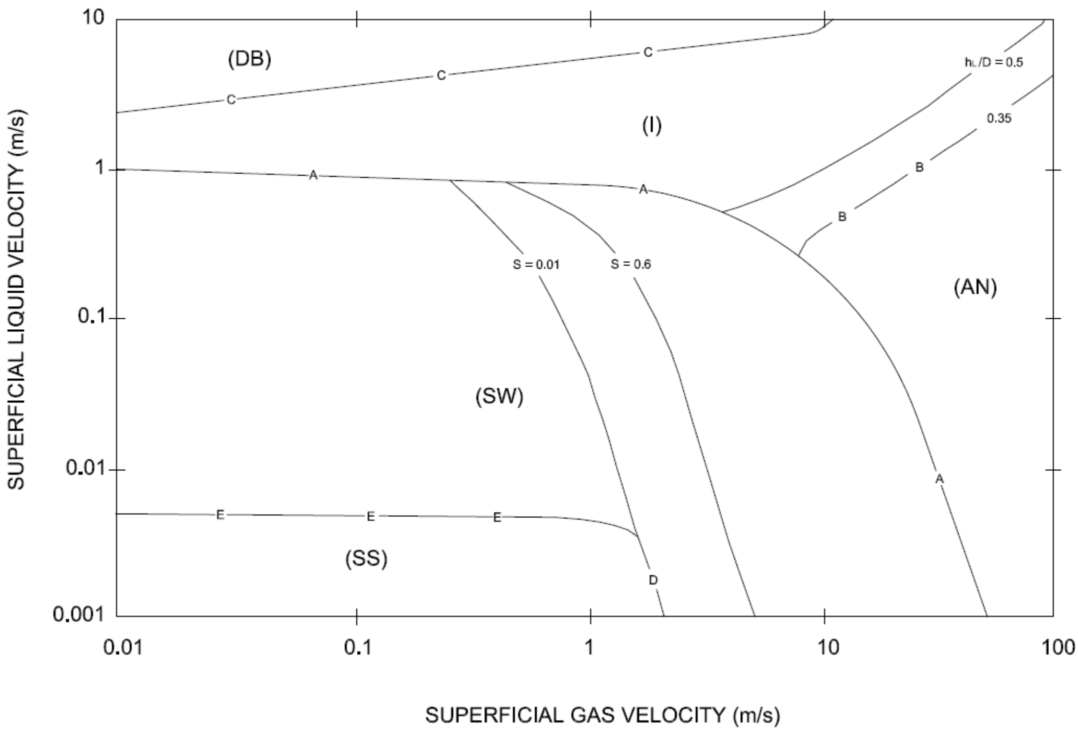


Figure 3: Flow pattern map indicating flow pattern transitions in an air-water system in a 5cm pipe of 1° inclination acc. to Xiao *et al.* (1990).

3.2. Pressure gradient calculation[5]

The basic idea of a mechanistic model is to describe the phenomenon of multiphase flow in pipes by providing a mechanistic analysis of the flow behaviour. In contrast to many empirical correlations, mechanistic models aim for solving momentum equations with attention to the flow pattern. The only empiricism that is involved in mechanistic models are closure relationships which are empirical relations for the determination of unknown parameters occurring in the momentum equations. Derivations of the pressure gradient equations for the particular flow patterns are not presented here. The following sections provide the main pressure gradient equation of the particular flow pattern only. If further details are required refer to Xiao *et al.* (1990).

3.2.1. Stratified flow model

In stratified flow liquid flows in the bottom portion of the pipe due to its higher density while the gas phase is traveling usually at higher velocities in the upper section of the pipe. The stratified flow pattern is shown in Figure 4.

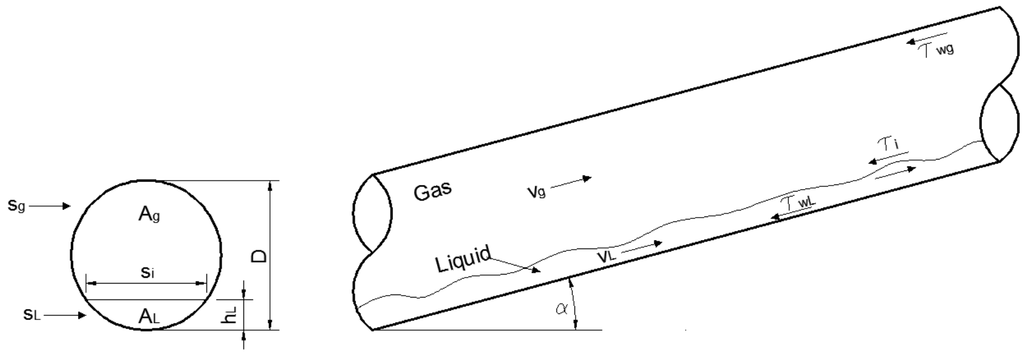


Figure 4: Stratified flow pattern according to Xiao *et al.* (1990).

The pressure gradient can be obtained from

$$-\frac{dp}{dx} = \frac{\tau_{wL}S_L + \tau_{wg}S_g}{A} + \left(\frac{A_L}{A}\rho_L + \frac{A_g}{A}\rho_g\right) g \sin\alpha \qquad \text{Equation 3-1}$$

where dp/dx = pressure gradient in Pa/m; τ_{wL} = shear stress of the liquid at the pipe wall in Pa; S_L = wetted liquid periphery in m; τ_{wg} = shear stress of the gas

[5] (Xiao, Shoham, & Brill, 1990)

at the pipe wall in Pa; S_g = wetted gas periphery in m; A = pipe cross section in m^2; A_L = cross section occupied by liquid in m^2; A_g = cross section occupied by gas in m^2; ρ_L = liquid density in kg/m^3; ρ_g = gas density in kg/m^3; g = acceleration due to gravity in m/s^2 and α = inclination in deg;

The first term on the right hand side in Equation 3-1 represents the frictional pressure gradient and the second term represents the gravitational pressure gradient.

3.2.2. Intermittent flow model

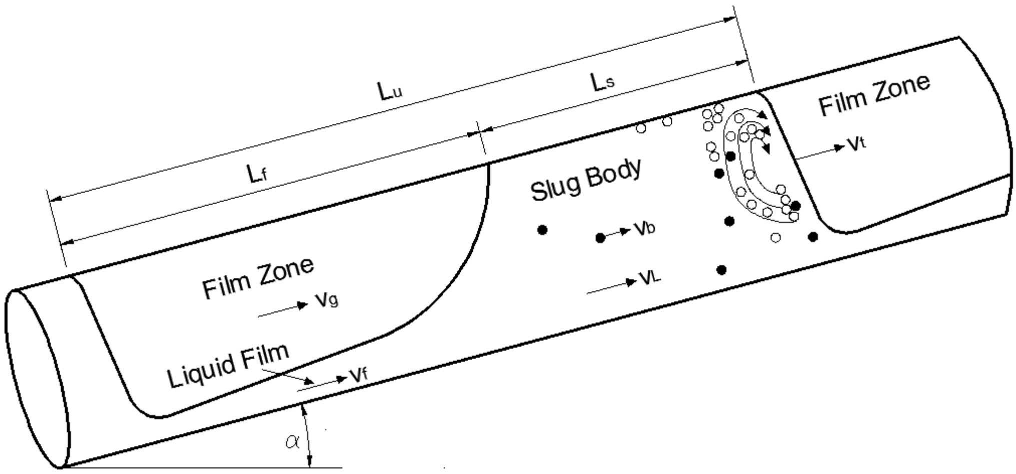


Figure 5: Intermittent flow pattern according to Xiao *et al*. (1990).

As can be seen in Figure 5, in intermittent flow a liquid slug body is followed by a gas pocket. While the slug fills the entire pipe cross section the gas pocket is flowing above a stratified liquid layer which is moving relatively slowly compared to the slug body. Conversely, the slug is moving at high velocities and at the front and the top side of the slug gas accumulates in form of small gas bubbles.

The average pressure gradient for intermittent flow is determined from Equation 3-2.

$$-\frac{dp}{dx} = \rho_u g \sin\alpha + \frac{1}{L_u}\left[\frac{\tau_f S_f + \tau_g S_g L_f}{A} + \frac{\tau_s \pi D L_s}{A}\right] \qquad \text{Equation 3-2}$$

where ρ_u = density of slug unit kg/m3; L_u = length of slug unit in m; τ_f = shear stress of the film in Pa; S_f = wetted liquid periphery in film in m; τ_g = shear stress of the gas in Pa; D = diameter in m; S_g = wetted gas periphery in m: L_f = film length in m and L_s = slug length in m;

Equation 3-2 consists of a gravitational pressure gradient (first term on the right hand side) and a frictional pressure gradient (second term) which is resulted by friction loss in the slug body and in the film zone.

3.2.3. Annular flow model

As described previously, in annular flow a gas core flowing through the pipe is surrounded by a liquid film attached to the pipe wall. Furthermore, the liquid phase is dispersed in the gas core in form of droplets. Annular flow is schematically shown in Figure 6. While in vertical flow case the liquid film is circumferentially uniform in case of horizontal flow the liquid film is usually thicker at the bottom than at the top of the pipe.

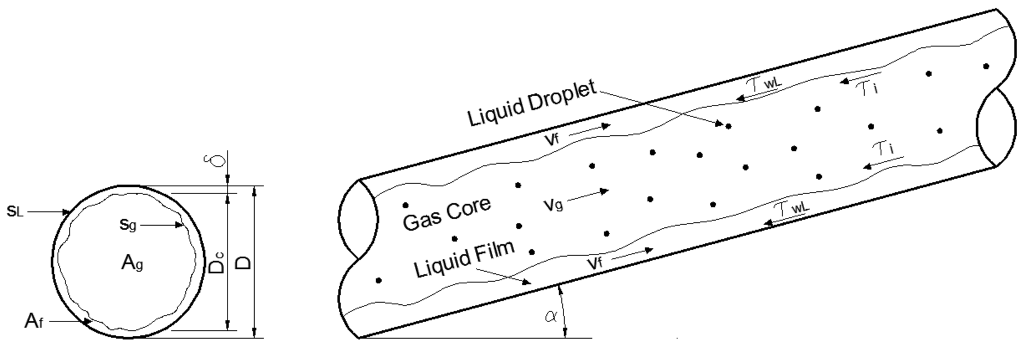


Figure 6: Annular flow pattern according to Xiao *et al*. (1990).

The pressure gradient for annular flow is calculated from

$$-\left(\frac{dp}{dx}\right) = \frac{\tau_{wL} S_L}{A} + \left(\frac{A_f \rho_L}{A} + \frac{A_C \rho_C}{A}\right) g \sin\alpha \qquad \text{Equation 3-3}$$

where τ_{wL} = shear stress of the liquid at pipe wall in Pa; S_L = wetted liquid periphery in m; A = pipe cross section in m^2; A_f = cross section occupied by film in m^2; A_C = cross section occupied by gas core in m^2; ρ_L = liquid density in kg/m^3 and ρ_C = gas core density in kg/m^3;

The first term of the right hand side is the frictional pressure gradient while the second term is the gravitational gradient.

3.2.4. Dispersed bubble flow model

Since there is no slippage between the phases in dispersed bubble flow a pseudo-single phase with average properties can be assumed. Thus, the calculation of the pressure gradient is simplified considerably.

The pressure gradient consists of a friction loss component (first term on the right hand side) and a gravitation loss component (second term on the right hand side):

$$-\left(\frac{dp}{dx}\right) = \frac{2 f_m \rho_m v_m^2}{d} + \rho_m g \sin\alpha \qquad \text{Equation 3-4}$$

where f_m = mixture friction factor; ρ_m = mixture density in kg/m^3; v_m = mixture velocity in m/s and d = diameter in m.

3.3. Temperature gradient calculation[6]

The theoretical considerations in this chapter which are limited to stratified flow are based on the PipeSim user guide. In analogy to the user guide relationships to calculate the heat transfer coefficients are provided (section 3.3.1). Section 3.3.2 gives the enthalpy balance for a fully exposed pipe in order to predict the temperature profile for a pipe segment.

3.3.1. Heat transfer coefficients

Heat transfer occurs between the environment and the pipe, in the pipe wall and between the pipe wall to the fluid (gas or liquid). Therefore, three different heat transfer coefficients (sections 3.3.1.1 - 3.3.1.3) have to be determined which are then used to calculate an overall heat transfer coefficient (section 3.3.1.4).

[6] (Schlumberger, 2010)

3.3.1.1. Internal fluid film heat transfer coefficient

The internal fluid film is considered to be a homogenous mixture of liquid and gas flowing through the pipe. It is defined as

$$\alpha_{film} = \frac{k_{mix} N_{Nu,film}}{L} \qquad \text{Equation 3-5}$$

where k_{mix} = mixture thermal conductivity in W/m/K; N_{Nu} = Nusselt number of the fluid film and L = a given length in m.

3.3.1.2. Conductive heat transfer coefficient

Considering the heat transfer through a pipe wall one can obtain the conductive heat transfer coefficient from

$$k_{pipe} = \frac{2\pi\lambda_{wall}}{A} \cdot \frac{1}{ln(r_{out}/r_{in})} \qquad \text{Equation 3-6}$$

where λ_{wall} = thermal conductivity of the pipe wall in W/m/K; A = pipe cross section in m^2; and r_{out} and r_{in} = pipe radii in m.

3.3.1.3. Convective heat transfer coefficient for a fully exposed pipe

Assuming that the pipe is fully exposed to air (or water) heat transfer due to free convection takes place on the outside of the pipe. The heat transfer coefficient for this case can be calculated from

$$\alpha_{amb} = \frac{k_{amb} N_{Nu}}{D} \qquad \text{Equation 3-7}$$

where k_{amb} = fluid conductivity of the ambient medium in W/m/K; N_{Nu} = Nusselt number and D = pipe diameter in m.

3.3.1.4. Overall Heat Transfer Coefficient

After having evaluated the partial heat transfer coefficients an overall heat transfer coefficient U can be calculated:

$$U = \frac{1}{\frac{1}{\alpha_{amb}} + \frac{1}{k_{pipe}} + \frac{1}{\alpha_{film}}} \qquad \text{Equation 3-8}$$

The overall heat transfer coefficient allows to estimate the total heat flow $\dot{Q}_{tot}$ which occurs between the fluid in the pipe and the ambient medium as follows

$$\dot{Q}_{tot} = U\pi r_{out}(T_{amb} - \overline{T_1}) \qquad \text{Equation 3-9}$$

where U = overall heat transfer coefficient in $W/m^2/K$; r_{out} = outer pipe diameter in m; T_{amb} = ambient fluid temperature in K and $\overline{T}_1$ = average fluid temperature in the pipe in K.

3.3.2. Enthalpy balance for a fully exposed pipe

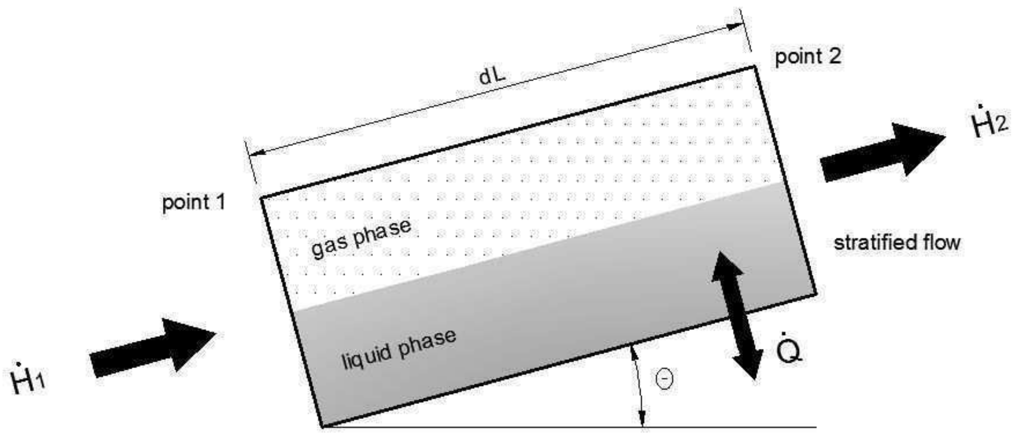


Figure 7: Heat flow in stratified flow pattern.

Figure 7 shows a considered pipe segment with equilibrium stratified flow at an inclination angle θ. The pipe is supposed to be fully exposed to air (or water) and the temperature profile between point 1 and point 2 is to be calculated.

The total enthalpy balance for the considered pipe segment in Figure 7 is defined as

$$\dot{H}_1 + \dot{Q} = \dot{H}_2 \qquad \text{Equation 3-10}$$

where $\dot{H}_1$ = enthalpy flow at point 1 in J/s; $\dot{Q}$ = transferred heat in J/s and $\dot{H}_2$ = resulting enthalpy flow at point 2 in J/s.

To determine the enthalpy flows at point 1 and point 2 it is recommended to refer to the "Black Oil Fluid enthalpy Model" (2009 Method) as introduced in PipeSim user guide.

3.4. Slug prediction

Slug flow in pipelines poses a challenge to operation of downstream oil or gas treatment facilities due to its unpredictability of liquid and gas flow rates. Different mechanisms such as hydrodynamic slugging, terrain slugging, operating fluctuations and pigging can trigger slugging in multiphase flow. These mechanisms are explained in section 3.4.1.

In order to minimize the effects of slugging on downstream facilities and to ensure stable operation conditions slug catchers are installed at the end of a field line. The design of a slug catcher requires the knowledge of the slug length. Methods to evaluate the slug length are presented in section 3.4.2.

3.4.1. Mechanisms of slugging[7]

In slug flow the liquid completely fills the cross section of the pipe with a certain frequency. This flow pattern can be resulted by different mechanisms. The most common mechanisms are hydrodynamic slugging, terrain slugging, slugging due to operational fluctuations and pigging. Additionally, severe slugging can also occur in offshore risers (Hall).

3.4.1.1. Hydrodynamic slugging

As a result of intensive impulse transfer between gas and liquid phase waves can develop and grow until they fill the entire cross section of the pipe (Stapelberg & Mewes, 1991). Such conditions for slug formation can even be observed in smooth horizontal pipes at constant inlet velocity (Hall). Once the liquid occupies the entire cross section it

[7] (Hall)

can be further accelerated by the gas phase drawing and emitting liquid from the film on the bottom of the pipe.

3.4.1.2. Terrain slugging

Changes in inclination can cause so called “terrain slugging” where liquid is accumulated in dips or hollows in a pipeline. Consequently, the gas flow is interrupted leading to an increase in gas pressure until the hydrostatic pressure of the liquid is overcome. The liquid is then transported further downstream. Even at shallow inclination angles terrain slugging can occur if the predominant flow pattern is stratified and if the pipeline is at turndown conditions.

3.4.1.3. Slugging due to operational fluctuations

Operational changes such as increases in gas velocity can cause so called “sweep out” or “upset” slugs. Due to higher gas velocities the liquid hold up is lower than before. Thus, the excess liquid is swept out of the pipeline causing intermittency in the outlet flow. Other reasons for slugging due to operational fluctuations are changes in well allocations or pressure fluctuations in the gas/liquid separator.

Turndown and ramp-down/ramp-up operations can result in slugging and therefore need to be considered carefully. As flow rates are decreased the lower velocities result in a higher tendency of the liquid to settle down at low points (Transient Analysis Report, 2011). Therefore, it is recommended that ramp-down operations are carried out in a relative short period of time to avoid too much liquid settling down. This is shown in Figure 8. On the contrary, in ramp-up operations the production should be increased gradually because as the flow rates increase again accumulated liquid is dragged by the coming fluid what can result in overloading of the downstream facilities.

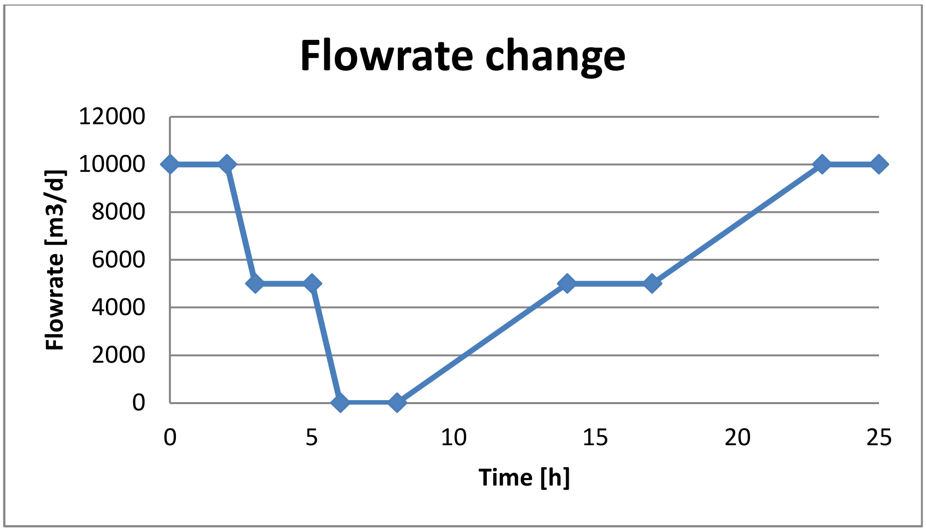


Figure 8: Hypothetical ramp-down/ramp-up scenario (Transient Analysis Report, 2011).

3.4.1.4. Severe slugging

As shown in Figure 9 severe slugging occurs in horizontal pipelines which slope vertically to a receiving platform, commonly a separator.

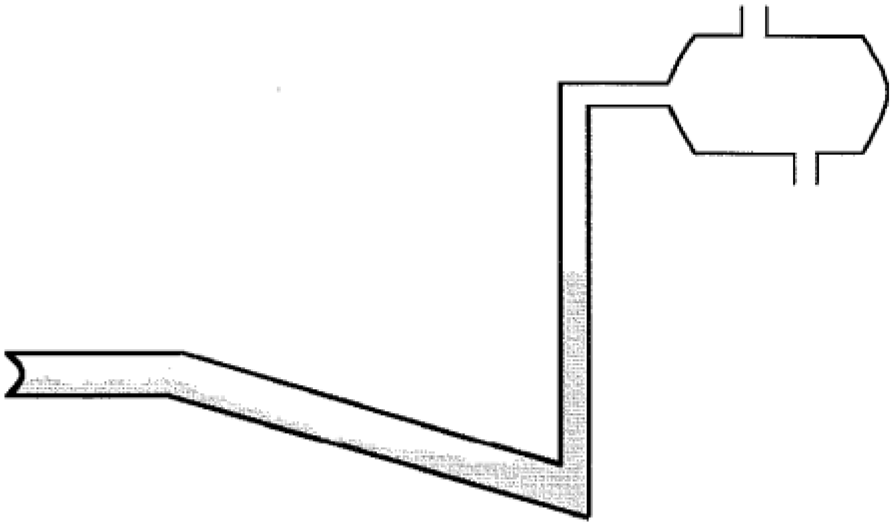


Figure 9: Slug formation in a riser connected with a platform (Brill & Sarica, 2010).

If gas velocities are low the liquid cannot be carried up the riser and therefore liquid is accumulated in the riser. The length of so formed slugs is continually increased and can exceed the height of the riser up to several times.

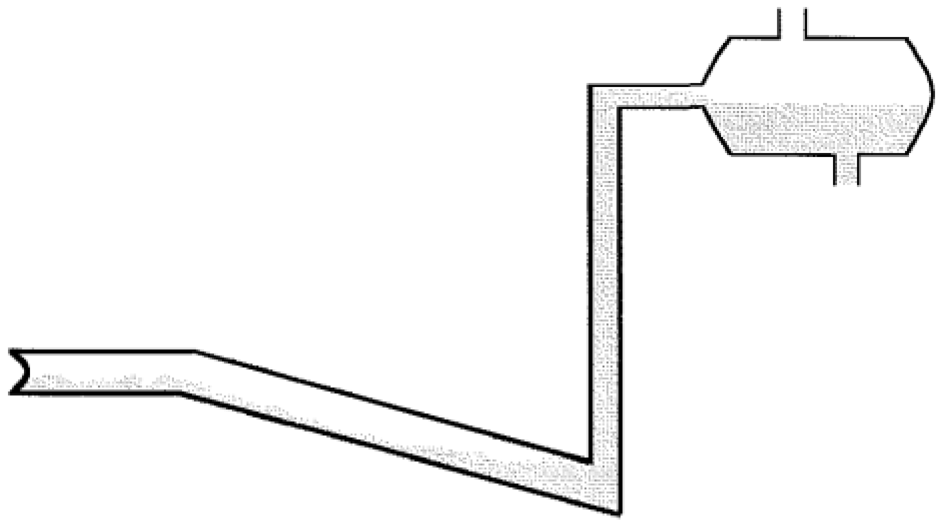


Figure 10: Slug production in a riser connected with a platform (Brill & Sarica, 2010).

This type of slugging causes periods where no liquid or gas is transported to the separator followed by very high gas and liquid flow rates. Thus, flooding can occur, resulting in a liquid carryover of the separator as shown in Figure 11.

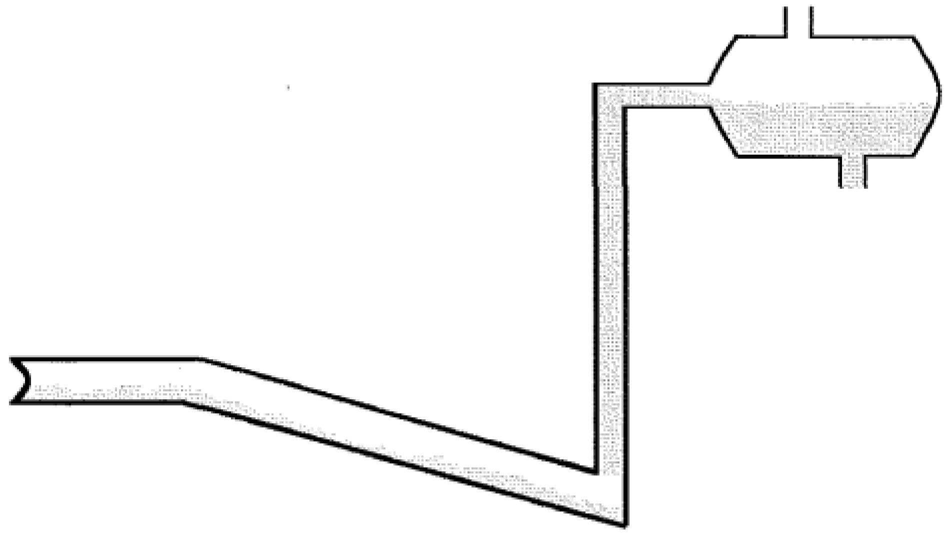


Figure 11: Blowout in a riser connected with a platform (Brill & Sarica, 2010).

The exerted forces on the pipeline are very large and may stress fittings severely, hence this type of slugging is termed “severe slugging”.

Bøe (1981) suggested a criterion (“PI-SS”) to predict the occurrence of severe slugging. Assuming no mass transfer between the phases, large differences in densities $\rho_L \gg \rho_G$ and neglecting liquid fallback in the riser the criterion is evaluated from

$$\Pi_{SS} = \frac{\left.\frac{\partial p}{\partial t}\right|_{Pipeline}}{\left.\frac{\partial p}{\partial t}\right|_{Riser}} < 1 \qquad \text{Equation 3-11}$$

The pipeline pressure build up rate and the hydrostatic pressure increase rate are defined as:

$$\left.\frac{\partial p}{\partial t}\right|_{Pipeline} = \frac{ZRT}{M}\frac{v_g}{LA(1-H_{LF})} \qquad \text{Equation 3-12}$$

$$\left.\frac{\partial p}{\partial t}\right|_{Riser} = \frac{\rho_L g v_L}{A\rho_L} = \frac{g v_L}{A} \qquad \text{Equation 3-13}$$

where Z = gas compressibility factor; R = gas constant, 8,314 J/K/kmol; M = molecular weight of gas in kg/kmol; L = pipeline length in m; v_g and v_L are the phase velocities in m/s; and A = cross sectional area of pipeline and riser in m^2.

The following Figure 12 shows the region in which severe slugging occurs in dependence of superficial velocities.

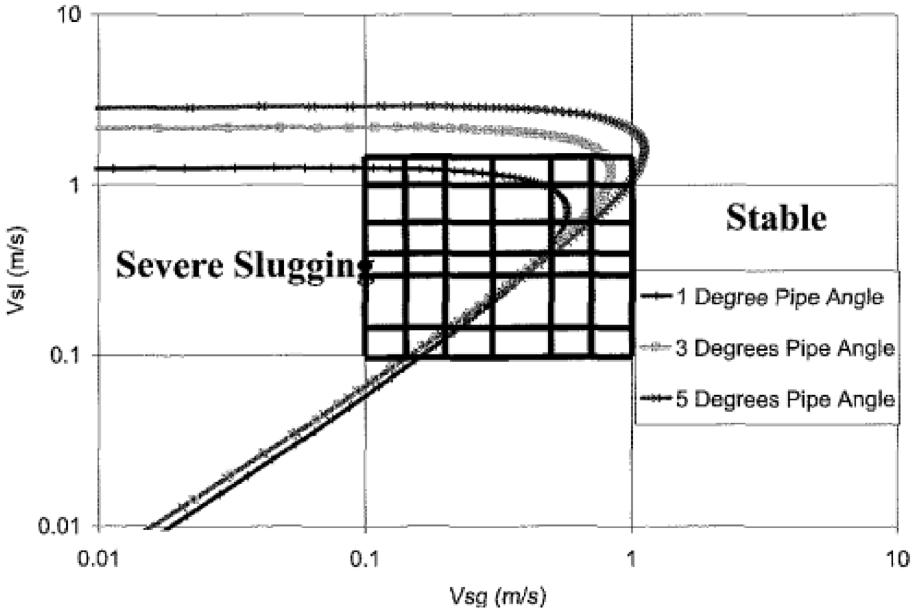


Figure 12: Severe slugging region in a riser (Brill & Sarica, 2010).

3.4.1.5. Pigging

Slugs can also be produced during pigging operations where pigs or spheres are injected into pipelines in order to remove accumulated liquid volumes. Pigging can be carried out at full but also at reduced flow rates. As indicated in Transient Analysis Report (ILF, 2011) pigging operations at lower flow rates can lead to smaller surge liquid volumes compared to operations at full flow rate. Therefore, a smaller slug catcher could be installed.

3.4.2. Slug length

Several correlations for slug length prediction were proposed but with limitations in terms of the pipe diameter. The following table shows examples of slug length correlations with related pipe diameters for horizontal flow.

Table 1: Correlations for predicting slug length in horizontal flow with related pipe diameter.

Correlation	Pipe diameter
Norris Correlations $$Ln(\bar{L}_S) = -2.099 + 4.859\sqrt{Ln(d)}$$ where d = pipe diameter, in.	≤ 24”
Brill *et al.* Correlation $$Ln(\bar{L}_S) = -2.663 + 5.441\sqrt{Ln(d)} + 0.059Ln(v_m)$$ where d = pipe diameter, in; and v_m = mixture velocity, ft/s.	≤ 16”
Scott *et al.* Correlation $$Ln(\bar{L}_S) = -26.6 + 28.5[Ln(d) + 3.67]^{0.1}$$ where d = pipe diameter, ft.	≤ 24”

Slugs can vary widely in length and it is not possible to predict an accurate slug length because the lengths follow a statistical distribution. Usually the 1 in 1000 slug is of interest for designing a slug catcher and it is common to assume a log normal distribution for the slug lengths. After determining the mean slug length with one of the mentioned correlations the (design) slug length can then be calculated from

$$X_\alpha = e^{\left(\sigma Z_\alpha + \ln(\bar{L}_S)\right)} \qquad \text{Equation 3-14}$$

where α = probability (1 in 10, 1 in 100, 1 in 1000 etc.); Z_α = mean and σ = standard deviation.

4. Fluid property approximation

The aim of this chapter is to introduce two models which can be used to approximate fluid physical properties under given conditions. While section 4.1 discusses the "Compositional Model" briefly section 4.2 explains the "Black Oil Model" in more detail which was also implemented in the code of the multiphase flow program.

4.1. Compositional Model [8]

If the composition of a fluid mixture is known a vapour/liquid equilibrium (VLE) calculation determines the amount that exists in vapour and liquid phases and the composition of each phase. Once the composition is calculated physical properties such as density, viscosity and surface tension can be determined. Especially in case of volatile oils the compositional model is more accurate to describe mass transfer compared to the Black Oil Model (section 4.2).

4.1.1. Density

Based on equations of state (EOS's) the compressibility factor Z_i for vapor and liquid phase can be determined. The compressibility factor is required to obtain vapor and liquid density ρ_i as follows

$$\rho_v = \frac{p\sum_{1=1}^{n} y_i M_i}{Z_V RT} \quad \text{Equation 4-1}$$

and

$$\rho_L = \frac{p\sum_{1=1}^{n} x_i M_i}{Z_L RT} \quad \text{Equation 4-2}$$

where y_i = mole fraction of component i in vapour phase; x_i = mole fraction of component i in liquid phase M_i = molecular weight of the i^{th} component in kg/kmol; R = gas constant in 8,314 J/K/mol and T = temperature in °C.

[8] (Brill & Mukherjee, 1999)

4.1.2. Viscosity

Gas viscosity μ_g can be calculated according to Lee *et al.* (1966).

Liquid viscosity μ_L can be expressed as

$$\mu = f(p, T, x_i), i = 1, \dots, n$$

If hydrocarbon liquids are considered the composition of the liquid must be known.

4.1.3. Surface tension

The surface tension σ can be obtained from

$$\sigma^{1/4} = \sum_{i=1}^{n} [(P_{ch})_i (Ax_i - By_i)] \qquad \text{Equation 4-3}$$

where P_{ch} = a temperature-dependent parameter and A and B = density-dependent parameters.

4.1.4. Heat capacity

The constant pressure heat capacity C_p^0 is defined as

$$C_p^0 = \left(\frac{dH}{dT}\right)_{p^0} \qquad \text{Equation 4-4}$$

where the superscript 0 indicates that the fluid is an ideal gas.

4.2. Black Oil Model [9]

The „Black Oil Model" is a model used to predict physical properties of any liquid phase that contains dissolved gas. It usually refers to oils which are dark in colour and have gravities less than 40° API (Brill & Mukherjee, Multiphase Flow in Wells, 1999). Such fluid systems typically undergo small changes in composition within the two-phase envelope. Therefore, the Black Oil Model is also termed "Constant Composition Model". The following sections illustrate how to determine gas and oil physical properties.

4.2.1. Gas physical properties

In its simplest way the physical properties of gases such as pressure p, temperature T and volume V are correlated by the ideal gas law

$$pV = nRT \qquad \text{Equation 4-5}$$

where p = absolute pressure in bara; V = volume of gas in m^3; T = absolute temperature in °C; n = number of moles of gas in mol; and R = universal gas constant = 8,314 J/mol/K.

However, the ideal gas law fails to predict the properties of real gases since interactions such as repulsive and attractive forces become significant for real gases. In order to meet deviations of real gases from the ideal gas the ideal gas law is modified into the real gas law:

$$pV = ZnRT \qquad \text{Equation 4-6}$$

where Z = compressibility factor.

[9] (Brill & Mukherjee, 1999)

4.2.1.1. *Gas specific gravity*

The specific gravity of gas is defined as

$$\gamma_g = \left(\frac{\rho_g}{\rho_a}\right)_{sc} = \frac{M_g}{M_a} \qquad \text{Equation 4-7}$$

where γ_g = specific gravity of gas in kg/m^3; ρ_a = density of air at standard conditions = 1,225 kg/m^3 (@ T=15°C, P=1,013 bar) and M_a = molecular weight of air = 28,96 kg/kmol.

4.2.1.2. *Gas density*

From the real gas law the gas density can be expressed as

$$\rho_g = \frac{m}{V} = \frac{pM_g}{ZRT} \qquad \text{Equation 4-8}$$

where ρ_g = density of gas in kg/m^3; m = weight of gas in kg and M_g = molecular weight of gas in kg/kmol.

4.2.1.3. *Gas viscosity*

Lee *et al.* (1966) supposed a method to predict gas viscosity empirically as follows

$$\mu_g = 10^{-4} K exp\left[X\left(\frac{\rho_g}{62,4}\right)^Y\right] \qquad \text{Equation 4-9}$$

where K = (9,4+0,02*M_g)$T^{1,5}$/(209+19*M_g +T); X = 3,5+(986/T)+0,01*M_g; Y = 2,4-0,2*X; ρ_g = gas density at reservoir pressure and temperature in lbm/ft^3; T = reservoir temperature in °R and M_g = molecular weight of the gas mixture in lbm/mol.

4.2.1.4. *Gas formation volume factor*

The gas formation volume factor B_g is the volume occupied by the gas at given conditions (at standard conditions this factor = 1). It is defined as

$$B_g = 0{,}0283\frac{ZT}{p} \qquad \text{Equation 4-10}$$

where B_g = gas formation volume factor in ft^3/scf; p = pressure in psia and T = absolute temperature in °R.

4.2.2. Oil Physical Properties

The following sections briefly explain how physical properties such as oil gravity, gas solubility, oil formation volume factor, oil density, oil viscosity and surface tension are correlated in the Black Oil model.

4.2.2.1. *Oil gravity*

The API oil gravity can be obtained from

$$\gamma_{API} = \frac{141{,}5}{\gamma_O} - 131{,}5 \qquad \text{Equation 4-11}$$

where γ_O = specific gravity of stock tank oil at 60°F.

4.2.2.2. *Gas solubility*

Gas Solubility is used to determine the mass transfer between the liquid and gaseous phase. It is defined as the volume of gas dissolved in one stock tank barrel of oil at a given pressure and temperature. Brill and Mukherjee (1999) presented several empirical correlations to determine gas solubility.

4.2.2.3. Oil formation volume factor

The Oil Formation Volume Factor B_o is defined as the volume in barrels which is occupied by one stock tank barrel of oil containing dissolved gas at any given pressure and temperature.

$$B_O = \frac{(V_O)_{p,T}}{(V_O)_{sc}} \qquad \text{Equation 4-12}$$

where B_O = oil formation volume factor in m^3/Sm^3; $(V_O)_{p,T}$ = volume of oil (at pressure p, and temperature T) in m^3 and $(V_O)_{sc}$ = volume of oil at standard conditions in Sm^3.

4.2.2.4. Oil density

If the oil is below the bubble point pressure the density at a specified pressure and temperature can be calculated from

$$\rho_O = \frac{62.4\gamma_O + 0.0136 R_S \gamma_{gd}}{B_O} \qquad \text{Equation 4-13}$$

where γ_O = stock-tank oil specific gravity and γ_{gd} = dissolved-gas gravity. Correlations to determine γ_{gd} are also found in Brill and Mukherjee (1999).

4.2.2.5. Oil viscosity

Empirical relationships were developed in order to correlate oil viscosity for distinct temperature and pressure. The “dead oil viscosity” has to be determined first. The expression “dead oil” is indicating that there is no gas dissolved in the oil. Oil that contains dissolved gas is lower in viscosity and is termed “live oil”. If the pressure is increased the oil viscosity decreases due to the dissolved gas until the bubblepoint pressure is reached (“saturated crude oil”). If the system pressure exceeds the bubblepoint pressure the viscosity of the oil is increased due to its compressibility and it is then termed “undersaturated crude oil”.

Correlations to determine dead oil, saturated and undersaturated oil viscosity can be found in Brill and Mukherjee (1999).

4.2.2.6. Surface tension

Surface tension values are important for the determination of flow patterns and liquid hold up in multiphase flow. Based on the work of Baker and Swerdloff (1956), the dead-oil surface tension σ_{od} can be obtained graphically as a function of temperature and the API gravity (for details refer to Brill and Mukherjee (1999)).

4.2.3. Volumetric flow rates

Depending on whether the black oil model or the compositional model is used to calculate the physical and thermodynamical properties the in-situ volumetric flow rates are determined in different ways. In case of the black oil model the flow rates are defined as

$$q_o = q_{o,sc} B_o \qquad \text{Equation 4-14}$$

where q_O = oil flow rate in m^3/s at a given pressure and temperature; $q_{O,SC}$ = oil flow rate in Sm^3/s at standard conditions and B_O = oil formation volume factor in Sm^3/m^3.

$$q_w = q_{w,sc} B_w \qquad \text{Equation 4-15}$$

where q_w = water flow rate in m^3/s at a given pressure and temperature; $q_{W,SC}$ = water flow rate in Sm^3/s at standard conditions and B_W = water formation volume factor in Sm^3/m^3.

and

$$q_g = \left(q_{g,sc} - q_{o,sc} R_s - q_{w,sc} R_{sw}\right) B_g \qquad \text{Equation 4-16}$$

where q_g = gas flow rate in m^3/s at a given pressure and temperature; $q_{g,SC}$ = gas flow rate in Sm^3/s at standard conditions; R_s = gas/oil ratio in m^3/m^3 at a given pressure and temperature; R_{SW} = gas/water ratio in m^3/m^3 at a given pressure and temperature and B_W = water formation volume factor in Sm^3/m^3.

In case of the compositional model the flow rates are determined from

$$q_L = \frac{w_t(1-x_g)}{\rho_L} \qquad \text{Equation 4-17}$$

where q_L = liquid volume flow rate in m^3/s; w_t = mass flow in kg/s; x_g = gas mass fraction and ρ_L = liquid density in kg/m^3.

and

$$q_g = \frac{w_t x_g}{\rho_g} \qquad \text{Equation 4-18}$$

where q_g = gas volume flow rate in m^3/s; w_t = mass flow in kg/s; x_g = gas mass fraction and ρ_g = gas density in kg/m^3.

5. Multiphase flow calculation

The purpose of chapter 5 is to carry out multiphase flow calculations with PipeSim and with the developed multiphase flow tool. Based on a given multiphase flow example which is introduced in section 5.1 it is the purpose of section 5.2 to explain the configuration and to demonstrate the performance of the multiphase flow tool. Furthermore, a comparison of several multiphase flow correlations available in PipeSim is made in section 5.3. Finally, the sensitivity of different variables on pressure loss and slug length is investigated in section 5.4.

5.1. Basic data[10]

The problem which served as basis for the calculations in this chapter is summarized in Table 2. Data were taken from the course "Fluid Flow Projects" instructed by Dr. Brill and Dr. Sarica (2010). All required fluid properties were determined on the basis of the Black Oil model.

Table 2: Input data used for the pressure calculation.

Parameter	Data
Gas gravity @ P_{sep} = 8 bara and T_{sep} = 15°C	0,75
Water specific gravity	1
API gravity	30°API
Produced gas-oil ratio	267 sm^3/sm^3
Water cut	0 %
Pipe diameter	0,5 m
Pipe roughness	0,0183 mm
Liquid flow rate	1,6 * 10^4 sm^3/d

[10] (Brill & Sarica, 2010)

5.2. Multiphase flow tool

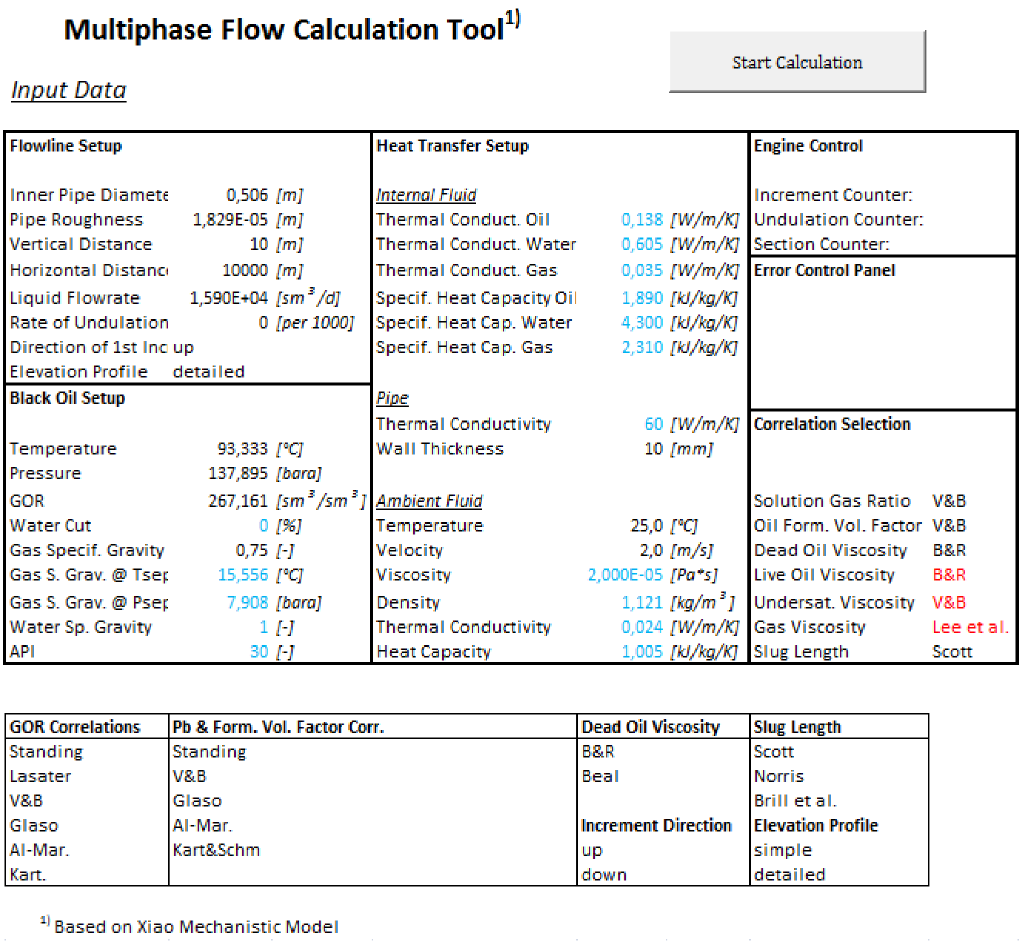
In this section the performance of the developed multiphase flow tool is demonstrated by solving the multiphase flow problem as introduced in the previous section. In order to understand the functions of the tool section 5.2.1 contains a configuration description introducing the main algorithms of the program. The results of the calculations are then presented in 5.2.2.

5.2.1. Configuration of multiphase flow tool

The main module of the program consists of five loops as indicated by the following sections and each of these loops is depicted in single flow diagrams. When opening the tool the user can see three spreadsheets. The first one is termed “Input sheet” where all required input data is specified. The second sheet called “Elevation-Profile” is required if “detailed elevation profile” option is selected and it contains a pressure plot. The “Output” sheet provides the user with the calculated data.

Figure 13 shows the main user interface of the multiphase flow tool where input sections for “Flow line setup”, “Black oil setup”, “Heat transfer setup” and options for correlations can be found. Values printed in blue are set as default values. Correlations highlighted in red are the only correlations available for the specific physical property.

The small table (also found in Figure 13) beyond the input spreadsheet is not linked with the required input data. It is a listing of correlations available in the program.



Multiphase Flow Calculation Tool[1)]

Start Calculation

Input Data

Flowline Setup		
Inner Pipe Diamete	0,506	[m]
Pipe Roughness	1,829E-05	[m]
Vertical Distance	10	[m]
Horizontal Distanc	10000	[m]
Liquid Flowrate	1,590E+04	[sm³/d]
Rate of Undulation	0	[per 1000]
Direction of 1st Inc	up	
Elevation Profile	detailed	
Black Oil Setup		
Temperature	93,333	[°C]
Pressure	137,895	[bara]
GOR	267,161	[sm³/sm³]
Water Cut	0	[%]
Gas Specif. Gravity	0,75	[-]
Gas S. Grav. @ Tsep	15,556	[°C]
Gas S. Grav. @ Psep	7,908	[bara]
Water Sp. Gravity	1	[-]
API	30	[-]

Heat Transfer Setup		
Internal Fluid		
Thermal Conduct. Oil	0,138	[W/m/K]
Thermal Conduct. Water	0,605	[W/m/K]
Thermal Conduct. Gas	0,035	[W/m/K]
Specif. Heat Capacity Oil	1,890	[kJ/kg/K]
Specif. Heat Cap. Water	4,300	[kJ/kg/K]
Specif. Heat Cap. Gas	2,310	[kJ/kg/K]
Pipe		
Thermal Conductivity	60	[W/m/K]
Wall Thickness	10	[mm]
Ambient Fluid		
Temperature	25,0	[°C]
Velocity	2,0	[m/s]
Viscosity	2,000E-05	[Pa*s]
Density	1,121	[kg/m³]
Thermal Conductivity	0,024	[W/m/K]
Heat Capacity	1,005	[kJ/kg/K]

Engine Control	
Increment Counter:	
Undulation Counter:	
Section Counter:	
Error Control Panel	
Correlation Selection	
Solution Gas Ratio	V&B
Oil Form. Vol. Factor	V&B
Dead Oil Viscosity	B&R
Live Oil Viscosity	B&R
Undersat. Viscosity	V&B
Gas Viscosity	Lee et al.
Slug Length	Scott

GOR Correlations	Pb & Form. Vol. Factor Corr.	Dead Oil Viscosity	Slug Length
Standing	Standing	B&R	Scott
Lasater	V&B	Beal	Norris
V&B	Glaso		Brill et al.
Glaso	Al-Mar.	**Increment Direction**	**Elevation Profile**
Al-Mar.	Kart&Schm	up	simple
Kart.		down	detailed

[1)] Based on Xiao Mechanistic Model

Figure 13: User interface of the multiphase flow tool.

5.2.1.1. Temperature loop

The temperature loop is the core loop of the multiphase flow tool and it is used to determine the temperature at the end of a considered pipe increment. First the temperature at the increment end (T2) is assumed (for example: T2 = T1) to determine the heat transfer coefficients (note that the ambient heat transfer coefficient also needs an initial guess of the heat flow) and the enthalpy flow at the increment end (refer to section 3.3.2). An enthalpy balance for the pipe increment can then be solved according to Equation 3-10. If the left hand side and right hand side of Equation 3-10 do not differ within a limit range the estimated temperature is selected to calculate Black Oil (BO) properties at the increment end. If the estimated temperature is too low (or too high), the program loop increases (or decreases) the estimated temperature until convergence is reached within the defined ranges. Note that the calculation of all required parameters (for example heat capacities, mass flow rates and densities) to determine the enthalpy flow at the increment end are based on the first estimation of T2. Therefore, the temperature loop is run through again with the predicted Black Oil parameters for point two. Afterwards the flow pattern (FP) and the pressure gradient (PG) for the increment end are calculated. Figure 14 summarizes this procedure.

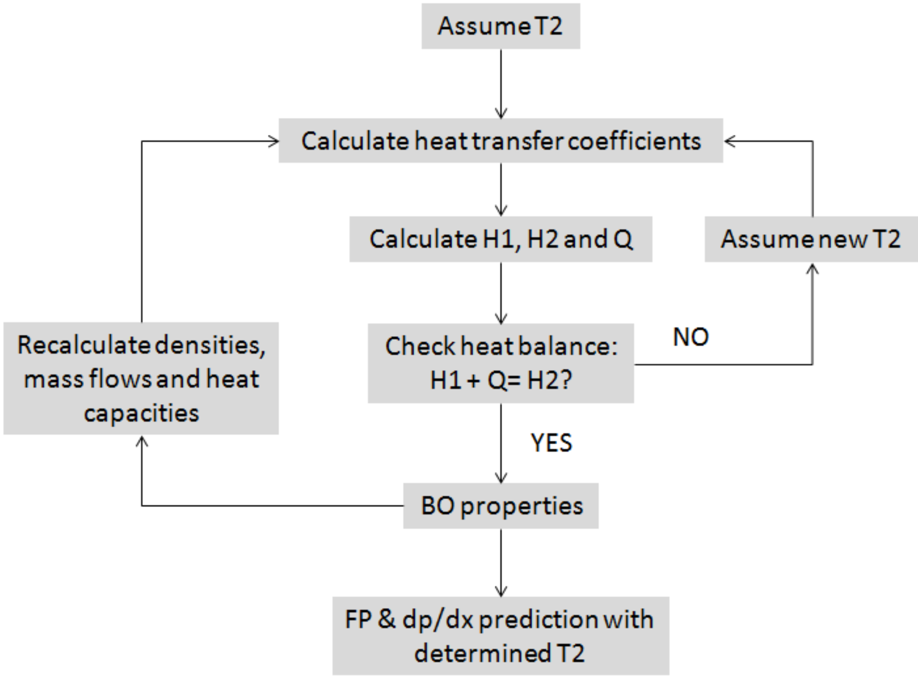


Figure 14: Main calculation procedure of the temperature loop.

5.2.1.2. Acceleration pressure gradient loop

The determination of the pressure gradient due to acceleration requires both the velocities at the increment entrance and at the end. To calculate the velocities at the increment exit it is necessary to know the liquid hold up. It requires a flow pattern prediction for the end. An iterative solution procedure is developed in which the acceleration pressure gradient in a considered pipe increment is neglected in a first guess. The pressure at the increment end can then be obtained with the pressure gradient at the pipe entrance. One can then solve for the liquid hold up at the pipe increment end and the velocities. Since the velocities at the increment start have already been calculated before the acceleration gradient can now be obtained. If the acceleration pressure gradient is exceeding a certain limit it is added to the calculated pressure gradient and the entire procedure is repeated one more time (restricted by variable “acceleration count”). Figure 15 indicates the main calculation steps in the acceleration pressure gradient loop.

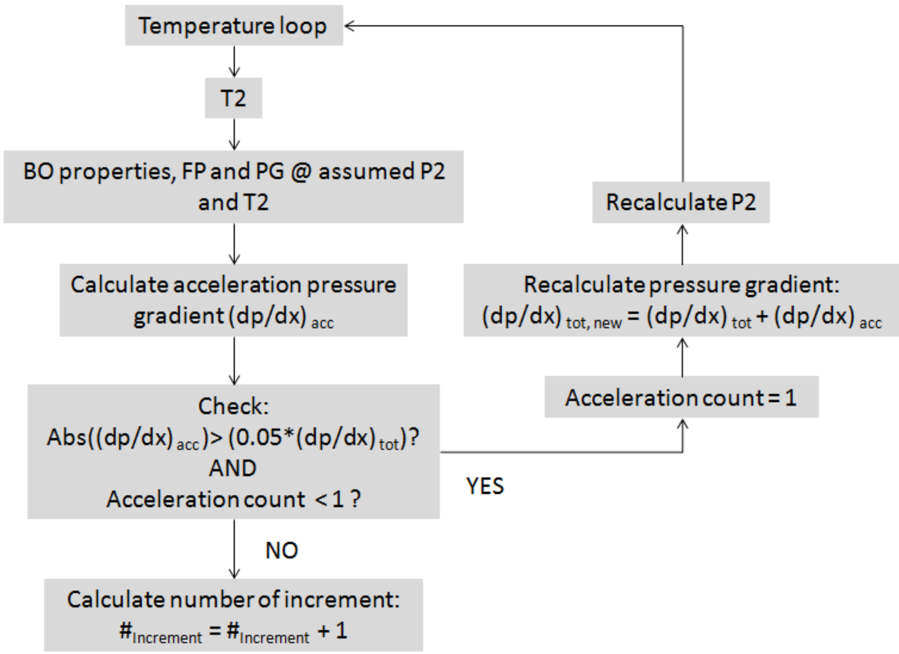


Figure 15: Main calculation procedure in the acceleration pressure gradient loop.

5.2.1.3. *Increment loop*

Afterwards the Black Oil properties, the flow pattern and the pressure gradient are recalculated with T2 and P2. The determined properties are now transferred to the start of the following increment. This variable transfer is the main purpose of the "Increment Loop". The number of increments is determined on the basis of the total horizontal distance as shown in Table 3.

Table 3: Default increment lengths depending on the horizontal distance.

Horizontal Distance [m]	Increment Length [m]
<= 1	0,1
<= 10	1
<= 500	2
<= 1000	4
<= 2000	8
<= 5000	16
<= 10000	32
> 10000	64

The increment loop is needed to determine all required parameters at the increment start and at the end. That's why the temperature as well as the acceleration gradient pressure loop are key tools within the increment loop. A summary of the main procedure is given in Figure 16.

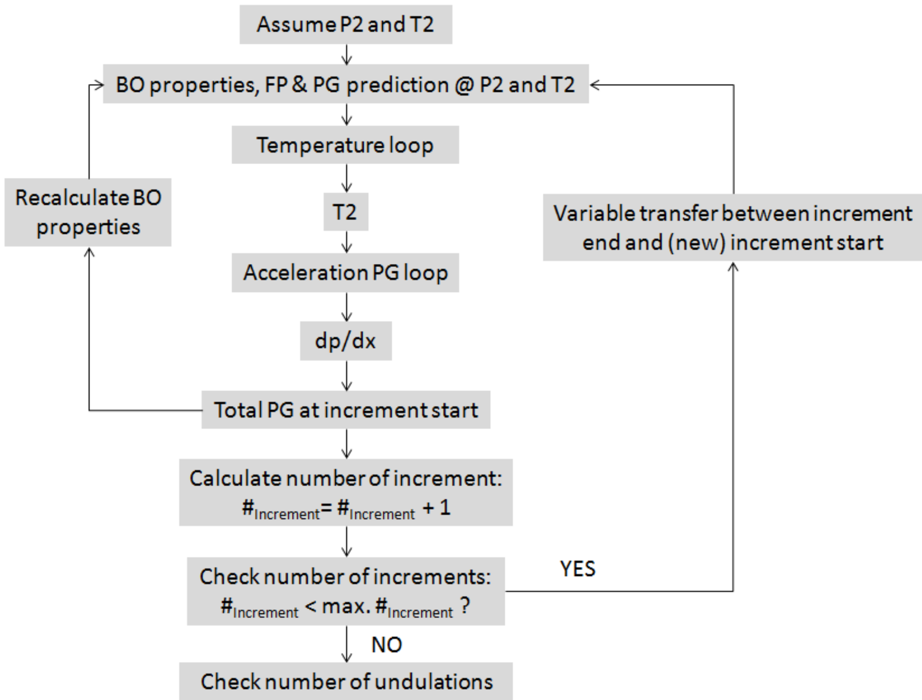


Figure 16: Calculation steps of the increment loop.

5.2.1.4. Undulation loop

A "rate of undulations" can be defined by the user to account for a simplified elevation profile if detailed data is not available. The entered value is equal to a total change in elevation for every 1000 units. A rate of 10, for example, on a flow line of 1000 m would be equal to an elevation of 5 m after a distance of 500 m.

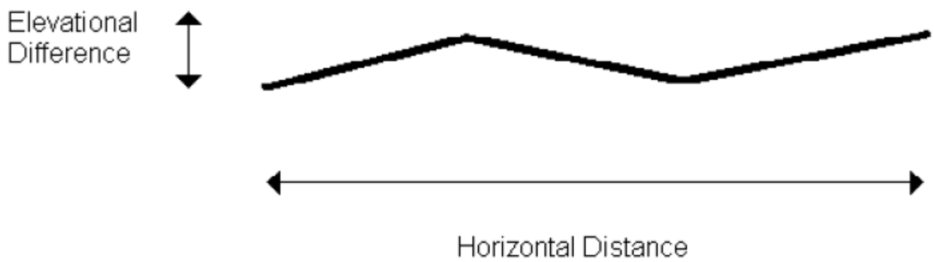


Figure 17: Undulations for a simplified elevation profile (Schlumberger, 2010).

This loop becomes active as soon as a rate of undulation greater than zero is specified. Even though a detailed elevation profile is available a rate of undulation can still be entered but results in longer calculation times.

A considered flow line section (the term flow line section refers to a segment of the flow line with constant inclination) is first divided into a certain number of increments according to Table 3. If a rate of undulations greater than zero is defined, each of the increments is inclined either up or down. The direction of the first increment can be defined by the user on the input sheet. Since each undulated increment needs to be solved individually it is necessary to subdivide them into smaller increments. Solving of each inclined increment individually is done by using the pipe increment loop again. A variable called "undulation-counter" is responsible for counting the solved inclined increments which were originally defined on basis of the flow line section horizontal length. This procedure is depicted in Figure 18.

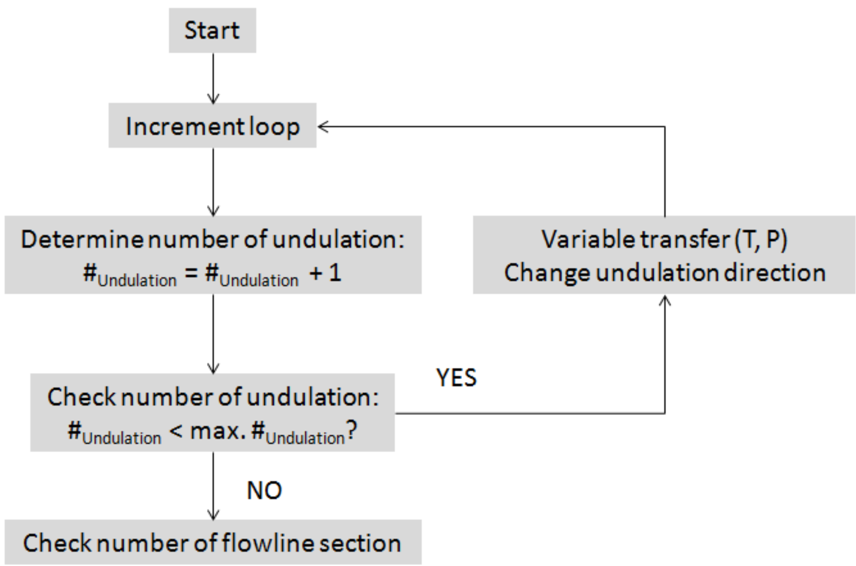


Figure 18: Increment loop with main calculation steps.

5.2.1.5. *Flow line section loop*

The flow line section loop is the covering loop and is especially designed for the case where a detailed elevation profile is known. This loop is only activated when the user selects "detailed elevation profile". The required data for the elevation profile is entered in the "elevation-profile" sheet which is limited to twelve sections. Each section is then divided into increments which can be inclined in a certain angle (according to the rate of undulations). After completing a flow line section the program transfers the parameters at the end of the section to the next one until all of the defined sections are completed. All of the previously introduced loops are subjected to the flow line section loop.

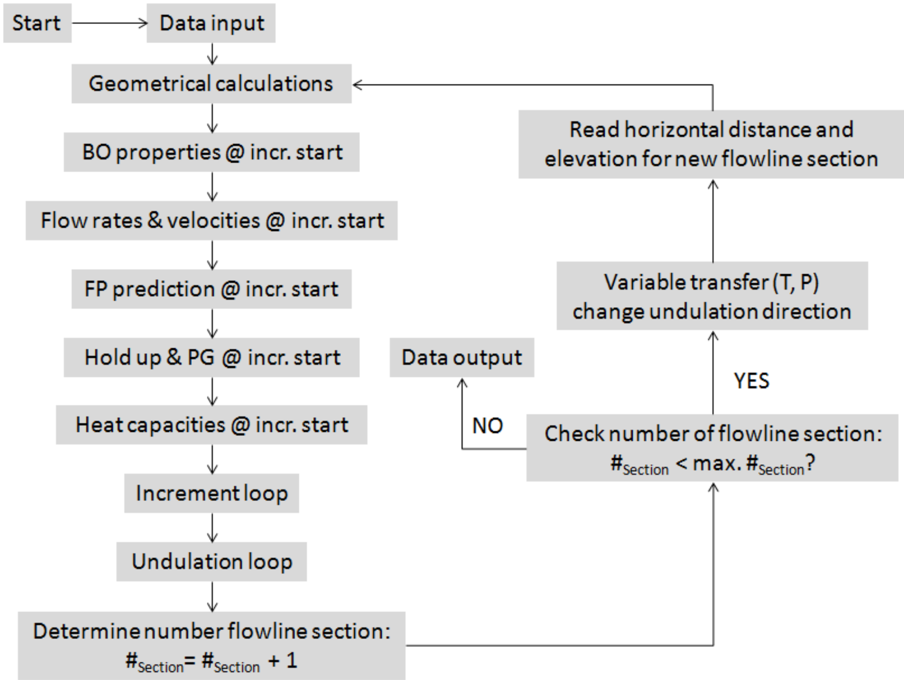


Figure 19: Main calculation procedure in the flow line section loop.

After completing the last section a PI-SS Number (ref. to section 3.4.1.4) is calculated and the data output is produced.

5.2.2. Calculation results and tool validation

On the basis of the problem explained in section 5.1 multiphase flow calculations were performed with the developed tool and compared with the results obtained in PipeSim (Xiao mechanistic model). At first, simple profiles (no undulations) of different lengths and inclinations were calculated. The results of the simple profiles can be found in the Appendix 2.

Afterwards the same problem with an elevation profile as given in Figure 20 was solved. The exit temperature and pressure of the flow line are calculated with the multiphase tool and with PipeSim (Xiao mechanistic model).

Multiphase Flow Calculation Tool

Elevation Profile

	Distance [m]	Elevation [m]	amb. Temp. [°C]	Pressure [bara] (Output)
Section1	0	0		137,90
	1500	10	25	136,97
Section2	1500	-10		
	3000	0	25	136,46
Section3	1500	10		
	4500	10	25	135,53
Section4	900	-10		
	5400	0	25	135,51
Section5	1100	15		
	6500	15	25	134,38
Section6	200	-10		
	6700	5	25	134,49
Section7	4900	0		
	11600	5	25	132,31
Section8	900	15		
	12500	20	25	131,04
Section9	200	-15		
	12700	5	25	131,21
Section10	700	25		
	13400	30	25	129,46
Section11	800	-30		
	14200	0	25	129,74
Section12	1300	25		
	15500	25	25	127,99

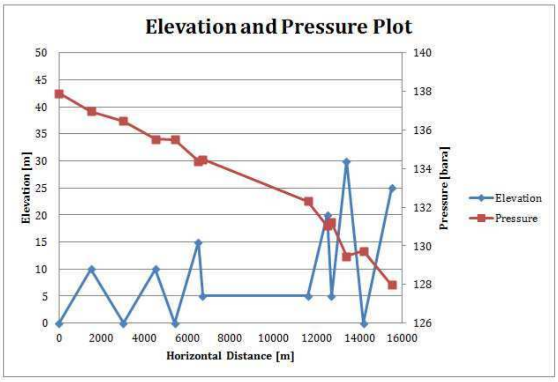


Figure 20: Elevation profile and predicted pressure plot for an 15,5 km transfer line (Abu Dhabi).

Table 4 presents the calculated exit pressure and temperature for the considered flow line.

Table 4: Results for the multiphase flow problem for exit pressure and temperature.

Parameter	Multiphase flow tool	PipeSim (Xiao model)
Flow line exit P. [bar]	128,0	129,7
Flow line exit T. [°C]	58,7	57,3

5.3. Correlation comparison

In order to demonstrate that the correlation selection is a crucial step in multiphase flow calculations parameters such as outlet pressure and pressure drop respectively, the liquid hold up and the mean slug length at the end of a multiphase flow line are determined with different correlations. The performance of multiphase flow correlations is examined for a horizontal case only.

The outlet pressure calculation was performed with various correlations in PipeSim. Table 5 shows the selected correlations and the abbreviations which also appear in the figures below. Note that the same correlation can appear with different data sources. “BJA” indicates that the data is based on research of Schlumberger (1987) while “TULSA” indicates that data were obtained from experiments carried out by TULSA University.

Table 5: Sources and abbreviations of horizontal multiphase flow correlations.

Correlation	Source	Abbreviation (acc. to PipeSim)
Beggs & Brill (Original)	BJA	BBO
Beggs & Brill (Revised)	BJA	BBR
Beggs & Brill (Original), Taitel & Dukler map	BJA	BBOTD
Beggs & Brill (Revised), Taitel & Dukler map	BJA	BBRTD
Baker Jardine (Revised)	BJA	BJA
Dukler, AGA & Flanagan	BJA	DKAGAF
Dukler, AGA & Flanagan (Eaton Hold up)	BJA	DKAGAD
Mukherjee & Brill	BJA	MB
No Slip Assumption	BJA	NOSLIP
Oliemans	BJA	OLIEMANS
Xiao	BJA	XIAO
TUFFP Unified 2-phase v2007.1	BJA	TU2P
Beggs & Brill	TULSA	TBB
Dukler	TULSA	TDUK
Mukherjee & Brill	TULSA	TMB
OLGA-S 2000 V5.3. 2-Phase	OLGA5.3.2	olga2pb
Segregated Flow - GRE Mechanistic Model BP	BPD	BP1

5.3.1. Comparison study in PipeSim

Table 6 summarizes the basic flow line data for the comparison case performed in PipeSim. A simplified elevation profile specified by a rate of undulations is assumed for this case in order to guarantee slug flow.

Table 6: Flow line data for the basic horizontal case.

Horizontal flow line data	
Liquid flow rate	$1{,}6 * 10^4$ Sm^3/d
Gas gravity @ P_{sep} = 8 bara and T_{sep} = 15°C	0,75
Water specific gravity	1
API gravity	30°API
Produced gas-oil ratio	267 Sm^3/Sm^3
Water cut	0 %
Pipe diameter	0,5 m
Pipe roughness	0,0183 mm
Rate of undulations	10
Distance	10000 m
Inlet pressure	137,9 bar
Inlet temperature	93 °C
Pipe wall thickness	12,7 mm

5.3.2. Calculation Results

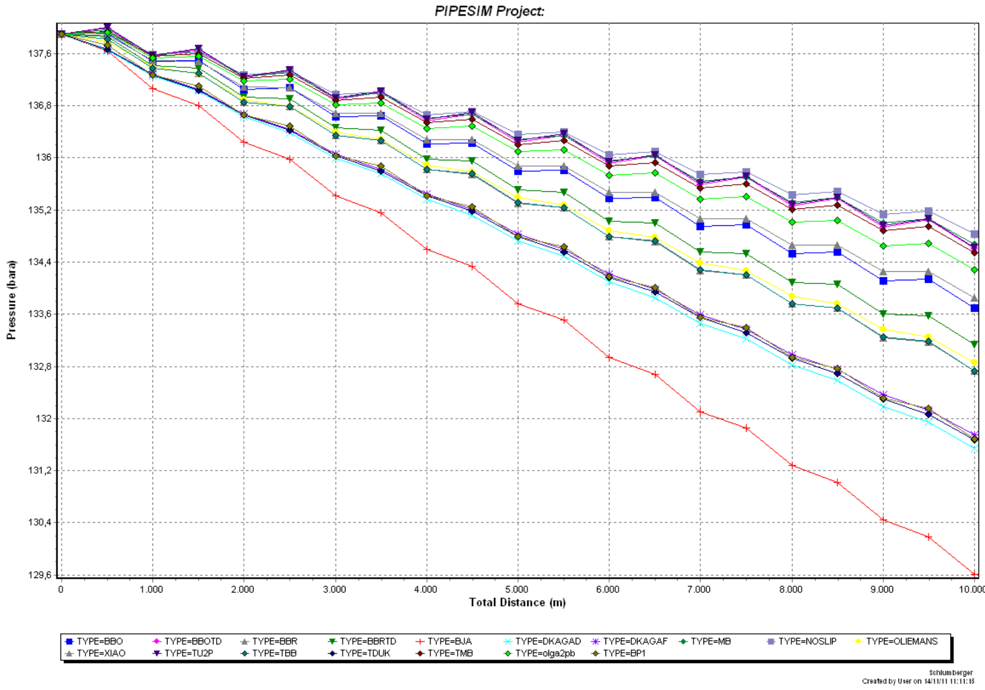


Figure 21: Outlet pressure calculation with different correlations.

As can be seen from Figure 21 the outlet pressure varies widely depending on the selected multiphase flow correlation. The highest outlet pressure is determined with NOSLIP (134,8 bar, Δp = 3,1 bar) while the BJA-correlation calculates the lowest outlet pressure (129,6 bar, Δp = 8,3 bar). Among pipeline engineers OLGA-S (134,3 bar, Δp = 3,6 bar) is known to give the most satisfying results. Correlations such as TMB (134,5 bar, Δp = 3,4 bar), TU2P (134,6 bar, Δp = 3,3 bar), BBOTD (134,6 bar, Δp = 3,3 bar) and MB (134,7 bar, Δp = 3,2 bar) are close to OLGA-S. The effect of undulations on the pressure gradient is also reflected in Figure 21.

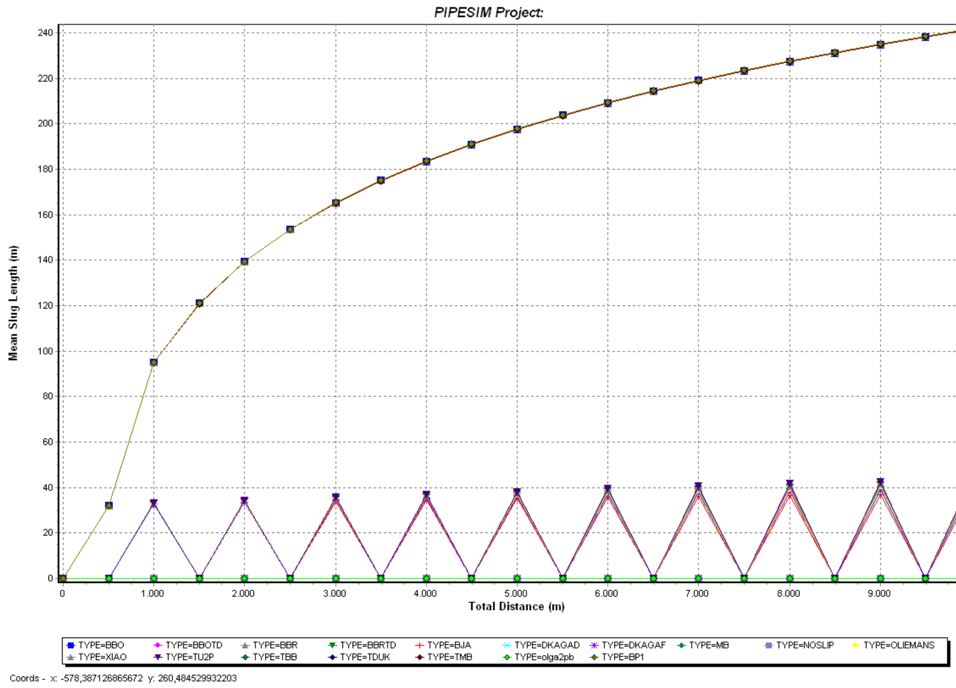


Figure 22: Mean slug length with different correlations.

According to Figure 22 multiphase flow correlations can differ significantly in the prediction of slugs. There are some correlations which suggest hydrodynamic slugging (BP1, BBR, TBB, BBO and TMB) and there are some correlations which predict only terrain induced slugging (TU2P, BBOTD, XIAO, BBRTD, OLIEMANS, DKAGAF, DKAGAD, BJA). And, there are correlations (OLGA-S, TDUK, NOSLIP, MB) which suggest that slugs do not occur. Hydrodynamic slugs are uniformly predicted with 240 m. However, the mean lengths of hydrodynamic slugs can also vary in other cases and may not be uniformly predicted as in this case. The mean length of terrain slugs is constantly increasing and reaches a length of 43,5 m (TU2P, BBOTD) or 37 m (BJA) respectively.

As indicated in Table 7 there is a broad variety in terms of liquid hold up. After 10.000 m the predicted hold ups range from 0,45 (NOSLIP) to 0,63 (XIAO). BJA predicts a hold up of 0,60 which comes close to OLGA-S (0,61). As far as empirical models are concerned the liquid hold up is predicted with different correlations. Therefore, it can be expected that the models suggest different values for the liquid hold up. Furthermore, the deviations can also be explained due to differences in the predicted flow patterns.

Table 7: Summary of calculated pressure drop, mean slug length and liquid hold up by various horizontal flow correlations.

Correlation	Flow pattern[1])	Pressure drop [bar]	Mean slug length [m][1])	Liquid hold up
BBO	Intermittent	4,2	241,5	0,55
BBR	Intermittent	5,2	241,6	0,51
BBOTD	Intermittent	3,3	43,6	0,55
BBRTD	Intermittent	4,8	41,6	0,51
BJA	Intermittent	8,3	37,2	0,60
DKAGAD	Intermittent	6,4	39,5	0,47
DKAGAF	Intermittent	6,2	39,7	0,51
MB	Mist	3,2	0	0,52
NOSLIP	Gas	3,1	0	0,45
OLIEMANS	Intermittent	5	41,2	0,51
XIAO	Intermittent	4	42,5	0,63
TU2P	Intermittent	3,3	43,6	0,55
TBB	Intermittent	5,2	241,6	0,51
TDUK	2-Phase	6,2	0	0,52
TMB	Slug	3,3	241,3	0,52
OLGA-S	Strat. Wavy	3,6	0	0,61
BP1	Intermittent	6,2	241,8	0,55

[1]) @ pipeline end

Since the method to calculate the temperature profile is not dependent on the multiphase flow correlation the outlet temperature is uniformly predicted with 72°C.

5.3.3. Correlation selection

As shown in the previous section the calculated pressure loss may differ significantly depending on the selected multiphase flow correlation. Consequently, in real cases selection criteria to determine the most suitable multiphase flow correlation for the specific application are required.

Since every real case differs from others a comprehensive study would be necessary to recommend a suitable correlation for a particular case. Carrying out such a study is beyond the scope of this thesis. Therefore, this section provides a correlation decision tree (Figure 23) which is created according to the PipeSim user guide. The suggested correlations found in the PipeSim user guide are based on the experience of Schlumberger.

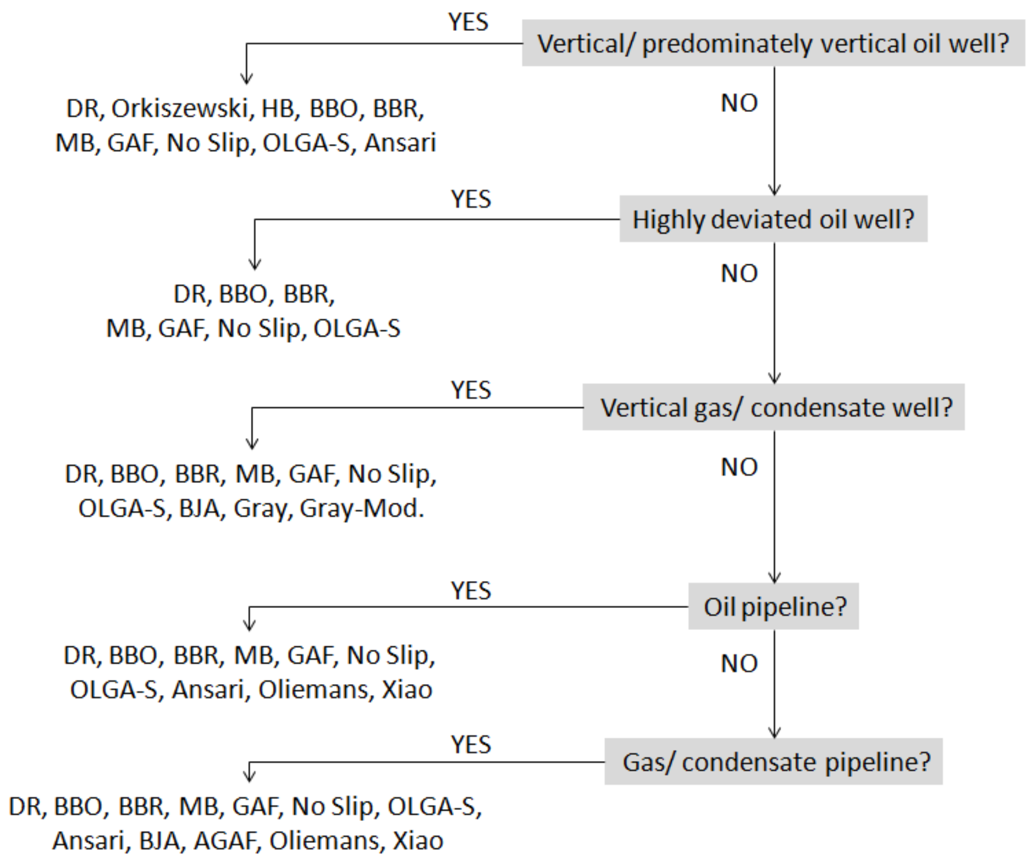


Figure 23: Decision tree for multiphase flow correlations based on PipeSim user guide.

According to PipeSim user guide measured data is to be used if available to find an appropriate correlation for the particular application. If not possible one can proceed as suggested by the decision tree. There might be some cases where more than one correlation leads to satisfying results. Depending on the availability of multiphase flow correlations one can also select a suitable method from Table 8 which gives further recommendations based on the PipeSim user guide.

Table 8: Suggested correlations for different multiphase flow cases acc. to PipeSim user guide.

Correlation	**Vertical and Predominately Vertical Oil Wells**	**Highly Deviated Oil Wells**	**Vertical Gas/ Condensate Wells**	**Oil Pipelines**	**Gas/ Condensate Pipelines**
Duns and Ros	yes	yes	yes	yes	yes
Orkiszewski	yes	no	yes	no	no
Hagedorn and Brown	yes	no	yes	no	no
Beggs & Brill (Revised)	yes	yes	yes	yes	yes
Beggs & Brill (Original)	yes	yes	yes	yes	yes
Mukherjee & Brill	yes	yes	yes	yes	yes
Govier, Aziz & Forgasi	yes	yes	yes	yes	yes
No Slip	yes	yes	yes	yes	yes
OLGA-S	yes	yes	yes	yes	yes
Ansari	yes	no	yes	no	no
BJA for Condensates	no	no	yes	no	yes
AGA & Flanigan	no	no	no	no	yes
Oliemans	no	no	no	yes	yes
Gray	no	no	yes	no	no
Gray Modified	no	no	yes	no	no
Xiao	no	no	no	yes	yes

5.4. Sensitivity study

After selecting an appropriate correlation it is recommended to analyse the system performance for varying operation conditions in order to identify sensitive parameters which can play a role in further considerations, for example, slug catcher design. A horizontal example case is presented in this section. In a first step parameters for the sensitivity study are selected. Afterwards, the influence of these parameters on the multiphase flow behaviour is examined and illustrated in diagrams.

As mentioned earlier OLGA-S is believed to give the most satisfying results in multiphase flow calculations. However, XIAOs model is selected for performing the sensitivity analysis since OLGA-S does not predict slug flow for this particular case (ref. to Figure 22).

The pipe roughness, pipe diameter, oil flow rate, gas flow rate, API gravity and the gas gravity are selected as sensitivity variables. The influence of the selected sensitivity variables are examined on system parameters such as pessure loss and mean slug length.

5.4.1. Sensitivity study in PipeSim

The sensitivity study is carried out with the same input data as specified in Table 6. Table 9 illustrates the range and increments of the selected sensitivity variables.

Table 9: Selected sensitivity variables with their range and increments.

Sensitivity variable	Range	Increment
Pipe diameter	300 - 1500 mm	100 mm
Pipe roughness	$1*10^{-6}$ m - $1*10^{-5}$ m	$5*10^{-6}$ m
	$1*10^{-5}$ m - $1*10^{-4}$ m	$2*10^{-5}$ m
	$1*10^{-4}$ m - $1*10^{-3}$ m	$1*10^{-4}$ m
	$1*10^{-3}$ m - $1*10^{-2}$ m	$1*10^{-3}$ m
Gas flow rate	8000 - 28000 Sm^3/d	2000 Sm^3/d
Oil flow rate	6000 - 24000 Sm^3/d	2000 Sm^3/d
Gas gravity	0,6 - 1	0,1
API gravity	10-60	4

5.4.2. Calculation results

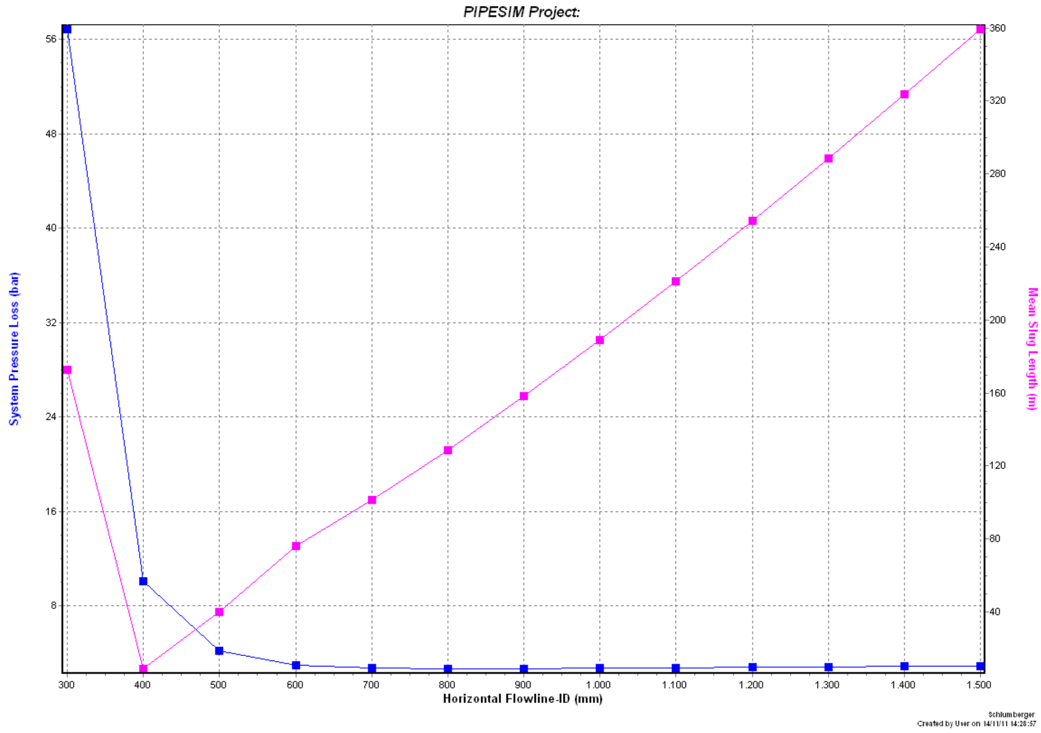


Figure 24: Pressure loss and mean slug length as a function of inner pipe diameter.

Figure 24 shows that the pressure loss decreases rapidly until a pipe diameter of 400 mm is exceeded while for pipe diameters larger than 600 mm the pressure drop remains constant. The rapid decrease in pressure loss between 300 mm and 400 mm can be explained due to a transition in flow pattern. While for small diameters (300 mm) the “Intermittent” flow pattern is predominant throughout the entire pipeline for diameters larger than 400 mm the flow pattern changes from “Stratified Wavy” at the pipeline inlet into “Intermittent” at the outlet. Consequently, if the flow pattern is conserved over the entire length of the flow line the pressure loss is significantly influenced by varying the pipe diameters. If the flow pattern changes in the flow line the effect of increasing diameters on the pressure drop is negligible in this case.

The mean slug length is also affected by the flow pattern transition as indicated by the decreasing slug length for pipe diameters smaller than 400 mm. If pipe diameters are larger than 400 mm the flow regime changes along the flow line from “stratified wavy” into “intermittent”. Therefore, slugs are generated which constantly grow when the pipe diameter is increased.

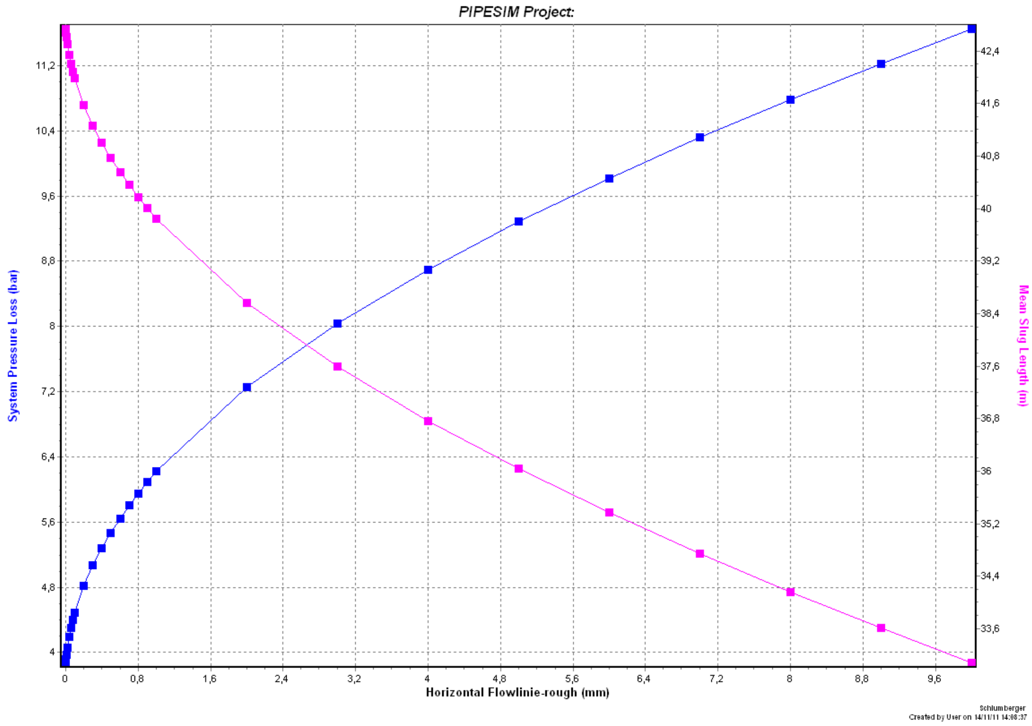


Figure 25: Pressure loss and mean slug length as a function of pipe roughness.

The influence of the increasing pipe roughness on the pressure loss and mean slug length is illustrated in Figure 25. While the mean slug length is decreasing the system pressure loss is increasing. The biggest change in pressure loss and mean slug length for a pipe roughness is beyond 0,8 mm. At a pipe roughness above 0,8 mm the pressure loss increases and the mean slug length decreases at a constant rate. Consequently, the impact of increasing pipe roughness on both the mean slug length and the system pressure loss is considerable.

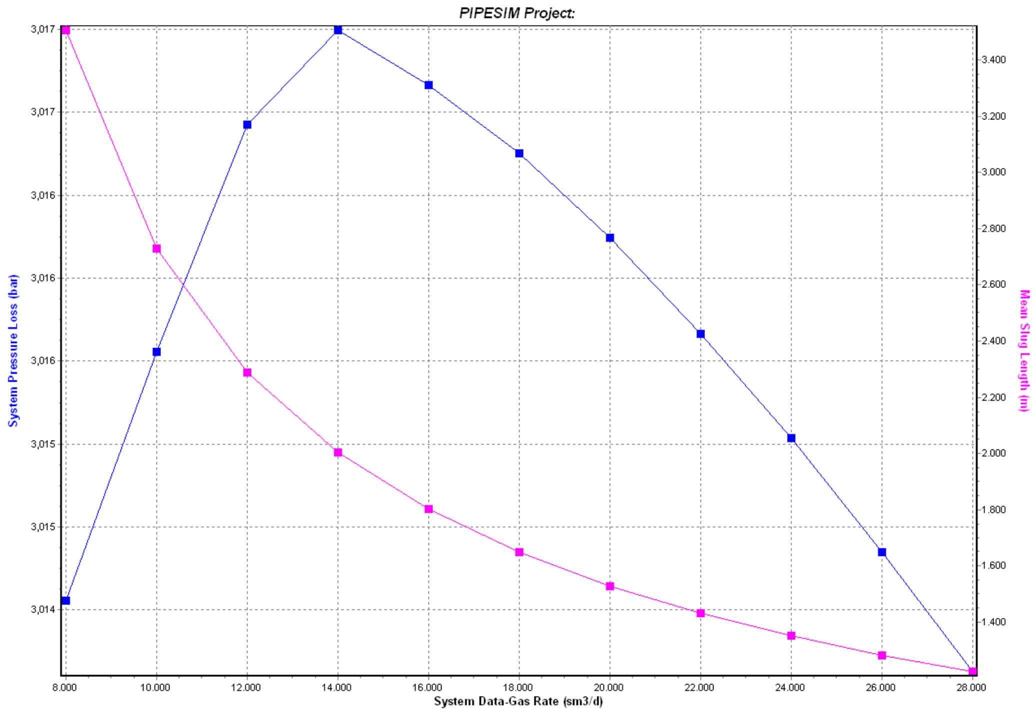


Figure 26: Pressure loss and mean slug length as a function of gas flow rate.

From Figure 26 it can be concluded that the influence of rising gas flow rate on the pressure loss is negligible. On the contrary, the mean slug length is significantly dependent on the increasing gas flow rate. Especially increases at low gas flow rates (for example from 8.000 sm^3/d to 10.000 sm^3/d) drastically enlarge the predicted slug length (Δl = 800 m). Low gas flow rates (8.000 sm^3/d) produce slugs of about 3500 m while high gas flow rates (28.000 sm^3/d) reduce the slug length to 1200 m. Apparently, the gas flow rate has a major influence on the slug length and therefore plays an important role in slug catcher design.

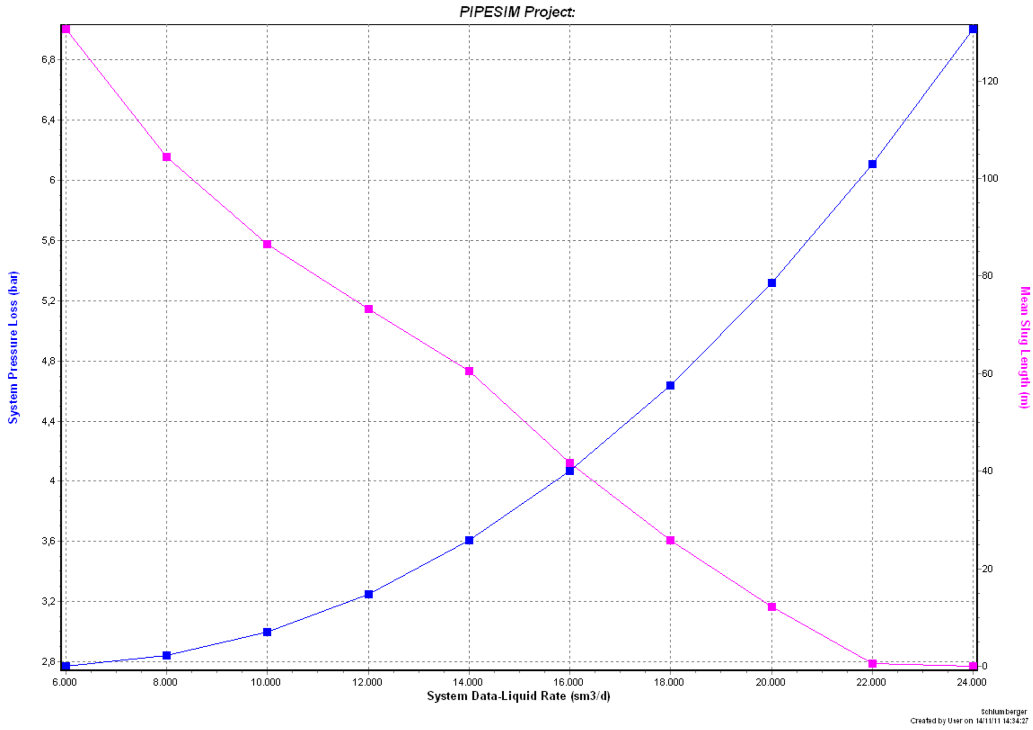


Figure 27: Pressure loss and mean slug length as a function of liquid flow rate.

It can be seen from Figure 27 that a rising liquid flow rate increases the system pressure drop. While at flow rates beyond 18.000 sm^3/d the effect on the pressure drop is negligible changes in flow rates above 20.000 sm^3/d can moderately affect the system pressure loss. On the contrary, the mean slug length is considerably dependent on the liquid flow rates up to 22.000 sm^3/d. If, for example, the liquid flow rate is increased from 6.000 sm^3/d to 8.000 sm^3/d the predicted slug length differs as much as almost 30 m. If the flow rate is increased from 18.000 sm^3/d to 20.000 sm^3/d the difference in slug length is still more than 10 m which can affect the slug catcher design considerably.

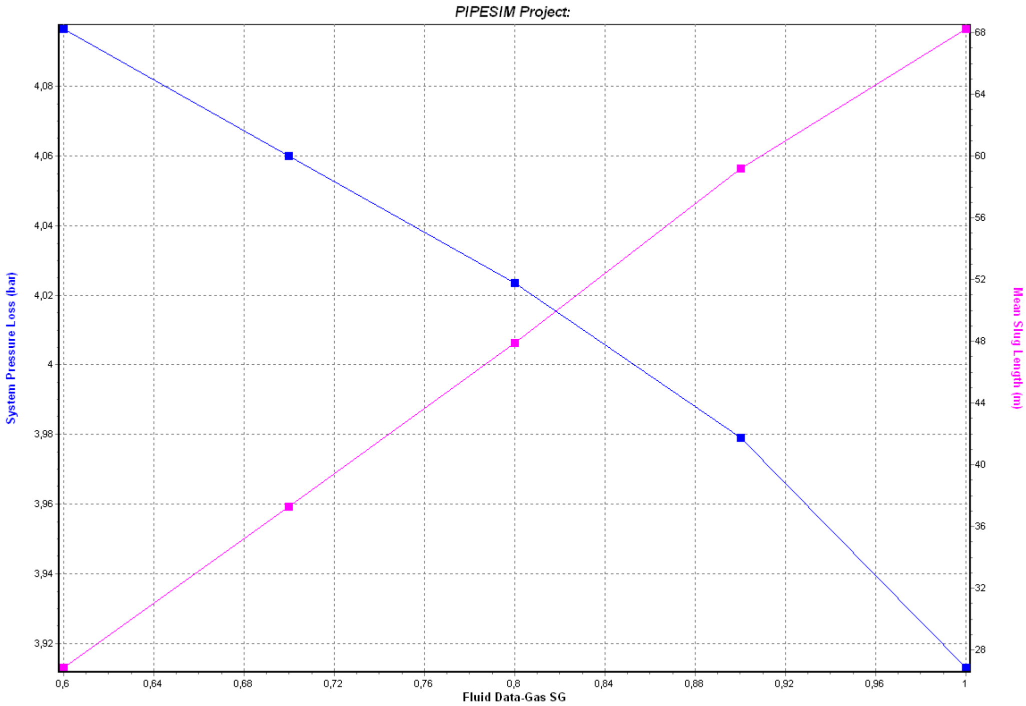


Figure 28: Pressure loss and mean slug length as a function of gas specific gravity.

As illustrated in Figure 28 the effect of specific gas gravity on the system pressure loss in a horizontal pipeline is negligible. This can be explained by the fact that the pressure loss remains almost constant over the considered range. Slugs are constantly getting longer as the specific gas gravity is increased. Different gas gravities influence both the mean slug length and the pressure loss considerably. A specific gravity of gas equal to 0,6 results in a slug length of less than 30 m while slugs with about 70 m are resulted by a gas gravity of 1. Therefore, the influence of gas gravity on the slug length should not be neglected.

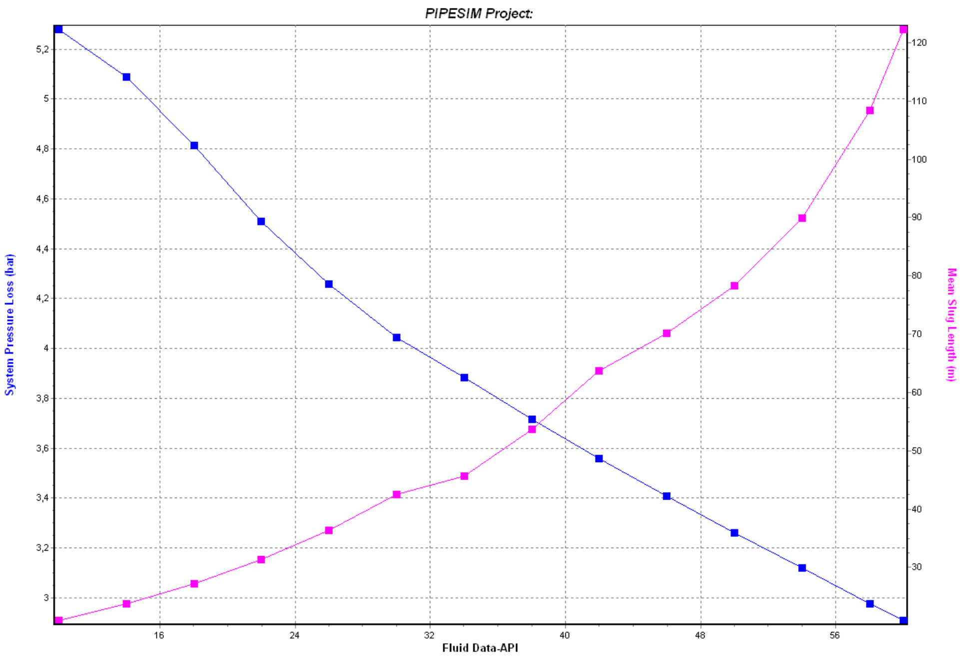


Figure 29: Pressure loss and mean slug length as a function of API oil gravity.

As can be seen in Figure 29 changes in the API gravity have a considerable influence on the mean slug length. While low API oils (API = 10°) generate slugs of approximately 20 m slugs in this particular case mean lengths of more than 120 m are predicted for oils with high API gravities (API = 60°). It can be seen that the largest growth in slug length occurs when the system contains oils lighter than 50° API. Figure 29 indicates that the pressure loss in the pipeline is constantly reduced at API gravities above 30°. Oils with an API gravity around 60° produce a pressure loss 4 bar lower than those of heavy oils (API around 10°) in this case. Therefore, the API gravity can have a considerable influence on pressure loss.

According to Table 10 only the pipe diameter and the gas flow rate have a significant impact on the mean slug length. Regarding the pressure loss only the pipe diameter seems to have a significant influence.

Table 10: Summary of the influence regarding the examined variables on the system parameters “mean slug length” and “pressure loss” for a horizontal 10000m pipeline.

Sensitivity variable	Pressure loss	Slug length
Pipe diameter	significant (D < 400 mm)	significant
Pipe roughness	considerable	considerable
Gas flow rate	negligible	significant
Oil flow rate	negligible	considerable
Gas gravity	negligible	considerable
API	considerable	considerable

6. Slug catcher design

The purpose of this section is to introduce theoretical (section 6.1) and design concepts (section 6.2) of slug catchers.

6.1. Theoretical concepts of slug catcher

Slug catchers are basically separation units installed at the end of a field line where stable flow conditions are required for subsequent facilities. Such separators cause a reduction in fluid velocities by enlarging the pipe diameter in order to promote a stratified flow regime. They have to be able to handle and absorb the largest slug volume. Thus, the determination of the appropriate slug catcher length and diameter are important steps in slug catcher design. Slug catchers are commonly classified into two categories:

6.1.1. Finger (or multiple-pipe) type slug catcher[11]

In case of finger type slug catchers a splitter can be found immediately behind the inlet which directs the two phase stream to the inlet header. The inlet header is connected to downcomers leading to primary bottles or "fingers" which consist of a separation part and a storage part. In order to reduce the flow velocity the finger diameters are usually larger than the diameters of the downcomers. While the liquid is settled in the fingers the gas is displaced through the gas risers to the gas outlet header. The liquid level in the fingers is reaching a maximum when the slug has completely entered the storage. The slug is then said to be "absorbed" by the finger storage. Therefore, the slug catcher has to provide sufficient volume for the absorption of the liquid slug. The separation of liquid from gas is driven by different mechanisms such as:

- Stratification: If the velocity is reduced stratification is likely to occur allowing the separation of liquid and gas.
- Droplet settling: Due to gravitation droplets can settle. It requires velocities less than 2 m/s.
- Tee-junction separation: Especially if a tee-junction is orientated in the upward direction, for example at the gas riser, efficient separation can be guaranteed.

[11] (SHELL, 1998)

6.1.2. Vessel type slug catcher

The vessel type slug catcher is basically a conventional vessel, simple in design and maintenance. It is preferably installed when liquid volume < 100 m^3 has to be stored and where space is limited (for example: offshore platform).

6.2. Design concepts of slug catcher

After having decided which type of slug catcher is most suitable for the particular case, slug catcher diameter and slug catcher length have to be calculated.

6.2.1. Finger (or multiple-pipe) Type Slug Catcher[12]

According to SHELL Design and Engineering Practice (1998) the following steps are recommended for sizing a finger-type slug catcher:

- Determine slug volume
- Choose finger diameter
- Choose number of fingers
- Determine length required for droplet separation
- Determine length required for slug storage

The slug volume includes the maximal possible slug volume predicted by a multiphase flow analysis as well as a buffer volume according to process requirements.

It is recommended to select a finger diameter > $3/2 * D_{Downcomer}$ resulting in further separation of the gas and liquid due to expansion. The downcomer should be installed at an angle of -45° allowing for the optimal development of stratified flow and the number of downcomers and fingers respectively per inlet manifold should be limited to eight. It is furthermore recommended that only an even number of fingers (2, 4, 6 or 8) is selected. Possible future extensions and a mal-distribution within the fingers (assume that the most heavily loaded finger receives 20% more than in case of an even distribution) should also be considered in deciding the finger number.

Since the gas liquid separation takes place in the first section of the finger upstream of the first gas riser this length depends on the desired separation efficiency and on the droplet size. To guarantee efficient separation the gas velocity at the finger entrance should not exceed 2 m/s. Appendix 2 in SHELL

[12] (SHELL, 1998)

Design and Engineering Practice (1998) includes a detailed procedure to determine a L/D ratio which is a required input parameter in the developed slug catcher tool.

Attention should also be paid to the slope of the fingers. Since stratified flow is desired the finger should slope downward. Otherwise liquid blockage due to choking is likely to occur. The Kelvin-Helmholtz criterion can therefore be used to check whether the selected slope is sufficient to maintain stratified flow. From a practical point of view it is recommended to choose a finger slope between 1% and 3%.

For further details on gas riser design also refer to SHELL Design and Engineering Practice (1998).

A basic scheme of a finger-type slug catcher is presented in Figure 30.

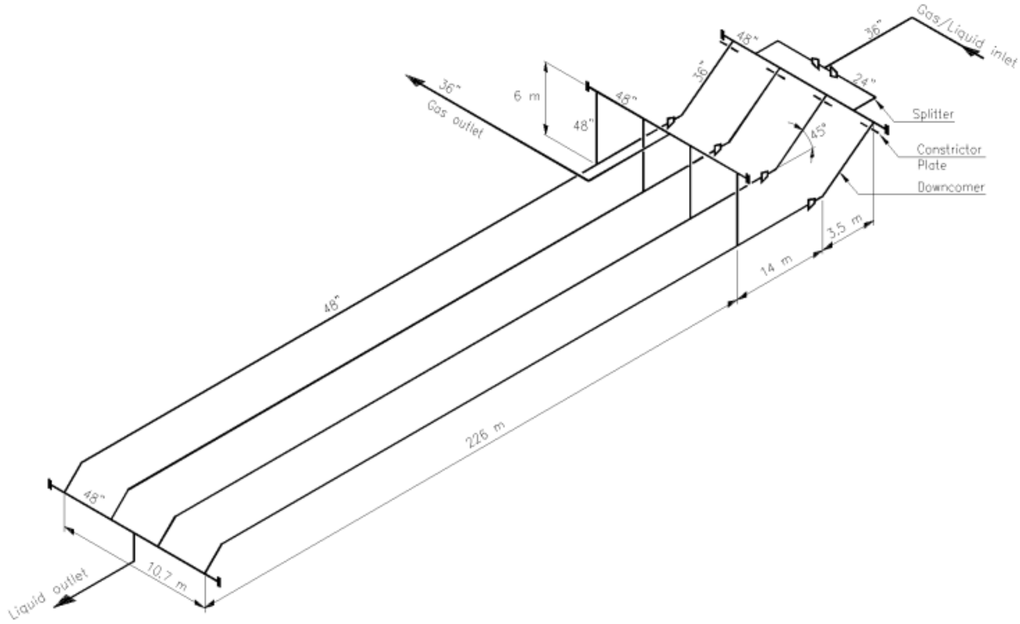


Figure 30: Scheme of finger-type slug catcher (Nogat slug catcher at Den Helder, from SHELL Design and Engineering Practice (1998)).

6.2.2. Vessel Type Slug Catcher[13]

In this section design concepts for a horizontal knock-out vessel are considered only. If design concepts for other vessel type slug catchers (for example vertical knock-out vessel) are required refer to SHELL Design and Engineering Practice (2002). Recommendations for the appropriate type selection can also be found there.

[13] (SHELL, 2002)

The diameter and the tangent-to-tangent length of a horizontal knock-out drum have to be defined in an iterative procedure which is explained in detail in APPENDIX VII of SHELL Design and Engineering Practice (2002). As stated there the tangent-to-tangent length/diameter ratio should be between 2,5 and 6 meters.

The iterative procedure also employs the following two criteria which have to be met to find an appropriate vessel diameter:

De-gassing criterion:

$$D \geq \left\{4.5 \cdot 10^{7} Q_{L,max} \eta_L / (\rho_L - \rho_G)\right\} / L \qquad \text{Equation 6-1}$$

De-foaming criterion:

$$D \geq 7000\, Q_{L,max} \{\eta_L / (\rho_L - \rho_G)\}^{0.27} / L \qquad \text{Equation 6-2}$$

The minimum vessel cross-sectional area for gas flow $A_{G,min}$ can be calculated from

$$\lambda_{max} = Q^{*}_{max} / A_{G,min} = 0.07 \qquad \text{Equation 6-3}$$

with

$$Q^{*}_{max} = Q_{G,max} \sqrt{\rho_G / (\rho_G - \rho_L)} \qquad \text{Equation 6-4}$$

where λ_{max} = maximum allowable gas load factor in m/s; Q^{*}_{max} = highest value of volumetric gas load factor in m^3/s; $Q_{G,max}$ = highest envisaged gas flow rate in m^3/s and includes a margin which is typically between 15% and 50% to account for surging, uncertainties in basic data, etc.

Note that λ_{max} = 0.07 applies under moderate conditions only. For further details refer to SHELL Design and Engineering Practice (2002).

For further design instructions (for example nozzle sizing) also refer to SHELL Design and Engineering Practice (2002).

A sketch of a horizontal knock-out drum is found in Figure 31. Furthermore, a sketch showing the liquid level control in a gas/liquid separator can be found in the Appendix 1. This sketch might be required to understand the vessel type slug catcher routine introduced in section 6.3.2.

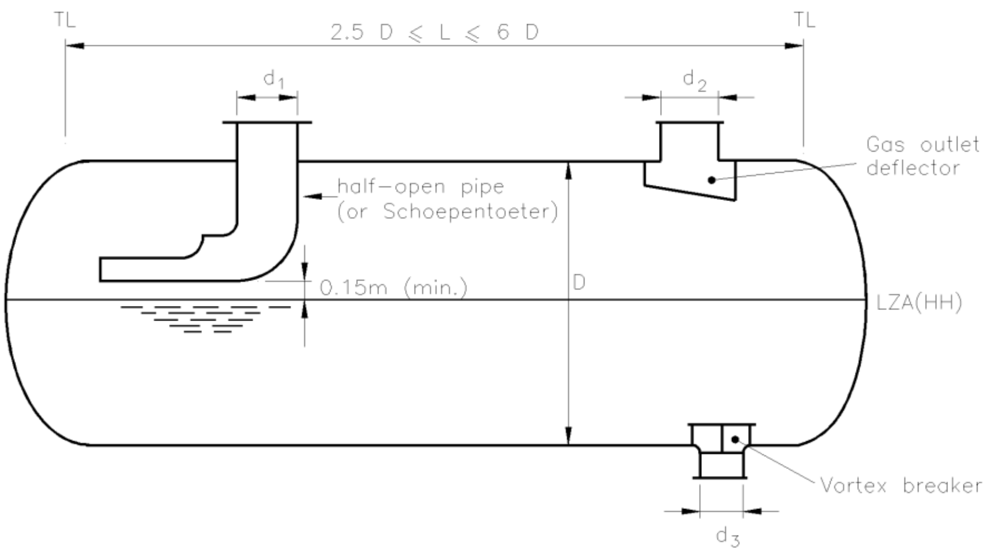


Figure 31: Horizontal knock-out drum from SHELL Design and Engineering Practice (2002).

6.2.3. Vessel type slug catcher according to Machado

The approach according to Machado (1977) is based on the Taitel-Dukler flow pattern model. If a multiphase flow analysis predicts intermittent flow at the end of the pipeline the diameter has to be increased until the flow pattern changes to stratified.

$$\mathrm{D_{min}} = \mathrm{H_{L,Transition}} / \tilde{\mathrm{h}}_{\mathrm{Transition}} \qquad \text{Equation 6-5}$$

where $H_{L,Transition}$ is the liquid hold up where the intermittent-stratified transition occurs.

$\tilde{h}_{Transition}$ is determined by the following criterion:

$$\mathrm{v_g} = \left(1 - \tilde{\mathrm{h}}_{\mathrm{Transition}}\right)\left[\frac{(\rho_L - \rho_g) g \cos\alpha A_g}{\rho_g S_i}\right]^{1/2} \qquad \text{Equation 6-6}$$

where v_g = gas velocity in m/s at the pipeline end; ρ_g and ρ_L = gas and liquid densities in kg/m^3 at the pipeline end; A_g and S_i = geometrical parameters and α = Inclination of slug catcher which is assumed = 0.

After selecting a diameter (based on the minimal diameter) the actual liquid hold up can be obtained from

$$\mathrm{H_L} = (\theta - \sin\theta)/(2 \cdot \pi) \qquad \text{Equation 6-7}$$

where

$$\theta = 2 \cdot \cos^{-1}\left(1 - 2 \cdot \tilde{\mathrm{h}}\right) \qquad \text{Equation 6-8}$$

with $\tilde{h}$ being determined from geometrical relationships as stated by Taitel and Dukler (1976).

Based on the selected diameter the transition (or critical) liquid hold up H_{LC} has to be recalculated from Equation 6-6. Note that v_g also changes since the selected diameter may differ from the minimum diameter.

After calculating the actual and the critical liquid hold up for the selected diameter the slug catcher length can be determined from

$$\mathrm{L_{SC}} = \mathrm{V_{Slug}}/\left(2 \cdot \mathrm{A_{SC}}(\mathrm{H_{LC}} - \mathrm{H_L})\right) \qquad \text{Equation 6-9}$$

where V_{Slug} = the expected slug volume (including buffer volume) in m^3 and A_{SC} is the cross-sectional area based on the selected diameter in m^2.

Figure 32 shows a slug absorption process and indicates actual and critical liquid hold up in a slug catcher.

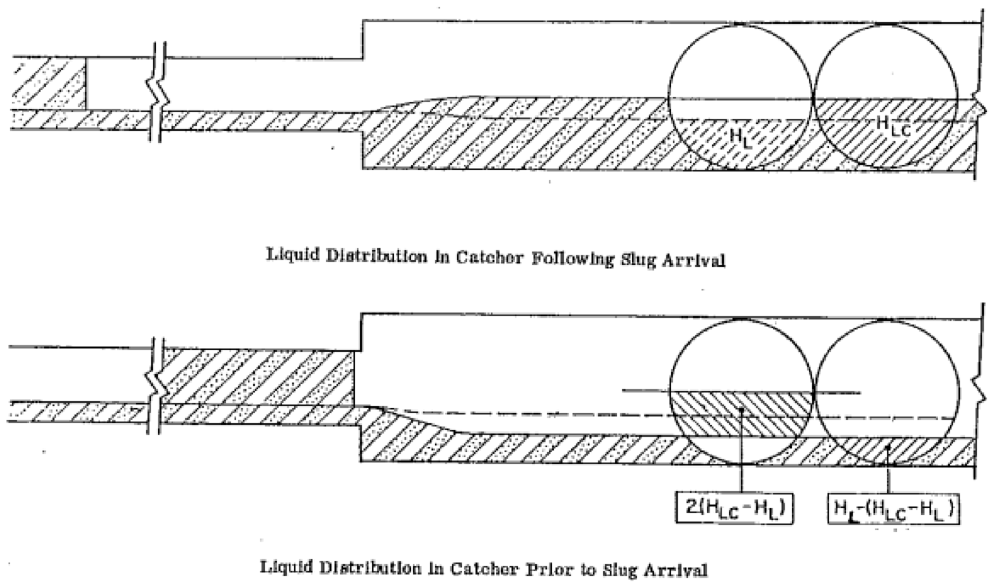


Figure 32: Slug absorption process in a slug catcher according to Machado (1977).

6.3. Configuration of the slug catcher tool

This section is meant to provide an overview of the developed slug catcher tool with respect to the configuration of the program. The program gives design recommendations for a finger-type slug catcher as well as a vessel type slug catcher. Note that two different approaches for sizing a vessel type slug catcher are incorporated in the program.

As can be seen from Figure 33 the user interface of the slug catcher tool is divided into three sections. In the upper part the user can specify fluid properties, flow rates and slug volume. If only the slug length is known the slug volume can also be estimated with the tool. In the middle section of the user interface design data concerning the particular slug catcher type have to be specified. Finally, the slug catcher sizes are produced and stored in bottom section after pressing the “calculate” bottom.

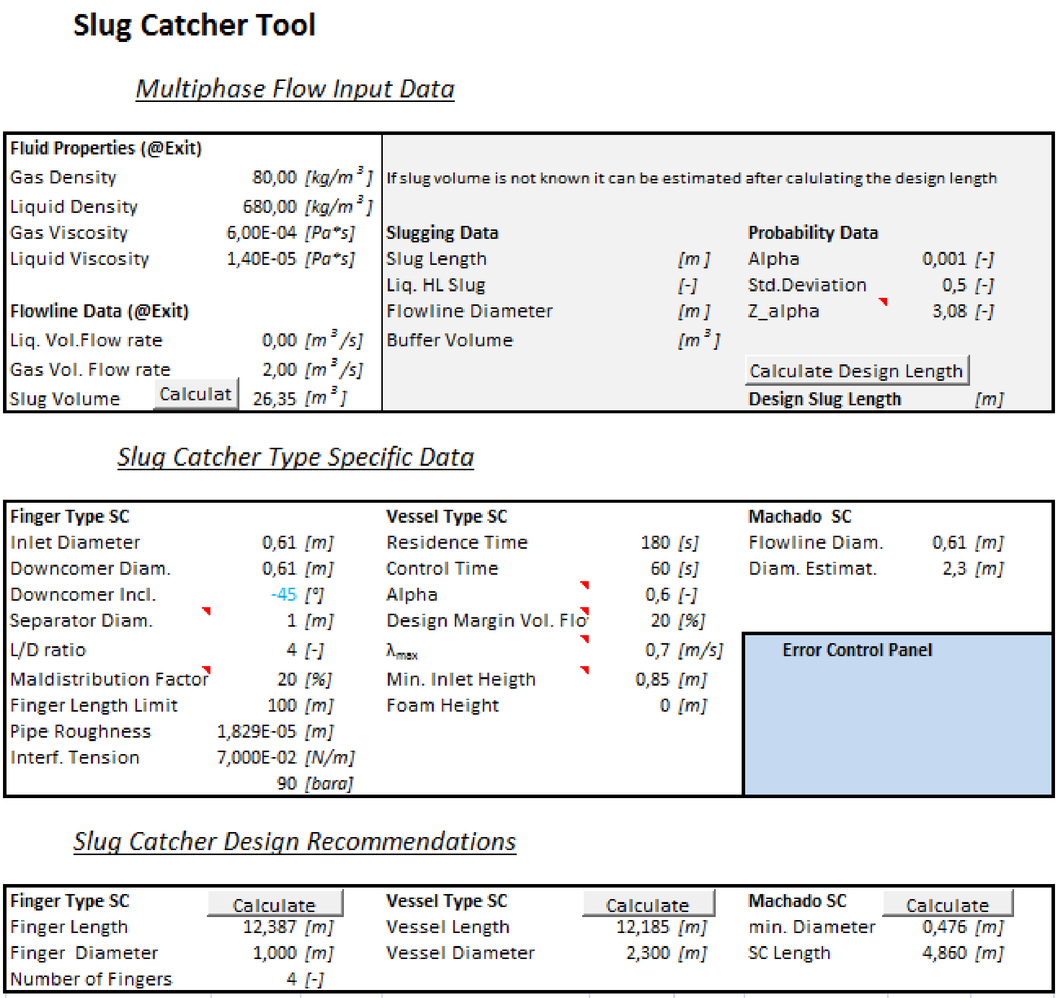


Figure 33: User interface of the slug catcher tool.

The finger-type and one of the vessel type routines are developed on the basis of SHELL Design and Engineering Practice (1998) and SHELL Design and Engineering Practice (2002). The second method for sizing a vessel-type slug catcher is based on the approach according to Machado (1977). Therefore, the program basically consists of three main routines. Each of them is presented below.

6.3.1. Finger-type slug catcher routine[14]

In a first step the program estimates the number of separation bottles (finger) and assumes an inclination of -1%. The velocity of the gas phase at the finger entrance can be determined (note that this also involves a hold up prediction for the finger). If the velocity does not meet the stratified - intermittent transition criterion as shown in section 3.1 the inclination is reduced by 1 % until a limit of -30%. If the change in inclination does not achieve stratified flow pattern the number of fingers has to be increased.

Determining the L/D ratio is not part of this routine and therefore has to be provided as input data. On the basis of the L/D ratio the length required for the separation process is calculated. Adding up this length to the length required for the storage of the slug volume the total length of the finger is obtained. If the total length should not exceed a certain limit, for example 100 m, the program recalculates the length with an increased number of fingers. The length limit can be specified by the user. The whole procedure is summarized in Figure 34.

[14] (SHELL, 1998)

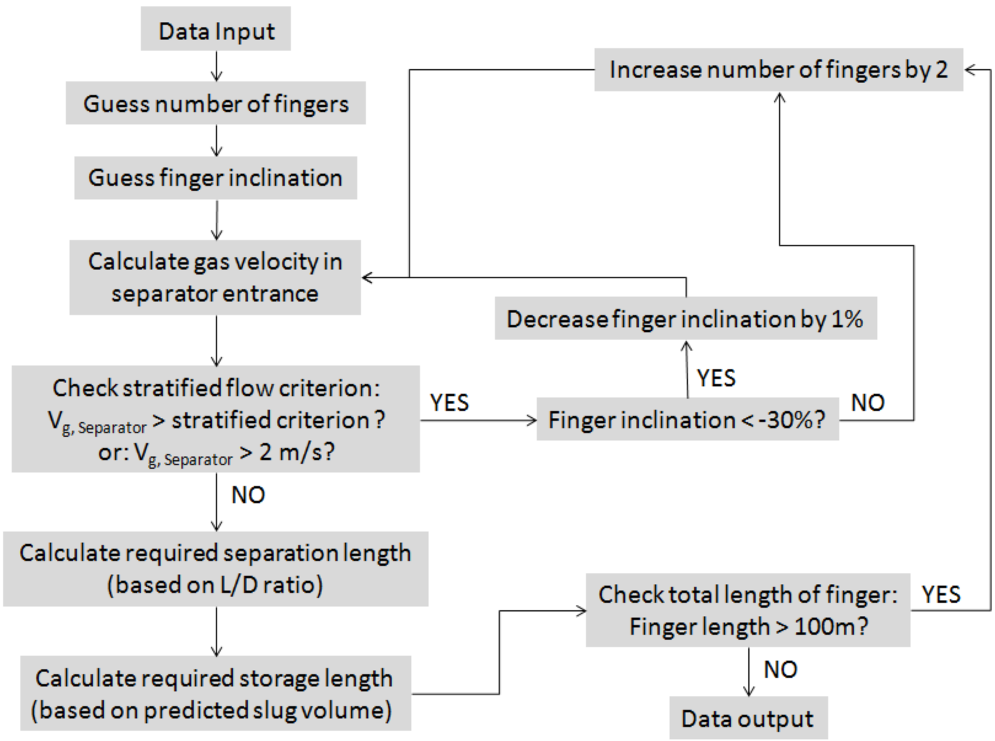


Figure 34: Calculation procedure of the finger-type slug catcher routine according to SHELL Design and Engineering Practice (1998).

6.3.2. Vessel type slug catcher routine[15]

For a first guess of the vessel diameter the program needs a minimum required height for level control as well as the minimal diameter depending on $A_{g,min}$. These parameters have to be specified on the input spreadsheet. The program is then estimating initial values for LA(H) and LA(L). From geometrical relationships the cross sectional areas depending on the estimated LA(H) and LA(L) can be obtained. Taking into account possible slugs the vessel length is then determined. Together with the initial guess of the diameter the length is used to calculate a L/D ratio. Now, depending on the size, the program checks the ratio against three criteria as shown in Figure 35. If the ratio is larger than 6 the diameter is increased and the entire procedure is repeated. In the other two cases the control times are checked. If the times are met the width of the control band(s) are increased (by 10% of the ratio specified to calculated control time) and LA(L) and LA(H) respectively have to be recalculated. If the calculated control times are close enough to the specified control times the program checks the final three criteria as indicated in Figure 35. If one of those criteria is not met the program repeats the calculation of LA(L) and LA(H) until the right diameter is found. Figure 35 illustrates the entire routine.

[15] (SHELL, 2002)

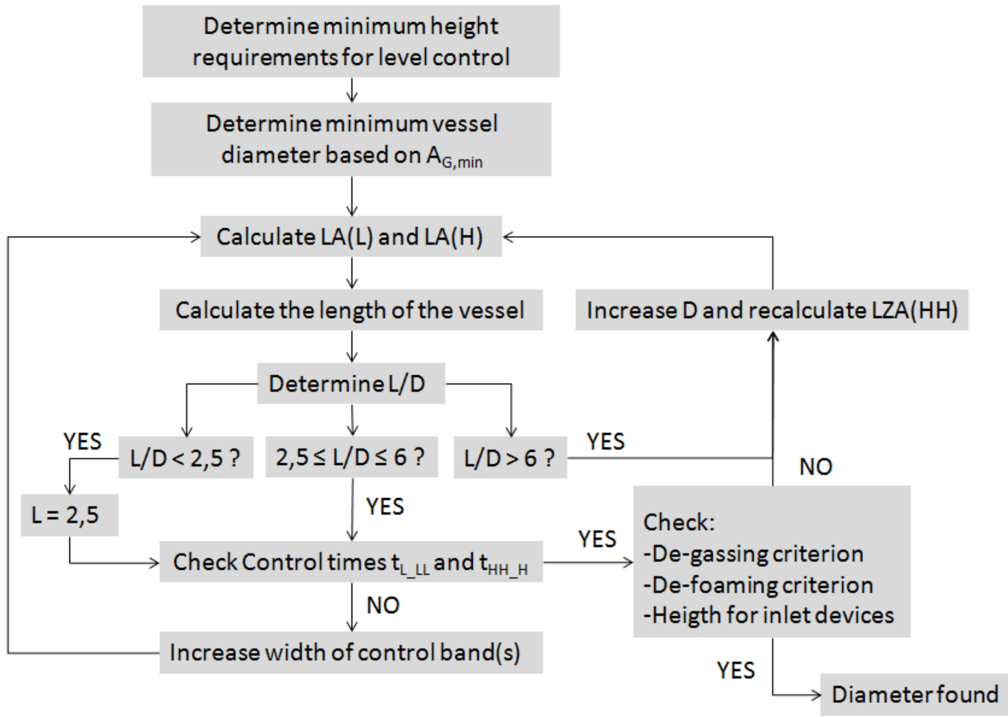


Figure 35: Calculation procedure of the vessel type slug catcher according to SHELL Design and Engineering Practice (2002).

6.3.3. Vessel type slug catcher routine based on the approach of Machado *(1977)*

The vessel type slug catcher routine based on the approach of Machado is an iterative procedure where the vessel diameter is determined on basis of multiphase flow analysis. An initial guess of the vessel diameter is required. The diameter predicted with the previously explained routine serves as a good approximation for estimating the initial diameter. The first calculation step is to determine the transition hold up in the slug catcher, that is, the highest allowable liquid hold up where intermittent flow is still not occurring. Based on this hold up the program calculates a minimum diameter, sufficiently large to guarantee stratified flow. This diameter can be seen as output in the user interface of the tool. This diameter is probably much smaller than the initial guess. However, to avoid large slug catcher lengths it is recommended to chose a large diameter. Afterwards, the actual hold up and the critical hold up (at which intermittent flow would appear) depending on the selected diameter are determined. Finally, the slug catcher length is calculated as a function of the slug volume, the critical hold up and the actual hold up. The main steps of the routine are summarized in Figure 36.

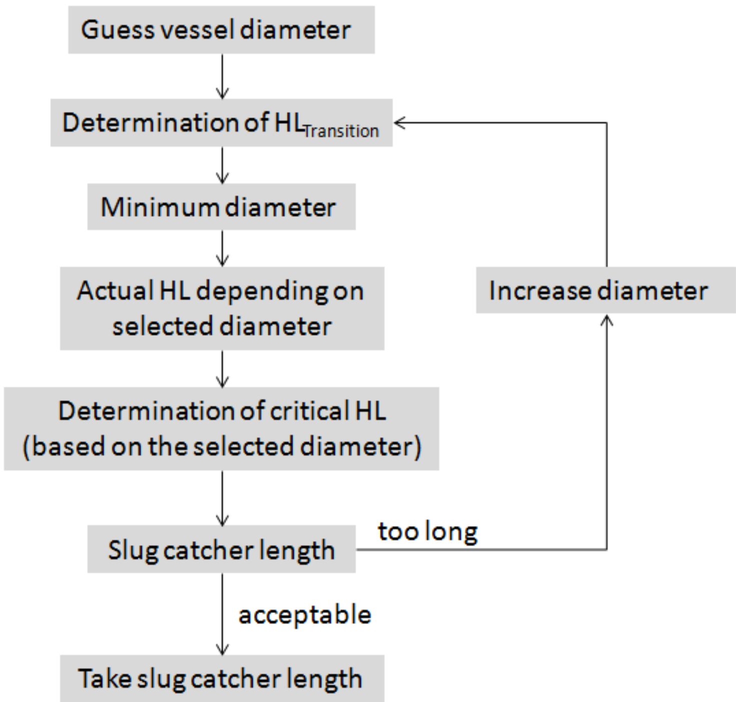


Figure 36: Calculation procedure for a vessel type slug catcher according to Machado (1977).

6.4. Slug catcher tool validation

The purpose of this section is the validation of the developed slug catcher tool. Based on a slug catcher design study a finger-type and a vessel slug catcher are sized with the slug catcher tool. The results are then compared with the slug catchers designed in the study.

6.4.1. Finger-type slug catcher

The design specification as well as the fluid properties required for the finger-type slug catcher sizing are taken from a slug catcher design study carried out by Contreras and Foucart (2007). For the development of the Margarita gas field (Bolivia) a finger-type and a vessel type slug catcher are compared in terms of economic aspects. A slug volume of 26,35 m^3 is predicted and it requires a finger-type slug catcher with four fingers where each finger has a diameter of 1 meter and a length of 10 meters. Alternatively, a vessel type slug catcher with a diameter of 2,6 meters and a length of 14 meters could be installed.

An oil flow rate (or liquid flow rate respectively) is required to size a slug catcher. Since the study does only state a gas flow rate (18 $MMSm^3/d$) the liquid flow rate has to be assumed. Due the fact that a gas field is basis of this study it is assumed that the Oil-Gas Ratio does not exceed 10 Sm^3/Sm^3 and that the oil gravity is about 50° API. Although this is a rough assumption it should be mentioned that the liquid properties such as density and viscosity do not have a significant effect on the number and length of fingers. Based on calculations with PipeSim (Black Oil model) the physical properties for the oil at a design pressure of 90 bar and a design temperature of 50 °C (given in the study) can be obtained and are shown in Table 11.

The gas physical properties and the gas flow were also calculated with PipeSim (Black Oil model) at the design pressure of 90 bar and the design temperature of 50 °C.

In terms of design specifications only the inlet diameter (0,61 m) and the separator (=finger) diameter (1 m) are given in the design study. A mal-distribution factor of 20% is assumed based on SHELL Design and Engineering Practice (1998). A L/D ratio of 4 is estimated from the drawing of the finger-type slug catcher which can be found in the study.

Table 11: Fluid properties, flow rates and design specifications for finger-type slug catcher sizing.

Fluid properties and flow rates (@ P=90bar, T=50°C)		
Gas density	80	*[kg/m³]*
Liquid density [1])	680	*[kg/m³]*
Gas viscosity	$6*10^{-4}$	*[Pa*s]*
Liquid viscosity [1])	$1,4*10^{-5}$	*[Pa*s]*
Liquid flow rate [1])	$2*10^{-3}$	*[m³/s]*
Gas flow rate	2,0	*[m³/s]*
Slug volume	26,35	*[m³]*
[1]) These parameters were not given in the study provided by Contreras and Foucart (2007) and had to be assumed (P=90 bar, T=50°C).		
Design specifications		
Inlet diameter	0,61	*[m]*
Separator diameter	1	*[m]*
L/D ratio [2])	4	*[-]*
Mal-distribution factor [2])	20	*[%]*
[2]) This parameter is assumed based on SHELL Design and Engineering Practice (1998).		

The following Table 12 contains the design recommendations obtained with the slug catcher tool and the sizing recommendations provided by the study.

Table 12: Comparison of slug catcher sizing according to the slug catcher design study and slug catcher sizing performed with the slug catcher tool.

Slug Catcher sizing acc. to slug catcher design study		
Storage length	10	*[m]*
Separator diameter	1	*[m]*
Separator inclination	-	*[°]*
Capacity slug volume	28,5	*[m^3]*
Number of fingers	4	*[-]*
Slug catcher sizing with slug catcher tool		
Separator length	4	*[m]*
Separator diameter	1	*[m]*
Separator inclination	-0,6	*[°]*
Storage length	17	*[m]*
Finger length	21	*[m]*
Capacity slug volume	26,35	*[m^3]*
Slug catcher total volume	32,6	*[m^3]*
Number of fingers	2	*[-]*

6.4.2. Vessel type slug catcher

The fluid properties and flow rates required for the vessel type slug catcher sizing are taken from a slug catcher design study (Contreras & Foucart, 2007) as summarized in Table 11 . Since detailed design specifications are not stated in the study they have to be assumed according to SHELL Design and Engineering Practice (2002).

Table 13: Design data and specifications for vessel type slug catcher sizing.

Vessel type design specifications		
Residence time	180	*[s]*
Control time	60	*[s]*
Alpha (semi-eliptical head)	0,5	*[-]*
Design margin liq. flow rate	20	*[%]*
λ_{max} (max. allowable gas load factor)	0,7	*[m/s]*
Min. inlet height	0,85	*[m]*
Foam height	0	*[m]*

The following Table 14 contains the design recommendations obtained with the slug catcher tool and the sizing recommendations given by the design study.

Table 14: Slug catcher sizing according to a slug catcher design study and slug catcher sizing performed with the slug catcher tool.

Slug catcher sizing acc. to slug catcher design study		
Vessel length	16	*[m]*
Vessel diameter	2,6	*[m]*
Capacity slug volume [1])	35,8	*[m^3]*
Vessel volume	85	*[m^3]*
[1]) Note that the capacity slug volume differs from the predicted slug volume of 26,35 m^3. The design study does not state any reasons for his deviation.		
Slug catcher sizing with slug catcher tool		
Vessel length	12,3	*[m]*
Vessel diameter	2,3	*[m]*
Vessel volume	51	*[m^3]*

The approach according to Machado (1977) is also validated based on the design study (ref. to Table 11). Table 15 contains the predicted vessel diameter and length.

Table 15: Slug catcher sizing according to a slug catcher design study **(Contreras & Foucart, 2007)** and according to the approach of Machado **(1977)**.

Slug Catcher sizing according to Machado		
Vessel length	5	*[m]*
Vessel diameter	2,3	*[m]*

7. Summary

In order to gain knowledge regarding multiphase flow the purpose of this thesis was to develop a multiphase flow analysis- and a slug catcher-tool.

Therefore, specific information on multiphase flow phenomena was collected and a multiphase flow calculation model (Xiao *et al.*) was introduced. Models for determining the temperature profile and slug characteristics were discussed. These models served as basis for the multiphase flow analysis and they are part of the program code.

The Compositional model and Black Oil model as two different ways of approaching fluid property approximation were explained. The latter one was also implemented in the program code of the developed multiphase flow analysis tool.

Multiphase flow calculations were performed to validate the developed multiphase flow tool. PipeSim was used to perform a correlation comparison and a sensitivity study. Furthermore, a decision tree was developed based on selection criteria from the PipeSim user guide.

Finally, theoretical and design concepts of slug catcher were introduced and the developed slug catcher tool was validated on basis slug catcher design study.

8. Conclusions and outlook

Regarding the correlation comparison (section 5.3.) and sensitivity study (section 5.4.) it must be mentioned that the results obtained in these sections refer to a specific case (ref. to Table 6) and may not be valid for different cases. This is especially true for the correlation comparison since - depending on the considered case - some correlations can lead to sufficient approximations in terms of pressure and slug length prediction while other correlations can completely fail. Therefore, Table 8 and Figure 23 are meant to serve as basic support whenever appropriate correlations have to be selected.

As can be seen from the validation spreadsheet for simple profiles (refer to Appendix 2) the developed multiphase flow tool tends to under predict the temperature as far as inclinations of less than -10° are concerned. It was found that the relative error in predicting the temperature drop may exceed 500% (Horizontal distance = 1000m. Inclination = -38,7°). Such high deviations were only found in case of downward flow while the relative error in the predicted temperature for upward flow did not exceed 10%. The temperature prediction in case of inclinations less than 5° could be approximated most satisfyingly.

One reason for the deviation of the predicted temperature can be found in the calculation of the Joule-Thompson coefficient for the gas phase. While the equation provided by PipeSim user guide for determining the Joule-Thompson coefficient involves a partial differentiation of the compressibility factor with respect to temperature (p = const) this term was approximated by $\Delta Z/\Delta T = (Z_2 - Z_1)/(T_2 - T_1)$ in the multiphase flow program.

It was furthermore found that the developed tool is predicting the pressure drop incorrectly between -1° and 1°. This deviation can be explained by analysing the friction pressure gradient which becomes dominant at small inclinations. The multiphase flow tool tends to over predict the friction pressure gradient. Due to the fact that the gravitational pressure gradient controls the total pressure loss at larger inclinations the program predicts the pressure drop satisfyingly at inclinations smaller than -5° and larger than 5°.

The validation with a given elevation profile (15,5 km) as indicated in Figure 20 showed that the multiphase flow tool predicts temperature and pressure accordingly to PipeSim with a deviation of 2,4 % in terms of temperature and 1,3 % in terms of pressure. Even though the considered flow line consisted mainly of sections with small inclinations the tool provided a satisfying approximation of the exit pressure. Larger deviations may be found if longer flow lines are considered. The temperature is also approximated with sufficient accuracy because of the high liquid fraction in the considered case (liquid hold up > 60 %). Consequently, the enthalpy balance is dominated by the liquid phase where Joule-Thomson expansion is negligible.

As can be seen from Table 12 the developed slug catcher tool predicts 2 fingers instead of 4 as stated by Contreras and Foucart (2007). Due to the low liquid flow rate the tool suggests single phase flow where slugs are not likely to occur. Therefore, more than 2 fingers may only be required if the liquid flow rate is increased until the gas velocity exceeds the stratified-intermittent transition (ref. to Figure 34 and section 3.1 respectively). Apart of the separator diameter the transition criterion plays a crucial role in the determination of the number of fingers. Since Contreras and Foucart (2007) suggest 4 fingers in their study the transition criterion is probably defined more conservative so that the transition occurs at lower gas velocities what would result in a higher number of separation bottles.

In terms of the vessel-type slug catcher it can be concluded that the tool approximates the vessel diameter and length with sufficient accuracy. Since the slug catcher design study (Contreras & Foucart, 2007) suggested a diameter of 2,6 m and a vessel length of 16 m (ref. to Table 14) it must be noted that these sizes refer to a slug volume of 35 m^3 (instead of 26,35 m^3 as originally stated in the study) and that there is no information on specific vessel design parameters such as residence time, control time, inlet height and foam height. Therefore, these parameters had to be assumed what can be identified as a source of the deviations (23 % relative error in terms of length and 11,5 % relative error in terms of diameter).

As can be seen from Table 15 the Machado (1977) approach cannot predict the vessel length with sufficient accuracy. Since this approach does not account for design requirements such as level control, inlet height or foam height the application of this tool is only recommended for cases where a detailed vessel sizing based on SHELL Design and Engineering Practice (2002) cannot be made. In such cases the vessel length calculated with the Machado approach may be multiplied at least by a factor 2.

Additionally, it is recommended that both the slug catcher tool as well as the multiphase flow tool may further be validated with different cases to get a more comprehensive understanding of the performance of the tools.

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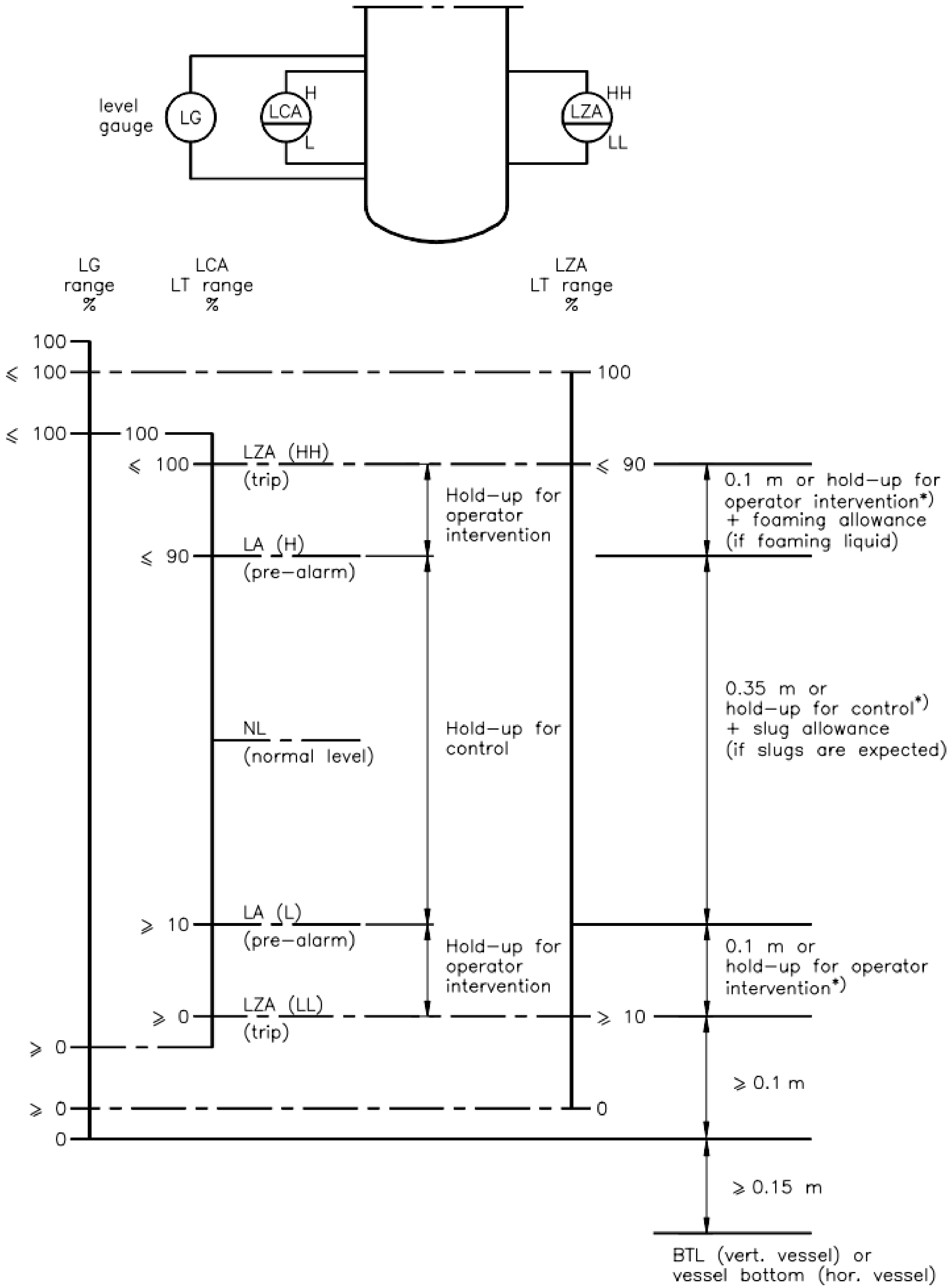
10. List of tables

11. List of figures

12. Appendix

1) Liquid level control in a gas/liquid separator (SHELL, 2002)
2) Validation of multiphase flow tool based on simple profiles



*) whichever is larger

Validation of multiphase flow tool based on simple profiles

P_Start	137,895	*bara*
T_Start	93,333	*°C*

Hor. Dist. [m]	Vert. Dist [m]	Incl. [°]	Pipe Sim		MPF Tool		ΔP PipeSim	ΔP Tool	ΔP rel. Error [%]	ΔT PipeSim	ΔT Tool	ΔT rel. Error [%]
			Temp. [°C]	Pressure [bara]	Temp. [°C]	Pressure [bara]						
1000	-800	-38,7	92,78	147,3	89,98	147,36	9,41	9,47	0,64	0,55	3,35	506,33
	800	38,7	85,71	99,75	86,03	99,49	38,15	38,41	0,68	7,62	7,30	4,20
	-400	-21,8	91,61	142,51	90,25	142,52	4,61	4,63	0,22	1,72	3,08	78,93
	400	21,8	88,13	118,06	88,21	117,89	19,84	20,01	0,86	5,20	5,12	1,54
	-200	-11,3	90,88	140,16	90,24	140,15	2,26	2,26	0,44	2,45	3,09	26,09
	200	11,3	89,16	127,69	89,45	127,62	10,21	10,28	0,69	4,17	3,88	6,95
	-100	-5,7	90,47	139	90,22	138,98	1,10	1,08	1,81	2,86	3,11	8,73
	100	5,7	89,67	132,57	89,79	132,53	5,33	5,37	0,75	3,66	3,54	3,28
	-10	-0,6	90,07	137,88	90,08	137,78	0,02	0,12	666,67	3,26	3,25	0,31
	10	0,6	89,99	137,1	90,06	137,09	0,80	0,81	1,26	3,34	3,27	2,09
	-1	-0,1	90,03	137,65	90,05	137,49	0,25	0,41	65,31	3,30	3,28	0,61
	1	0,1	90,02	137,58	90,07	137,67	0,31	0,23	28,57	3,31	3,26	1,51
2000	-1600	-38,7	92,22	157,42	86,89	157,67	19,53	19,78	1,28	1,11	6,44	478,89
	1600	38,7	78,1	67,41	79,44	67,16	70,49	70,74	0,35	15,23	13,89	8,80
	-800	-21,8	89,98	147,33	87,32	147,39	9,44	9,49	0,64	3,35	6,01	79,33
	800	21,8	83,08	99,42	83,16	99,15	38,48	38,75	0,70	10,25	10,17	0,78
	-400	-11,3	88,56	142,5	87,28	142,49	4,60	4,60	0,22	4,77	6,05	26,82
	400	11,3	85,16	117,72	85,56	117,61	20,18	20,29	0,55	8,17	7,77	4,89
	-200	-5,7	87,75	140,12	87,17	140,1	2,22	2,20	0,90	5,58	6,16	10,39
	200	5,7	86,08	127,3	86,46	127,2	10,60	10,70	0,94	7,25	6,87	5,24
	-20	-0,6	86,97	137,86	87,06	137,66	0,03	0,24	571,43	6,36	6,27	1,41
	20	0,6	86,81	136,3	87,01	136,28	1,60	1,62	1,25	6,52	6,32	3,07
	-2	-0,1	86,9	137,41	86,95	137,09	0,49	0,81	65,98	6,43	6,38	0,78
	2	0,1	86,88	137,26	87,05	137,45	0,64	0,45	29,92	6,45	6,28	2,63
5000	-4000	-38,7	90,44	192,54	79,54	194,16	54,65	56,27	2,96	2,89	13,79	376,77
	4000	38,7	Pressure is too low				137,90	137,90	0,00	#WERT!	93,33	#WERT!
	-2000	-21,8	85,54	163,13	79,64	163,47	25,24	25,58	1,35	7,79	13,69	75,71
	2000	21,8	68,91	52,56	69,63	52,35	85,34	85,55	0,25	24,42	23,70	2,95
	-1000	-11,3	82,27	149,94	79,4	149,98	12,05	12,09	0,33	11,06	13,93	25,94
	1000	11,3	74,2	89,52	74,6	89,33	48,38	48,57	0,39	19,13	18,73	2,09
	-500	-5,7	80,42	143,67	79,07	143,64	5,77	5,74	0,52	12,91	14,26	10,45
	500	5,7	76,48	111,7	77,24	111,51	26,20	26,39	0,73	16,85	16,09	4,51
	-50	-0,6	78,62	137,84	78,82	137,38	0,06	0,52	836,36	14,71	14,51	1,36
	50	0,6	78,25	133,88	78,65	133,86	4,02	4,04	0,50	15,08	14,68	2,65
	-5	-0,1	78,45	136,69	78,58	135,9	1,21	2,00	65,56	14,88	14,75	0,87
	5	0,1	78,41	136,31	78,8	136,82	1,59	1,08	32,18	14,92	14,53	2,61
10000	-8000	-38,7	87,11	268,46	75,95	273,52	130,57	135,63	3,88	6,22	17,38	179,33
	8000	38,7	Pressure is too low				137,90	137,90	0,00	#WERT!	93,33	#WERT!
	-4000	-21,8	79,36	194,35	70,16	195,86	56,46	57,97	2,67	13,97	23,17	65,84
	4000	21,8	Pressure is too low				137,90	137,90	0,00	#WERT!	93,33	#WERT!
	-2000	-11,3	73,75	163,87	68,88	164,16	25,98	26,27	1,12	19,58	24,45	24,87
	2000	11,3	58,89	49,72	58,88	49,52	88,18	88,38	0,23	34,44	34,45	0,03
	-1000	-5,7	70,43	150,13	68,17	150,13	12,24	12,24	0,00	22,90	25,16	9,87
	1000	5,7	63,31	86,88	64,09	86,63	51,02	51,27	0,49	30,02	29,24	2,60
	-100	-0,6	67,32	137,85	67,57	137,03	0,05	0,87	1822,22	26,01	25,76	0,96
	100	0,6	66,66	129,82	67,31	129,77	8,08	8,13	0,62	26,67	26,02	2,44
	-10	-0,1	67	135,51	67,15	133,92	2,39	3,98	66,67	26,33	26,18	0,57
	10	0,1	66,94	134,73	67,53	135,84	3,17	2,06	35,07	26,39	25,80	2,24

$$\Delta P = abs(P_{Start} - P)$$

$$\Delta T = abs(T_{Start} - T)$$

$$Err_{\Delta P,rel} = \frac{abs(\Delta P_{PipeSim} - \Delta P_{Tool})}{\Delta P_{PipeSim}}$$

$$Err_{\Delta T,rel} = \frac{abs(\Delta T_{PipeSim} - \Delta T_{Tool})}{\Delta T_{PipeSim}}$$