

Assessment of Possible Flow Assurance Issues in a Production Well

Master Thesis in Production Engineering



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"We usually find gas in new places with old ideas. Sometimes, also, we find gas in an old place with a new idea, but we seldom find much gas in an old place with an old idea. Several times in the past we have thought that we were running out of gas, whereas actually we were only running out of ideas."

Parke A. Dickey

Acknowledgements

Chania, September, 2015

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ABSTRACT

Hydrocarbon exploration becomes complicated due to recent developments in offshore, deepsea and permafrost zones. Field developments in areas never explored before introduce new issues that have to be tackled from the production engineers to achieve economic production rates. Flow assurance in pipelines and specifically hydrate formation, becomes an issue of major importance in the system design of hydrocarbon fields.

The objective of this thesis is the assessment of the hydrate forming conditions in a hydrocarbons production system when the system design parameters are varied. The first part is introductory and deals with the flow assurance issues occurring in petroleum industry operations. Subsequently, it focuses on one of the most important flow assurance issues, the hydrate formation. A review on hydrate thermodynamics and probable areas of formation in gas and oil-dominated systems is given, followed by a brief analysis on how hydrate plugs are formed. Finally a short discussion about the hydrate potential as a hydrocarbon resource is given followed by a presentation of all possible remedial applications on the hydrate problem, be it prevention or treatment.

The second part deals with hydrate formation on two hydrocarbon systems (Light Oil and Gas condensate) at two developed fields which represent an onshore arctic and a deep offshore region. A sensitivity analysis was done to study how the system parameters affect the hydrate forming conditions. For this task, PROSPER by PetEx, the industry standard production simulation software was utilized to simulate all possible scenarios. The sensitivity analysis is split into two parts. The first discusses the effect of system parameters that affect the well deliverability (VLP curve) and the second deals with the effect of two thermodynamic inhibitors in various concentrations.

Concluding, the effect of all system parameters on the hydrate formation is given. The parameters that stand out are: the water percentage of the flow stream and the pipeline insulation. Both are equally important parameters that can deteriorate or eliminate the hydrate formation problem regardless the environmental conditions prevailing at the field development.

Classical design against hydrate formation is targeting to the avoidance of the hydrate stability region, however new methods are under development, which shift their targeting in producing the hydrate particles and further research must stand up to that task.

1. INTRODUCTION

Hydrocarbons are produced from wells that penetrate geological formations rich on oil and gas. As the wells are perforated in the oil and gas bearing zones, the hydrocarbons can flow to the surface provided the reservoir pressure is high enough to overcome the back pressure from the flowing fluid column in the well and the surface facilities.

Production wells are drilled and completed to move the oil or gas from its original location inside the reservoir to the stock tank or to the sales point. Movement or transport of these fluids to the surface requires energy to overcome friction losses in the system. The fluids must travel through the reservoir and the piping system and ultimately flow into a separator for gas-liquid separation. Depending on the reservoir properties, the production system can be relatively simple or it may include many components in which energy or pressure losses occur. As a result, the production rate or deliverability of a well can often be severely restricted by the performance of system components or other external parameters.

The most important of those restrictions belong to the flow assurance domain. Severe production rate decreases occur in pipe and well systems due to flow assurance issues such as hydrate, waxes, asphaltenes and scale deposition. Flow assurance problems are not directly connected to uncontrollable parameters such as fluid composition and environmental conditions. However the system design can actively affect the forming conditions and the area of accumulation of the deposits

From the discussion above, the necessity to design the production system so as to avoid potential flow assurance issues becomes critical. For this purpose PROSPER has been developed by PetEx to model and simulate all possible production scenario. The process is relatively straightforward. Selection of system design -> Simulation in PROSPER -> Evaluation danger of hydrates.

In this chapter an introduction to Flow Assurance is given. A description on how hydrate formation affects hydrocarbon production operations is later discussed. Details on hydrates structure, properties and forming conditions are given in chapter 2. All possible applications aimed on the avoidance or removal of hydrate formation is given in chapter 3.

The objective of this thesis is the assessment of the hydrate forming conditions in a hydrocarbons production system when the system design parameters are altered or chemical substances are introduced in the fluid stream. To be able to answer if the hydrate formation risk is possible, the pressure, temperature and fluid composition must be known along the pipeline. In chapter 4 the principles and required tools that make these calculations possible are given.

Two hydrocarbon systems (Light Oil and Gas condensate) at two developed fields which represent an onshore arctic and a deep offshore region are described in chapter 5 along with all the required information regarding these scenarios.

A sensitivity analysis was done to study how the system parameters affect the hydrate forming conditions. For this task, PROSPER by PetEx, the industry standard production simulation software was utilized to simulate all possible scenarios in chapter 6. The sensitivity analysis is split into two parts. The first discusses the effect of system parameters that affect the well deliverability (VLP curve) and the second deals with the effect of two thermodynamic inhibitors in various concentrations.

In the last chapter a conclusion is given on which system parameters affect most the hydrate formation, which are the main differences between the two fluid types and which region displays the worst behavior and why. Finally some thoughts on further research is given.

1.1 INTRODUCTION IN FLOW ASSURANCE IN THE PETROLEUM INDUSTRY

Flow assurance is the technical discipline that guarantees achievement of lifting and transporting a system's lifetime production targets from the near-wellbore to offloading tanks by predicting, preventing, and solving problems originated by the behavior of the transported substances (i.e., gases, liquids, and solids either separated or in multiphase conditions) (Lullo, 2012).

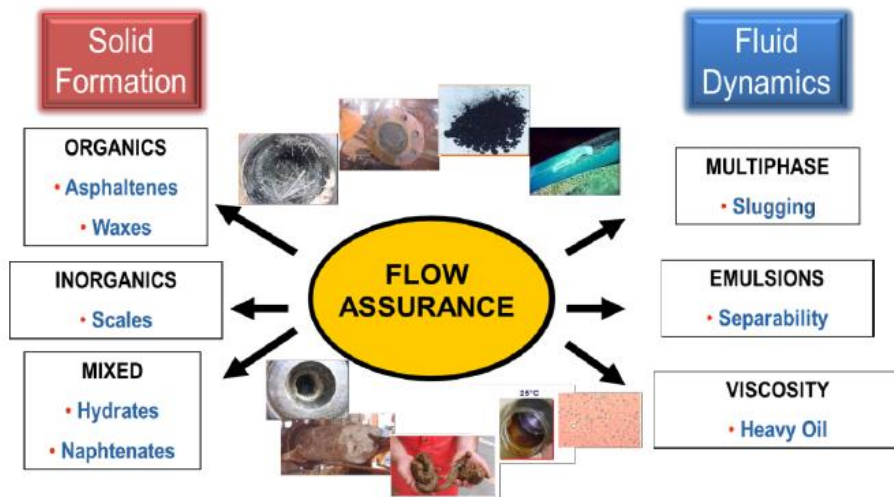


Figure 1: Flow assurance issues during production (Lullo, 2012)

A focus on flow assurance has some physical boundaries. Qualitatively, whenever flow from the reservoir is confined within a very limited space, flow assurance starts becoming an issue. This may happen in pipes of any sort (e.g., well tubings, flowlines, and pipelines), but also in plant facilities (e.g., valves, manifolds, separators, and slug catchers) and in the area called the near-wellbore, where the fluids must accelerate and squeeze close together to enter the well. The property common to all these places is that even tiny transformations produced by the fluids, may have drastic effects on the capability to produce at the desired hydrocarbon flow.

This leads to the real heart of flow assurance: the behavior of what moves in those restricted spaces. Hydrocarbons are there, either in vapor or liquid form (or both), but so will reservoir water, reinjection water, sand, corrosion products, formation debris, asphaltenes, particles, and other components. Combinations of immiscible substances, such as emulsions (liquid/liquid) and foams (vapor/liquid), have properties of their own, quite different from those of the composing fluids, and deserve special description. In this complex context, the objectives of flow assurance is making sure we deeply understand the fluid mechanics, oil

field chemistry, heat transfer, process instrumentation and control so a well's production targets can be achieved. It is important that we can predict fluid pressure and temperature as a function of reservoir behavior over field life, the performance of energy boosting methods and means of reducing pressure and temperature losses. We need to manage corrosion, erosion, wax deposition, scale deposition, and hydrate formation (Figure 1). The effect of unsteady flow, on the stability of process controls and equipment, continues to limit the operating range of subsea systems.

In shallow water or onshore, operators often got a lot of solid deposits in the production system, but gaining access to the wellbore and the flowlines was easier. As operators move into more remote regions, deeper offshore depths, and into regions that yield more challenging reservoir fluids, they face difficulty in keeping the produced fluids and gas flowing through the lines. Deepwater wellbores are more expensive to access and present more complexity in mitigating blockage of flowlines (McMullen, 2006), (Lullo, 2012).

Profitability, especially in deep-water developments often hinges on the ability of the designers to ensure consistent and controllable flow interventions. This requires costly equipment, and equally costly downtime. Additionally, deep-water operations amplify environmental and safety concerns due primarily to inaccessibility of the flowline or pipeline. (Wasden, 2003).

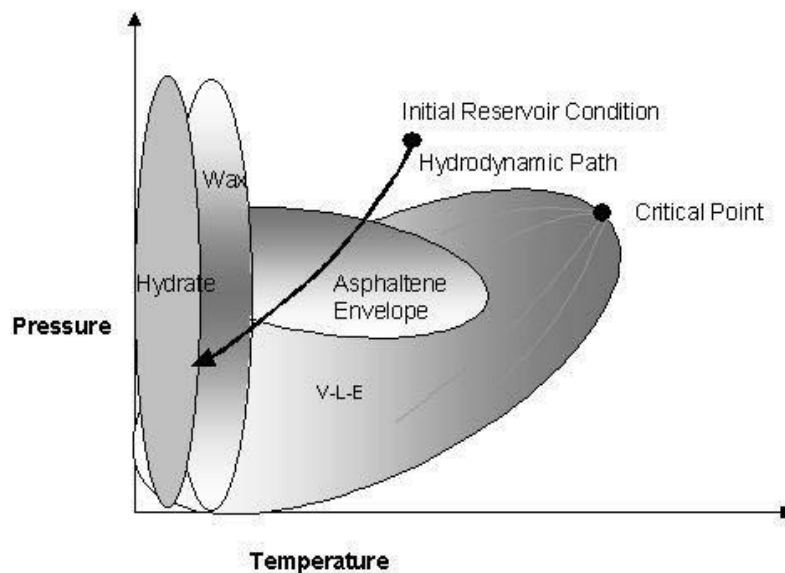


Figure 2: Pressure – Temperature along production line, passing areas exhibiting Flow assurance issues

1.2 GENERAL FLOW ASSURANCE ISSUES

The major flow assurance issues for the multiphase production, through pipelines and risers for both offshore and onshore oil and gas field developments, are summarized as follows:

Lack of reservoir energy: Lack of reservoir energy leads to insufficient hydrocarbon flowrate, which could render the production not economic or even result in a non-flowing well. Remediation is needed in this case by changing the system design (artificial lift, reservoir stimulation) in order to achieve the desired flowrate.

Hydrates: Natural gas hydrates are ice-like solids that form when free water and natural gas combine with high pressure and low temperature. This can occur in gas and gas/condensate wells, as well as in oil wells and restrict hydrocarbon flow or even plug the pipeline entirely. Location and intensity of hydrate accumulations in a well vary and depend on:

- Operating flow regime
- Well design
- Geothermal gradient of the system
- Fluid composition
- Other factors



Figure 3: Flow restriction due to hydrate formation and agglomeration on the pipeline wall

Wax: Waxes are high molecular weight, highly saturated organic substances. The main organic compound in crude oil that precipitates wax at production operating conditions is paraffin compounds, which are insoluble in the oil. Paraffins consisting of carbon numbers greater than C_{20} are considered potential problems for the production system.

The prediction of the potential for wax deposition problems is fundamentally based on the determination of the physical characterization of the oil (paraffin content) and WAT (Wax Appearance Temperature). The formation of wax crystals depends mostly on temperature change, while pressure and composition also affect their formation but not to a significant extent. Large quantities of wax deposition can require a major shutdown operation to clear the blockage with associated economic penalties to the development and consequential loss in production and revenue (Joshi et al, 2003).



Figure 4: Paraffinic wax precipitation on the pipeline wall

Asphaltenes: Asphaltenes are defined as high molecular weight, aromatic, polar compounds that are soluble in toluene but are precipitated by alkanes. Generally, asphaltenes tend to remain in solution or in colloidal suspension under reservoir temperature and pressure conditions. They may start to precipitate once the stability of the colloidal suspension changes, which is caused by alterations in temperature and/or pressure during primary depletion. On the other hand, asphaltenes have been reported to become unstable as a result of fluid streams blending as well as by gas injection (Jamaluddin et al, 2001).

During production, asphaltenes are also known to precipitate as a result of change in pressure, temperature and or composition of the fluid. From literature and past history, it

is known that asphaltene precipitation is more likely to occur in an under-saturated, light reservoir fluid than a heavy hydrocarbon system. It is also noteworthy that problems due to asphaltenes occur in a two-step process:

- a) Precipitation from the reservoir fluid
- b) Deposition of the precipitated particles, which causes plugging.



Figure 5: Asphaltene deposition on the pipeline wall

Slugging: Slug flow in production pipelines and risers has been a major operational issue, associated with subsea field developments. This creates problems associated with instability in production flow due to pressure fluctuations, which can be caused by any of the following:

- Severe slugging at the riser base.
- Terrain induced slugging.
- Operational based slugging.

Corrosion: The risks associated with the use of common carbon steel for transporting wet oil and multiphase fluids obliges the operator to implement an internal corrosion management program through all phases of the system life.

Various mechanisms have been postulated for the corrosion process but all involve either carbonic acid or the bicarbonate ion formed on the dissolution of CO_2 in water.



Figure 6: Internal view of a corroded oil pipe

Erosion: Erosion due to sand production has been seen as the cause of a number of problems associated with separation efficiency, material loss and flow path blockages.

Sand screens or gravel packs are generally installed in the horizontal sections of production well bores to minimize sand production. Failure of these results and therefore allowance of sand production needs to be included into the systems design.

Emulsions: Under a combination of low ambient sea temperatures and high fluids viscosity, tight emulsions can occur between the water and oil phase. This can impair separation efficiency at the processing facility and thus cause loss in production.



Figure 7: Water/oil emulsion and separated phases after enough retention time

Scale: Scale compounds will precipitate out of water when their individual solubility in water is exceeded. The solubility is dependent upon the temperature, pressure and salinity conditions along the production flow path. Most troublesome scales are Barium Sulphate, Calcium Carbonate and Strontium Sulphate (Johal, 2007).

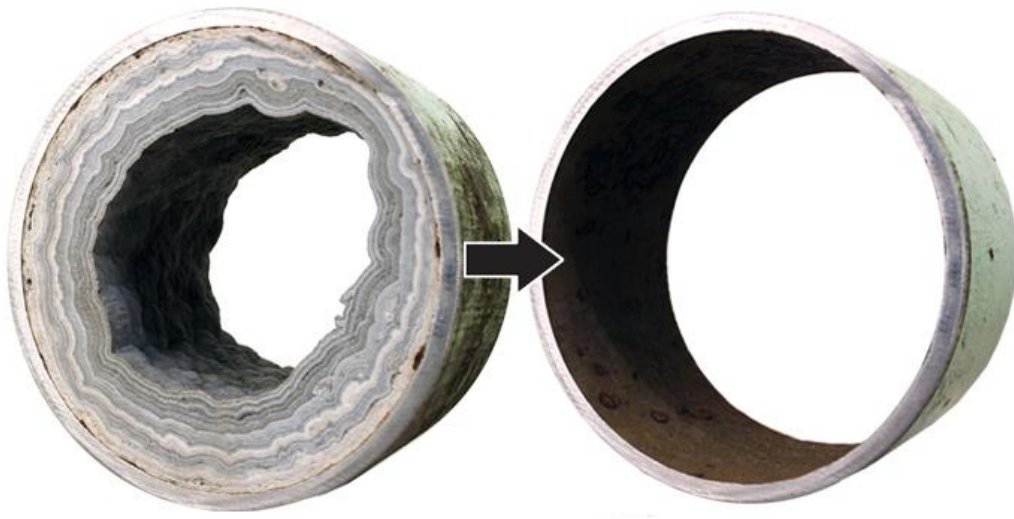


Figure 8: Steel tubing with and without scale deposition

Liquid loading of pipelines and wells: In a multiphase flow the area of the pipeline that each phase occupies is governed by the velocity of each phase. Generally gas has very high velocity inside the production pipelines and is potentially possible to lead in a pipeline filled with liquid. The hydrostatic head of the accumulated liquid can be high enough as to restrict the gas flow underneath causing intermittent (slug flow), abnormally high internal pressures to the pipeline or even killing the well. This problem should be addressed by implementing in the system design all possible multiphase flows that can occur and also ensure such pressure differential as to avoid severe slippage effect.

Chemicals compatibility: The potential for problems related to chemicals compatibility is due to the blending of various chemicals existing in the system. These chemical are used as a solution of different flow assurance problems, although possible interaction between them may result in negative effects in production. In this case, flow assurance considerations must extend beyond the well bore and subsea production equipment to the annulus maintenance systems, umbilicals and chemical distribution system. Interactions of seawater, production fluids, chemical treatment fluids, drilling and completion fluids, well treatment fluids, and hydraulic fluids must be considered to the extent that interaction is reasonably expected in the production environment. Interaction with the materials in the subsea system must also be considered.

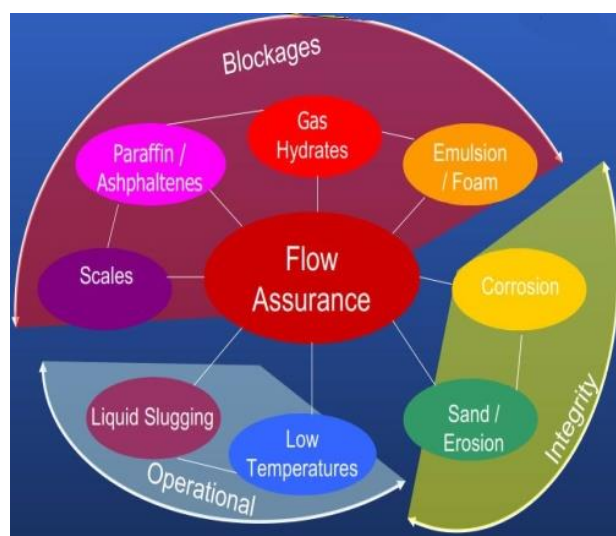


Figure 9: Flow assurance issues

The following are typical solutions in the flow assurance toolbox that help remediate the problems occurring in the production system:

- Optimal diameter, route, and number of pipes
- Artificial lift technologies, such as pumping/compression technologies, multiphase pumps, and gas lift
- Thermal insulation or active heating, to prevent hydrates and wax deposition during steady production and delay the formation of hydrates during shutdowns
- Use of internal pipe coatings
- Active flow control (by valves or chemical products)
- Regular use of inhibitors, pour point depressants, or solvents
- Injection of hydrate inhibitors (thermodynamic, kinetic, or colloidal)
- Use of emulsion breakers or foamers/de-foamers in wells and pipelines
- Pigging, for liquids or deposits removal

Flow assurance is a discipline in which innovations are often relatively easy to prototype and transfer to the field and can provide a competitive advantage. The combination of the flow assurance solutions is usually case specific, so engineering work is done within a company aiming for the optimal system design for a specific field.

1.3 THESIS CONTENT

In the following chapters a general aspect of the flow assurance issues will be given (do not use displayed!), coupled with a view of hydrate issues in the Petroleum industry. A thorough analysis of hydrates from a thermodynamic view is presented along with techniques that can prevent or remediate the hydrate formation. Later a description of the software, along with its principals, is given. The program is then used to realize a sensitivity analysis of system design parameters, connected to hydrate formation and the effect of different concentrations of thermodynamic inhibitors. The results will be presented and a discussion on the conclusions will be given.

2. GAS HYDRATES IN THE PETROLEUM INDUSTRY

In the oil and gas industry natural gas hydrates are comprised of small molecules and water. Hydrates are crystalline solid compounds formed from water and smaller molecules. They are a subset of compounds known as clathrates or inclusion compounds. Gas (clathrate) hydrates are dirty-looking crystalline structures formed when small non-polar gas molecules are enclathrated within water molecules. The process is concomitant with the presence of both host (water) and guest molecules (gas). Hydrate formation takes place at low temperatures (above or below ice point, 0 °C) and high pressures when light hydrocarbons and water are present and can occur anywhere and anytime (Mogbolu et. al, 2014).



Figure 10: Hydrate plug inside a pipeline

Sir Humphrey Davy is credited for discovering hydrates; chlorine in the early 19th century. Faraday, his assistant, reported the composition of chlorine hydrate in 1823 (Carroll, 2003). Hydrates became an intellectual curiosity in the 18 hundreds, and a lot of work was done on the subject.

Reservoir gas consists mainly of light alkanes, hydrocarbons, like methane, ethane, propane, butane and so on. Other components often found in the natural gas mixture are non-hydrocarbons like carbon dioxide, hydrogen, nitrogen, sulfur and water.

Natural gas hydrates, whether occurring in gas/condensate or oil systems, often represent the most dramatic flow assurance problem for a deep-water project. In many cases, ambient water temperatures surrounding flowlines and pipelines fall below those needed to prevent hydrate crystal growth in hydrocarbons for the (typically) high flowing pressures, leading to potential for forming large, solid plugs (Figure 10).

Identification of pressure and temperature conditions conducive to hydrate formation is a mature technology, with abilities to accurately model impact of dissolved salts and inhibitors available in many commercial simulation packages (Wasden, 2003).

In this chapter, the flow assurance aspect of the hydrate formation inside pipelines is presented. A detailed description on their thermodynamic properties is given along with a presentation on the procedure of agglomeration and deposition on the pipeline. In the final chapter the possible use of gas hydrates as hydrocarbon reserve is given along with the environmental danger that their production poses.

2.1 GAS HYDRATES AS A FLOW ASSURANCE ISSUE

In 1999 flow assurance was ranked at the top of major technical problems during offshore developments by 110 energy companies (Welling et. al, 1999) (Macintosh, 2000). The importance of flow assurance problems listed after decreasing significance are: Hydrates, wax, scale, corrosion and asphaltenes. The level of importance varies over the world, but in the Gulf of Mexico hydrates are a much larger concern than any of the other ones.

Hydrates can form onshore or offshore in a relatively short period of time, plug up lines quickly, and disrupt production. Within flowlines, hydrates can form along the walls or can flow through the production, forming solid slurry (Borden, 2014). The hydrate problem is addressed during normal production (steady conditions) and during shut-in and start-up (transient conditions). These issues are considered below analytically.

Normal production

The obvious constraint that arises for normal production is to maintain tubing head temperatures and pressures at sufficient values to avoid hydrate formation. Minimum rates may be calculated that correspond to hydrate formation conditions at the tubing head. While these considerations typically do not drive design, the condition is worth documenting, particularly for early life when care for the completion may drive low start-up rates over long durations.

During shut-in

This constraint requires that the production rate be maintained such that well fluids remain outside the hydrate formation region during the cool-down period (defined later). When this requirement is not met, alternate operating procedures may be effective at mitigating risk.

Following shut-in

The immediate concern following unplanned shut-in should be determining the cause of the shut-in and understanding how operations may be quickly resumed.

In the case where production fluids are continuously treated for hydrate management, operations staff may do this without further consideration. Where continuous chemical treatment is not used it is convenient to build into the design of the well a period of time following the shut-in, when the equipment is allowed to cool (without entering the hydrate formation region) without the need for operator intervention. This period is often termed the “no-touch” time. Often, the design allows that a restart within the no-touch time requires little of the additional procedures that are required for long-term shut-ins and subsequent start-ups (considered below).

In addition to the “no-touch” period, additional time to perform operations is required to stabilize the well and subsea equipment for a long-term shut-in. These operations could include any treatment or displacement of fluids in the well bore and/or subsea equipment. It is also desirable to allow the fluid to cool during this period without entering the hydrate formation region. The sum of the “no-touch time” and the time required to perform the shut-in operations is termed the “cool-down” period. The specification of the cool-down period is subject to CAPEX/OPEX expense tradeoffs. Shorter cool-down periods (particularly no-touch periods) require more frequent full shut-ins, with implications on availability and operating costs. Longer cool-down periods may require additional insulation on subsea trees or other subsea equipment. Operational experience indicates that relatively short duration shut-ins are common and, thus, designing for a “no-touch” period of several hours is typically justified. The design implication for cool-down is that a “no-touch” time and a “cool-down” period must be defined for systems that are not continuously treated. Insulation configuration decisions are made based on how these times can be achieved. Similarly, operations required following shut-in will follow the schedule limitations defined by these times. A minimum rate can be defined (for normal production) below which the cool-down time for the well and subsea equipment cannot be achieved ([Hudson et al, 2000](#)).

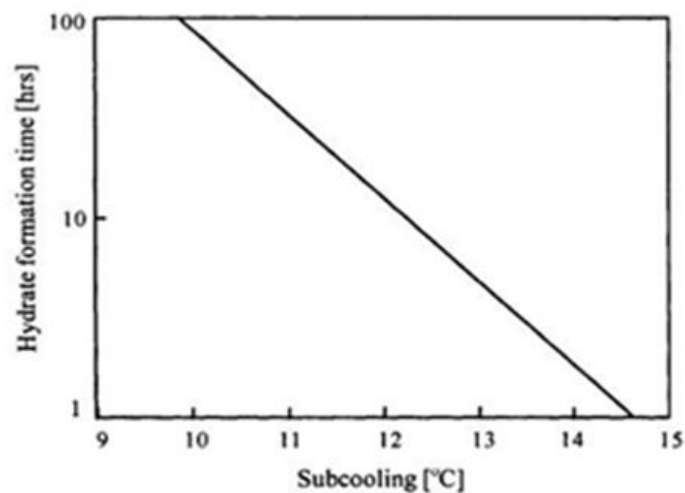


Figure 11: Variation of hydrate formation time with subcooling

Time (hrs)	Activity	
0	shut-in	
3	no touch time	
6	blowdown (gas lift assisted)	Jumper/tree, MeOH injection
12	dead oil displacement	

Figure 12: Unplanned shutdown sequence of events

During Startup

The discussion related to start-up is considered in two parts: “cold-earth” start-ups, and start-up after short or intermediate shut-in durations.

“Cold-earth” start-ups

At the initial start-up or after the well has been shut-in for some period of time, the temperature of the fluids in the well bore and other subsea equipment will approach the ambient conditions. While “live” fluids have either been displaced or treated to prevent hydrate formation following shut-in, as cold fluids are produced into the upper well bore and subsea equipment, there may be some risk of hydrate plug formation. The cold fluids are quickly replaced with warm fluids from deeper in the well, but the cold tubing, casing, and subsea equipment will continue to warm during the initial production period, cooling the fluids that are exposed to these surfaces. Often production fluids are exposed to hydrate forming conditions for some period during start-up. This certainly implies risk in terms of hydrate plug formation; however, this risk may be tolerable due to short durations in the hydrate formation region and limitations to hydrate plug formation

based on plug formation kinetics. Certainly, longer periods of production at higher subcooling values (differences between the hydrate dissociation temperature and actual temperature) indicate higher risk. Thus, slow start-ups at high pressure would typically impose higher risk than fast startups at low pressure. Because of this, hydrate management concerns may compete with completion integrity issues in establishing initial startup rates.

Because the risk of hydrate plug formation during start-up needs to be understood, the duration of exposure to hydrate formation on start-up must be determined. Transient thermal hydraulics calculations for the well will predict the time required for the tubing head temperature to reach the hydrate dissociation temperature (HDT) for a particular production rate or ramp-up schedule. The predicted temperatures and pressures during the period between initiation of production and the time that the HDT is reached will determine the time and subcooling under which the fluids are exposed to hydrate formation conditions (Figure 11). An assessment can then be made, based on operational experience with similar conditions, whether or not this risk is tolerable (Hudson et al, 2000).

Start-up after short/intermediate duration shut-ins

The cold-earth start-up is a convenient, worst-case design scenario in that the well temperature initially follows the geothermal profile and the subsea equipment is initially at ambient temperature. However, it is often the case that such harsh start-up conditions are atypical beyond the initial startup. Indeed, an assessment of risk where every startup is assumed to have the cold-earth start-up risk may lead to an overly conservative design. For this reason, it is often useful to consider start-up from finite duration shut-ins. The most obvious such case is a very short shut-in duration that corresponds to the “no-touch” period defined above. If the exposure to hydrate plug formation risk is small for this startup, production can typically be resumed without the additional measures that are required to prepare the well and subsea equipment for longer shut-ins. Similarly, start-up measures are also typically avoided.

It may also be useful to consider start-up from intermediate shut-in durations. An example of this could be when “cold-earth” start-ups imply significant risk, but are expected to occur rarely, if ever, in life. In these cases, consideration of intermediate shut-in durations can provide additional guidance in operating procedures, indicating that start-up after periods longer than that considered “intermediate”, could be subject to greater risk and/or be required to follow different procedures (Hudson et al, 2000).

In addition to flow assurance issues, potential problem of hydrate formation during Petroleum industry operations can occur while drilling. When drilling through natural

deposited gas hydrates they may cause wellbore instability and other drilling hazards such as: severe mud gasification, low quality logging and cementing, casing collapse due to high pressure, gas accumulation behind the casing, casing subsidence due to Natural gas hydrates base sediment failure and consequently instability of the ocean floor, drilling fluid gasification, wellbore instability (wall diffusion), casing running difficulties, wellbore widening, formation failure, casing subsidence and personnel health risks. These are examples of dangers which may occur during drilling operations in hydrate bearing sediments in deep waters. Another concern is the risk of hydrate dissociation during surface casing cementing. Negative effects on cement quality may result if gas bubbles flow into fresh cement. Escaped gas from dissociated hydrates may find its way to the sea floor, or to the other formations which may bring hazardous consequences to personnel and the drilling platform (Fereidounpour, 2014).

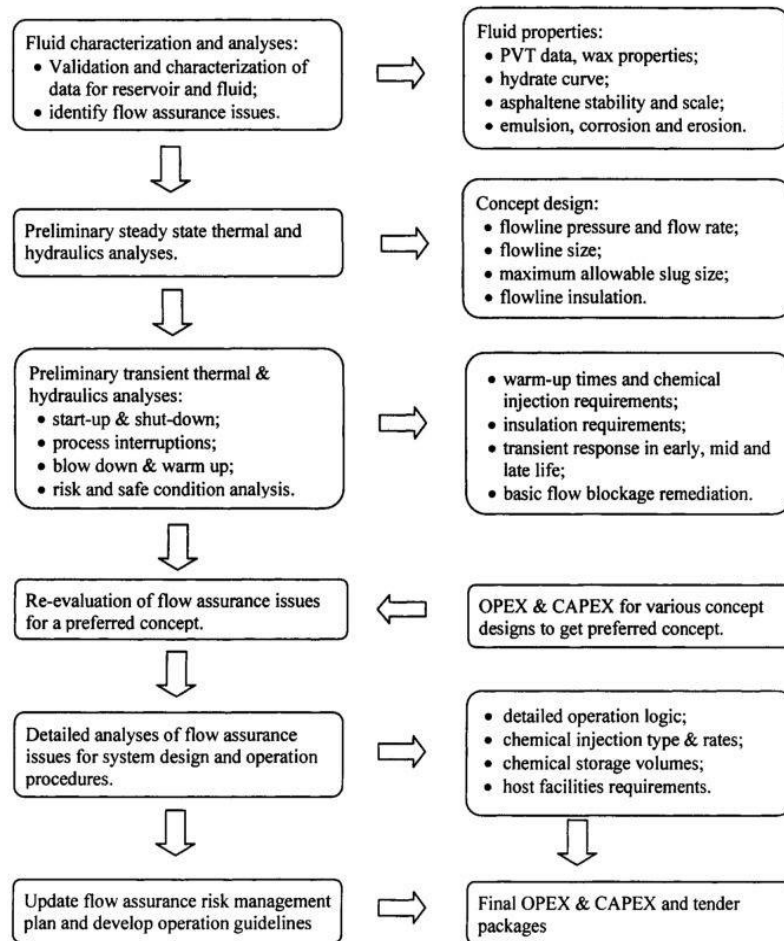


Figure 13: Production system design flowchart for Flow assurance (Carroll 2003)

2.2 THERMODYNAMIC VIEW ON GAS HYDRATES

In the petroleum industry, the term hydrate is reserved for substances that are usually gaseous at ambient conditions. The common chemical compounds involved into the hydrocarbon production include methane, ethane, carbon dioxide, and hydrogen sulfide. This leads to the term gas hydrates and to one of the popular misconceptions regarding these compounds. It is commonly believed that non-aqueous liquids do not form hydrates. However, liquids may also form hydrates (Carroll, 2003).

When natural gas and free water together are subjected to high pressures and relatively low temperatures hydrates may form. In the early days of the gas industry this was not known. Not until the natural gas expansion in the 20th century, when the gas was transported under high pressures that the first experience with hydrates in pipelines and processing equipment occurred. Hammerschmidt demonstrated first in the 1930s that the ice found blocking pipes actually was gas hydrates. His argument was that the temperature was not sufficiently low for water to freeze (Carroll, 2003).

In the oil industry, hydrate formation may occur during drilling operations such as collapsed tubing, in water-based mud, during production of target fluids or pipeline foundation damage in situ hydrated mounds, back-flush operations in water injection, oil and gas processing as well as in chemical inhibition umbilicals. Offshore production operations could encounter the hydrate stable region as illustrated in Figure 14 (red line from subsurface to surface facilities) when temperature and pressure changes and prior analysis is important to avoid hydrate formation (Mogbolu et. al, 2014).

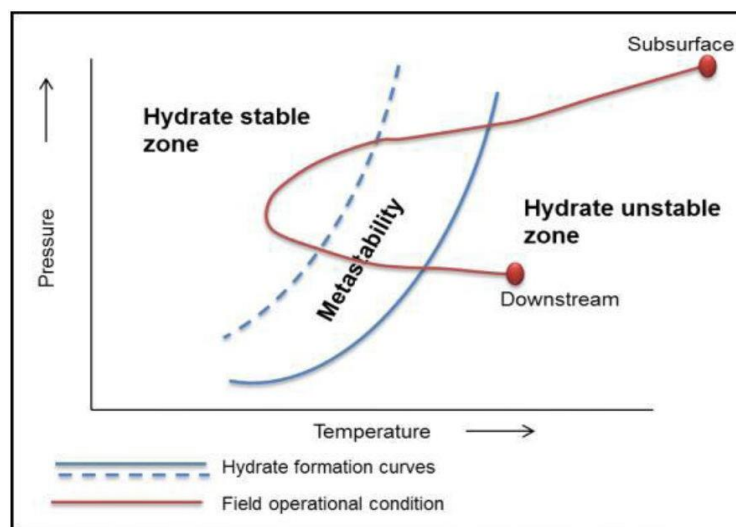


Figure 14: Pressure-Temperature diagram of an offshore production line and hydrate stability zone

2.2.1 WATER AND HYDROGEN BONDS

It is the structure of the water molecule that creates the foundation for hydrate formation. The water molecule consists of one oxygen atom and two hydrogen atoms. It is polar and has four electrons, but only two of these are shared with the hydrogen atoms. The angle between each line is 109.5° , similar to the methane molecule, CH_4 . The water molecule has two free electrons on the opposite side of where the hydrogen atoms are connected. And the angle between the hydrogen atoms is only 104.5° .

The induced charges on the water molecule that result in hydrogen bonding and the angle between the hydrogen atoms are showed in [Figure 15](#) (Carroll, 2003). This can be explained in a simplified way as the pair of free electrons repulses each other and the hydrogen atoms with a larger force than the hydrogen atoms repulse each other. The free electrons induce a negative charge on the oxygen molecule and a weak positive charge on the hydrogen atoms.

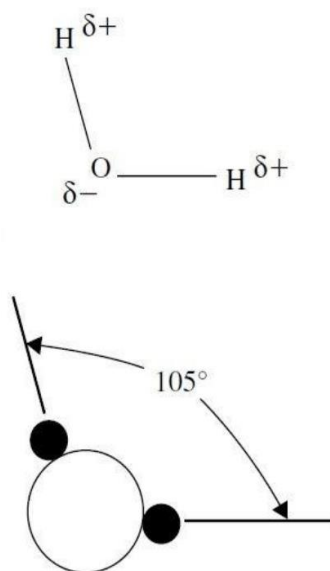


Figure 15: Shape of the water molecule

The hydrogen atoms share electrons with the oxygen atoms. These binding forces are very strong and are called covalent bonds. Since the water molecule is polar, the negative side will be attracted to another water molecule's positive side. This causes each of the hydrogen atoms to attract a new water molecule. These bonds are called hydrogen bonds and are stronger than van der Waals forces, which connects regular non-polar molecules.

Hydrogen bonds are electro static forces. They are strong and explain the special properties of water compared to other molecules that consist of elements from the same area of the periodic table. Elements that possess similar properties are organized together in the periodic table. The water molecule stands out from the others with its high boiling temperature. It also requires much more energy to break up the net of water molecules when the water is boiled.

2.2.2 *HOST AND GUEST MOLECULES*

Hydrogen bonds are the reason that water can form hydrates. Hydrogen bonds cause the water molecules to organize in specific patterns. The presence of some compounds can cause these structures to stabilize and cause solids to precipitate.

The water molecules are often referred to as the host molecules, while the stabilizing compounds are called guest molecules. The water molecules form three dimensional cages with complex geometry and room for guest molecules.

Van der Waals forces between the guest molecule and the water molecules are thought to stabilize the cage. Van der Waals forces are attraction between molecules caused by other things than electro static forces. The guest molecule is not tied to the host molecules, and has space to rotate freely inside the cage. That is why these components are best described as solid-solution.

2.2.3 *HYDRATE FORMING CONDITIONS*

The formation of hydrates requires the combination of three critical criteria:

- The right combination of pressure and temperature (High pressure and low temperature)
- The presence of hydrate formers (for example methane, ethane and carbon dioxide)
- The presence of sufficient amount water.

The hydrate formation temperature is very dependent on the gas composition, but is always higher than the freezing point of water, 0°C. Due to the requirements mentioned above it may seem trivial to avoid hydrate formation. By removing one condition is clear that hydrates will not form, but in reality it is not easy. The hydrate formers are the gas that the energy companies are after. The focus in the natural gas industry is therefore on the other two forming conditions requirements. Two phenomena that enhance the growth of hydrates are turbulent flow and nucleation sites.

Hydrates form more rapidly in places of high velocity like choke valves and other flow restriction spots. The diameter reduction in those areas causes the gas to accelerate. Mixing of water and hydrocarbons in flowlines, process vessels, etc increase also the rate of hydrate formation.

Nucleation sites can in general terms be described as a point where phase transition is favored. In the case of hydrates; the formation of a solid from a fluid phase. Fine nucleation sites for hydrate formation can be an imperfection in the flowline, a weld spot, or a flowline fitting (elbow, tee, valve, etc.). In addition scale, sand and other solid depositions all make good nucleation sites. The gas-water transition provides a good nucleation site as well. The points above are not required for hydrate formation, but will increase speed of the deposition.

The accumulation of hydrates does not necessarily occur at the same place as they are formed. Hydrates may be carried along with liquid phase. They tend to accumulate at same locations as the liquid. A typical place is at the bottom of a U-shapes pipe. This can block the pipe and cause damage to equipment and endanger the safety of humans and the environment.

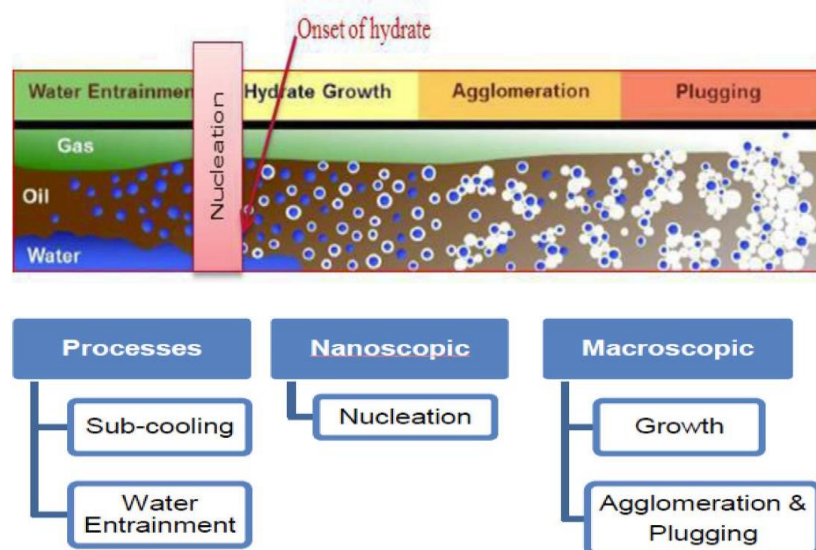


Figure 16: Steps of hydrate formation that leads to plugging (Sloan & Koh, 2008)

2.2.4 HYDRATE STRUCTURE TYPES AND FORMERS

There are three different structure types of hydrates; Type 1, Type 2 and Type H (Figure 18). They are classified by how the water molecules are arranged in the lattice/crystal. In the oil and gas industry it is most common to see hydrates of type 1 and type 2. Table 1 shows a comparison of the different hydrate types.

Table 1: Comparison of Type 1, Type 2 and Type H Hydrates (Carroll, 2003)

	Type I	Type II	Type H
Water Molecules per Unit Cell	46	136	34
Cages per Unit Cell			
Small	6	16	3
Medium	—	—	2
Large	2	8	1
Theoretical Formula [†]			
All cages filled	$X \cdot 5 \frac{3}{4} \text{H}_2\text{O}$	$X \cdot 5 \frac{2}{3} \text{H}_2\text{O}$	$5X \cdot Y \cdot 34 \text{H}_2\text{O}$
Mole fraction hydrate former	0.1481	0.1500	0.1500
Only large cages filled	$X \cdot 7 \frac{2}{3} \text{H}_2\text{O}$	$X \cdot 17 \text{H}_2\text{O}$	—
Mole fraction hydrate former	0.1154	0.0556	—
Cavity Diameter (Å)			
Small	7.9	7.8	7.8
Medium	—	—	8.1
Large	8.6	9.5	11.2
Volume of Unit Cell (m ³)	1.728×10^{-27}	5.178×10^{-27}	
Typical Formers	CH ₄ , C ₂ H ₆ , H ₂ S, CO ₂	N ₂ , C ₃ H ₈ , i-C ₄ H ₁₀	See text

[†] Where X is the hydrate former and Y is a Type H former.

The simplest hydrate structure is Type 1. It is composed of two types of cages:

- Dodecahedron, a twelve sided polyhedron where each surface is a regular pentagon.
- Tetrakaidecahedron, a fourteen sided polyhedron with twelve pentagonal surfaces and two hexagonal surfaces. The dodecahedron cages are often referred to as the small cages because they are smaller than the tetrakaidecahedron cages. For the opposite reason the tetrakaidecahedron cages are called the big cages.

Type 1 hydrates consist of 46 water molecules. The theoretical formula for the number of water molecules in a type 1 hydrate is $X \cdot 5 \frac{3}{4} \text{H}_2\text{O}$, where X is the guest molecule (Carroll, 2003). Hydrates are non-stoichiometric, that means that not all the cages need to have a guest molecule to make the hydrate stable. The saturation amount is a function of pressure and temperature. This means that the real hydrate

composition is not always equal to the theoretical. Methane, ethane, carbon dioxide and hydrogen sulfide are common type 1 formers. With the exception of ethane, which can only occupy large cages, the other ones can occupy both cages (Figure 17).

The structure of type 2 hydrates is much more complex than of that of type 1. It consists of two cage types:

- Dodecahedron
- Hexakaidecahedron, a sixteen sided polyhedron with twelve pentagonal sides and four hexagonal sides.

Type 2 consists of 136 water molecules. The theoretical formula for the maximum number of water molecules is $X \times 5 \frac{2}{3} \text{H}_2\text{O}$. If only the large cages are filled then the formula is $X \times 17 \text{H}_2\text{O}$. Nor this hydrate type is stoichiometric, and the real composition will deviate from the theoretical. Common type 2 formers are iso-butane, nitrogen and propane. Nitrogen can occupy both the small and large cages, while iso-butane and propane can only occupy the large cages (Figure 17).

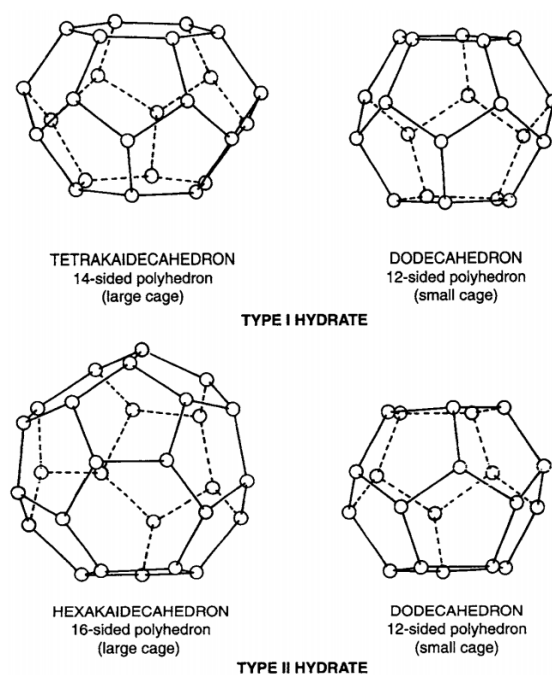


Figure 17: Type I and type II hydrate cages (Carroll, 2003)

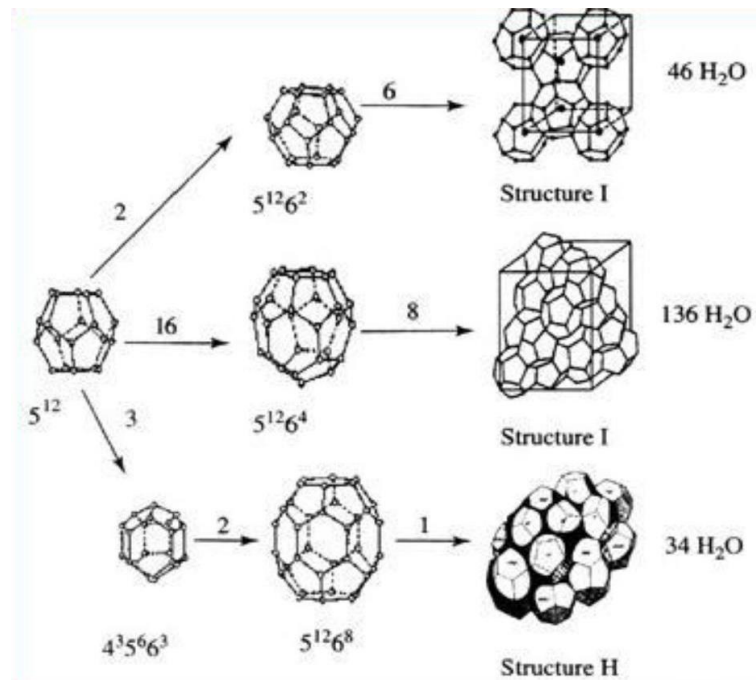


Figure 18: Hydrate structure types (Carroll, 2003)

Von Stackelberg discovered that there is a relationship between the size of the guest molecule and type of hydrate formed. He made an overview that shows which hydrate type are formed by guest molecules of increasing size (Figure 19: Comparison of Guest Size, Hydrate Type, and Cavities Occupied for Various Hydrate Formers (Stackelberg, 1949)). The molecule size is given in Ångström: $1 \text{ Å} = 1\text{E-}10 \text{ meter}$).

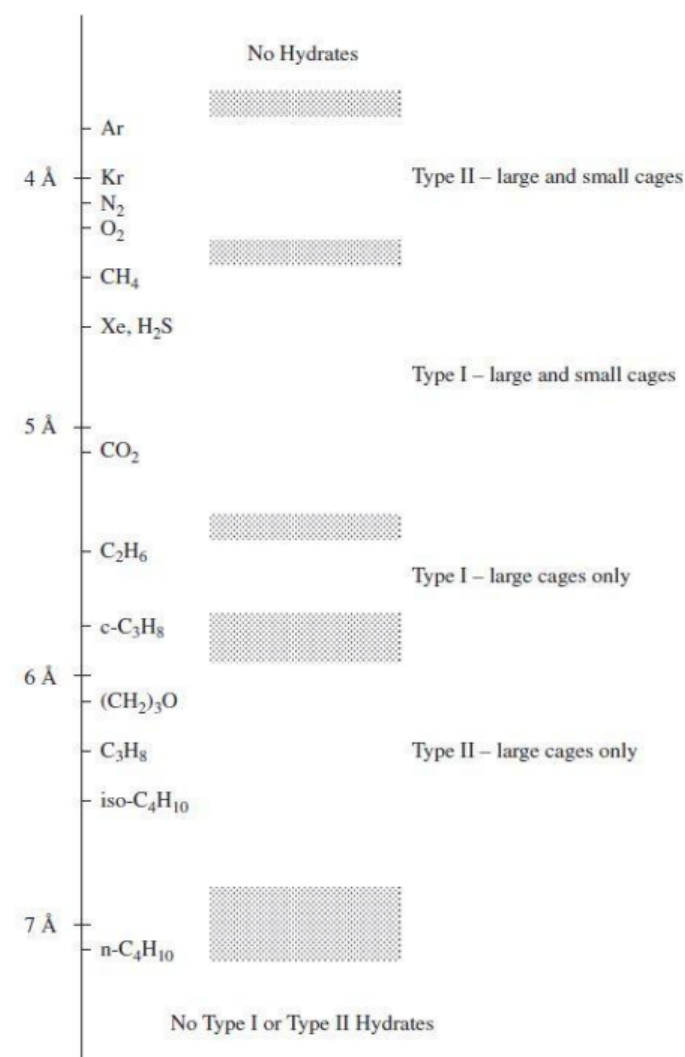


Figure 19: Comparison of Guest Size, Hydrate Type, and Cavities Occupied for Various Hydrate Formers (Stackelberg, 1949)

The classes of hydrate formers are the following:

- Molecules with smaller diameter than 3.8 Å which do not form hydrates.
- Molecules with diameters between 3.8 and 4.2 Å which are small enough to enter both small and large cages and form type 2 hydrates.
- Molecules with diameter between 4.4 Å and 5.4 Å (Carbon dioxide, hydrogen sulfide and methane) which they form hydrates of type 1 and are small enough to occupy both cage sizes.
- Molecules from 5.6 Å to 5.8 Å like ethane. They form type 1 hydrates, but can only fit in the large cages.
- Molecules with diameter from 6.0 Å to 6.9 Å, like iso-butane and propane form type 2 hydrates, and can only fit inside the large cages.
- Molecules larger than about 7 Å do not form either a Type I or Type II hydrate.

Therefore, molecules such as pentane, hexane, and larger paraffin hydrocarbons are non-formers.

From the chart, we can see that cyclopropane ($c\text{-C}_3\text{H}_8$) and n-butane are in the hatched regions. These special components are discussed in more detail later. Slightly larger molecules can form Type H hydrates, but the maximum size for these compounds to form a hydrate is about 9 Å (Carroll, 2003).

N-butane is a transition molecule. Molecules larger than n-butane do not form type 1 and type 2 hydrates, but smaller do. What makes n-butane so special is that it does not form hydrates alone, but in the presence of another hydrate former it can occupy cages/a cage.

There are other types of hydrocarbon that are sufficiently small to form hydrates. Compounds like acetylene, ethylene, propylene, and propyne are hydrate formers under conditions. Cyclopropane can form both type 1 and type 2 hydrates. The type of hydrate it forms is dependent on pressure and temperature.

2.2.5 CHEMICAL PROPERTIES OF POTENTIAL GUEST MOLECULES

The right molecule size is not adequate enough for a molecule to be hydrate former. It has to possess the right chemical properties. Components easily soluble in water usually do not form hydrates. Carbon dioxide and hydrogen sulfide are fairly soluble in water and however form hydrates. These molecules can, as a rule of thumb, be thought of as being in the transition when it comes to solubility in water.

Hydrates will not also form if the molecule interferes with the hydrogen bond. The small molecule methanol is an example of this. Its own hydrogen bond interferes with the hydrogen bond in the water molecules. Methanol is also very soluble in water. Methanol plays an important role in the oil and gas industry as a hydrate inhibitor.

2.2.6 LIQUID HYDRATE FORMERS

Opposite to the common belief that Hydrates are only formed from gaseous hydrocarbons it is observed that they can also be formed by liquid hydrocarbons. The only thing that matters is whether or not the three hydrate forming factors are present; hydrate formers, sufficient amount of water, and the right combination of pressure and temperature. The phase of the fluid is not a limiting factor. The confusion may be because hydrates only form from light components which are more common in natural gas.

There have been done experiments with all the common components in natural gas to find out when they are forming hydrates. Sloan has collected all the results in a book (Sloan, 1998). A pressure-temperature table for methane has been included to give an example for the forming conditions (Table 2). Tables for other natural gas components can be found easily in literature.

Table 2: Hydrate Forming Conditions of Methane

Temp. (°C)	Press. (MPa)	Phases	Composition (mol %)		
			Aqueous	Vapor	Hydrate
0.0	2.60	L _A -H-V	0.10	0.027	14.1
2.5	3.31	L _A -H-V	0.12	0.026	14.2
5.0	4.26	L _A -H-V	0.14	0.026	14.3
7.5	5.53	L _A -H-V	0.16	0.025	14.4
10.0	7.25	L _A -H-V	0.18	0.024	14.4
12.5	9.59	L _A -H-V	0.21	0.024	14.5
15.0	12.79	L _A -H-V	0.24	0.025	14.5
17.5	17.22	L _A -H-V	0.27	0.025	14.5
20.0	23.4	L _A -H-V	0.30	0.027	14.6
22.5	32.0	L _A -H-V	0.34	0.028	14.6
25.0	44.1	L _A -H-V	0.37	0.029	14.7
27.5	61.3	L _A -H-V	0.41	0.029	14.7
30.0	85.9	L _A -H-V	0.45	0.029	14.7

Notes: Composition for aqueous phase and for the hydrate is the mole percent of the hydrate former (CH₄). For the vapor, the composition is the mole percent water.

The table for methane is limited to 30 °C, possibly because the hydrate formation pressure at this temperature is 85.9 MPa (12500 psia). A pressure not exceeded in regular petroleum operations. The values in the table are among others presented in a plot shown in Figure 20. In every case, the three-phase loci involving two liquid

phases are very steep. That means small changes in temperature have dramatic effect on the pressure. As seen from the figure the methane does not have such a locus.

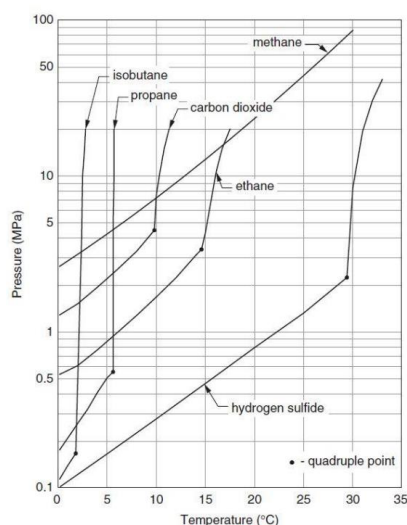


Figure 20: Hydrate Loci for Several Compounds Found in Natural Gas (Sloan, 2009)

For the purpose of comparing, a common data base exists for several hydrate formers. The temperature variable was eliminated by using the hydrate pressure at 0 °C as reference. Carroll has presented the hydrate pressure for multiple components at 0°C together with several of their physical properties in Table 3.

Table 3: Physical Properties and Hydrate Formation of Some Common Natural Gas Components (Carroll, 2003)

	Hydrate Structure	Molar Mass (g/mol)	Hydrate Press. at 0°C (MPa)	Normal Boil. Point (K)	Density (kg/m ³)	Solubility (mol frac × 10 ⁴)
methane	Type I	16.043	2.603	111.6	19.62	0.46
ethane	Type I	30.070	0.491	184.6	6.85	0.80
propane	Type II	44.094	0.173	231.1	3.49	0.74
isobutane	Type II	58.124	0.113	261.4	3.01	0.31
acetylene	Type I	26.038	0.557	188.4	6.70	14.1
ethylene	Type I	28.054	0.551	169.3	7.11	1.68
propylene	Type II	42.081	0.480	225.5	9.86	3.52
c-propane	Type II	42.081	0.0626	240.3	1.175	2.81
CO ₂	Type I	44.010	1.208	194.7 ²	25.56	13.8
N ₂	Type II	28.013	16.220	77.4	196.6	0.19
H ₂ S	Type I	34.080	0.099	213.5	1.50	38.1

A rule of thumb says that if the mixture only consists of guest molecules that form the same type of hydrate, that hydrate type will be formed. A mixture of carbon dioxide, hydrogen sulfide and methane, all type 1 formers, will form type 1 hydrate. However the hydrates behavior may vary in different situations.

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2.2.8 AZEOTROPY

A very interesting phenomenon is azeotropic hydrates. These forms either at lower or higher pressure compared to the pure components. An example from Natural Gas Hydrates (Carroll, 2003) is the hydrate that is formed from a mixture of hydrogen sulfide and a sweet natural gas mixture. This mixture forms hydrates at higher temperature than the pure components (Figure 22). The opposite behavior is exhibited by other sour gas components like CO₂ in a gas mixture. In that case it forms hydrate in lower temperatures for the same pressure conditions (Figure 22).

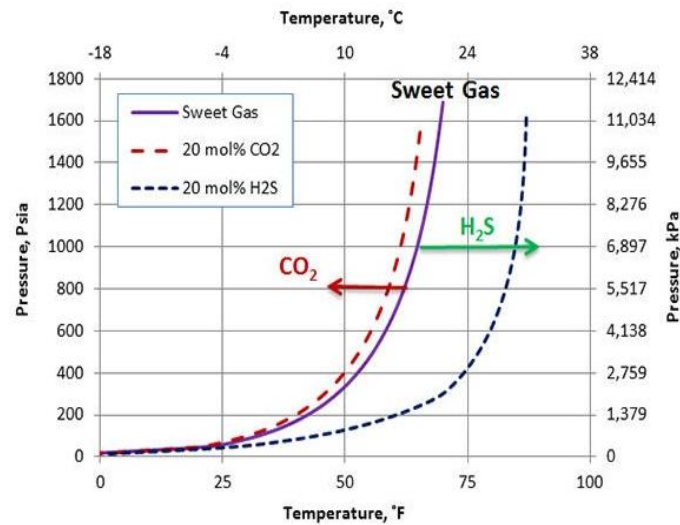


Figure 22: Effect on inorganic substances on hydrate phase envelope

2.2.9 HYDRATE FORMATION DUE TO EXPANSION THROUGH RESTRICTIONS OR VALVES

When water saturated gas flows through a valve or a flow restriction, it causes a rapid pressure drop. The expansion that occurs is adiabatic (the change in enthalpy is equal to zero, $\Delta H = 0$). For the enthalpy to remain the same on the other side of the valve the temperature must be reduced. This is called the Joule-Thomson effect. When the temperature decreases, the water condenses and the basis for hydrate formation is present. The expansion of two gases with gravity of 0.6 is showed in [Figure 23](#). The reduction in pressure causes a temperature reduction that leads both gases into the hydrate region.

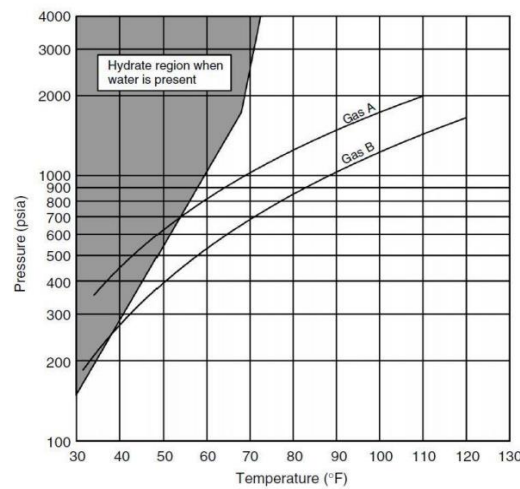


Figure 23: Expansion of Two Gases into Hydrate Formation Region (Sloan & Koh, 2008)

Generally the upstream pressure and temperature is known, and downstream pressure can be found if the pressure drop across the flow restriction is known. The cooling curves in [Figure 23](#) are designed for constant enthalpy (Joule-Thomson) expansion. They are taken from the first law of thermodynamics for systems that flow in the steady-state ($\Delta H = Q + W_s$), where one disregards the change in kinetic and potential energy.

2.2.10 HYDRATE FORMATION IN OIL DOMINATED PIPELINES

A conceptual illustration of the hydrate formation in an oil dominated system is shown in Figure 24. The formation is depicted in six steps.

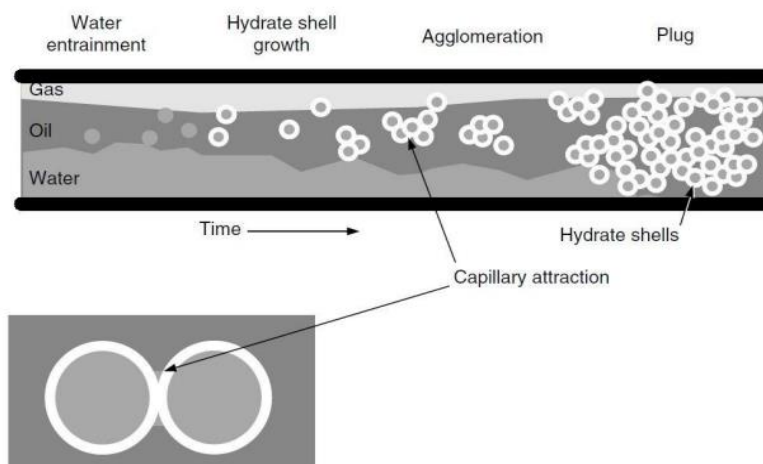


Figure 24: Conceptual Figure of Hydrate Formation in an Oil-Dominated System (Sloan & Koh, 2008)

In the first stage the water phase emulsifies into the oil phase. As a rule there is less water than oil. The water droplets are typically a couple of microns across.

In the second stage a thin hydrate starts to form on the outside of the water droplets (maybe even less than 6 microns thick). In the beginning the particles are very malleable. Whilst the particles still are malleable they form a diffusional barrier between the oil phase and the water phase. Usually the shell does not grow very thick; however it may if there is enough time for the hydrates to grow.

The droplets are drawn together by capillary forces. These forces have varying strength, which is dependent on the temperature. The magnitude of the forces is reduced when the temperature falls (Taylor, 2006).

The accumulation of particles causes an increase in the apparent viscosity. Hydrate structures that breaks down can be identified as spikes in pressure drop measurements. In the end the accumulation the hydrate particles grows large, which results in a large pressure drop that will stop the flow. This is at this point where the

hydrate plug is formed. With time the porosity and permeability of the plug is reduced due to particle growth and exposure in high pressure (Borden, 2014).

Agglomeration of hydrate particles is, indicated by the figure, the limiting factor for plug formation. This has made scientists wondering if it possible to prevent the particles from agglomerating. And the result of these ideas is anti-agglomerates, which make it possible for the oil to transport the hydrate particles as slurry.

2.2.11 HYDRATE FORMATION IN GAS DOMINATED PIPELINES

The amount of liquid hydrocarbons is much smaller in gas dominated systems. And for that reason the previous concept of water in oil emulsions is not valid for gas systems. Figure 25 is divided into two parts. The upper part is the depiction of the pressure drop upstream, before a hydrate plug is formed as a function of time. The pressure scale is semi-logarithmic. The lower part of the figure depicts stepwise the chain of events when a hydrate plug is formed in a gas dominated system. The upstream pressure response corresponding to each step is illustrated in the upper part of Figure 25.

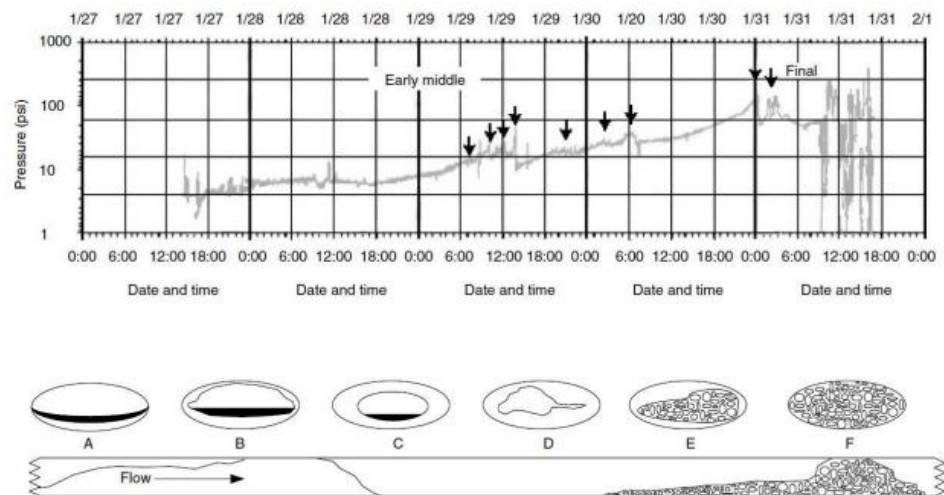


Figure 25: Hydrate Blockage Formation (bottom) and Corresponding Pressure Build-up (top) in A Gas-Dominated Pipeline (Sloan & Koh, 2008)

To make it easier to follow the description of the chain of events, each stage has been given a letter. Water in the flowline originates from the reservoir either as produced formation water, or as condensed vapor (point A). Hydrates are usually formed in water, condensed or splashed, at the pipe wall (B). In a hot gas stream the temperature falls radially from the center of the pipe to the wall. The pipe wall has the lowest temperature due to heat exchange with surroundings. The inner pipe diameter is reduced when hydrates accumulates at the pipe wall (C). They are deposited unevenly on the wall, and this causes irregular pipe diameter and this result in an increased frictional pressure drop.

After a while the hydrate accumulations breaks from the wall due to its own weight and the stress caused by passing flow (E). The event can be recognized as pressure reduction on the reading. With time the concentration of broken wall accumulations grows large in the liquid fraction and the hydrates start to gather in lumps. Eventually they will plug the flow line (F) (Borden, 2014).

2.3 GAS HYDRATES AS HYDROCARBON RESERVES

Production from gas hydrate deposits can be traced back to 1963 when the Markhinskaya well was drilled in Siberia. Gas hydrate formation conditions were recognized from rock sections at 0°C and 1450m. This was first doubted until later proved by Makogon in 1969. Gas has also been exploited from gas hydrates in the Messoyakha field in Siberia upon dissociation of the gas hydrate cap at effective rates. Since then commercial gas exploitation from ocean hydrates has been on the frontier of research and this was recently achieved in March 2013 in the Nankai Trough gas field in Japan, 50km away from the main island. Engineers used a depressurization method that turns methane hydrate into methane gas (Mogbolu et. al, 2014).

Natural gas hydrates have been discovered in permafrost and sea bed and postulated to exist in outer space. The estimated volume of natural methane hydrate deposits is about 120 quintillion cubic meters at standard temperature and pressure, twice that of known fossil fuel resources. Also ocean hydrates volume surpasses that in the permafrost by two orders of magnitude, but permafrost hydrates may be more accessible and frequently have higher concentrations. Hydrocarbons are also concentrated in these gas

hydrates and can be as high as 180 m³ of gas in 1 m³ of hydrate.

Relative to the amount of conventional gas available, methane from hydrates is considered pre-commercial; therefore, industry/government partnerships are required for development. Countries with a high energy demand such as Japan, India, China, and South Korea are mounting large campaigns to develop hydrated energy, the goal for Japan is commercial productivity by 2015 and for the Gulf of Mexico and other permafrost areas the year 2020 (Sloan et al, 2009) (Vedachalam et al, 2015).

Development of unconventional hydrocarbons will help to replace and displace coal based power production with clean methane exploited from shale gas, tight gas, coalbed methane and natural gas hydrates. Natural gas produces less CO₂ per BTU than any other combustible hydrocarbon, which can help countries to attain the target CO₂ emission levels.

Natural gas is the most environment friendly combustible resource, and is also commercially the best to use as an energy gas backup to renewable energy and as a peak spiker fuel; because natural gas electrical generating stations can be started and stopped most rapidly to meet demand (Vedachalam et al, 2015).

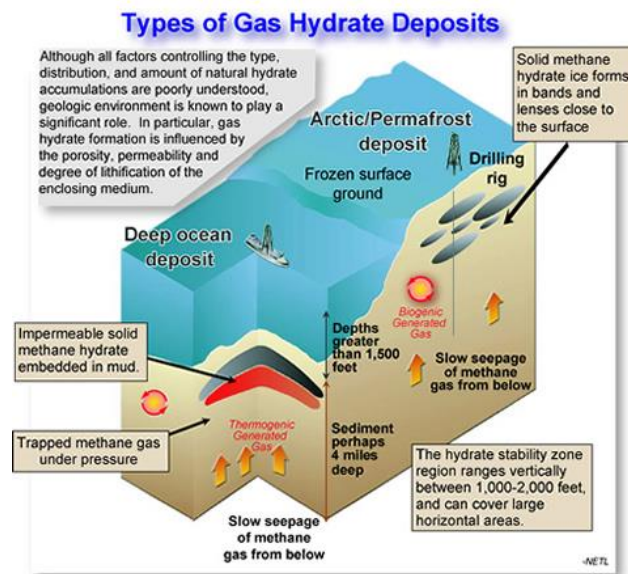


Figure 26: Natural gas hydrates reserves distribution

Arctic - Permafrost Hydrates.

- There are no great technical deterrents to recovery of energy.
- Hydrates have been produced for short periods in the 2007 Mt. Elbert well ([Boswell et al. 2008](#)), and in the 6-day 2008 Mallik depressurization, which had average flow rates of 70 Mscf/D, with peak rates as high as 160 Mscf/D ([Grace et al. 2008](#)).
- Hydrates have, as the largest technical concern, wellbore/reservoir mechanical stability during production.
- Hydrates require continuous multiyear production testing to enable reservoir modelers to eliminate transient effects and to assess commercial feasibility ([Kleinberg 2007](#)).
- Hydrates provide an opportunity, during such production tests, for innovative technologies to be assessed such as CO₂ displacement of CH₄.
- Hydrates provide an acceptable place (ease of access, high concentrations at 'sweet spots') for developments that can be transferred to the ocean in the future.
- Hydrates may be recoverable economically, particularly in places where there is access to existing infrastructure ([Walsh et al. 2009](#)).

Marine Gas Hydrates

- These resources are less advanced developmentally than Arctic-permafrost hydrates because gas from marine hydrates has not yet been produced.
- These resources have the major technology challenge of developing a reliable method to find hydrates ([Kleinberg 2007](#)). The common bottom-simulating reflector is a first order detection method, which frequently is unreliable.
- These resources require a multisite drilling expedition for reliable assessment and recovery, which would have a very high expense. The 2006, 113-day offshore Indian hydrate exploratory expedition required USD 36 million. International cooperation is required to share expenses and results.
- These resources as unconventional are two to three times more expensive than conventional offshore gas, when existing infrastructure is unavailable for either ([Walsh et al. 2009](#)).

Over 230 NGH potential deposits have been identified globally (Figure 27). NGH could exist at subsurface depths ranging from about 130 to 1100 m in permafrost regions, and at water depths between 800 and 4000 m, below the mean sea level in offshore continental margins.

Changing the pressure and temperature conditions of the NGH reservoir, results in methane gas dissociation. Various methods for exploiting the gas hydrates such as thermal stimulation, depressurization and inhibitor injection are in the conceptual or field testing stages. However, a suitable technology for extraction on a commercial basis is yet to be achieved in practice, although depressurization appears to be most promising, taking into account the environmental and techno-economic challenges.

Economic considerations, including the mapping of economically extractable NGH locations, production costs, proximity to large energy markets, and available pipeline networks, form the basis of the economic modeling of gas production from NGH and are required to define the baseline economics of production.

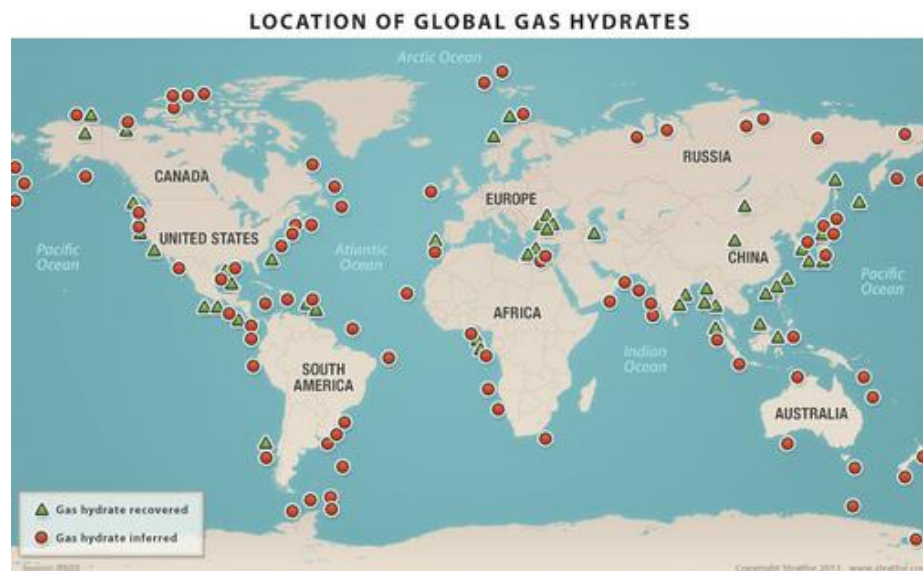


Figure 27: Map of identified Natural Gas Hydrate deposits globally

2.4 GAS HYDRATES AS ENVIRONMENTAL PROBLEM

The environmental challenges of the NGH production include seafloor stability, as uncontrolled hydrate dissociation could result in the subsidence or collapse of the seafloor, (Rutqvist et al., 2009), and excessive release of methane into the atmosphere and absorption by sea water could add to global warming.

Methane is a stable greenhouse gas and thus poses as an environmental threat as a huge volume has the capacity to change the earth's climate and have an increased impact on global warming. Methane is anywhere from 25 to 72 times as powerful an agent of global warming as carbon dioxide, depending on the given time horizon. It is estimated that if the vented methane exceeds 3.2% of the produced methane, natural gas would exceed the same amount of coal as a warming agent (Mogbolu et. al, 2014). The impact on greenhouse effect by in situ gas hydrate has however been determined to be probably inconsequential when small amount of methane is released. Although deliberated methane from natural gas hydrate exploitation in large scale could pose a potential environmental hazard.

Isotopic records support global warming from hydrated methane evolution approximately 600 million years ago. More recently, there is conflicting evidence from analysis of the isotopic record from the late Quaternary. It appears that hydrates may have been relatively stable for the last 10,000 years (Grace et al. 2008). Little is known about methane evolution from hydrates in nature. Methane may be oxidized before reaching the upper atmosphere, but any methane evolution is likely to be chronic, rather than abrupt (Sloan et al, 2009).

3. REMEDIAL SOLUTIONS ON HYDRATE FORMATION AND AGGLOMERATION

The design of oil and gas pipeline systems for deep-water developments is invariably based on the worst-case hydrates formation scenario, which is the cool down period during the production facilities unplanned shut down. To treat such cases, one should be able to predict the hydrate formation temperature of the production fluids which, in turn, depends on the reservoir fluid composition. Calculations become even more complex by further taking into account the fact that the composition of the production fluids along the length of the pipeline varies during both a planned, as well as an unplanned shut down. This is due to the phase slippage between the produced fluids. As a result the hydrate temperature will vary along the length of the pipeline, not only due to the change in pressure but also due to the composition change. Such information is used to optimize the cool down time to the hydrate dissociation point along the length of the pipeline and also to minimize the pipeline insulation requirements. This principle is used to mitigate the hydrate formation potential during an unplanned production system shutdown (Johal, 2007) and it can be achieved through various remedial applications which will be described analytically in this Chapter.

3.1 THERMODYNAMIC HYDRATE INHIBITORS

As already stated many times hydrate formation is a significant problem to the industry. Flow assurance engineers have been searching for solutions to the issue since the development of the first offshore fields. One of the initial solutions, which are still used today, is the use of chemical substances called thermodynamic inhibitors. The inhibitors are usually polar solvents or ionic salts (existing in brine solutions – Figure 28, Figure 29) which they do not prevent the hydrate formation, but instead they reduce the formation temperature or increase the formation pressure (i.e. changing the phase envelope inside which the hydrate is stable).

This way they prohibit the hydrocarbon fluid from entering the hydrate stability zone for the whole length of the pipeline. The most commonly used thermodynamic inhibitors (THI) are methanol and ethylene-glycol (MEG). These are added at high concentrations (10-60 w/w %) and alter the chemical potential of the aqueous or hydrate phase so that the hydrate dissociation curve is displaced to lower temperatures or higher pressures (Figure 30, Figure 31) (Kelland, 1995).

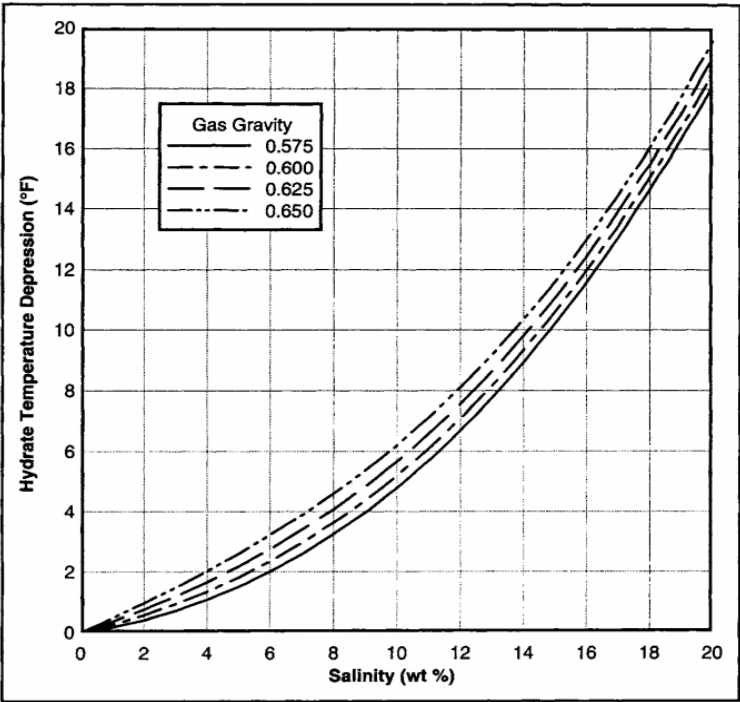


Figure 28: Depression of hydrate temperature due to brine (Carroll, 2003)

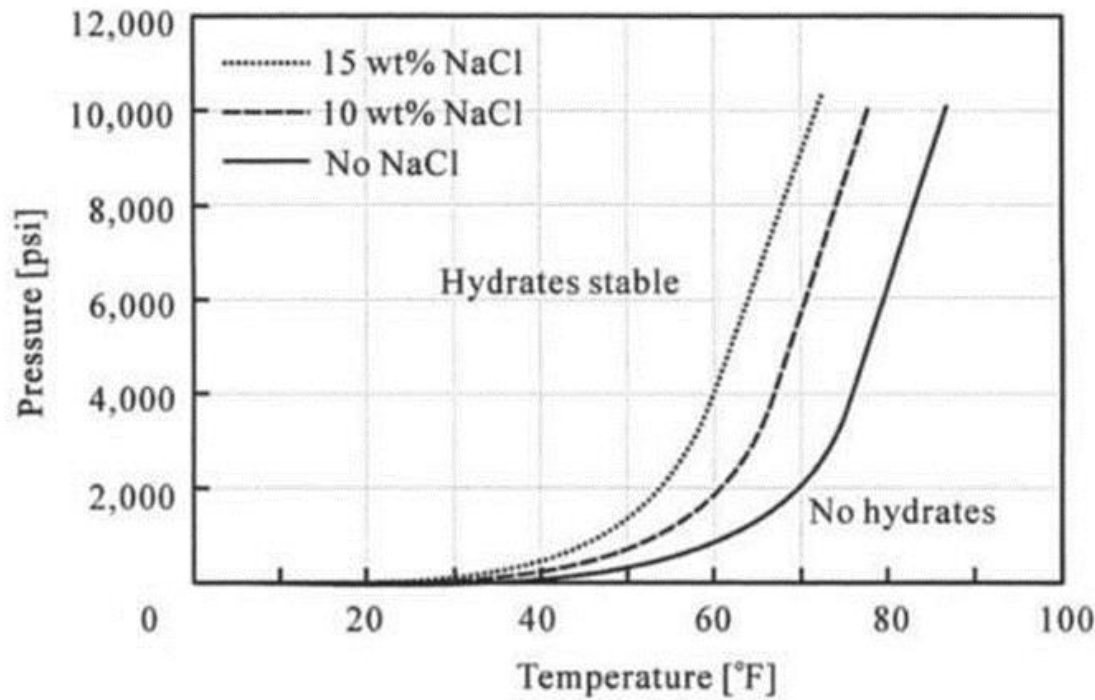


Figure 29: Hydrate dissociation temperature reduction with various ionic salt concentrations (Carroll, 2003)

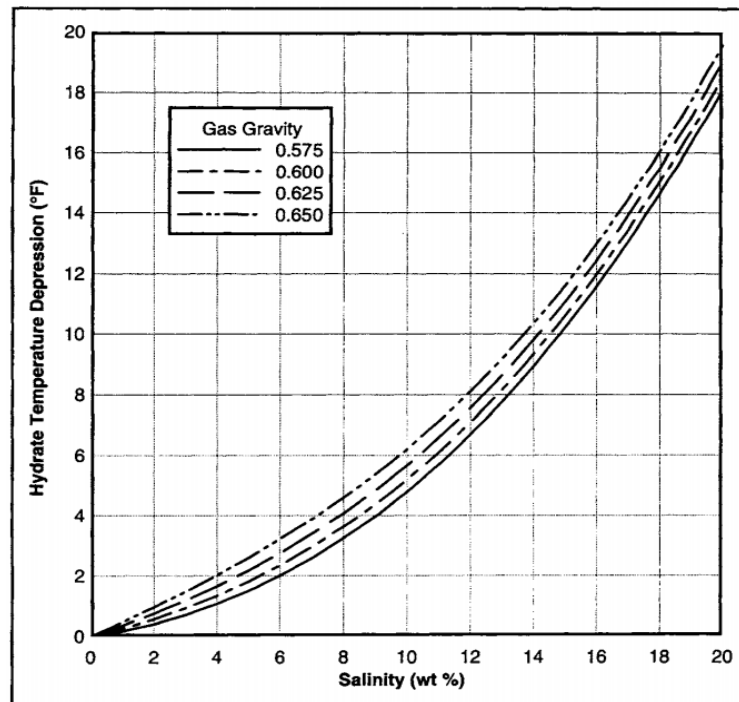


Figure 30: The inhibiting effect of methanol on the methane hydrate (Carroll, 2003)

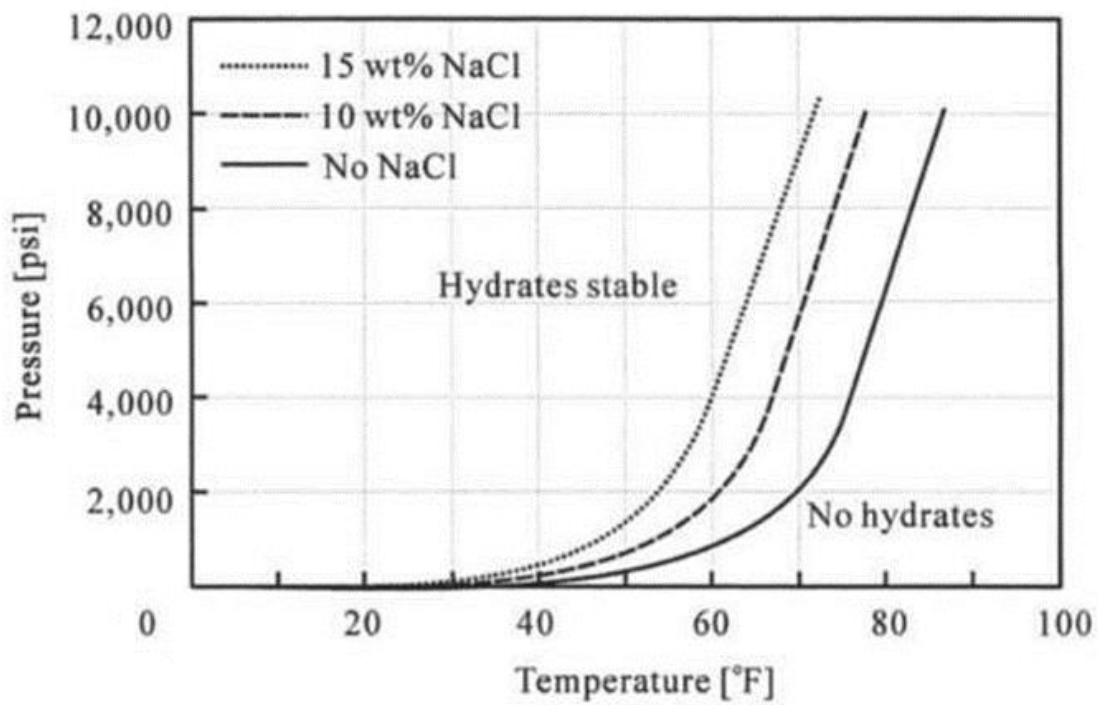


Figure 31: The inhibiting effect of ethylene glycol (MEG) on the methane hydrate (Carroll, 2003)

Scientific knowledge of hydrates has significantly advanced in the last ten years. Simulators predict the hydrate propensity against the design options. We can predict hydrate disassociation within one to three degrees with the exception of brines that have high salt concentration. The hydrate disassociation curves typically provide conservative limits for hydrate management design. The effects of thermodynamic hydrate inhibitors, THIs, such as methanol and ethylene glycols can be predicted with acceptable accuracy (Figure 32).

As a result, methanol and glycols are used as the primary hydrate inhibitors. Methanol and glycol recovery units today are incrementally improving the basic technology. However, the units require appreciable heat to recover the THI.

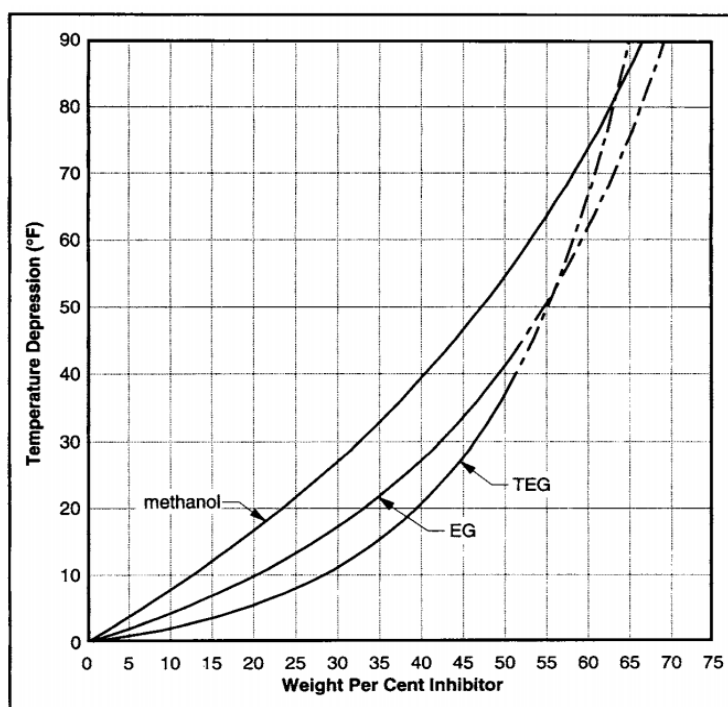


Figure 32: Comparative diagram of inhibiting effect of MeOH, MEG, TEG (Carroll, 2003)

Glycol recovery units can be designed to remove the salts that have traditionally limited the glycol quality. To reduce the methanol and glycol to the needed oil quality target, crude washing requires large volumes of water that must be treated to the seawater injection quality. The recovery units and wash units have a significant footprint, weight, and operability impact on the project and operations design. These units are large and heavy. The units are designed against one of the greatest unknowns in our industry and projects, water rates and water chemistry.

Table 4: Properties of some hydrate thermodynamic inhibitors (Carroll, 2003)

	Methanol	Ethanol	EG	TEG
Empirical formula	CH ₄ O	C ₂ H ₆ O	C ₂ H ₆ O ₂	C ₆ H ₁₄ O ₄
Molar mass, g/mol	32.042	46.07	62.07	150.17
Boiling point, °C	64.7	78.4	198	288
Vapor press. (at 20°C), kPa	12.5	5.7	0.011	<0.001
Melting point, °C	-98	-112	-13	-4.3
Density (at 20°C), kg/m ³	792	789	1,116	1,126
Viscosity (at 20°C), cp	0.59	1.2	21	49
EG = ethylene glycol, HO-CH ₂ -CH ₂ -OH. TEG = triethylene glycol, HO-CH ₂ -CH ₂ -O-CH ₂ -CH ₂ -O-CH ₂ -CH ₂ -OH.				

	Advantages	Disadvantages
Methanol	<ul style="list-style-type: none"> • move hydrate formation temperature more than MEG in a mass basis; • less viscous; • less likely to cause salt precipitation; • relative cost of regeneration system is less than for MEG; • approximate GoM cost of 1.0 \$/gal; 	<ul style="list-style-type: none"> • losses of methanol to gas and condensate phases can be significant, leading to a lower recovery (<80%) ; • impact of methanol contamination in downstream processing; • low flash point; • environmental limitation on overboard discharge.
MEG	<ul style="list-style-type: none"> • easy to recover with recovery of 99%. • low gas & condensate solubility • approximate GOM cost of 2.5\$/gal 	<ul style="list-style-type: none"> • high viscosity, impacts umbilical and pump requirements; • less applicable for restarts, stays with aqueous phase at bottom of pipe; • more likely to cause salt precipitation.

Figure 33: Advantages and disadvantages of commonly used THI's (Methanol – MEG) (Carroll, 2003)

The amount of the inhibitor needed is case specific and depends greatly on water cut. In some cases is reported to reach numbers as high as 100% of the produced water. Methanol can reach even higher values due to its increased volatility that leads to entering in both the aqueous and gaseous phase of the fluid. It would be best if the possibility of removing the water from a subsea separation unit and eventually reducing the cost of chemicals, because in general terms injection rates higher than 1.5 m³/day are not economic.

Equations from correlations exist that facilitate the calculations of the required inhibitor. The most common is the Hammerschmidt equation (Figure 34).

$$\Delta T = \frac{KW}{M(100 - W)}$$

where, ΔT : temperature shift, hydrate depression [$^{\circ}\text{C}$]

K : constant [-],

W : concentration of the inhibitor in weight percent in the aqueous phase.

M : molecular weight of the inhibitor divided by the molecular weight of water.

Inhibitor	K value
Methanol	2335
Ethanol	2335
Ethylene glycol (MEG)	2700
Diethylene glycol (DEG)	4000
Triethylene glycol (TEG)	5400

Figure 34: Hammerschmidt equation for calculation of temperature reduction of hydrate dissociation (Carroll, 2003)

The case for downhole hydrate inhibitor injection is also based on hydrate risk during start-up. The primary benefit of downhole injection is that hydrate risk may be averted altogether at start-up, regardless of start-up rate. In the start-up with downhole injection, injection begins as the well is brought on-line and continues until the safe condition time. Thus, there is no risk of hydrate plug formation and no requirement for urgent intervention. A secondary benefit of downhole injection is seen at a planned shut-in of the well, since the well fluids may be treated prior to shut-in so that displacement (bullheading) is not required.

Inhibitor deliverability from injection points may be extremely limited, especially in high pressure wells. This in turn may lead to greatly extended start-ups for significant water cut wells (Hudson et al, 2000).

3.2 LOW DOSAGE HYDRATE INHIBITORS (LDHI) - KINETIC INHIBITORS, ANTI-AGGLOMERANTS

This technology was established as a breakthrough when LDHI chemicals came available as a substitute for methanol. First LDHI products were marketed in 1998 and these products have been used in over 100 projects worldwide however some countries do not allow the use of these chemicals due to toxicity. Although now a prevalent chemical technology, LDHIs are currently under further development to address some areas such as offshore regulations, corrosion, temperature, water cut, and other factors (Kondapi et al, 2013).

Low dosage hydrate inhibitors are divided into two categories with different interaction mechanisms. These are described as follows:

Kinetic Inhibitors (KI)

These are polymers (polyvinylpyrrolidone, polyethyloxazoline and low and high Mw of polyethylene glycol) added at low concentrations (<1 w/w %) and do not affect the thermodynamics of hydrate formation. However, they do delay hydrate nucleation and/or crystal growth. The inhibition mechanism is understood to occur by the polymer interacting with certain faces of the ice crystals limiting the crystal growth (Kelland, 1995).

In some gas fields, where the process conditions are considerably severe, high dosage of KHIs (around 2%) is used to provide protection against hydrate formation (Lavallie et al, 2009). However, under high subcooling temperature (above 8 °C) and at high pressure in the presence of sour gas, thermodynamic inhibitor is indispensable because KHIs cannot provide protection even if 5% is used (Kelland, 2006). Therefore, rational design of polymers based on fundamental understandings of polymer properties is required to develop high performance inhibitors (Hould et al, 2015).

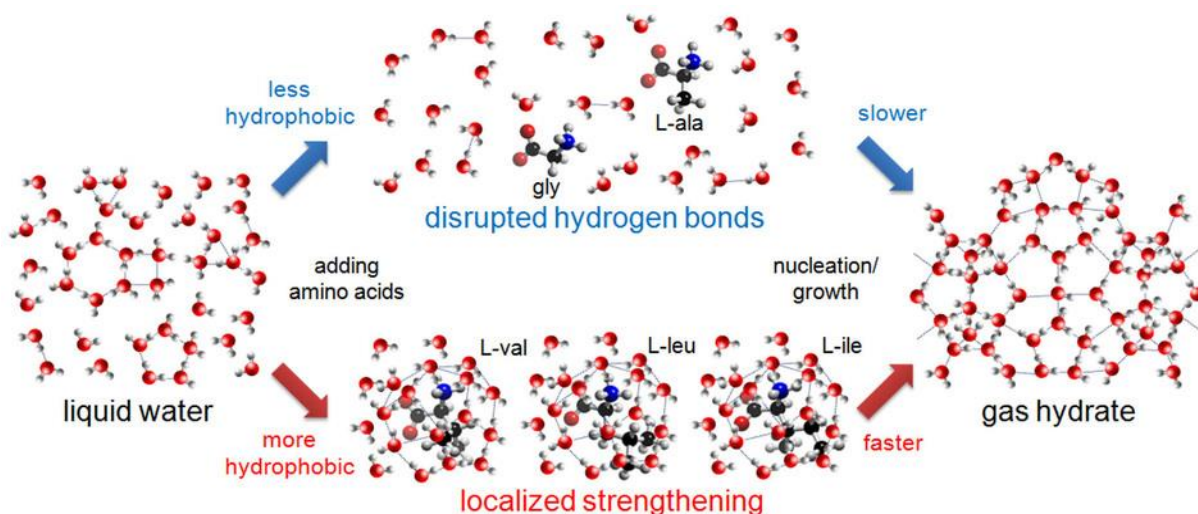


Figure 35: Chemical process of Kinetic Inhibitors

Anti-Agglomerates (AA)

These are polymers, also added at low concentrations (< 1 w/w %) and prevent the agglomeration of hydrates so that all the hydrate crystals are transportable and do not build up in the pipe.

Generally, the effectiveness of AAs does not appear to be dependent on the degree of subcooling, as are KIs. Hence, they have a much wider range of pressure-temperature applications. However, known AAs appear to work only in the presence of a hydrocarbon phase and their effectiveness is affected by the type of oil/condensate, the salinity of the water and the water cut (Kelland, 1995).

Use of KI or AA Chemicals in Shut-down Situations

During shutdown, the fluids are not moving and will cool down to the seabed temperature. For some KIs, this temperature may represent too great a subcooling for them to be effective. However, the lack of turbulence during shutdown will greatly reduce the rate of hydrate nucleation and formation as the gas diffusion to the aqueous phase will be minimal. Problems may occur during start-up when the fluids become turbulent and gas diffusion and hydrate formation rapid. To overcome this, a thermodynamic inhibitor could be pumped into the line just before start-up, then switched back to KI injection during normal production. To avoid the use of THIs completely, the better solution may be to use the AA type inhibitor which is much less dependent on the subcooling. The AA inhibitor must ensure that the pre-formed slurry of hydrate crystals

does not agglomerate during shutdown, so that slurry transportation can begin again at start-up.

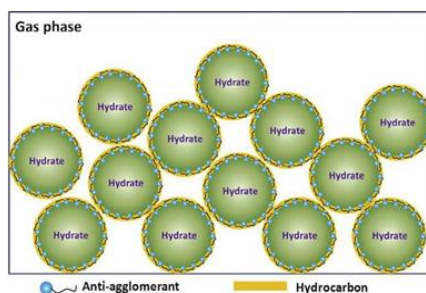


Figure 36: Proposed mechanism of Anti-Agglomerant in hydrate formation

Anti-agglomerates (AAs) can be used at any line pressure and temperature. However, they have several possible drawbacks. The effect of the AA appears to be dependent on a) the hydrocarbon fluid composition, b) the brine concentration and c) the water cut. This may mean different AAs need to be developed for different crudes. The AA must also remain active at varying brine concentrations and increasing water cuts as water breakthroughs occur and the field matures (Kelland, 1995).

LDHIs offer some advantages to the typical THIs (e.g. methanol and glycols) due to the lower concentrations required. Anti-agglomerates are exhibiting protection at higher subcooling than kinetic hydrate inhibitors. However, low dosage hydrate inhibitors are not recoverable and they are expensive.

Impact on Cost Effectiveness, Exploration and Production Activities

- Requires lower dosage rates than thermodynamic hydrate inhibitors such as methanol and MEG. Lower dosage rates decrease the chemical usage volume and operating expenses (OPEX) from lower chemical costs.
- Allows for higher production rates where methanol or MEG injection rate is limited by pumping capacity
- Enables faster start-up times for systems with many tiebacks
- Avoids excessive methanol content in the oil and its problems that methanol poses to refining operations
- Reduces chemical storage of methanol or MEG at processing facility, especially for offshore platforms
- Uses smaller pumps and chemical injection line capacities than with thermodynamic hydrate inhibitors such as methanol and MEG Reduces flammability risk associated with methanol storage

- Anti-agglomerate chemistries can eliminate the need for methanol or MEG injection prior to an extended shut-in
- Kinetic hydrate inhibitor chemistries can reduce the methanol / MEG injection rate when used together, reducing OPEX costs
- LDHI can reduce the insulation requirement and save CAPEX costs

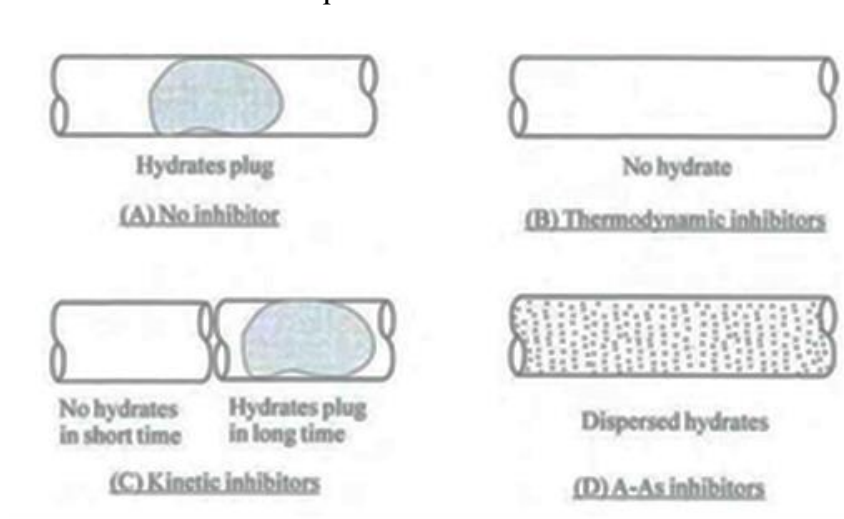


Figure 37: Summary of applications benefits of chemical inhibitors

Thermodynamic Hydrate Inhibitors	Kinetic Hydrate Inhibitors	Anti-Agglomerant Inhibitors
Applications		
1. Multiphase	1. Multiphase	1. Multiphase
2. Gas & Condensate	2. Gas & Condensate	2. Condensate
3. Crude Oil	3. Crude Oil?	3. Crude Oil
Benefits		
1. Robust & effective	1. Lower OPEX/CAPEX	1. Lower OPEX/CAPEX
2. Well understood	2. Low volumes (< 1wt%)	2. Low volumes (< 1wt%)
3. Predictable	3. Environmentally friendly	3. Environmentally friendly
4. Proven track-record	4. Non-toxic	4. Non-toxic
	5. Tested in gas systems	5. Wide range of subcooling
Limitations		
1. Higher OPEX/CAPEX	1. Limited subcoolings (<10°C)	1. Time dependency?
2. High volumes (10-60 wt%)	2. Time dependency	2. Shutdowns?
3. Toxic / hazardous	3. Shutdowns	3. Restricted to lower watercuts
4. Environmentally harmful	4. System specific – testing	4. System specific – testing
5. Volatile – losses to vapour	5. Compatibility	5. Compatibility
6. 'Salting out'	6. Precipitation at higher temps	6. Limited experience
	7. Limited exp. in oil systems	7. No predictive models
	8. No predictive models	

Figure 38: Summary of applications benefits and limitations of chemical inhibitors

3.3 PASSIVE THERMAL MANAGEMENT

Thermal insulation systems help to avoid hydrate precipitation by keeping the flowing fluids above the hydrate formation temperature throughout the length of the production line and at the desired production rate. Also state of the art insulating techniques provides extended cool down times for the production pipelines. The typical insulation technique involves filling the annulus completely or partially with a low thermal conductivity fluid in order to prevent heat loss (wet pipe). Lately other methods such as extended reach drilling, burial, Pipe in Pipe insulation and phase changing materials are currently emerging as well. Insulation offers the potential of a guaranteed solution provided economics and reliability can be significantly improved, particularly for deep-water applications (Hudson et al, 2000) (Lloyd, 2002) (Kondapi et al, 2013).



Figure 39: Silicone-based Syntactic Insulation

Typically brine is present in the annulus as completion fluid but most brine have high thermal conductivity and easily facilitate heat loss. Various fluids have been used as insulation. Silicate foams were amongst the first fluids to be documented as insulating packer fluids in steam- injection applications. The foam is an excellent insulator, with thermal conductivity of about 0.017 Btu/ (hr-ft²-oF/ft). Also, gelatinous oil based fluids have been developed for this purpose. (Purdy et al, 1991) employed vacuum insulated tubing (VIT) for wells near the Arctic Circle and (Javora et al, 2004) developed various water based insulating fluids for deep-water riser applications. As for today the insulation methods applied to deep-water oilfields are as follows:

Phase Change Material Systems

Most of the high inertia systems use phase change materials (PCM), i.e. materials which release a high latent heat during their transition from a solid state to a liquid state.

The PCM is located between the hot flowline and an insulation material, ensuring that it will be above its temperature of phase transition during operation. The thickness of the external insulation material provides most of the OHTC (Overall Heat Transfer Coefficient) of the system. High performance insulation material should be selected in order to limit the layer thickness and minimize the diameter increase of the outer pipe.

During the shutdown phase, the latent heat release will extend typical cool down times by 3 to 5 days. Such transient thermal performance can prevent the requirements for prevention measures for a significant percentage of the shutdowns throughout field life. However, in the event of a very long shut down period in excess of the cool-down time performance, the re-start of the pipeline may become a major issue. As the system is able to store and release a significant amount of heat, it will also need to retrieve the same amount of energy before reaching stable steady state operating conditions. Consequently, the start up of a PCM insulated pipeline will be longer than other types of pipeline. This may become a major issue in the case of a longer tie-back, where the risks of localized trapped pockets of gas are more prone to hydrate formation ([Denniel et al, 2004](#)).

Pipe in Pipe (PiP)

The choice of PiP system is critical for flow assurance in deep water environments as it provides a highly reliable and well proven method for achieving exceptional thermal performance in a passive manner. Insulation materials and construction methods used in the system allow tuning of the PiP system to the specifics of the field development. As a thermal management tool, PiP is developed to protect the line(s) from many operational hazards, such as hydrates, corrosion, and turbulent flow along with emulsion/foam, scale, liquid slugging and Paraffin/Asphaltenes. A Pipe-in-Pipe system consists of the flowline (the inner pipe) inserted inside a carrier, or jacket pipe, often referred to as the outer pipe ([Figure 40](#)). The created intermediate annulus is used to place so called dry insulation material with high insulating efficiency, protected by the outer pipe from the hydrostatic pressure and from water penetration.

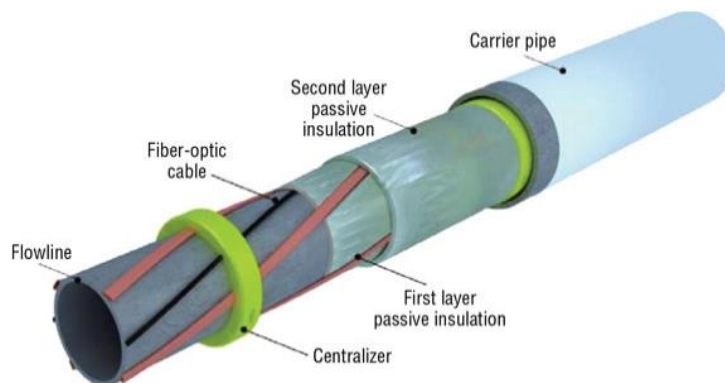


Figure 40: Typical Pipe in Pipe (PiP) configuration schematic

PiP systems employ one of the following insulation materials: traditional fiberglass and PU foam insulation or more modern Izoflex and nanoporous materials, also used in the space industry. The latter category provides thermal conductivity values down to $7 \text{ W/m}^2\text{K}$ and requires only a thin layer of insulation material to achieve low U-values. This reduces the pipeline weight due to the reduction in size of the outer pipe and, in turn, typically produces lower as-installed system costs while allowing extended tie-backs with longer cool-down time (Dixon, 2013).

Vacuum-Insulated Tubing (VIT)

VIT is a special case of PiP system but instead of dry insulating material in the annular space it utilizes vacuum conditions (Figure 41). Its use has become common in deep-water oil wells, and has even occurred in some cases for DVA oil wells. The use of VIT has a significant impact on the thermal performance of the well, and thus on the flow assurance design. While the increase in flowing tubing head temperature during normal production in early-life is typically very modest, significant increases are often observed at low rates, or in the late-life condition. Thermal transients at startup and shut-in are also impacted.

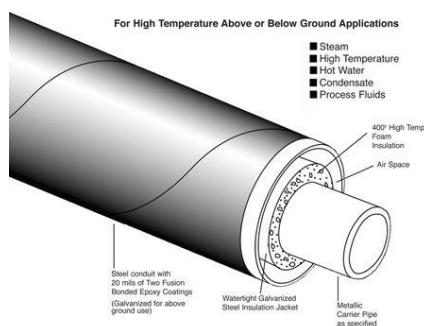


Figure 41: VIT pipeline schematic

From a flow assurance perspective, the most significant benefits of VIT use are:

- More turndown ratio – Because the flowing tubing head temperature is higher at low rates with VIT, the flow assurance minimum rates for cool-down (hydrate management) are decreased. The benefits may include extended field life and additional operational flexibility.
- Faster warm-up – Because the tubing head is well insulated from the surroundings, produced fluids are not cooled as much during start-up. The result is shorter times to HDT and shorter safe condition times. Overall, the risk of hydrate plug formation in the well during startup is typically decreased with VIT usage.

Of course, there are drawbacks to its use as well. The most significant of these include:

- Faster cool-down – Insulating the tubing from the surroundings also means that at shut-in, the surroundings will be cooler and long cool-down times may be difficult to achieve.
- Costs / Project Schedule – VIT costs often exceed \$100 \$/ft.
- Mechanical integrity issues – The increased complexity of the well means more opportunities for failure. This includes collapse of the VIT, but also other hazards of having less room in the well due the spatial requirements of the VIT (Hudson et al, 2000).

3.4 ACTIVE THERMAL MANAGEMENT

Active Heating – Hot Fluid and Electric Heating

Actively heated systems provide the next level of thermal management. They generally use hot fluid (Figure 42) or electricity (Figure 43) as a heating medium. The main attraction of active heating is its flexibility. It can be used to extend the cool down time indefinitely, i.e. continuously maintaining a uniform flowline temperature above the critical levels of hydrate formation. It is also capable of warming up a line from seawater temperature to a target operating level and avoid the requirements for complex and risky start-up procedures. Consequently, it is not necessary to rely on a circulation loop and dual lines to manage flow assurance issues.

Active heating by circulation of hot fluid is generally more suited to a bundle configuration because large pipeline cross sections are required. There is a length limitation to hot fluid heating because it generally involves a fluid circulation loop along

which the temperature of the heating medium decreases. In order to be sufficiently efficient, it is necessary to inject large volumes of fluid at a relatively high temperature. This involves storage facilities and significant energy to heat the fluid and to maintain the flow.

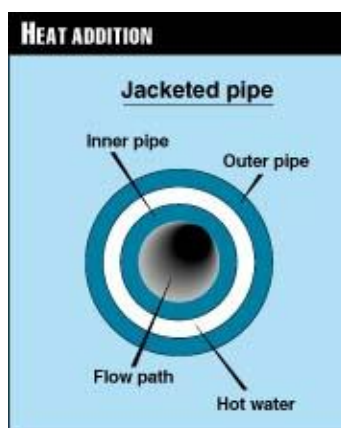


Figure 42: Active pipeline heating with hot fluid circulation

Electrical heating does not need these storage facilities and is applicable to smaller size pipeline systems such as pipe-in pipe or wet insulated flowlines. It also offers the advantage of providing a uniform heating input to the flowline. With safeguards in-place to prevent possible shorting of the power line to the production line, the efficiency of electric power to heat power is between 25% and 30% (Lloyd, 2002).

A number of qualified electrically based systems are available to the market meeting different heating performance and linear heating requirements. Trace heating is believed to provide the highest level of heating efficiency and can be applied to bundles or to pipe-in-pipe system. The length capability of an electrical system depends on the linear heat input required and the admissible voltage. Trace heating can be applied for very long tie-backs (several 10s of kilometers) by either using higher voltage or by introducing intermediate power feeding locations (Denniel et al, 2004).

Direct Electrical Heating (DEH) technology is matured for shallow water applications and pipelines that are shorter than 50 km but for risers it is still a growing technology. This technology is still a challenge for plug removal as it is being currently investigated for risk factors such as local pressure buildup and pipe rupture. The technology needs to be extensively tested for deep water applications and longer pipelines, risers and manifolds. The fields that have been using DEH open loop technology are Statoil Asgard, Huldra, Kristin, Urd, Tyrihans, Alve, Ormen Lange, Morvin, BP Idun and Skarv. Other

fields using closed loop technology are Shell Serrano and Oregano, Nakika and Habanero (Kondapi et al, 2013).

However evolutionary extension of the technology; it carries premium costs as well as increased contingency risks in the active operational phase. This type of system benefits from a systems engineering approach so that operators and contractors are able to select and implement active PiP systems safely, reliably and cost-effectively (Dixon, 2013).

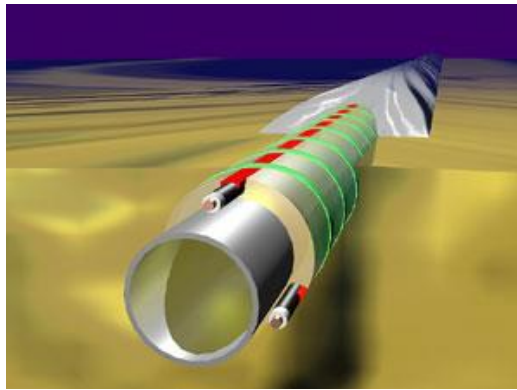


Figure 43: Actively heating pipeline with electrical cable

Impact of Active Heated Systems on Cost Effectiveness of E&P Activities

- Active control of flowline operating temperatures
- Eliminate or reduce inhibitor consumption rates
- Potential tool for hydrate remediation (field-specific)
- Supplement or replace other hydrate management strategies: depressurization, dead oil circulation, etc.
- Allow for quick restart after a shutdown, no need for remediation actions; depressurization, dead oil circulation, etc.
- Reduce capital expenses (CAPEX) by use of single flowline
- High reliability leading to reduced maintenance costs.

3.5 PIGGING

A pig, flowline investigation gauge, is a cylinder shaped tool that hugs the wall inside the flowline (Figure 46). Its outer diameter is almost the same as the flowlines inner diameter. The pig is transported through the flowline from high to low pressure where it cleans the pipe along the way (Figure 45). Modern pigs are advanced diagnostic tools. It is important not to let the pig intervals become too large. The deposits accumulate over

time, and may become too large for the pig to transport. The result is a stuck pig, and this may lead to the abandonment of the flowline.

A form of flow assurance for oil and gas pipelines and flowlines that ensures the line is running smoothly is called pipeline pigging. The maintenance tool, pipeline pigs are introduced into the line via a pig trap, which includes a launcher and receiver (Figure 44).

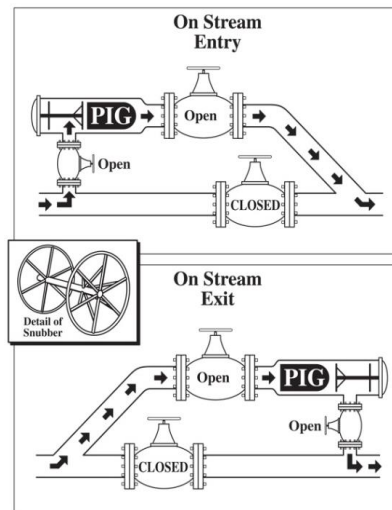


Figure 44: Pigging launcher and receiver on a pipeline

Without interrupting flow, the pig is then forced through it by product flow, or it can be towed by another device or cable. Usually cylindrical or spherical, pigs sweep the line by scraping the sides of the pipeline and pushing debris ahead. As the travel along the pipeline, there are a number functions the pig can perform, from clearing the line (utility pigs), inspecting the interior (smart pigs) and specialty pigs for situations that repair must be done in the pipe.

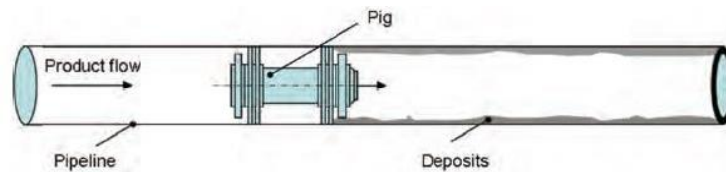


Figure 45: Pig running through pipeline

In the specific case of hydrate remediation pigging is used to either remove the hydrate deposition from the pipeline wall before it plugs the line or for locating the plugs so they can be treated with other methods. A typical design is the dual pipeline from a wellhead which renders the system more flexible and easier to stop production from one of the lines to perform remedial work.



Figure 46: Pig after remedial work on a pipeline

Benefits of dual pipelines:

- Turndown (operation through only one line at low production rates)
- Allowing operation at two independent pressure levels (allowing high pressure wells to flow in one line and low pressure in the other). Maximizing utilization of transport capacity.
- Dynamic pigging (periodically increasing the production through one line at the time to sweep out liquid).
- Impact on slug catcher size (through better liquid inventory control).
- Opening up for gas recycle.
- Round trip pigging.
- De-pressurization from both sides.
- Simplifies de-watering & first start-up.

3.6 BLOWDOWN

Depressurization is one method to move the fluid outside the stable region of the hydrate phase envelope. Blowdown can be applied locally for plug dissociation but it is prone to accidents because partially dissociated plugs move quickly in the flowline with the danger of accidents either to personnel or equipment, if the operation is not performed correctly. The procedure and the safety concerns are described analytically.

Normally the water temperature at the sea bed is above the freezing point of water. This excludes the possibility of ice plugs forming. When a hydrate plug is formed the system quickly cools down to the ambient temperature. Pressure and temperature conditions are illustrated in Figure 47. To the left of the three-phase line (Lw-H-V) hydrates can form, whilst on the right hand side only fluids can exist. The figure shows how rapid pressure relief may hurl the system further into the hydrate region.

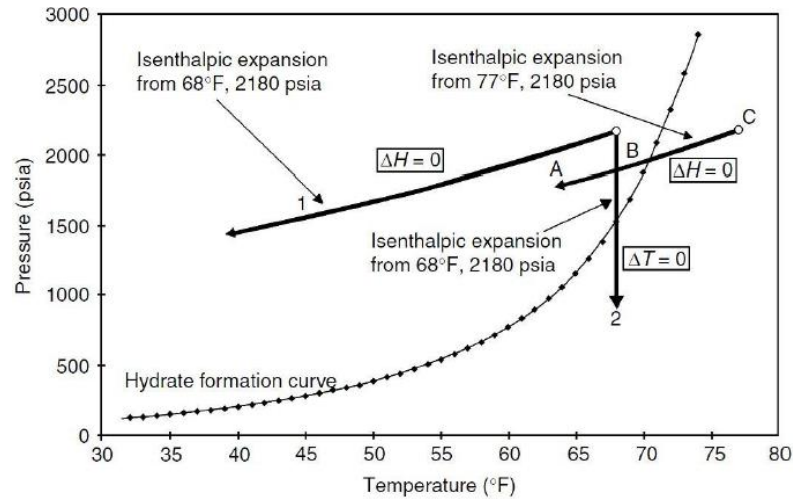


Figure 47: Pressure Dissociation (Sloan & Koh, 2008)

Two different depressurization scenarios are depicted in Figure 47. Point 1 illustrates how the temperature drops when the gas is passing through a restriction like a valve. Rapid expansion, where $\Delta H=0$, will cool the gas quickly. In point 2 a large volume of gas is depressurized very slowly at a constant temperature. On the right in the figure it is shown that expanding gas may move from outside the hydrate region and into it due to expansion.

Generally the flowline may not be depressurized fast enough for the Joule-Thompson effect to occur. If the flowline is depressurized slowly it will be an isothermal process, and the temperature will not change, $\Delta T=0$. Most often an intermediate pressure relief causes the hydrate temperature to fall below the ambient temperature. This results in heat being transferred from the surroundings to the plug. This results in radial heat transfer from the surroundings to the center of the flowline. That means that the temperature is at its highest at the flowline wall. And it causes melting along the entire plug length in contact with the wall.

Melting of a hydrate plug after one, two and three hours is displayed in Figure 48. Radial hydrate melting controls the plug removal in the flowline because the pipe diameter is at least an order of size smaller than the length of the plug.

How the heat flows radially towards the center of the plug when the hydrate temperature is less than ambient temperature is illustrated in Figure 48. And it causes dissociation along the entire length of the plug. Melting also occurs at the ends of the plug, just in a smaller tempo. It is the dissociation at the wall which controls the speed of the plug removal.



Figure 48: Radial Dissociation of Hydrate Plugs (Peters, 1999)

Two sided pressure relief is the recommended method, both from a safety and technical point of view. It may be difficult to implement if the liquid head on the hydrate plug is larger than the dissociation pressure. Typical scenario for this incident to happen is in ultra-deep waters. The second method is one sided pressure relief (usually from the separator side) which can be performed easily, but leads to extreme pressure differential at the hydrate plug. When the plug gets dissociated it will move as a projectile with speeds up to 60 m/s (Figure 49).

A partly dissociated plug may move down the flowline when the system is restarted. The partly dissociated plug may get stuck and form a new plug at pipe bend, valve or other flow restrictions.

If the momentum is large enough it may make the plug more compact. In extreme incidents the plug may form a moving projectile, and be a serious safety concern. To avoid this, the flowline is normally pumped full of methanol when the annulus is large enough to allow for fluid flow past the plug. The methanol dissociates the rest of the hydrate plug.

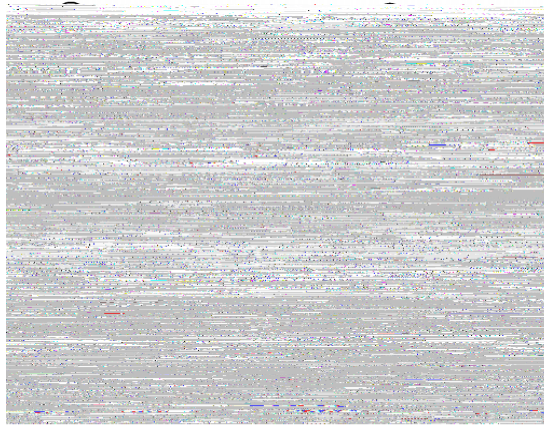


Figure 49: One side depressurization resulting to hydrate plug projectile

A hydrate plug should be dissociated as soon as possible due to physical/chemical and economic reasons. Younger plugs have larger porosity and permeability and less hydrates. Hydrate plug depressurization should always be done carefully because hydrate projectiles create a large safety risk, and can cause damage to equipment and personnel. Joule-Thomson effect is avoided by two-sided depressurization. Also two-sided depressurization can cause radial dissociation. This can reduce the down time by 50% compared to one-sided depressurization.

To apply two-sided depressurization another production flowline or umbilical should be used to depressurize upstream. Sometimes the liquid head on the hydrate plug may be too large to perform depressurization. In such events, direct electrical heating may be a good alternative.

There are several incidents where flowlines have ruptured and people have been seriously injured, or killed. Safety problems are caused by three types of characteristics. The density of hydrate is close to that of ice, which combined with a large upstream pressure gradient leads to a detached hydrate plug with very high velocity.

The ruptures are most likely to happen at a flow restriction (orifice), obstruction (flange/valve), or by an immediate change in direction (bend, tee) as shown in [Figure 50](#).

As the hydrate projectile travels down the flowline the gas in front of it is being compressed and can lead to a burst. A direct impact could also cause the pipe to explode.



Figure 50: Two Ways a Hydrate Plug Can Rupture a Pipe (Sloan & Koh, 2008)

When it is discovered that the system is blocked by hydrates, it is not possible to know how many plugs there are. There is a risk that high pressure gradients are trapped between plugs. Hydrates contain approximately 164 Sm³ gas per cubic meter hydrate. Hydrate plugs that are dissociated by heating release a lot of gas. If this gas is trapped between two plugs, there may be a rapid increase in the gas pressure, resulting in serious damage in the pipe (Caroll, 2003).

3.7 COILED TUBING REMEDIATION

In the oil and gas industry, coiled tubing refers to a very long metal pipe; normally 1" to 3.25" in diameter which is supplied spooled on a large reel.

Coiled tubing is used for various downhole operations such as fishing, gas lift initiation, injection of chemicals, etc. Regarding flow assurance and specifically hydrate plug remediation coiled tubing can be used inside the tubing to drill through the hydrate plug and thus remove it from the line. Also coiled tubing can be used for depressurization methods locally around the plug.

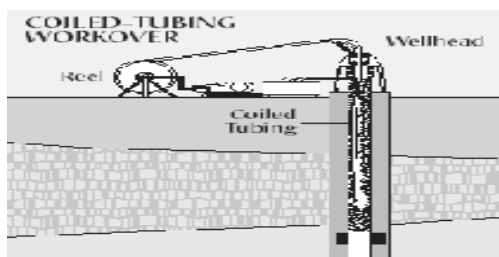


Figure 51: Coiled tubing workover application

3.8 SUBSEA SEPARATION

Subsea processing offers promise for reducing the hydrate inhibition requirements. Both multiphase (gas/oil/water, and two phase (gas/liquid) separation systems are in development. Three-phase separation with water re-injection would be ideal. Water quantity and therefore chemical concentrations would be minimized if required at all. The amount of overboard water disposal would be significantly reduced. However, achieving injection quality water can be a challenge even on topsides. Working over a deep-water injection well due to can be very expensive ([Mogbolu et al, 2014](#)).

Subsea Separation technology is growing at an increasing rate due to its huge potential to increase recoverable reserves and to accelerate production. The challenges that still exist for subsea separation are:

- Achieve liquid-liquid separation and gas-liquid separation from heavy oils
- Realize the optimum combination of pump acceptance criteria with respect to gas-liquid separator design for heavy oil applications
- Cost and installation challenges and opportunities to reduce bulky and heavy equipment
- Disposal of the separated water
- Improving and maturing an efficient compact design

Subsea Separation projects have been installed in North Sea, GoM, West Africa and Brazil with many other subsea field developments in the process of considering Subsea Separation systems. The first pilot-separation system was installed on Troll field in 1999 for liquid-liquid separation and in 2001 for gas-liquid separation ([Kondapi et al, 2013](#)). Projects that have been delivered to date are:

- Oil-Water Separation projects: Statoil Tordis and Petrobras Marlim
- Gas-Liquid Separation projects: Shell Perdido, Shell BC-10 and Total Pazflor



Figure 52: Petrobras Marlim Subsea Separator



Figure 53: Shell Perdido Caisson Separator

Impact of Subsea Separation on Cost Effectiveness during E&P Activities

- Accelerated and/or increased recovery achieved by reducing back-pressure on wells
- Reduced risk due to reservoir uncertainty versus multiphase pumping only
- Improved flow assurance such as hydrates, wax, slugging, erosion, less chemical injection
- Reduced CAPEX on topside processing equipment and number of flowlines and insulation
- Useful in harsh Environments (Kondapi et al, 2013)

3.9 COLD FLOW TECHNOLOGY

In flow assurance, there is a hydrate plug prevention shift under way: from avoidance to management of hydrate formation. In addition to avoiding the region of hydrate stability time dependent studies enable flow assurance engineers to better address such concerns as flowline restarts, cold (stabilized) flow, low-dosage hydrate inhibitors, and plug remediation.

In a growing number of flow assurance situations, hydrate risk management is more economical than avoidance. One aspect of hydrate risk management is to allow hydrate particles to form, but to prevent hydrate particle aggregation to form a blockage by ensuring that the particles will flow, and remain entrained in the oil phase. To move from avoidance to risk management, it is essential to quantify the hydrate formation time dependence. From experimental work the following seem to be the essential factors that can prevent the hydrate particle agglomeration ([Borden, 2014](#)).

- The formation of emulsions/dispersions to keep the water/hydrates suspended in the oil phase is one key to prevent blockage formation.
- Particle aggregation may be prevented by high shear stress.

There are two requirements to prevent hydrate-plug formation in oil-dominated systems:

- Hydrate particles must be in low concentration (<50 v/v %).
- Particle aggregation is prevented by particles being oil-wet through oil chemistry, or by application of anti-agglomerate chemicals.

The closer the operating conditions are to the hydrate dissociation conditions, the stronger the interactions between hydrate particles, because of attractive capillary forces from a quasi-liquid layer at the particle interface. The formation and dissociation of hydrates can cause coalescence of water drops in water-in-oil systems. This coalesced free-water phase is prone to hydrate-blockage formation.

Like other deposits, freshly formed hydrates are more porous and malleable than are hydrates that have time to age and solidify. The aging process causes a more-dense crystal mass, making dissociation of the plug increasingly difficult. There are two main, patented, cold or stabilized flow concepts ([Lund et al. 2004](#); [Talley et al. 2007](#)). In each process, the key principle is to emulsify and convert free water to hydrate as entirely and rapidly as possible. Without a free water phase to encourage hydrate particle aggregation, hydrate particles will not aggregate but will flow with the oil phase, much like dry snow is difficult to compact/aggregate into a snowball ([Sloan et al, 2009](#)).

Cold Flow Technology is still in the development stage and yet to be tested for a commercial application. Since the technology is relatively new and not validated at industry scale, large scale testing and R&D efforts are still needed to bring qualification of this technology before a field application can be planned. Development target is to implement in a large scale pilot testing facility and validate technical challenges such as (Kondapi et al, 2013):

- a) Long distance transport of high fluid viscosity slurry with high pressure drops and boosting capacity
- b) High heat capacity because of more hydrate formation with high water cuts

Impact of Cold flow Technology on Cost Effectiveness during E&P Activities

- Reduce CAPEX and OPEX on heating system hardware and operation.
- Reduce CAPEX on pipeline insulation.
- Reduce OPEX on production chemical costs by reducing need for chemical injection.
- Reduce OPEX by reducing MEG regeneration needs.
- Reduced handling of bulk and specialty harmful chemicals.
- CAPEX and OPEX for topside processing facilities may be increased to process hydrate slurries.
- Higher costs for subsea cooling systems and multiphase/liquid pumps may be anticipated.

4. PRINCIPLES AND METHODS USED FOR SENSITIVITY ANALYSIS CALCULATIONS

In the next chapters an evaluation of the sensitivity analysis scenaria of the production system parameters will be given. For accomplishing this task appropriate software is used that facilitates the field development, design and production simulation. To predict whether hydrates formation could take place or not, the software needs to determine the fluid composition and the prevailing pressure and temperature conditions at each point along the length of the pipeline. For this task, a series of calculations are required to determine the pressure loss (gravitational and frictional), the temperature profile of the pipeline and the thermodynamic properties of the fluids (PVT). The determination of these parameters is done by use of the nodal analysis method, combined to thermal calculations along with the use of an equation of state. The implementation of these methods is further discussed in this chapter.

4.1 PVT DATA

As already discussed, hydrate formation depends only on three parameters; the presence of water, a suitable combination of temperature and pressure and the abundance of hydrate forming molecules. All these parameters need to be calculated (analytically or from correlations) at any point along the production line.

The water is considered always present in the flowline and no further calculations are needed to prove it, even in early production dates where the water cut is low. The other three parameters, pressure, temperature and hydrate forming molecules, must be determined as accurately as possible to provide reliable results on hydrate formation possibility and location.

Accurate information regarding the fluid composition is necessary for running hydrate calculations. For production system calculations three types of PVT data can be used. Starting from the simplest, a “black oil” model can be used to simulate the thermodynamic behavior of the fluids. In the black oil model water is modeled explicitly together with two hydrocarbon components, one (pseudo) oil phase and one (pseudo) gas phase. The data therefore needed for such calculations consists only from non-compositional data, such as GOR (Gas-Oil Ratio), Bo (Oil Formation Volume factor), Rs (Gas in Oil solution), Bw (Water Formation Volume Factor) and Bg (Gas Formation Volume Factor). This model can be used only for fast calculations to get a rough estimation of how the fluid behaves, but is not accurate for further modelling and developing. A more advanced version of the black oil model is the tuned black oil model,

using real fluid data. The analytic calculation of the bubble (or dew) point pressure, Constant Composition Expansion and a Differential Vaporization study helps the reliability of the model in every pressure encountered. This model ultimately works well for material balance calculations and simple reservoir simulation applications. Also gas lifting applications are often performed using this type of PVT data.

For more demanding applications the black oil model is not adequate to get accurate calculations. For this reason fully compositional fluid models, utilizing an Equation of State (ex. Peng- Robinson cubic equation), are needed to predict the composition of every flowing phase under specific P-T conditions. These models can use the fluid composition as found from fluid analysis in a gas chromatogram. To calculate the hydrate forming conditions (i.e. the hydrate dissociation curve) the use of a fully compositional model is obligatory as the exact concentration of the hydrate forming molecules is needed. In the sensitivity analysis scenaria discussed later, two fully compositional models were used. One of a gas condensate and one of a light oil, both of which were already tuned against laboratory data.

4.2 NODAL ANALYSIS – INFLOW PERFORMANCE RELATIONSHIP – VERTICAL LIFT PERFORMANCE

Necessary parameters for computing hydrate forming conditions, other than the fluid composition, are the pressure and temperature along the flowline. One cannot simply measure the reservoir pressure from a well test (under “no-flow” conditions) and assume the pressure drop inside the tubing from an analytic expression. In fact, pressure calculation is dependent on the gravitational losses for the lifting of the hydrostatic column and the frictional losses due to fluid movement through the pipeline. Under “no-flow” conditions the second term cannot be measured and also the first is not accurately described, as slippage effects are not taken into account.

The flow of reservoir fluids from the reservoir to the stock tank requires an understanding of the principles of fluid flow through the porous medium and the well tubulars. As the fluid moves through the production flowline, there will be an associated pressure drop along the fluid flow. The total pressure drop, from the perforations level to the wellhead or to the separator, will be the sum of the pressure drops through the various components in the production system.

As a result, the design of a production system requires an integrated approach, since the system cannot be separated into various components and handled independently. The production flow rate depends on the total pressure drop in the production system, and vice versa. Consequently, the entire production system must be analyzed as a unit or system and this is ultimately the purpose of the nodal analysis method.

Systems analysis has been used for many years to analyze the performance of systems composed of multiple interacting components. Gilbert was the first to introduce the approach to oil and gas applications but Mach, Proano and Brown and Brown popularized the concept, which is typically referred to as Nodal Analysis within the oil and gas industry. The objective of this analysis is to combine the various components of the production system for an individual well to estimate production rates and optimize the components of the production system.

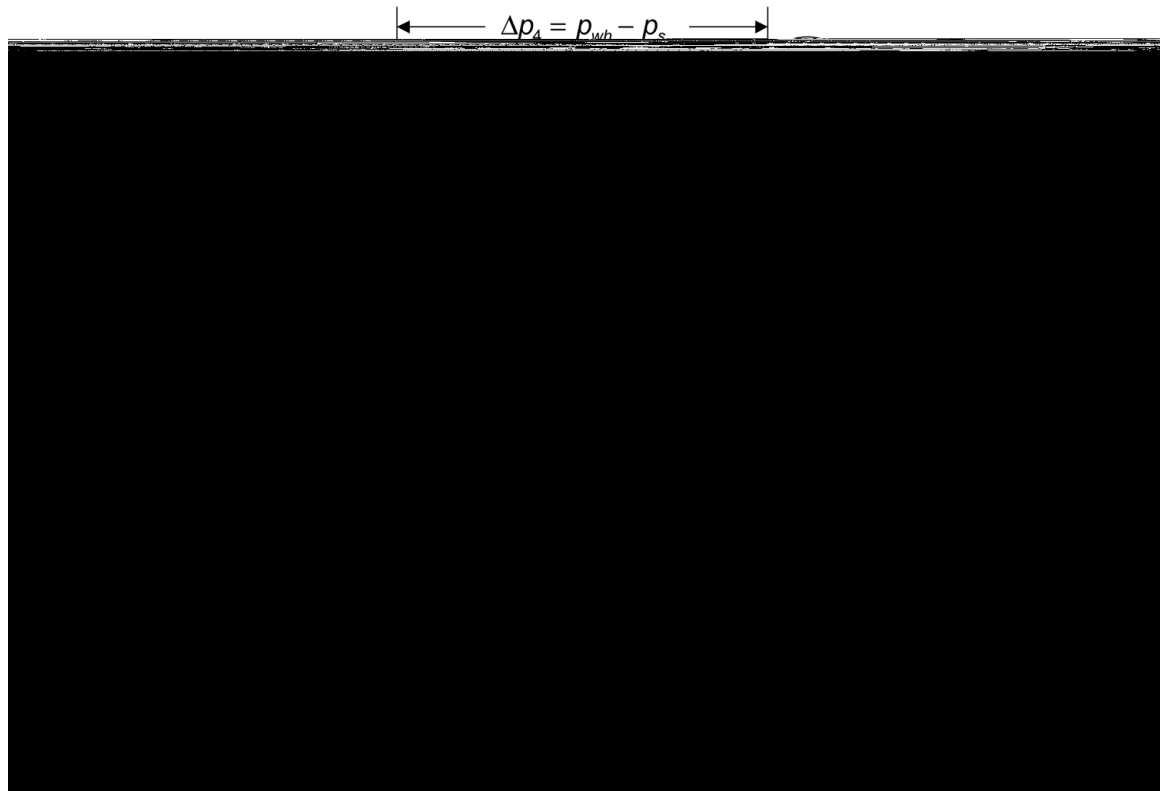


Figure 54: Production System pressure losses

There is a total pressure drop from the reservoir pressure to the surface, as depicted in Figure 54. When the separator is the end of the production system, the total pressure drop in the system is the difference between the average reservoir pressure and the separator pressure:

$$\Delta p_T = \overline{p_r} - p_s$$

The total pressure drop however is composed of individual pressure drops as the fluid flows to the surface. These pressure drops occur as the fluid flows through the reservoir and completion, up the tubing, through the wellhead equipment and choke, and through the surface flowlines to the separator. Thus, the total pressure drop can be represented as:

$$\Delta p_T = \Delta p_1 + \Delta p_2 + \Delta p_3 + \Delta p_4$$

The above pressure drops can be divided into yet additional ones to account for restrictions, subsurface safety valves, tubing accessories that increase pressure losses locally. At every point of the production system there exist a particular pressure and production rate associated with that point for certain conditions. If there is any change in the system, then there will be an associated change in pressure and production rate at that point. This concept allows the division of the production system at a point of interest for evaluation of the two portions of the system. This evaluation determines the conditions of continuity of pressure and production rate at the division point.

The nodal analysis approach provides the flexibility to divide the production system at any point of interest within the system. The most common points of interest are at the wellhead, the perforations and the reservoir sandface. The terminal ends of the system are the reservoir on the upstream end and the separator at the downstream end of the system. In the cases studies in later chapters the solution node is set at the perforations level, as it divides the losses from the reservoir (Inflow Performance Relationship – IPR) to those in the wellbore and the pipeline from the perforations to the separator (Vertical Lift Performance –VLP).

The components upstream of the division point comprise the inflow section (IPR) of the system, while the downstream components represent the outflow section (VLP). Once the system has been divided into inflow and outflow sections, models are used to describe the rate-pressure relationship within each section. The solution flow rate through the system is determined once the following conditions are satisfied, that is:

- The incoming flow into the division point equals to the flow out of the division point ($Q_i = Q_o$)
- The pressure at the division point is the same at the corresponding endpoints of both the inflow and outflow sections of the system ($P_i = P_o$)

After the division point has been selected, pressure relationships are developed for the inflow and outflow sections of the system to estimate the node pressure. The pressure in the inflow section of the system is determined from:

$$\overline{p_R} - \Delta p_u = p_n$$

where:

p_n : pressure at the node

Δp_u : pressure losses in the reservoir

p_R : average reservoir pressure

while the outflow section pressure is determined from:

$$p_s + \Delta p_d = p_n$$

where:

p_n : pressure at the node

Δp_d : pressure losses in the wellbore and surface pipeline

p_s : separator pressure

The pressure drop at any component, and consequently both the inflow and outflow section of the system, varies as a function of flow rate. As a result, a series of flow rates is used to calculate node pressures for each section of the system. Then, plots of node pressure vs. production rate for the inflow section and the outflow section are made. The curve representing the inflow section is called the inflow performance relationship (IPR) curve, while the curve representing the outflow section is the vertical lift performance (VLP) curve. The intersection of the two curves provides the point of continuity required by the system analysis approach and indicates the anticipated production rate and pressure for the system being analyzed.

Vertical Lift Performance (VLP) – Inflow Performance Relationship (IPR) Curves

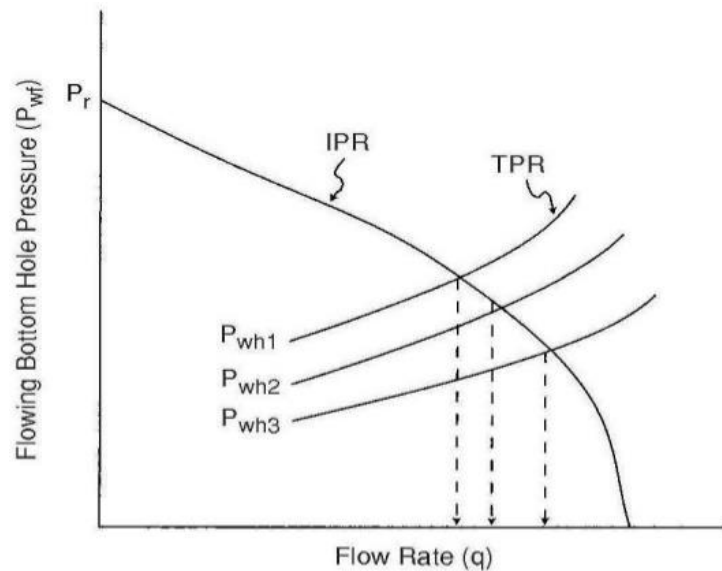


Figure 55: IPR and Tubing Performance Relationship (TPR or VLP) curves intersection

The effect of varying the outflow component (labeled 1, 2, and 3 respectively) while IPR remains unchanged is shown in Figure 55. In this example, the inflow and outflow performance curves do intersect for all three cases, implying that the design ensures flow. However, it could be the case when the reservoir does not have enough pressure to lift the fluids to the surface, where no solution rate can be found by these curves (no intersection). If an intersection exists the system satisfies continuity and the well will be expected to produce at a rate and pressure indicated by the intersection of the IPR and VLP curves. System 3 would be expected to produce at a higher rate and lower bottom hole pressure than System 2 and 3, as indicated by the graph.

The VLP curve of System 3 has a rapidly decreasing pressure at low flow rates and increase with increasing rate. This is typical for many outflow curves, which, in some cases, will yield two intersection points with the inflow curve; however, the intersection at the lower rate is not a stable solution and is meaningless. The proper intersection of the inflow and outflow curves should be the intersection to the right and several pressure units higher than the minimum pressure on the outflow curve.

The P_r line in Figure 55 depicts the Inflow Performance relationship curve. It is very important in petroleum engineering to formulate the relation between production rate (q) and flowing wellbore pressure (p_{wf}) over a practical range of production conditions. This relation is commonly known as Inflow Performance Relationship (IPR).

Productivity index is a measure of the ability of a well to produce hydrocarbon fluids and water. It is the ratio of the liquid flow rate to the pressure drawdown:

$$PI = \frac{q_o}{P_R - P_{wf}}, q_o \frac{kh(p_{av} - P_{wf})}{141.2\mu_o B_o [\ln \frac{r_e}{r_w} - S]}$$

where

k: rock permeability (mD)

h: perforation height (ft)

p_{av}: average reservoir pressure (psig)

p_{wf}: pressure at the perforations (psig)

μ_o: oil viscosity (cP)

B_o: oil formation volume factor (dimensionless)

r_e: external radius of well drainage area (ft)

r_w: wellbore radius (ft)

S: skin factor (dimensionless)

Productivity index, also denoted by J, is a reliable indication of the well production capacity only if the well is producing under steady and usually pseudo steady state flow regime. Hence, it is important to allow the well to flow at constant flow rate for a sufficiently long period of time, during a well test, to reach the pseudo steady state. When these conditions are achieved a point of the IPR can be obtained (P-Q). To obtain a second point the well must be shut-in for enough time for the pressures between the sandface and the reservoir to be equalized. With this procedure the average reservoir pressure is obtained (pressure for no flow conditions). This set of P-Q data is the origin of the IPR curve from the left side.

For analyzing the performance of oil wells, engineers often assume that flow rate is directly proportional to the pressure drawdown, with productivity index J being the proportionality constant. The resulting IPR is a straight line passing through the origin and is only applicable to undersaturated oils. The effect of two phase flow on the shape of IPR curve was first observed in 1920s and 1930s during field testing with a curved rather than straight line IPR. In this case, the productivity index is not constant and decreases with increasing drawdown (Figure 55). Several empirical methods have been proposed in the literature to generate the present and future IPRs in solution-gas drive reservoirs (Fetkovich). The most commonly used of these methods is the one proposed by Vogel

In Vogel's method, the IPR curve for a well producing saturated oil from a solution gas drive reservoirs can be approximated by the dimensionless quadratic equation:

$$\frac{Q}{Q_{\max}} = 1 - 0.2 \frac{P_{wf}}{P_{ws}} - 0.8 \left(\frac{P_{wf}}{P_{ws}} \right)^2$$

where

Q = oil flow rate, bbl/day

Q_{\max} = maximum oil flow rate at BHP = 0 psig, bbl/day

P_{wf} = FBHP = flowing bottom hole pressure, psia

P_{ws} = reservoir pressure, psia

The effect of changing any component of the system can be evaluated by recalculating the node pressure for the new configuration. If a change is made in an upstream component of the system, then the inflow (IPR) curve will change and the outflow (VLP) curve will remain unchanged. On the other hand, if a change in a downstream component is made, then the inflow curve will remain the same and the outflow curve will change. Both the IPR and VLP curves will be shifted if either of the fixed pressures in the system is changed, which can occur when evaluating the effects of reservoir depletion or considering different separator conditions or wellhead pressures.

Nodal analysis is used for many purposes in analyzing and designing production oil and gas wells. The approach is applied for evaluating both flowing wells and artificial lift applications. The technique provides powerful insight in the design of an initial completion. Even with limited data, various scenarios can be evaluated to estimate the well behavior. This process is very useful in analyzing current producing wells and increasing their performance. Typical applications include:

- Estimation of flow rates
- Selection of tubing diameter
- Selection of wellhead operational pressures
- Estimation of the effects of reservoir pressure depletion
- Identification of flow restrictions
- Evaluating perforation density
- Gravel pack design
- Artificial lift design
- Optimizing injection gas-liquid ratio for gas lift
- Evaluating the effects of lower wellhead pressures or installation of compression
- Evaluating well stimulation treatments

4.3 THERMAL CALCULATIONS

The last necessary parameter for evaluating hydrate formation is the flowing temperature conditions along the pipe. In this section, a short description of the thermal calculations required for determining the temperature profile is given.

When calculating the temperature profile of a well fluid, the friction is not so important, but the temperature in the surrounding formation is of great importance for the temperature development in the well fluid. Moreover, the heat flow properties of the surrounding formations and of the casing, cement are significant factors.

Two different methods for the temperature calculations are available. The simpler of the two is called the Rough Approximation method and the other one is the Enthalpy Balance method.

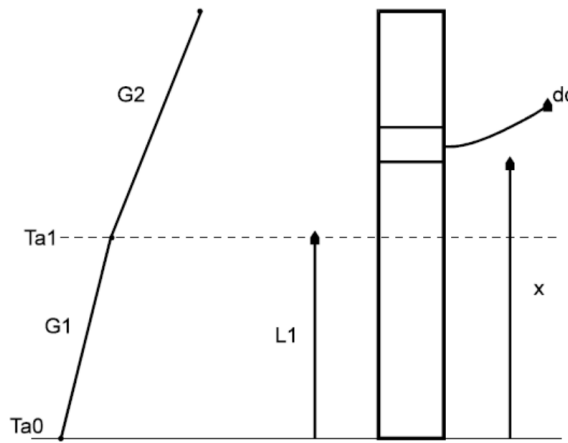


Figure 56: Geothermal gradient and surface surrounding environment temperature gradient

The above diagram shows the ambient temperature along the length of the well and the surface pipeline. From this gradient the ambient temperature of any point can be calculated easily by the equation:

$$T_a(x) = T_{a1} - G_2(x - L_1) \cos \theta$$

where:

$T_a(x)$: temperature of a node at x distance from the perforations

T_{a1} : temperature of the node at L_1 distance from the perforations

x : distance from the perforations

G_2 : pipeline length above point L_1

$\cos \theta$: angle of temperature gradient

Rough Approximation Model

The Rough Approximation temperature model assumes that the heat exchanged between the fluid and the surrounding environment by different heat transfer mechanisms can be captured using an Overall Heat Transfer Coefficient (U Value). This U value has units BTU/h/ft²/°F, where the area parameter refers to the pipeline wall. The heat transferred is given by the equation below:

$$Q_H = UA(T_e - \overline{T}_{fl}) = U\pi Dx(T_e - \overline{T}_{fl})$$

where:

A: area of the inner tubing or casing wall

D: internal diameter of tubing or casing

x: segment height

T_e : surrounding temperature of the pipeline (from the temperature gradient)

T_{fl} : average fluid temperature within the segment

Q_H : heat transferred per hour

The expression above can be used to compute the heat loss from the flowing fluids to the surroundings for any infinitesimal pipe segment. Such a length ensures that the temperature values are constant all along the segment. Subsequently we assume that the energy transferred is obtained from the heat generated when the fluid's temperature drops. This heat is calculated by multiplying the average heat capacity of each phase by the mass flow rate of the phase and then by the temperature drop in the fluid, as shown below:

$$Q_H = (m_{oil}C_{p,oil} + m_{gas}C_{p,gas} + m_{water}C_{p,water})(T_{in} - T_{out}) = mC_p T$$

where:

m_x : the mass of each phase inside the segment

$C_{p,x}$: the thermal capacity of each phase

With the equation of heat transferred and heat obtained from the temperature drop, a single equation is derived which gives the rate of temperature change along depth:

$$\frac{dT}{dx} = -\frac{U\pi D}{\dot{m}C_p}[T - T_{a1} + G_2(x - L_1)\cos\theta]$$

where:

T_{a1} : ambient temperature at L_1

T: average fluid temperature in the segment

θ : deviation angle

\dot{m} : fluid mass flow rate

Cp: weighted average specific heat capacity for all the phases

U: overall heat transfer coefficient referred to pipe inside diameter

D: pipe inside diameter

G: geothermal gradient

By integrating the above equation at a desired depth the temperature can be found and used in further calculations.

Overall Heat Transfer Coefficient

This value is used to calculate heat loss in surface or downhole equipment. It accounts for the heat flow through the pipe, annulus (inside the well) and insulation (if present) to the surroundings (referred to the pipe inside diameter). Heat transfer by convection, conduction and radiation are all accounted in this value.

The Rough Approximation model is used in the calculations of the following chapters mainly for the reason that the properties of the surrounding sediments were not available. .

Enthalpy Balance Model

The enthalpy balance model utilizes the classic energy equation for flowing fluid (Bernoulli's equation) in terms of enthalpy. The heat transfer with the surroundings is neglected and the equation is solved by considering the enthalpy balance across incremental pipe lengths. The enthalpy term includes the effects of pressure (Joule-Thomson effect⁴) and phase changes. This process is iterative and the total heat transfer coefficient is estimated for the temperature and pressure of each iteration step to calculate the heat exchanged. The method solves the energy equation simultaneously for temperature and pressure, while the heat transfer coefficient is used to calculate the heat loss differential within the calculations. The emerging issue is that the heat transfer coefficient is itself a function of temperature of the fluid and the pipeline material. Iterations are therefore needed to find P, T, the heat transfer coefficient and satisfy the enthalpy balance. Finally the surrounding environment heat transfer is inserted by calculating the temperature profile near the wellbore. The profile is dependent on time

and the thermal diffusivity of the formation. Subsequently the equation used is the heat diffusivity equation which account for localized heating of the formation by the fluids.

4.4 HYDRATE CALCULATIONS

With all the required information known the hydrate dissociation curve calculation is possible. To obtain enough pairs of P-T values two methods, of varying accuracy, can be used for calculating the actual temperature for a given pressure. The first method available is a correlation developed by [Munck \(1988\)](#) and the second, more accurate, method is based on the Cubic-plus Association equation of state developed by [Kontogeorgis et al 1999](#). A brief review of the methods is given below and in the Appendix.

Both methods can provide a dissociation curve for any fluid composition. Practically, they are applied to each pipeline segment for which the exact composition of the oil and gas phases together with the pressure and temperature values is available. Then, one can determine whether hydrate forming conditions are achieved. This calculation is done for every segment of the pipeline to examine whether favorable hydrate formation conditions exist to at least one of them.

Munck Correlation

For gas mixtures the [Munck et al, 1988](#) correlation uses the algorithms developed by [Michelsen \(1982\)](#). These algorithms are used to check the stability of the mixture, ensuring correct prediction of the number of present phases. The phase envelope calculations are using algorithms developed also by [Michelsen \(1980\)](#). Finally, for the heavy hydrocarbon fraction (C7+) the method utilizes the characterization procedure developed by [Pedersen \(1985\)](#). Summarizing the method combines existing models for gas hydrates with well-known liquid models like UNIQUAC and the Soave-Redlich-Kwong equation of state.

The correlation is a simple but very quick and robust method for checking complex flow regimes for the risk of hydrate formation. At any conditions the method provides answers to important questions such as the probability of hydrate formation and the possible change of required conditions due to the inclusion of inhibiting molecules.

The method can also include various inhibitors in its calculations, such as methanol (MEOH), ethanol, ethylene glycol (MEG) and triethylene glycol (TEG). Moreover the inclusion of ionic salts like NaCl is also a possibility.

Cubic-Plus-Association Equation

CPA (Cubic-Plus-Association) is an equation of state, developed by [Kontogeorgis et al 1999](#), which is based on a combination of the Soave–Redlich–Kwong (SRK) for describing the physical interactions with the Wertheim’s first-order perturbation theory, which can be applied to different types of hydrogen-bonding compounds.

The development of CPA started in 1995 as a research project funded by Shell (Amsterdam), and the model was first published in 1996. Since then, it has been successfully applied to a variety of complex phase equilibria, including mixtures containing alcohols, glycols, organic acids, water, and hydrocarbons. Focus has been placed on cases of industrial importance, e.g., systems with gas-hydrate inhibitors (methanol, glycols), glycol regeneration and gas dehydration units, oxygenate additives in gasoline, alcohol separation, etc.

In summary, the thermodynamics model uses the CPA-EoS and classical mixing rules for fugacity calculations in all fluid phases. The CPA-EoS in terms of pressure P is given by [Kontogeorgis et al, 1995](#).

$$P = \frac{RT}{v-b} - \frac{a}{v(v+b)} - \frac{1}{2} \frac{RT}{v} \left(1 + \rho \frac{\partial \ln(g)}{\partial \rho} \right) \sum_i x_i \sum_{A_i} (1 - X^{A_i})$$

a : the energy parameter

b : the co-volume parameter (assumed to be temperature independent, in agreement with most published equations of state)

ρ : the molar density of the fluid

g : the simplified expression of the radial distribution function as suggested by Kontogeorgis et al.

X_{Ai} : the mole fraction of pure component i not bonded at site A

x_i : is the mole fraction of component i .

5. CASE STUDIES DESCRIPTION AND SETUP

The objective of this thesis is the sensitivity analysis of the production system characteristics, regarding the hydrate formation conditions under steady state flow. To examine the hydration formation issue two reservoir fluids were utilized. The first one is a typical oil of high API value whereas the other one is a high yield retrograde condensate. Both reservoir fluids are combined to two extreme conditions regions, one onshore arctic and one deep-sea offshore with extended tieback.

The analysis will be run by use of the PROSPER software by PetEx. In the next chapters a brief description of the capabilities of PROSPER will be given. Subsequently, the inputs required for the calculations described above will be introduced. Fluid parameters, hydrate dissociation curves together with the well and surface equipment in the two areas will be discussed.

5.1 PROSPER

Hydrate formation inside a production pipeline implies money and time spent in remediating it. Therefore for the oil industry it is important to develop design and simulation programs able to predict hydrate formation. In the production flowlines it is of ultimate importance to know if, when and where hydrate formation is probable.

IPM (Integrated Production Modeling) is a program suite developed by the **PetEx** (Petroleum Experts) company. As its name indicates this suite provides programs for analyzing, simulating and developing objectives of Petroleum Engineering, like fluid PVT analysis, reservoir simulation, field development and production simulation. In the flow assurance domain and specifically the hydrate formation issue, one of the suite's programs called **PROSPER** is extremely useful.

PROSPER is a well performance, design and optimization program for modeling most types of well configurations. **PROSPER** can assist the production or reservoir engineer to select tubing sizes, pipeline hydraulics and temperatures with accuracy and speed so as to optimize the well deliverability. **PROSPER's** sensitivity calculation feature enables existing well designs to be optimized and the effects of future changes in system parameters to be assessed.

PROSPER is designed to allow building of reliable and consistent well models, with the ability to address each aspect of well bore modeling; PVT (fluid characterization), VLP correlations (for calculating pipeline pressure loss) and IPR (reservoir inflow). By modeling separately each component of the producing well system, the user can verify

each model's subsystem by performance matching. Once a well system model has been tuned to real field data, **PROSPER** can be confidently used to model the well under different scenarios and to make forward predictions based on surface production data.

PROSPER gives the user the ability to predict the exact temperature and pressure distribution along the system (from the perforations level to the separator). That is provided the given PVT data for the fluid, the reservoir and well design parameters are accurate and tuned against real production data. Up until IPM 9.5 version the design and prediction in **PROSPER** can include only steady state flow. Transient phenomena, more important in designing the system (like shut-in and start-up) especially against hydrate formation will be able to be modeled in the IPM 10 version.

Another program used indirectly during **PROSPER** calculations is IPM's **PVTp**. **PROSPER** uses this program to calculate fluid properties of a given fluid composition as well as to calculate hydrate dissociation curves for mixtures with salts and thermodynamic inhibitors, by use of an Equation of State model.

5.2 MODELS SETUP ON PROPER

5.2.1 GAS CONDENSATE

Fluid Description	
Fluid	Retrograde Condensate
Method	Equation of State
Eq. of State	PROSPER Internal EOS model
Separator	Multi-Stage Separator
EOS Setup	
PVT Warnings	Enable Warning
Water Viscosity	Use Default Correlation
Water Vapour	No Calculations

Figure 57: Prosper options on fluid description (Style for images and captions?)

The retrograde gas condensate's composition is given in [Table 5](#) (Red). Over 75% of the gas condensate consists of hydrate forming molecules (CO_2 , CH_4 , C_2H_6 (subscripts), C_3H_8). Moreover, the heavy end concentration (C_7+) is more than 6% of the total gas composition which results to significant condensate production and therefore multiphase flow in the pipeline. The salinity of the condensed water is 20,000 ppm ([Table 5 - Orange](#)) which is enough to impose a slight inhibiting effect to the hydrate formation conditions. The viscosity models were calculated using the Lohrenz, Bray Clark¹ correlation for both

the oil and gas phases (Table 5 - Blue). Finally the fluid's phase envelope has been calculated by use of the Peng-Robinson² cubic Equation of State (Table 5 - Red) and it is shown in Table 6 (Blue) and Figure 59 (Blue).

Table 5: Retrograde gas condensate composition

The screenshot shows the 'Edit Composition' window with the following data and settings:

	Name	Mole Percent (percent)	Critical Temp. (deg F)	Critical Pressure (psig)	Critical Volume (ft ³ /lb.mole)	Acentric Factor	Molecular Weight	Specific Gravity (sp. gravity)	Boiling Point (deg F)	Volume Shift	Ome
1	N2	0.31	-233.124	477.326	1.43842	0.039	28.01	1.026	-320.37	0	0.4
2	CO2	2.33	87.672	1058.26	1.50409	0.239	44.01	1.101	-109.23	---	0.4
3	C1	68.73	-103.716	615.435	1.58899	0.011	16.04	0.415	-258.81	0	0.4
4	C2	12.37	20.8662	635.504	2.37547	0.099	30.1	0.546	-127.41	0	0.4
5	C3	5.01	23.255	335.154	3.25166	0.153	44.1	0.585	-43.71	0	0.4
6	NC4	2.71	434.907	385.869	4.08459	0.199	58.1	0.6	31.17	0	0.4
7	NC5	1.4	445.786	385.403	4.86948	0.251	72.2	0.63	96.87	0	0.4
8	C6	0.96	429.689	370.622	5.92667	0.299	86.2	0.664	155.73	0	0.4
9	C7+:12	3.43	476.292	496.024	7.4259	0.2904	121.581	0.7628	228.102	0	0.4
10	C13+:C35	2.75	906.115	262.76	16.0286	0.6232	252.73	0.8692	590.221	0	0.4

Options:

- Change
- EOS Model: Peng Robinson
- Optimisation Mode: Medium
- Separator Calc Mode: Use Separator Train
- Oil Viscosity Mode: Lohrenz, Bray Clark
- Gas Viscosity Mode: Lohrenz, Bray Clark
- Volume Shift: Full Composition: No Volume Shift

Separator Information:

Stage	Pressure (psig)	Temperature (deg F)
1	500	100
2	100	90
3	0	60
4		

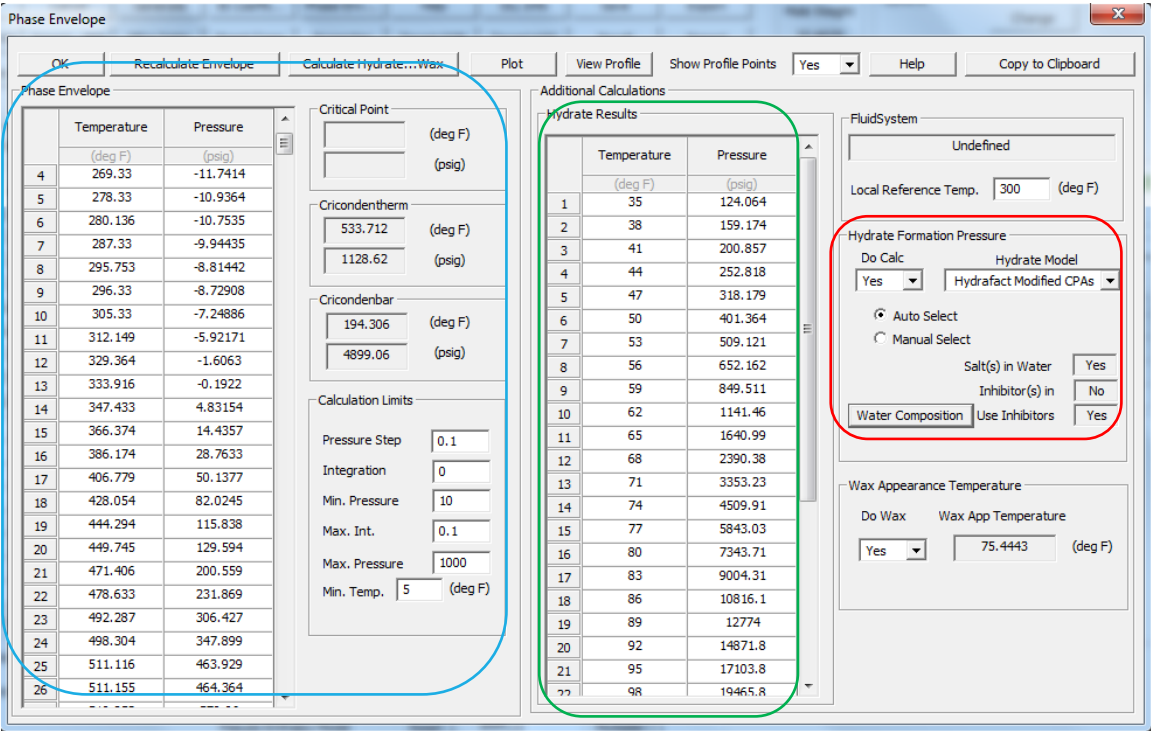
Calculation Settings:

- Viscosity Mode OIL: Lohrenz, Bray Clark
- Viscosity Mode GAS: Lohrenz, Bray Clark
- Pseudo Enthalpy Model
- Reset
- Shift: 0
- Multiplier: 1
- Calculation Type: Calculated From EOS Model
- Reservoir Temperature: 300 (deg F)
- Salinity: 20000 (ppm)
- Water Composition

The hydrate dissociation (P-T) curve (Table 6 - Green) has been calculated by use of the Hydract Modified CPA³ equation (Table 6- Red) and is displayed in Figure 60.

PROSPER provides an option to include various inhibitors to the hydrate formation curve calculations. All common thermodynamic inhibitors are available (Figure 58 - Red), as well as various kinds of salts (Figure 58 - Green) that act as inhibitors to the producing fluid. The program also takes into account the fact that some inhibiting molecules, such as ethanol may also act as hydrate formers (Figure 58). One disadvantage of the program is that it cannot provide reliable results when high percentages (over 30%) of thermodynamic inhibitors are used. More specifically the number of the dissociation curve points produced are not enough for a reliable phase envelope.

Table 6: Retrograde gas condensate phase envelope and hydrate dissociation curve



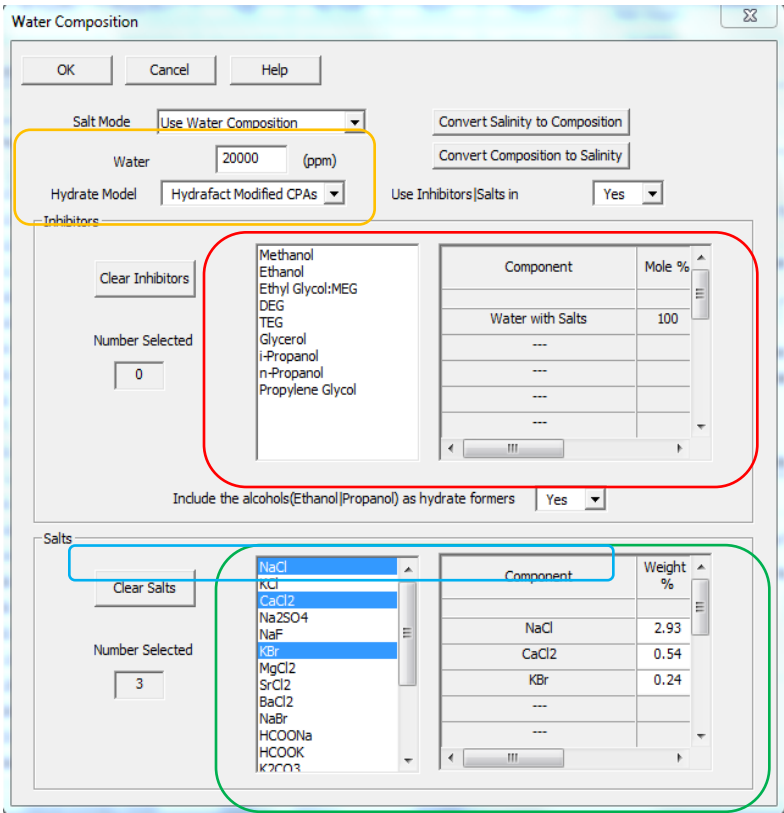


Figure 58: Inclusion of hydrate thermodynamic inhibitors in PROSPER

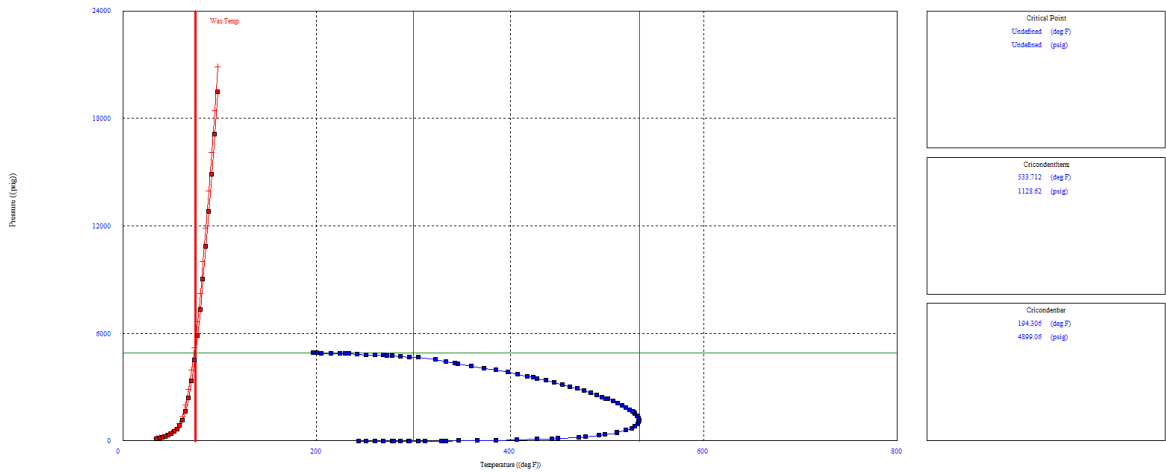


Figure 59: Phase envelope of the retrograde gas condensate (blue) and hydrate dissociation curve for the same fluid (red)

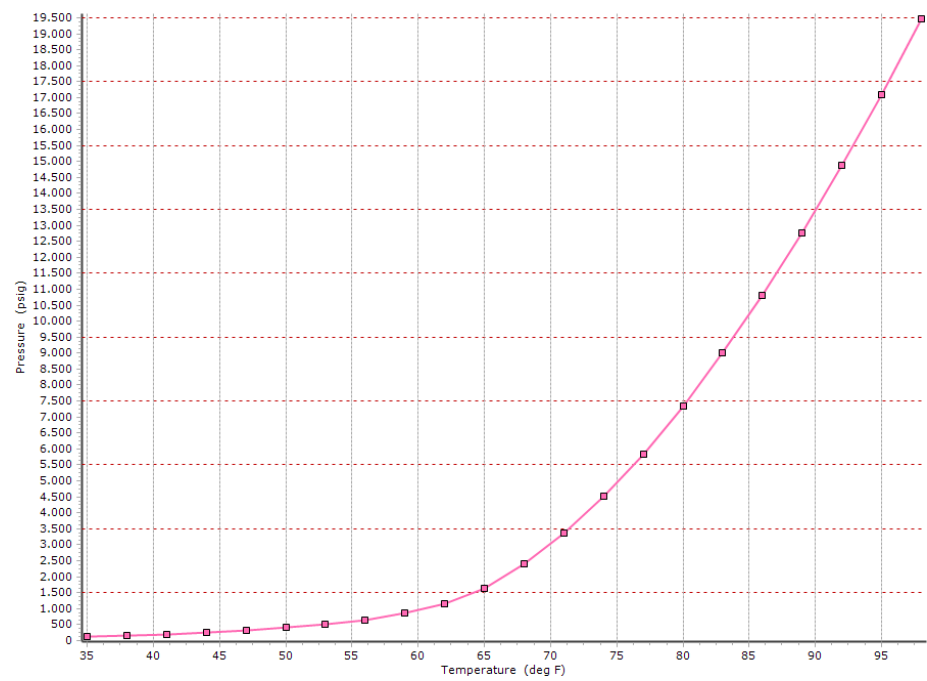


Figure 60: Hydrate dissociation curve (hydrates are stable on the left side of the line)

5.2.2 OIL

Fluid Description	
Fluid	Oil and Water
Method	Equation of State
Eq. of State	PROSPER Internal EOS model
Separator	Multi-Stage Separator
EOS Setup	
PVT Warnings	Enable Warning
Water Viscosity	Use Default Correlation

Figure 61: Prosper options on fluid description

The reservoir fluid composition is given in [Table 7](#) (Red) and [Table 8](#). More than 55% of the oil consists of hydrate forming molecules (CO_2 , CH_4 , C_2H_6 , C_3H_8) and this illustrates the importance of designing with hydrate management in mind. The heavy end's (C_7+) concentration is also over 30% of the total oil composition ([Table 8](#)). The large percentage of light components results to additional problematic conditions, such as high gas oil ratio (GOR) as this leads to multiphase flow in the pipeline with very high gas phase occupancy. This apparently leads to increased hydrate formation danger. The salinity of the produced fluid is 20,000 ppm ([Table 7 - Orange](#)), which is enough to have a slight inhibiting effect on the hydrate formation conditions (23 – 1800 psig formation pressure decrease for temperatures ranging from 35°F to 80°F respectively). As it was the case with the gas condensate, the viscosity models were calculated from the Lohrenz, Bray Clark¹ correlation for both oil and gas phase ([Table 7 - Blue](#)). Similarly, the fluid's phase envelope has been calculated by use of the Peng Robinson² cubic Equation of State ([Table 7 - Red](#)) and it is displayed in [Table 8](#) (Blue) and [Figure 63](#) (Blue).

Assessment of Hydrate Formation Parameters in Production Wells

Table 7: Reservoir fluid composition (Light Oil)

Done Cancel Generate BI Coeffs... Phase Env... Help DLL Info Save Export Mole Weight 92.9836

Export...PRP Import...PRP Fill in Table Reset Comp Properties Target GOR Target H2O Recall Report

Interpolate Hydrate

	Name	Mole Percent (percent)	Critical Temp. (deg F)	Critical Pressure (psig)	Critical Volume (ft ³ /lb.mole)	Acentric Factor	Molecular Weight	Specific Gravity (sp. gravity)	Boiling Point (deg F)	Volume Shift	Ome
1	N2	0.38	-233.104	477.326	1.43842	0.039	28.01	1.026	-320.35	-0.154	0.4
2	CO2	1.37	87.692	1058.26	1.50409	0.239	44.01	1.101	-109.21	---	0.4
3	C1	43.45	-116.518	658.381	1.58899	0.011	16.04	0.415	-258.79	-0.22029	0.4
4	C2	7.84	89.798	693.651	2.37547	0.099	30.1	0.546	-127.39	-0.35649	0.4
5	C3	5.3	206.006	602.683	3.25166	0.153	44.1	0.585	-43.69	-0.34962	0.4
6	IC4	1.14	274.694	514.36	4.21274	0.183	58.1	0.6	10.85	-0.9292	0.4
7	NC4	2.93	305.294	535.963	4.08459	0.199	58.1	0.6	31.19	-0.36177	0.4
8	IC5	1.16	369.806	468.362	4.90151	0.227	72.2	0.621	82.13	-0.47502	0.4
9	NC5	1.6	385.592	474.828	4.86948	0.251	72.2	0.63	96.89	-0.32487	0.4
10	C6	2.23	454.1	425.008	5.92667	0.299	86.2	0.664	155.75	-0.13409	0.4
11	C7::C8	5.12848	595.316	410.04	7.84176	0.32717	126.36	0.77843	272.794	0.014424	0.4
12	C9::C9	2.85067	692.923	345.416	10.3317	0.4095	149.021	0.8084	368.285	0.036054	0.4
13	C10::C11	5.13645	753.634	314.709	11.6649	0.46148	172.048	0.82869	428.334	0.08753	0.4
14	C12::C13	4.28607	834.22	273.816	13.8026	0.54218	202.554	0.85181	512.849	0.11869	0.4
15	C14::C15	3.50238	922.585	232.241	16.6042	0.64465	232.416	0.87445	610.701	0.14422	0.4
16	C16::C17	2.83665	991.091	204.081	19.0287	0.7326	261.612	0.89134	689.134	0.16437	0.4
17	C18::C19	2.28886	1041.45	185.663	20.921	0.80166	290.135	0.90352	747.883	0.18021	0.4
18	C20::C21	1.84484	1082.77	172.8	22.4446	0.85906	317.981	0.91405	795.885	0.19268	0.4

Options: Change
EOS Model: Peng Robinson
Optimisation Mode: Medium
Separator Calc Mode: Use Separator Train
Oil Viscosity Mode: Lohrenz, Bray Clark
Gas Viscosity Mode: Lohrenz, Bray Clark
Volume Shift: Full Composition: Use Volume Shift

Separator Information

Stage	Pressure (psig)	Temperature (deg F)
1	500	150
2	100	120
3	0	60
4		

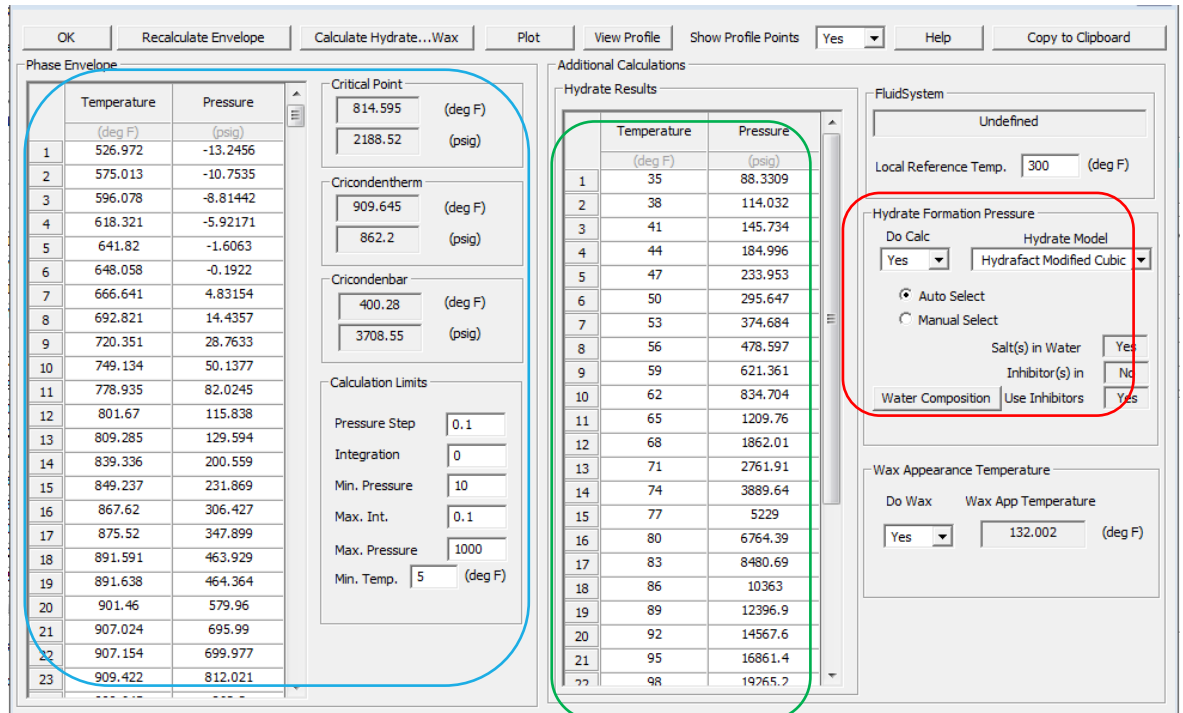
Viscosity Mode OIL: Lohrenz, Bray Clark
Viscosity Mode GAS: Lohrenz, Bray Clark
Pseudo Enthalpy Model
Calculation Type: Calculated From EOS Model
Reservoir Temperature: 300 (deg F)
Salinity: 20000 (ppm)
Water Composition

Oil composition	
N2	0,38
CO2	1,37
C1	43,45
C2	7,84
C3	5,30
IC4	1,14
NC4	2,93
IC5	1,16
NC5	1,60
C6	2,23
C7::C8	5,12848
C9::C9	2,85067
C10::C11	5,13645
C12::C13	4,28607
C14::C15	3,50238
C16::C17	2,83665
C18::C19	2,28886
C20::C21	1,84484
C22::C24	2,12030
C25::C30	2,60531

Table 8: Oil composition

The hydrate dissociation (P-T) curve (Table 9 – Green) is calculated by use of the Hydract Modified CPA³ equation (Table 9 - Red) and is displayed in Figure 64.

Table 9: Light Oil phase envelope and hydrate dissociation curve



OK Cancel Help

Salt Mode: Use Water Composition (ppm) 20000
Hydrate Model: Hydract Modified Cubic
Use Inhibitors/Salts in: Yes

Convert Salinity to Composition
Convert Composition to Salinity

Inhibitors

Clear Inhibitors
Number Selected: 0

Methanol
Ethanol
Ethyl Glycol:MEG
DEG
TEG
Glycerol
i-Propanol
n-Propanol
Propylene Glycol

Component Mole %
Water with Salts 100

Include the alcohols(Ethanol|Propanol) as hydrate formers: Yes

Salts

Clear Salts
Number Selected: 3

NaCl
KCl
CaCl2
Na2SO4
NaF
KBr
MgCl2
SrCl2
BaCl2
NaBr
HCOONa
HCOOK
K2CO3

Component Weight %
NaCl 3.52
KCl 0.95
KBr 0.56

As stated previously, **PROSPER** provides an option to include several different inhibitors in the hydrate formation curve calculations. All the common thermodynamic inhibitors are available (Figure 62 – Red), as well as different kinds of salts that act as inhibitors in the produced fluid. In Figure 62 (Green) the composition of salts in the produced fluid, that act as inhibitors, is inserted.

Figure 62: Inclusion of hydrate thermodynamic inhibitors in PROSPER

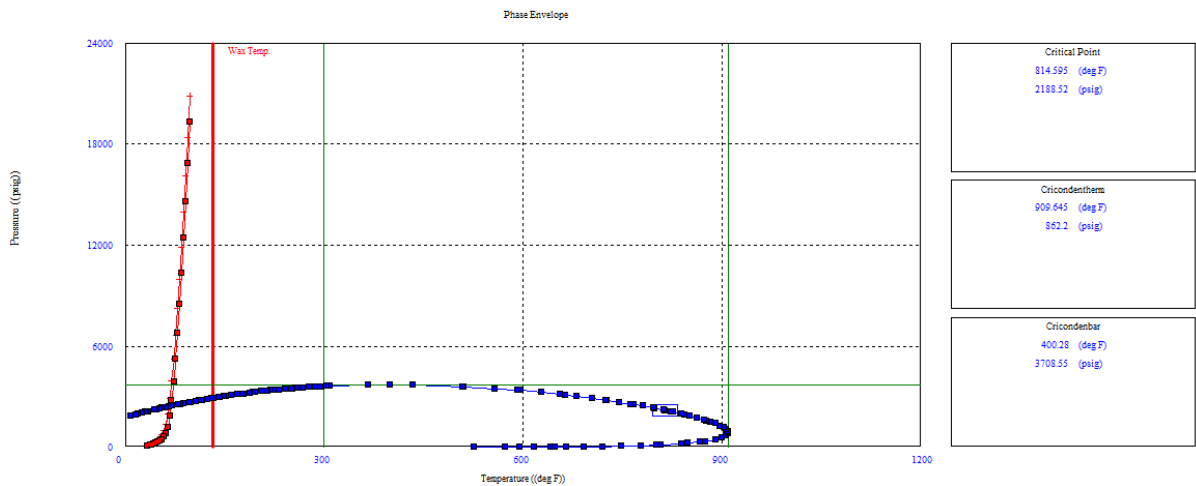


Figure 63: Phase envelope of the light oil (blue) and its hydrate dissociation curve (red)

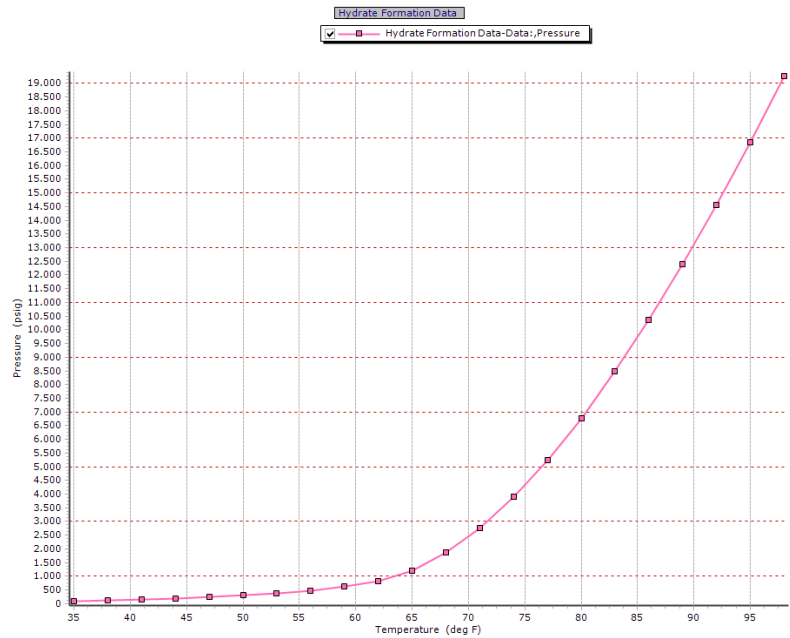


Figure 64: Hydrate dissociation curve (hydrates are stable on the left side of the line)

5.2.3 ARCTIC ONSHORE REGION

The areas chosen were one onshore arctic and one deep-sea offshore with extended tiebacks, as they both exhibit increased risk of hydrate formation. The setup of the arctic onshore model in **PROSPER** will be described in this section.

To set up the arctic well's model, some parameters regarding the equipment installed and the pipelines needed to be inserted. From the System menu and through the Equipment tab these parameters are configured.

Before inserting the characteristic of the arctic region the values of the hydrate dissociation curve need to be inserted into the program. The values for pressure and temperature of the hydrate dissociation curve were inserted in the PVT warnings interface ([Table 10](#)), where also other values can be entered for several flow assurance issues. As it can be seen from [Table 10](#) the range of P-T values is reaching very high pressure values (19,465 psia) that are never encountered in the production flowline. The temperature range of the dissociation curve, suspect for hydrate forming, encountered inside the pipelines is between 0-80 °F. Higher temperatures do not pose danger as the required pressures are very high and never encountered inside the pipeline. The condition that prevents hydrate forming at steady state conditions is the velocity of the fluid inside the flowline. This limits the contact time for the fluid heat to be transferred to the environment, thus not allowing the temperature to fall below the dissociation one.

The PVT warning for hydrate forming conditions was enabled to show, for the solution flow rate, where in the flowline the conditions are favorable for hydrate formation ([Figure 65](#)).

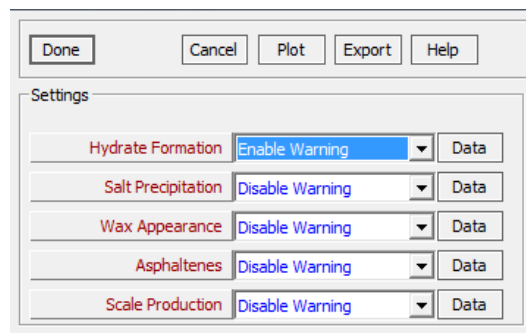


Figure 65: PVT warnings introduction window

Table 10: Hydrate dissociation curve P-T for uninhibited flowline (Oil - Retrograde gas condensate)

Done Main Cancel Import Export Plot Help		
Point	Pressure (psig)	Temperature (deg F)
1	88.3309	35
2	114.032	38
3	145.734	41
4	184.996	44
5	233.953	47
6	295.647	50
7	374.684	53
8	478.597	56
9	621.361	59
10	834.704	62
11	1209.76	65
12	1862.01	68
13	2761.91	71
14	3889.64	74
15	5229	77
16	6764.39	80
17	8480.69	83
18	10363	86
19	12396.9	89
20	14567.6	92
21	16861.4	95
22	19265.2	98

Done Main Cancel Import Export Plot Help		
Point	Pressure (psig)	Temperature (deg F)
1	124.064	35
2	159.174	38
3	200.857	41
4	252.818	44
5	318.179	47
6	401.364	50
7	509.121	53
8	652.162	56
9	849.511	59
10	1141.46	62
11	1640.99	65
12	2390.38	68
13	3353.23	71
14	4509.91	74
15	5843.03	77
16	7343.71	80
17	9004.31	83
18	10816.1	86
19	12774	89
20	14871.8	92
21	17103.8	95
22	19465.8	98
23		
24		

For the description of the arctic region the first important parameter needed to be set is the Top Node Pressure. In the specific case the top node was set to the pressure of the first separator of a two-stage separator train. This pressure was initially set 500 psig for the first separator and 100 psig for the second (Table 5 – Green) but it is subject to sensitivity analysis scenaria and it varies per case.

One of the required configurations is the Deviation Survey. From this option the user can determine the measured depth (MD) of the well, along the true vertical depth (TVD) reached (Table 11). This is paramount as to allow the software to understand the path of the well and be able to introduce values regarding angle and length into the calculations. Several points can be entered, whenever the inclination angle changes, until the wellbore reaches the payzone of the reservoir. Then from the program's interface a plot of the well trajectory can be drawn. In this specific scenario the well is a vertical one, with the measured depth being equal to the true vertical one all along. The well's depth is set at 12,000 ft (Figure 66).

Table 11: Deviation survey window (Arctic region)

Done

Cancel

Main

Help

Filter

Plot

MD <-> TVD

Calculate

Input Data

Point	Measured Depth (feet)	True Vertical Depth (feet)	Cumulative Displacement (feet)	Angle (degrees)
1	0	0	0	0
2	12000	12000	0	0
3				

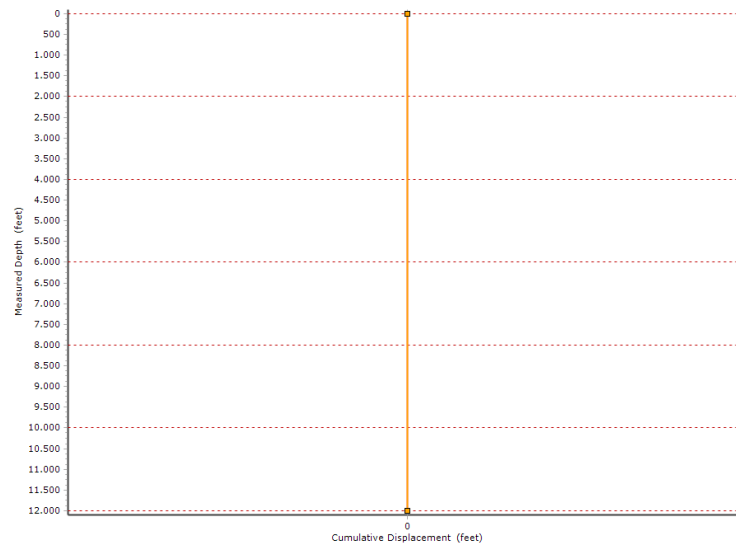


Figure 66: Plot of Measured depth versus cumulative displacement from the vertical axis (Arctic region)

The next configuration option is the description of the installed Downhole Equipment, i.e. casing tubing and restrictions. The description of this equipment is vital because it is related to the significant pressure drops or to intense thermal effects due to the Joule-Thomson⁴ effect on the downstream of these points. The effect of such flow restricting equipment is not calculated by use of mechanistic models. Instead, the pressure drop is calculated from correlations provided by the software (Beggs and Brill⁵ and Petroleum Experts 2⁶). In this specific scenario the equipment installed consists of: the wellhead at 200 ft (MD), a tubing of 4 inches inner diameter from 200 down to 1,000 ft, a subsurface safety valve (SSSV) at 1,000 ft depth with internal diameter of 3.5

inches, another tubing of 4 inches inner diameter through to 11,800 ft and finally the cased hole of 5.4 inches inner diameter through the payzone, from 11,800 to 12,000 ft (Table 12). The formation is isolated with packers from the wellbore’s annular space and all flow occurs through the tubing solely.

A sketch of the configuration is given in Figure 67 where the previously described equipment is depicted.

Table 12: Installed Downhole Equipment (Arctic region)

Input Data									
Point	Label	Type	Measured Depth (feet)	Tubing Inside Diameter (inches)	Tubing Inside Roughness (inches)	Tubing Outside Diameter (inches)	Tubing Outside Roughness (inches)	Casing Inside Diameter (inches)	Casing Inside Roughness (inches)
1		Xmas Tree	200						
2		Tubing	1000	3.992	0.0018				1
3		SSSV		3.5					1
4		Tubing	11800	3.992	0.0018				1
5		Casing	12000					6.4	0.0006
6									
7									
8									
9									
10									
11									
12									
13									
14									
15									
16									
17									
18									
19									
20									

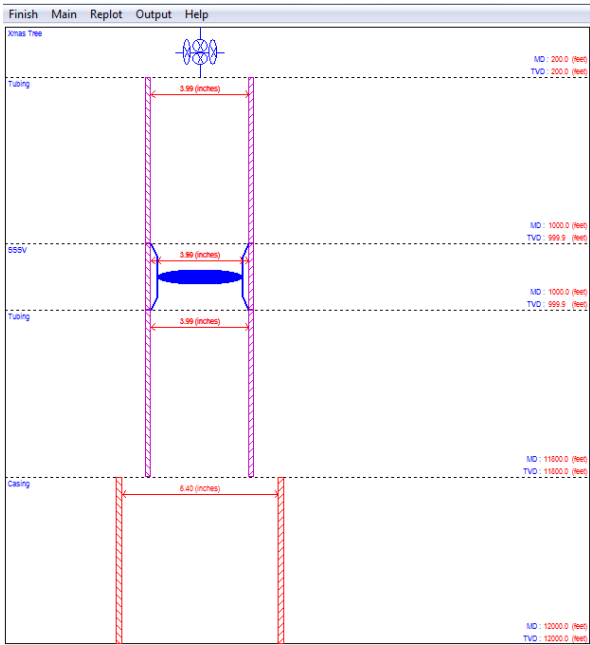


Figure 67: Sketch of Downhole Equipment configuration (Arctic region)

The system configuration description continues with the surface equipment design (after the wellhead, till the separator). In Table 13 the pipelines' length and inclination are inserted via their measured depth (MD) and true vertical depth (TVD). The scenario described has a surface pipeline of 16,000 ft (MD) total length, total vertical depth of 365 ft (TVD) and consequently very low inclination angle of $1.26^\circ - 1.34^\circ$ (Table 13). Also from Table 13, the initial overall heat transfer coefficient value of the pipeline is set at 3 BTU/h/ft²/°F (due to the insulation effect provided by the pipe material). Finally the average ambient temperature of the surrounding environment is set at 0 °F (arctic environment). From this information a deviation diagram of the surface pipeline is shown in Figure 68 and a sketch of the pipeline diameter and equipment in Figure 69.

Table 13: Surface Equipment and Pipelines configuration (Arctic region)

Done

Cancel

Main

Import

Export

Report

Plot

Pipe Schedule

Help

Coordinate System

TVD, Length

Temperature of Surroundings

0

deg F

Choke Method

ELF

Overall Heat Transfer Coefficient

3

BTU/h/ft²/F

Input Data

Point	Label	Type	Pipe Length	True Vertical Depth	Pipe Inside Diameter	Pipe Inside Roughness	Rate Multiplier
			(feet)	(feet)	(inches)	(inches)	
1		Manifold		0			
2		Pipe	7500	165	4	0.0006	1
3		Pipe	8500	200	4	0.0006	1

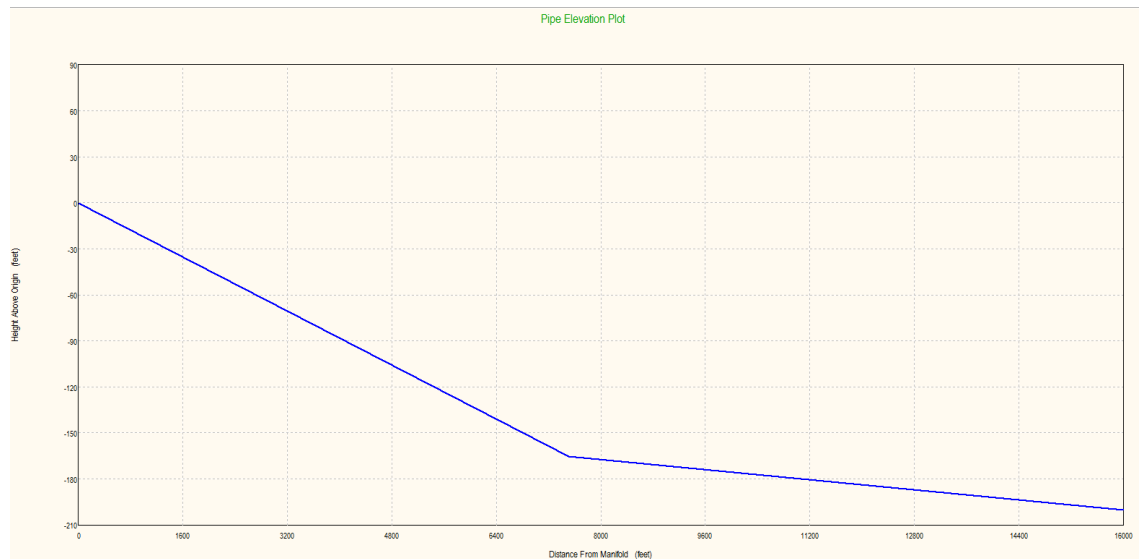


Figure 68: Deviation Survey diagram of the Surface Equipment (Arctic region)

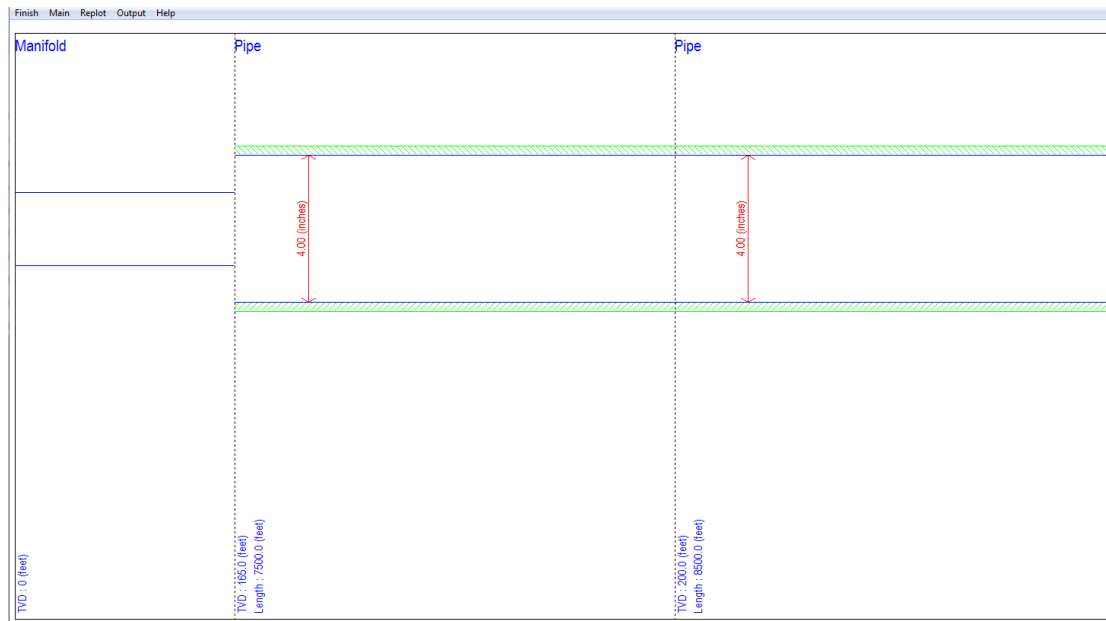


Figure 69: Sketch of the Surface Equipment (Arctic region)

Another piece of information about the system is the temperature gradient in the sedimentary rocks, from the reservoir to the surface, as well as the overall heat transfer coefficient of the tubing inside the wellbore (5 BTU/h/ft²/°F - Table 14). In the arctic region the temperature gradient follows the profile shown in Figure 70, derived by use of the temperature information measured at each depth ($T_{\text{res}} = 300$ °F, $T_{\text{sediment}} = 150$ °F and $T_{\text{surf}} = 0$ °F), provided from Table 14. Finally the average heat capacities of the production fluids are entered. As shown in Figure 71 the oil has a value of 0.53 BTU/lb/°F, the gas 0.51 BTU/lb/°F and the water 1 BTU/lb/°F.

Table 14: Geothermal Temperature Gradient (Arctic region)

Point	Formation TVD (feet)	Formation Measured Depth (feet)	Formation Temperature (deg F)
1	200	200	0
2	4000	4000	146.67
3	9500	9500	253
4	12000	12000	300
5			

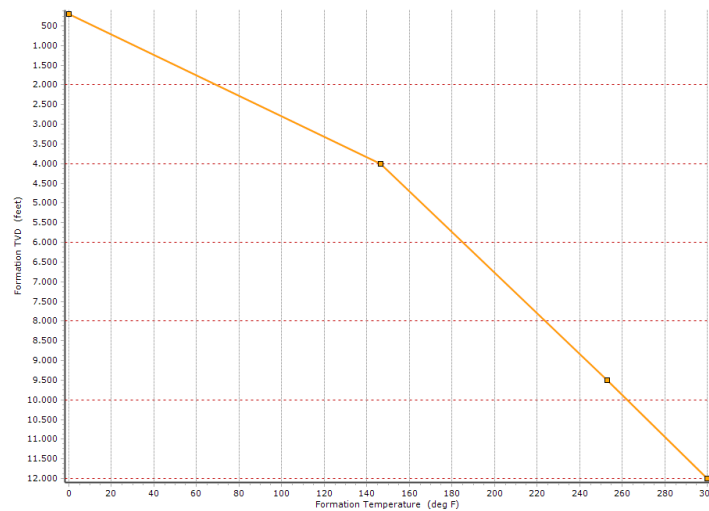


Figure 70: Geothermal Gradient diagram (Arctic region)

Input Parameters		
Cp Oil	0.53	BTU/lb/F
Cp Gas	0.51	BTU/lb/F
Cp Water	1	BTU/lb/F

Figure 71: Average Heat capacities of the production fluids

The last piece of information needed to describe the arctic system configuration is the description of the Inflow Performance Relationship (IPR) of the reservoir. As explained in Chapter 4, IPR is required for calculating the production rate as the latter can be obtained from the intersection of Inflow Performance Relationship (IPR) and the Vertical Lift Performance (VLP) curves by use of the nodal analysis method.

The necessary information required for the IPR calculation is related to the reservoir properties. As explained before these parameters are needed to calculate the Productivity index. To obtain a reliable IPR line the reservoir must be described accurately. This is done from the System menu of the **PROSPER** under Inflow Performance (Figure 72). The reservoir model, depending on the information available, can be generated from various correlations and/or theoretical models. If the information is sparse it is done with less accurate correlations that require less information. Accordingly, more detailed correlations provide better results but also require a big amount of information. Options are also available to test the correlation versus real reservoir data from well tests and tune the model against them. As this

configuration was done for experimental purpose and it does not apply to a real case scenario the reservoir was described with one of the simpler correlation methods implemented in the program (Petroleum Experts 4⁷) without matching. The reservoir is single branched, with a cased hole completion and no sand control implemented. Also the skin factor is entered by hand and set at 5 (Figure 72 – Bottom). The numerical data inserted regarding the reservoir (Figure 72) for this correlation are:

- Initial Reservoir Pressure (6,000 psig)
- Reservoir Temperature (300 °F)
- Initial Water Gas ratio (10 STB/MMscf)
- GOR (6,943.98 scf/STB)
- Permeability (25 md)
- Reservoir Thickness (100 ft)
- Drainage Area (340 acres)
- Dietz Shape factor (31.6)
- Wellbore radius (0.354 ft)
- Perforation Interval (100 ft)
- Time since production started (50 days)
- Reservoir Porosity (0.25)
- Connate Water Saturation (0.2)

The IPR curve calculated from the software, is displayed in the results window (Figure 72) and plotted graphically in Figure 73 (Reservoir Pressure – Flow rate).

5. Case Studies Description and Setup

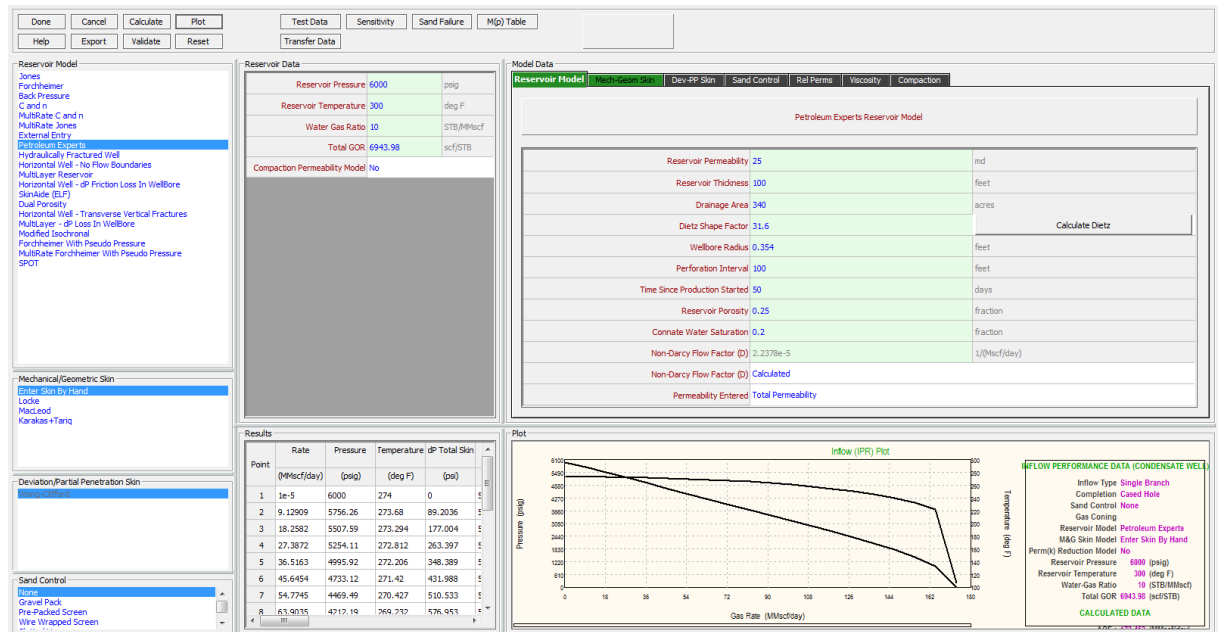


Figure 72: Reservoir characteristics configuration and IPR curve calculation from correlations (Arctic region)

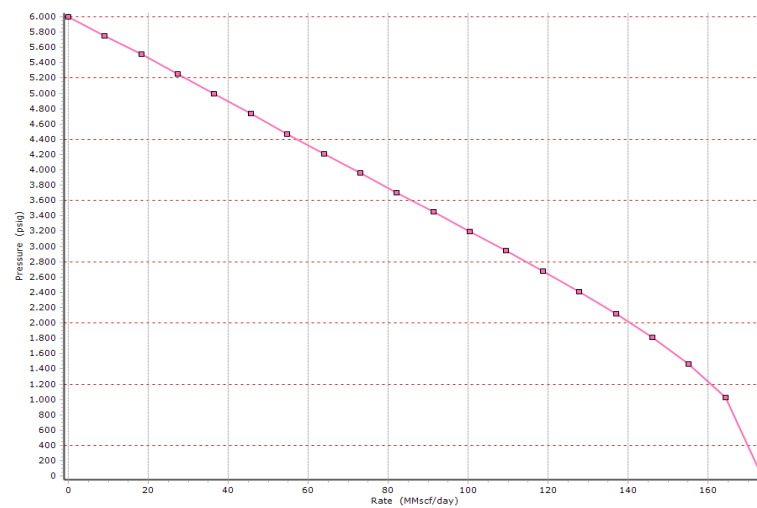


Figure 73: IPR curve diagram (Arctic region)

5.2.4 DEEP OFFSHORE REGION

The second region chosen is a deep-sea offshore area with extended tiebacks, due to their increased risk of hydrate formation. The set up in **PROSPER** of the offshore model will be explained in this chapter.

To set up the offshore well's model, some parameters regarding the equipment and the pipelines installed need to be inserted. From the System menu and through Equipment tab these parameters are configured.

Same as in the previous chapter, before inserting the characteristic of the offshore region the values of the hydrate dissociation curve need to be inserted into the program. The values for pressure and temperature of the hydrate dissociation curve were inserted in the PVT warnings interface ([Table 15](#)), where also other values can be entered for several flow assurance issues. As displayed in [Table 15](#) the P-T values have a value range exhibiting very high pressures (19,465 psia) that are never encountered in the system. The important temperature range associated with pressure values encountered in the system is 0 to 80°F. The condition that prevents hydrate forming, in steady state conditions, is the velocity of the fluid inside the flowline. High velocity implies less time spent inside the production tubing that does not allow the heat transfer from the fluid to the environment to be enough for the temperature to fall below the dissociation curve.

The PVT warning for hydrate forming conditions was enabled where the flowline conditions are favorable for hydrate formation ([Figure 74](#)). The plot of the hydrate dissociation curve and the respective fluid Pressure-Temperature diagram is shown in [Figure 75](#).

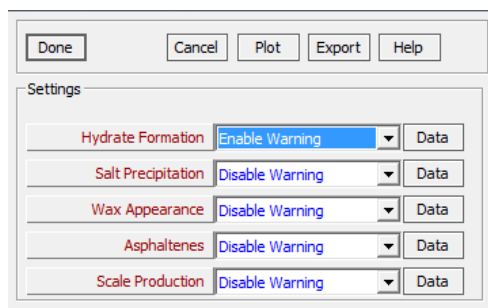


Figure 74: PVT warnings introduction window

Table 15: Hydrate dissociation curve P-T for uninhibited flowline (Retrograde gas condensate - Oil)

Point	Pressure (psig)	Temperature (deg F)
1	124.064	35
2	159.174	38
3	200.857	41
4	252.818	44
5	318.179	47
6	401.364	50
7	509.121	53
8	652.162	56
9	849.511	59
10	1141.46	62
11	1640.99	65
12	2390.38	68
13	3353.23	71
14	4509.91	74
15	5843.03	77
16	7343.71	80
17	9004.31	83
18	10816.1	86
19	12774	89
20	14871.8	92
21	17103.8	95
22	19465.8	98
23		
24		

Point	Pressure (psig)	Temperature (deg F)
1	88.3309	35
2	114.032	38
3	145.734	41
4	184.996	44
5	233.953	47
6	295.647	50
7	374.684	53
8	478.597	56
9	621.361	59
10	834.704	62
11	1209.76	65
12	1862.01	68
13	2761.91	71
14	3889.64	74
15	5229	77
16	6764.39	80
17	8480.69	83
18	10363	86
19	12396.9	89
20	14567.6	92
21	16861.4	95
22	19265.2	98



Figure 75: Hydrate dissociation curve and gas condensate phase envelope

For the description of the offshore region the first parameter needed to be set is the Top Node pressure. In the specific case the top node was set to the pressure of the first separator, belonging in a two-stage separator train. This pressure is initially 500 psig for the first separator and 100 psig for the second (Table 5 – Green) but it is subject to sensitivity analysis scenarios and varies per case.

The next required configuration is the Deviation Survey description. From this option the user can determine the measured depth (MD) of the well along the true vertical depth (TVD) (Table 16). Several points can be entered, whenever the inclination angle changes, until the wellbore reaches the payzone of the reservoir.

Then from the program's interface, a plot of the well trajectory can be drawn. In this specific scenario the well is a vertical one, with the measured depth being equal to the true vertical one. The depth of the well is at 20,200 ft (Figure 76).

Table 16: Deviation survey window (Offshore region)

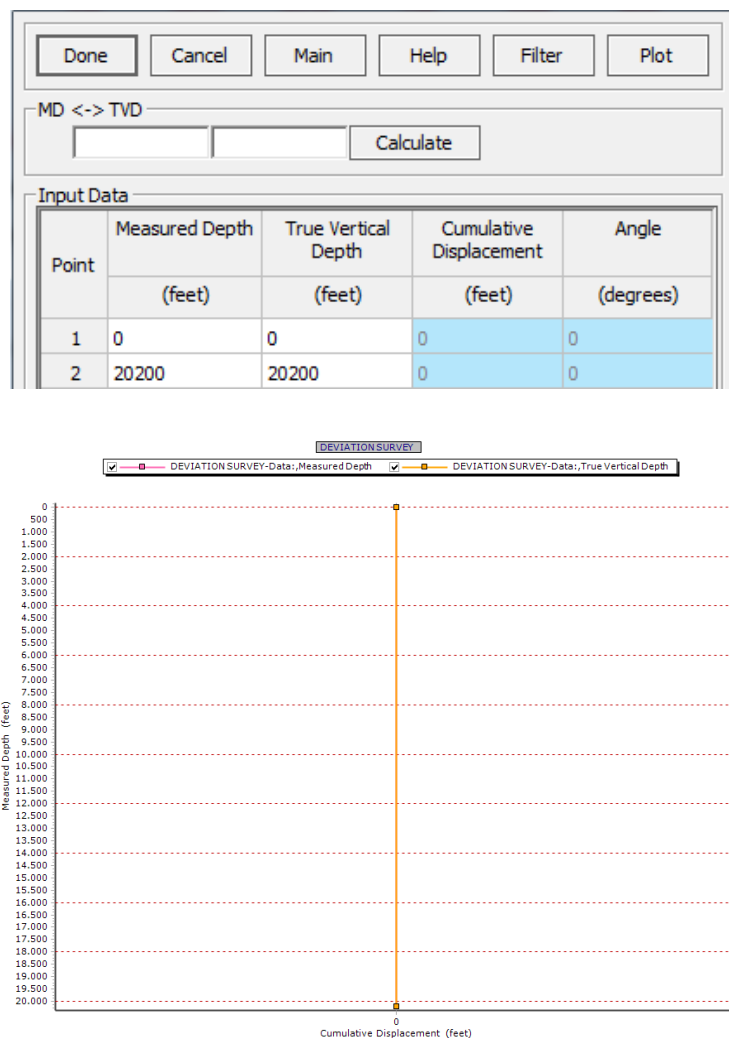


Figure 76: Plot of Measured depth versus cumulative displacement from the vertical axis (Offshore region)

The next configuration option is the insertion of the installed Downhole Equipment i.e. casing tubing and restrictions. This equipment is vital because it is related to the significant pressure drops or to intense thermal effects due to the Joule-Thomson⁴ effect on the downstream of these points. The effect of those flow restricting equipment is not calculated by use of mechanistic models. Instead, the pressure drop is calculated from correlations provided by the software (Beggs and Brill⁵). In the specific scenario the equipment installed are: the Xmas Tree (wellhead) at 8,200 ft (MD), a

tubing of 4 inches inner diameter from 8200 ft to 9200 ft, a subsurface safety valve at 9,200 ft depth with internal diameter of 3.5 inches, another tubing of 4 inches inner diameter through to 20,000 ft and finally a casing with 5.4 inches inner diameter from 20,000 to 20,200 ft (Table 17). The formation is isolated with packers from the wellbore's annular space and the flow occurs solely through the tubing.

A sketch of the configuration is given in Figure 77 where the previously described equipment is depicted.

Table 17: Installed Downhole Equipment (Offshore region)

<div> <div>Done</div> <div>Cancel</div> <div>Main</div> <div>Import</div> <div>Export</div> <div>Report</div> <div>Tubing DB</div> <div>Casing DB</div> <div>Help</div> </div>										
Input Data										
Point	Label	Type	Measured Depth (feet)	Tubing Inside Diameter (inches)	Tubing Inside Roughness (inches)	Tubing Outside Diameter (inches)	Tubing Outside Roughness (inches)	Casing Inside Diameter (inches)	Casing Inside Roughness (inches)	Rate Multiplier
1		Xmas Tree	8200							
2		Tubing	9200	3.992	0.0018					1
3		SSSV		3.5						1
4		Tubing	20000	3.992	0.0018					1
5		Casing	20200					6.4	0.0006	1

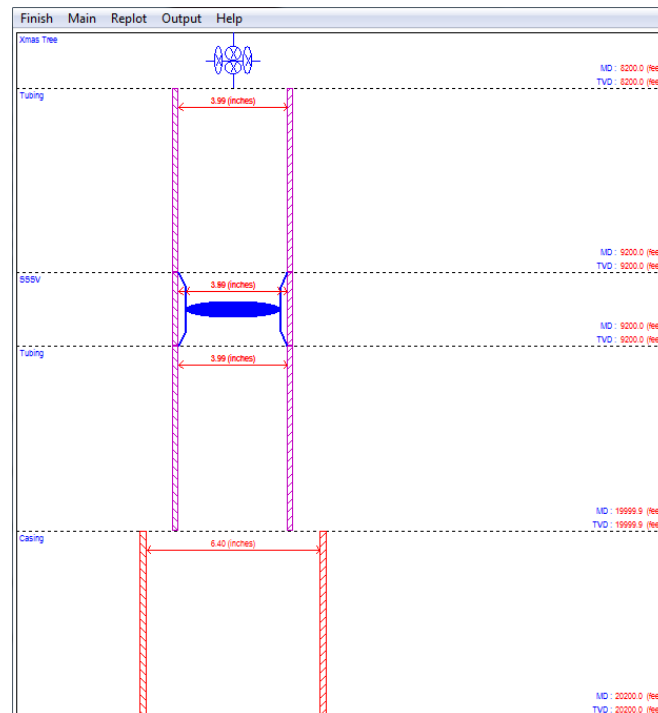


Figure 77: Sketch of Downhole Equipment configuration (Offshore region)

The system configuration continues to the surface equipment design (after the wellhead, till the separator). In Table 18 the pipelines' length and inclination are inserted via the measured depth (MD) and true vertical depth (TVD). The scenario described has a surface pipeline of total length of 26,650 ft (MD), a total vertical depth of 9,850 ft (TVD) and consequently increased angle around 30° (Table 18). Also from Table 18 the initial overall heat transfer coefficient of the pipeline is inserted at 3 BTU/h/ft²/°F (effect of the insulation provided by the pipe material). Finally the average ambient temperature of the surrounding environment is set at 39 °F (typical ocean subsea environment). From this information a deviation diagram of the surface pipeline is shown in Figure 78 and a sketch of the pipeline diameter and equipment in Figure 79.

Table 18: Surface Equipment and Pipelines configuration (Offshore region)

Done

Cancel

Main

Import

Export

Report

Plot

Pipe Schedule

Help

Coordinate System

TVD, Length

Choke Method

ELF

Temperature of Surroundings

39

deg F

Overall Heat Transfer Coefficient

3

BTU/h/ft2/F

Input Data

Point	Label	Type	Pipe Length	True Vertical Depth	Pipe Inside Diameter	Pipe Inside Roughness	Rate Multiplier
			(feet)	(feet)	(inches)	(inches)	
1		Manifold		0			
2		Pipe	1650	1650	4	0.0006	1
3		Pipe	25000	8200	4	0.0006	1



Figure 78: Deviation Survey diagram of the Surface Equipment (Offshore region)

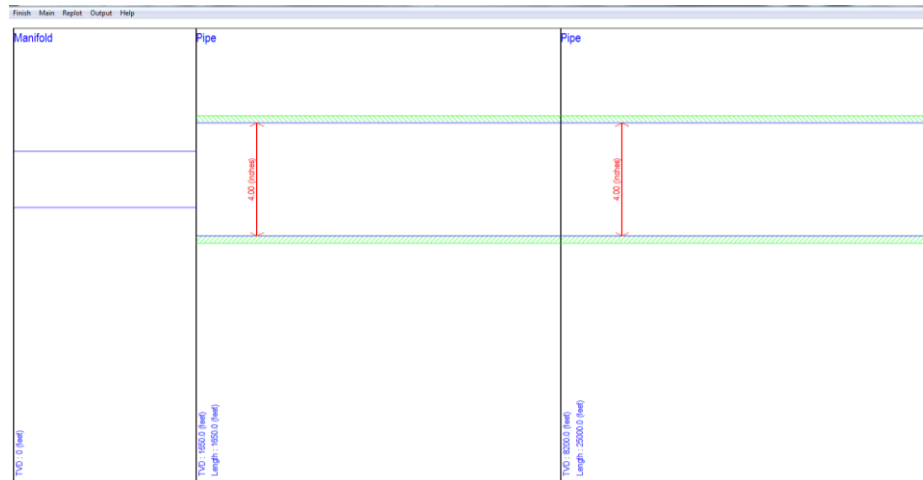


Figure 79: Sketch of the Surface Equipment (Offshore region)

Another piece of information required for the calculations is the temperature gradient in the sedimentary rocks from the reservoir to the surface, as well as the overall heat transfer coefficient of the tubing inside the wellbore (5 BTU/h/ft²/°F - Table 19). For this case the subsurface environment and pipeline is similar to that of the arctic region with the exception of the temperature around the wellhead. The temperature gradient follows the profile shown in Figure 80, made with the information provided in Table 19, from the temperature measured at each depth ($T_{\text{res}} = 300$ °F, $T_{\text{sediment}_1} = 253$ °F, $T_{\text{sediment}_2} = 146.67$ °F and $T_{\text{wellhead}} = 39$ °F). Finally the average heat capacities of the production fluids are entered. As shown in Figure 81 the oil has a value of 0.53 BTU/lb/°F, the gas 0.51 BTU/lb/°F and the oil 1 BTU/lb/°F.

Table 19: Geothermal Temperature Gradient (Offshore region)

Point	Formation TVD (feet)	Formation Measured Depth (feet)	Formation Temperature (deg F)
1	8200	8200	39
2	12200	12200	146.67
3	17700	17700	253
4	20200	20200	300

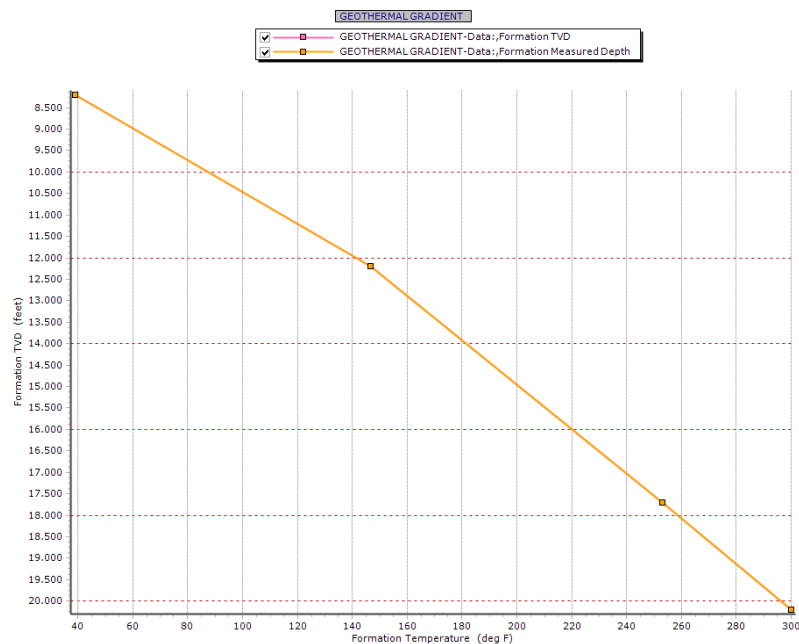


Figure 80: Geothermal Gradient diagram (Offshore region)

Parameter	Value	Unit
Cp Oil	0.53	BTU/lb/F
Cp Gas	0.51	BTU/lb/F
Cp Water	1	BTU/lb/F

Figure 81: Average Heat capacities of the production fluids

The last piece of information needed to describe the arctic system configuration is the description of the Inflow Performance of the reservoir (IPR) of the reservoir. As explained in Chapter 4.2, IPR is required for calculating the production rate as the latter can be obtained from the intersection of Inflow Performance Relationship (IPR) and the Vertical Lift Performance (VLP) curves by use of the nodal analysis method.

The necessary information required for the IPR calculation is related to the reservoir properties. As explained before these parameters are needed to calculate the Productivity index. To obtain a reliable IPR curve the reservoir must be described accurately. This is done from the System menu of the **PROSPER** under Inflow Performance (Figure 82). The reservoir model, depending on the information available, can be done from different correlations and/or theoretical models. If the information is sparse it is done with less accurate correlations that require less information. Accordingly more detailed correlations provide better results but also require a big

amount of information. Options are also available to test the correlation versus real reservoir data from well tests and tune the model against them. As this configuration was done for experimental purposes and it does not apply to a real case scenario, the reservoir was described with one of the simpler correlation methods implemented in the program (Petroleum Experts 4⁷) without matching. The reservoir is single branched with a cased hole completion and no sand control implemented. Also the skin factor is entered by hand and set at 5 (Figure 82).

The numerical data inserted regarding the reservoir for this correlation are:

- Initial Reservoir Pressure (6,000 psig)
- Reservoir Temperature (300 °F)
- Initial Water Gas ratio (10 STB/MMscf)
- GOR (6,943.98 scf/STB)
- Permeability (25 mD)
- Reservoir Thickness (100 ft)
- Drainage Area (340 acres)
- Dietz Shape factor (31.6)
- Wellbore radius (0.354 ft)
- Perforation Interval (100 ft)
- Time since production started (50 days)
- Reservoir Porosity (0.25)
- Connate Water Saturation (0.2)

The IPR curve calculated from the program is displayed in the results window (Figure 82) and plotted graphically in Figure 83 (Reservoir Pressure – Flow rate).

Assessment of Hydrate Formation Parameters in Production Wells

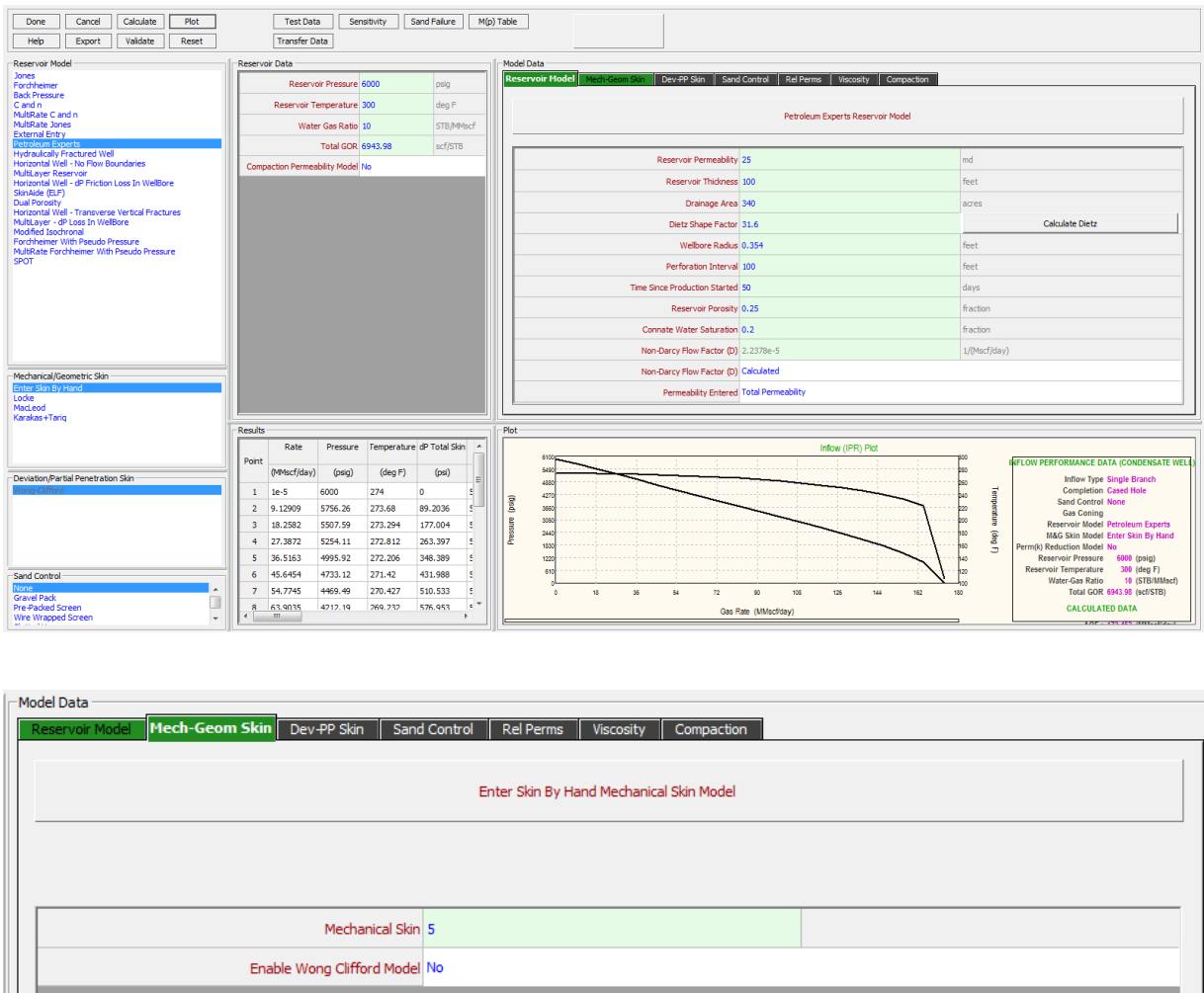


Figure 82: Reservoir characteristics configuration and IPR curve calculation from correlations (Arctic region)

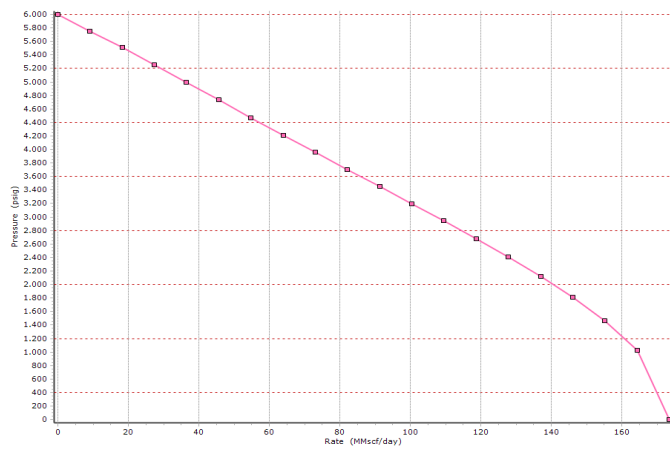


Figure 83: IPR curve diagram (Arctic region)

6. SENSITIVITY ANALYSIS

The objective of this chapter is to determine which production system parameters influence the hydrate forming conditions and why, by using the PROSPER software. The principles and the calculations used were thoroughly described in the Chapter 4 and in Chapter 5 the properties of the production fluids and the system configuration is introduced. Using this information one can proceed to the sensitivity analysis scenarios that will be described in this Chapter. Two main sensitivity analysis scenarios are presented. The first is the sensitivity analysis of the system parameters that affect the VLP curve and the second is the sensitivity analysis of two thermodynamic inhibitors with various concentrations on the effect of hydrate forming conditions. All the cases are tested with both PVT fluids and both well and surface system configurations, for each region, already presented.

6.1 SENSITIVITY ANALYSIS ON SYSTEM PARAMETERS AND EFFECT ON HYDRATE FORMATION WITHOUT USE OF INHIBITORS

In [Figure 84](#) the window dialog for the calculation of the intersection between the Inflow Performance Relationship (IPR) and the Vertical Lift Performance (VLP) curve is shown. As described in Chapter 4 the IPR curve for the specific reservoir configuration is given by the Petroleum Experts 4⁷ correlation. The VLP curve demonstrates the ability of the designed production system to produce fluids (Pressure – Gas rate). It depends on the equipment, pipeline characteristics and environmental conditions and in this thesis is subject to Sensitivity Analysis. The objective is to determine which characteristics, that affect the VLP curve, have a negative or positive effect on the hydrate forming conditions. In [Figure 84](#) the configuration of the sensitivity analysis cases is introduced and the gas rates of each scenario are calculated. The calculation rates are defined by the user from 0.1 to 50 MMscf/day ([Figure 84 - Green](#)) to facilitate the calculation time. If the solution rate lies between the values that are calculated the program determines the value by interpolating between the two nearest points. The correlations used for the determination of pressure - temperature along depth for the surface equipment and the VLP calculations are the Beggs and Brill⁵ and Petroleum Experts 2⁶ respectively ([Figure 84 - Red](#)). The program uses as solution node the bottom hole one, so it can match the pressure calculated from the correlations with the true bottom hole pressure that can be measured. The sensitivity analysis cases are entered by the user from cases tab. The

variables checked in this sensitivity analysis are the tubing diameter, the insulation of the pipe, the separator pressure and the water cut.

A total of 120 different scenarios for each different fluid and region combination are provided. These scenarios will be explained in the following chapters analytically.

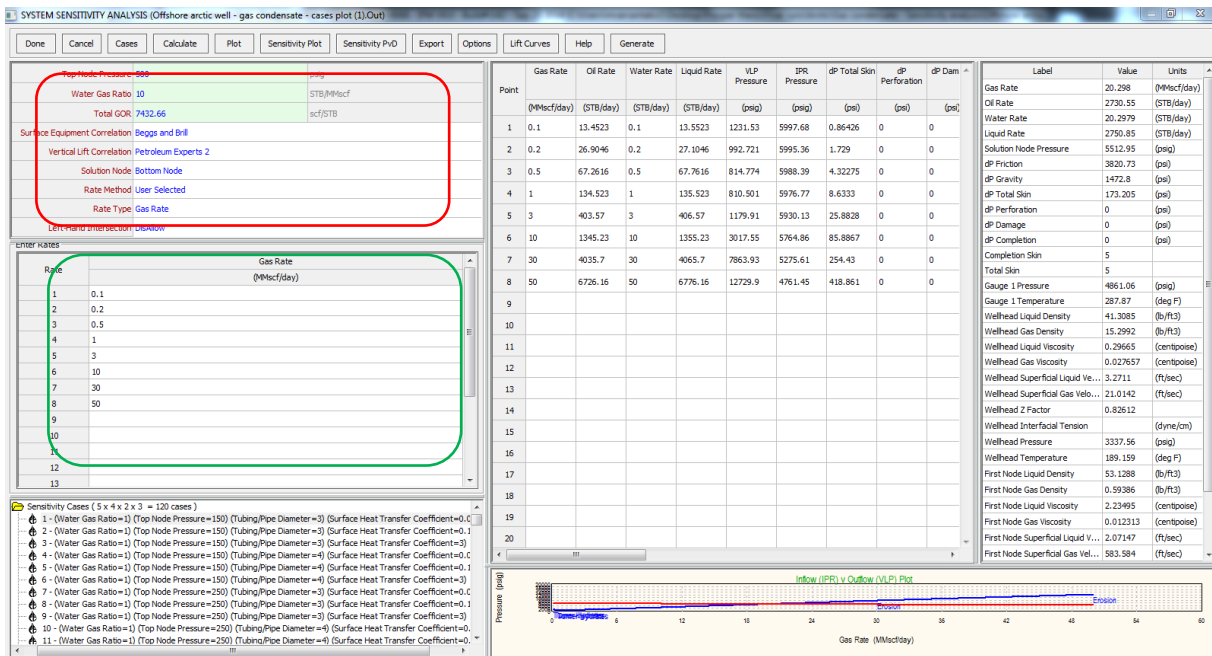


Figure 84: System Sensitivity Analysis dialog (IPR-VLP curves intersection calculation)

6.1.1 EFFECT OF TUBING DIAMETER

The tubing effect as all the variables checked in Chapter 5 does not alter the hydrate dissociation curve but change the Vertical Lift Performance (VLP) curve, which in turn affects the flow rate and finally the velocity of the fluid. This is of utmost importance in steady conditions as the velocity determines the time the fluid has to interact and transfer heat to the environment. The tubing size affect in nonlinear way the flow, the exact relation can be found with momentum balance equations. Larger diameters of tubing result in higher flow rates but not proportionally higher velocities. Sometimes these velocities can be even less compared to those of smaller tubing diameters and less flowrate. In the case of oil generally smaller diameters facilitate the maintenance of a high fluid speed, especially if the reservoir does not have much energy. Things are the other way around with gases, which have generally high velocities and can maintain them or move even faster in larger tubings.

The window dialog in Figure 85 is used to define the cases needed to be calculated in the sensitivity analysis scenarios for a specific variable. In this chapter the tubing diameter will be discussed. Two cases were selected one with tubing of internal diameter (ID) 3 and one with 4 inches.

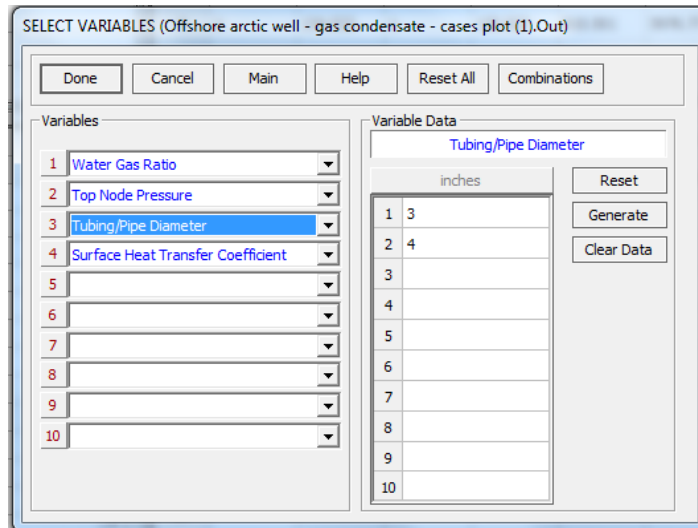


Figure 85: Sensitivity Analysis window – Selection of Tubing diameter possible cases (3 and 4 inches)

6.1.1.1 Gas Condensate – Arctic Region

Because of the large number of the total scenarios calculated by the sensitivity analysis it is difficult to visualize all of them. Thus, the diagrams regarding the effect of the tubing's Internal Diameter ([Figure 86](#) and [Figure 87](#)) depict the combination of the different tubing diameters only with the worst and best case scenario regarding the rest variables. In [Figure 87](#) the best case scenario is displayed. The rest of the system parameters are set at the values that facilitate flow outside the hydrate stability region. In particular the water rate is set at 1 STB/MMscf. This value can either appear in early stages of production or with the use of a subsea two-phase separator during later stages of production. The insulation used is a Vacuum Installed Tubing (VIT) which provides the best heat transfer coefficient (0.03 BTU/h/ft²/°F) and the top node (separator) pressure is set at the minimum possible operational pressure, 150 psig. The pink VLP line corresponds to 3 inch ID tubing and the purple one to 4 inch ID. The red line displays the so called hydrate flag line. This line determines if the solution flow rate is operating inside or outside the hydrate stability region. If the solution node is on the right side then no hydrate problem will occur, the opposite is true for the left side.

In [Figure 86](#) the worst case scenario is shown. The system parameters are set at water rate 250 STB/day, which increases the hydrostatic column of the fluid and can occur at later stages of production with no subsea separation. The surface pipeline is uninsulated and its heat transfer coefficient is 3 BTU/h/ft²/°F, while the separator pressure (top node) has a high value of 500 psig. The pink VLP line corresponds to 3 inch ID tubing and the purple one to 4 inch ID. The red line displays the so called hydrate flag line. This line determines if the solution flow rate is operating inside or outside the hydrate stability region. If the solution node is on the right side no hydrate problem can occur, the opposite is true for the left side.

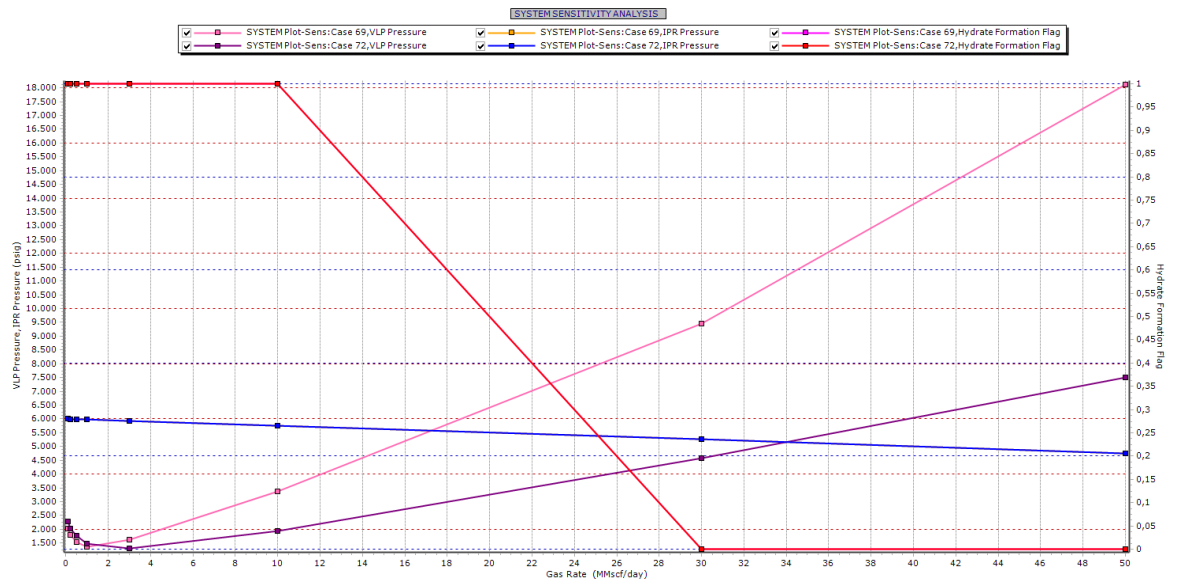


Figure 86: IPR - VLP curves intersection for 4 inches tubing and 3 inches tubing with 250 STB/scf water rate, 3 BTU/h/ft²/°F overall heat transfer coefficient and 500 psig separator pressure. Arctic region, Gas condensate - Worst case scenario

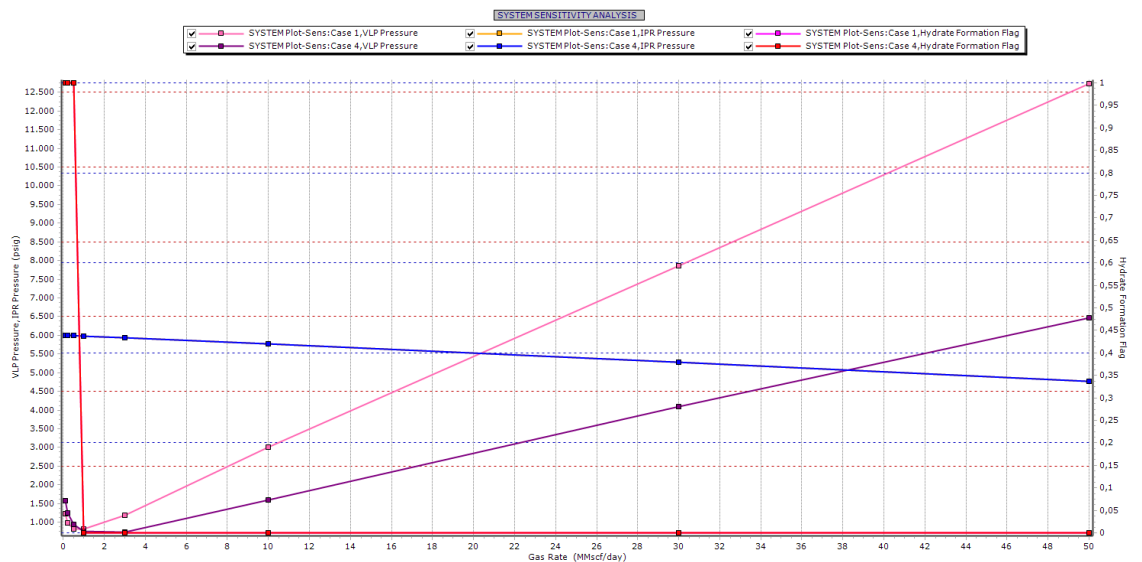


Figure 87: IPR - VLP curves intersection for 4 inches tubing and 3 inches tubing with 1 STB/scf water rate, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 150 psig separator pressure (top node). Arctic region, Gas condensate - Best case scenario

In both cases the bigger internal diameter results in higher production gas rate by twofold. With higher gas rate, for the specific chosen tubing diameters, the fluid velocity is greater (2,500 ft/sec compared to 600 ft/sec, almost 4.5 times faster) and the fluid remains less time inside the production tubing (11 sec compared to 47 sec). With less time available for the heat exchange between the fluid and environment,

the fluid travels having higher temperatures compared to the 3 inches tubing. Ultimately the risk of hydrate formation is reduced with the higher temperatures of the fluid.

In the best scenario (Figure 87), during steady state conditions, the fluid maintains enough temperature to be well outside the hydrate forming region (left of the hydrate flag –red line) disregarding the tubing diameter used (3 inch VLP – pink, 4 inch VLP – purple).

In Figure 86 however when the parameters do not facilitate the flow outside the hydrate forming region, the low velocity maintained by the fluid inside the 3 inch tubing (pink) forces the system to operate inside the hydrate stability region (left of the hydrate flag –red line). By using the larger diameter the problem can be easily avoided or alternate apply inhibitor dosage in the stream.

6.1.1.2 Oil – Arctic Region

In the same manner, as in previous chapters, the tubing diameter sensitivity analysis for an oil system in an arctic region is discussed. In Figure 88 the best case scenario is displayed. The rest of the system parameters, disregarding the sensitivity variable of the tubing, are set as: water cut (0%), heat transfer coefficient (0.03 BTU/h/ft²/°F) and top node (separator) pressure (150 psig).

In Figure 89 the worst case scenario is shown. The system parameters are set at water rate 80%, the surface pipeline's heat transfer coefficient is 3 BTU/h/ft²/°F, while the separator pressure (top node) has a value of 500 psig.

In the best case scenario (Figure 88) the bigger internal diameter results in higher oil production rates for operating conditions (from 1450 to 1600 STB/day, around 10% increase). This happens due to the increase via the tubing diameter of the well deliverability as already explained. Also with 0% water cut the reservoir provides fluids of high energy and low density that can easy travel to the surface. With both tubings (3 inch VLP – pink, 4 inch VLP – purple) the fluid maintains enough temperature to be well outside the hydrate forming region (left of the hydrate flag – red line).

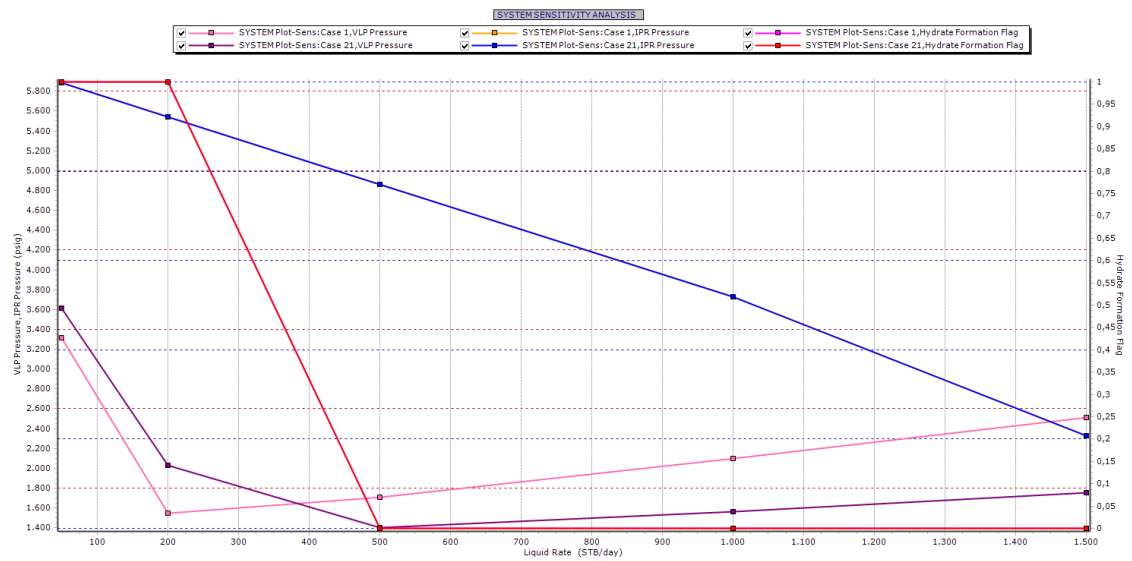


Figure 88: : IPR – VLP curves intersection for 4 inches tubing (purple) and 3 inches tubing (pink) with 0% water cut, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 150 psig separator pressure (top node). Arctic region, Oil – Best case scenario

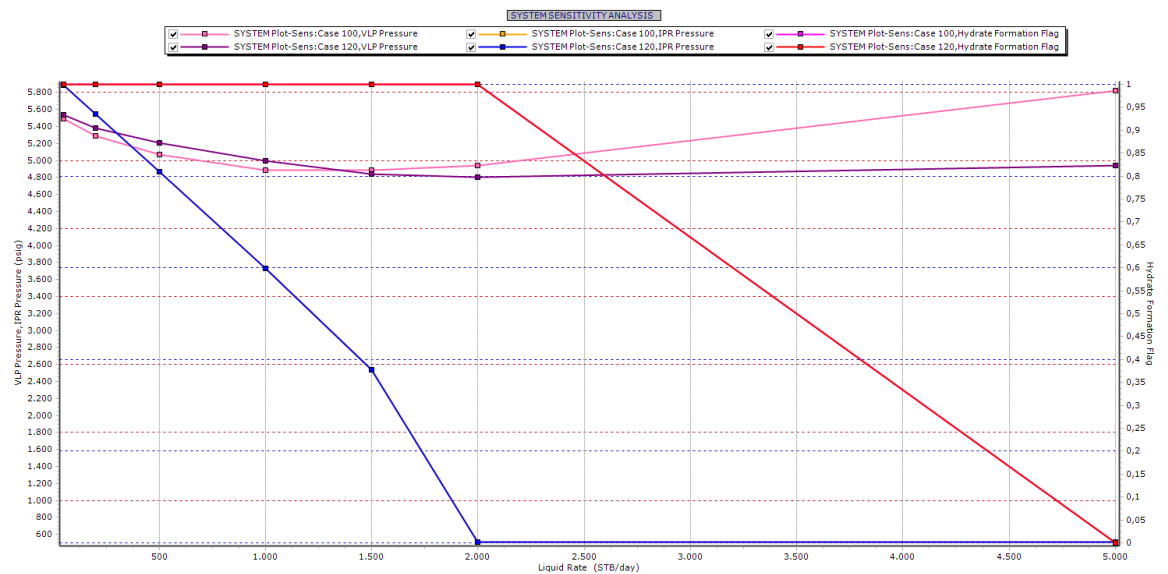


Figure 89: IPR – VLP curves intersection for 4 inches tubing (purple) and 3 inches tubing (pink) with 80% water cut, 3 BTU/h/ft²/°F overall heat transfer coefficient and 500 psig separator pressure (top node). Arctic region, Oil – Worst case scenario

In the worst case scenario the well deliverability is extremely reduced and as shown from Figure 89 needs very high bottom hole pressures to maintain the flow. The increased water cut in this scenario (80%) results in higher density fluid from the reservoir. That decreases the deliverability of the well by increasing the hydrostatic head that needs to be beaten. In this specific case the smaller tubing (3 inch VLP –

pink) has a beneficial effect in the small oil rates and the fluids exhibit higher velocities. That is the reason it is closer to the hydrate dissociation curve, as higher velocities imply reduced heat transfer. Even so, in the specific case both solution nodes lay deep inside the hydrate stable region, as the rates are extremely low and so the velocities.

6.1.1.3 Gas Condensate - Deep Offshore Region

The same method in presenting the cases is followed as in chapter 5.1.1.2 and will continue in the rest of the chapters of the sensitivity analysis. In the present chapter a gas condensate system in an offshore region will be discussed. In Figure 85 the best case scenario is displayed. The rest of the system parameters disregarding the sensitivity variable are set at a water rate of 1 STB/MMscf, the heat transfer coefficient (0.03 BTU/h/ft²/F) and the top node (separator) pressure set at 150 psig.

In Figure 91 the worst case scenario is shown. The system parameters are set at water rate 250 STB/day, the surface pipeline's heat transfer coefficient is 3 BTU/h/ft²/F, while the separator pressure (top node) has a value of 500 psig.

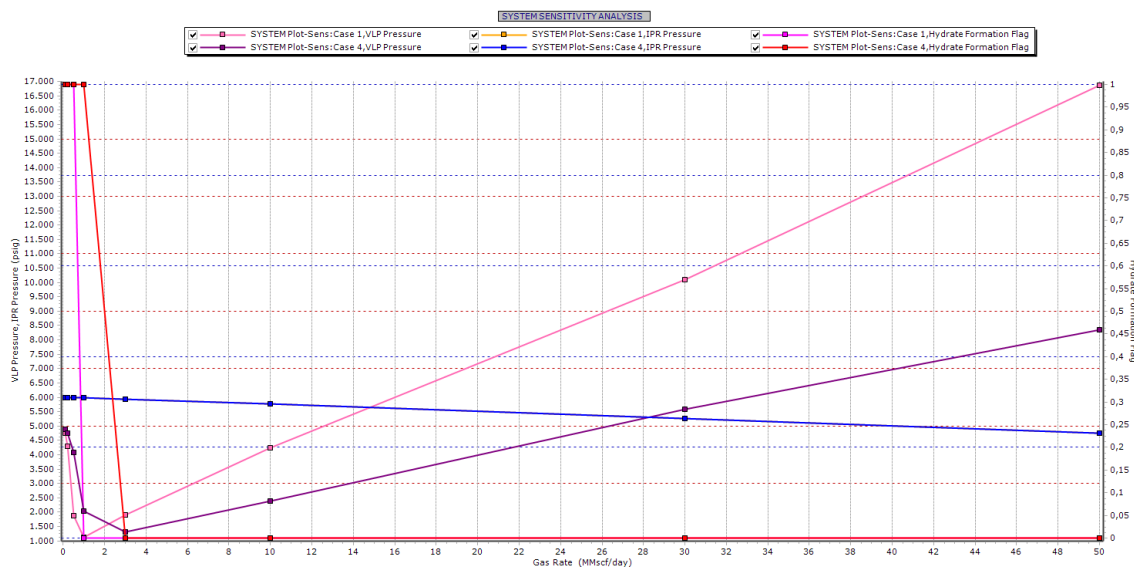


Figure 90: IPR – VLP curves intersection for 4 inches tubing (purple) and 3 inches tubing (pink) with 1 STB/scf water rate, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 150 psig separator pressure (top node). Offshore, Gas condensate – Best case scenario

As in chapter 5.1.1.1 in both cases the bigger internal diameter results in increased production gas rate by twofold. With higher gas rate, for the bigger tubing diameters, the fluid velocity is greater and the fluid remains less time inside the production

tubing. This ultimately leads to decreased heat exchange between the fluid and environment.

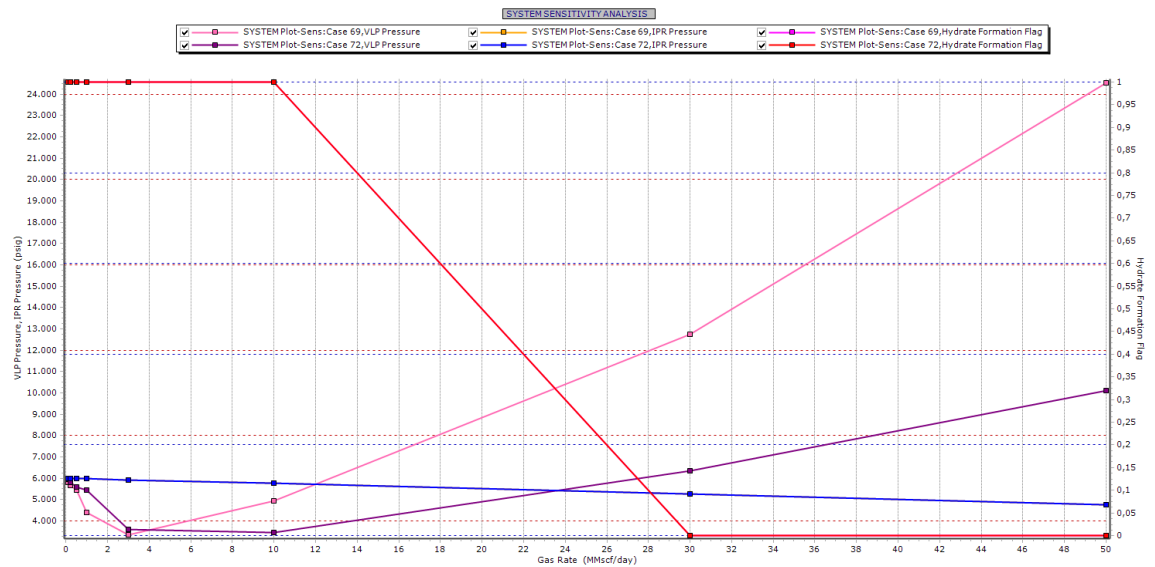


Figure 91: IPR – VLP curves intersection for 4 inches tubing (purple) and 3 inches tubing (pink) with 250 STB/scf water rate, 3 BTU/h/ft²/°F overall heat transfer coefficient and 500 psig separator pressure (top node). Offshore, Gas condensate – Worst case scenario

In the best scenario (Figure 90), during steady state conditions, the fluid maintains enough temperature to be well outside the hydrate forming region (left of the hydrate flag –red line) disregarding the tubing diameter used (3 inch VLP – pink, 4 inch VLP – purple).

In Figure 91 however the situation is worse than in chapter 5.1.1.1 due to the more extended tie-back of the offshore production system. Despite the smaller temperature differential that exists in the offshore system (higher ambient temperature) the system operates in conditions closer to the hydrate stability region. The time spent to bring the fluids to the surface affects more the heat transfer and the gas rate that must be maintained to avoid the hydrate problem is over 30 MMscf/day. Thus disregarding the tubing diameter the pipeline operates inside the hydrate forming region, and remedial solution must be implemented, such as thermodynamic inhibitors (THIs).

6.1.1.4 Oil – Deep Offshore Region

The most challenging system in terms of well deliverability is described in this chapter. The sensitivity analysis is done for the production tubing as already described previously.

In Figure 92 the best case scenario is displayed. The rest of the system parameters disregarding the sensitivity variable checked are set as: water cut 0%, overall heat transfer coefficient (0.03 BTU/h/ft²/°F) and top node (separator) pressure 150 psig.

In Figure 93 the worst case scenario is shown. The system parameters are set at water cut 80%, the surface pipeline's heat transfer coefficient is 3 BTU/h/ft²/°F, while the separator pressure (top node) has a value of 500 psig.

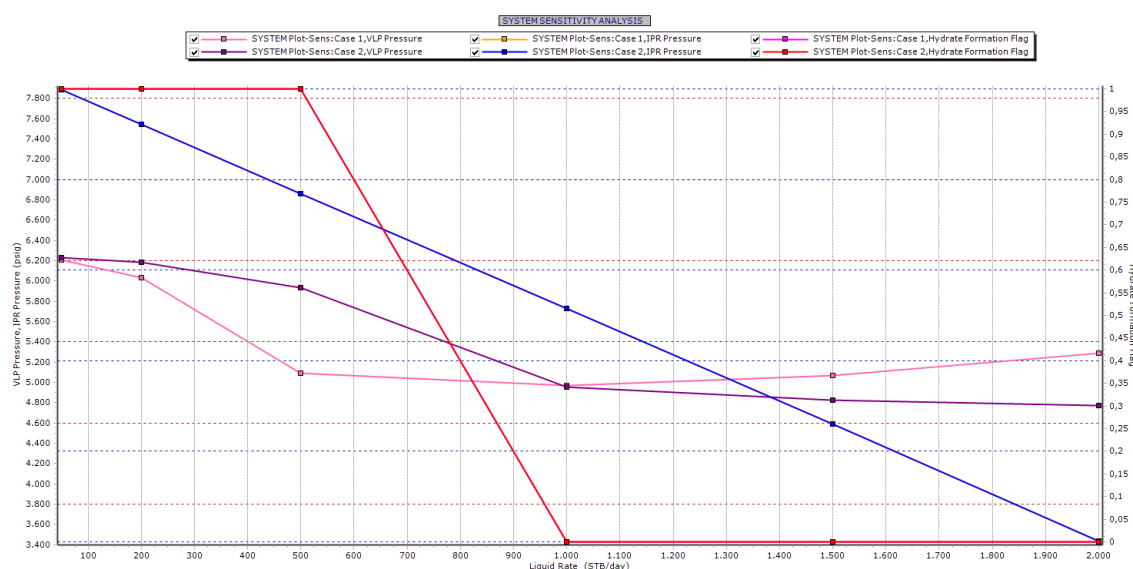


Figure 92: IPR – VLP curves intersection for 4 inches tubing (purple) and 3 inches tubing (pink) with 0% water cut, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 150 psig separator pressure (top node). Arctic region, Oil – Best case scenario

In the best case scenario (Figure 92) the bigger internal diameter results in higher oil production rates (from 1300 to 1380 STB/day around 6% increase). Also with 0% water cut the reservoir provides fluids of low density that can easier travel to the surface. With both tubing's diameters (3 inch VLP – pink, 4 inch VLP – purple) the fluid maintains enough temperature to be well outside the hydrate forming region (left of the hydrate flag –red line).

In the worst case scenario (Figure 93) no intersection between VLP and IPR can occur. The well does not have the ability to deliver fluids to the surface with this configuration and it is pointless further discussion.

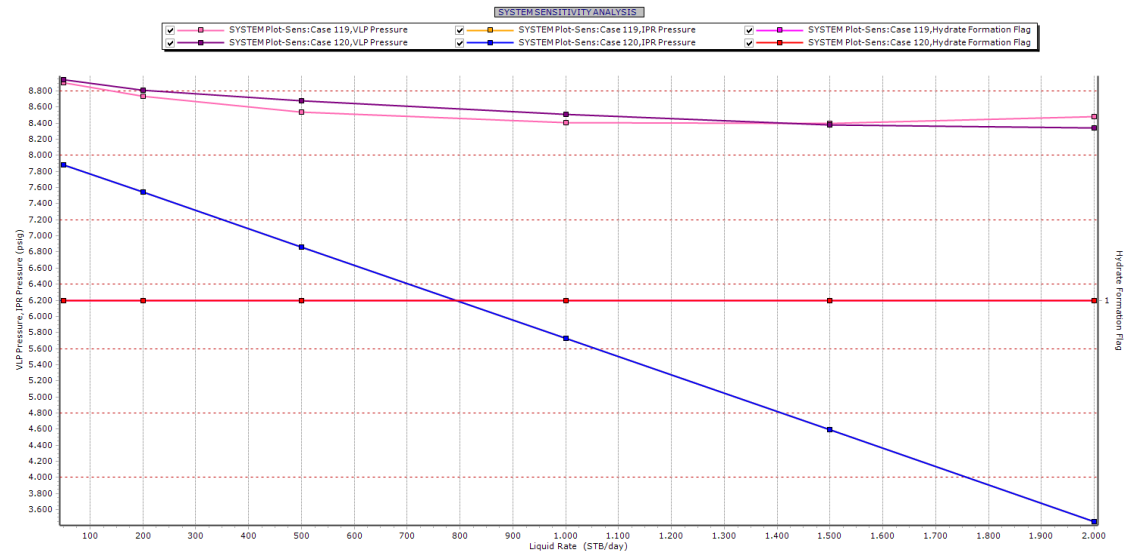


Figure 93: IPR – VLP curves for 4 inches tubing (purple) and 3 inches tubing (pink) with 80% water rate, 3 BTU/h/ft²/°F overall heat transfer coefficient and 500 psig separator pressure (top node). Offshore region, Oil – Worst case scenario (no flow)

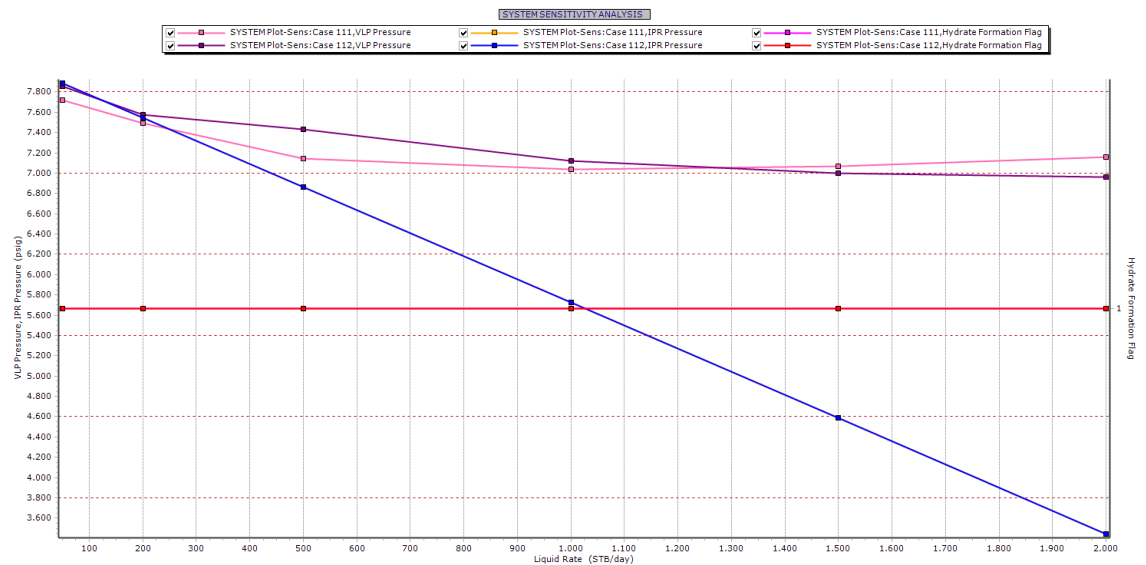


Figure 94: IPR – VLP curves for 4 inches tubing (purple) and 3 inches tubing (pink) with 40% water cut, 3 BTU/h/ft²/°F overall heat transfer coefficient and 500 psig separator pressure (top node). Offshore region, Oil – Worst scenario

If the water cut is reduced to 40% a scenario exists where the IPR intersects with the VLP, although with very low oil production rates (250 STB/day). The well deliverability is extremely reduced and as shown from [Figure 94](#) very high bottom hole pressure is needed to maintain flow. The high water cut results in highly dense fluid with heavy hydrostatic column. It cannot easily travel to the production facilities even though the bottom hole pressure is high. In this specific case the smaller tubing (3 inch VLP – pink) has a beneficial effect in the small oil rates and the fluids exhibit higher velocities. That is the reason it is closer to the hydrate dissociation curve (from the left side). As already discussed before higher velocities imply reduced heat transfer. In the specific case both solution nodes lay deep inside the hydrate stable region, as the rates are extremely low and so are the velocities.

6.1.2 EFFECT OF TOP NODE PRESSURE (SEPARATOR)

The window dialog in Figure 95 is used to define the cases needed to be calculated in the sensitivity analysis scenarios. In this chapter the Top Node (separator) pressure will be discussed. Four cases were chosen with 150 (just above the 2nd separator pressure), 250, 400 and 500 psig pressure respectively (Figure 96). The Top Node pressure is a significant factor that affects the VLP curve more intensively in low energy systems, such as heavy oils. Lower separator pressure increases the pressure differential of the system and therefore results in increased flow rate, despite the increase of the pressure losses due to higher rate and velocity of the fluid, according to the Bernoulli equation.

$$\frac{p_1}{\rho g} + \frac{u_1^2}{2g} + z_1 = \frac{p_2}{\rho g} + \frac{u_2^2}{2g} + z_2$$

In terms of VLP curve it can be said that it moves downwards, the lower the pressure at the separator is.

SELECT VARIABLES (Offshore arctic well - gas condensate - cases plot (1).Out)

Done Cancel Main Help Reset All Combinations

Variables

1	Water Gas Ratio
2	Top Node Pressure
3	Tubing/Pipe Diameter
4	Surface Heat Transfer Coefficient
5	
6	
7	
8	
9	
10	

Variable Data

Top Node Pressure

psig

1	150
2	250
3	400
4	500
5	
6	
7	
8	
9	
10	

Reset Generate Clear Data

Figure 95: Sensitivity Analysis window – Selection of Top Node (separator) pressure possible cases

6.1.2.1 Gas Condensate – Arctic Region

As already stated previously due to the big number of cases implemented in the sensitivity analysis, the selection of cases is therefore depicting only the worst and best case scenario. In Figure 96 and Figure 97 the best and the worst scenario are displayed respectively. The VLP curve of 500 psig separator pressure is displayed with brown color, 400 psig curve with green, 250 psig with purple and 150 psig with pink. The pink hydrate flag line corresponds to the lowest separator pressure tested and the orange one to the highest one.

For the best case scenario (Figure 96) the Top Node pressure has more significant effect on the VLP curve as the gas rates get lower. This effect for operating conditions (in the specific case 38 MMscf/day) is very marginal. One reason is that the top node pressure difference leads to an increased pressure differential which is not yet significant compared to the high pressure already prevailing in the reservoir. Also the increased friction losses for the increased flow rate counterbalance the pressure increase.

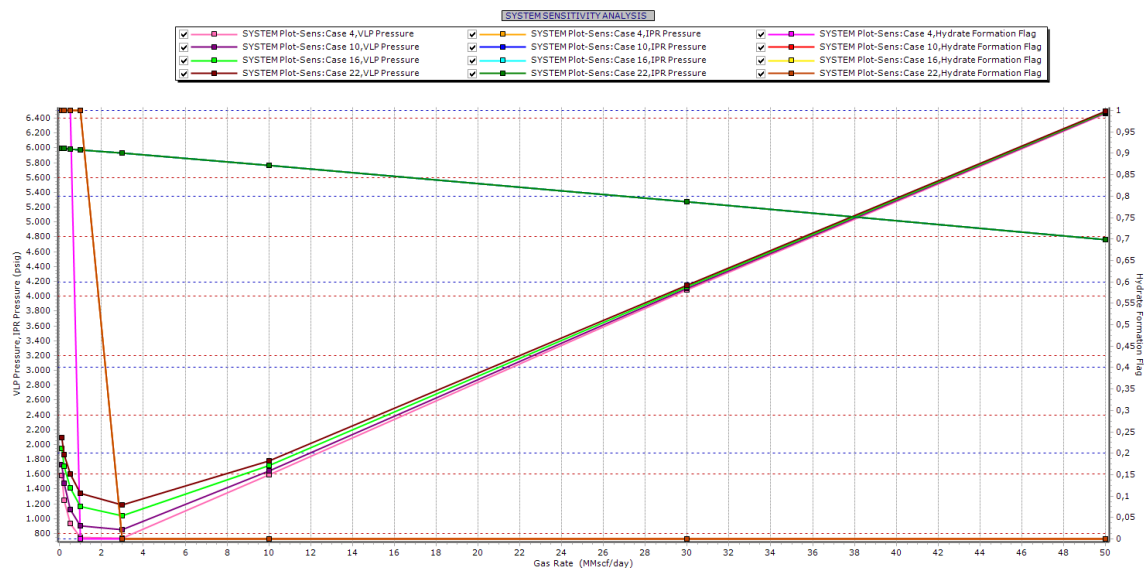


Figure 96: IPR - VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water rate is 1 STB/scf, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 4 inches tubing diameter. Arctic region, Gas condensate – Best case scenario

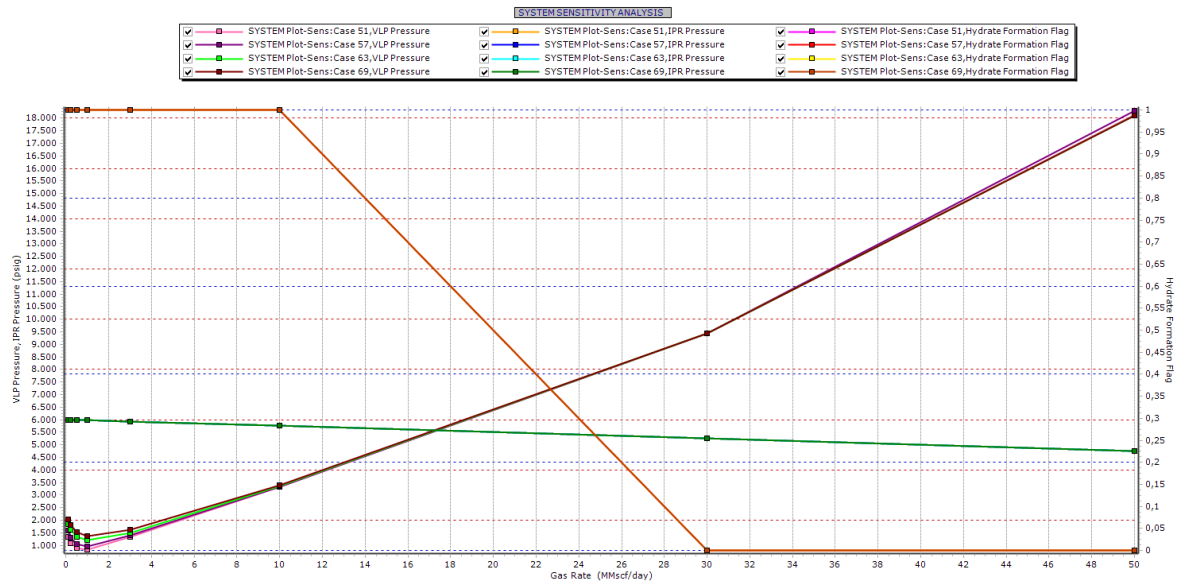


Figure 97: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water rate is 250 STB/scf, 3 BTU/h/ft²/°F overall heat transfer coefficient and 3 inches tubing diameter. Arctic region, Gas condensate – Worst case scenario

Similar behavior as previously is exhibited in the worst case scenario (Figure 97). In most operational gas rates the difference on the VLP curve, as effect of the separator pressure, is insignificant and other parameters affect the curve more. Same as before only in very low gas rates the curves do not coincide. The solution gas rate is around 17 MMscf/day and operates inside the hydrate stability region during steady state conditions. The production fluids exhibit the possibility of forming hydrate particles, with all the risk this implements.

6.1.2.2 Oil – Arctic Region

In Figure 98 and Figure 99 the best and the worst scenaria are displayed respectively for an oil system at an arctic region in the same manner as in previous chapters.

The Top Node pressure has very important effect on the VLP curve for an oil system. The bottom hole pressure, required to deliver fluids to the surface, can decrease almost by 50% by decreasing the separator pressure from 500 to 150 psig (Figure 98). For operating conditions with all other parameters set at optimal values the oil production rate ranges from 1400 (500 psig separator pressure) to over 1600 STB/day (150 psig separator pressure). Even small pressure differential differences are important in oil systems because the energy stored in the reservoir fluids is significantly lower. In the other hand the hydrostatic head has a paramount effect in

the well deliverability (VLP curve), due to the higher density of the oil. The solution oil rates are well afar from the hydrate stability region and during steady state conditions no hydrate problem can occur.

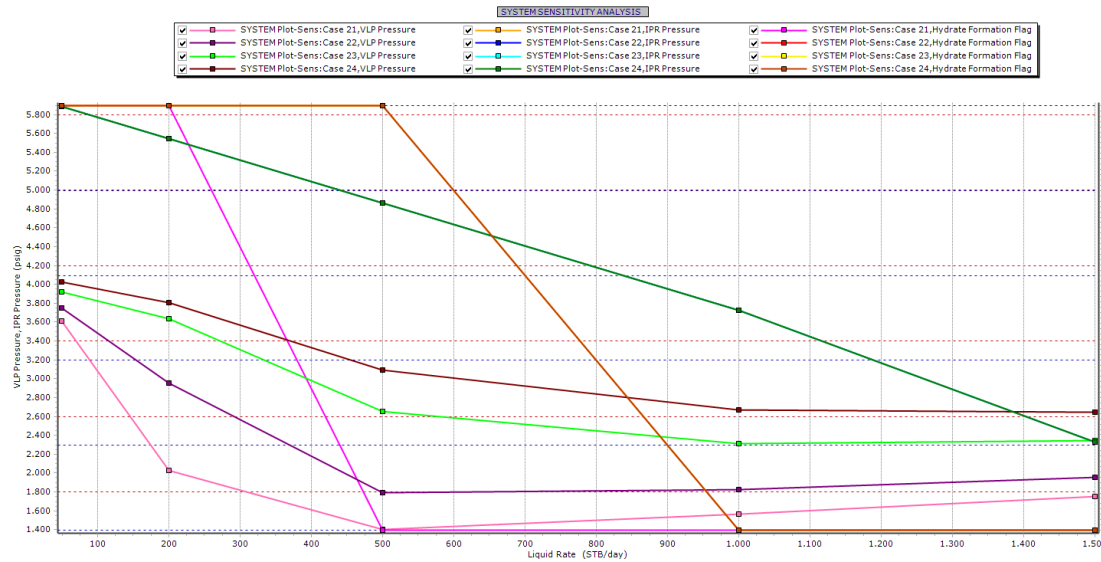


Figure 98: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water cut is 0%, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 4 inches tubing diameter. Arctic region, Oil – Best case scenario

Similar behavior is exhibited in the worst case scenario (Figure 99). The difference on the required bottom hole pressure required to deliver fluids to the surface is smaller than previously but still has a difference of 800 psig (around 20%). The well deliverability is extremely decreased in this case and the pressure needed for the well to flow is increased compared to the best case scenario. The outcome is that the solution nodes are at low rates (500 – 800 MMscf/day) and thus the system operates under hydrate forming conditions, due to the low fluid velocity. During steady state conditions the production fluid will start forming hydrate particles.

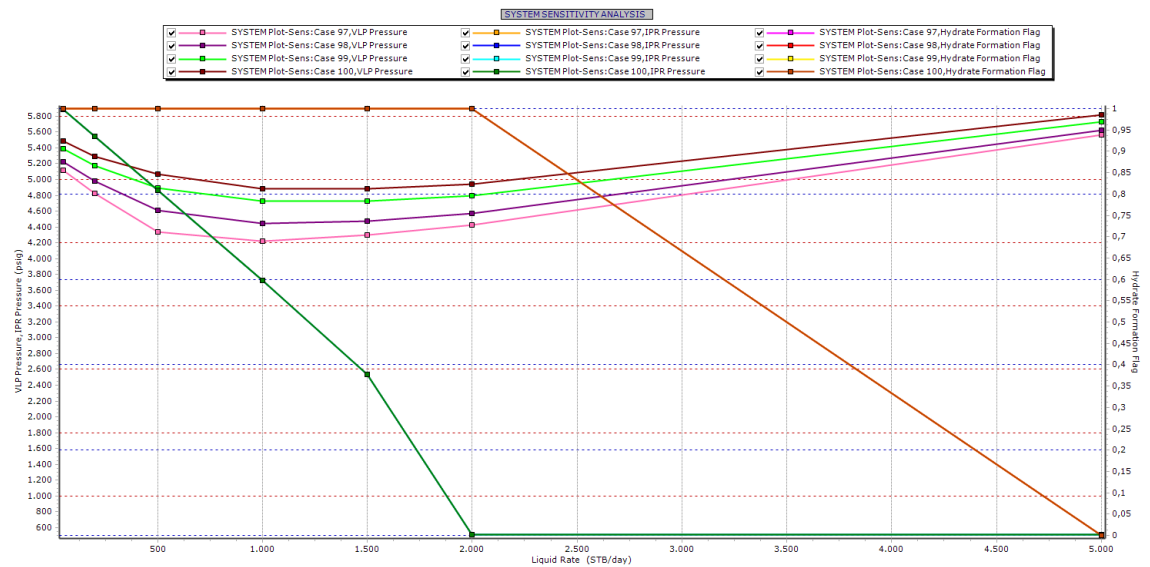


Figure 99: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water cut is 80%, 3 BTU/h/ft²/°F overall heat transfer coefficient and 3 inches tubing diameter. Arctic region, Oil – Worst case scenario

6.1.2.3 Gas Condensate - Deep Offshore Region

In [Figure 100](#) and [Figure 101](#) the best and the worst scenario are displayed respectively, regarding the Top Node pressure sensitivity analysis of a gas condensate, offshore system.

For the best case scenario ([Figure 100](#)) the Top Node pressure has greater effect on the VLP curve the lower the gas rate is. The deviation exhibited is far larger than that of a gas condensate in an arctic region (1400 psig compared to 800 psig). The main reason is that for the same low gas rates the distance the fluid has to travel is 8200 ft longer and with higher inclination. This produces a significant hydrostatic head to be beaten when liquid loading occurs (for small production rates mostly). Also the friction losses are higher simply because the distance is longer. Although this effect, for operating conditions (in the specific case 28 MMscf/day), is marginal due to the high travelling velocities.

Similar behavior as in chapter 5.1.2.1 is exhibited in the worst case scenario ([Figure 101](#)). In most operational gas rates the difference on the VLP curve, as effect of the separator pressure, is insignificant. Only in very low gas rates the curves do not coincide. The solution gas rate around 11 MMscf/day is inside the hydrate stability

region and during steady state conditions the production fluids exhibit the danger of forming hydrate particles.

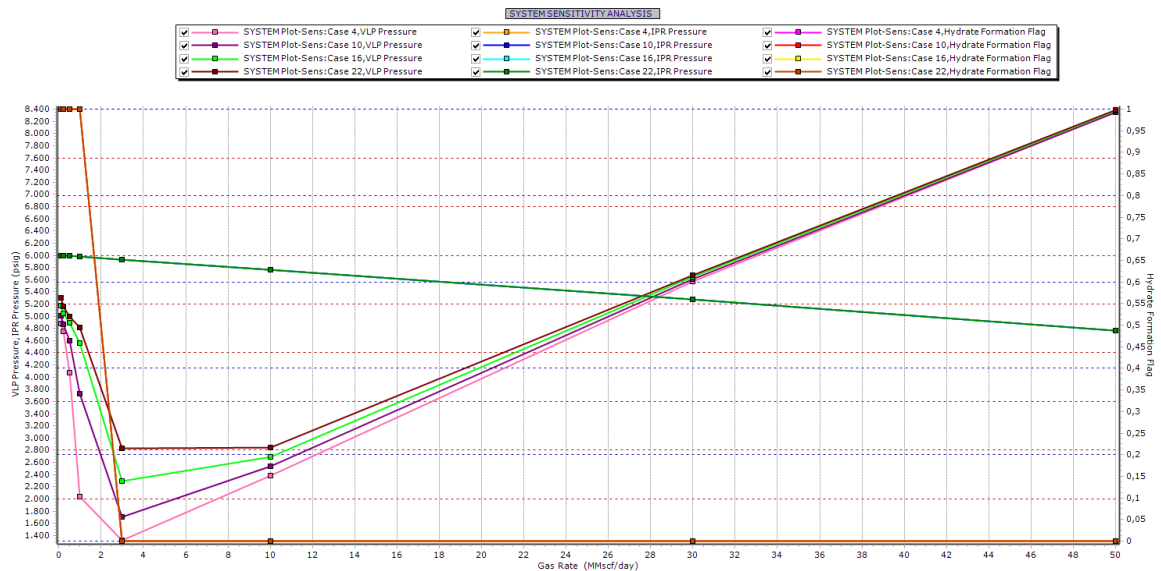


Figure 100: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water rate is 1 STB/scf, 0.03 BTU/h/ft²/°F overall heat transfer coefficient and 4 inches tubing diameter. Offshore region, Gas condensate, – Best case scenario

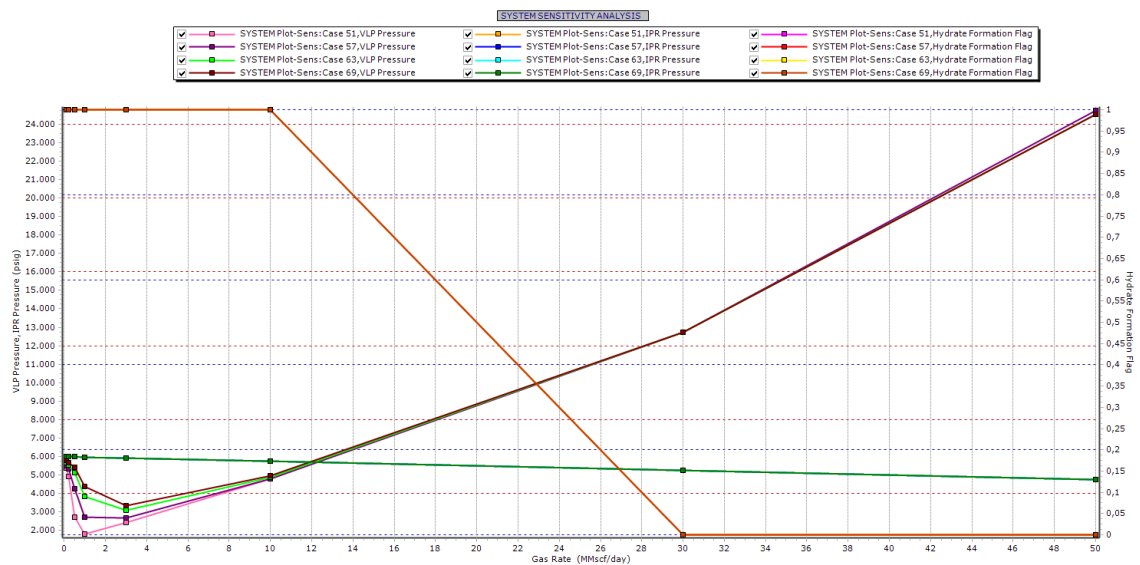


Figure 101: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water rate is 250 STB/scf, 3 BTU/h/ft²/°F overall heat transfer coefficient and 3 inches tubing diameter. Offshore region, Gas condensate – Worst case scenario

6.1.2.4 Oil – Deep Offshore Region

The Top Node pressure affects in an important manner the VLP curve for an oil system. The bottom hole pressure, required to deliver fluids to the surface, can decrease almost by 1000 psig by decreasing the separator pressure from 500 to 150 psig (Figure 102). For operating conditions, with all other parameters set at optimal values, the oil production rate ranges from values of 980 to over 1320 STB/day. Even small pressure differential differences are important in oil systems because the energy stored in the reservoir fluids is low. In the other hand the hydrostatic head has an important effect in the well deliverability (VLP curve), due to the higher density of the oil. The solution oil rates are well far from the hydrate stability region and during steady state conditions no hydrate problem can occur.

In the worst case scenario (Figure 103) the water cut is very high and the hydrostatic head cannot be beaten. The well deliverability is not enough for the well to able to flow and further discussion about hydrate formation is not possible.

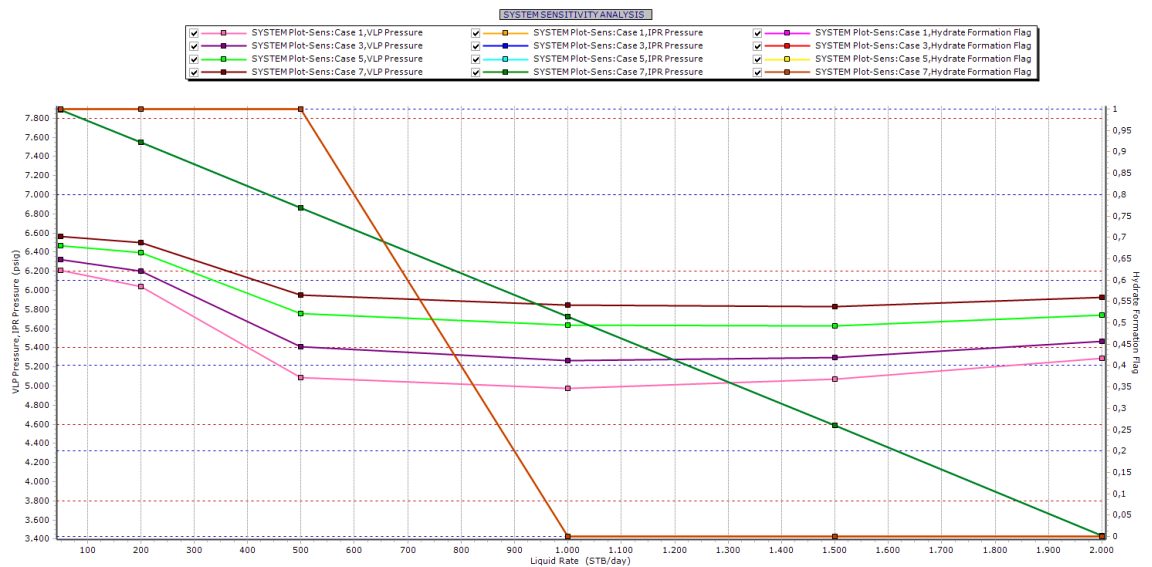


Figure 102: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water cut is 80%, 0.03 BTU/h/ft²/°F the overall heat transfer coefficient and 4 inches tubing diameter. Offshore region, Oil – Best case scenario

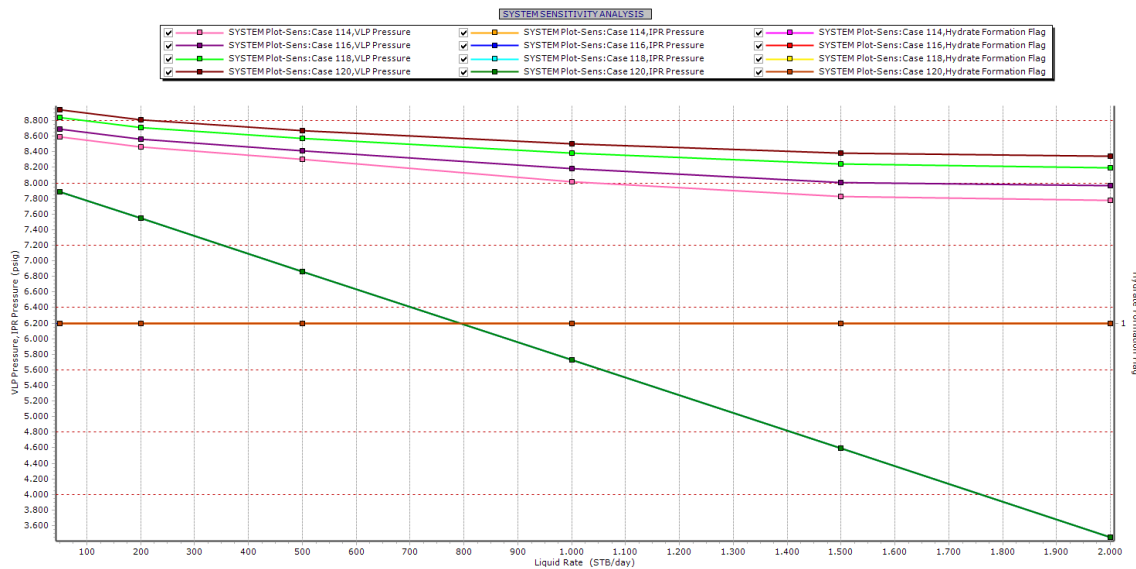


Figure 103: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water cut is 40%, 3 BTU/h/ft²/°F the overall heat transfer coefficient and 3 inches the tubing diameter. Offshore region, Oil – Worst case scenario (no flow)

With water cut set at 40% and the rest of the system parameters at their worst values the well is able to deliver fluids at the surface, although the rate is very low (Figure 104). The difference on the required bottom hole pressure to deliver fluids to the surface, as effect of the separator pressure is smaller than previously but still has a difference of 800 psig (around 20%). The well deliverability, due to the heavy column created, is extremely decreased in this case and the pressure needed to produce flow is far greater compared to the best case scenario. The outcome is that the solution nodes are at low rates (200 – 400 STB/day). Apparently the system operates under hydrate forming conditions, due to the low fluid velocity. During steady state conditions the production fluids is certain that will start forming hydrate particles.

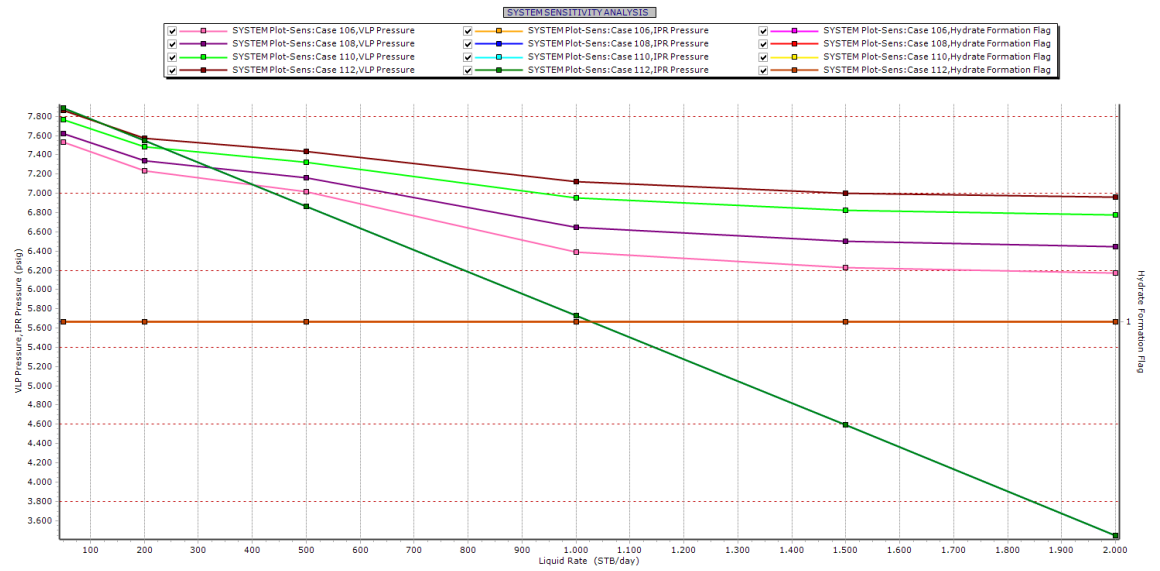


Figure 104: IPR – VLP curves intersection for 150, 250, 400 and 500 psig top node (separator) pressures. The water cut is 40%, 3 BTU/h/ft²/°F the overall heat transfer coefficient and 3 inches the tubing diameter. Offshore region, Oil – Worst case scenario

6.1.3 EFFECT OF SURFACE PIPELINE INSULATION

The effect of the pipeline insulation is expressed with the Overall Heat Transfer Coefficient. This term simplifies the effect of all the materials that consist the system and provide insulation to the pipeline such as valves, joints, steel and insulation material. Insulation does not affect significantly the VLP curve as it only maintains the temperature at high levels. This has an increasing effect in the viscosity of the fluids, thus the linear pressure drop becomes less. An opposite effect can happen when gas exists in the flowpath, where its molecules obtain higher energy and velocity. With higher velocity, the molecules hit the pipeline walls more often and lose more energy. The result is an observed linear pressure drop. The main benefit of the insulation is that it maintains the temperature of the pipeline at levels above of the hydrate forming conditions and other flow assurance issues, such as wax.

Three cases were chosen corresponding to different insulation material or type. The pipeline with 3 BTU/h/ft²/°F is not insulated and the heat transfer coefficient is given only by the protection that the steel itself provides to the fluid (green VLP curve). The next pipeline utilizes a Pipe in Pipe system (PiP) with the use of fiberglass as insulating material. This results in a heat transfer coefficient of the order of 0.12 BTU/h/ft²/°F (purple VLP curve). Finally the last pipeline has a vacuum insulation system (VIT - described in chapter 4) (Cases entered in Figure 105). This configuration provides the system with very low heat transfer coefficient values, in this case 0.03 BTU/h/ft²/°F (purple VLP curve). The hydrate flag lines for the above cases are respectively yellow (3 BTU/h/ft²/°F), red (0.12 BTU/h/ft²/°F) and pink (0.03 BTU/h/ft²/°F).

SELECT VARIABLES (Offshore arctic well - gas condensate - cases plot (1).Out)

Done Cancel Main Help Reset All Combinations

Variables

- Water Gas Ratio
- Top Node Pressure
- Tubing/Pipe Diameter
- Surface Heat Transfer Coefficient
-
-
-
-
-
-

Variable Data

Surface Heat Transfer Coefficient

BTU/h/ft²/F

1	0.03
2	0.12
3	3
4	
5	
6	
7	
8	
9	
10	

Reset Generate Clear Data

Figure 105: Sensitivity Analysis window – Selection of Overall Heat Transfer coefficient (insulation material) possible cases

6.1.3.1 Gas Condensate - Arctic Region

The selection of cases displayed is depicting only the worst and best case scenarios regarding the other parameters. In Figure 103 and Figure 107 the best and the worst scenario are displayed for a gas condensate system in an arctic region.

For the best case scenario (Figure 103) the insulation material has small effect on the VLP curve. This effect for operating conditions (in the specific case 38 to 40 MMscf/day) happens because the increased fluid temperature (purple) increases the gas molecules energy and velocity, which leads to increased frictional losses and subsequently the pressure needed to maintain flow is more. The possible positive effect that the temperature could have on the pressure losses is minimal. Despite the reason that increased temperature decreases the viscosity of the oil phase the condensate is not a large portion of the fluid stream and does not affect much. The fluid stream of lower temperature (green) has increased hydrostatic head due to early condensation; however the total losses are less compared to the frictional losses of the other fluid streams (purple).

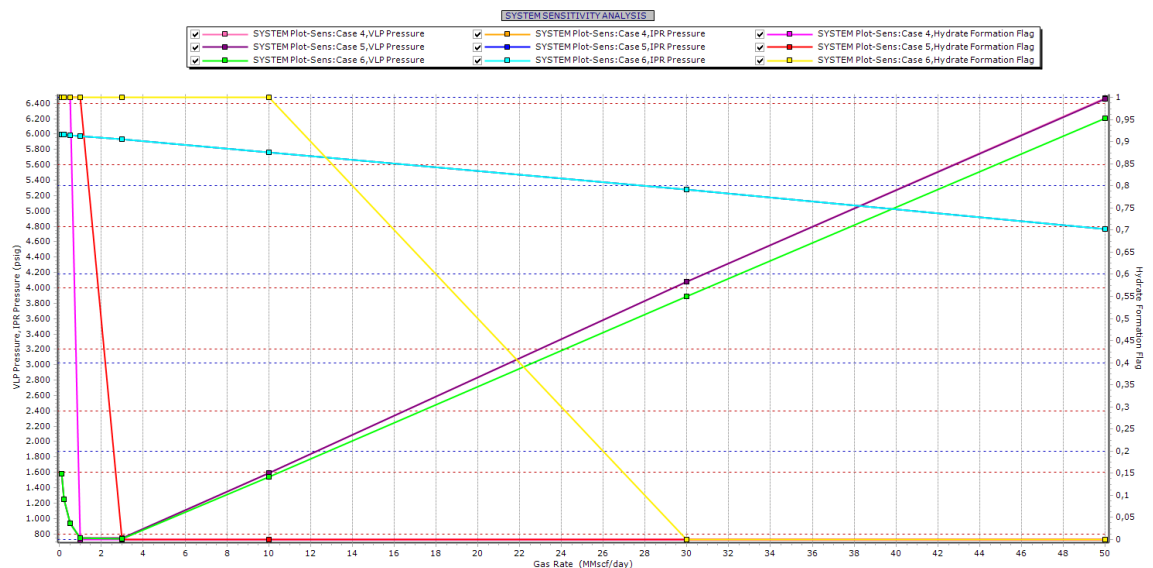


Figure 106: IPR - VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water rate is 1 STB/scf, 150 psig separator pressure and 4 inches tubing diameter. Arctic region, Gas condensate - Best case scenario

In the worst scenario (Figure 107) it is observed that the difference on the VLP curve, as effect of the insulation, is limited for the same reasons described previously. The solution gas rate is around 23 MMscf/day for all the cases but is inside the hydrate stability region only for when utilizing the non-insulated pipeline. On the contrary for insulated pipelines the same problem does not occur and it is safe to

operate in these conditions. The significance of the insulation in the system configuration is evident, because it overshadows the effect of the other parameters by being able to deliver fluids to the surface without hydrate problems.

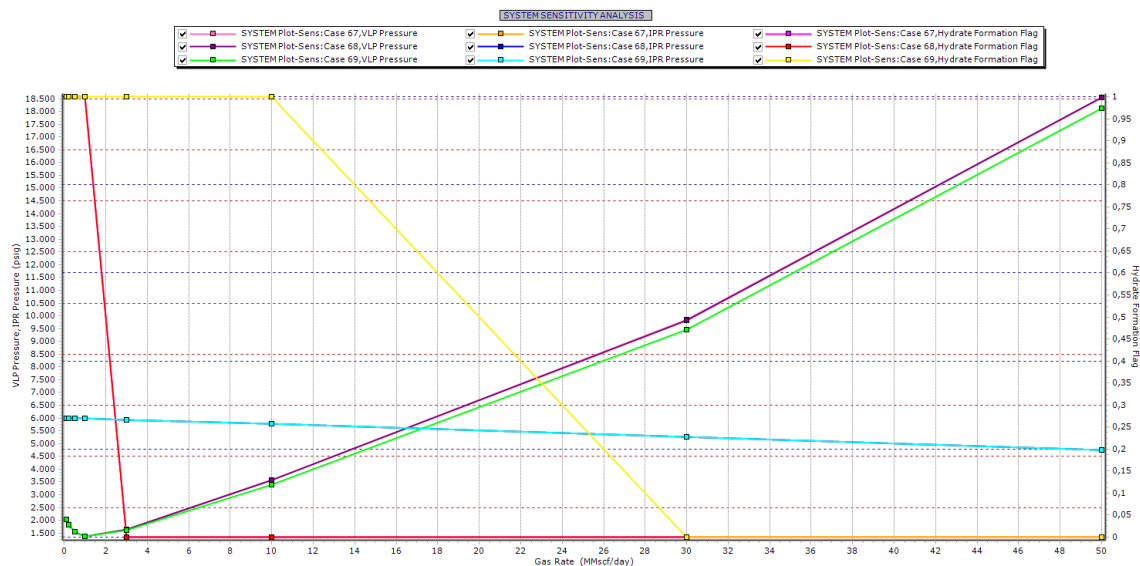


Figure 107: IPR – VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water rate is 250 STB/scf, 500 psig separator pressure and 3 inches tubing diameter. Arctic region, Gas condensate - Worst case scenario

6.1.3.2 Oil – Arctic Region

In Figure 108 and Figure 109 the best and the worst scenario, regarding the insulation sensitivity analysis, are displayed respectively for an oil system in an arctic region.

For the best case scenario (Figure 108) the insulation difference has no effect on the VLP curve. In oil systems, where the fluid is dense and viscous, small changes in temperature do not change drastically the critical attributes that lead to increased or decreased pressure losses due to gravity or to friction. The solution node (intersection of VLP and IPR) lies at 1600 STB/day for all cases. The difference is on the temperature conditions that prevail inside the tubing for each case. This difference is very important for the initial formation of the hydrate particles and can be seen in Figure 108, with the difference of the hydrate flag line for each scenario. When the flowline is not insulated the fluid enters the hydrate stability zone (below the yellow line). In the cases where some degree of insulation is achieved, the solution oil rate is well outside the hydrate forming conditions. This difference between the hydrate flag lines can be observed also for the gas condensate in arctic region (Figure 106). Although the gas travels much faster than the oil and manages to avoid entering into

the hydrate stability conditions. The density and viscosity of the oil are paramount factors for the VLP curve and for flow assurance issues.

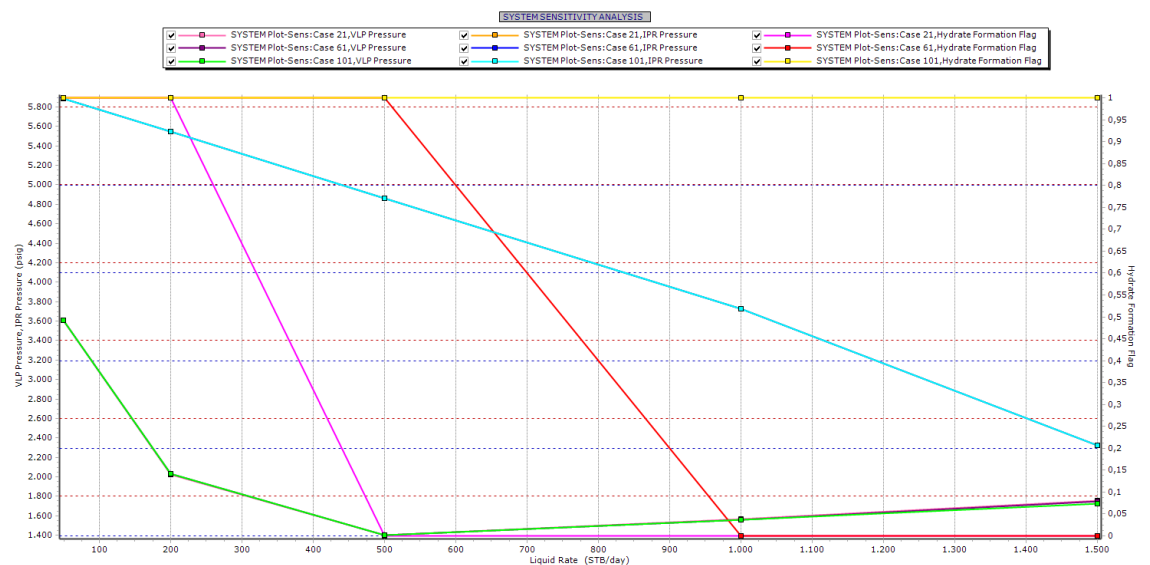


Figure 108: IPR - VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water cut is 0%, 150 psig separator pressure and 4 inches tubing diameter. Arctic region, Oil - Best case scenario

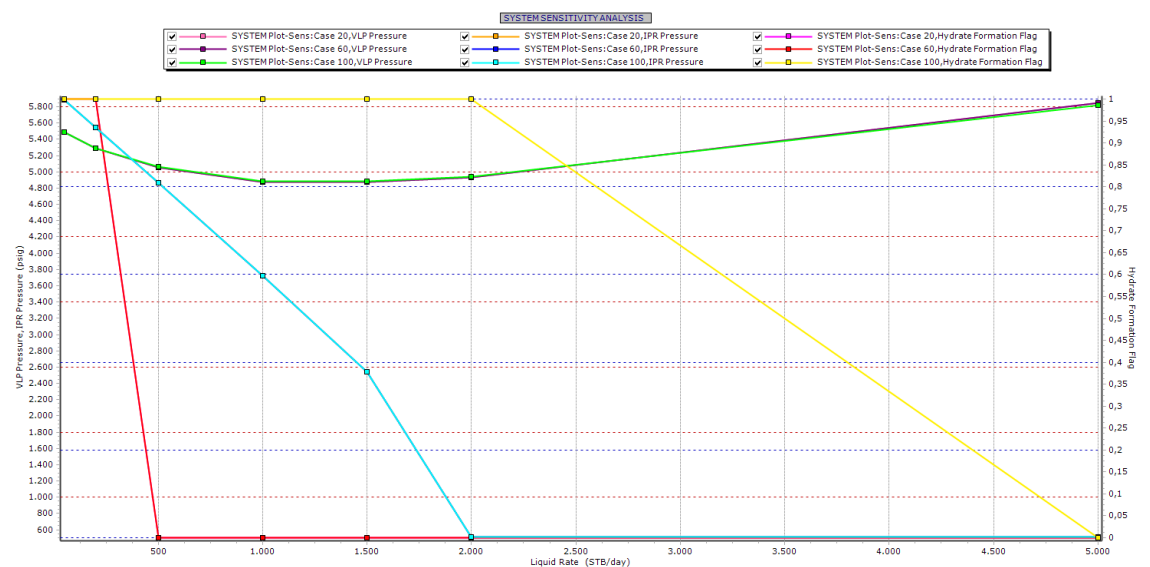


Figure 109: IPR - VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water cut is 40%, 500 psig separator pressure and 3 inches tubing diameter. Arctic region, Oil - Worst case scenario

In Figure 109 the worst case is displayed. The difference on the VLP curve as effect of the insulation is insignificant, as the temperature difference is not so much to display phenomena where the viscosity of the fluid clearly changes and leads to decreased pressure drop. The solution oil rate is around 400 STB/day for all the cases but is inside the hydrate stability region only when utilizing the non-insulated

pipeline. On the contrary for insulated pipelines this problem does not occur and it is safe to operate in these conditions. The significance of the insulation in the system configuration is evident, because it masks the effect of the other parameters by being a reliable parameter that can deliver fluids to the surface without encountering hydrate problems.

6.1.3.3 Gas Condensate - Deep Offshore Region

For the best case scenario (Figure 110) the insulation difference has a small effect on the VLP curve mainly for high gas rates. The increased temperature decreases the viscosity of the fluid and subsequently the pressure needed to maintain flow. On the other hand, the gas expands and obtains higher velocity which increases the friction losses. The result is that the flowline operates with higher temperature, having more pressure losses and subsequently less flow rate. The phenomenon is more intense in longer tie back systems where linear pressure losses increase dramatically, due to the long distance that have to be traveled (purple line insulated – green line uninsulated). This effect for operating conditions results in a difference from 28 to 30 MMscf/day.

In the worst scenario (Figure 111) the same behavior, regarding the VLP, is observed with the insulated (Purple) and uninsulated (Green) pipeline. The solution gas rate is around 12 MMscf/day for all the cases, but is inside the hydrate stability region only when utilizing the non-insulated pipeline (Yellow hydrate flag line). On the contrary for insulated pipelines (Red hydrate flag line) the same problem does not occur and it is safe to operate under these conditions. The significance of the insulation in the system configuration is evident, because it overshadows the effect of the other parameters by being able to deliver fluids without encountering hydrate problems.

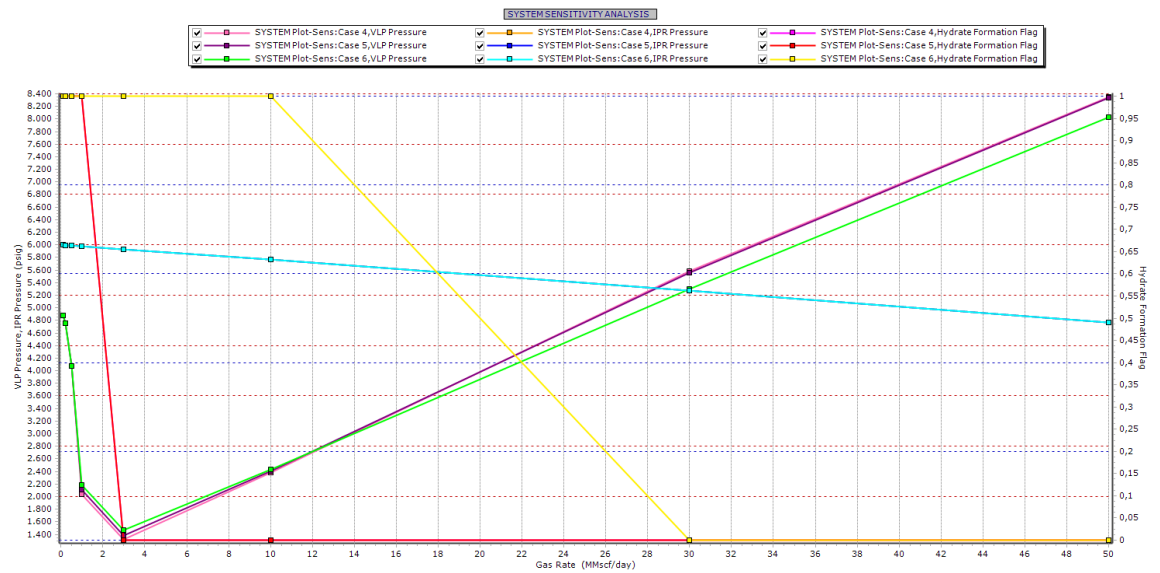


Figure 110: IPR – VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water rate is 1 STB/scf, 150 psig separator pressure and 4 inches tubing diameter. Offshore region, Gas condensate – Best case scenario

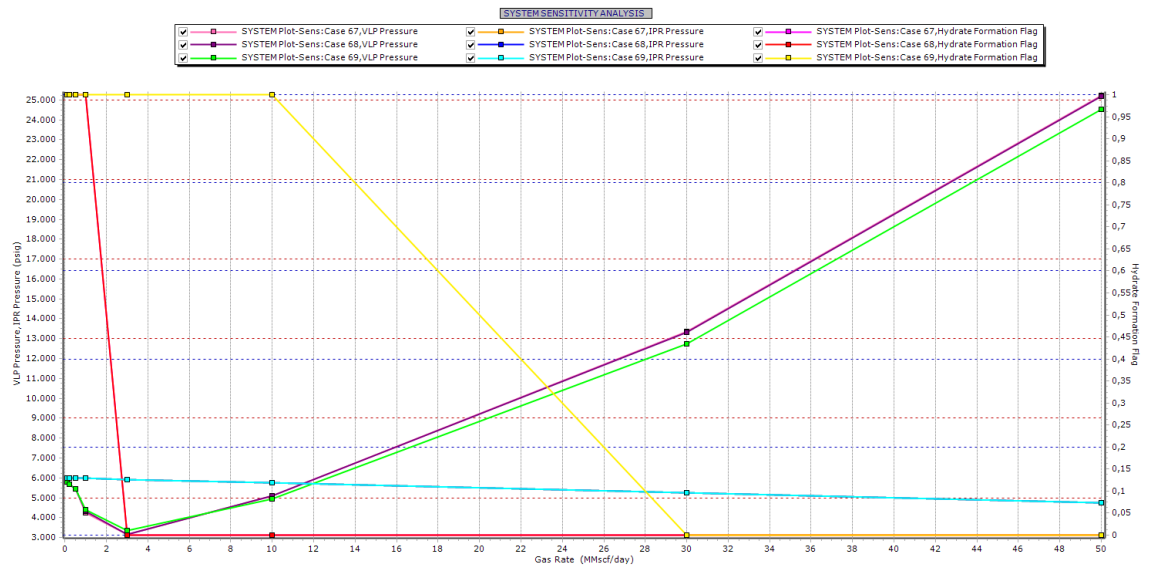


Figure 111: IPR – VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water rate is 250 STB/scf, 500 psig separator pressure and 3 inches tubing diameter. Offshore region, Gas condensate – Worst case scenario

6.1.3.4 Oil – Deep Offshore Region

In [Figure 112](#), [Figure 113](#) and [Figure 114](#) the best and the worst scenaria of insulated pipeline of an oil system in an offshore area are displayed.

For the best case scenario ([Figure 112](#)) the insulated pipeline requires less pressure to maintain the same production rate with the uninsulated one. The pressure losses are less in the scenario of an oil system in an offshore area. The reason this difference is observed is that temperature increase decreases the oil viscosity, which in turn decreases friction losses. Coupled with extended pipeline length, the total pressure needed is significantly less. Also delayed occurrence of the bubble point, initiates multiphase flow later and the frictional losses of the gas phase are minimized that way. The solution node (intersection of VLP and IPR) lies at 800 (uninsulated – green) and 820 STB/day (insulated –purple). The difference of the temperature conditions, which prevail inside the tubing for each case, is very important in the initial formation of the hydrate particles. This difference can be seen in [Figure 112](#) as the difference of the hydrate flag line for each scenario. When the flowline is not insulated the fluid enters the hydrate stability zone (below the yellow line). In the cases where some degree of insulation is achieved the solution oil rate is well outside the hydrate forming conditions. This difference between the hydrate flag lines can be observed also for the gas condensate in offshore region ([Figure 110](#)). The gas however travels much faster than the oil and manages to avoid entering into the hydrate stability conditions. The density and viscosity of the oil are paramount factors for the VLP curve and for flow assurance issues.

In [Figure 113](#) the worst case scenario is displayed. The water cut has its higher value (80%) and the hydrostatic pressure that is required to lift the fluids to the surface is greatly increased, so that no solution oil rate can be found. Thus, no flow can occur with this well design and the specific characteristics of the production fluids.

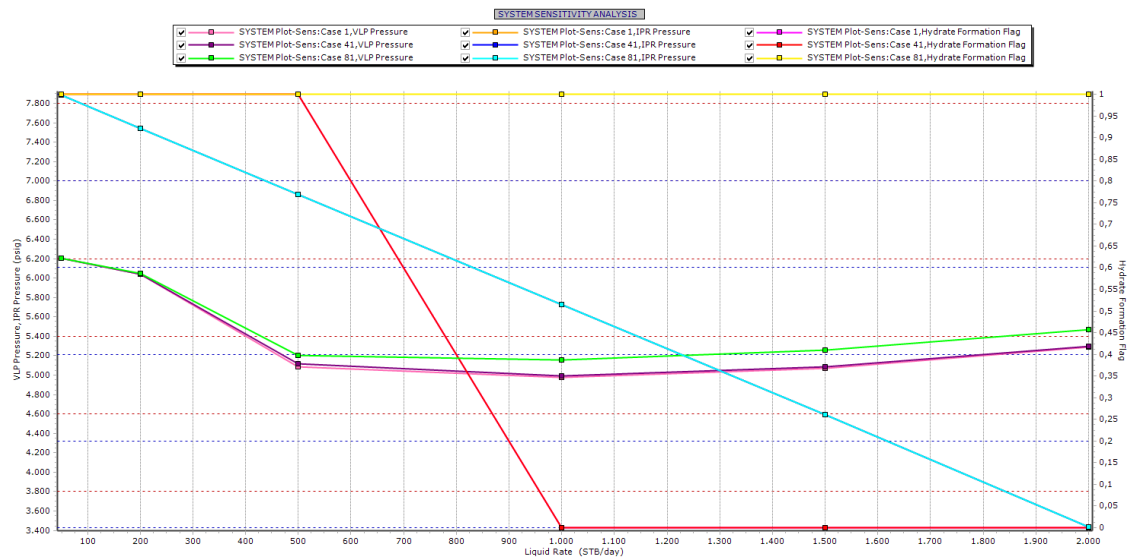


Figure 112: IPR – VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water cut is 0%, 150 psig separator pressure and 4 inches tubing diameter. Offshore region, Oil – Best case scenario

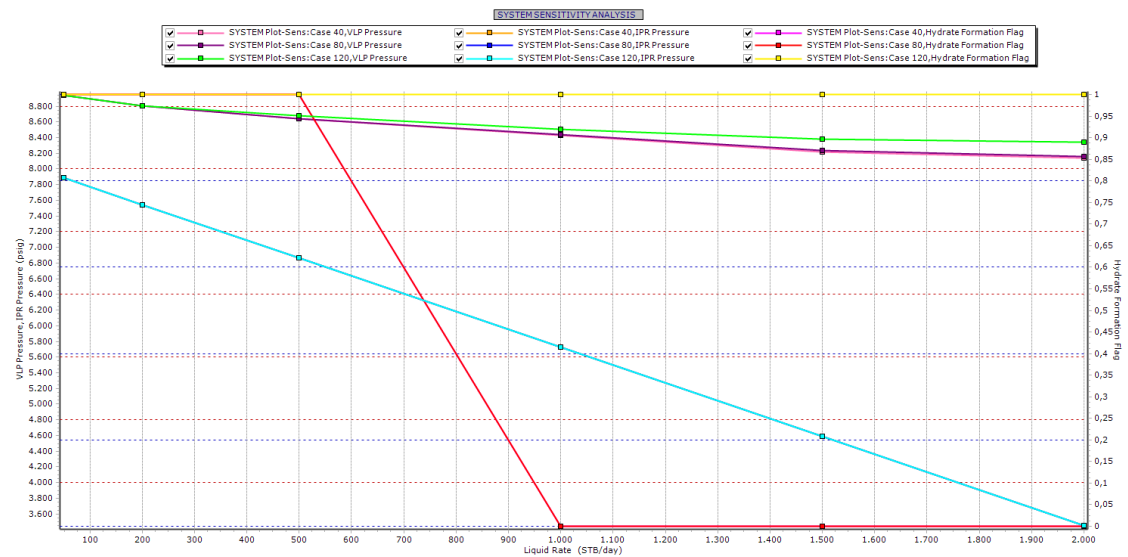


Figure 113: IPR – VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water cut is 80%, 500 psig separator pressure and 3 inches tubing diameter. Offshore region, Oil - Worst case scenario (No flow)

As worst scenario the case in Figure 114 is selected, where the water cut is lower than before (40%) and flow can be initiated and maintained. The difference on the VLP curve as effect of the insulation is the same, as already described in the previous paragraphs. The solution oil rate is around 200 STB/day for all the cases but is inside the hydrate stability region disregarding the pipeline insulation used.

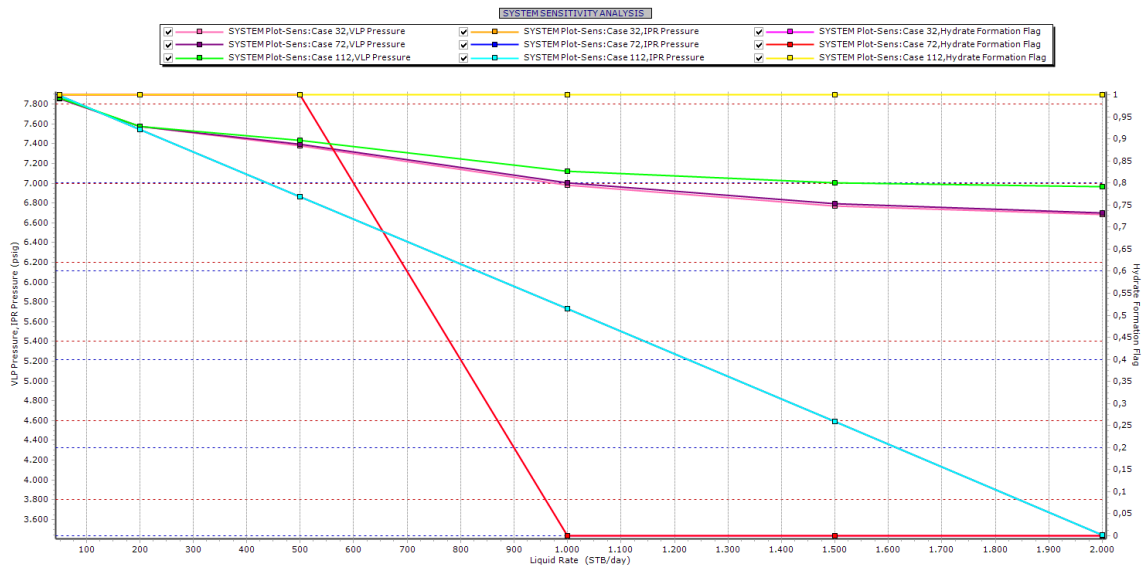


Figure 114: IPR - VLP curves intersection for 3, 0.12 and 0.03 BTU/h/ft²/°F Overall Heat Transfer coefficient. The water cut is 40%, 500 psig separator pressure and 4 inches tubing diameter. Offshore region, Oil - Worst case scenario

6.1.4 EFFECT OF WATER CONCENTRATION PERCENTAGE

Water percentage in the production fluid stream is a crucial parameter which affects the deliverability of the well and its production rate. The presence of water in the flowline is a phenomenon that occurs in every production system. Connate water always exists in the reservoir so it will be produced after depletion starts. Usually reservoir pressure drop goes hand to hand with the increase in water production except when certain precautions, such as subsea separators, are in place. In the specific scenario illustrated in this chapter the effect of water on the VLP curve will be examined, disregarding the potential pressure drop. It is well known that water has far greater density than oil or gas. With this in mind it is clear that the fluid column will be heavier the more water content it has. The hydrostatic pressure of the fluid column will be greater and the pressure differential needed to maintain flow and produce the fluids will be increased. This leads to an elevated VLP curve and subsequently lower production rates. Another issue is the compressibility of the water. As it is less compressible than oil or gas, its B_w is very low. Low B_w means that the amount initially displaced from the reservoir expands marginally. It occupies therefore an important volume downhole that if it would have been oil or gas it would have expanded and increased dramatically the rate of the hydrocarbon stream. Finally water has one property that can act beneficially for the hydrate formation during steady state conditions, but usually is masked from the problems discussed before. The heat capacity of the water is higher than oil or gas by twofold, that means that water can store more heat and therefore maintain higher temperature while travelling inside the flowline.

Five cases were chosen corresponding to different water percentages when producing oil and another five when producing gas condensate. The cases are as follows:

- For the oil system: Water of 0% (Pink), 10% (Purple), 20% (Green), 40% (Brown) and 80% (Blue).
- For the gas condensate system: Water of 1 (Pink), 10 (Purple), 50 (Green), 150 (Brown), 250 (Blue) STB/MMscf.

The figure shows two side-by-side windows for sensitivity analysis. The left window is titled 'Water Cut' and the right window is titled 'Water Gas Ratio'. Both windows have a 'Variables' list on the left and a 'Variable Data' table on the right. The 'Water Cut' window has a 'percent' column in the table, while the 'Water Gas Ratio' window has an 'STB/MMscf' column. Both windows have buttons for 'Reset', 'Generate', and 'Clear Data'.

Figure 15: Sensitivity Analysis window – Selection of water percentage and water gas ratio (WGR) scenarios for oil and gas condensate.

6.1.4.1 Gas Condensate - Arctic Region

The selection of cases displayed below is depicting only the worst and best case scenario regarding the system parameters. In [Figure 116](#) and [Figure 117](#) the best and the worst scenario for a gas condensate system in an arctic region are described.

The water to gas rate (WGR) has an important effect on the VLP curve of a gas condensate. This effect for operating conditions results in large differences on the solution flow rate (in the specific case 22 (Blue) to 37 (Pink) MMscf/day) ([Figure 116 – Best case scenario](#)). Increased water concentration inside the pipeline implies heavier fluid column, due to its increased density. The hydrostatic head of the column is increased and higher pressure differential is required to maintain the solution gas rate.

For extreme cases of water rates other phenomena also appear in the system. One of them is the liquid hold-up effect. The multiphase fluid inside the pipeline occupies a certain area of the tubing cross section. When two phases have very different densities, the one with lower density (gas) moves faster than the other and needs less area to maintain the flowrate. The fluid of higher density (water & condensate) occupies a large area and may obstruct the flow by filling the whole cross section. If all the area is filled with liquid, gas will be trapped underneath and the well will not flow until a certain pressure differential is developed. The flow regime in that case is intermittent (slug flow) and needs larger pressures differentials periodically. The average pressure differential needed is higher and the flow rate due to intermissions is lowered. Also it is dangerous regarding hydrate formation due to the increased time the fluid spends inside the flowline (enough for the temperature to drop).

Higher gas rates lead to increased deviation between the VLP curves. The reason is that the gas condensate has to obtain high velocities to achieve these high rates. High velocities increase the frictional pressure losses of the fluid. The more water is inside the pipeline the gas has to achieve even higher velocities to achieve the flowrate while travelling in a fraction of the flowline (slip effect).

The possible positive effect of the water presence is that it can store heat at double efficiency than hydrocarbon streams. By increasing the total heat capacity of the system, the total time that a fluid can spend inside a cold flowline (before reaching the hydrate stability region) is longer than the same fluid without the water. Despite that increased water rate increases the heat capacity of the production fluid, its effect is masked from the negative effects water has.

For the specific case, the pipelines under every water gas ratio (WGR) checked is operated well outside the hydrate stability region (Red – Orange hydrate flag line).

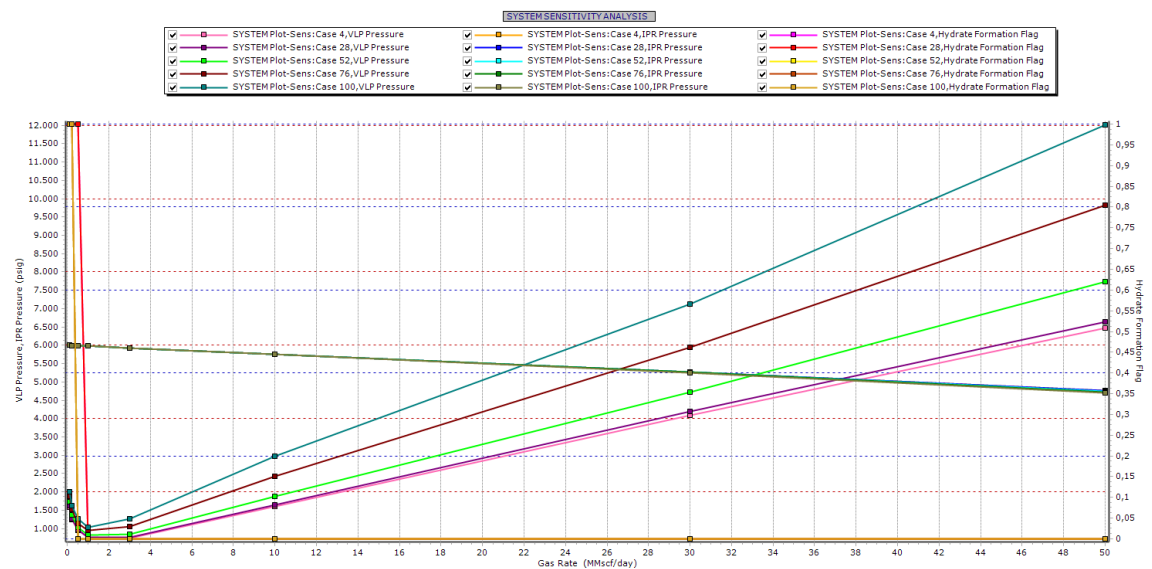


Figure n6: IPR – VLP curves intersection for 1, 10, 50, 150 and 250 STB/MMscf water to gas ratio (WGR). The overall heat transfer coefficient is 0.03 BTU/h/ft²/°F, 150 psig separator pressure and 4 inches tubing diameter. Arctic region, Gas condensate – Best case scenario

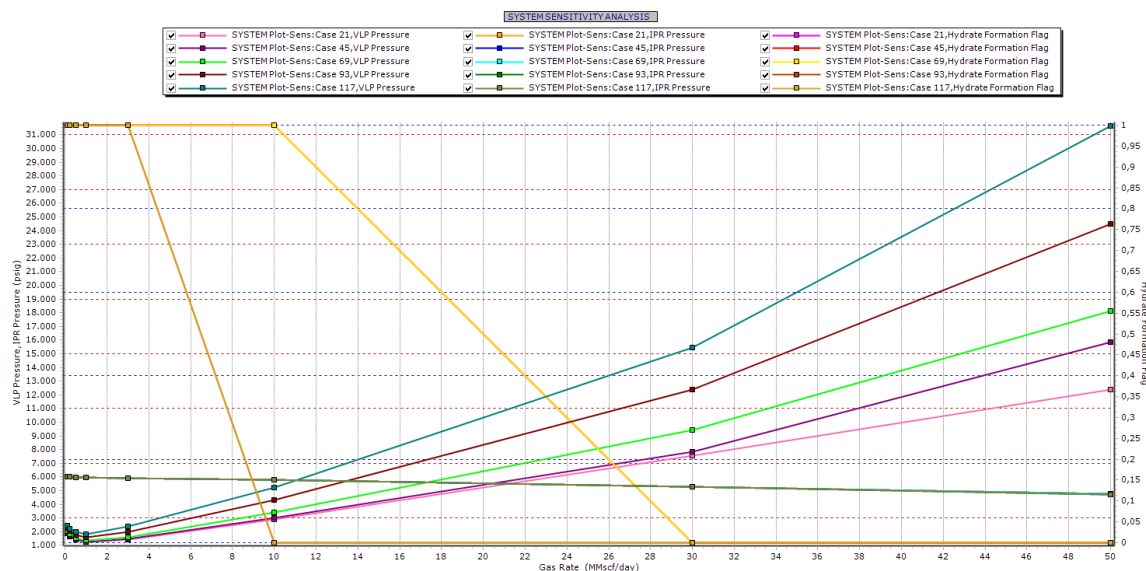


Figure 117: IPR – VLP curves intersection for 1, 10, 50, 150 and 250 STB/MMscf water to gas ratio (WGR). The overall heat transfer coefficient is 3 BTU/h/ft²/°F, 500 psig separator pressure and 3 inches tubing diameter. Arctic region, Gas condensate – Worst case scenario

In the worst scenario (Figure 117) the same behavior, already explained, is observed. The solution gas rate ranges from 10 (Blue) to 21 (Pink) MMscf/day. The fluid passes inside the hydrate stability region only when the highest water rate case is tested. 250 STB/MMscf provides a significant water production in the surface (2750 STB/day). It is clear that a large cross section of the flowline is occupied by water and liquid hold up effect increases the required hydrostatic pressure of the column, which in turn needs to be beaten. On the contrary for cases with lower water rate the same problem does not occur and it is safe to operate in these conditions.

6.1.4.2 Oil – Arctic Region

The selection of cases displayed below is depicting only the worst and best case scenarios regarding the system parameters. In Figure 118 and Figure 119 the best and the worst scenario for an oil system in an offshore region are described.

The water cut of an oil stream is one of the most important parameters affecting a VLP curve. In oil systems, where the pressure losses due to gravity are more important than friction losses water plays a critical role. The high density of the water coupled with high concentration results in extremely high hydrostatic head of the fluid column. The pressure needed to lift such fluid is extreme and even then will result in low production rates. This can be seen in the Figure 115 where the pressure difference of the VLP curves is present regardless the production rate. For operating conditions the

solution flow rate ranges from 700 (Blue) to 1600 (Pink) STB/day) (Figure 118 – Best case scenario).

The possible positive effect of the water presence is that it can store heat at double efficiency than hydrocarbon streams. By increasing the total heat capacity of the system, the total time that a fluid can spend inside a cold flowline (before reaching the hydrate stability region) is longer than the same fluid without the water. Despite that increased water rate increases the heat capacity of the production fluid, its effect is masked from the negative effects water has.

For the specific case, the pipelines under every water cut checked are operated well outside the hydrate stability region (Orange hydrate flag line). So no problem regarding hydrate formation is probable.

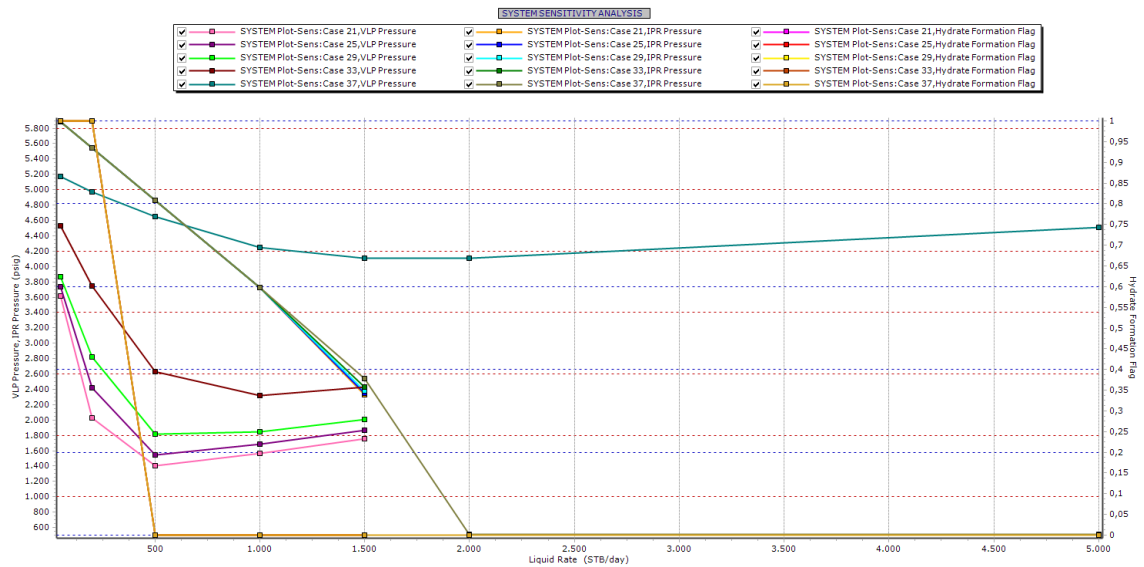


Figure 118: IPR – VLP curves intersection for 0, 10, 20, 40 and 80% water cut. The overall heat transfer coefficient is 0.03 BTU/h/ft²/°F, 150 psig separator pressure and 4 inches tubing diameter. Arctic region, Oil – Best case scenario

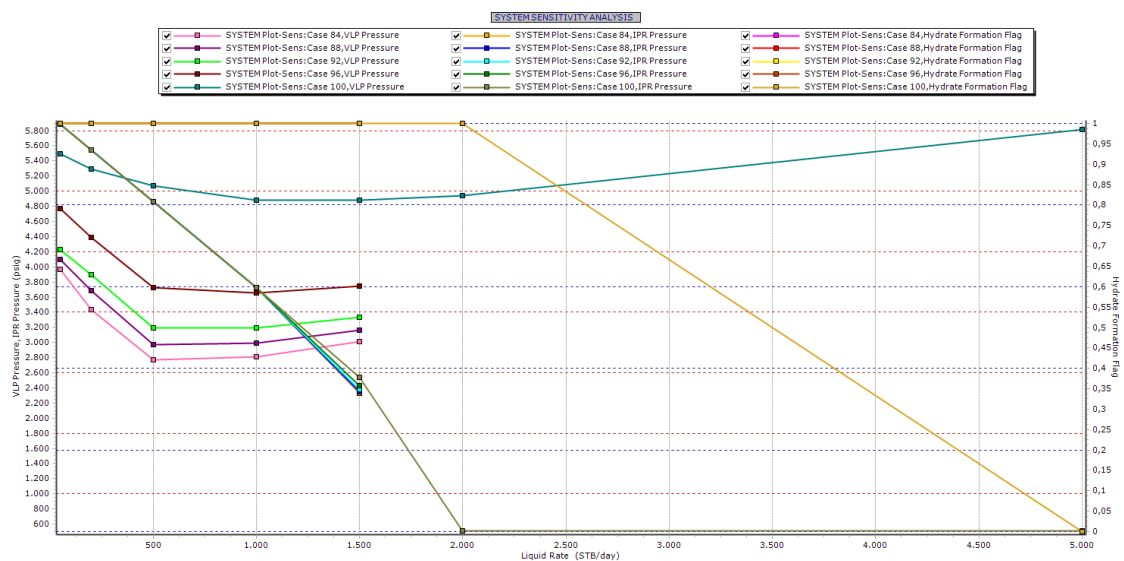


Figure 119: IPR – VLP curves intersection for 0, 10, 20, 40 and 80% water cut. The overall heat transfer coefficient is 3 BTU/h/ft²/°F, 500 psig separator pressure and 3 inches tubing diameter. Arctic region, Oil – Worst case scenario

In Figure 119 the worst case is displayed. The difference on the VLP curve is the same for all oil rates, as explained already the main difference is the increased density that leads to gravitational losses. The solution oil rate is around 400 STB/day for water cut of 80% and reaches to 1300 STB/day for hydrocarbon stream with 0% water cut. It is evident how much the concentration affects the gravitational losses in oil systems (Increase of the column hydrostatic head from 50% to 180%, 50% water cut and 80% respectively). All solution rates force the pipeline to operate inside the hydrate stability region regardless the water rate. The reason is mostly the insulation presence and not so much the flow rate decrease in the specific case, as the oil travels with similar speed in the best and worst scenarios.

6.1.4.3 Gas Condensate - Deep Offshore Region

The selection of cases displayed below is depicting only the worst and best case scenarios regarding the system parameters. In Figure 120 and Figure 121 the best and the worst scenario for a gas condensate system in an arctic region are described.

As already discussed the water to gas rate (WGR) has an important effect on the VLP curve of a gas condensate. This effect for operating conditions results in large differences on the solution flow rate (in the specific case 13 (Blue) to 28 (Pink) MMscf/day) (Figure 120 – Best case scenario). The difference compared with the gas condensate scenario in the arctic region is that the pipeline is longer and that

dramatically increases the frictional pressure losses. Finally it results to lower solution rates and operating conditions closer to the hydrate stability zone.

Liquid hold-up effect is more important in the offshore scenario where risers are more elevated and if severe slippage is encountered between phases the pipe is filled with liquid and operate with intermittent flow regime increasing that way the pressure losses and decreasing the production rate. This phenomenon gets worse as the water rate increases.

Higher gas rates also lead to increased deviation between the VLP curves. The reason is that the gas condensate has to obtain high velocities to achieve these high rates. High velocities increase the frictional pressure losses of the fluid and the more water is inside the pipeline the gas has to achieve even higher velocities to achieve the flowrate while travelling in a fraction of the flowline (slip effect). The effect is again worse in the offshore pipeline where the length increases the frictional losses.

The possible positive effect of the water presence is that it can store heat at double efficiency than hydrocarbon streams. Although the increased water rate increases the heat capacity of the production fluid, its effect is masked from the negative effects water has.

For the specific case, the pipelines under every water gas ratio (WGR) checked is operated well outside the hydrate stability region (Red – Orange hydrate flag line).

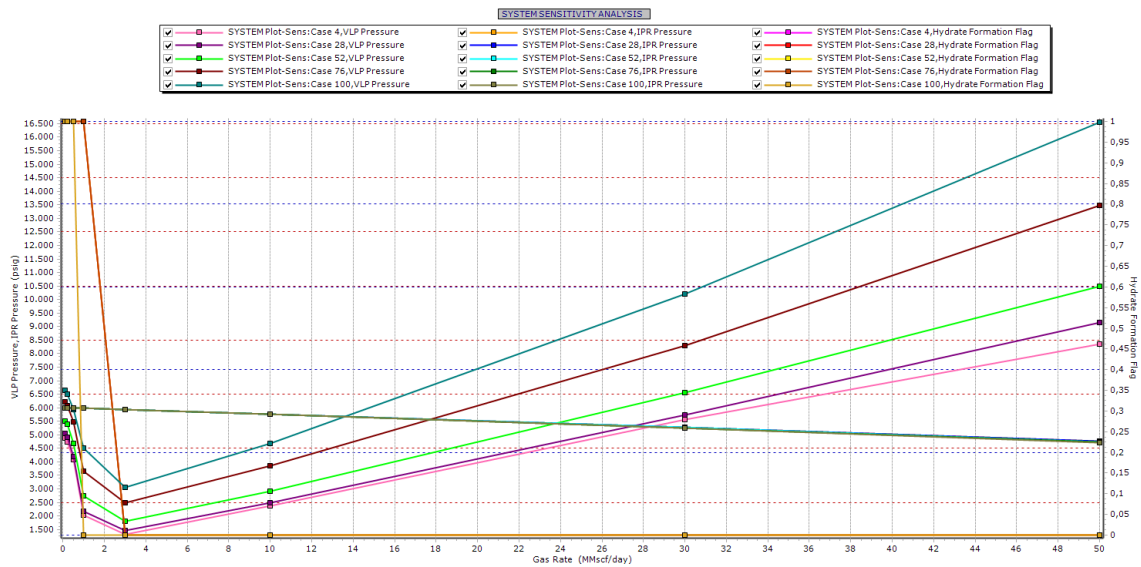


Figure 120: IPR – VLP curves intersection for 1, 10, 50, 150 and 250 STB/MMscf water to gas ratio (WGR). The overall heat transfer coefficient is 0.03 BTU/h/ft²/°F, 150 psig separator pressure and 4 inches tubing diameter. Offshore region, Gas condensate– Best case scenario

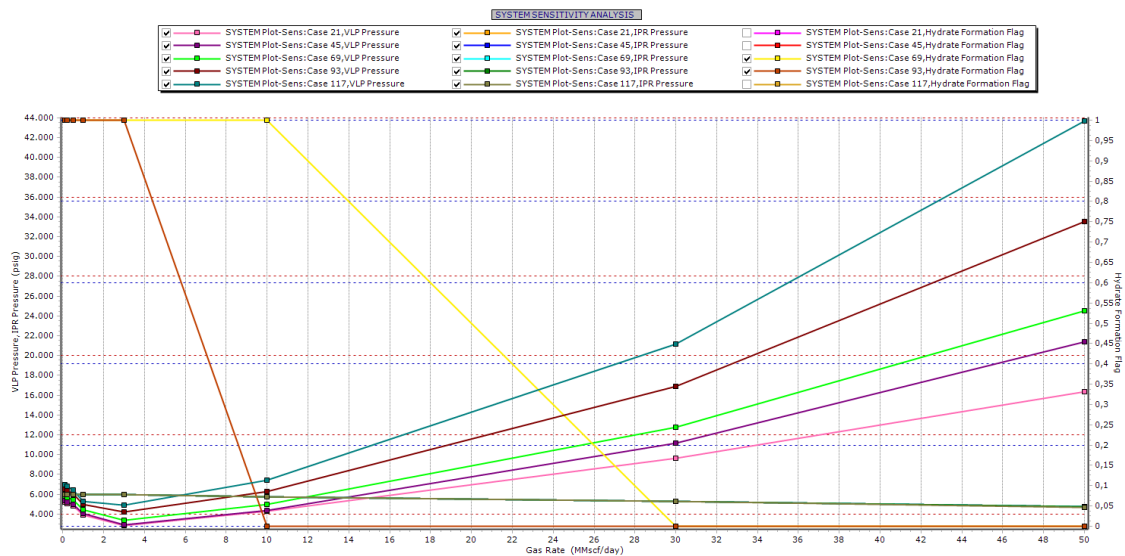


Figure 121: IPR – VLP curves intersection for 1, 10, 50, 150 and 250 STB/MMscf water to gas ratio (WGR). The overall heat transfer coefficient is 3 BTU/h/ft²/°F, 500 psig separator pressure and 3 inches tubing diameter. Offshore region, Gas condensate – Worst case scenario

In the worst scenario (Figure 121) the same behavior that is already explained is observed. The solution gas rate ranges from 6 (Blue) to 14 (Pink) MMscf/day. The fluid passes inside the hydrate stability region only when the highest water rates of 150 and 250 STB/MMscf are tested. In those cases increased water production in the surface (2750 STB/day) imply that a large cross section of the pipe is filled with water

and liquid hold up effect increases the required hydrostatic head, which needs to be beaten. In those cases very low rates are achieved and the fluid velocity is severely low. Hydrate formation under these conditions is a certain risk. On the contrary for cases with lower water rate the same problem does not occur and it is safe to operate in these conditions.

6.1.4.4 Oil – Deep Offshore Region

The cases displayed below are depicting only the worst and best case scenarios regarding the system parameters. In [Figure 122](#) and [Figure 123](#) the best and the worst scenario for an oil system in an offshore region are described respectively.

The water cut of an oil stream is one of the most important parameters affecting a VLP curve. In oil systems gravity pressure losses are more important than friction losses because water plays a critical role on the total fluid density. The pressure differential required for the production of the fluid is extreme and the solution oil rate is low if any. This can be seen in the [Figure 122](#) and [Figure 123](#), where the pressure difference of the VLP curves is steady regardless the production rate. The extended length and elevation of the offshore production line result in greatly increased pressure needs. For very high water cut percentage (80%, [Figure 122 - Blue](#)) the well will not flow. Under operating conditions the solution flow rate ranges from 800 (Blue) to 1300 (Pink) STB/day for water cuts of 40% and 0% respectively ([Figure 122](#)).

By increasing the total heat capacity of the system, the total time that a fluid can spend inside a cold flowline (before reaching the hydrate stability region) is longer than the same fluid without the water. The possible positive effect of increased heat storage that water presence has effect is masked from the negative effects water has on gravitational and frictional losses.

For the specific case, the pipelines under every water cut checked are operated well outside the hydrate stability region ([Figure 122 - Orange hydrate flag line](#)). So no problem regarding hydrate formation is probable, except the obvious scenario of 80% water cut where the well does not flow.

In [Figure 123](#) the worst case is displayed. The difference on the VLP curve is the same for all oil rates, as explained already the main difference is the increased gravitational losses due to density change. The solution oil rate is around 100 STB/day for water cut of 40% and reaches to 800 STB/day for hydrocarbon stream with 0%

water cut. It is obvious how much the concentration affects the gravitational losses in oil systems (Increase of the column hydrostatic head from 50% to 180%, 50% water cut and 80% respectively). For water cut of 80% the well is not able to deliver fluids as expected. All solution rates for the rest of the cases force the pipeline to operate inside the hydrate stability region regardless the water rate. The reason is mostly the insulation absence and secondly the flow rate decrease.

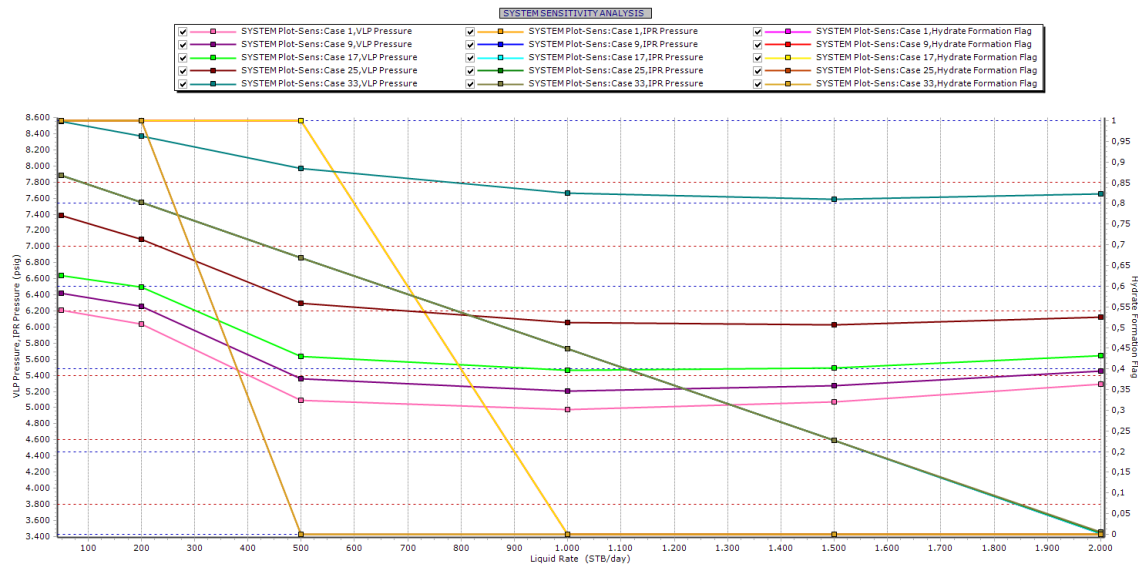


Figure 122: IPR – VLP curves intersection for 0, 10, 20, 40 and 80% water cut. The overall heat transfer coefficient is 0.03 BTU/h/ft²/°F, 150 psig separator pressure and 4 inches tubing diameter. Offshore region, Oil – Best case scenario

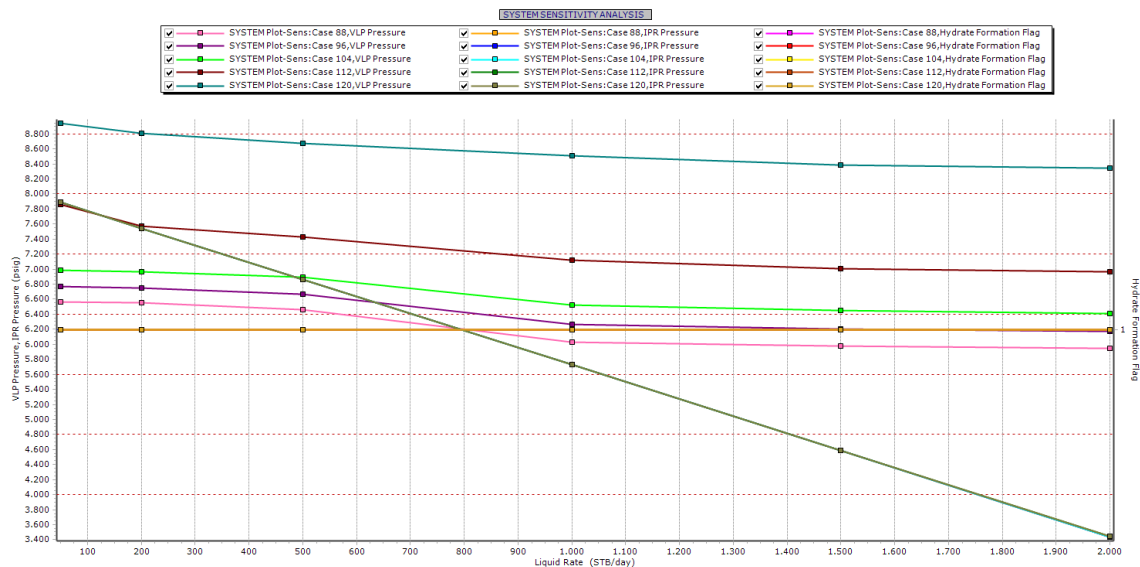


Figure 123: IPR – VLP curves intersection for 0, 10, 20, 40 and 80% water cut. The overall heat transfer coefficient is 3 BTU/h/ft²/°F, 500 psig separator pressure and 3 inches tubing diameter. Offshore region, Oil – Worst case scenario

6.2 EFFECT OF THERMODYNAMIC INHIBITORS ON THE HYDRATE FORMING CONDITIONS

In this chapter the effect of thermodynamic hydrate inhibitor in the flow stream will be discussed. In the scenarios described in chapter 5 (Arctic and offshore region, oil and gas condensate systems) PROSPER's option to include hydrate inhibiting substances in the flow will be used. For thermodynamic inhibitors Methanol Glycol (MEG) will be used, as well as traditional Methanol. When testing the system behavior for hydrate formation it was observed that for the best case scenarios no hydrate formation was present. The use of hydrate inhibitors in these cases is without a purpose, so only for the worst cases the THI's will be introduced in the flow stream. The results regarding the hydrate stability region (dissociation curve) change will be discussed by using different concentration of THI's. Finally the risk for hydrate problem will be assessed again and Pressure versus Temperature along Depth diagrams will be plotted. In reality the THI concentration changes the fluid composition and its properties, but PROSPER does not alter the properties of the original stream whatsoever. It only uses the THI to change the hydrate dissociation curve. That introduces an important error in calculations.

The composition of the hydrate inhibitors used is 10 and 20% (v/v of water in the flow stream) for MEG and Methanol in gas condensate systems. For oil systems 10, 20 and 30% (v/v of water in flow stream) of MEG and Methanol will be used. The reason higher percentage values are not tested is that PROSPER has a weak representation of the hydrate dissociation curve. For high percentages it provides very few points of the curve, so it is not accurate to proceed designing it.

In [Figure 124](#) and [Figure 125](#) the pressure and temperature values of the hydrate dissociation curves are given for oil and gas condensate respectively. More analytically the cases for oil systems are: stream without inhibiting substances, stream with inhibiting salts at a concentration of 20000 ppm (more details given in chapter 4.1.2.), Methanol-Glycol (MEG) with concentration 10-20-30% (v/v of water) and finally Methanol of concentrations 10-20-30% (v/v of water) ([Figure 124](#)). For gas condensate systems the cases are: stream without inhibiting substances, stream with inhibiting salts at a concentration of 20000 ppm (more details given in chapter 4.1.1.), Methanol-Glycol (MEG) with concentration 10-20% (v/v of water) and finally Methanol of concentrations 10-20% (v/v of water) ([Figure 125](#)).

Assessment of Hydrate Formation Parameters in Production Wells

Oil															
No salts - No inhibitors		Salts - No inhibitors		Salts - MEG 10%		Salts - MEG 20%		Salts - MEG 30%		Salts - Methanol 10%		Salts - Methanol 20%		Salts - Methanol 30%	
T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)
35	65.8273	35	90.1759	35	339.174	35	2290.01	35	13343.3	35	296.204	35	1707.18	35	11952.3
38	86.6278	38	116.710	38	434.922	38	3524.09	38	15744	38	379.271	38	2755.53	38	14358.6
41	112.410	41	149.716	41	570.227	41	5009.10	41	18355.5	41	488.679	41	4109.82	41	16850
44	144.430	44	191.035	44	757.064	44	6723.87	44	21064.3	44	642.814	44	5687.30	44	19496
47	184.416	47	243.183	47	1060.91	47	8649.95			47	870.036	47	7487.48	47	22252.6
50	234.707	50	309.653	50	1692.73	50	10732.6			50	1251.94	50	9493.21		
53	298.622	53	395.592	53	2672.28	53	13053.2			53	2058.15	53	11687.4		
56	380.866	56	508.742	56	3915.83	56	15511.1			56	3142.61	56	14068.5		
59	488.545	59	661.855	59	5424.93	59	18124.9			59	4495.99	59	16571.4		
62	633.118	62	878.615	62	7159.82	62	20849.3			62	6091.40	62	19226		
65	835.088	65	1191.23	65	9137.20					65	7900.70	65	21999.1		
68	1107.68	68	1831.27	68	11264.4					68	9931.73				
71	1682.94	71	2678.30	71	13633.4					71	12183				
74	2477.66	74	3707.96	74	16067.7					74	14559.9				
77	3456.12	77	4899.47	77	18746.8					77	17121.5				
80	4601.11	80	6238.02	80	21520.5					80	19822				
83	5897.78	83	7710.05							83	22643.3				
86	7329.37	86	9301.97												
89	8883.61	89	11002.2												
92	10549.4	92	12799.8												
95	12312.8	95	14683.8												
98	14167.1	98	16642.6												

Figure 124: Oil - Hydrate dissociation curves with their respective inhibitor percentage

Gas condensate											
No salts - No inhibitors		Salts - No inhibitors		Salts - MEG 10%		Salts - MEG 20%		Salts - Methanol 10%		Salts - Methanol 20%	
T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)	T (oF)	P (psig)
35	100.940	35	124.064	35	365.361	35	3021.61	35	322.756	35	802.683
38	129.006	38	159.174	38	464.629	38	4214.15	38	409.154	38	1108.40
41	163.587	41	200.857	41	597.132	41	5578.79	41	522.615	41	1695.83
44	206.410	44	252.818	44	781.254	44	7103.98	44	676.190	44	2577.03
47	259.794	47	318.179	47	1056.97	47	8779.55	47	895.047	47	3670.82
50	326.917	50	401.364	50	1550.56	50	10598	50	1240.05	50	4954.91
53	412.417	53	509.121	53	2337.60	53	12552.6	53	1870.25	53	6415.22
56	523.174	56	652.162	56	3343.43	56	14637.6	56	2752.23	56	8040.98
59	670.331	59	849.511	59	4542.48	59	16847.6	59	3841.64	59	9823.28
62	873.748	62	1141.46	62	5919.28	62	19177.8	62	5119.18	62	11754.4
65	1175.73	65	1640.99	65	7462.28	65	21623.3	65	6571.39	65	13827
68	1691.11	68	2390.38	68	9162.85			68	8188.43	68	16035.7
71	2446.57	71	3353.23	71	11013.3			71	9961.77	71	18373.7
74	3414.18	74	4509.91	74	13007.2			74	11884.4	74	20837.2
77	4571.63	77	5843.03	77	15133.6			77	13944.4		
80	5906.56	80	7343.71	80	17390.4			80	16140.5		
83	7408.15	83	9004.31	83	19774.1			83	18466.6		
86	9068.78	86	10816.1	86	22278.6			86	20917.6		
89	10880.3	89	12774								
92	12836.1	92	14871.8								
95	14931.4	95	17103.8								
98	17160.9	98	19465.8								

Figure 125: Gas condensate - Hydrate dissociation curves with their respective inhibitor percentage

6.2.1 GAS CONDENSATE - ARCTIC REGION

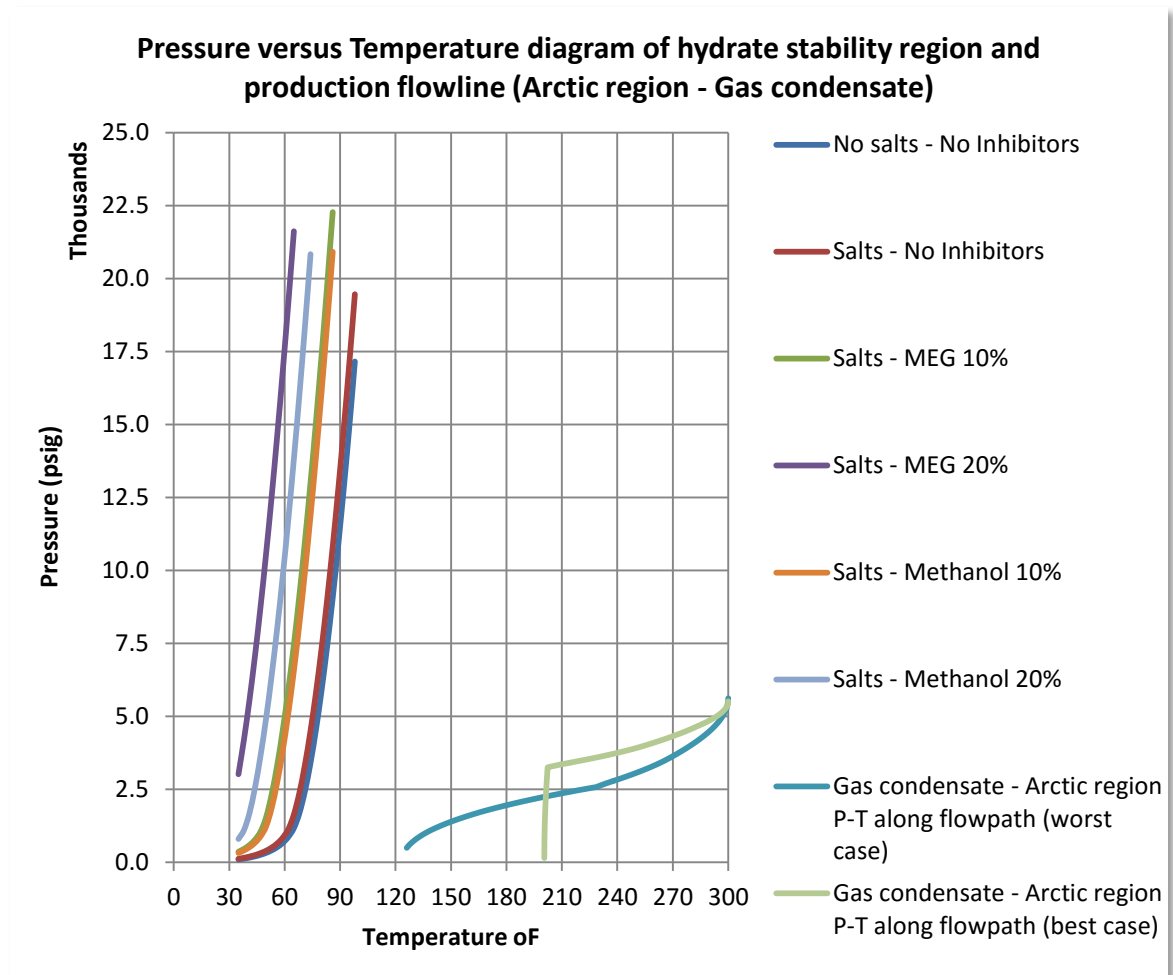


Figure 126: Effect of Thermodynamic Inhibitors (THIs) on the hydrate dissociation curve and P-T diagrams of the production flowlines from separator to the perforations (Arctic region - Gas condensate).

In Figure 126 and Figure 127 the hydrate dissociation curves are displayed in the left hand side of the diagrams. The dissociation curves are the same for the arctic and offshore region as the same gas condensate composition exist in the pipelines. The curves further to the right require less pressure for the same temperature to form stable hydrate particles compared to the curves that are on the left side. So the hydrate problem is more dangerous for these cases. As the inhibitor percentage increases the curve moves to the left by maintaining its slope. The uninhibited flow stream has a hydrate dissociation curve 3°F increased for the same temperature compared to the flow stream with 20,000 ppm salt content. An increase of 10% MEG decreases the hydrate formation temperature further by 15°F, while the same increase in Methanol gives a slightly worse result with a decrease in required temperature of 13°F. So far

MEG and Methanol have nearly the same inhibiting behavior. However when the concentration of THIs is further increased the two chemical compounds provide different results. A percentage of 20% MEG reduces the required temperature further by 18°F, while the same amount of Methanol decreases the temperature only by 8°F further. If the percentages would have been further increased this difference would have increased further. The reason higher concentrations were not tested is that PROSPER does not provide enough points of the dissociation curve. Also it does not give values below 35°F where low pressures, in which we are interested, are present.

The other two lines in the diagrams describe the production flow path of the best and worst case scenario in the arctic and offshore region respectively. More specifically they show the pressure and temperature prevailing in the pipeline at every depth, from the perforations to the separator. The separator (surface) conditions are given by the left node of the lines and the reservoir conditions by the right node, which is the reason they start from the same point. The green and purple lines showcase the best case scenarios where all system parameters have optimal values that facilitate the flow outside the hydrate phase envelope. The blue and red lines on the contrary show the worst case ones.

The fluid of the best case scenario ([Figure 126 - Green](#)) starts from reservoir conditions (6000 psig and 300°F) and maintains high temperature till it reaches the separator (200°F). This happens due to the high velocity the fluid has, compared to the worst case (Blue), which does not allow heat exchange to the environment to be high. Also from the diagram can be seen that the finishing point (separator) has lower pressure value (150 psig) than the worst case (500 psig). The reason a sharp drop only in pressure is observed, before reaching the separator conditions, is because the surface pipeline (in contrast with the downhole one, has very good insulating material and coupled with high fluid velocity it prevents temperature drop. The pipeline operates very far outside the hydrate stability region and does not require the help of thermodynamic inhibitors to produce fluids without problems.

The worst case scenario pipeline ([Figure 126 - Blue](#)) operates from reservoir conditions (6000 psig and 300°F) and its temperature drops steadily till it reaches the separator (no insulating material). The fluid enters the separator at 130°F temperature. However neither in this case hydrate problems are present because the pipeline operates outside the hydrate stability region, thanks to the high velocity the fluid has.

6.2.2 GAS CONDENSATE – OFFSHORE REGION

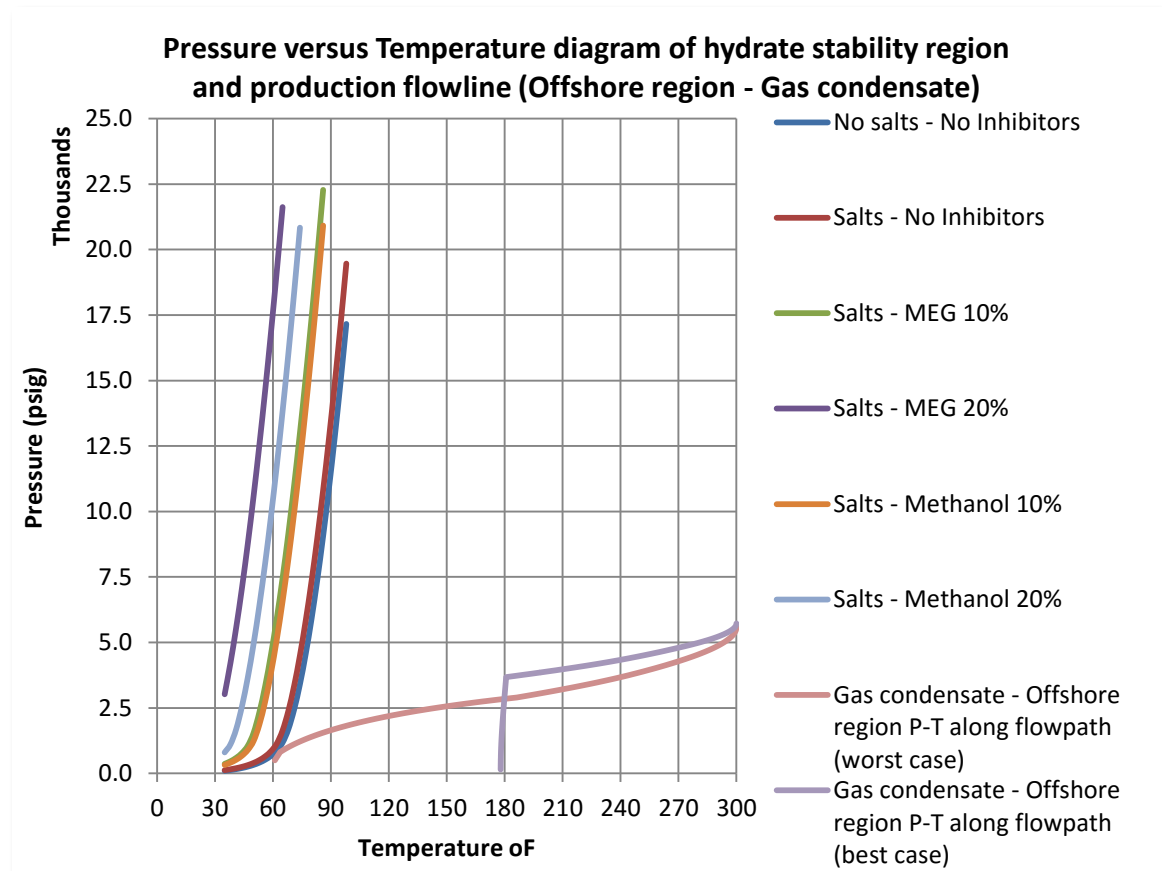


Figure 127: Effect of Thermodynamic Inhibitors (THIs) on the hydrate dissociation curve and P-T diagrams of the production flowlines from separator to the perforations (Offshore region – Gas condensate).

The fluid of the best case scenario (Figure 127 - Purple) starts from reservoir conditions (6000 psig and 300°F) and maintains high temperature till it reaches the separator (180°F). This happens due to the high velocity the fluid has, compared to the worst case (Red), which does not allow heat exchange to the environment to be high. Also from the diagram can be seen that the finishing point (separator) has lower pressure value (150 psig) than the worst case (500 psig). The reason a sharp drop only in pressure is observed, before reaching the separator conditions, is the same described previously in the arctic region pipeline.

The worst case scenario pipeline (Figure 127 - Red) operates from reservoir conditions (6000 psig and 300°F) and its temperature drops steadily till it reaches the separator (no insulating material). The fluid enters the separator at 60°F temperature. Neither in this case hydrate problems are present because the pipeline operates outside

the hydrate stability region, but the distance of the finishing with the dissociation curve when no inhibitors are used is marginal. Although numerically it seems that the pipeline can operate under these conditions it is wise to use a small amount of thermodynamic inhibitor to prevent possible problems.

6.2.3 OIL - ARCTIC REGION

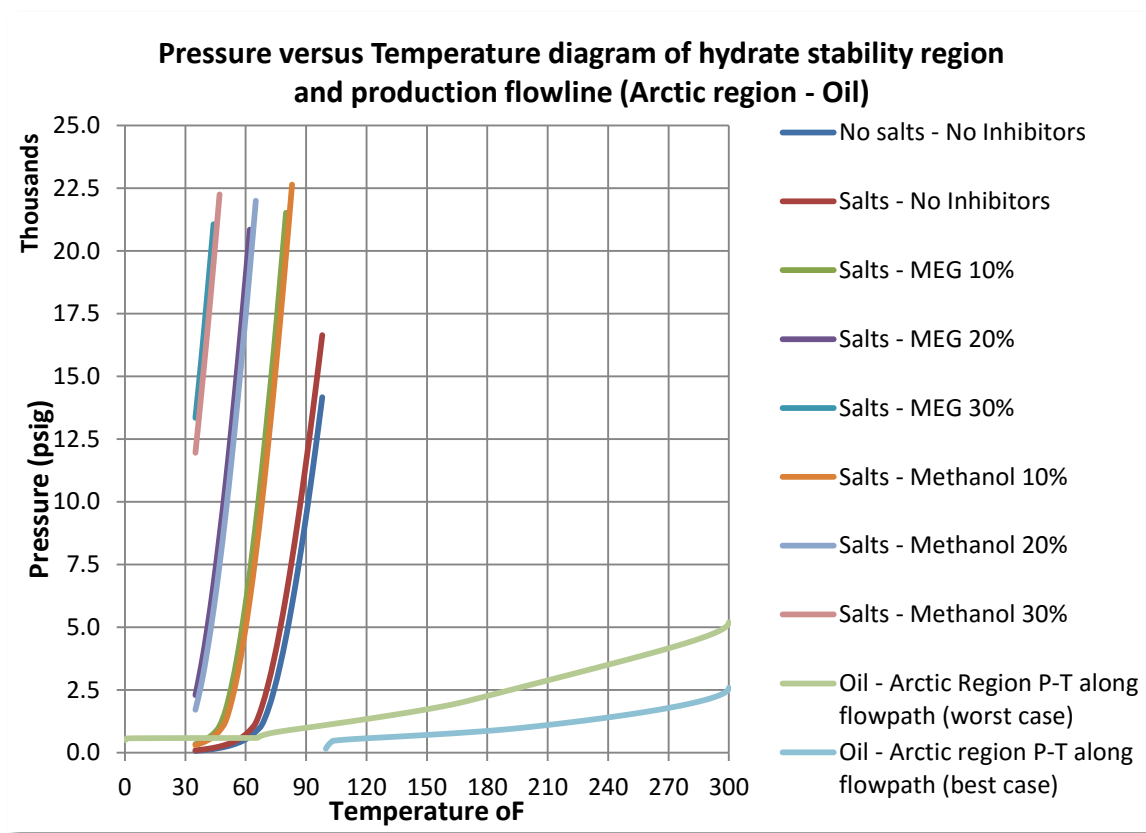


Figure 128: Effect of Thermodynamic Inhibitors (THIs) on the hydrate dissociation curve and P-T diagrams of the production flowlines from separator to the perforations (Arctic region - Oil).

In Figure 128 and Figure 129 the hydrate dissociation curves are displayed in the left hand side of the diagrams. The dissociation curves are the same for the arctic and offshore region as the same oil composition exist in the pipelines. The curves to the right require more temperature for the same pressure to form stable hydrate particles compared to the curves that are on the left side. So the hydrate problem is more dangerous for these cases. As the inhibitor percentage increases the curve moves to the left as it increases its slope. The uninhibited flow stream has a hydrate dissociation curve 3°F increased for the same temperature compared to the flow stream with 20,000 ppm salt content. An increase of 10% MEG decreases the hydrate formation temperature further by 18°F, while the same increase in Methanol gives a slightly worse

result, with a decrease in required temperature of 17°F. So far MEG and Methanol have nearly the same inhibiting behavior. In contrast with the behavior of THIs in a gas condensate stream, when the concentration of THIs is further increased, the two chemical compounds provide similar results in the case of oil stream. A percentage of 20% MEG reduces further the required temperature by 18°F, while the same amount of Methanol decreases the temperature by the same number. The same is true for further increase of the inhibitor percentage. Both MEG and Methanol of 30% percentage further decrease the temperature required for hydrate formation by 18 °F. The reason higher concentrations than 30% were not tested, is that PROSPER does not provide enough points of the dissociation curve. Also it does not give values below 35°F where low pressures, in which we are interested, are present.

The other two lines in the diagrams describe the production flow path of the best and worst case scenario in the arctic and offshore region respectively. More specifically they show the pressure and temperature prevailing in the pipeline at every depth, from the perforations to the separator. The separator (surface) conditions are given by the left node of the lines and the reservoir conditions by the right node, which is the reason they start from the same point. The blue lines showcase the best case scenarios where all system parameters have optimal values that facilitate the flow outside the hydrate phase envelope. The green on the contrary show the worst case ones.

The fluid of the best case scenario ([Figure 128 - Blue](#)) starts from reservoir conditions (6000 psig and 300°F) and maintains high temperature till it reaches the separator (100°F). This happens due to the high velocity the fluid has, compared to the worst case (Green), which does not allow heat exchange to the environment to be high. Also from the diagram can be seen that the finishing point (separator) has lower pressure value (150 psig) than the worst case (500 psig). The reason a small drop only in pressure is observed, before reaching the separator conditions, is because the surface pipeline (in contrast with the downhole one, has very good insulating material and coupled with high fluid velocity it prevents temperature drop. The pipeline operates far outside the hydrate stability region and does not require the help of thermodynamic inhibitors to produce fluids without problems.

The worst case scenario pipeline ([Figure 128 - Green](#)) operates from reservoir conditions (6000 psig and 300°F) and its temperature drops steadily till it reaches the separator (no insulating material). The fluid enters the separator at 0°F temperature and 500 psig. In the arctic region the environment has very low temperature. The oil

with not favorable system configuration has a very low solution production rate and moves with very low velocity. The time required to travel from the reservoir to the separator is high and without insulating material to obstruct heat losses the fluid temperature reaches the environmental one. In this case hydrate problems are present because the pipeline operates inside the hydrate stability region after a certain point. The use of THIs in this case yields no results as the fluid is so deep in the hydrate zone. Possible solution in the form of inhibitors could be kinetic one or anti-agglomerants to maintain with cold flow and produce the hydrates formed.

6.2.4 OIL – OFFSHORE REGION

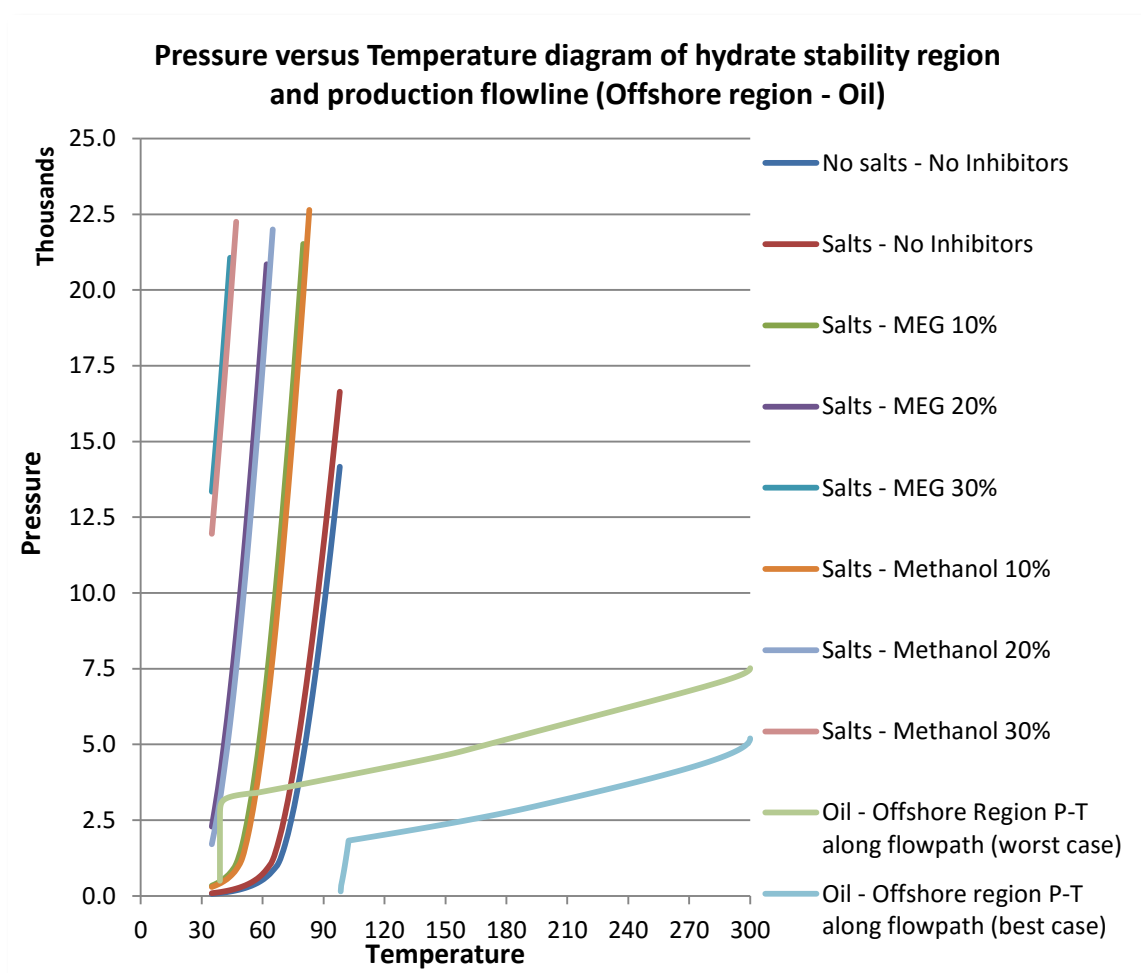


Figure 129: Effect of Thermodynamic Inhibitors (THIs) on the hydrate dissociation curve and P-T diagrams of the production flowlines from separator to the perforations (Offshore region – Oil).

The fluid of the best case scenario (Figure 129 - Blue) starts from reservoir conditions (6000 psig and 300°F) and maintains quite high temperature till it reaches the separator (100°F). This happens due to the high velocity the fluid has, compared to

the worst case (Green), which does not allow heat exchange to the environment to be high. Also from the diagram can be seen that the finishing point (separator) has lower pressure value (150 psig) than the worst case (500 psig). The reason a sharp drop only in pressure is observed is because the surface pipeline, in contrast with the downhole one, has very good insulating material. This coupled with high fluid velocity prevents temperature drop. The pipeline operates far outside the hydrate stability region and does not require the help of thermodynamic inhibitors to produce fluids without problems.

The worst case scenario pipeline ([Figure 129 – Green](#)) operates from reservoir conditions (6000 psig and 300°F) and its temperature drops steadily till it reaches the environment temperature. Then the temperature remains constant at 35°F and only pressure drops until the separator pressure (500 psig) is reached (no insulating material). The fluid enters the separator at environmental temperature. In the offshore region the environment may not have extremely low temperature although hydrate problems still persist. Longer tie backs and high angle risers imply lower rates due to the increased pressure differential required to produce fluids. The oil with not favorable system configuration has a very low solution production rate and moves with very low velocity. The time required to travel from the reservoir to the separator is high and without insulating material to obstruct heat losses the fluid temperature reaches the environmental one. Also the high pressures needed facilitate the hydrate formation. In the specific case hydrate problems are present because the pipeline operates inside the hydrate stability region after a certain point. The use of THIs in this case yields results when over 20% (v/v of water) is used of either Methanol or MEG. As seen from the diagram with lower concentrations the dissociation curve of the hydrate has not moved enough for the pipeline to be operating under safe conditions.

7. CONCLUSIONS – PROPOSALS

After observing the parameters that affect hydrate formation in an offshore and an onshore arctic region the conclusions made can be divided into three distinct categories:

- 1) **System Parameters**
- 2) **Regional – Environmental parameters**
- 3) **Parameters that can alter the hydrate stability curve.**

In the first category the system parameters discussed in the sensitivity analysis section 6.1 are presented. These are parameters of the well, pipeline and surface facilities (VLP) that can be designed and modified if necessary to best achieve the project goals. The most important are:

- **Effect of Tubing Diameter**

Larger tubing diameter has the potential to facilitate operating conditions outside the hydrate phase envelope but is case specific. For gases the size acts beneficially by leading to higher rates and velocities (less time available for heat transfer to the environment). The contrary can be true for low energy reservoirs, where slightly higher rate can be achieved with smaller velocity in the larger tubing.

- **Effect of Separator Pressure**

Decreased separator pressure has a beneficial effect in the production flow rate and subsequently in the flowing velocity. Larger pressure differential increases the flow rate of a production system. In high energy reservoirs (gas) it can yield no important effect as the total pressure needed for the fluid transfer is low. The fluid travels at high speed and the friction losses along the pipeline mask the effect of the increased pressure differential. In low energy systems (oil) the difference in the flow rate is substantial and the hydrate risk minimized.

- **Effect of the Insulation Material of the Surface Pipeline**

Insulation of the surface pipeline could be the most important factor affecting hydrate formation regardless the system configuration. It delays the heat transfer from the fluid to the environment, provides longer “no-touch” periods during shut-ins and during steady

state conditions allows the fluid to travel in the pipeline without reaching the hydrate formation conditions. The longer the tie-back the better results an insulated pipeline can exhibit. From the cases tested it is shown that the well is able to produce fluids without encountering problems only by using an advanced insulation material; no other system parameter needs to be modified. The disadvantages are the high cost per kilometer and the integrity issues that could occur.

- **Effect of the Annular Space Brine**

In arctic regions it is possible to utilize a fluid of high heat capacity that acts as insulation for the downhole pipeline. In arctic regions the geothermal gradient of sedimentary rocks takes very low values up to 1500ft from the surface. To avoid heat losses to the surrounding rock such solution can be implemented.

- **Effect of the Water Rate**

Water affects the hydrate formation in three ways. First of all, water presence is one necessary parameter for hydrate formation. Secondly, increased water percentage increases the gravitational losses of the pipeline. Pipelines and risers of high elevation filled with water need high pressure differentials to operate and even then the production rate is minimal. Low rate imply low velocity and the hydrate formation can be a real issue. Finally high water rates usually is coupled with reservoir depletion, so the pressure differential of the system gets lower than the desired one adding more difficulty in operating outside the hydrate envelope. The only way to deal with this condition is to remove the water before it enters the surface pipeline by the use of subsea separators or by fluid injection inside the reservoir. In the cases studied water concentration is the most important factor for hydrate formation both in retrograde gas condensate and oil systems. High water percentages lead to severely reduced flow rates and in some cases no flow at all. The risk is significant for the regions tested when the water cut is more than 20%

- **Effect of the Surface Pipeline Length**

Longer surface pipeline lengths are usually observed in offshore systems where many wells deliver fluids to the same platform with long tie-backs. In these cases the problem is double. The first is that increased length means increased travelling time for the fluid. Hydrate formation risk is important due to the increased heat transfer to the environment. The second is the increased frictional pressure losses (decreased flow rate) which are proportional to the length of the traveling medium. This is a condition in offshore systems not easily avoided.

- **Effect of the Total Height of Surface Pipeline**

Increased elevation of the surface pipeline increases the gravitational losses due to the increased hydrostatic head. This again leads to lower production rates and lower fluid velocity. This condition as previously cannot be avoided.

- **Reservoir Properties**

Skin factor as all other reservoir properties (porosity, permeability, wellbore radius and reservoir thickness) affect production rate by the same way they affect reservoir pressure in relation to flow rate. Increasing these parameters (except skin factor which acts the other way) the IPR slope decreases and the solution rate of the VLP and IPR is at a higher rate. This subsequently means higher velocity and less time available for heat transfers and hydrate formation potential. These parameters however are designed with optimal well design in mind and not production flow assurance issues.

- **Effect of Active Heating methods**

Active pipeline heating can be applied by circulating a fluid of high heat capacity from the surface facilities to the wellbore, surrounding the pipeline and maintaining its temperature at high levels. The other possible way of active heating is by introducing electrical current to resistances surrounding the pipeline. Both ways imply procedures quite difficult to implement and high cost with potential failures. If successful the hydrate problem is eliminated but the costs are substantial and can render the project sub-economic.

- **Effect of Thermodynamic Inhibitors**

THIs are the most commonly used method for hydrate problem remediation. They are chemical compounds mixed with the fluid stream that alter the hydrate stability zone and allow operations in previously hydrate stable conditions. They can be used regardless location or characteristics of the well and do not need sophisticated system design. They can although exhibit high costs if large quantities are needed (when high water percentages are present) and injection pressures may be not enough to maintain the these percentages. This solution is most commonly used in fields nowadays but new methods tend to render it obsolete.

- **Effect of Low Dosage Inhibitors: (Kinetic, AA)**

LDHIs are the future of the hydrate inhibitors although the way they act is case specific and

no simulation modeling is yet available in commercial scale. The benefit of these inhibitors is that they act at low percentages and do not prevent hydrate formation. Instead they obstruct plugging and allow hydrate particles to be produced. In many systems where large THI quantities are not an option and other solutions render the project uneconomical they can be the key.

In the second category the regional (case specific) parameters are discussed. These are parameters of the sediments, the drilled wellbore, reservoir properties (IPR) and environmental conditions that usually cannot be altered or they may but at extreme costs. The most important are:

- **Effect of Ambient Temperature after the Wellhead**

Environmental temperature around the surface pipeline is a critical parameter for hydrate formation together with the total length of the pipeline. The heat transfer rate is determined by the total difference of the fluid and the environmental temperature, along with the system's overall heat transfer coefficient. Generally lower temperature is worse for hydrate formation but there are cases where pipeline length (time exposed) matters more). This can be seen from cases in Chapter 4 where the hydrate formation risk, especially in the oil systems, is increased in the offshore environment compared to the arctic one.

- **Effect of Geothermal Gradient**

Geothermal gradient is the temperature profile of the sedimentary rocks. In arctic regions it can pose problems because hydrate formation temperatures can be encountered even from 1500ft depth.

- **Effect of Reservoir Depth – Wellbore Length**

Reservoir depth affects the reservoir fluid's initial temperature and also leads to an increased well length. The first parameter acts beneficially as the fluid has larger margin for temperature losses during its travel time to the surface. On the other hand longer wells increase both gravitational and frictional pressure losses and result to low rates.

- **Effect of the Fluid Composition**

Fluid composition is the critical parameter that can affect the hydrate phase envelope. From fluid composition the hydrate risk can be assessed by recognizing if hydrate formers exist in the stream and their respective percentages. Fluid composition is the cornerstone of the

system design when its target is set at tackling the hydrate formation problem. It is the critical parameter that determines the usage of THI, LDHI or even system hardware such as insulation and change of pipeline diameter. The properties and the modeling start with the fluid composition, because the hydrate forming conditions are always case specific for a reservoir fluid.

- **Effect of Hetero-Compounds**

Hetero compounds existing in the fluid stream can act either to our advantage or against regarding hydrate formation. H_2S in sour gases or oils shifts the hydrate phase envelope to the right and renders the hydrate formation easier. This way is more probable to operate inside the envelope and this multiplies the risk. The opposite is true with high percentages of CO_2 in the fluid stream. Either way the exact composition of the hetero compounds must be calculated so a reliable model can be developed that describes accurately the true fluid stream.

- **Effect of Salinity**

The effect of salinity on thermodynamic equilibrium condition for hydrate formation is similar to that of the THIs. It was observed that there was a slight reduction in hydrate equilibrium temperature at all pressures between fresh water (0% salts) and 20,000 ppm of salts. The effect of salts in the fluid stream and in hydrate forming conditions should be kept in mind for cases with increased concentrations.

Comparison of Hydrate Formation Risk (Oil - Gas Condensate)

In literature gas streams are considered more dangerous regarding the hydrate formation problem. The increased concentration of hydrate forming components is the reason that the possibility of hydrate particle formation is increased. Also high pressures existing in gas production systems facilitate the fluids to pass through hydrate stability conditions. However high pressure differentials result in high travelling speeds and less time spent inside the flowlines. This can be beneficial in two ways. The first is that the fluid has less time to transfer heat to the environment and is possible that it will never reach the required temperature of the hydrate formation. The second is that even if hydrate stability conditions are reached, the particles do not have the time to agglomerate and form plugs. This way the hydrate particles will be produced and dissociated at the surface facilities. The only way for a gas condensate flowline to plug is the production under low gas rates (travelling velocity) and the presence of extended risers with locations that facilitate liquid loading. High gas

rates are the obvious solution, however other problems appear with high gas rates, such as high erosional velocities that can destroy the material of the pipeline. Fortunately such rates are very high and the majority of production systems never reach these values.

On the other hand oil systems are generally more “heavy” due to the increased density of the fluids. High densities increase dramatically the hydrostatic column of the fluid that needs to be produced and require increased pressure differential to obtain economical flow rate. Even if high values of pressure differential are viable, the achieved production rate usually results in fluids that travel significantly slower than gas systems. The heavier an oil is the slower its flowing rate is and this renders critical the liquid loading phenomenon. Oil may not contain as much hydrate formers in its composition as a gas stream, however the increased time it spends inside the flowline increases dramatically the risk of hydrate formation and agglomeration. This coupled with many possible areas of accumulation, where liquid loading phenomena are exhibited, leads to risky situations.

Generally oil systems exhibit increased risk compared with gas ones, despite the popular belief of the contrary. However remediation is easier with the use of traditional hydrate inhibitors. The lower pressures existing in the system facilitate the injection, which needs less powerful pumps to provide adequate rates.

Proper design of the production system, regardless being oil or gas, can eliminate the risk of hydrate formation in the majority of developments without the need of inhibitor injection.

Comparison of Hydrate Formation Risk (Arctic Onshore – Offshore)

As the demand for energy resources is rising, the industry has to develop fields in harsh regions to achieve the production goals. Arctic and deep offshore fields are being developed and new challenges appear during production. Hydrocarbon flow in such regions deal with many flow assurance issues. Most common are the hydrates and wax problems occurring in the flowline.

Arctic regions exhibit lower environmental temperatures, most time of the year, compared to offshore regions. Pipelines with large lengths are very difficult to operate in these regions without serious investments into high end insulation and active heating systems. In most cases a simple injection of chemicals will not work, as the conditions are extreme and even a shift of the hydrate formation envelope is not enough for the pipeline to

operate without problems. Cold flow is a possible solution in arctic regions but must be coupled with extended tests that prove its viability for each specific case. Blow down and pigging are considered passive solutions that are used after hydrate particle formation has occurred. They can yield positive results regarding the hydrate plug avoidance but can lead to safety concerns, such as integrity issues and personnel injuries. Also they stop production and this can prove to be sub-economical.

In offshore developments the temperature is steady around the year at 35°F approximately. However this temperature coupled with high pressures, existing inside the pipeline, results in production passing through the hydrate stability region. Hydrate problem is more severe in offshore systems because long tie-backs increase the travel time for the fluid. This way a significant amount of heat is transferred to the environment and the temperature could equalize to the environmental inside the pipeline.

Hydrate risk is evident in those cases and remedial action must be taken. Insulation is extremely expensive in those systems as the installation is technically difficult and the pipelines very long. Also it has integrity issues, which are very difficult to repair underwater if they occur. Active heating is used in very few cases because most of the time is sub-economical and poses threat in terms of integrity. Other remediation techniques, such as pigging and blow down are difficult to implement as double flowlines are required and highly trained personnel needed. Hydrate inhibitor injection is a typical solution, implemented in the majority of cases, but it can be proven insufficient to achieve the high pressure values and increased injection rates required when high water cut is present. Low dosage inhibitor usage is increasing exponentially as it does not pose many problems and can be a reliable solution if enough testing is done for a specific case. Last the use of subsea separation can eliminate many problems such as high water cut that increases pressure losses and THI required quantity. It is possible to even achieve flow free of water, with minimal risk for hydrate formation. The technology is on the rise but a solution regarding treatment of the disposed water is yet to be found.

Further Research Proposals

The hydrate formation problem is a well discussed issue and a lot of research has been focused in this area. The mechanisms acting are well known and development is already in place in many fields around the globe. However there are still aspects of the matter that demand a closer look and further discussion. More specifically:

- **Transient phenomena modeling and design:**

During common well procedures, as shut-ins and start-ups with various production stoppage times, the danger of hydrate formation is more severe compared to production under steady state conditions. Popular commercial software does not address these issues yet nor does it reliably, as time does not enter into the calculations directly. However theoretical models exist and system modeling during transient phenomena will be more accessible and reliable in the future.

- **Cold flow production modeling and implementation:**

“Cold flow” production is a prominent design that allows producing the hydrate particles with some risk. It is not yet considered “mainstream” as many issues need to be resolved. Some of them are that the design is entirely case specific and only general guidelines exist outside the use of “flow loops”. Also the way kinetic inhibitors and anti-agglomerants act must be fully understood as this method takes advantage of their properties. Finally real production data have to be available from developments that implement this method.

- **Subsea separation:**

Subsea separation is a very promising way to get rid of excess water underwater and produce clean hydrocarbon stream with minimal risk of hydrate formation. The remediation needed in this case would be minimal and very economic (use of THIs for example to low water cuts). Many issues need to be resolved however. Some of them are the treatment of disposed water, the successful separation of liquid and gas in a very compact facility and finally the successful separation of gas-oil-water that has yet to be achieved.

Conclusion

Flow assurance plays a critical role both from a technical and economical perspective. Design problems have become greater and the cost of solving them went up. Lower seabed temperatures (1-4°C) and increased depths (1500 – 3000m) made the hydrate problem worse.

Large elevation differences across deep-water risers not only make evacuation of liquids for hydrate management difficult, but also intensify slugging to the point where riser operating ranges have become significantly narrowed. Concerns about the effect of chemical additives on downstream processing are significant. At the same time, insulation costs rise in deep-water, and any form of intervention in the event of a mechanical failure becomes almost prohibitively expensive. Little headway has been made in intervention cost-reduction. Flow assurance still has an important effect on field development design. Alongside drilling costs and reservoir complexity, it is one of the key considerations in making the decision on whether to develop a field. Development costs per barrel are generally lower for subsea developments compared to stand alone hubs, so it is clearly important to maximize tiebacks.

The economic viability of a deep-water subsea tieback is driven by the cost of drilling and pipelines. Pipelines can be at least 25% or more of tieback costs. Adoption of a traditional approach (hydrate avoidance) to flow assurance leads to sub economic development schemes, because pipeline costs are too high. Additional risk is taken on with these designs and must be mitigated. When prospects start off as sub-economic in view of rising oil prices, radical rethinking on flow assurance questions can transform field architecture and hence economics.

The future deep-water multiphase systems will need to work with predictable, but significantly higher levels of risk than implemented today. This means working within the hydrate formation region, or below the wax appearance temperature, or inside the asphaltene deposition region, or near to or within severe slugging conditions. The challenge is therefore how the industry can keep pushing toward rapid and cheap intervention. If low-cost ways to quickly remove plugs in systems are found, the designs will likely consider having a higher risk of plugging.

Sources of risk relate to:

- **Understanding fluid properties and obtaining quality fluid samples:**

It is absolutely essential to retrieve quality fluid samples. Most samples are contaminated with drilling fluids that interfere with accurate assessment of wax deposition and hydrate inhibitor qualification. Improvement in sampling techniques, better downhole equipment and improved laboratory procedures are needed to reduce risk associated with contaminants.

- **Complexity of modeling fluid mechanics and heat transfer:**

Complexity of design in deep-water requires mathematical models that do not exist or require specialized model construction. Complex heated bundles and specialized bundles require computational tools that may not be the best representation of physical systems – leading to potential problems in installed systems. Flow Assurance considerations tend to drive design to complexity. Greater risk tolerance may lead to simpler, cleaner designs that are more predictable and less costly.

- **Understanding and predicting plugging:**

The underlying principle of all subsea designs is driven by the mandate to not plug the system; hence the logical approach is avoidance through design. Improved understanding of mechanisms leading to plugging may unlock new possibilities in cleaner, simpler design that would appear to be taking on greater risk when compared to today's approach. Where plugging is a certainty, rapidly deployed cheap intervention would be a valid alternative to complex and expensive design alternatives.

- **Insufficient field data:**

In the past there has been a concentrated effort to collect multiphase flow, wax deposition, and hydrate inhibitor performance data from field operations. Future deep-water development success depends on the willingness to risk current production to obtain key performance data and operating experience for future production gain. In view of new field initiatives within the industry and a willingness to instrument production systems, opens many new opportunities for understanding and comparing actual operations to the models.

Expansion to harsher environments has promoted research and field testing. Multiphase flow technology had to develop rapidly to support system design in arctic and deep-water

environments. Systems became more complex by utilizing S-shaped risers, free-standing risers and coil tubing gas lift in operations. In the recent years field tests have been conducted and data collected to validate the simulation models. The intention was to increase confidence in addressing multiphase flow issues in ultra-deep-water developments. The technology on flow assurance issues advanced rapidly in the recent decades and one could possibly assert that little more development is needed. In parallel, on the production chemistry front, hydrate research focuses into kinetic modeling, potential cold flow systems, and usage of anti-agglomerant technology in the field. Advancements have been made in the development of kinetic inhibitor technology, but subcooling requirements of deep-water developments still cannot be met. Anti-agglomerate development is currently being implemented in black oil systems, and is often included in emerging hydrate management strategies.

Subsea processing has yet to make an important impact as a solution to flow assurance challenges. These systems were to be developed to increase the flow assurance engineer's options. Still the most challenging operation is water-oil separation at the deep-water wellhead, as a design alternative to reduce thermodynamic inhibitor usage and reduce hydrostatic columns in the flowline and riser. There is still much to be learned about wax deposition and additional flow assurance concerns may be faced should future subsea systems be built with less thermal protection in the face of potential cold flow systems.

Active heating of surface pipelines is starting to make headway as a reliable system of controlling the fluid temperature along the flowline, avoiding flow assurance issues with costly remediation. However issues must be addressed regarding pipeline integrity, reliable transfer of electrical energy to great lengths and high pressure pumps for hot fluid circulation.

Each of these challenges involves a combination of processes, instrumentation, control, and electrical technologies. Ultimately successful development of seabed processing capability will unlock currently uneconomic reserves and maximize the value of existing infrastructure.

Nevertheless, many deep-water production systems are operating around the world and providing valuable field data. The central challenge in deep-water flow assurance is how the industry can more effectively use that data collectively to keep improving the system design. Building confidence is important to being able to assess risk associated with reduced-cost systems.

APPENDIX – CORRELATIONS USED BY PROSPER

(1) Peng – Robinson Equation of State

$$p = \frac{RT}{V_m - b} - \frac{a \alpha}{V_m^2 + 2bV_m - b^2}$$

$$a = \frac{0.457235 R^2 T_c^2}{p_c}$$

$$b = \frac{0.077796 R T_c}{p_c}$$

$$\alpha = \left(1 + \kappa \left(1 - T_r^{0.5}\right)\right)^2$$

$$\kappa = 0.37464 + 1.54226\omega - 0.26992\omega^2$$

$$T_r = \frac{T}{T_c}$$

In polynomial form:

$$A = \frac{a \alpha p}{R^2 T^2}$$

$$B = \frac{bp}{RT}$$

$$Z^3 - (1 - B) Z^2 + (A - 2B - 3B^2) Z - (AB - B^2 - B^3) = 0 ,$$

where ω is the acentric factor of the species, R is the universal gas constant and $Z=PV/(RT)$ is compressibility factor. The Peng–Robinson equation was developed in 1976 at The University of Alberta in order to satisfy the following goals:

- The parameters should be expressible in terms of the critical properties and the acentric factor.
- The model should provide reasonable accuracy near the critical point, particularly for calculations of the compressibility factor and liquid density.
- The mixing rules should not employ more than a single binary interaction parameter, which should be independent of temperature pressure and composition.
- The equation should be applicable to all calculations of all fluid properties in natural gas processes.

For the most part the Peng–Robinson equation exhibits performance similar to the Soave - Redlich equation, although it is generally superior in predicting the liquid densities of many materials, especially nonpolar ones.

(2) Lohrenz, Bray & Clark Viscosity Correlation

The most commonly used oil viscosity correlations are those of Beggs-Robinson and Vasquez-Beggs. Corrections must be applied for under-saturated systems and for systems where dissolved gas is present in the oil. However, in compositional simulation, where both gas and condensate compositions are known at every point of the reservoir, it is customary to calculate condensate viscosity using Lohrenz, Bray & Clark correlation. This serves the purpose of guaranteeing that the gas phase and condensate phase converge to the same value of viscosity as they approach near-critical conditions.

Lohrenz, Bray and Clark (1964) proposed an empirical correlation for the prediction of the viscosity of a liquid hydrocarbon mixture from its composition. Such expression, originally proposed by Jossi, Stiel and Thodos (1962) for the prediction of the viscosity of dense-gas mixtures, is given below:

$$\mu = \mu^* + \frac{1}{\epsilon_m} \left[(0.1023 + 0.023364p_r + 0.058533p_r^2 - 0.040758p_r^3 + 0.0093724p_r^4)^4 - 1 \cdot 10^{-4} \right]$$

where:

μ = fluid viscosity (cp),

μ^* = viscosity at atmospheric pressure (cp),

ξ_m = mixture viscosity parameter (cp-1),

ρ_r = reduced liquid density (unitless),

(3) CPA (Cubic-Plus-Association) EoS

CPA (Cubic-Plus-Association) is an equation of state, developed by Kontogeorgis et al 1999, which is based on a combination of the Soave–Redlich–Kwong (SRK) for describing the physical interactions with the Wertheim’s first-order perturbation theory, which can be applied to different types of hydrogen-bonding compounds.

The development of CPA started in 1995 as a research project funded by Shell (Amsterdam), and the model was first published in 1996. Since then, it has been successfully applied to a variety of complex phase equilibria, including mixtures containing alcohols, glycols, organic acids, water, and hydrocarbons. Focus has been placed on cases of industrial importance, e.g., systems with gas-hydrate inhibitors (methanol, glycols), glycol regeneration and gas dehydration units, oxygenate additives in gasoline, alcohol separation, etc.

In summary, the statistical thermodynamics model uses the CPA-EoS and classical mixing rules for fugacity calculations in all fluid phases. The CPA-EoS in terms of pressure P is given by Kontogeorgis et al, 1995.

$$P = \frac{RT}{v-b} - \frac{a}{v(v+b)} - \frac{1}{2} \frac{RT}{v} \left(1 + \rho \frac{\partial \ln(g)}{\partial \rho} \right) \sum_i x_i \sum_{A_i} (1 - X^{A_i})$$

a : the energy parameter

b : the co-volume parameter (assumed to be temperature independent, in agreement with most published equations of state)

ρ : the molar density of the fluid

g : the simplified expression of the radial distribution function as suggested by Kontogeorgis et al.

X^{A_i} : the mole fraction of pure component i not bonded at site A

x_i : is the mole fraction of component i .

(4) Joule – Thomson Effect

In thermodynamics, the Joule–Thomson effect describes the temperature change of a gas or liquid when it is forced through a valve or porous plug while kept insulated so that no heat is exchanged with the environment. The effect is named after James Prescott Joule and who discovered it in 1852.

The adiabatic (no heat exchanged) expansion of a gas may be carried out in a number of ways. The change in temperature experienced by the gas during expansion depends not only on the initial and final pressure, but also on the manner in which the expansion is carried out.

If the expansion process is reversible, meaning that the gas is in thermodynamic equilibrium at all times, it is called an isentropic expansion. In this scenario, the gas does positive work during the expansion, and its temperature decreases.

In a free expansion, on the other hand, the gas does no work and absorbs no heat, so the internal energy is conserved. Expanded in this manner, the temperature of an ideal gas would remain constant, but the temperature of a real gas may either increase or decrease, depending on the initial temperature and pressure.

The method of expansion discussed, in which a gas or liquid at pressure P_1 flows into a region of lower pressure P_2 via a valve or porous plug under steady state conditions and without change in kinetic energy, is called the Joule–Thomson process. During this process, enthalpy remains unchanged.

If the pressure starts out high enough, the temperature increases as the pressure drops, until an inversion temperature is reached and a phase transition occurs, called the inversion point. After this, as the fluid continues its expansion, the temperature begins immediately to drop.

The rate of change of temperature T with respect to pressure P in a Joule–Thomson process (that is, at constant enthalpy H) is the Joule–Thomson (Kelvin) coefficient μ_{JT} . This coefficient can be expressed in terms of the gas's volume V , its heat capacity at constant pressure C_P , and its coefficient of thermal expansion α as:

$$\mu_{JT} = \left(\frac{\partial T}{\partial P} \right)_H = \frac{V}{C_P} (\alpha T - 1)$$

The value of μ_{JT} is typically expressed in °C/bar (SI units: K/Pa) and depends on the type of gas and on the temperature and pressure of the gas before expansion. Its pressure dependence is usually only a few percent for pressures up to 100 bar.

All real gases have an inversion point at which the value of μ_{JT} changes sign. The temperature of this point, the Joule–Thomson inversion temperature, depends on the pressure of the gas before expansion.

(5) Beggs and Brill Correlation for Surface Equipment Pressure Losses

Beggs and Brill (1973) correlation, is one of the few correlations capable of handling all flow directions encountered in oil and gas operations, namely uphill, downhill, horizontal, inclined and vertical flow for two phase fluid.

Total pressure gradient is described by following relation.

$$dP/dZ = [(dP/dZ)_{\text{Fric.}} + (dP/dZ)_{\text{Ele.}}]/(1-E_k)$$

where, $(dP/dZ)_{\text{Fric.}}$ is pressure gradient due to friction, $(dP/dZ)_{\text{Ele.}}$ is hydrostatic pressure difference and E_k estimates pressure loss due to acceleration.

Flow Pattern Map

A flow regime is identified based on the Froude number of the mixture (Fr_m) and input liquid content (no slip liquid holdup C_L).

$$Fr_m = v_m^2 / g \cdot D$$

where, v_m is mixture velocity, D is pipe inside diameter and g is gravitational constant.

$$C_L = Q_L / (Q_L + Q_G)$$

where, Q_L is liquid volumetric flow and Q_G is gas volumetric flow.

The transition lines for correlation are defined as follows:

$$L_1 = 316 C_L^{0.302}$$

$$L_2 = 0.0009252 C_L^{-2.4684}$$

$$L_3 = 0.1 C_L^{-1.4516}$$

$$L_4 = 0.5 C_L^{-6.738}$$

Segregated Flow

$$C_L < 0.01 \text{ and } Fr_m < L_1$$

$$\text{OR } C_L \geq 0.01 \text{ and } Fr_m < L_2$$

Intermittent Flow

$$0.01 \leq C_L < 0.4 \text{ and } L_3 < Fr_m \leq L_1$$

$$\text{OR } C_L \geq 0.4 \text{ and } L_3 < Fr_m \leq L_4$$

Distributed Flow

$$C_L < 0.4 \text{ and } Fr_m \geq L_4$$

$$\text{OR } C_L \geq 0.4 \text{ and } Fr_m > L_4$$

Transition Flow

$$L_2 < Fr_m < L_3$$

Liquid Holdup, $E_L(\theta)$

Once flow type has been determined, liquid holdup for horizontal flow $E_L(o)$ is calculated.

$$E_L(o) = a C_L^b / Fr_m^c$$

Flow Regime	a	b	c
Segregated	0.98	0.4846	0.0868
Intermittent	0.845	0.5351	0.0173
Distributed	1.065	0.5824	0.0609

$E_L(o)$ must be greater than C_L , if $E_L(o)$ is smaller than C_L , then $E_L(o)$ is assigned a value of C_L . Actual liquid volume fraction is obtained by multiplying $E_L(o)$ by a correction factor, $B(\theta)$.

$$E_L(\theta) = B(\theta) \times E_L(o)$$

$B(\theta)$ is obtained as -

$$B(\theta) = 1 + \beta(\sin(1.8\theta) - (1/3)\sin^3(1.8\theta))$$

where θ is the angle of inclination of pipe with horizontal.

Correction factor β is calculated as following -

$$\beta = (1 - C_L) \ln(d.C_L^e.N_{LV}^f.Fr_m^g)$$

Uphill	d	e	f	g
Segregated	0.011	-3.768	3.539	-1.614
Intermittent	2.96	0.305	-0.4473	0.0978
Distributed	$\beta = 0$			
Downhill	d	e	f	g
All	4.7	- 0.3692	0.1244	- 0.5056

Liquid velocity number, N_{LV} is given by:

$$N_{LV} = 1.938 V_{sl}(\rho_L / (g \cdot \sigma))^{1/4}$$

V_{sl} is no slip liquid velocity, ρ_L is liquid density, g is gravitational constant and σ is surface tension.

For transition flow,

$$E_L(\theta)_{\text{transition}} = A E_L(\theta)_{\text{segregated}} + B E_L(\theta)_{\text{intermittent}}$$

where A and B are as following -

$$A = (L_3 - Fr_m) / (L_3 - L_2)$$

$$B = 1 - A$$

Liquid holdup, $E_L(\theta)$ is used to calculate mixture density ρ_m .

$$\rho_m = \rho_L \cdot E_L(\theta) + \rho_G \cdot (1 - E_L(\theta))$$

$$(dP/dZ)_{\text{Elevation}}$$

Pressure change due to the hydrostatic head of the vertical component of the pipe is given by:

$$(dP/dZ)_{Ele.} = \rho_m \cdot g \cdot \sin(\theta) / (144 \cdot g_c)$$

$$(dP/dZ)_{Friction}$$

Calculate no slip Reynold's number using no slip mixture density and viscosity.

$$Re_{NS} = \rho_{NS} \cdot V_m \cdot D / \mu_{NS}$$

No slip friction factor, f_{NS} is then calculated using Colebrook-White equation.

Ratio of friction factor is defined as

$$f_{TP} / f_{NS} = e^S$$

Value of S is governed by following conditions -

$$S = \ln(2.2y - 1.2)$$

if $1 < y < 1.2$, otherwise -

$$S = \ln(y) / (-0.0523 + 3.182 \cdot \ln(y) - 0.8725 \cdot (\ln(y))^2 + 0.01853 \cdot (\ln(y))^4)$$

where y is defined as

$$y = C_L / E_L(\theta)^2$$

Pressure loss due to friction is:

$$(dP/dZ)_{Fric.} = 2 \cdot f_{TP} \cdot V_m^2 \cdot \rho_{NS} / (144 \cdot g_c \cdot D)$$

Pressure loss due to acceleration, factor E_k is given by:

$$E_k = \rho_m \cdot V_m \cdot V_{sg} / (g_c \cdot P)$$

where, V_{sg} is no slip gas velocity and P is gas pressure.

(6) Petroleum Experts 2

The Petroleum Experts 2 correlation includes the features of the original Petroleum Experts correlation with more emphasis on predicting low-rate VLPs and well stability.

The Petroleum Experts correlation combines the best features of existing correlations for VLP calculations. It uses the Hagedorn Brown correlation in slug flow and Dun and Ross for mist flow. In the transition regime a combination of these is utilised.

The Hagedorn Brown correlation performs well for slug flow at moderate to high production rates but well loading is poorly predicted. It should not be used for condensates and whenever mist flow is the prevailing flow regime. Hagedorn Brown underpredicts VLP at low rates and should not be used for calculating stable low rates.

Duns and Ros defined a flow map together with a series of correlations for calculating the boundaries between the flow regimes as well as the slip velocity (V_s).

The Friction factor is calculated from the liquid Reynolds Number when flow is in the Bubble or Slug regions; while the gas Reynolds number is used in the Mist region. Finally, calculation of the pressure drop is completed by adding an acceleration term for flow in the Mist region only.

(7) Petroleum Experts 4

Petroleum Experts 4 is an advanced mechanistic model for any angled wells, suitable for any fluid. It uses features of all the correlations described before plus original work from **PetEx** Company.

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