

PSIG 2317

## Incorporating Condensation and Evaporation into the Simulation of Multiphase Pipelines

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This paper was prepared for presentation at the PSIG Annual Meeting held in San Antonio, Texas, 16 May – 19 May 2023.

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### ABSTRACT

During the operation of multiphase and single-phase pipelines, gas/liquid phase transitions can significantly influence the efficiency of the pipeline. Gas/liquid phase changes occur when pipes are exposed to particularly high or low temperatures. The energy absorbed or released during the phase change will impact the temperature and pressure of the fluid. Failing to account for that impact can result in inaccurate predictions for temperatures and pressures.

The formation of a 2nd phase in single-phase pipelines can also introduce slugging, interfere with the operation of the compressors/pumps, and damage the equipment that runs the pipeline. Incorporating gas/liquid equilibrium into the modeling of single-phase pipelines allows users to identify conditions that produce significant amounts of condensation/evaporation.

This paper describes a procedure that was developed to simulate the impact that evaporation and condensation have on the behavior of pipelines. It also discusses the results of two case studies that were performed to test the accuracy and efficiency of the new method.

In one of the case studies, the new method was used to simulate the behavior of a commercial pipeline. The data obtained from that simulation was compared to field data that was collected for the pipeline under the same operating conditions.

### INTRODUCTION AND BACKGROUND

Condensation and evaporation are physical processes that can occur in pipelines that have large sections of pipes which are exposed to high or low temperatures. This includes some deep-sea pipelines as well as pipelines that traverse large areas of desert and tundra.<sup>[1],[2]</sup> Another situation that can introduce condensation in gas pipelines are locations in the pipeline where there are sudden, significant drops in pressure (such as valves or fittings). Those conditions can also produce drops in temperature (due to the Joule-Thomson effect) which are large enough to form droplets of liquid.<sup>[3]</sup> In liquid pipelines, the pressure drops and turbulent conditions that occur at valves and fittings can also produce small, temporary gas bubbles due to transient cavitation.<sup>[4]</sup>

For single-phase pipelines, the unexpected formation of a secondary phase will hamper the performance of the pumps/compressors. It can also introduce slugging. Ultimately those effects will lower the efficiency of the pipeline and could even damage the pipes, valves, and equipment that operate the pipeline.<sup>[5]</sup> For single-phase pipelines, reliable predictions of condensation/evaporation can help engineers determine when and where pipes are at risk and determine how to mitigate that risk.

For multi-phase pipelines (which should be designed to handle the problems listed above) condensation and evaporation can still impact how the pressures and temperatures will change across the pipeline.<sup>[5]</sup> Including those effects might be necessary to accurately simulate the behavior of the pipeline.

The next section of this paper describes a procedure that was developed to account for the impacts that condensation and evaporation have on the transportation of multiphase fluids. The method uses thermodynamic equilibrium calculations to determine how the relative amounts of gas and liquid change across the pipeline.<sup>[6]</sup> Adjustments are also made to the fundamental conservation equations to account for the energy absorbed/released during the phase changes.<sup>[7]</sup>

The third section of this paper discusses the results of two case studies that were performed to test the accuracy and usefulness of the new procedure. The first case study considers the transportation of a binary fluid mixture across a small pipeline when it is under conditions that thermodynamically support gas/liquid equilibrium. The purpose of this case study is to verify that changes to the operating conditions of the pipeline will impact the performance of the pipeline in a manner that seems qualitatively correct.

The second case study uses the new method to simulate the transportation of a gas mixture across a commercial pipeline. The results of the simulation are compared to field results and to results that were computed without using the new method. The results highlight the importance of accounting for the impact that condensation has on the temperature.

## THEORETICAL BACKGROUND

Computing the Change in the Relative Amounts Gas and Liquid. At their core condensation and evaporation are thermodynamic processes that determine how the addition/removal of heat alters the compositions and properties of the gas and liquid phases of a multiphase fluid mixture. The extent of those changes is governed by two factors: 1. Gas and liquid phase fugacities (*i.e.*, the compositions of the gas and liquid phases and the relative amount of gas and liquid in the multiphase mixture will adjust to make the gas and liquid fugacities of each component equal). 2. The total amount of each component that is present in the multiphase fluid will not change (*i.e.*, even if the amounts of the component that are present in the gas and liquid phases change, the combined amount of the component in the whole mixture will not change).<sup>[8]</sup>

Based on those requirements the following system of equations can be derived to determine the ratio of the total moles of gas to the total moles of gas and liquid (for a fluid that has reach thermal equilibrium),  $\beta_{g,eq} = \frac{\text{moles gas}}{\text{moles gas} + \text{moles liquid}}$ .<sup>[6],[8]</sup>

$$\begin{cases} Z_i = \beta_{g,eq} Y_{i,eq} + (1 - \beta_{g,eq}) X_i \\ Y_{i,eq} = K_i X_{i,eq} \\ \sum_i Y_{i,eq} = \sum_i X_{i,eq} = 1 \end{cases} \quad (1)$$

where:

- $Y_{i,eq}$  is the mole fraction of the  $i^{\text{th}}$  component of the gas phase mixture once it reaches thermal equilibrium
- $X_{i,eq}$  is the mole fraction of the  $i^{\text{th}}$  component of the liquid phase mixture once it reaches thermal equilibrium

- $Z_i$  is the overall mole fraction of the  $i^{\text{th}}$  component of the multiphase fluid (unlike  $Y_i$  and  $X_i$  the starting value of  $Z_i$  is equal to its equilibrium value)
- and  $K_i$  is the equilibrium constant of the  $i^{\text{th}}$  component of the multiphase fluid

The value of each component's equilibrium constant depends on the temperature and pressure of the mixture. The values of the equilibrium constants are determined using standard equations of state.<sup>[9],[10],[11]</sup> The values of  $Z_i$  for each component are obtained from the starting gas and liquid phase compositions. The remaining properties ( $Y_{i,eq}$ ,  $X_{i,eq}$ , and  $\beta_{g,eq}$ ) are determined by solving the system of equations shown in Eq. 1. In our software, the Rachford-Rice procedure is used to solve Eq. 1.<sup>[6]</sup>

Unfortunately, the values that are computed for  $Y_{i,eq}$ ,  $X_{i,eq}$ , and  $\beta_{g,eq}$  do not directly correspond the gas/liquid compositions and total gas mole fraction that are observed in the pipeline. The values for  $Y_{i,eq}$ ,  $X_{i,eq}$ , and  $\beta_{g,eq}$  that are obtained from Eq. 1 specify the compositions and total gas mole fraction that the multiphase fluid would obtain if it was left undisturbed long enough to reach thermodynamic equilibrium.

In a pipeline, the conditions will change before the equilibrium compositions are reached, causing the multiphase fluid to start moving towards a new set of equilibrium compositions before it reaches the original set of equilibrium compositions. Furthermore, even if the conditions didn't change the fluid cannot instantly transition from its original composition to the equilibrium composition. Its gas and liquid phase compositions (and total gas fraction) will gradually change as it travels across the pipeline.

The following expressions can be used to simulate the gradual changes that occur in  $Y_i$ ,  $X_i$ , and  $\beta_g$  as the multiphase fluid travels across the pipeline<sup>[8]</sup>

$$\begin{cases} Y_i(t_n) = Y_i(t_{n-1}) + C_{vap} \cdot \{Y_{i,eq}(t_n) - Y_i(t_{n-1})\} \cdot \Delta t \\ X_i(t_n) = X_i(t_{n-1}) + C_{vap} \cdot \{X_{i,eq}(t_n) - X_i(t_{n-1})\} \cdot \Delta t \\ \beta_g(t_n) = \beta_g(t_{n-1}) + C_{vap} \cdot \{\beta_{g,eq}(t_n) - \beta_g(t_{n-1})\} \cdot \Delta t \end{cases} \quad (2)$$

where:

- $t_n$  and  $t_{n-1}$  denote two moments in time that are  $\Delta t$  seconds apart, (*i.e.*,  $t_n = t_{n-1} + \Delta t$ )
- $Y_i(t_n)$  and  $Y_i(t_{n-1})$  denote the mole fractions of the  $i^{\text{th}}$  component of the gas phase mixture at times  $t_n$  and  $t_{n-1}$
- $X_i(t_n)$  and  $X_i(t_{n-1})$  denote the mole fractions of the  $i^{\text{th}}$  component of the liquid phase mixture at times  $t_n$  and  $t_{n-1}$

- $\beta_g(t_n)$  and  $\beta_g(t_{n-1})$  denote the total gas mole fraction of the multiphase fluid at times  $t_n$  and  $t_{n-1}$
- $Y_{i,eq}(t_n)$ ,  $X_{i,eq}(t_n)$ , and  $\beta_{g,eq}(t_n)$  denote the values of  $Y_{i,eq}$ ,  $X_{i,eq}$ , and  $\beta_{g,eq}$  at time  $t_n$
- and the coefficient  $C_{vap}$  regulates the rate that the multiphase mixture transitions towards the current equilibrium conditions

The value of  $C_{vap}$  can either be determined using an empirical formula<sup>[12]</sup> or by optimizing it against field readings.

**Simulating Flow Rate and Pressure Changes.** Once a procedure has been developed to determine the change in the liquid and gas compositions and the change in the relative amounts of liquid and gas that are present in the multiphase fluid, it is necessary to determine how those changes will influence the temperatures, pressures, and flow rates across the pipeline. In principle, separate equations of motion should be specified for the gas and liquid phases. Unfortunately, even using the most recent advancements in computer design, a direct solution of the full set of differential equations cannot be applied to real world pipelines. Instead, a single set of conservation equations were developed that regulate the changes to the average density, momentum, and internal energy of the multiphase fluid.<sup>[13]</sup>

$$\left\{ \begin{array}{l} \frac{\partial \rho}{\partial t} + \frac{\partial \rho v}{\partial x} = 0 \\ \rho \cdot \frac{\partial v}{\partial t} + \rho v \cdot \frac{\partial v}{\partial x} = -\frac{\partial P}{\partial x} - \frac{f \rho v^3}{2D} - g \rho \sin \theta \\ \rho C_p \cdot \frac{\partial T}{\partial t} + \rho v C_p \cdot \frac{\partial T}{\partial x} = \frac{f \rho v^3}{2D} - \frac{4h \cdot \{T - T_{amb}\}}{D} + \rho \Delta H_{vap} \end{array} \right. \quad (3)$$

where:

- $x$  and  $t$  specify the position and time of all the properties (density, velocity, etc.) that appear in Eq. 3
- $P$  and  $T$  denote the pressure and temperature of the multiphase fluid
- $\rho$ ,  $C_p$ , and  $v$  denote the average density, average specific heat capacity, and average velocity of the multiphase fluid
- $D$  and  $\theta$  denote the pipe's diameter and angle of elevation
- $T_{amb}$  denotes the ambient temperature of the environment that surrounds the pipe
- $f$  and  $h$  denote the values of the friction factor and heat transfer coefficient,
- and  $\Delta H_{vap}$  denotes the amount of heat released/absorbed per unit mass due to condensation/evaporation

In Eq. 3 most of the impact that evaporation/condensation has on the changes to multiphase fluid's pressure, temperature, and flow rate are indirectly accounted for through the changes that occur to its density and velocity. The one exception is the influence that the phase change has on the temperature. During gas/liquid phase changes, significant amounts of heat are absorbed or released. The value of  $\Delta H_{vap}$  can be determined

using the following expression<sup>[7]</sup>

$$\Delta H_{vap} = H_{int} \left( Y_i(t_n), X_i(t_n), \beta_g(t_n) \right) - H_{int} \left( Y_i(t_{n-1}), X_i(t_{n-1}), \beta_g(t_{n-1}) \right) \quad (4)$$

where  $H_{int} \left( Y_i(t_{n-1}), X_i(t_{n-1}), \beta_g(t_{n-1}) \right)$  and  $H_{int} \left( Y_i(t_n), X_i(t_n), \beta_g(t_n) \right)$  specify the enthalpy per unit mass of the multiphase fluid *before* and *after* its compositions and total gas mole fraction have been adjusted by condensation/evaporation, respectively.

Equations of state are used to compute the specific enthalpies of the multiphase fluid.<sup>[9],[10],[11]</sup> Similarly, well established empirical formulas are employed to determine the other chemical and physical properties ( $\rho$ ,  $C_p$ ,  $h$ , etc.) that appear in Eq. 3.<sup>[14],[15]</sup>

Under most circumstances (including conditions that produce significant amounts of condensation/evaporation), the changes to  $Y_i$ ,  $X_i$ , and  $\beta_g$  are gradual enough that standard methods<sup>[16],[17]</sup> can be used to solve Eq. 3 (based on the previous values of  $Y_i$ ,  $X_i$ , and  $\beta_g$ ). The new pressures, temperatures, and flow rates that were determined from Eq. 3 can then be used in Eqs. 1 and 2 to update  $Y_i$ ,  $X_i$ , and  $\beta_g$ .

Following that procedure, improved predictions can be made for the temperatures, pressures, and flow rates of pipelines that are exposed to long durations of high/low temperatures. This method can also be employed in single-phase pipeline simulations to detect the formation of unexpected secondary phases.

When used in conjunction with methods for computing the liquid holdup, it can also identify conditions that introduce slugging (even slugging that occurs in single-phase pipelines due to the formation of a secondary phase). Unfortunately, the assumption that the changes to  $Y_i$ ,  $X_i$ , and  $\beta_g$  occur gradually does prevent this method from simulating transient cavitation and its effects on the performance of the pipeline.<sup>[18]</sup>

## CASE STUDIES

**Pentane/Octane Transportation Pipeline.** To verify that the new method produces qualitatively correct results, a set of simple simulations were performed which focused on the transportation of a binary fluid mixture of n-pentane and n-octane across a short pipeline at operating conditions that produce gas/liquid equilibrium.

The simulated pipeline was 415 m long and did not include any changes in elevation. The pipes had 0.4064 m outer diameters and were 6.35 mm thick. Standard values of  $5e^{-5}$  m and

2.5 J sec<sup>-1</sup> m<sup>-2</sup> K<sup>-1</sup> were used for the roughness and heat transfer coefficients of the pipes, respectively.

The composition of the mixture was chosen to be 50 % n-pentane and 50 % n-octane (by moles) and preliminary gas/liquid equilibrium calculations were performed to determine the temperature and pressure ranges that generate both gas and liquid phases. From those calculations it was determined that at 36.85 °C a 50/50 liquid mixture of n-pentane/n-octane will form a secondary gas phase once the pressure is less than or equal to 120 kPaa (the exact dew point pressure was not computed).

Based on those calculations three different sets of operating conditions were chosen for the simulation. The first set was selected to ensure that the mixture remains completely liquid across the whole pipeline. The other two operating conditions were chosen to cause the mixture to enter the pipeline with a single liquid phase and exit with both gas and liquid phases. Table 1 lists the three operating conditions that were used for the simulation.

**Table 1 – Operating Conditions (O.C.) for the n-Pentane / n-Octane Pipeline Simulation**

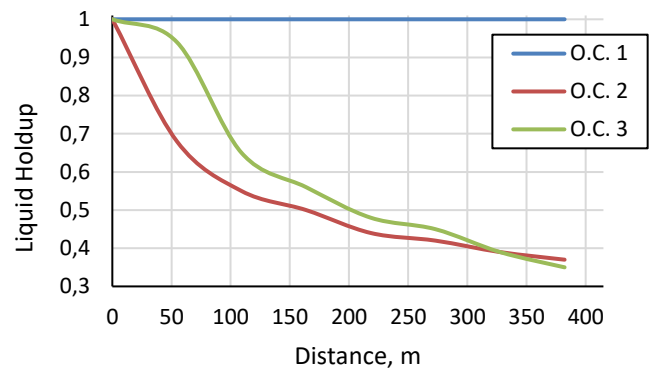
Property	O.C. 1	O.C. 2	O.C. 3
Ambient Temperature, °C	36.85	36.85	36.85
Entering Temperature, °C	36.85	36.85	36.85
Entering Mass Flow Rate, kg/min	4200	4200	5200
Exiting Pressure, kPaa	150	110	110

Figures 1 - 3 show the changes in the liquid holdup, temperature, and pressure that were computed at several points along the pipeline for each of the operating conditions. As was expected, the temperatures and liquid holdups that were computed using the 1<sup>st</sup> set of operating conditions did not show any signs of the formation of a secondary gas phase. At those conditions the liquid holdup remained close to 1 and the temperature only had a slight increase due to friction.

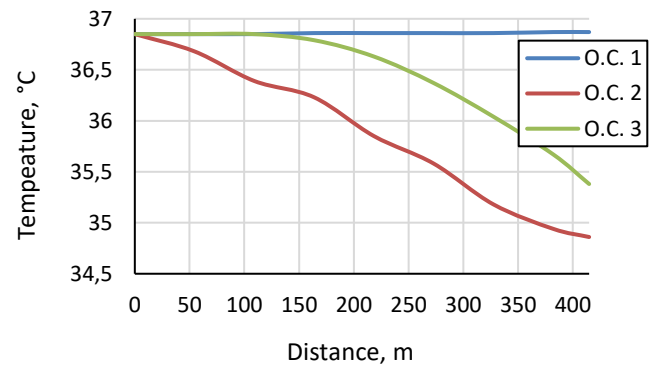
The 2<sup>nd</sup> set of operating conditions was expected to produce an entering pressure that was near the dew point pressure. The pressure would then quickly drop below the dew point pressure allowing a secondary gas phase to form. As expected, the calculated values show a drop in the liquid holdup due to the formation of a gas phase. The calculated temperatures also show a drop in temperature due to the heat that is used to evaporate the liquid. Finally, the change in pressure across the pipeline that was computed for the 2<sup>nd</sup> set of operating conditions was larger than the change in pressure that was

computed for the 1<sup>st</sup> set of operating conditions. This is consistent with the fact that gases and multiphase mixtures typically generate larger frictional pressure drops than liquids.

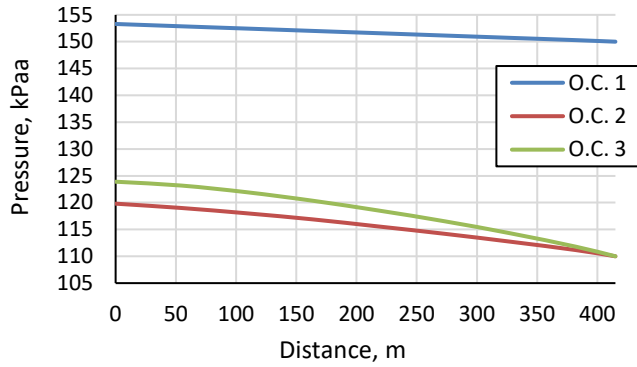
The final set of operating conditions were designed to exit at a pressure that would allow the gas phase to form near the end of the pipeline. However, a larger flow rate was used to produce high pressures near the start of the pipeline which would prevent the formation of a gas phase near the start of the pipeline. Consistent with this prediction, we see that at the start of the pipeline the liquid holdup remains close to 1 and the temperature does not significantly change. After 50 m, the pressures are low enough to form enough gas to impact the liquid holdup. Finally, after 150 m enough evaporation has occurred to lower the temperature of the mixture.



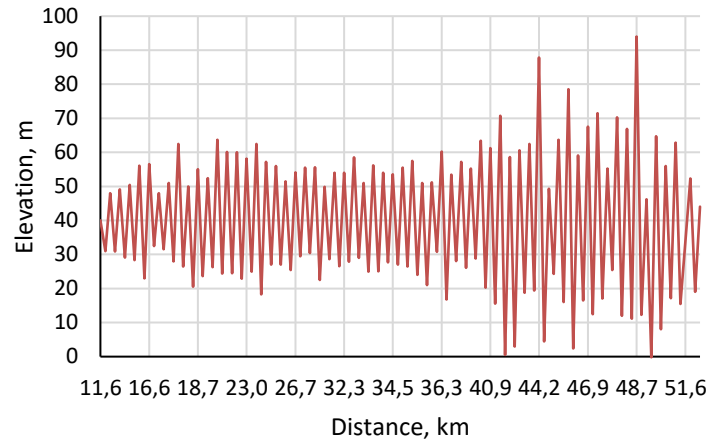
**Figure 1 – Liquid Holdups across the Pipeline for each set of Operating Conditions (O.C.)**



**Figure 2 – Temperatures across the Pipeline for each set of Operating Conditions (O.C.)**

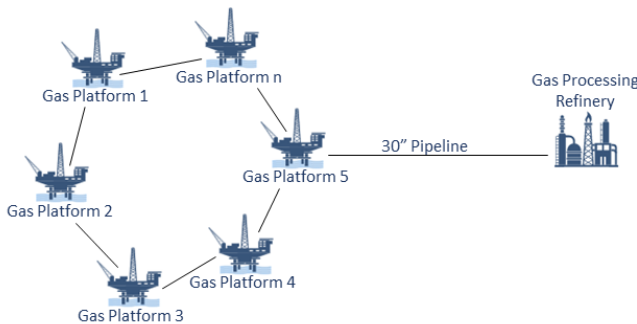


**Figure 3 – Pressures across the Pipeline for each set of Operating Conditions (O.C.)**



**Figure 5 – Elevation vs. Pipeline Length**

Application to a Commercial Pipeline. The proposed modeling and simulation techniques were also applied to a commercial pipeline that was having issues with our previous calculation of condensation. The gas input was not well defined but has gas meters upstream from several producers. In this instance the gas is mixed along the pipeline network from offshore into an onshore facility where heavy hydrocarbons are taken off that are liquid and then piped to a gas processing facility. The pipeline from onshore to the gas processing facility is a 30” pipeline that is 55.6 km. The offshore has anywhere from 10 to 15 platforms that are on based on gas demand and based on contracts for transportation. Figure 4 shows the setup of the system.



**Figure 4 – Field and Pipeline in Discussion**

Figure 5 shows the topography of the pipeline, and it does have some ups and downs to it over its 56 km length. The highest elevation change being about 100 meters over a 1 km distance. Tables 2 and 3 list the gas composition and operating conditions that were used for the simulation.

**Table 2 – Gas Composition for the Pipeline Simulation**

Name	Cp/Cv	Cv (kJ/kg/K)	Mole Fraction
Methane	1.31	1.7	0.963500
Carbon Dioxide	1.32	0.655	0.005560
Ethane	1.19	1.48	0.017400
Propane	1.14	1.48	0.006090
i-Butane	1.1	1.53	0.001530
n-Butane	1.09	1.53	0.001472
i-Pentane	1.08	1.5259	0.000756
n-Pentane	1.08	1.5507	0.000617
Nitrogen	1.4	0.743	0.000709

**Table 3 – Operating Conditions for the Pipeline Simulation**

Elements	Temp	Press	Flow
Eng Unit	°F	PSIG	Kg/sec
Producer	80	750	2.5
Consumer 1	91	633	2.5

At the operating conditions listed in Table 3, the original method predicted that 1,345 barrels of liquid would accumulate within the pipeline. However, pigging measurements that were performed on the pipeline found that zero barrels of liquid were collected. Using the new algorithm and the same set of operating conditions, the simulation correctly determined that zero barrels of liquid condensate should be formed. More specifically, the simulation identified locations in the pipeline that had conditions which supported gas condensation. However, the simulation also determined that the transported gas was not exposed to those conditions long enough to generate measurable amounts of liquid condensate.

The original method that was used to simulate the performance of this pipeline ran gas/liquid equilibrium calculations to determine locations within the pipeline where condensation could occur, but it did not perform any calculations to determine the amount of condensate that formed and propagated across the pipeline. It also did not determine how condensation would influence the temperatures, pressures, and composition of the gas mixture.

However, the original method did use the equilibrium values that were computed for gas/liquid ratios to obtain a rough estimate of the amount of liquid condensate that was formed in the pipeline. That estimate was the amount listed above (1,345 barrels).

Three factors contribute to the original method's inaccurate measurement of the liquid condensate: 1. The method did not account for the increase in temperature that was produced during condensation. Accounting for the increase in temperature would reduce the size of the locations that favored condensation. 2. The equilibrium gas/liquid estimates fail to account for the gradual build up in the size of the liquid condensate that occurs in the pipeline. 3. The original method is not capable of determining how much of the liquid condensate would reevaporate into gas at a later point within the pipeline.

## RESULTS

When running the new algorithm with the composition and operating conditions that are listed in tables 2 and 3 the

simulation calculates zero barrels of liquid condensate in the pipeline, matching what was expected with the previous pigging operations.

## CONCLUSIONS

The results that were determined for the two case studies show that: 1. accurate gas/liquid equilibrium calculations are needed to simulate the influence that condensation/evaporation have on the behavior of the pipeline; 2. Additional calculations must be performed to determine how gas/liquid equilibrium properties correlate with the fluid's properties inside the pipeline.

The second case study also shows that this new algorithm produces a more accurate prediction of the liquid drop off due to condensation and helps reduce the error in calculation of our larger pipelines.

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## FIGURES

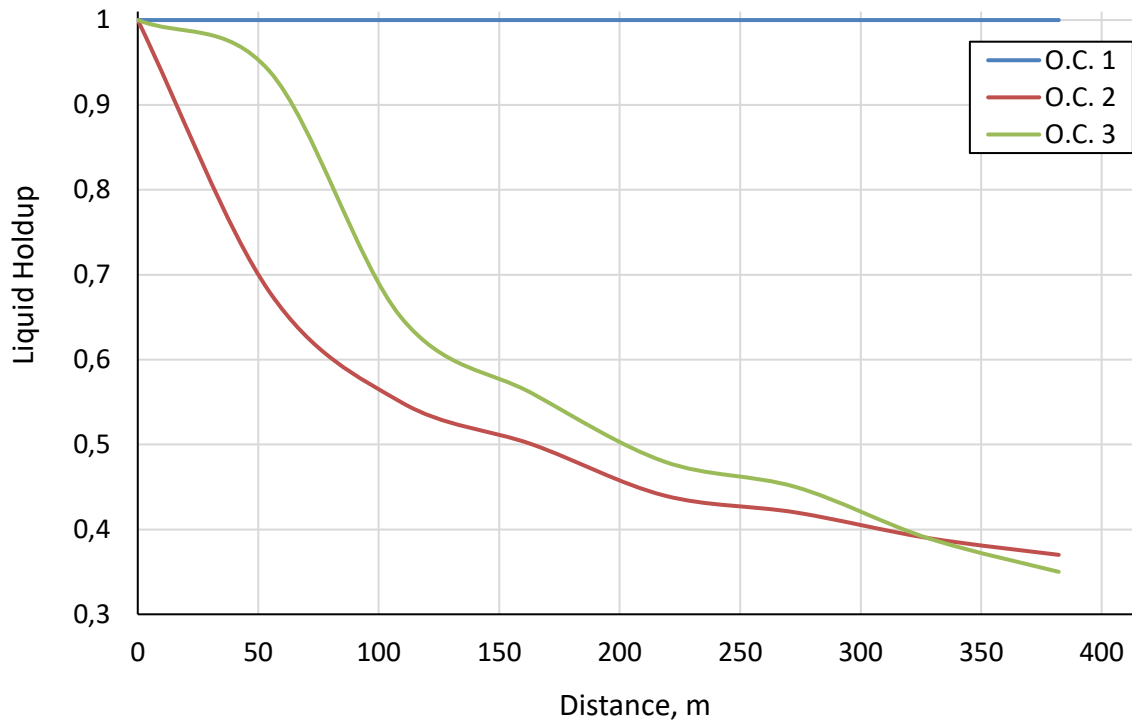


Figure 1 – Liquid Holdups across the Pipeline for each set of Operating Conditions (O.C.)

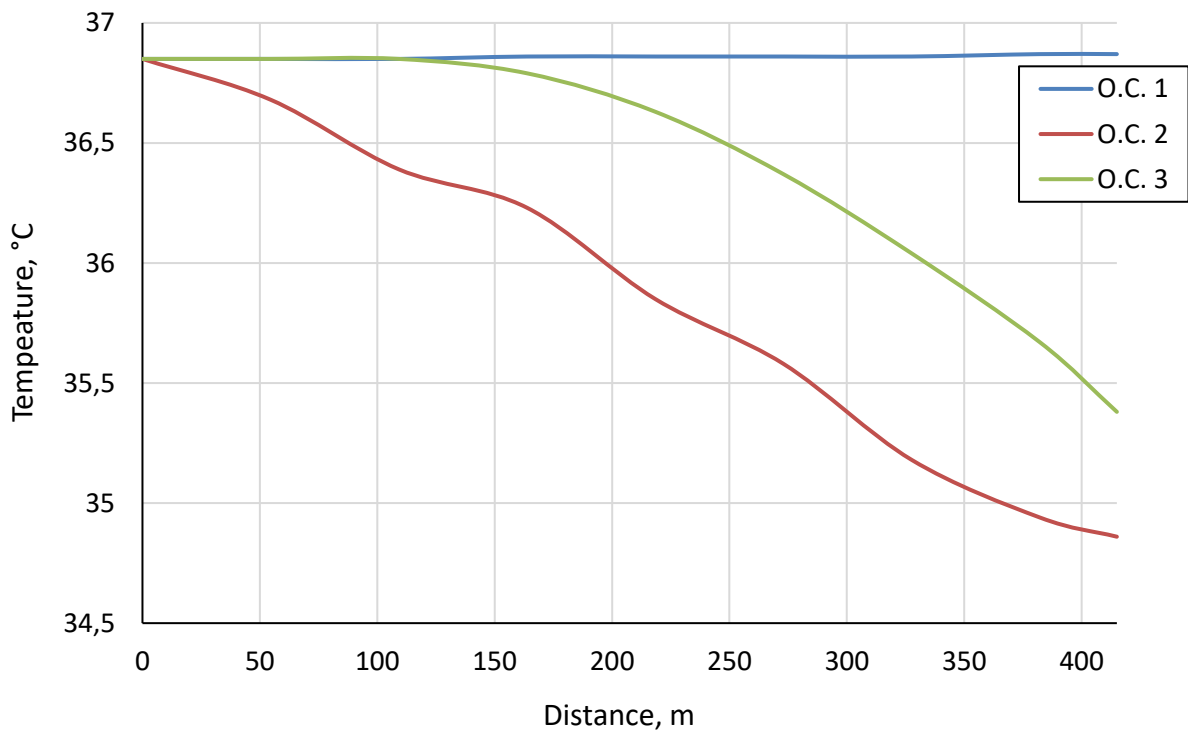


Figure 2 – Temperatures across the Pipeline for each set of Operating Conditions (O.C.)



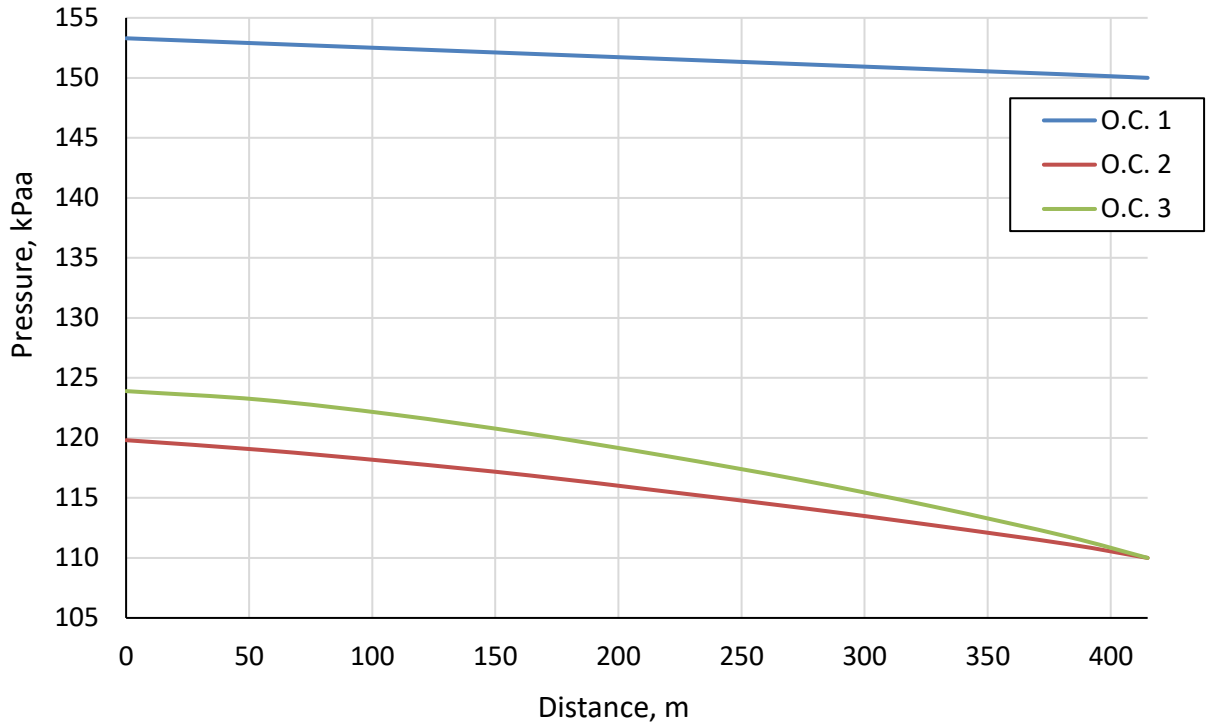


Figure 3 – Pressures across the Pipeline for each set of Operating Conditions (O.C.)

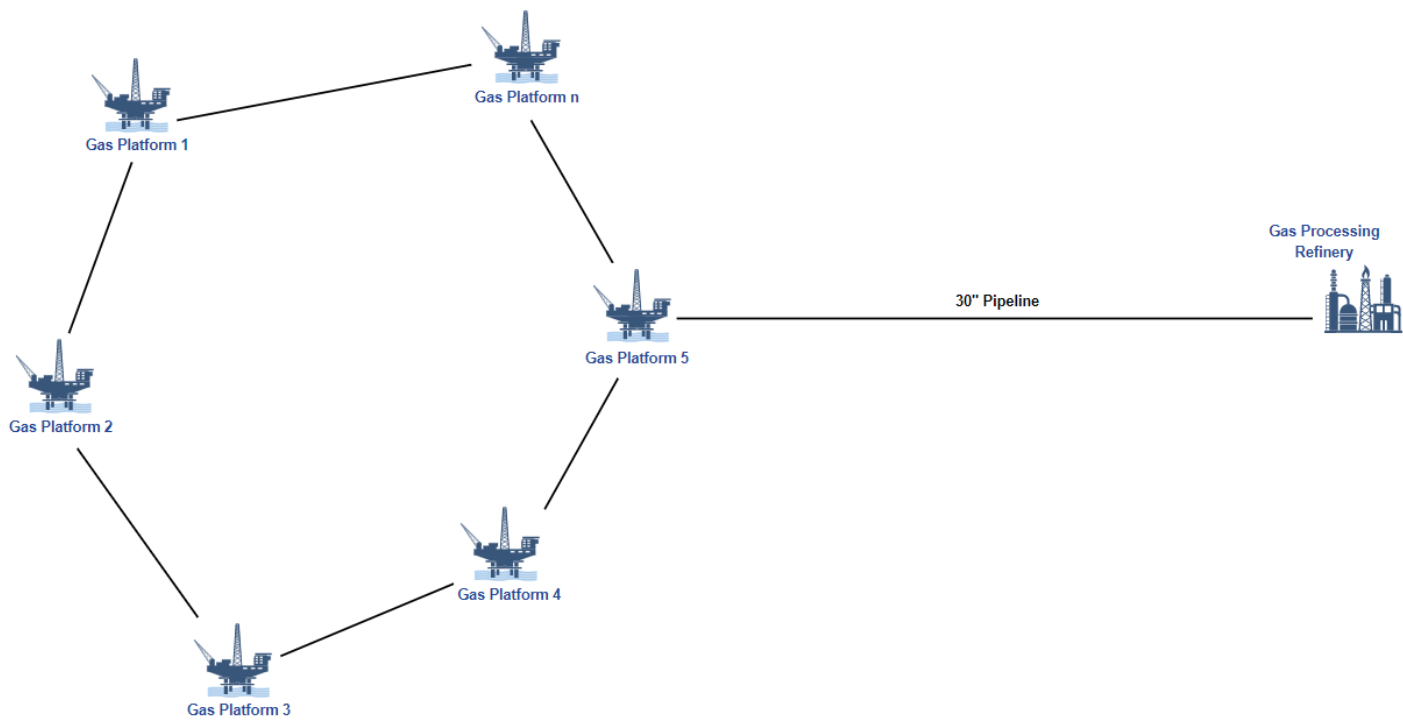
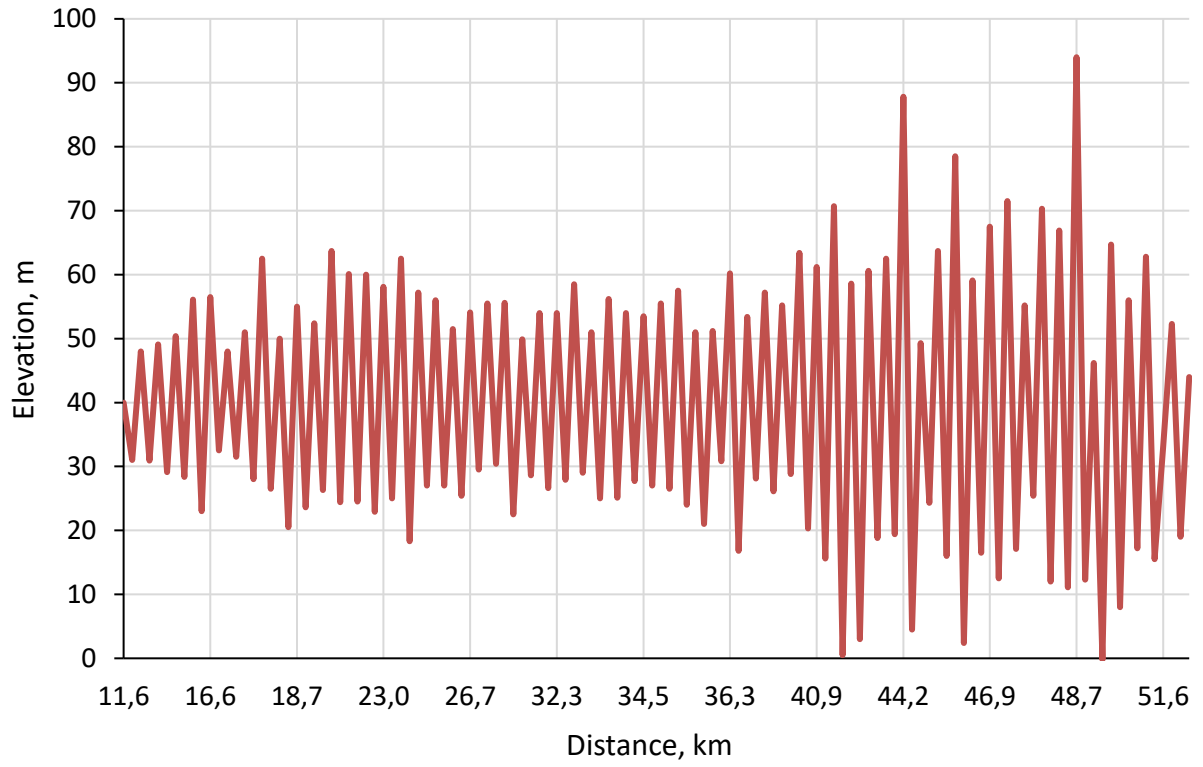


Figure 4 – Field and Pipeline in Discussion



**Figure 5 – Elevation vs. Pipeline Length**