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Real-Time Congealing and Pipeline Monitoring System

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ABSTRACT

In recent years, pipeline operators have faced reduced production environments caused by declining brownfield operations and capital constraints induced by oil prices, among other factors, which have led to pipelines operating well under their designed capacity and challenges such as congealing—the precipitation of wax solids in a crude oil pipeline. This paper discusses how models are built using scientific principles and how simulation may be used to predict where congealing is or may occur inside a pipeline. Finally, a case study from a major oil and gas company's site demonstrates how these modeling and simulation techniques may be effectively applied in the field.

INTRODUCTION AND BACKGROUND

Pipeline operators are currently challenged with operating pipelines safely in reduced production environments, which have been caused by declining brownfield operations, capital constraints brought on by oil prices, and the lack of drilling rigs to keep pipelines full. These present conditions result in pipelines operating well under their designed capacity and challenges such as congealing.

Congealing refers to the precipitation and nucleation of wax solids in a crude oil pipeline. It is initiated by a temperature gradient between the pipe wall and the centerline flow, leading to high-yield flow stress and causing changes in flow behavior.

This paper discusses the physical considerations that contribute and are necessary to detect congealing followed by a series of modeling steps to accurately simulate when and where congealing occurs in a pipeline while accounting for multiphase flow of differing compositions from multiple producers. In turn, this information can automatically be displayed as a visual pipeline profile, allowing operators to understand their entire pipeline operation from remote locations and view critical parameters and events, such as congealing, leak detection, and slugging.

These modeling and congealing algorithms were implemented and validated at a major oil and gas company's site on a 150-km (~93.2 mi) commercial pipeline network used to transport roughly 50,000 BOPD (7,949 m3/day) from 11 gathering stations to a distribution tank farm. The main transportation pipeline was designed to transport 500,000 BOPD (79,490 m3/day). Congealing events were detected and verified by comparing the simulated and assayed pipeline data. Prediction time averaged between three and six hours in advance of the congealing event, allowing the pipeline operator take appropriate mitigation actions and reduce lost production opportunity (LPO).

SCIENTIFIC FOUNDATION

Monitoring pipelines for congealing, or the formation of wax solids in a pipeline, is dependent on numerous factors, including reservoir fluid temperatures, temperature gradients, pressure, elevation, product composition, phase differences, and flow behavior. Wax deposits are typically composed of n-paraffins (normal linear alkane chains) as well as small quantities of branched paraffins and aromatic compounds. Temperature and the concentration of paraffins in the fluid highly affect wax precipitation.

Wax Appearance

Wax solids appear due to temperature or other thermodynamic factors. Figure 1 in the appendix illustrates a typical wax precipitation envelope, which shows a strong dependence on temperature.

Wax appearance in a pipeline is characterized using the following temperature-related terminology:

- Wax formation temperature (WAT) or cloud point: Temperature at which wax crystals form and the fluid takes on a cloudy appearance
    - Below the WAT, a mixture of wax and liquid are present (wax zone)
    - Above the WAT, the fluid is a liquid (wax-free zone)
- Bubble point: Temperature at a given pressure where the first "bubble" of vapor is formed; below the bubble point curve, vapor is present as part of the mixture
- Wax dissolution temperature (WDT): Temperature at which all wax precipitation has been dissolved in heated oil
- Pour-point temperature: Lowest temperature at which oil is mobile
    - In crude oil, a high pour point typically corresponds to high paraffin content and vice versa
    - In crude oil with a lower paraffin content, wax crystals form slower

Modeling these characteristics and creating the wax precipitation envelope is based on the fluid composition, which is determined through a comprehensive crude oil assay performed regularly by the field and pipeline operators.

Thermodynamic factors that may cause wax appearance include

- Fluid temperature is below the WAT
- Temperature gradient between the pipe wall and centerline flow is high
- High-yield flow stress occurring due to changes in flow behavior such as paraffin content, fluid viscosity, flow rates, gas/oil ratio, and heat transfer coefficient

Oil Assay

Crude oil assays define physical and chemical characteristics of crude oil feedstocks and are used for marketing, forecasting, and business decisions, including whether the feedstock is compatible for a specific refinery [2]. Comprehensive crude oil assays may include properties such as molecular composition, viscosity, WAT, pour point temperature, wax dissolution temperature, and specific gravity [3]. This information is used to calculate pure component properties.

Wax Residue Transport

The physical processes that have been investigated [4] as contributing to wax residue lateral transport and deposition in pipelines are molecular diffusion, Brownian diffusion, shear dispersion, and gravity settling. Molecular diffusion is the primary mechanism for transporting paraffins [5]; the other processes may be ignored because the wax crystals are very small compared to the volume flow. The following steps (illustrated in Figure 2 in the appendix) describe the molecular diffusion and wax precipitation process in the pipe [6].

1. During fluid flow, the temperature is cooler near the pipe wall; paraffin molecules are dispersed in the fluid based on temperature, size, and fluid viscosity
2. When at WAT, a wax precipitation (deposit) layer begins to form on or near the colder pipe wall surface, causing a concentration gradient [7]
3. Convective mass flux moves paraffin molecules from the liquid fluid toward the deposit
4. Internal diffusive flux diffuses paraffin molecules into the deposit layer, causing the wax deposit thickness to grow
5. Paraffin molecules in the wax deposit precipitate, increasing the deposit's solid wax fraction
6. Oil molecules (paraffin molecules above a critical carbon number) counter-diffusion out of the deposit

THERMODYNAMIC MODELING

The thermodynamic behavior of diffused wax existing in equilibrium with the liquid enables model development for the real-time congealing and pipeline monitoring system using thermodynamic equations.

When the liquid phase exists in equilibrium with the solid phase at a fixed temperature and pressure, the fugacity (f_i) of component i in both the phases can be determined using equation (1) [8], [9], [10].

$$f_i^L = f_i^S \tag{1}$$

We assume that each component precipitating out of the mixture forms a pure component; therefore, $f_i^S = f_{i_0}^{Pure\ S}$.

In an oil mixture, component precipitation will occur when equation (2) is true.

$$f_i^L \geq f_i^{Pure\ S}, \tag{2}$$

where f_i^L is the fugacity of the ith component in the liquid mixture and $f_i^{Pure\ S}$ is the fugacity of the pure solid formed by that component. According to reaction thermodynamics when a precipitate is formed, the component amount remaining in the solution will be determined by the equilibrium condition in equation (1).

Molecular thermodynamics and regular solution theory have been used to describe solid-liquid equilibrium with the following assumptions:

- Pressure (P) effect is only considered while evaluating fugacity coefficients and is usually neglected except at very high pressure and/or low temperature
- Heat capacity and thermal conductivity are considered

independent of temperature
- Solid/liquid molar volume difference is very low and is, therefore, considered constant
- Chemical potential of each component in all phases is equal
- Several immiscible solid phases of pure components i are formed but are cumulatively considered as one pure solid component for simplification
- At a given temperature, the total amount of precipitated wax is the sum of the contribution of all solid phases that exist in equilibrium with the liquid at that temperature
- No vapor phase exists and only solid-solution equilibrium is considered

Based on solution theory, the liquid-phase fugacity can be calculated using equation (3), while the pure solid fugacity can be calculated using equation (4).

$$f_i^L = \gamma_i^L x_i^L f_i^{Pure\ L} e^{\int_0^P \frac{V_i^L dP}{RT}} \qquad (3)$$

$$f_i^S = \gamma_i^S x_i^S f_i^{Pure\ S} e^{\int_0^P \frac{V_i^S dP}{RT}}, \qquad (4)$$

where, x_i^L is the mole fraction of ith component in the liquid mixture and γ_i^L and γ_i^S are the liquid- and solid-phase activity coefficients, respectively.

Solubility Model

The solubility model uses the molecular solubility approach to describe heavy hydrocarbon mixtures containing fluid with solute (paraffins) and solvent (bulk oil) in a homogeneous liquid state. Based on the solid-liquid equilibrium defined in equation (1), liquid- and solid-phase fugacity can be used to define the molar ratio by dividing equation (3) by equation (4), as given in equation (5).

$$\frac{x_i^S}{x_i^L} = \frac{\gamma_i^L}{\gamma_i^S} \frac{f_i^{Pure\ L}}{f_i^{Pure\ S}} e^{\int_0^P \frac{(V_i^L - V_i^S)dP}{RT}} \qquad (5)$$

In crude oil, the volume differences between the liquid and dispersed wax solid is less than 10%; therefore, the volume difference has negligible influence on the equilibrium. Therefore, equation (5) may be written as shown in equation (6).

$$\frac{x_i^S}{x_i^L} = \frac{\gamma_i^L}{\gamma_i^S} \frac{f_i^{Pure\ L}}{f_i^{Pure\ S}} \qquad (6)$$

To define the fusion of a pure liquid of component i in a solid-liquid equilibrium, an equilibrium constant, K_i^{SL}, is defined as $K_i^{SL} = \frac{x_i^L}{x_i^S}$ [8].

Equation (7) is used to calculate the fusion equilibrium constant for each component in the liquid mixture.

$$\left(\frac{f_i^l}{f_i^S}\right)_{Pure} = exp\left(\frac{\Delta H_{f,i}}{RT}\left(1 - \frac{T}{T_{f,i}}\right) + \frac{\Delta H_{t,i}}{RT}\left(1 - \frac{T}{T_{t,i}}\right) + \frac{1}{R}\int_T^{T_{f,i}} \frac{\Delta C_{P,i} dT}{T} - \frac{1}{RT}\int_T^{T_{f,i}} \Delta C_{P,i} dT\right) \qquad (7)$$

Using the regular solution theory for liquid phase given in equation (7) and expanding it for the multi-wax solid model gives equation (8).

$$K_i^{SL} = \frac{x_i^L}{x_i^S} = \frac{\gamma_i^S}{\gamma_i^L} exp\left(\frac{\Delta H_{f,i}}{RT}\left(1 - \frac{T}{T_{f,i}}\right) + \frac{\Delta H_{t,i}}{RT}\left(1 - \frac{T}{T_{t,i}}\right) + \frac{1}{R}\int_T^{T_{f,i}} \frac{\Delta C_{P,i} dT}{T} - \frac{1}{RT}\int_T^{T_{f,i}} \Delta C_{P,i} dT\right), \qquad (8)$$

where $\Delta C_{P,i}$ is the change in the heat capacity of component i when it melts, $\Delta C_{P,i} = \Delta C_{P,i}^L - \Delta C_{P,i}^S = 1.2739m_i - 0.0019467m_i T$, and m_i is the molecular mass of component i.

Activity Coefficient

Crude oil mixtures contain a variety of microscopic species interacting chemically. While ideal mixtures have enthalpy change of solution values of zero, in real-time crude oil transportation, enthalpy changes are encountered due to species transfer, viscosity, and diffusion. To compensate for these changes and deviations, solid- and liquid-phase activity coefficient values, γ, have been empirically determined in equations (9) and (10), respectively, using solubility parameter values.

$$\ln \gamma_i^L = V_i^L \frac{\left(\delta_m^L - \delta_i^L\right)^2}{RT} \qquad (9)$$

$$\ln \gamma_i^S = V_i^S \frac{\left(\delta_m^S - \delta_i^S\right)^2}{RT}, \qquad (10)$$

where $\delta_m^L = \sum \phi_i^L \delta_i^L$ and $\delta_m^S = \sum \phi_i^S \delta_i^S$.

Equation of State (EOS)

To determine the relative amount of wax precipitate along the pipe surface of a pipe, the fugacity coefficients, ϕ_i, for each component, i, in the oil mixture must be determined. The Peng-Robinson EOS is used to determine the fugacity coefficients, ϕ_i^L and $\phi_i^{Pure\ L}$, for each component in the liquid mixture.

The Peng-Robinson EOS states that the pressure (P), volume (V), and temperature (T) of a pure liquid or gas of component i are related to one another in single-phase mixtures using equation (13) [11], [12].

$$P = \frac{RT}{V - b_{mix}} - \frac{a_{mix}}{V^2 + 2Vb_{mix} - b_{mix}^2}, \qquad (13)$$

where $b_{mix} = \sum_{i=1}^N x_i b_i$, $a_{mix} = \sum_{i,j=1}^N x_i x_j a_{ij}$, $a_{ij} = \left(1 - k_{ij}\right)\sqrt{(a_i a_j)}$, N is the number of components in the mixture, and k_{ij} is the binary interaction parameter between components i and j.

Based on equation (13), the fugacity coefficient of the ith component of the fluid mixture is calculated using equation (14)

$$\ln \phi_i = (Z_{mix} - 1)\frac{b_i}{b_{mix}} - \ln(Z_{mix} - B_{mix}) -$$
$$\frac{A_{mix}}{\sqrt{2B_{mix}}}\left(\frac{1}{a_{mix}}\sum_{i=1}^{N}x_i a_{ij} - \frac{b_i}{2b_{mix}}\right)\ln\left(\frac{Z_{mix} + (1+\sqrt{2})B_{mix}}{Z_{mix} + (1-\sqrt{2})B_{mix}}\right), \quad (14)$$

where Z_{mix} is the fluid mixture compressibility, $A_{mix} = \frac{a_{mix}P}{R^2 T^2}$, and $B_{mix} = \frac{b_{mix}P}{RT}$.

Substituting fugacity equations (3) and (4) into equation (2) produces the inequality given in equation (15).

$$x_i^{Feed} \geq \frac{\phi_i^{Pure\,L}K_i^{SL}}{\phi_i^L}, \quad (15)$$

where x_i^{Feed} is the mole fraction of ith component of the liquid mixture entering the pipe prior to wax precipitation.

Similarly, the equilibrium mole fractions for components in the liquid mixture that produce wax precipitates are determined empirically using equation (16).

$$x_i^L = \frac{\phi_i^{Pure\,L}K_i^{SL}}{\phi_i^L} \quad (16)$$

Using the x_i^{Feed} and x_i^L values determined in equations (15) and (16), the relative amount of wax precipitate (c_s) is calculated using equation (17).

$$c_S = \frac{Moles\,Solid}{Moles\,Feed} = \frac{\sum_i^{Waxes}\left(x_i^{Feed} - x_i^L\right)}{1 - \sum_i^{Waxes}x_i^L} \quad (17)$$

The wax precipitate (c_s) threshold may be used during simulation to alert pipeline operators to congealing in the pipeline when the calculated c_s value is greater than the defined threshold. Therefore, pressure (P), temperature (T), and initial oil mixture composition (x_i^{Feed}) are required model and simulation parameters. Additionally, because wax deposits form along the pipe's inner surface, the oil mixture temperature will be approximately equal to the pipe inner surface temperature, T_W.

CASE STUDY: MODELING PERFORMANCE OF AN OIL PIPELINE

The modeling and simulation techniques presented were implemented and validated at a major oil and gas company's site on a 150-km (~93.2 mi) commercial pipeline network used to transport 50,000 BOPD (7,949 m3/day) from 11 gathering stations (referred to as Producers) to a distribution tank farm. This pipeline was originally designed to transport 500,000 BOPD (79,490 m3/day). Figure 3 in the appendix graphically illustrates the pipeline system model used in this study.

In this pipeline system, oil from the gathering stations flows through a single, 30-inch trunk line to the tank farm, encountering multiple T-junctions; tie-ins; elevation, flow rate, and pressure changes; control values; and measuring devices. Pipeline diameters from the gathering stations to the main trunk line ranged from 6 inches to 10 inches.

Problem Statement

A simulation of the pipeline system model (in a pipe without wax precipitation) experimentally calculated the tank farm to be receiving 53,140 BOPD (8,449 m3/day) from the gathering stations flowing to it. However, the real-time value was significantly lower than this simulated value, leading to the conclusion that congealing was occurring in the pipeline. Goals of this study were to confirm the modeling and simulation techniques described and to determine where congealing was occurring in the pipeline so that appropriate actions could be taken.

Initial Data

Initial data included the following:

- Crude oil assay to determine wax precipitation potential and for oil properties, including temperature and flow rate
- Oil composition (compositionally the same for all gathering stations)
- Transient flow rate from the field
- Pressure data from the field
- Ambient temperature at the station

Table 1 provides the initial data for each gathering station (Producer) collected from the crude oil assay.

Table 1 – Crude oil properties at each gathering station

Station	Density kg/m3 (slug/ft3)	Viscosity(cP)		Oil Temperature °C (°F)	Oil Flow m3/d (BOPD)	Ambient Temperature °C (°F)
		μ1	μ2			
Producer 1	862.6 (1.67)	23.3	7.2	65 (149)	5,002 (31,461.6)	23.9 (75)
Producer 2	862.6 (1.67)	23.3	7.2	64.4 (148)	0 (0)	23.9 (75)
Producer 3	870.2 (1.69)	31.8	9.9	64.4 (148)	361.7 (2,275)	23.9 (75)
Producer 4	853.6 (1.66)	26.2	6	48.9 (120)	60.4 (380)	23.9 (75)
Producer 5	857.4 (1.66)	19.6	5.1	64.4 (148)	292.5 (1,840)	24.4 (76)
Producer 6	863.5 (1.68)	22	6.6	66.1 (151)	241.7 (1,520.3)	23.9 (75)
Producer 7	852.2 (1.65)	12.5	5.3	62.8 (145)	1,123 (7,061)	23.9 (75)
Producer 8	868.4 (1.68)	25.7	6.2	68.3 (155)	370.8 (2,332)	25 (77)
Producer 9	941.4 (1.83)	260.4	65.2	36.1 (97)	146.3 (920)	24.4 (76)
Producer 10	922.6 (1.79)	242.9	42.5	41.7 (107)	244.8 (1,540)	23.9 (75)
Producer 11	861.8 (1.67)	19.4	6.5	57.2 (135)	601.4 (3,783)	23.9 (75)

Simulation Results

Congealing algorithms were applied to the model simulation shown in Figure 4 in the appendix. In this scenario, all gathering stations contributed to the tank farm flow rate except for Producer 2. Additionally, real-time values for oil flow rate and temperature, and ambient temperatures were used in the simulation. In this simulation, congealing was detected in the Producer 9 branch, shown in Figure 4 as the orange and yellow icon.

The oil temerature in the Producer 9 branch was substantially lower than the other gathering stations, leading to congealing. The lower temperature also changed the viscosity in the pipeline, which was visualized as a pipeline profile (trend) across the pipeline length. Figure 5 in the appendix shows the oil's viscosity change as it flowed from Producer 9 to the tank farm as a cumulative average of the net mass flowing through the pipe. The disturbance in the pipeline profile shows the location in the pipeline where congealing was occuring.

Figure 6 shows the pipeline profile for the entire trunkline, from Producer 1 to the tank farm. Because of the contributions of oil from the remaining gathering stations, the pipeline profile uses a cumulative average of the net mass flowing through the pipe. The disturbance in Figure 6 in the appendix is the transmitted increase of viscosity in the congealed branch to the trunk line due to wax precipitation in fluid from Producer 9.

CONCLUSIONS

Based on the results of the case study, the described modeling and simulation techniques effectively detect congealing in real time in pipelines with single-phase and multiphase flow. Congealing events were detected and verified by comparing the simulated and assayed pipeline data. Prediction time averaged between three and six hours in advance of the congealing event, allowing the pipeline operator to respond with the appropriate mitigation techniques and thereby reduce the LPO.

This paper discussed the physical considerations contributing and necessary to detect congealing followed by a series of modeling steps to accurately simulate when and where congealing occurs in a pipeline while accounting for multiphase flow of differing compositions from multiple producers. In turn, this information can be visually displayed as a pipeline profile (as shown in the example case given), allowing operators to understand their entire pipeline operation from remote locations and view critical parameters and events, such as congealing, leak detection, and slugging.

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AUTHOR BIOGRAPHY

Ullas Pathak is a systems engineer working at Statistics & Control, Inc. He currently works on developing automation archit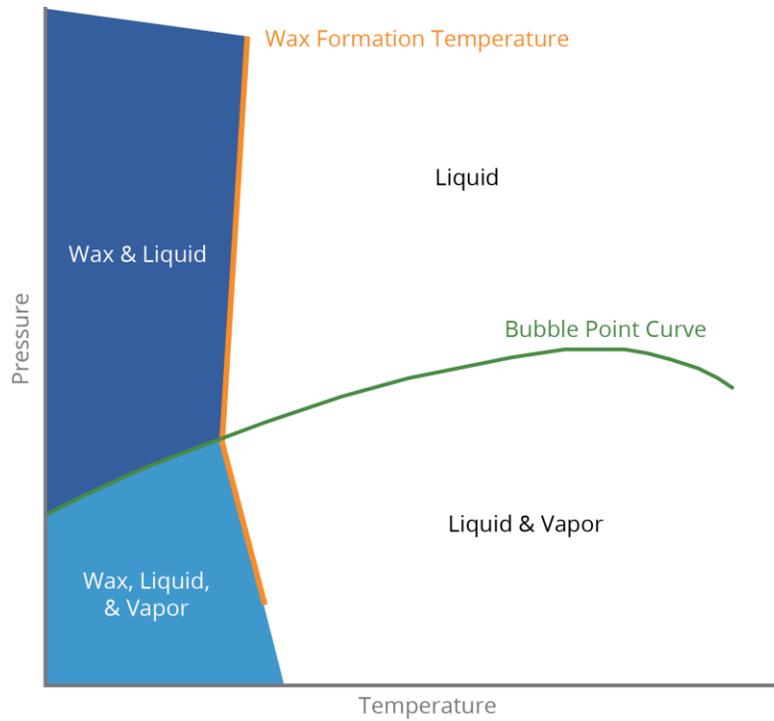ecture, quantitative modeling, optimization strategies for unit control systems, and process dynamics. Ullas has a Bachelor of Technology in Chemical Engineering from the National Institute of Technology in Tiruchchirappalli, India; a Master of Science in Chemical Engineering from Arizona State University, and a Master of Engineering in Chemical Engineering from Iowa State University, specializing in the relaxation behavior of molecules, multicomponent analysis, and biorenewable process technologies.

Dr. Daniel P. Theis is a software engineer working at Statistics & Control, Inc. He has a Bachelor of Science in Chemistry from Bethel College, a Master of Science in Physical Chemistry from the University of Minnesota, and a Doctorate in Computational Chemistry from the University of North Dakota. Daniel has over 10 years of experience in the development and use of computer software designed to predict the properties and behavior of complex chemical systems.

John Hooker is a systems engineer and the *OptiRamp* group manager at Statistics & Control, Inc. He has a Bachelor of Science in Aerospace Engineering from Iowa State University. Over the past 6.5 years, John has participated in several global projects in the oil & gas and power generation sectors and has implemented numerous applications for control solutions, equipment monitoring, design testing, data analysis, and brownfield optimization.

FIGURES

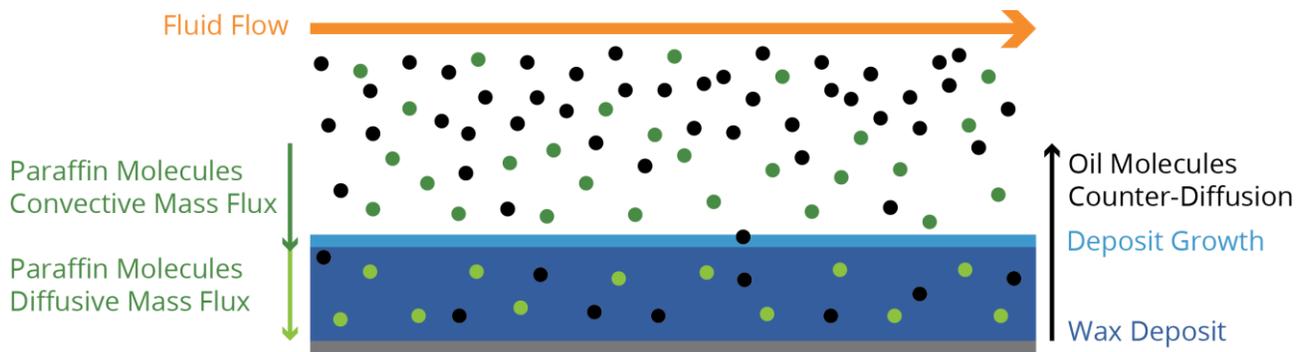**Figure 1 – Typical wax precipitation envelope [1]**
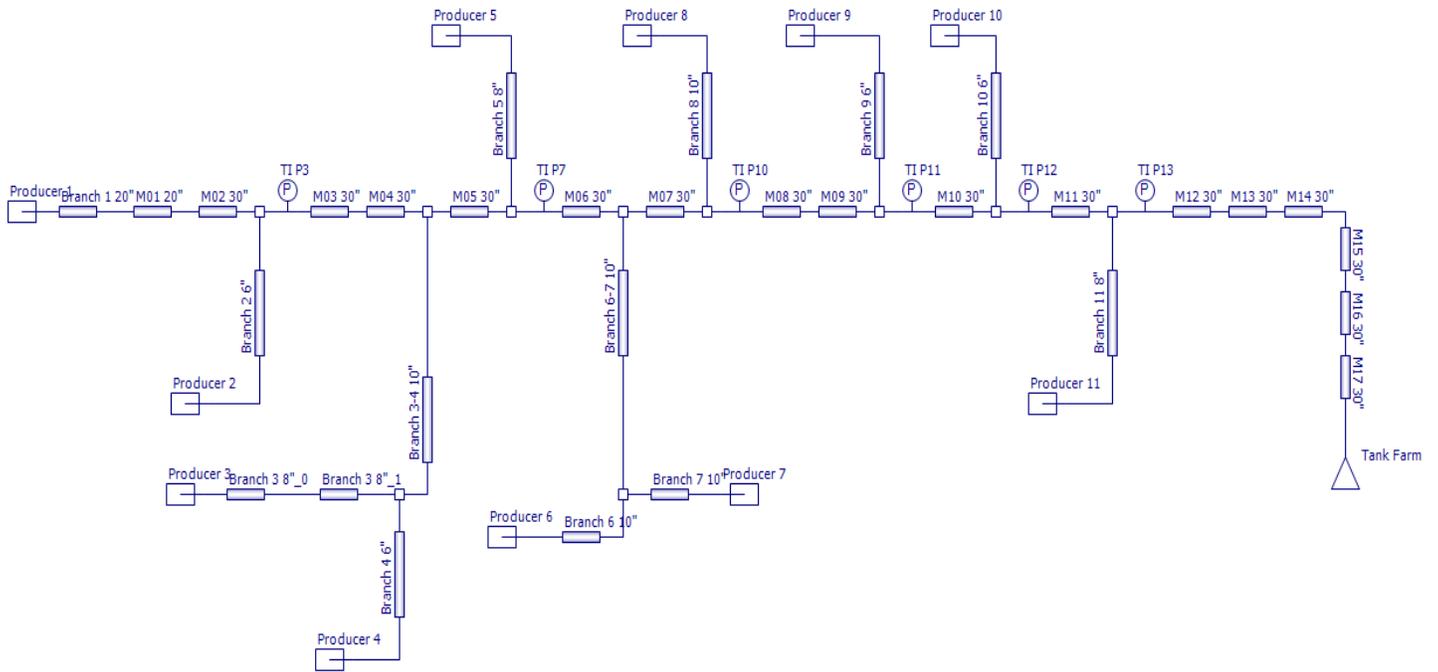

Figure 2 – Molecular diffusion and wax precipitation
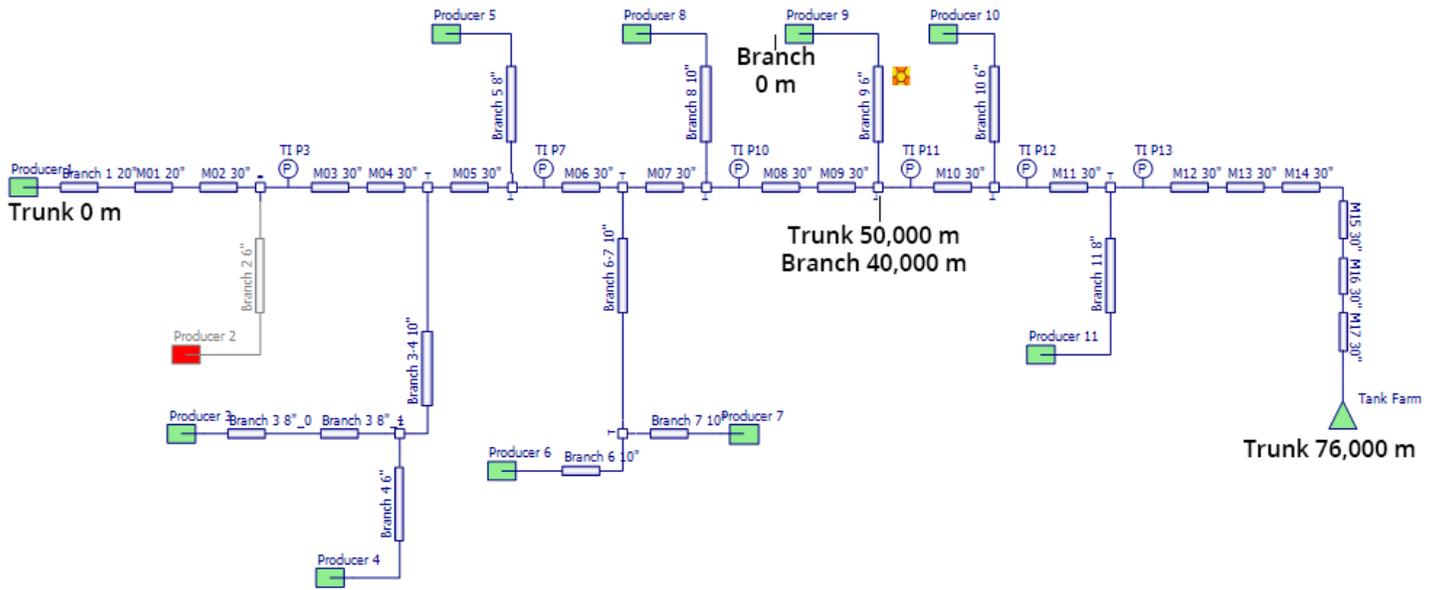
Figure 3 – Pipeline system model

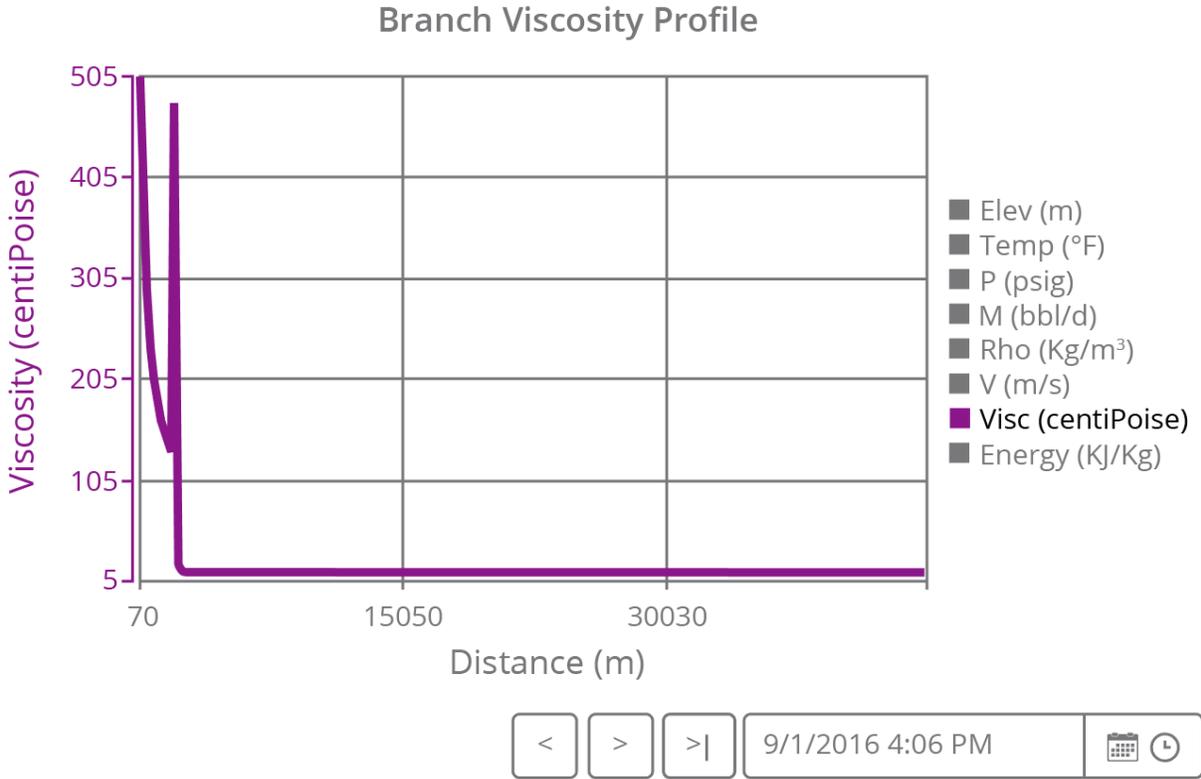**Figure 4 – Pipeline simulation with real-time congealing**

Branch Viscosity Profile

Figure 5 – Pipeline profile highlighting viscosity for Producer 9 pipeline branch
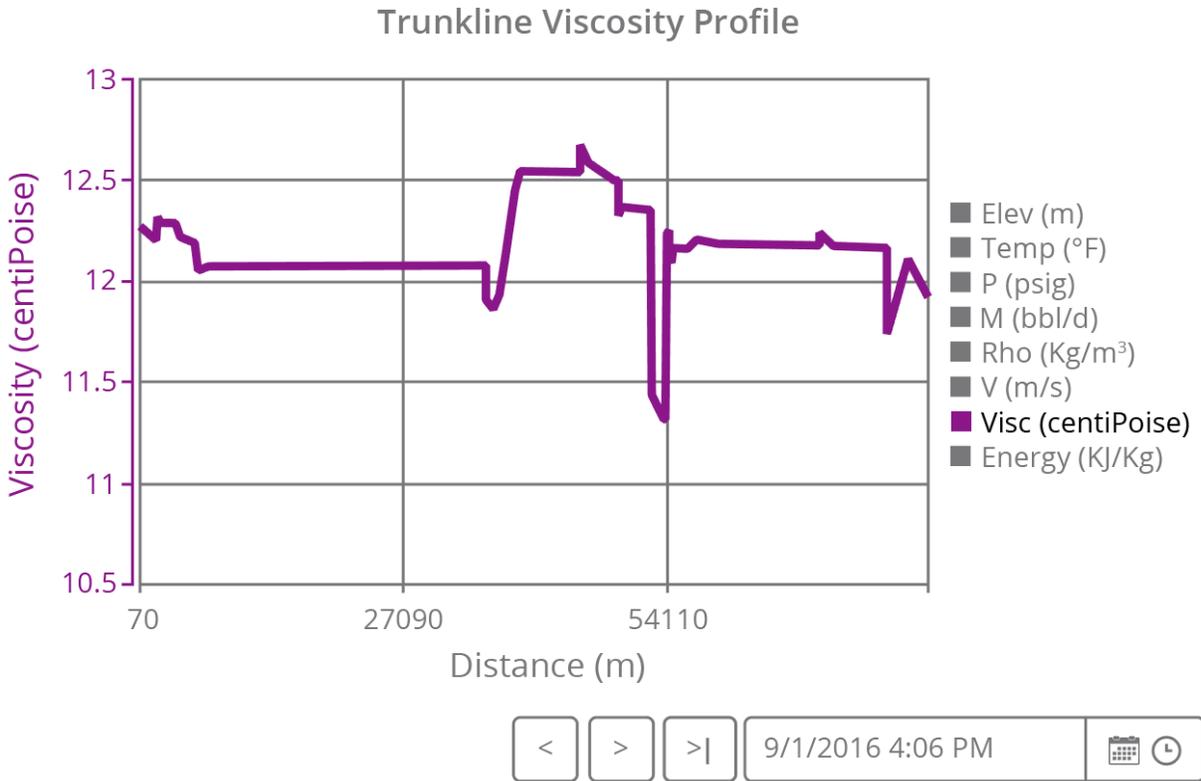
Trunkline Viscosity Profile

Figure 6 – Pipeline profile highlighting viscosity for Producer 9 pipeline branch