A Methodology for Fast Thermodynamic Computation of Mixed Composition Streams: Application to Subsea Pipeline Modelling

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ABSTRACT

We consider the problem of simulating a multiphase pipeline network with streams of various compositions, and subsea separators. An explicit, algebraic thermodynamic model has been developed for mixture and flash calculations. This obviates the need for expensive compositional simulation and yields the speedups necessary for operator training and engineering simulator systems. The model calculates the gas mass fraction in terms of a few “characterization” parameters, determined a priori from compositional data. Predictions are accurate to within 10% over large ranges of pressure and temperature. The model is incorporated into a pipeline simulator and applied to a real-world, deepwater oil and gas system. Relevant operational scenarios involving transients due to stream mixing, varying subsea separator conditions, and gas breakout and absorption in post-separation liquids are illustrated.

I. INTRODUCTION

Multiphase pipeline simulation technology has matured to the point where it is frequently used to tackle flow assurance problems during design as well as day-to-day operation of oil and gas pipelines. Simulations are used, for example, in sizing pipes and thermal insulation; developing procedures for start-up, shutdown, and process upsets; and avoidance of wax and hydrate formation. An emerging application area is “online” simulators [1], which are integrated with field data acquisition systems (DCS/SCADA) and provide real-time information on holdup, pressure, flowrate, and temperature profiles; arrival slug size; and proximity to hydrate or wax formation. A very useful functionality is forecasting: simulations initiated from current or user-specified conditions to predict the consequences of planned operational actions. For example, safe pigging campaigns can be planned through look-ahead simulations to ensure that the swept liquids will not flood the slug-catchers. Online simulators are key to implementing model-based leak/restriction detection systems, which are the only viable leak detection systems for subsea multiphase pipeline networks [2].

Online models are also finding application as operator training (OTS) and engineering simulator systems, especially for offshore and subsea facilities. Here tight safety constraints and complex multiphase operations dictate that the operators be adequately trained to handle process upsets and emergency shut-down procedures, where inappropriate control actions can have severe consequences. The simulator must represent the process with adequate fidelity while retaining the look and feel of the actual control room interfaces with which the operators are familiar. OTS can also be used by process engineers to test control loops, verify the coherence of DCS/SCADA databases, and develop new operational procedures using historical field data [3].

The demands on online systems differ considerably from design simulation tools. Design tools can assume idealized conditions, with precise knowledge of process parameters (inlet/outlet pressures and flowrates, fluid compositions, etc.). Corresponding measurements are often unavailable in the field. Numerical models must
therefore be flexible to use such data as available. Models must handle instrument drop-out, noise, bias and drift, and incorporate filtering to stabilize the numerical model. Compositional uncertainties must be tackled by building in tuning factors, so that predictions match the measured outlet phase rates. In forecasting and OTS applications, the model must run orders of magnitude faster than real time. Often this requires balancing the need for rigor with the demands on robustness, tuneability, and speed.

This paper is a case study of an OTS application that required an innovative approach to compositional simulation and tracking. The pipeline network, shown in Figure 1, is a deepwater system with multiple wells of differing composition. Well streams are mixed and transported via a flowline to a subsea separator. The separated vapour and liquid phases pass to topsides through individual risers. There are additional pipes (service line and downcomer) to circulate dead oil prior to shutdown and start-up operations. The network is operated with several controllers in place (not shown, but included in the model): level controls on the subsea and topsides separators, pressure controllers at riser outlets, and on valves and pumps for dead oil recirculation.

The need for some form of compositional simulation is evident. Individual wells may be operated at variable choke settings, so the downstream blend varies dynamically. Topsides controllers determine the pressures and temperatures at the outlet, and thence throughout the network. Therefore, the model must be capable of flashing the flowline fluids to dynamically changing conditions. Separator flashes determine not only the riser flowrates, but also the composition of riser fluids. This in turn means that gas breakout in the liquid riser (as the fluid loses pressure) and liquid dropout in the gas riser must be dynamically simulated. The extent of gas in the liquid riser is hydrodynamically significant – determining the hydrostatic pressure, the multiphase flow regime, propensity to slugging, phase rates at the outlet, and the operation of topsides equipment.

The conventional approach would be to integrate the pipeline solver with a full compositional simulation model. Compositional simulations are expensive, typically taking up 50-70% of total computational time [4]. The burden increases with the number of compositions handled, to the extent that the addition of further wells to the network would reduce model speed too close to real time. This was considered unacceptable for this OTS application. Speedup was the driving motivation for the thermodynamic model developed in this study. A secondary motivation was the need to incorporate tuning factors so that discrepancies between actual and design-time compositions could be reconciled in the field. This is vital to allow the OTS to have a shelf-life past first oil period of the field.

The remainder of this paper is organized as follows: section II presents an outline of the thermodynamic model used for compositional analyses; section III discusses validation tests comparing model predictions with those obtained from rigorous compositional simulation; section IV presents results of thermal-fluidic simulations using the pipeline solver; section V illustrates the gains in computational speed; concluding remarks appear in section VI.
II. THERMODYNAMIC MODEL

The goal of the work described herein is to allow thermal-fluidic simulation of pipeline networks without having to resort to full compositional simulation. Toward this objective an approximation model was empirically developed and tested.

The model exploits the fact that baseline compositions of production fluids from individual wells are known \textit{a priori}. These compositions can be subjected to rigorous analysis before pipeline simulation is initiated; the variation of thermodynamic parameters over the requisite range of pressures and temperatures is tabulated and available for use during pipeline simulation. The key parameters needed for hydrodynamic simulation are gas mass fraction (GMF), free water fraction, phase enthalpies, average molecular weight of the vapour phase, and heat capacities. When dealing with dynamically varying blends of the known compositions, excellent approximation to some parameters (e.g. phase enthalpy) is obtained simply by weighted blending of the thermodynamic properties of individual compositions. The thermodynamics of vapour-liquid equilibria (VLE), however, imply that GMF cannot be similarly approximated. GMF is used extensively in the pipeline model to calculate phase mass transfer, bubble point transition, and liquid holdup. Inaccurate estimation of holdup can result in incorrect prediction of multiphase flow regime, slugging characteristics, and hydrostatic pressure drop. GMF prediction for dynamically varying blends and post-separation fluids, therefore, forms the focus of this modelling effort.

The features desired in a compositional model for the OTS include the following:

1. Computational burden much smaller than conventional phase equilibrium analysis.
2. GMF prediction for separated fluids under varying separator conditions.
3. GMF prediction for varying blends of known compositions.
4. Tuning of composition in the field (to match measured phase rates, available typically at the outlet), without requiring re-analysis.

To meet these objectives, we target three sub-problems: (a) characterization: describing a composition using only a few parameters in such a way that GMF can be recovered at arbitrary pressure ($P$) and temperature ($T$); (b) composition shift: predicting
GMF behaviour of the liquid obtained by phase-separation of the characterized fluid at a prescribed $P$, $T$; (c) mixing: predicting GMF behaviour of an arbitrary blend of two characterized fluids.

Model predictions were validated by comparison with results from the Grand Unified Thermodynamic Simulator (GUTS). GUTS was developed as a joint-industry project (JIP) at the University of Tulsa with companies such as Conoco, BP, Amoco, and Arco. GUTS has been extensively validated and used at MSi and partner companies in the original JIP.

Characterization. Compositional analysis is simplified by modelling the mixture as a two-component fluid, with one pseudo-component playing the role of a volatile hydrocarbon liquid or gas, and the other pseudo-component modelling heavy hydrocarbon liquid.

Recall that classical phase equilibrium analysis derives from species conservation laws and individual-component phase equilibrium behaviour described in terms of $K$-values (the ratio of gas to liquid moles). The problem is usually solved by framing an objective function to be minimized, resulting in the Rachford-Rice equation for a multicomponent mixture [5]. The equation can be solved using Newton-Raphson iterations. When there are only two components, the solution is available in closed form:

$$G = \frac{Z_i(1-K_i)+(1-Z_i)(1-K_2)}{(1-K_i)(1-K_2)},$$

where $G$ is the ratio of gas moles to feed moles, $Z_i$ is the mole fraction of the $i$-th component in the feed, and $K_i$ is the $K$-value of the $i$-th component. In Rachford-Rice analysis, initial estimates for $K_i$ can be obtained from the Wilson approximation [6], which requires knowledge of critical temperatures, pressures and acentric factors.

The model employs an analogous functional form for GMF, but treats $Z_i$ and $K_i$ as unknown, fitting parameters that must be estimated from known compositional data. In effect, $Z_i$ loses its original meaning of a component mole fraction. $K_i$ is assumed to be a function of $P$ and $T$, but $Z_i$ is held fixed. The functional dependence of $K_i$ on $P$ and $T$ is similar to the Wilson formula, but with an additional power-law dependence on pressure for the heavy component:

$$K_1 = C_1 \exp(B_1/T)/P, \quad K_2 = C_2 \exp(B_2/T)/P^n.$$

This power coefficient $n$ accommodates the pressure dependence of the fugacity and activity coefficients.

The characterization procedure, in all, can involve six scalar parameters: $Z_i$, $n$, $B_i$, and $C_i$ ($i = 1, 2$). In practice, it was found adequate to keep $n$ fixed; the results in section III are obtained using only five variable parameters. These parameters are solved using an iterative optimization procedure to minimize the difference between modelled and calculated (using GUTS) GMF over the entire domain. Note that this calculation proceeds in offline mode (i.e., not as part of the pipeline simulation), and the calculated parameters are utilized by the pipeline solver during simulation.
**Composition Shifts.** Shifting refers to the change in composition after the multiphase hydrocarbons have been phase separated. The goal of the shifting procedure is to re-characterize each phase using only the originally characterized parameters of the feed.

Shifting was achieved by allowing only $Z_1$ to vary. A physical interpretation is that phase-separation can be modelled by changing the quantity of the light and heavy pseudo-components in the separated phases, with the thermodynamic properties of these pseudo-components remaining invariant. In this model, the new phase $Z_1$s can be obtained in closed form by inverting the original GMF expression. Thus,

$$
\Delta Z_1 = \frac{(GMF)K_1'K_2' - (K_1 - K_2)Z_1 - K_2}{K_1 - K_2},
$$

where $K_1' = (1 - K_1)$. Here $K$-values are evaluated at the known separator $P$ and $T$.

**Mixing.** The mixing problem is to derive new characterization parameters for a blend of known fluids using only the feed characterization parameters. The blend ratio is assumed to be known in terms of the mass rates of the incoming streams containing the known fluids.

Mixing is achieved by combining $Z_{i}^{(j)}$ and $C_{i}^{(j)}$ on a mass-basis (here the superscript is the mixture index in the blend). $B_i$ is calculated by requiring the single-composition $K$ formulas to hold at a reference condition:

$$
B_i = T_{ref} \ln(K_i P / P_{ref} C_i),
$$

where $K_i$ is the blend $K$-value for the light component at the reference condition. This ensures that the mixing rules remain valid in the limit of 100-0, 0-100 blend ratios.

These modelling procedures have been applied in simulating the pipeline network. Note that a key feature is that all three aspects of the model provide closed-form expressions for GMF. This is crucial in reducing the model’s computational effort.

Before discussing results from this model, we note that there are some similarities between the present model and existing “black oil” models (e.g. Lasater, Standing correlations; see [7]). Both use a two-component approximation to accelerate phase equilibria calculations. Indeed we initially attempted a black oil modelling approach, but found it to be unsatisfactory. First, errors in the predicted gas volume fraction were upwards of 25% for some relevant ranges of $P$ and $T$. More importantly, solution of the black oil modelling parameters requires more reference points (e.g. gas volume fractions at two reference points) than are available in practice. Existing black oil models would therefore have to be extended to meet the characterization, shifting, and mixing functionalities described above. These shortcomings contributed the development of the present modelling approach.

**III. RESULTS**

We present here results validating the thermodynamic model. Predictions are compared with GUTS data. $P$ and $T$ were varied over 20-200 bara and 15-100 °C, respectively. These ranges are adequate for the pipeline network at hand.
Three compositions have been analyzed (Table 1): a high-GOR fluid (composition M1), a low-GOR fluid (M2), and dead oil (M3). M3 was obtained by flashing 50-50 blend of M1 and M2 to standard conditions and extracting the liquid.

The characterization procedure involves multi-parameter constrained optimization to minimize the global (least-squares) error between modelled and calculated GMF. The problem can be solved using standard optimization programs; here, the fitting procedure has been implemented in a Microsoft® Excel® spreadsheet.

**Table 1.** (a) Compositions (in %) used for model validation. (b) Representative properties at standard conditions (1 bara and 15.5 C).

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<th>M1</th>
<th>M2</th>
<th>M3</th>
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<table>
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**Characterization.** Figure 2 shows modelled and calculated GMF for M1 and M2. It can be seen that P-T dependence is captured quite accurately most of the domain for both compositions. The global maximum of the error is approximately 7% for M1, occurring at the low-pressure/high-temperature and high-pressure/low-temperature extrema of the domain. Over the bulk of the domain errors are less than 2%. For M2, the most severe normalized errors occur at the high-pressure/low-temperature extrema, where the fluid is close to being single-phase liquid. In these regions, the unnormalized GMF error is less than 0.005. Again, over the bulk of the domain, errors are approximately 2%. At the high-pressure/high-temperature ends, errors for M1 are approximately 2% and for M2 (fluid close to single-phase liquid) the unnormalized GMF error is 0.0025. In regions close to the bubble point, the model slightly over-predicts the amount of gas.

**Shifting.** The shifting procedure was applied to compositions M1 and M2. We consider two cases. First, the compositions are perturbed by a relatively small amount: the original composition is flashed to a prescribed pressure and temperature; the liquid and vapour phases are separated and recombined in a different molar ratio. The mixture
was such that the GMF of the recombined fluid differed from the original by approximately 20%. The model is then tested by seeking the new GMF behaviour using only the original characterization and knowledge of the recombined fluid’s GMF at the single reference $P$, $T$ condition. The purpose of this exercise was to test the tuneability of the model. In the field, actual compositions are often found to differ from those assumed before field start-up, and it is important to be able to re-calibrate the model without requiring extensive full-fluid sampling and analysis, and associated thermodynamic calculations.

Figure 2: Comparison of calculated (lines) and modelled (symbols) GMF with varying pressure, temperature, and composition. (a) GMF vs. $P$ for $M1$, (b) GMF vs. $T$ for $M1$, (c) GMF vs. $P$ for $M2$, (d) GMF vs. $T$ for $M2$.

Figure 3 shows results for perturbed compositions differing by +20% and -20% from the original $M2$ composition. Again, errors are small over the bulk of the $P$-$T$ domain, testament to the robustness of the characterization and shifting procedures. At the high GMF end (low-pressure, high-temperature) the error is approximately 7%. At the low GMF end (high-pressure, low-temperature), when the fluid is close to single-phase liquid, the raw GMF error is less than 0.0015. At the high-pressure/high-temperature end (upstream conditions in the network), the errors for the +20% composition are negligible (< 0.01%). The -20% composition is close to its bubble point in this region, and consistent with the characterization errors, the model slightly over-predicts the amount of gas; the unnormalized GMF error here is approximately 0.0025. For $M1$ (not shown), model predictions are similar, with errors less than 8% everywhere except at
the highest temperatures and pressures. Note that the error generally increases with decreasing proximity to the reference \(P, T\) point. In practice, points of interest in the pipeline network will be localized near the reference point, and therefore we can expect small errors (<3%) over the domain.

Figure 3: Comparison of calculated (lines) and modelled (symbols) GMF for compositions obtained by “shifting” M2 so that the GOR is increased by 20% (a, b), and decreased by 20% (c, d): (a) GMF vs. \(P\), (b) GMF vs. \(T\), (c) GMF vs. \(P\), (d) GMF vs. \(T\). Solid circle indicates the \(P, T\) condition to which the composition is flashed.

In the second test of the model, a much more drastic shift is applied for predicting the GMF behaviour of liquid obtained from a separator. Again, the separator pressure and temperature are assumed to be known, but now the liquid GMF is trivially 0.

Figure 4 shows results for an M1 fluid separated at \(P = 101\) bara and \(T = 70\) °C. Model predictions are expectedly less accurate since the liquid composition is significantly different from the feed fluid. In the present application, liquid from the subsea separator loses heat to the ambient; there is significant pressure drop as well. The operative range of the model is therefore over pressures and temperatures lower than the separation point. Errors generally increase with decreasing proximity from the separator \(P, T\) values, and at the lowest \(P\) and \(T\), errors are in the 35-50% range. This is an area where the model performance should be improved. It should be noted that the characterization procedure is amenable to further tuning to reduce this discrepancy by selective optimization, and by reducing the \(P, T\) domains to the operational ranges. Of particular interest is the bubble point pressure, since this determines the gas breakout.
location in the riser, and hence the extent of the pipeline susceptible to slugging. The model prediction of the bubble point is within 2-3 bara for the range of riser pressures encountered in the present application.

**Figure 4:** Comparison of calculated and modelled GMF for the liquid phase of an M1 mixture separated at $P = 101$ bara and $T = 70$ C. (a) GMF vs. $P$, (b) GMF vs. $T$.

**Mixing.** The mixing procedure was applied to several combinations of M1 and M2. Figure 5 shows results for a 25-75 mixture (by mass) of M1 and M2. Predictions are within 3% error over the bulk of the domain, with the global maximum error of approximately 15% occurring at the low-pressure/high-temperature, high-temperature/low-pressure extrema. Again, these regions are outside the normal operating range of the pipeline network, and thus model performance is adequate.

**Figure 5:** Comparison of calculated and modelled GMF for a 25-75 mixture of M1 and M2. (a) GMF vs. $P$, (b) GMF vs. $T$.

**IV. PIPELINE SIMULATION**

The thermodynamic model was integrated into MSi’s pipeline solver, the Pipeline Integrity Management System (PIMS). PIMS is a transient pipeline model capable of handling converging or diverging networks with one-, two-, or three-phase (gas, hydrocarbon liquid, and water) flows. PIMS is designed for online operation with algorithmic emphasis on robustness, speed and configuration flexibility. The pipeline model can be integrated into a generalized modelling architecture to simulate process
and control components. Here the generalized model included the pipeline network, well models (using the inflow performance relationship), subsea and topsides separators and pumps, and various controllers. The model was run with the subsea separator under level control (maintained by running the subsea pump at variable speeds) and a choke valve at the gas riser outlet under pressure control.

For deployment, the pipeline simulator is supported by a number of auxiliary software components to handle communication with SCADA interfaces, data historizing, alarms management, graphical user interface (GUI), etc. The end product is a fully customized software package designed for installation in control rooms. Operators are presented with several overview and drill-down screens providing access to all aspects of model functionality. A screenshot of the simulator’s workstation GUI is shown in Figure 6. Some of the simulation results below are plots taken from this workstation interface.

![Figure 6: Screenshot from the pipeline simulator GUI, showing pop-ups for controllers and process parameter trends.](image)

Thermodynamic calculations within PIMS are normally based upon lookup-table data generated using GUTS. (In most cases, GUTS data is generated at configuration time, although there is provision for including full compositional simulation at run time as well.) Here, characterization is performed during model configuration. The modelling parameters are included in the lookup tables inputted to PIMS. Shifting and mixing calculations are performed by PIMS at each time step and at each of the spatial discretization points (pipe sections) in the network.

We discuss below four specific simulations, chosen to illustrate model performance under operationally relevant flow conditions.

**Simulation I: Varying outlet pressure.** To test the characterization procedure, the network is operated with only the M1 well producing at a fixed choke position. Outlet pressure variation causes changes in pressure throughout the network, and also affects the flowrate produced by the well. Figure 7(a) shows pressure variations at the separator and the flowline inlet (downstream of well choke). With decreasing (increasing) pipeline pressure, the amount of gas produced at every section of the flowline is increased (decreased). This is reflected in the flowline liquid holdup (volume-integrated liquid content) trend (Figure 7b). A decrease in the separator
pressure causes more of the lighter components to escape into the gas phase, rendering the separated liquid less volatile and reducing the amount of gas breakout in the liquid riser. The model captures this trend, as seen from the liquid holdup profiles at outlet pressures of 30 and 50 bara (Figure 7c). GMF error at the flowline outlet (Figure 7d) is obtained by comparing flashes within PIMS with GUTS data. The agreement is quite good, with errors restricted to ~ 1%.

Figure 7: Simulation I results. Trends of (a) pressure at gas riser and flowline outlets; (b) flowline total liquid holdup; (d) GMF error (PIMS versus GUTS). (c) shows profiles of liquid riser holdup fraction during operation at 50 bara (solid) and 30 bara (dashed).

Simulation II: Mixing. To test the mixing procedure, the network is operated with varying settings for the M1 and M2 well chokes. Figure 8(b) shows the produced liquid volume rates at the flowline inlet. In terms of mass rates, the successive states correspond to 0.51/0.49, 0.76/0.24, and 0.22/0.78 blends of M1 and M2. The compositional blend is dynamically tracked as the fluids move down the pipeline. An example of the blend profiles (during 0.22/0.78 operation) is shown in Figure 8(a), where it can be seen that fluid near the pipeline outlet is mostly M1 and near the inlet is mostly M2. These blend values are used in the local mixing calculations to estimate sectionwise GMF. A comparison of PIMS and GUTS GMF predictions (Figure 8c) shows that, in accordance with expectation, there is less gas with increasing content of the low-GOR M2 fluid. Model error at the flowline outlet (Figure 8d) is generally within 2%. Larger values occur during blend transitions, when the blend values used in
GUTS did not exactly match PIMS. These errors are artefacts of the comparison procedure.

Figure 8: Simulation II results. (a) Profiles of blend along flowline during the final phase of simulation. (b) Flowrate trends from each well. (c) Modelled and calculated GMF at flowline outlet. (d) GMF error.

Simulation III: Flash Calculation. A simulation similar to case I is performed with only the M1 well producing. The pressure setpoint at the gas riser outlet is successively varied between 60, 50, and 70 bara. Outlet pressure variation causes the well to produce at different rates, which in turn results in flowrate and temperature variations throughout the network. Figures 9(a, b) plot the temporal variation of pressure and temperature at various points in the network. As the subsea separator pressure and temperature change, so do the compositions of the gas and liquid phases produced at the separator. In particular, when the separator operates at high pressure, more of the volatile components are retained within the liquid phase at separator conditions; vice versa at low pressure.

This variation in post-separation compositions is reflected in the temporal variation of gas at the liquid riser outlet (Figure 9c). The first pressure transient (from 60 bara to 50 bara) results in increased production from the wells and the consequent increase in the liquid riser flowrate. This causes a transient spike in the gas rate at the riser outlet. The lowered separator pressure means that the separated liquid is less volatile; this eventually increases the liquid riser holdup (see also figure 7e) and reduces the outlet gas rate: note the drop in gas flowrate at liquid riser outlet during separator operation at 50 bara, even though the outlet pressure/temperature conditions are nearly invariant.
Finally, the second separator pressure transient (from 50 to 70 bara) increases the content of lighter components in the liquid phase, contributing to a rising gas rate at the outlet.

For quantitative estimation of model performance, Figure 9(d) shows a comparison between GMF calculated within PIMS and that obtained from GUTS. GUTS GMF is obtained by flashing the M1 composition to various $P, T$ (74.5 bara, 37 $^\circ$C; 87.5 bara, 40 $^\circ$C; and 62.1 bara, 37 $^\circ$C) occurring at the separator, as predicted by PIMS. The trends are presented at a location two-thirds of the way up the liquid riser (1600m from the separator). The trends qualitatively agree with the expectation that gas breakout should increase with increasing separator pressure. Note in particular that PIMS correctly predicts no gas breakout at 50 bara. The largest error occurs during 70 bara operation, when PIMS underpredicts the GMF by approximately 7%.

It is worth pointing out that we have neglected PIMS GMF errors in the flowline and well tubing, resulting potentially in incorrect $P, T$ at the separator. While the actual total error can be quantified only through comparisons with a full compositional simulation, we can note that GMF errors in simulation I were rather small (~1%). Thus, it is reasonable to expect that the actual errors are of the same order as seen in Figure 9(d).

![Figure 9: Simulation III results: (a) Pressure variation at different points from separator to liquid riser outlet, changing in response to varying gas riser outlet pressure. (b) Corresponding temperature variations. (c) Variation of outlet gas mass rate from liquid riser. (d) GUTS (symbols) vs. PIMS-calculated GMF at a point on the liquid riser.](image-url)
**Simulation IV: Dead oil recirculation.** Recall the basic modelling assumption that the fluid mixture comprises one volatile and one heavy pseudo-component. Despite this drastic approximation, it is possible to capture the effects of complete gas absorption by heavy (dead) oil. To illustrate this, a simulation was performed in which the volatile M1-M2 mixture is evacuated from the pipelines by M3 fluid pumped down from topsides storage vessels. The simulation is initiated from a shut-in condition (no flow) with the entire network filled with M1-M2 fluid. Figure 10(a) shows trends of M3 blend at flowline and liquid riser outlets. As the volatile fluid is removed, gas mass rate drops to zero first at the flowline outlet and later at the liquid riser outlet as well (Figure 10b). GMF errors during this transition (not shown) are in line with levels seen in earlier cases.

![Figure 10](image.png)

**Figure 10:** Simulation IV results: (a) Trends of blend M3 fraction at flowline and liquid riser outlets. (b) Trends of gas mass rate at flowline and liquid riser outlets.

**V. SPEEDUP**

The replacement of iterative VLE and property calculation algorithms with an explicit algebraic model expectedly yields considerable computational speedup. Conventional analysis is a multistage process, comprising stability analysis to determine the number of phases and initial guesses for $K$-values, followed by a root-finding procedure (e.g. by solving the Rachford-Rice objective function) to solve for the mole fractions, and finally the computation of phase molar volumes and other properties in terms of an assumed equation-of-state model. Additional iterations or algorithmic refinements are often required when the objective function associated with a particular composition has multiple poles [4]. The computational burden increases as a power of the number of components in the composition.

Table 2(a) presents timing results for a single flash and GMF calculation using GUTS, as a function of the number of components (NCOMP). Note that computation time increases 20-fold as the number of components increases from two to 16. The simplified thermodynamic model represents any composition as a mixture of two pseudo-components, and does not suffer from the scaling issue. A flash and GMF calculation involves approximately 20 mult-add and two exponentiation operations. It can be seen that a 30-fold speedup over GUTS would be obtained even if the actual composition only contained two components.

Within the pipeline solver, model calculations are performed in several stages. First, the compositional blend is determined at each pipeline section, and the mixing rules are
used to obtain the characterization parameters for the local blend. Second, these parameters are used to calculate pseudo-component $K$-values at the local pressure and temperature. Third, the $K$-values are used to determine sectionwise GMF and phase mass transfers. Since all calculations are local to a pipeline section, a single model time-step involves as many flash/property calculations as the number of sections in the network. Further computational benefit is obtained by performing mixing and $K$-value calculations only when blend, pressure, or temperature values have changed.

Table 2(b) shows timing results per pipeline network time-step, averaged over a transient PIMS simulation. The GUTS value is the estimated time (based on Table 2a) for a 16-component mixture (as used in this study), while retaining the pipeline solution algorithm. It is important to note that while the model thermodynamic calculations take up less than 0.1% of total PIMS time, the burden would be greater than 50% were full compositional simulations to be used. The net speedup benefit of using this model is more than 500-fold, in terms of the time required for thermodynamic calculations. In terms of the overall calculation time, the speedup benefit from using the model is 58%.

**Table 2:** (a) Timing results for a single flash/property calculation. (b) Time taken by thermodynamic calculations per time-step, in seconds and as fraction of PIMS time.

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<th>MODEL</th>
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**VI. CONCLUDING REMARKS**

We have presented a simplified approach to compositional simulation in complex multiphase pipeline networks that yields acceptable accuracy at a drastically reduced computational cost. Further work is underway to extend the model for calculation of phase densities and viscosities. More broadly, the work highlights the type of pragmatic computational approaches required to leverage existing simulation technologies in full-field simulators. Online simulators capable of running orders-of-magnitude faster than real time open up a wide range of applications, including advisory or expert systems for control room operators, training simulators, model-based leak detection and location, and inferential well metering systems.

**GLOSSARY**

- $B$: Fitting parameter in model
- $C$: Fitting parameter in model
- DCS: Distributed Control System
- GMF: Gas mass fraction
- GOR: Gas-oil ratio at standard conditions, usually expressed in scf/stb
GUI  Graphical User Interface
GUTS  Grand Unified Thermodynamic Simulator: the phase equilibrium solver used in this study

$K$  Ratio of gas to liquid moles of a species in a mixture

M1 – M3  Compositions used in this study (see Table 1)

$n$  Fitting parameter in model

NCOMP  Number of components in a mixture

OTS  Operator Training System

$P$  Pressure

PIMS  Pipeline Integrity Management System: the pipeline solver used in this study

SCADA  Supervisory Control and Data Acquisition system

$T$  Temperature

VLE  Vapour-Liquid Equilibrium

$Z$  Mole fraction; fitting factor describing contribution of light or heavy pseudo-component to the gas mass fraction.

REFERENCES


