COMBINING MIXED-COMPOSITION PETROLEUM STREAMS USING PROXY MODELS OF EQUATION OF STATE FOR COMPOSITIONAL INTEGRATED PRODUCTION MANAGEMENT

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ABSTRACT

When two distinct offshore reservoirs produce in the same region, the same gathering system for both reservoirs may be used. It is recommended that the gas/oil composition is modeled because its effects can impact reservoir forecasting. An incorrectly-modeled Integrated Production Management (IPM), i.e. one with insufficient data, can impact system management negatively due to the lack of a rigorous compositional modeling impact assessment. To fully study this problem, there need to be a method to model mixed-composition petroleum streams. Most proposed methods to carry this task are Pseudo-Compositional (based on K-values) or are based on Black-Oil models. The objective of this work is to develop a method for mixed-composition petroleum streams to be used in future compositional IPM optimization studies. This paper implements, validates, and discusses the limitations and reproducibility of a mixed-composition petroleum stream. It models a methodology based on an Equation of State (EoS) and PVT data, and combines concepts proposed by Carpio (2012) to automate the process. This work compares the proposed method with two mixing methods from the literature. The three mixing methods studied are based on concepts of tuned EoS and the well-known methods of EoS from Peng-Robinson (1978), and volume translation from Jhaveri-Youngren (1988). The analysis and comparison of the methods are based on conventional data and simulated tuned experiments, when applicable. The results of this study show a small but significant variability of mixed-stream properties from the three proxy methods with the potential to impact optimal reservoir control conditions. Therefore, the deviation potential of mixed-stream models is worth further investigation, thus, justifying a subsequent sensitivity study, with focus on closed-loop IPM.

KEYWORDS

Reservoir Engineering; Compositional Reservoir Simulation; Equation of State (EoS); Integrated Production Modeling (IPM); Petroleum

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1. INTRODUCTION

Oil and gas field management and operational optimization are complex because of many factors: multi-component fluid production, sour or hazardous gas releases, complexity in determining any interconnection between operational variables and operational limitations, multiple, and sudden temporary modifications (Giorgio et al., 2012). In this context, mixing production streams affect all the issues listed above.

A reoccurring situation in projects of integrated production systems is the joining of two or more streams with different petroleum characteristics, such as those from different fields (Carpio, 2012).

The reliability of Integrated Asset Modeling results can be compromised as the applied fluid-dynamic models neglect various intrinsically transient and/or thermal phenomena, such as liquid loading, slug/unstable flow, and Joule-Thomson cooling (Lullo et al., 2011).

Modeling the contour conditions for reservoir production systems can be an “uncoupled” (standalone) architecture from the wellhead to the platform separator using Black-Oil models (Mattax & Dalton, 1990). This uncoupled Black-Oil type production modeling may be unsuitable in several scenarios. For example, scenarios comprising the mixture of different offshore reservoir oils in integrated oil production such as reservoir systems under high pressure, or reservoirs rich in carbon dioxide and with a mix of inlet streams from reservoirs under reinjection of production gases. Both scenarios are possible in Brazilian offshore projects, thus, requiring the development of methods to assess whether rigorous compositional modeling is critical to a project, affecting reservoir contour conditions and reservoir management significantly.

Stream junctions occur at variable temperatures and flow rates throughout the production period. The exit stream is usually not characterized experimentally. For this reason, in some cases, such as mixing very different fluids, Black-Oil based fluid mixing volumetric estimates might differ from compositional estimates and from real values. If Black-Oil model estimates are used for the gathering system design, they can affect volumetric distribution of phases and the flow rate of the resulting stream. As a result, potential deviations

from expected flow rates and well-head pressures can affect reservoir contour conditions directly. Whenever these contour conditions change considerably, the profitability is affected negatively as reservoir management does not reflect reservoir conditions. Ultimately, it hinders final oil recovery and reduces cash flow.

To address this issue, this paper implements three proxy methods for a compositional mixture of streams. The proposed methodology is valid for high-pressure mixtures at non-critical conditions and not yet subject to reinjection of production gases. As such, subsequent works can automate the compositional modeling of the resulting stream, allowing one, through the use of closed-loop reservoir management and Integrated Production Management (IPM), to determine this mixing impact on reservoir contour conditions and its subsequent impact on reservoir drainage strategy.

To develop a methodology to assess the impact of a rigorous compositional modeling on IPM and reservoir development, we start by implementing and evaluating simpler proxy compositional methods. These methods are used for mixed streams based on Equations of State (EoS), when different inlet reservoir streams are mixed in integrated oil production and reservoir systems.

The first step for the proposed closed-loop composition-based workflow is to reproduce and evaluate different methods. The aim of this work is to improve reservoir design as well as the structuring, planning, and operationalizing the reservoir through the improved exit-stream modeling from multiple inlet streams at a definite junction section (for instance, separator or subsea manifold).

The methods described in this work facilitate the future evaluation of the impact of compositional methods on the contour conditions of coupled reservoirs and production system models. A more rigorous representation of the phase behavior of the exit stream will allow future work to assess whether this behavior significantly impacts the reservoir production strategy. Measuring the influence of the modeling impact on optimal reservoir parameters of the integrated reservoir and production system also becomes possible.

The main objective of this work is to implement, validate, and discuss the reproducibility and limitations of a mixed-composition petroleum stream method based on an EoS and PVT data combined with concepts proposed by **Carpio (2012)**.

A secondary objective is to check the consistency of the parameters that affect the capability of an EoS to reproduce the physical behavior of the hypothetical mixture constituents separately.

2. LITERATURE REVIEW

2.1 Standard EoS PVT characterization and consistency checks required for this study

Standard EoS PVT characterization is an essential tool for the mixed-composition petroleum stream methods evaluated in this work. PVT characterization can be improved with experimental data using various different fine-tuning techniques matched with the experiments to improve accuracy associated to inherent experimental errors and to reproduce the measured behavior. First, however, data must be checked for consistency. To the best of our knowledge, there are no rigorous classical methods for thermodynamic consistency for petroleum containing heavy fractions, developed using the Gibbs–Duhem equation for petroleum, like those for pure components and other types of mixtures from **Won and Prausnitz (1973)**.

Won and Prausnitz (1973) proposed an integrated thermodynamic consistency check for high pressures. Since Won and Prausnitz studies, validating data through thermodynamic consistency checks have become increasingly important in the thermodynamic sciences. For every certified databank of pure component properties and chemical process engineering, these tests are required and constitute industry's best practices standard. **Van Ness (1995)** theoretically reviewed the foundations of some thermodynamic consistency checks and confirmed that an integrated area test is suitable to discard inconsistent and failed data, but, since it presents exceptions, it is insufficient to choose strictly

consistent data. Van Ness proved, with examples, that thermodynamically inconsistent data may be approved by area consistency check. He proposed the use of a direct, residue-based, thermodynamic consistency check as a second analysis. **Miklós et al. (1995)** performed a bibliographic review of available thermodynamic consistency checks and suggested a statistical approach to verify systematic errors in data without using an EoS. **Bertucco et al. (1997)** proposed a thermodynamic consistency check for high pressures and supercritical conditions. **Christiansen and Fredenslund (1975)** proposed a modified form of **van Ness' (1995)** consistency check, designed specifically for high pressures.

All the above tests require a general form for multicomponent hydrocarbon mixtures that have not yet been developed. Despite the identification of this gap, the present paper does not focus on that problem. Also, one of the most important terms for calculating multicomponent petroleum mixtures is obtaining the standard state properties of the heavy fraction. These standard state properties remain unknown and undetermined, creating a significant gap between the practical extension of this theory to petroleum EoS, which is yet to be explored. Considering the above reasons, there is a lack of available models for thermodynamic consistency tests for cases of petroleum mixtures containing heavy fractions. As this is not the focus of this study, only the three tests discussed in the following methodology section will be considered in the evaluation of the quality of mixing methods.

Another popular, but less effective, approach is to check the properties of the mixtures with correlations, based on works such as the one proposed by **Vazquez and Beggs (1977)**. We have discounted this method for the present study because most available correlations are developed for oil databases, mainly Black-Oil fluid. Unless the correlation is developed specifically for a databank of similar oils, this approach becomes case dependent and vulnerable to error. Nevertheless, there are other useful tools to detect and discard troublesome data like those described by **Pedersen and Christensen (2007)**, who called this procedure "Quality Control of PVT data."

Once PVT data are consistent with quality control, and fluid compositions are considered

representative of the material analyzed at the PVT laboratory, the EoS parameters are modified to improve fluid characterization. The need of consistent petroleum PVT data also applies to the description of a mixed-composition fluid formed from a pair of streams.

We used the concept of quality control from **Pedersen and Christensen (2007)**, including several criteria-based concepts on experimental data and the simulation of tests for comparison. These data are detailed in the methodology section.

2.2 EoS tuning and inverse problem statements

After PVT data validation, the EoS is calibrated. Many bibliographic references use methods to modify the cubic EoS parameters to improve fluid characterization and reproduce PVT experimental data. Most of these methods modify the properties of the heavy plus fraction (e.g., C_{7+} or C_{20+}), such as T_c , P_c , ω , and V_c ; or direct multipliers of EoS constants Ω_a and Ω_b ; and binary interaction coefficients of hydrocarbon pairs between methane and the heavy fraction. Some popular methods use the **Levenberg-Maarqdt (1963)** or **Dennis Jr. et al. (1981)** non-linear regression by least-squares to modify EoS parameters.

Judicious EoS tuning requires that only parameters with enough experimental uncertainties are used for regression, such as the critical parameters of a heavy fraction, P_c , T_c , and V_c , and binary interaction parameters (BIPs or k_{ij}). Tuning must not modify random or arbitrary parameters without a theoretical foundation, especially from hydrocarbon fractions other than those of the residual heavy fraction. EoS tuning and the resulting petroleum characterization are an inverse problem. The large number of different acceptable solutions with the same answer comes from different combinations of EoS parameter modifications possible within the uncertainty of properties. For this reason, the mixed-composition petroleum stream also becomes an inverse problem, with input variations with magnitudes depending, among other factors, on the uncertainty of heavy fraction parameters, on the engineering choice involved, on calibration tolerance, on heavy fraction critical properties correlation, and also on the mixing method of streams or mixing rules chosen.

Another factor of consistency is that one specific EoS regression solution cannot be considered realistic only because it has reproduced a given set of PVT experiments with good accuracy. EoS regression is a very complex process and undergoes subjective interpretations that make it sensitive and delicate. According to **Pedersen et al. (1988)**, often parameters “adjusted” for a specific property lead to unreliable predictions of other thermodynamic properties. Also, *“there is no reason a priori to believe that EoS parameters “adjusted” for PVT data of a certain composition, covering a limited temperature and pressure region will lead to enhanced K-values at other temperatures, pressures and compositions”* (**Pedersen et al., 1988**). Subcases of tuning combined without the specific mixture data may lead to unrealistic phase envelopes, unexpected or case-dependent formation of an additional phase, non-convergence of saturation pressures or non-physical behavior. Among the possible preventive measures to mitigate these problems, we selected two to apply to the current choice of mixed paired fluid.

2.3 Review of novel works relevant to this study

Ghassemzadeh and Charkhi (2016) used a novel fully dynamic approach to optimize a gas-lift system, using proxy models to minimize the resource usage objective function. While reviewing the literature, they found several works demonstrating that it is possible to apply the best scenario to enhance the production rate by integrating the whole reservoir and production system using, at some point, both integrated model and proxy methodologies. These works found by **Ghassemzadeh and Charkhi (2016)** try to incorporate the conventional criteria with the ongoing consideration of reservoir and well conditions.

According to **Ghassemzadeh and Charkhi (2016)**, although the integrated methods are very powerful and valuable, the computational costs are a disadvantage. To overcome these difficulties, there are two possible solutions: tabulate and interpolate the necessary parameters, or use proxy models.

Ghassemzadeh and Charkhi (2016) explained that proxy techniques, of increasing popularity in

recent years, are simplified models of a highly complicated process. Although their high computational efficiency is proven, using proxy models to optimize production is rare, although increasingly common in other fields of petroleum engineering.

Not every proxy model involves large amounts of data training, or employs artificial intelligence and optimization. Nevertheless, all of the models consider a specific, strict answer, and assume some kind of simplification on the overall framework. In light of this, there is an important distinction between the proxy approach employed by **Ghassemzadeh and Charkhi's (2016)**, in the works they reviewed, and in this current work. The present work integrates oil stream phase equilibrium consolidated complex algorithms, but employs a single simplification on the compositional framework, so the reservoir model and production system remain unaltered. Other kind of models could be defined as proxy models, such as the Inflow Performance Relationships and Reservoir surface response models, but this work does not use these models or perform optimizations.

3. METHODOLOGY

3.1 Assumptions and limitations of the proposed method

To achieve our objectives, input data is the standard measurements of oil phase behavior of real PVT for unmixed, separate samples of a hypothetically combined pair of real fluids from **Whitson and Brulé (2000)** and from **Pedersen et al. (1983)**. The data are input for a PVT simulator regression and complimentary modeling. Then, the results obtained are compared with the results of **Carpio (2012)** for a subsea manifold equimolar exit stream, with a mean exit temperature of 98.6 °C and an exit pressure of 100 Kgf/cm². This was modeled through two of the four mixing methods proposed by **Carpio (2012)**, and an additional variation of one of these methods based on the simpler **Kay's (1936)** combination rule (defined as reference method).

The present work makes the following considerations:

- Each set of PVT data is obtained at the same temperature as the reservoir and the mixing temperature for a subsea manifold is the exit stream average;
- The data from the resulting stream is unknown;
- The mixed stream has no significant water and no chemical reactions;
- There are no excess property models in the stream mixture methods;
- The equilibrium time is instantaneous, since it is based on flash expansions physics;
- There is an absence of deposition phenomena related to flow assurance issues that could affect phase behavior.

This work is unable to forecast solid formation (since it would need more robust models of PC-SAFT EoS or CPA EoS and additional data), multiphasic systems (VLLE) formation, or dynamic miscibility because of the simplifications assumed for the methods. Improvements can be made in subsequent works to address these issues, but these are not the focus of the present study.

3.2 Input data

PVT data were obtained from the literature for two samples: A light oil with a density of 40.5° API from **Whitson and Brulé (2000)**, and another light oil with a density of 42° API from **Pedersen et al. (1983)**. Tables 1 and 2 show the complete oil compositions. Table 3 shows heavy end oil properties.

Of the three data set of experiments used from **Whitson and Brulé (2000)**, the first was for differential liberation, the second a constant compositional expansion, and the third a complete oil chromatography up to C₂₀₊ and the associated heavy fraction characterization.

Out of the three from **Pedersen et al. (1983)**, the first was for differential liberation; the second, a constant composition expansion and saturation pressure determination; and, the third, a complete oil chromatography up to C₂₀₊ (with TBP distillation data) and the associated heavy fraction characterization.

Table 1. Oil Composition from **Whitson and Brulé (2000)** 40.5 °API, GOR 854 ft³/bbl, T= 102.78° C.

Component	Sample (%)	Component	Sample (%)
N ₂	0.16	C ₉	2.16
CO ₂	0.91	C ₁₀	2.12
C ₁	36.47	C ₁₁	1.85
C ₂	9.67	C ₁₂	1.61
C ₃	6.95	C ₁₃	1.40
I-C ₄	1.44	C ₁₄	1.23
N-C ₄	3.93	C ₁₅	1.08
I-C ₅	1.44	C ₁₆	0.95
N-C ₅	1.41	C ₁₇	0.85
C ₆	4.33	C ₁₈	0.76
C ₇	2.88	C ₁₉	0.70
C ₈	3.15	C ₂₀₊	12.55

Table 2. Oil Composition from **Pedersen et al. (1983)** 42 °API, GOR 132.9 m³/m³, T= 92.8° C.

Component	Sample (%)	Component	Sample (%)
N ₂	0.56	C ₉	2.14
CO ₂	3.55	C ₁₀	1.94
C ₁	45.33	C ₁₁	1.62
C ₂	5.48	C ₁₂	1.47
C ₃	3.70	C ₁₃	1.69
I-C ₄	0.70	C ₁₄	1.62
N-C ₄	1.65	C ₁₅	1.59
I-C ₅	0.73	C ₁₆	1.30
N-C ₅	0.87	C ₁₇	1.11
C ₆	1.33	C ₁₈	1.26
C ₇	2.73	C ₁₉	1.07
C ₈	3.26	C ₂₀₊	13.32

Table 3. Heavy Fraction Original Properties.

	Sample (%)	Molecular Weight	Density (g/cm ³) 15°C - 1 atm
C ₂₀₊ (A-1)	12.55	325	0.914
C ₂₀₊ (A-2)	13.32	421	0.875

Table 4 shows the constant composition expansion used for **Pedersen et al. (1983)**.

Table 4. **Pedersen et al. (1983)** Constant Composition Expansion.

P (psi _g)	P (psi)	V _{rel}
5000	5014.69	0.9639
4500	4514.69	0.9703
4000	4014.69	0.9771
3500	3514.69	0.9846
3000	3014.69	0.9929
2900	2914.69	0.9946
2800	2814.69	0.9964
2700	2714.69	0.9983
2620	2634.69	1.000*
2605	2619.69	1.0022
2591	2605.69	1.0041
2516	2530.69	1.0154
2401	2415.69	1.035
2253	2267.69	1.0645
2090	2104.69	1.104
1897	1911.69	1.1633
1698	1712.69	1.2426
1477	1491.69	1.3618
1292	1306.69	1.5012
1040	1054.69	1.7802
830	844.69	2.1623
640	654.69	2.7513
472	486.69	3.7226

*P_{sat}

Table 5 shows the Differential Liberation used for **Pedersen et al. (1983)**.

The constant composition expansion employed for **Whitson and Brulé (2000)** oil is displayed at Table 6.

Table 7 shows the Differential Liberation employed for **Whitson and Brulé (2000)**.

Table 5. Pedersen et al. (1983) Differential Liberation.

P (bar)	P (kPa)	B_o (m ³ /m ³ STD)	γ (g/cm ³)	Rs	μ_o (cP)	Z
	38980				0.469	
	34770				0.447	
	30210				0.425	
	28590				0.413	
	27560				0.406	
	27450				0.404	
267.9	26790	1.396	0.718	132.9	*	
	26090				0.424	
239.6	23960	1.365	0.726	119.4		0.889
225.1	22510				0.484	
209.4	20940	1.329	0.735	103.7		0.888
179.8	17980	1.297	0.744	89.6		0.876
150.0	15000	1.265	0.753	75.7		0.880
143.2	14320				0.651	
112.9	11290	1.227	0.765	58.6		0.894
107.9	10790				0.761	
72.4	7240	1.186	0.778	40.4		0.904
64.7	6470				0.911	
35.8	3580				1.051	
33.1	3310	1.144	0.791	21.9		0.949
1.0	100	1.067	0.816			

* P_{sat}

Table 6. Whitson and Brulé (2000) Constant Composition Expansion.

P (psi _g)	P (psi)	V _{rel}	P (psi _g)	P (psi)	V _{rel}
5000	5014.69	0.9639	2401	2415.69	1.0350
4500	4514.69	0.9703	2253	2267.69	1.0645
4000	4014.69	0.9771	2090	2104.69	1.1040
3500	3514.69	0.9846	1897	1911.69	1.1633
3000	3014.69	0.9929	1698	1712.69	1.2426
2900	2914.69	0.9946	1477	1491.69	1.3618
2800	2814.69	0.9964	1292	1306.69	1.5012
2700	2714.69	0.9983	1040	1054.69	1.7802
2620	2634.69	1.000*	830	844.69	2.1623
2605	2619.69	1.0022	640	654.69	2.7513
2591	2605.69	1.0041	472	486.69	3.7226
2516	2530.69	1.0154			

* P_{sat}

3.3 Equations used in this work

3.3.1 Peng Robinson (1978) Equation of State

The three mixing methods used in this current work all use the two-constant EoS from Peng and

Robinson (1978), because of its widespread use, and known capabilities. The PR EoS (1976) is given by:

$$p = \frac{RT}{v-b} - \frac{a}{v(v+b)+b(v-b)} \quad (1)$$

Table 7. Whitson and Brulé (2000) Differential Liberation.

P (psig)	P (psi)	B _o (m ³ /m ³ STD)	γ (g/cm ³)	Rs	Mo (cP)	B _g (m ³ /m ³ STD)	Mgas (cP)	drgas	Z
5000	5014.69				0.450				
4500	4514.69				0.434				
4000	4014.69				0.418				
3500	3514.69				0.401				
3000	3014.69				0.385				
2800	2814.69				0.379				
2620	2634.69	1.600	0.6562	854.00	0.373	*			
2350	2364.69	1.554	0.6655	763.00	0.396	0.00685	0.0191	0.825	0.846
2100	2114.69	1.515	0.6731	684.00	0.417	0.00771	0.018	0.818	0.851
1850	1864.69	1.479	0.6808	612.00	0.442	0.00882	0.0169	0.797	0.859
1600	1614.69	1.445	0.6889	544.00	0.469	0.01034	0.016	0.791	0.872
1350	1364.69	1.412	0.6969	479.00	0.502	0.01245	0.0151	0.794	0.887
1110	1124.69	1.382	0.7044	416.00	0.542	0.01552	0.0143	0.809	0.903
850	864.69	1.351	0.7121	354.00	0.592	0.02042	0.0135	0.831	0.922
600	614.69	1.320	0.7198	292.00	0.654	0.02932	0.0126	0.881	0.941
350	364.69	1.283	0.7291	223.00	0.783	0.05065	0.0121	0.988	0.965
159	173.69	1.244	0.7382	157.00	0.855	0.10834	0.0114	1.213	0.984
0	14.69	1.075	0.7892	0.00	1.286		0.0093	2.039	

*P_{sat}

Or, in terms of factor Z:

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

$$-(AB - B^2 - B^3) = 0$$

$$Z_c = \frac{p_c v_c}{RT_c} = 0.3074 \quad (3)$$

The constants of this PR EoS are given by:

$$a = \Omega_a^0 \frac{R^2 T_c^2}{p_c} \alpha \quad (4)$$

$$\text{Where } \Omega_a^0 = 0.45724$$

$$b = \Omega_b^0 \frac{RT_c}{p_c} \quad (5)$$

$$\text{Where } \Omega_b^0 = 0.07780$$

$$\alpha = [1 + m(1 - T_r^{0.5})]^2 \quad (6)$$

$$m = 0.37464 + 1.54226 \omega - 0.26992 \omega^2 \quad (7)$$

Robinson et al. (1978) and Robinson and Peng (1978) proposed a modified expression for “m” that is recommended for heavier compounds ($\omega > 0.49$).

$$m = 0.3796 + 1.485 \omega - 0.1644 \omega^2 + 0.01667 \omega^3 \quad (8)$$

Therefore, the fugacity expression is given by:

$$\ln \frac{f}{p} = \ln \phi = Z - 1 - \ln(Z - B) - \frac{A}{2\sqrt{2}B} \ln \left(\frac{Z + (1 + \sqrt{2})B}{Z - (1 - \sqrt{2})B} \right) \quad (9)$$

$$\ln \frac{f_i}{y_i P} = \ln \phi_i = \frac{B_i}{B} (Z-1) - \ln(Z-B) + \frac{A}{2\sqrt{2}B} \left(\frac{B_i}{B} - \frac{2}{A} \sum_{j=1}^N y_j A_{ij} \right) \ln \left(\frac{Z + (1+\sqrt{2})B}{Z - (1-\sqrt{2})B} \right) \quad (10)$$

Traditional mixing rules (Equations 11 to 13) were used to derive the expression for multicomponent expression. In this EoS model, the traditional quadratic mixing rule is used for A, and a linear mixing rule is used for B. For one vapor with composition y_i , is given by:

$$A = \sum_{i=1}^N \sum_{j=1}^N y_i y_j A_{ij} \quad (11)$$

$$B = \sum_{i=1}^N y_i B_i \quad (12)$$

$$A_{ij} = (1 - k_{ij}) \sqrt{A_i A_j} \quad (13)$$

3.3.2 Jhaveri and Youngren (1988) volume translation for Peng Robinson's Equation of State (1978)

Volume translation considerably improves the main problem: poor volumetric forecast for liquids of the two-constant PR (1978) EoS. A simple correction term is applied to the molar volume calculated by the EoS:

$$v = v^{EOS} - c \quad (14)$$

Where, v = corrected molar volume, v^{EOS} = EoS calculated volume, and c = compound specific constant. This volume variation is, in practice, equivalent to adding a third constant to the EoS, but is special because equilibrium conditions remain unaltered.

Multicomponent vapor-liquid equilibrium (VLE) is unaltered through the introduction of a correction term as an average of molar fractions:

$$v_L = v_L^{EOS} - \sum_{i=1}^N x_i c_i \quad (15)$$

$$v_v = v_v^{EOS} - \sum_{i=1}^N y_i c_i \quad (16)$$

Where, v_L^{EOS} and v_v^{EOS} = molar volumes

calculated by EoS for liquid and vapor, respectively; x_i and y_i = liquid and vapor composition, respectively; and c_i = volume shift parameters dependent on component.

When volume shift is introduced in the EoS for mixtures, the resulting expressions of fugacity are:

$$(f_{vi})_{modified} = (f_{vi})_{original} \exp \left(-c_i \frac{P}{RT} \right) \quad (17)$$

$$(f_{Li})_{modified} = (f_{Li})_{original} \exp \left(-c_i \frac{P}{RT} \right) \quad (18)$$

That implies that fugacity ratios are unaltered by volume shifts

$$\left(\frac{f_{Li}}{f_{vi}} \right)_{modified} = \left(\frac{f_{Li}}{f_{vi}} \right)_{original} \quad (19)$$

In fact, c_i can be determined by equalizing the EoS to any density value at one specific pressure and temperature. Jhaveri and Youngren write c_i as a ratio $s_i = c_i/b_i$, suggesting the following equation for heavy ends C_{7+} :

$$s_i = c_i/b_i = 1 - A_0/M_i^{A_1} \quad (20)$$

3.3.3 Kay and Lee-Kesler's mixing rules and definition of combination rules

Carpio (2012) discussed the definitions of a mixing rule and a combination rule. According to Carpio (2012), a mixing rule is a term within the cubic EoS that is used to calculate the multicomponent A and B terms. Combination rule is an equation of numerical similarity to the mixing rule, but used to average lumped components properties and to average EoS parameters in his methods.

The most simple and common mixing rule is offered by Kay (1936), used as a combination rule as well as a basis for lumping in most commercial software:

$$\theta_i = \frac{\sum_{i \in I} z_i \theta_i}{\sum_{i \in I} z_i} \quad (21)$$

Where θ = any property (T_c , p_c , ω or M) and z_i =

Figure 1. Scanavini et al.'s (2013) tuning method (PR EOS), adapted from original flowchart, reproduced with permission of authors.

original molar fraction for the components ($i=1,\dots,l$) that form the pseudo component l .

Based on arguments presented by Chueh and Prauznitz (1967), Lee-Kesler (1975) proposed the following mixing rules, also used by Carpio (2012) as combination rules:

$$v_{cl} = \frac{\left[\frac{1}{8} \sum_{i \in l} \sum_{j \in l} z_i z_j \left(v_{ci}^{1/3} + v_{cj}^{1/3} \right)^3 \right]}{(\sum_{i \in l} z_i)^2} \quad (22)$$

$$T_{cl} = \frac{\left[\frac{1}{8 v_{cl}} \sum_{i \in l} \sum_{j \in l} z_i z_j (T_{ci} T_{cj})^{1/2} \left(v_{ci}^{1/3} + v_{cj}^{1/3} \right)^3 \right]}{(\sum_{i \in l} z_i)^2} \quad (23)$$

$$\omega_l = (\sum_{i \in l} z_i w_i) / (\sum_{i \in l} z_i) \quad (24)$$

$$z_{cl} = 0.2905 - 0.085 \omega_l \quad (25)$$

$$p_{cl} = \frac{Z_{cl} R T_{cl}}{v_{cl}} \quad (26)$$

3.4 Equation of State tuning method

The regression method for the three mixing methods used in this work was adapted from both the methods proposed by Scanavini et al. (2013) and by Carpio (2012). Figure 1 displays the method.

3.5 General Methodology

The methodology section is divided in three parts: (1) Quality Control of compositional mixture modeling, (2) compositional mixture of two fluids, and (3) tuning the EoS prior to or simultaneously with the mixing method. The first section defines

quality checks for the second section, while the third comprises finer details not covered in the second section.

3.5.1 Methodology for quality control of compositional mixture modeling

As previously stated in the Literature Review section, **Pedersen and Christensen (2007)** described concepts of quality control including several criteria based on experimental data and comparison of simulated tests. Two of these criteria, plus an additional concept from **Li et al. (1984)** can be useful to detect inconsistent compositional mixtures modeling descriptions. They are discussed below:

1- Saturation pressure at the reservoir temperature must be lower than reservoir pressure, or the sample is not representative. This concept will be further extended to describe mixtures. When the highest reservoir pressure of either reservoir sources, at the temperature of the mixed stream, is lower than the saturation pressure (BBP in this case) of the mixed stream, we suggest that the mixture modeling is inconsistent. The saturation pressure of the reservoirs must be higher than the mixed stream.

2- **Li et al. (1984)** suggests that the use of phase diagrams and quality lines are useful tools to verify the quality of the lumped fluid description (as also supported by **Whitson and Brulé, 2000**). Extending this concept, we propose that an adequate simulated description of mixtures should behave in the following way: an acceptable mixture proxy model should have a 'coherent' phase diagram according to engineering judgment, when compared to both the original mixture constituent phase diagrams. This means that the phase envelope will be well posed, with behavior compatible with the expected type of petroleum mixture.

3- **Pedersen and Christensen (2007)** proposed a refined version of Hoffman's plot (**Hoffman et al., 1953**) by calculating the ratio of the component mole fractions in the separator gas and oil phases. They state that by plotting the logarithmic K-factor of simulated data against experimental data, the plotted points should be very close or identical to a straight line, especially in the first segment of the plot. Extending this concept, we suggest that a consistent simulated description of mixture

behaves in the following way: the simulated K-factor plot at mixing vessel conditions is located within the experimental K-factors plot for both constituent streams, for cases with only two predicted phases. Unfortunately, for the sample pair studied, there is no gas and oil compositional information allowing the calculation of the experimental K-values of Hoffman plots for any mixture constituents at mixture conditions. Incomplete information is available for only one of the samples about the separator gas composition at conditions different to those specified and limited to only below 7 carbon fractions, so this test will not be applied for the mixture chosen for this work.

3.5.2 Methodologies for compositional mixture of two fluids

This section describes the core part of the methodology, while the following section details the tuning method. Figure 2 describes the main steps for the proposed general workflow.

Three main methods to model the composition of mixed streams were compared (KCR, LKCR, and Weaved):

Method 1 – Combination rules based on Kay's mixing rule

Figures 3 and 4 illustrate the simplified flowcharts for this and the following method. Figure 3 details specific steps for Methods 1 and 2.

The following steps comprise mixing method 1:

1. All PVT and composition data and for each mixture component are input on separate files.

2. Heavy fraction for each fluid (C_{20+}) is subject to the Gamma function splitting (**Whitson, 1983**), following **Whitson and Brulé (2000)** recommended minimum number of pseudo components (5 for fraction C_{20+} , according to **Whitson and Brulé, 2000**). For this lumping we used the Gaussian quadrature (**Whitson, 1983**), critical properties correlations from **Lee-Kesler (1975)**, the correlations for SG, T_b and MW from **Twu (1984)**, and the Whitson Characterization Factor from (**Whitson, 1983**). The molecular weight interval was divided by matching weight fractions, using the sum of squared residuals to measure the discrepancy between data and estimated models.

Figure 2. General workflow.

3. Each fluid is, then, tuned individually for an initial complete characterization. Then, any necessary adjustments are made to fit within determined tolerance levels.

4. Subsequently, rheology is calibrated individually. Then, necessary adjustments are made to fit within determined tolerance levels.

5. Molar weighted Kay's combination rules are used in a spreadsheet to calculate mixture compositions, critical properties, molecular weight, acentric factors, volume shifts, Z , densities, boiling temperatures, parachor parameters, binary interaction parameters, and rheological coefficients.

6. PVT models are updated for mixture conditions, P-T phase diagrams are plotted, and consistency is checked.

7. Simulation of flash expansion at P and T of forecast (target reservoir temperature, target pressure, molar proportion feed) is performed. Check for results reproducibility.

Method 2 – Combination rules based on Lee-Kesler's (1975) mixing rule

Method 2 follows the same steps as for Method 1, with a single important difference in step 5: The critical properties of the mixture and the compressibility constant (P_c , T_c , Z and V_c) are calculated through Lee-Kesler's (1975) combination rules. This method and Method 3 are the original version of the methods described by Carpio (2012). Any other mixing rule could be chosen as a combination rule, but for the sake of brevity, we chose only this one and Kay's (1936).

Method 3 – Simultaneous regression: weaved

The flowchart summarizing the steps of this method is shown above in Figure 3. Figure 4 shows the steps specific to Method 3.

The following steps comprise this mixing methodology:

1. All data and compositions for each component in a mixture are input on two fields of the same file

Figure 3. Workflow with mixing rule for compositional mixture.

Figure 4. Method of compositional mixture with simultaneous regression workflow.

(main stream and enriched stream), PVT measurements are added to several appendixes of the same file (as if belonging to the same mean

characterization). The experiments are specified on the symmetrical feed ratio desired for each experiment and associated fluid (equimolar in this

case). The generated heavy fractions are inserted at this stage, as described in step 2 of the previous method, and combined as described in step 4 of this method, respecting the feed ratio.

2. A combined stream of the paired fluids is tuned simultaneously to the highest degree possible for an initial complete characterization (different fluid experiments for the same file, producing average simulated experiments), resulting in the same tuned parameters and values for both fluids. Necessary adjustments are made to minimize deviation, usually less restrictive than the conventional tuning.

3. Rheology is then calibrated simultaneously for both fluids. Necessary adjustments are made to minimize deviation.

4. Kay's molar weighted combination rules are used in a calculation spreadsheet for only compositional mixtures and not heavy fraction properties, with a single and important difference in this Step: the heavy fractions are not lumped or combined but are kept separate. We advise keeping the binary interaction parameters among heavy fractions at zero.

5. PVT model are updated for mixed feed composition and heavy fractions, P-T phase diagrams are plotted and, consistency checked.

6. Simulation of flash expansion at P and T for forecast (target reservoir temperature, target pressure, molar proportion feed). Check for reproducibility of results.

3.5.3 Methodology for EoS tuning prior to or simultaneously with mixing method

We used pure component data from the literature for some EoS components, as well as theoretical correlations to predict critical properties. The methodology consists of 3 different steps and is explained below:

Step 1 – PVT and Critical Properties Calculation

PVT data were obtained from the literature for two samples: A light oil with a density of 40.5° API from **Whitson and Brulé (2000)**, and another light oil with a density of 42° API from **Pedersen et al. (1983)**, previously mentioned in the input data section.

The viscosity values were calibrated for PVT models using the **Pedersen et al. (1988)** corresponding states model and were not reproduced. The three data sets of experiments used from **Whitson and Brulé (2000)** were for differential liberation, constant composition expansion (CCE), with determined bubble point and complete oil chromatography up to C₂₀₊; and associated heavy fraction characterization. The three sets used from **Pedersen et al. (1983)** were for, differential liberation, CCE with determined saturation pressure and complete oil chromatography up to C₂₀₊ (with TBP distillation data), and the associated heavy fraction characterization.

This step has two possible paths, depending on the source of characterization data: a software databank for pure compounds is used if the data is from the light fraction composition determined through gas chromatography (GC). For GC components heavier than C₅, different from the heavy fraction, the database is identical to those published by **Katz and Firozaabadi (1978)**. However, if the data is from a TBP characterization or indefinite residual heavy fraction, heavier than hexane, it is characterized through osmometric vapor pressure measurements. In this case, critical properties correlations are used instead, using input data for fraction densities, average molecular weights, and distillation fraction boiling points. The correlations used in this study were **Lee-Kesler (1975)** for critical properties and **Lee-Kesler (1975)** for Acentric Factor (ω).

3.6 Component cluster choice

28 Components: N₂; CH₄; CO₂; C₂H₆; C₃H₈; iso-butane; n-butane; iso-pentane; n-pentane; C₆; C₇; C₈; C₉; C₁₀; C₁₁; C₁₂; C₁₃; C₁₄; C₁₅; C₁₆; C₁₇; C₁₈; C₁₉; PC-1-C₂₀₊, PC-2-C₂₀₊, PC-3-C₂₀₊, PC-4-C₂₀₊, PC-5-C₂₀₊.

4. RESULTS AND DISCUSSION

The phase equilibria calculation provides the volumes of oil and gas for every part of the simulated coupled system, and also for each phase component and the molar composition of the stream. This information is relevant to the numerical simulation of the composition. Any significant variability in the mixed stream properties will directly impact reservoir management.

Figure 5. Phase diagram for mixture components from [Pedersen et al. \(1983\)](#) (quality lines are suppressed in this graphic).

Figure 6. Phase diagram for mixture components from [Whitson and Brulé \(2000\)](#) (quality lines and critical points are suppressed in this graphic).

4.1 Input fluid phase behavior

Figures 5 and 6 show the phase envelope of the entry mixture components.

Tables 8 and 9 show an example of characterization input.

4.2 Comparison of results for the three methods

After simulation, the base (proposed) method using the combination rules from [Kay \(1936\)](#) was

compared to Methods 1 and 2 assessing: phase volume fractions, molar fractions, molecular weights, densities, and phase viscosities.

Each set of PVT data relative to the chosen mixed pair of streams was initially adjusted for 28 components (23 usual components + 5 heavy fraction split components) through non-linear regression.

Table 8. Pedersen et al. (1983) oil EoS parameters.

Component	P_c (atm)	T_c (K)	ω	MW	Volume Shift Coefficient (L/°C)	Volume Shift T_{ref} (°C)	Z	Vc (l/mol)	Vc (viscosity)	Ω_a	Ω_b	SG	Tb (°C)	Parachor
CO ₂	72.8	304.2	0.225	44.01	0.14739	-0.007842	15.55556	0.27	0.094	0.4572	0.078	0.82	-78.45	78
N ₂	33.5	126.2	0.04	28.01	-0.16632	-0.000358	15.55556	0.29	0.09	0.4572	0.078	0.81	-195.75	41
CH ₄	45.4	190.6	0.008	16.04	-0.14924	-0.001152	15.55556	0.29	0.099	0.4572	0.078	0.3	-161.45	77
C ₂ H ₆	48.2	305.4	0.098	30.07	0.13549	-0.010678	15.55556	0.28	0.148	0.4572	0.078	0.36	-88.65	108
C ₃ H ₈	41.9	369.8	0.152	44.1	-0.06893	0.002715	15.55556	0.28	0.203	0.4572	0.078	0.51	-42.05	150.3
IC ₄	36	408.1	0.176	58.12	-0.07742	0.001133	15.55556	0.28	0.263	0.4572	0.078	0.56	-11.85	181.5
NC ₄	37.5	425.2	0.193	58.12	-0.06369	0.000764	15.55556	0.27	0.255	0.4572	0.078	0.58	-0.45	189.9
IC ₅	33.4	460.4	0.227	72.15	-0.05951	-0.000133	15.55556	0.27	0.306	0.4572	0.078	0.63	27.85	225
NC ₅	33.3	469.6	0.251	72.15	-0.03726	0.000079	15.55556	0.27	0.304	0.4572	0.078	0.63	36.05	231.5
FC ₆	32.46	507.5	0.275	86	-0.05917	-0.000818	15.55556	0.27	0.344	0.4572	0.078	0.69	63.85	250.11
C ₇	35.33	542.46	0.284	89.9	-0.04749	-0.000852	15.55556	0.27	0.336	0.4572	0.078	0.76	85.31	261.2
C ₈	31.4	577.98	0.281	103.2	0.01487	-0.000191	15.55556	0.26	0.4	0.4572	0.078	0.78	112.66	298.51
C ₉	29.48	605.2	0.373	117.7	-0.00553	-0.000275	15.55556	0.27	0.436	0.4572	0.078	0.8	142.1	338.29
C ₁₀	25.87	627.87	0.431	133	0.03106	-0.000285	15.55556	0.26	0.506	0.4572	0.078	0.8	167.95	379.27
C ₁₁	23.36	648.1	0.482	147	0.06147	-0.000289	15.55556	0.26	0.568	0.4572	0.078	0.8	190.69	415.87
C ₁₂	21.95	670.88	0.525	160	0.08352	-0.000293	15.55556	0.25	0.617	0.4572	0.078	0.82	213.2	449.08
C ₁₃	20.88	700.69	0.497	172	0.12099	-0.00026	15.55556	0.25	0.673	0.4572	0.078	0.83	234.59	479.08
C ₁₄	19.6	713.34	0.608	186	0.12548	-0.000291	15.55556	0.25	0.713	0.4572	0.078	0.84	255.91	513.27
C ₁₅	18.25	730.06	0.654	200	0.14691	-0.000289	15.55556	0.24	0.771	0.4572	0.078	0.85	275.12	546.6
C ₁₆	17.28	746.82	0.695	213	0.16505	-0.000286	15.55556	0.24	0.821	0.4572	0.078	0.86	293.33	576.78
C ₁₇	15.35	758.9	0.761	233	0.19272	-0.000284	15.55556	0.24	0.916	0.4572	0.078	0.85	312.9	621.75
C ₁₈	14.48	772.34	0.803	247	0.20914	-0.000281	15.55556	0.24	0.973	0.4572	0.078	0.86	328.96	652.19
C ₁₉	14.1	787.53	0.832	258	0.21986	-0.000278	15.55556	0.23	1.007	0.4572	0.078	0.87	344.1	675.5
HYP01	12.6	812.39	0.827	288.3	0.23695	-0.000263	15.55556	0.23	1.15	0.4572	0.078	0.86	368.93	736.93
HYP02	10.55	868.84	1.008	359.3	0.26956	-0.000262	15.55556	0.23	1.38	0.4572	0.078	0.89	434.17	865.1
HYP03	9.43	1038.3	1.203	592.9	0.11468	-0.000261	15.55556	0.25	1.708	0.4572	0.078	0.93	530.54	1047.42
HYP04	7.12	1055.2	1.378	851.4	0.0979	-0.000257	15.55556	0.25	2.027	0.4572	0.078	0.97	639.95	1172.83
HYP05	5.53	687.04	1.518	1263	-0.50571	-0.000711	15.55556	0.32	2.32	0.4572	0.078	1.03	761.82	951.11

Table 9. Pedersen et al. (1983) oil EoS non-HC/HC pairs K_{ij} parameters (Hydrocarbon pairs employ Chueh and Prausnitz's coefficient of 1.586)

	CO2	N2	CH4	C2H6	C3H8	IC4	NC4	IC5	NC5	FC6	C7	C8	C9	C10	C11	C12	C13	C14	C15	C16	C17	C18	C19	HYP01	HYP02	HYP03	HYP04	HYP05
CO2	0	0	0.11	0.13	0.13	0.1	0.12	0.1	0.12	0.1	0.1	0	0	0	0	0	0	0	0	0	0	0	0	0.4	0.4	0.4	0.32	0.2
N2	0	0	0.03	0.01	0.09	0.1	0.1	0.1	0.11	0.1	0.1	0	0	0	0	0	0	0	0	0	0	0	0	0.12	0.12	0.12	0.12	0.12

Figure 7. Comparison of the phase volume fraction for properties of each mixture, for each method, for the liquid (oil phase, blue) and gaseous phases (red).

Figure 8. Comparison of the phase molar fraction for the properties of each mixture, for each method, for the liquid (oil phase, blue) and gaseous phases (red).

Experimental data were unavailable for the final real mixture. As a base method for this work, we chose the simple molar average of tuned properties from Kay's combination rule. After flash simulation, the results from the generated properties were given in Figures 7 to 10. Figure 7 displays the volume fraction calculated for the properties of each mixture, for each method.

Figure 8, above, displays molar fractions calculated for the properties of each mixture for each method.

Figure 9 displays the molecular weight

calculated for the properties of each mixture, for each method.

Figure 10 displays the molecular weight calculated for the properties of each mixture, for each method.

Table 10 displays the relative percent differences for all methods.

Figures 7 to 10 show volume fractions, molar fractions, molecular weights and densities, a small but influential density, and molecular weight variability of the mixed fluid data compared

Figure 9. Comparison of the phase molecular weight for the properties of each mixture, for each method, for the liquid (oil phase, blue) and gaseous phases (red).

Figure 10. Comparison of the phase molar density for the properties of each mixture, for each method, for the liquid (oil phase, blue) and gaseous phases (red).

between methods. This data is sufficiently relevant to prompt further investigation of IPM and reservoir management studies. There are some smaller differences of data in relation to all methods considered, especially in columns

representing oil properties. For instance, oil volume fractions tended to vary by about 5%, and about 3% for gas volume fractions. Consequently, molar fractions tended to vary by about 2.5% for oil molar fractions, and 0.64% for gas molar

Table 10. Differences between properties generated at flash simulation, for both liquid and gas phases, for all methods studied, 24 components base description (base method is green, large discrepancies are highlighted in red).

T_{RAV} and P 100 kgf/cm ²	$P_{sat}(T_{RAV})$	Oil Volume Fraction	Gas Volume Fraction	Oil Mole Fraction	Gas Mole Fraction	Oil MW	Gas MW	Oil Density	Gas Density	μ Oil	μ Gas
Kay combination rules	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE
Lee-Kesler combination rules	FAILED	-5.07%	2.19%	-2.56%	1.86%	-0.71%	1.11%	2.35%	1.23%	-99.98%	30723.46%
Weaved Method	3.89%	-2.06%	3.02%	-0.34%	0.64%	-6.19%	-3.24%	-3.36%	-4.35%	11.14%	-0.62%

Table 11. Comparison of the saturation pressures for the properties of the mixture, for each method with maximum reservoir pressure, for consistency checks of proposed mixing methods (One description failed to converge, no fluid description passed the consistency check limits).

T_{RAV} and P_{100} kgf/cm ²	$P_{sat}(T_{RAV})$	Oil Volume Fraction	Gas Volume Fraction	Oil Mole Fraction	Gas Mole Fraction	Oil MW	Gas MW	Oil Density	Gas Density	μ Oil	μ Gas
Kay combination rules	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE	BASE
Lee-Kesler combination rules	FAILED	-5.07%	2.19%	-2.56%	1.86%	-0.71%	1.11%	2.35%	1.23%	-99.98%	30723.46%
Weaved Method	3.89%	-2.06%	3.02%	-0.34%	0.64%	-6.19%	-3.24%	-3.36%	-4.35%	11.14%	-0.62%

fractions. Also, molecular weights varied by 6.2% for oil, and over 3% for gas, while densities only varied by 3.4% for oil, and over 4% for gas. Finally, the forecasting for viscosities was unreliable with large differences, over 11% for the most consistent description.

Based on the consistency checks of saturation pressures, we concluded that the main cause for this disagreement was an inconsistent Method 2 with large deviations for most properties (based on Lee-Kesler mixing rule). The target saturation pressure for Method 2 did not converge, and caused the formation of a small volume in a third phase. Other consistency checks did not detect any larger discrepancies from the phase envelopes for this case. We believe that small discrepancies for properties, originating from a consistent description, will amplify any property deviations in future studies of Integrated Production

Management (IPM), and in closed-loop reservoir management. Therefore, in future studies, it will be possible to evaluate with two of the previous three tools (saturation pressures and phase envelopes). Through quality control checks, one will detect incorrect or unreliable descriptions, consequently, preventing strong modeling mismatches.

4.3 Discussion to discard data descriptions using quality check

In quality check number 1, the saturation pressures and maximum reservoir pressures of the two fluids comprising the mixture, at reservoir temperature, can be seen on Table 11. The consistency checks for every method are displayed for comparison with the reference method.

Table 11 also shows the calculated saturation pressures of each mixture, for every method. It is

Figure 11. Comparison of the phase envelopes for the properties of each mixture, each method, for each lumping scheme that converged (quality lines are suppressed in this graphic).

clear that Method 2 is inconsistent (saturation pressure did not converge) and also that no predicted saturation pressure for any method exceeded any reservoir pressure, so only the Kay-based and Weaved methods are initially consistent.

For quality check number 2, Figure 11 shows the phase envelopes.

There are no notable differences between the P-T diagrams. Since the third proposed quality check was non-viable for this study, the discussion will focus on the variability of forecasts for the properties of the mixture.

As stated by **Carpio (2012)**, in a previous study, the efficient use of both methods based on combination rules is related directly to the pair of mixed fluids and to the P and T conditions of the study. It is believed that fluids with greater volatility and at near-critical region conditions decrease the reliability of accurately forecasting the equilibrium of the mixture. The same concept is applicable to all the studied mixing methods. Because of this, simulating the fluid mixture is not possible using simple black-oil average additive properties for every composition, especially for lighter components and heavy fractions. The high variability of the inconsistent mixture method (Lee-Kesler based) was largely caused by the calculated (P_c , T_c , Z , and V_c) properties during the combining procedure.

Lastly, the most important analysis concerns the choice of mixing methods. Since the real answer is unknown, and two descriptions were successfully achieved passing two quality checks, the recommendation for future studies of Integrated Production Management (IPM) is to assess every studied mixing method to evaluate the impact on stream properties, reservoir contour conditions, reservoir management, and IPM.

Based on all the above observations, a further study of the impact of production systems on reservoir contour conditions might be useful using an automated black-oil method; the standard rigorous method and; two of the three proxy methods (Kay CR and Weaved) studied in this work. The variability of mixture properties from different methods resulted in a small but significant margin of deviation. Therefore, we suggest that when contour conditions are affected considerably, composition mix modeling may prove useful for

both IPM and closed-loop reservoir management to improve recovery and profit.

5. CONCLUSIONS

This work reviewed three proxy methods to simulate the compositional mixture of petroleum streams through EoS tuning and characterization (that are applicable to flow simulators). It also compared these methods in the absence of experimental mixture PVT data.

Behavioral analysis of mixed-composition base Method 1 (based on combinational rules, KMR) has shown to be a simple and practical solution to a critical problem. Method 2 (LKMR) had inconsistency issues which prevented further testing. We found Method 3 (Weaved) to be consistent.

There are small but significant variations in mixture properties (oil volume fractions, molar fractions, molecular weights, densities, and viscosities) that occur due to the inherent nature of this inverse problem. In some cases, a third phase forms making variability even more critical to IPM studies.

Compositional mixing can therefore be a source of uncertainty for both closed-loop Reservoir Management and IPM because of the variability of the properties in the mixture. If a mixing method is an influential factor on profit margins and it is not addressed, a systematic investigation is needed for closed-loop Reservoir Management, to mitigate losses and improve oil recovery.

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