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Non-Equilibrium Reservoir Simulation of Solvent-Steam Processes, Based on Mass and Heat Transfer Inside a Pore

Bitra Bayestehparvin, Jalal Abedi, and S. M. Farouq Ali, University of Calgary

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Abstract

Steam processes involving injection of a solvent with steam, such as CSS (Cyclic Steam Stimulation), SAGD (Steam-Assisted Gravity Drainage), and steam injection in isothermal mode are currently receiving a great deal of attention in Alberta. These combination processes are designed to reduce energy consumption and the emission of greenhouse gases over steam alone. Field results have been mixed and the original intent remains elusive. This paper addresses questions regarding the mechanism of solvent dissolution of bitumen and the expected improvement in oil recovery, if any, when a solvent is injected with steam.

The unique aspect of this work is that instantaneous phase equilibrium is not assumed, as is customary in numerical simulators for solvent-steam applications. Nor is partial equilibrium assumed based on an empirical factor or concept. Rather, equilibrium is based on an analytical model of the dissolution and mobilization of a drop of bitumen inside a pore by solvent and heat. The results of this part of the study show that solvent requires at least three times as long to mobilize bitumen as by heat conduction. Such a delay is implicit in the nature of diffusion and dispersion of a solvent. Several boundary conditions are tested for a spherical drop.

A new thermal compositional simulator with the single drop model was developed for this study and several thermal processes for solvent injection were investigated for non-equilibrium phase behaviour. The results were compared to the case of instantaneous equilibrium, confirming the reason for the lack of success with solvent injection.

The results and extensions of this work will be of great interest in heavy oil production because they serve to explain the expected performance and frequent lack of success in solvent-steam injection. Use of the developed simulator would make it possible to determine whether solvent injection is a good choice in a given situation.

Introduction

Published in 1856, Darcy's law is used to calculate the flow rate in a porous medium:

$$q_0 = \frac{kk_{ro}A\Delta P}{\mu_0 L} \quad (1)$$

where q_o is the flow rate of oil and k and k_{ro} are the absolute and relative permeabilities, respectively. The variables μ , A , and $\frac{\Delta P}{L}$ represent the viscosity, area open to flow, and pressure gradient, respectively.

Investigating the parameters in this equation shows that the only parameter that can be changed without limit is oil viscosity. This concept has inspired the development of numerous methods to reduce the viscosity of the oil in heavy oil and bitumen reservoirs. The oil viscosity can be reduced by either heat conduction/convection, as in steam-assisted recovery processes or solvent diffusion/dispersion as in solvent-based processes. Despite the advantages of steam-based processes, they require high energy consumption and water use/treatment and produce significant greenhouse gas emissions. It has been postulated that using solvent instead of steam to reduce oil viscosity could largely eliminate the energy losses inherent in steam processes. Following the first use of solvents in recovery processes in California (Jeffries-Harris and Coppel (1969)), alternative ways of using solvents have been developed. Although the idea dates back to 1960's, field results show large deviations from the analytical and numerical predictions.

One major shortcoming of the existing numerical models is the unrealistic assumption of equilibrium, viz. the complete mixing of the injected solvent with oil under field conditions. The focus of the present study is to investigate the validity of this assumption and propose new analytical and numerical approaches for non-equilibrium mixing.

Literature Review

Nghiem et al. (1989) modelled unstable miscible displacement by considering two zones: one of complete mixing and the other of bypassed fluid. In their model, the oil was not contacted by the gas but could move as a result of the pressure gradient. They used two variable parameters in their mass transfer equation between the two regions for different Sherwood numbers. The approach used arbitrary parameters.

Saha and Peng (1992) developed a procedure for non-equilibrium phase behaviour in displacement. They defined non-equilibrium as a function of K -values and effective mobility ratios in each grid blocks as:

$$K_i^* = K_i(1 - E_i) \quad (2)$$

where the parameter E_i is the measure of non-equilibrium. They used the mixing parameter model developed by Todd and Longstaff (1972) and the Murphree (1925) efficiency model to calculate the degree of incomplete mixing. Their model employed a matching parameter to predict non-equilibrium behaviour and the researchers compared the results of their simulations with those of slim-tube experiments.

In 1997, Nghiem and Sammon presented an approach to non-equilibrium equation of state modelling in gas injection experiments. They attempted to overcome modelling difficulties during bypassing. They used a transfer term in the flow equation to represent non-equilibrium behaviour between the oil and gas phases. The transfer term was defined as:

$$\tau_i = \tau_{io} - \tau_{ig} \quad (3)$$

where τ is defined as

$$\tau_{ik} = \left[\frac{A}{L} \right] \frac{1}{S_k} \rho_k \frac{D_{ik}}{F} (y_{ik} - y_{ik}^*) \quad (4)$$

where $\left[\frac{A}{L} \right]$ is the characteristic area/characteristic length. S_k , ρ_k , D_{ik} , and y_{ik} are saturation in phase k , molar density of phase k , the diffusion coefficient of component i in phase k , and the molar composition of phase k . F is the formation resistivity factor.

Ivory et al. (2009) investigated numerically and experimentally the performance of cyclic solvent injection as a follow up for the Cold Heavy Oil Production with Sand (CHOPS) process in the Cold Lake and

Lloydminster reservoirs. In their experiment, the solvent (28% propane and 72% CO₂) was injected during six cycles with 50% observed oil recovery after primary production. They modelled primary production using a foamy oil model and used a nonequilibrium model to match the results of the experiments through defining reactions. [Imran et al. \(2013\)](#) used the reaction type model to predict nonequilibrium behaviour in the Solvent Vapour Extraction (SVX) process. In the reaction type model, the gas solvent converts to a liquid solvent through a reaction.

[Ahmadloo et al. \(2011\)](#) experimentally investigated the performance of butane in the VAPEX process. They stated that the solvent content of the produced oil was much lower than the equilibrium values.

In 2015, [Wang et al.](#) conducted the constant composition expansion (CCE) test for methane and propane with heavy oil. They measured the non-equilibrium pressure-volume-temperature (PVT) data. Their tests showed different saturation temperatures at different pressure decline rates at constant temperature.

[Lobanova and Indrupsky \(2015\)](#) developed a non-equilibrium approach in a compositional simulator, based on a new flash calculation without changing the flow equations in the reservoir simulator. In their model, the component mass transfer parameter is defined as a function of the chemical potential difference. The non-equilibrium parameter was based on reservoir characteristics and dynamic flow parameters. In their approach, the fugacity of each component in each phase is not zero, which is the case in equilibrium calculations.

The analytical model for mass transfer and heat transfer in the case of a bitumen droplet, developed by the author, showed that the mobilization of bitumen by solvent dispersion is a much slower process than mobilization by temperature increase. In the solvent or solvent-steam simulation studies reported, instantaneous phase equilibrium is assumed. Many published numerical simulation studies are based on this concept and the results are considered optimistic. In some cases, the dispersion coefficient is increased ("upscaled") to obtain a desirable result. An example is numerical simulations by two large multicompany failed VAPEX tests, where predicted bitumen production rates were 100-170 m³/day but the actual rates were less than 10 m³/day.

Previous non-equilibrium mass transfer experiments and simulations reviewed modified the equilibrium state in some manner via an arbitrary factor. For example, a simplistic approach is to change the value of the equilibrium factor by a time-dependent multiplier. In this work, the state of equilibrium is based on the dissolution of a single drop of bitumen in a pore with or without the application of heat. For a particular application, it is only necessary to determine the average pore size and the diffusion coefficient. The rate of dissolution and mobilization of the bitumen droplet are derived from standard mathematical equations. It is possible to take a range of pore sizes and use a median or desired average value to generate the non-equilibrium phenomena for the compositional thermal simulations.

Methodology

It has been shown that solvent dissolution in bitumen is not an instantaneous process but a certain time is needed for the solvent to reach the equilibrium concentration in the bitumen-rich phase ([Bayestehparvin et al. \(2015\)](#)). In the next section, the methodology for modelling non-equilibrium behaviour in a pore will be explained, however the detailed analytical model will be discussed in another paper. The development of compositional thermal simulator is based on [Abou-Kassem \(1981\)](#) and will be discussed briefly in this paper.

Non-Equilibrium Modelling at Pore Scale

This model is based on the idea of dividing a droplet of bitumen into two distinct regions of mobile and immobile bitumen. The mobile region will flow while the immobile bitumen will be considered to be immobile part of a gridblock, e.g. by reducing the porosity of the grid block. As a result, four phases are considered inside a gridblock, solid (S), oleic (O), aqueous (A), and vapour (V). These phases behave as:

- The solid phase is immobile at all times.
- When the solvent concentration in bitumen at some radius is $<10\%$, or if the temperature is $<T_{lim}$, the bitumen is assumed to be immobile. A better criterion rather than considering temperature and/or concentration threshold is a critical viscosity. This critical viscosity is considered to be 200 cp based on field data and economical oil rate.
- The remaining bitumen is assumed to become part of the oleic phase.

Four components are considered in this simulator; bitumen (b), solvent (s), water (w), and gas (g). It is possible to characterize bitumen by defining different pseudo-components; however, the analytical model will become more computationally intensive. This modification will be discussed in future work. In a grid block, bitumen exists strictly as a solid phase however gas can be a non-condensable gas, such as methane, or condensable gas, such as hexane, etc., which may exist in liquid or vapour phase at block conditions. The schematic drop model is shown in Figure 1.

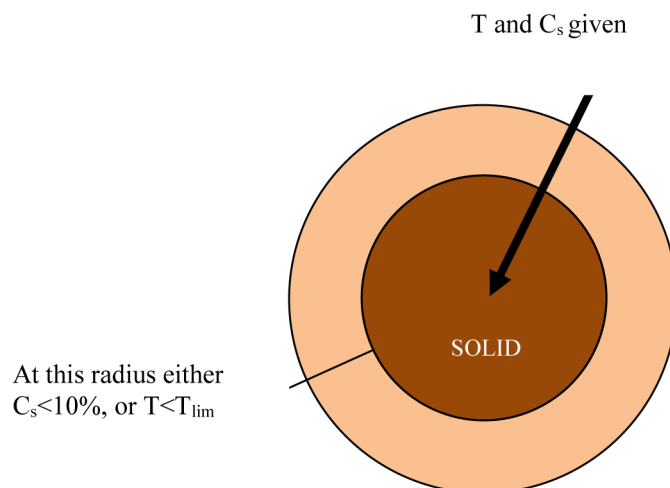


Figure 1—Drop Model. Arrow shows heat and solvent diffusion

Assumptions.

1. Solvent and gas may diffuse into bitumen.
The amount of solvent or gas diffusing into bitumen comes from the drop model, for a given concentration and temperature.
2. Bitumen is insoluble in water, soluble in solvent, but may not transfer into vapour phase components (water/steam and gas).
3. Water can only transfer into the vapour phase.
4. Solvent may diffuse into bitumen and transfer into the vapour phase.
5. Gas may diffuse into bitumen and solvent, but not into water.

What can be obtained from the Drop Model?. 1. K-Values

Given a drop, the concentration or temperature at any radius can be determined, as shown in the sketch, and the bitumen in the outer ring is assumed to mix with the solvent instantly.

In the simulation, a given block has a known amount (mass) of bitumen. Depending on the solvent concentration or temperature in the block, at the end of a time step, a certain amount of bitumen is left undissolved. We have to keep an account of the bitumen in each block, at the end of each time step.

Consider two extreme situations.

1. Bitumen is immediately miscible with the oleic (i.e. solvent) phase

$$\begin{aligned} \text{In that case } K_{OSb} &= x_{bO}/x_{bS} \\ &= \text{mole frac of bitumen in oleic phase/mole frac of bitumen in solid phase} \\ &= x_{bO}/0 = \infty \end{aligned}$$

For this case, instead of a separate solid phase, we can just define the oil as made up of two components, one a heavy fraction (bitumen), and the other one a light fraction (solvent). Then the oleic phase will always consist of two components in general.

2. Bitumen is totally insoluble in the oleic phase

$$\text{In that case } K_{OSb} = x_{bO}/x_{bS} = 0$$

For this case, bitumen will stay in the block, but may become mobile if there is a temperature increase. In that case, we can have four phases flowing in the general case.

For the case shown in the diagram, there is a certain amount of bitumen diffused into the surrounding oleic phase, and certain amount of solvent diffused into the solid bitumen phase.

So in this case $K_{OSb} = x_{bO}/x_{bS}$, as before, and

$$\begin{aligned} K_{OSs} &= x_{sO}/x_{sS} \\ &= \text{mole frac of solvent in oleic phase/mole frac of solvent in solid phase} \end{aligned}$$

Again, eventually the bitumen phase can become mobile, because of solvent concentration buildup and/or temperature increase. But in this case, we will specify that the bitumen solid phase is always immobile, regardless of solvent concentration or temperature, and the only way it can be transported is by diffusion into the solvent phase.

2. Viscosity and Property Data

We have two models, one for concentration, and one for temperature. We can generate curves like those shown in [Figure 2](#).

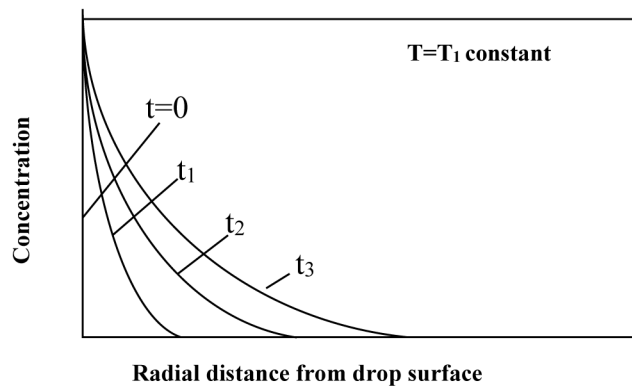


Figure 2—Concentration profile as functions of radius at constant temperature ($t_1 < t_2 < t_3$)

Similar curves can be generated for temperature in place of concentration. These data can either have a subroutine to generate the above data, or more conveniently, we can generate tables of concentration vs. time at different radii, for a given temperature. We can then interpolate for concentration at any time, and any radius, and temperature. This is much like K-value interpolation as functions of temperature and pressure, for different components. A similar table can be generated for temperature in place of concentration.

Overall Procedure

At a given simulation time, for a given temperature and solvent concentration, we determine the radius from the generated tables where viscosity is 200 cp. That will give the radius of the immobile bitumen. Or, in other words, the fraction of the solid phase that is immobile. Also, we know how much of the solid phase is transferred into the solvent phase. Given the concentrations, we can calculate the viscosity of the solvent phase. We can also calculate the solvent diffused into the solid bitumen phase, and its viscosity, but we are assuming that the solid phase is always immobile.

If there were complete equilibrium, and the bitumen were the heavy fraction of the oleic phase, and (solvent+bitumen) the oleic phase, we would not need to do any of the above. The oleic-vapour phase K-value would determine how much solvent is in the oleic phase and how much is in the vapour phase. Our non-equilibrium procedure means that not all the bitumen is transferred to the oleic phase, because now it is a separate phase.

Analytical Model. The analytical model will be discussed in detail for spherical coordinates (drop model) in another work. A simple linear model will be presented in this section to illustrate the procedure. This model is developed for concentration distribution in the two-phase system. In this case, we assume that the phases reach equilibrium at their interface. In this case, at infinite time, the phases reach a homogeneous concentration given by the K-value. However, they will not form a single phase.

In this case, K-values are required for the boundary conditions and are obtained using the phase behaviour equation of state by equalizing the fugacities of the components.

The following assumptions are implicit in the above equations:

1. There is no volume change due to the mixing of solvent and bitumen
2. The dissolution of bitumen in solvent is zero, the K-value of bitumen is zero and the equivalent concentration of the solvent in bitumen boundary is $1/K$ -value
3. The diffusion coefficient is constant
4. There is no external force (gravity)
5. There is no thermal effect

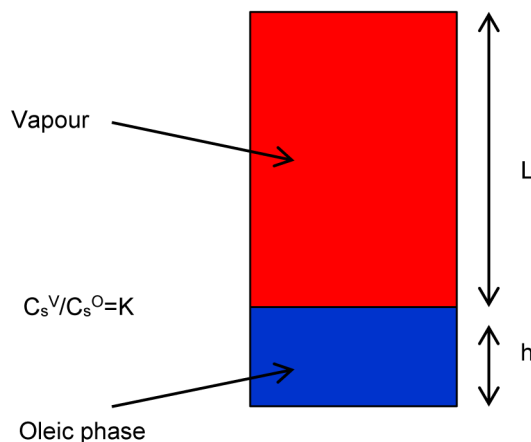


Figure 3—Schematic diagram of two phase mass transfer

Reservoir Simulator Development. A thermal compositional reservoir simulator was developed in C⁺⁺. Some simple fluid and rock properties are considered to validate the code.

Model Assumptions

The principal assumptions are as follows:

1. The simulator can handle three phases, oleic, aqueous, and vapour.
2. Hydrocarbon solubility in the aqueous phase is zero.
3. Water is insoluble in the oleic phase but can exist in the gas phase as steam.
4. Kinetic energy and viscous force are neglected.
5. The phases are assumed to reach equilibrium instantaneously. This assumption is invoked for the base case and is modified based on the developed pore scale analytical model for time-dependent partition coefficient.

Partial Differential Equations

In this simulator, we state the mass balance equations for all components, energy balance, and constraint equations.

1. Molar mass equation

The molar mass balance for component i in the system is:

$$-\vec{\nabla} \cdot \left[\sum_{p=o,w,g} x_{ip} \overline{\rho_p} \vec{u_p} \right] - q_i^* - i_i^* + r_i^* = \frac{\partial}{\partial t} \left[\phi \sum_{p=o,w,g} x_{ip} \overline{\rho_p} S_p \right] \quad (5)$$

This equation consists of conduction/convection, production, injection, reaction, and accumulation term. All of these terms have the units of mol/m³/s. The reaction term is ignored in this simulator. This equation is simplified for hydrocarbon components and water as follow:

$$\text{Hydrocarbon components: } -\vec{\nabla} \cdot [x_i \overline{\rho_o} \vec{u_o} + y_i \overline{\rho_g} \vec{u_g}] - q_i^* - i_i^* = \frac{\partial}{\partial t} [\phi (x_i \overline{\rho_o} S_o + y_i \overline{\rho_g} S_g)]$$

$$\text{Water: } -\vec{\nabla} \cdot [\overline{\rho_w} \vec{u_w} + y_i \overline{\rho_g} \vec{u_g}] - q_i^* - i_i^* = \frac{\partial}{\partial t} [\phi (\overline{\rho_w} S_w + y_i \overline{\rho_g} S_g)]$$

2. Darcy law for each phase:

$$\vec{u_p} = -k \frac{k_{rp}}{\mu_p} (\vec{\nabla} P_p - \gamma_p \vec{\nabla} Z) \quad (6)$$

This equation is used for velocity calculation of each phase.

3. Energy Balance Equation

$$\begin{aligned} & \vec{\nabla} \cdot [K_r \vec{\nabla} T] - \vec{\nabla} \cdot \left[\sum_{p=o,w,g} \overline{\rho_p} \vec{u_p} \sum_{i=1}^4 x_{ip} H_{ip} \right] - q_i^* - i_i^* + r_i^* - \sum_{j=1}^6 q_{Lj}^* - \frac{1}{2Jg_c} \vec{\nabla} \cdot \left[\sum_{p=o,w,g} \overline{\rho_p} \vec{u_p} u_p^2 \right] \\ & - \vec{\nabla} \cdot \left[\sum_{p=o,w,g} \tau_p \vec{u_p} \right] = \frac{\partial}{\partial t} \left[(1 - \phi) \rho_r U_r + \phi \sum_{p=o,w,g} \overline{\rho_p} S_p \sum_{i=1}^4 x_{ip} U_{ip} + \frac{1}{2Jg_c} \sum_{p=o,w,g} \overline{\rho_p} u_p^2 \right] \end{aligned} \quad (7)$$

This equation contains conduction, convection, production, injection, reaction, heat loss, and accumulation terms. Each term has unit of J/m³/s.

4. Phase Saturation and Composition Constraints

$$\sum_{p=o,w,g} S_p = 1 \quad (8)$$

$$\sum_{i=1}^3 x_i = 1 \quad (9)$$

$$\sum_{i=1}^4 y_i = 1 \quad (10)$$

5. *The phase pressure*

$$P_{cow} = P_o - P_w \quad (11)$$

$$P_{cgo} = P_g - P_o \quad (12)$$

Considering the large pore sizes in oil sands, the capillary pressures are negligible and the pressures of all phases would be equal.

6. *Phase Equilibrium Relations:*

$$y_i = x_{ig} = K_{iog} x_{io} \quad (13)$$

$$x_{iw} = \frac{y_i}{K_{iwg}} \quad (14)$$

In addition to these equations, initial and boundary conditions are needed for model solution. The initial conditions are pressure, temperature, and composition distribution at initial time. The boundary of the model needs to be derived from the condition that the reservoir boundaries are impermeable to fluid flow.

Rock and Fluid Properties

In this simulator, all of the derivatives are calculated analytically except for the data given in tabular form (e.g. steam tables).

Equilibrium Ratios

The water equilibrium ratio is calculated using steam tables by taking the saturated steam pressure at the grid block temperature. This value is a function of temperature and pressure and is independent of the component compositions in the oleic phase.

$$\widehat{K}_w = \frac{P^{sat}}{P} \quad (15)$$

$$K_w = \widehat{K}_w X_g X_w \quad (16)$$

$$X_g = \frac{S_g + \varepsilon_g}{S_g + 10^{-30}} \quad (17)$$

$$X_w = \frac{S_w}{S_w + \varepsilon_w} \quad (18)$$

The hydrocarbon K-values are expressed in tabular form. These values need to be modified in order to model phase disappearance at low saturations.

$$\widehat{K}_i = \frac{y_i}{x_i} \quad (19)$$

$$K_i = \widehat{K}_i X_o X_g \quad (20)$$

$$X_o = \frac{S_o}{S_o + \varepsilon_o} \quad (21)$$

In the case of negative saturation, the following modification is used:

$$S_p^v = \sqrt{\varepsilon_p S_p^n} \quad (22)$$

Capillary Pressures and Relative Permeability Curves

The capillary pressures can be presented in tabular form as a function of temperature and saturation. The following equations are used to determine capillary pressures (Coats 1980).

$$P_{cwo} = [a_1 + a_2(1 - S_w) + a_3(1 - S_w)^3][1 - a_4(T - T_i)] \quad (23)$$

$$P_{cgo} = [b_1 + b_2S_g + b_3S_g^3][1 - b_4(T - T_i)] \quad (24)$$

Parameters a_1 to a_4 and b_1 to b_4 are constants. The relative permeability curves are calculated by Stone's Model (II). In this method, the relative permeabilities of the water and gas are obtained from tables as functions of water and gas saturations, respectively.

$$k_{rw} = k_{rw}(S_w) \quad (25)$$

$$k_{rg} = k_{rg}(S_g) \quad (26)$$

The relative permeability of the oil is obtained from the following equation:

$$k_{ro} = k_{ro_{cw}} \left[\left(\frac{k_{row}}{k_{ro_{cw}}} + k_{rw} \right) \left(\frac{k_{rog}}{k_{ro_{cw}}} + k_{rg} \right) - (k_{rw} + k_{rg}) \right] \quad (27)$$

Molar Internal Energy of Phases

The molar internal energies of the phases are calculated considering ideal solution behaviour as follows:

$$U_g = \sum_{i=1}^3 y_i u_{ig} + y_w u_s \quad (28)$$

$$u_{ig} = c_{vi}(T - T_{in}) + L_{vi} \quad (29)$$

$$U_o = \sum_{i=1}^3 x_i u_{io} \quad (30)$$

$$u_{io} = c_{vi}(T - T_{in}) \quad (31)$$

The steam and water internal energies are taken from the steam tables.

Molar Density

The gas phase molar density is calculated by the real gas law.

$$\bar{\rho}_g = \frac{P}{ZRT} = \frac{P}{RT(k_4 Z_s(T) + Z_{hc} \sum_{i=1}^3 k_i x_i)} \quad (32)$$

$Z_{h.c.}$ is given in tabular form. Z_s is calculated from the steam tables at the saturated temperature as follows.

Viscosity

Vapour, oleic, and aqueous viscosities are calculated with the following equations.

$$\mu_g = y_w \mu_s + \mu_{hc} \sum_{i=1}^3 y_i \quad (33)$$

$$\ln(\mu_o) = \ln(\mu_o(T))x_o + \ln(\mu_s(T))x_s \quad (34)$$

$$\mu_w = \mu_w(T) \quad (35)$$

Diffusion Coefficient

The diffusion coefficient mostly considered to be constant. in most cases. However, this parameter changes by viscosity of the fluid and operating condition. The viscosity dependent diffusion coefficient suggested by Hayduk and Cheng (1971) was widely used for this purpose:

$$D(\mu) = a\mu^b \quad (36)$$

Das and Butler (1996) investigated propane and butane diffusion coefficient for the Peace River bitumen based on this correlation.

Production and Injection Models

Four different production scenarios are included in the present study, specified oil production rate, specified total liquid production rate, specified total production rate, and bottom-hole pressure. The rates are specified by the following equations for the oil components and water.

$$q_i = \begin{cases} x_i \rho_o q_o + y_i \rho_g q_g \\ \rho_w q_w + y_i \rho_g q_g \end{cases} \quad (37)$$

The volumetric production rates are dependent on the scenario considered. The fourth scheme is transformed into other schemes if the grid block pressure is lower than the bottom-hole pressure. The injection rates are calculated through two different schemes of constant injected components and specified bottom-hole pressure. For the injection scheme, the second approach is replaced by other approaches if the grid block pressure exceeds the maximum bottom-hole pressure.

Results and Discussion

The thermal simulator was developed in C⁺⁺ and is validated with Abou-Kassem steamflood model (1981). The injection, production, and accumulation terms were treated implicitly. The upstream approach has been chosen for mobility calculation. Thomas's method was used as the diagonal matrix solver. In the first step, a two dimensional model was considered. Two different recovery methods of steam injection and solvent-steam co-injection are applied in this reservoir. Butane was selected as the injected solvent. The injector and producer are handled through bottomhole constraints of 1000 psi and 400 psi, respectively. The selected oil has the typical properties of a California oil with a viscosity of 2000 cp and steamflooding is discussed in this work. The application of non-equilibrium assumption in SAGD is investigated in another study (Bayestehparvin et al. (2017)).

Butane is injected into reservoir with 5 mol% concentration in the injection fluid. In the first step, the equilibrium assumption was considered. The cumulative oil production is compared in Figure 4. As the figure indicates the solvent increases the oil production by viscosity reduction which is in agreement with the results of commercial simulators.

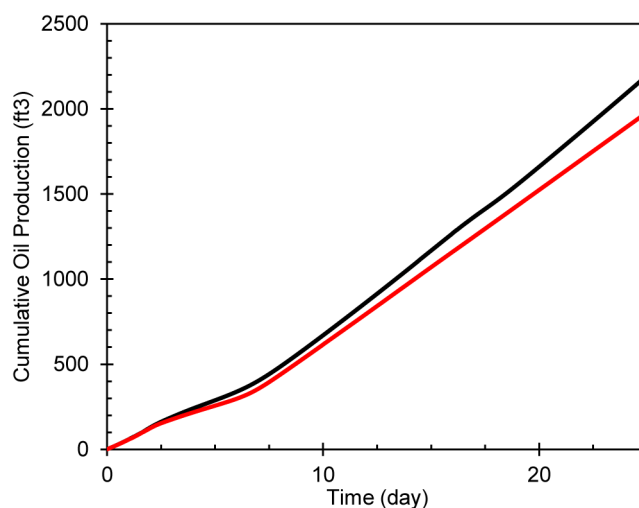


Figure 4—Cumulative oil production (ft³). Black and red lines are butane coinjection and steam injection, respectively.

In the next step, the analytical model was used to generate partition coefficient as a function of time. This nonequilibrium partition coefficient is a function of solvent concentration, pressure, temperature, and time. The oil concentration alteration with time is shown in Figure 5.

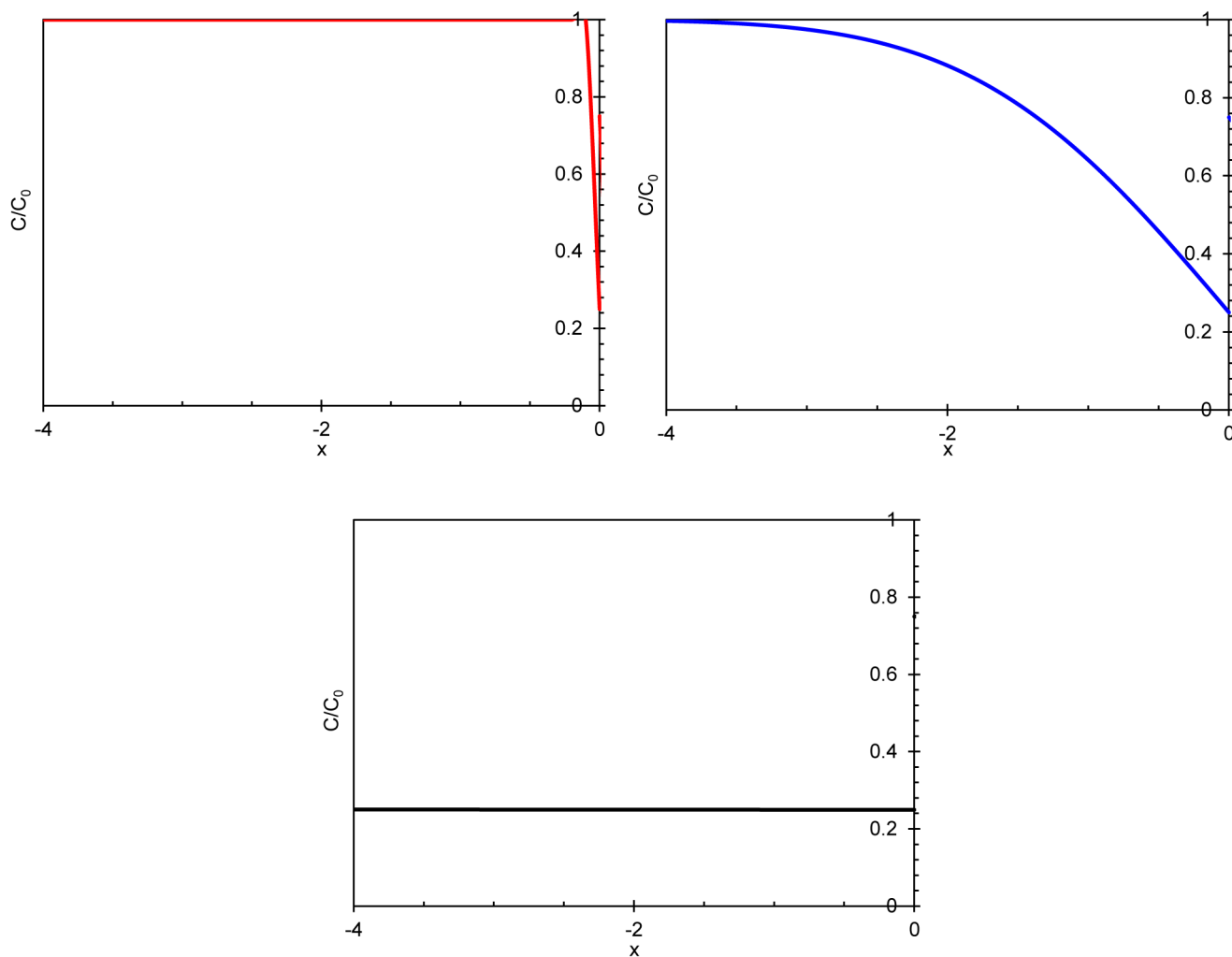


Figure 5—Oil concentration as a function of distance. Red, blue, and black lines are for \sqrt{Dt} of 0, 1, and ∞ , respectively.

At initial time, the oil-rich phase is pure oil except at boundary which is dictated by the K-value. Oil concentration in the oil-rich phase decreases with time. After some time, the oil composition will reach its equilibrium value. Based on the analytical equation, the transient partition coefficient is generated in Figure 6. The partition coefficient is reduced until it reaches its equilibrium value.

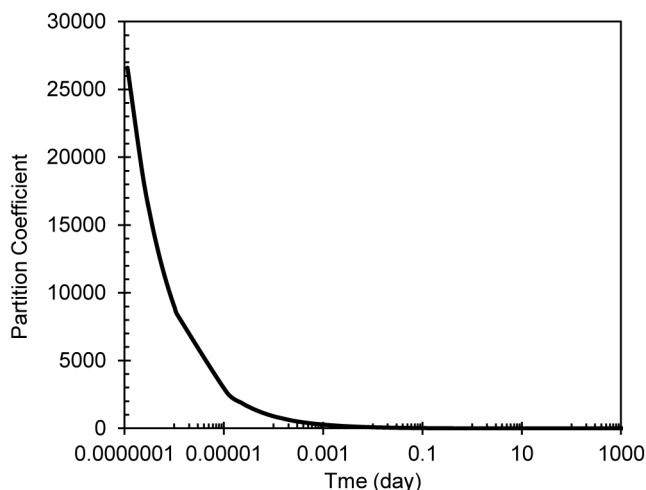


Figure 6—Partition coefficient of butane in butane-oil binary system as a function of time.

This transient partition coefficient is applied in the solvent co-injection case and the results compared in Figure 7. As the figure indicated, the oil production is lower in transient partition coefficient case compared to equilibrium K-value. It should be noted that this low oil production is not only due to lower oil viscosity in the equilibrium assumption case. When the partition coefficient is high, the solvent will remain in the vapour phase rather than dissolve in the oil phase. As a result, by having more solvent in the vapour phase, the pressure of the gridblock will be maintained and the steam injection will be lower for the constant bottomhole pressure constraint. By injecting less steam, less heat will be injected into the reservoir and the recovery will be lower. More solvent is dissolved in the oil-rich phase and the steam injection rate increases with time. In the SAGD case, the oil contacts fresh solvent at each time step and as a result, the oil does not have enough time to reach equilibrium at the chamber edge.

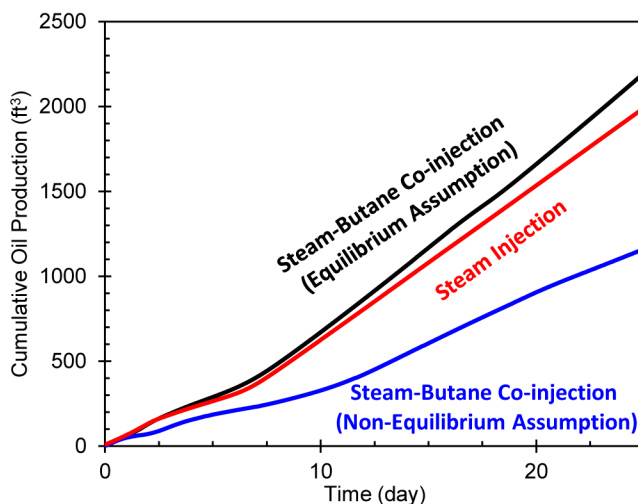


Figure 7—Cumulative oil production (ft³). Red, Black, and blue are steam injection, solvent co-injection (equilibrium assumption), and solvent co-injection (non-equilibrium assumption).

The cumulative steam-oil ratio is shown in Figure 8. As the figure shows, solvent co-injection reduces steam-oil ratio compared to the steam injection case. In the case of solvent-co-injection with equilibrium assumption, the injected steam is much lower than in other processes.

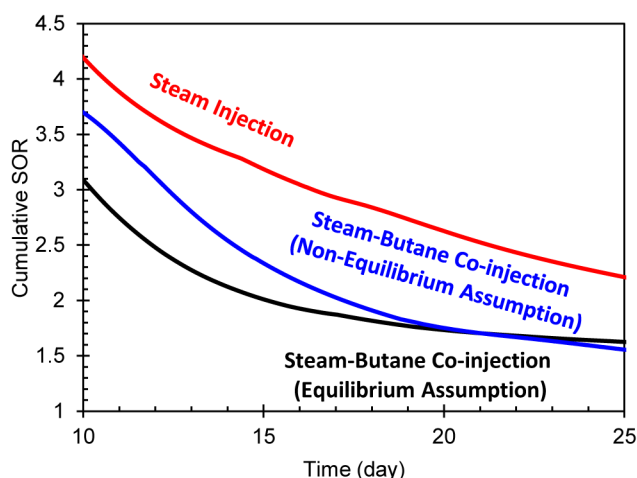


Figure 8—Cumulative steam oil ratio. Red, Black, and blue are steam injection, solvent co-injection (equilibrium assumption), and solvent co-injection (non-equilibrium assumption).

Conclusions

A non-equilibrium thermal-compositional simulator is developed based on heat and mass transfer at the pore scale. Based on the results for three scenarios of steam injection, solvent coinjection (equilibrium assumption), and solvent coinjection (non-equilibrium assumption) the following conclusions are obtained:

1. The equilibrium assumption in the simulator is not valid and a non-equilibrium approach needs to be applied.
2. The solvent co-injection in the equilibrium assumption can improve oil production by reducing the viscosity of the oil compared to steam injection.
3. The solvent co-injection with non-equilibrium assumption can have negative impact at the oil recovery. At early times, solvent cannot diffuse into the oil and cannot reduce oil viscosity. In addition, the presence of vapour solvent maintains the pressure which reduces the steam injection rate.
4. No convergence problem was observed by the non-equilibrium assumption in the simulator.

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