Effect of Temperature on Performance of Vapex Process for Heavy Oil

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Effect of Temperature on Performance of Vapex Process for Heavy Oil

by

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A THESIS
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Abstract

Incorporating some heat injection along with the solvent injection appears to be the most viable option for improving the oil drainage rate of Vapex (Vapor Extraction) in extra-heavy oil formations. This study was aimed at quantifying the upper limit of oil rates achievable with heated solvent injection. Results of the performed research can also be used to assess the applicability of Vapex to naturally warm reservoirs and reservoirs with mobile oil in place.

The experimental phase of current research involved conducting Vapex experiments in a high-pressure physical model, packed with 250 Darcy sand, saturated with Athabasca bitumen (Mackay River) using propane as the solvent. The physical model was pre-heated to 40, 50 and 60 °C and propane was injected within 0.817 to 1.55 MPa pressure to investigate the effect of temperature elevation and injection pressure on Vapex performance.

Due to the mobility of the original oil in place at elevated temperatures, the total rate of oil production is controlled by two mechanisms: 1. solvent dissolution and oil mobilization at the boundaries of the vapor chamber 2. free fall gravity drainage of undiluted oil within the sand matrix beyond the solvent penetration zone. To evaluate warm Vapex performance under different operating conditions, the contribution of free fall gravity drainage to total recovery was experimentally and numerically determined.

Experimental results were compared against Vapex analytical models and a reported empirical correlation. The empirical correlation was modified based on the findings of the current experimental study by incorporating the effect of live oil viscosity into the original correlation.

The experimental results of warm Vapex tests were numerically simulated with a commercial compositional simulator, Computer Modelling Group’s (CMG) GEM. Bitumen characterization and thermo physical modeling of bitumen/propane system were performed as part of the numerical study. The developed simulation model verified by the experimental production data was used to extend the study to wider ranges of temperature.

Results of the current study indicates that elevating the reservoir temperature to 50°C without increasing the injection pressure, can improve production rate only by 70% and reservoir temperature elevation to above 60 °C is required to achieve more impressive rate increase of nearly 200% when Vapex is performed in heavy oil reservoirs.
Acknowledgements

My first and profound appreciation goes to my advisor Dr. Brij Maini for his unconditional availability, continuous support and invaluable advices through years of my study at University of Calgary. The extensive knowledge, creative thinking and insightful vision of Dr. Maini have always been a source of inspiration for me and always will be.

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I would also like to extend my appreciations to Amin Badamchi Zadeh and Mohammad Tavallali for their constructive technical inputs on the numerical simulation aspects of current research.


The love, support and encouragement I’ve received all the way from my family are beyond what I can express through these solid words. I’m endlessly grateful to my parents for keeping hope and determination alive during all these years of my life far from home.
This Dissertation is Lovingly Dedicated to My Parents.
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<th>Symbol</th>
<th>Definition</th>
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<tbody>
<tr>
<td>a</td>
<td>Cubic EoS attractive term.</td>
</tr>
<tr>
<td>A</td>
<td>Empirical exponent in relative permeability-saturation exponential function</td>
</tr>
<tr>
<td>b</td>
<td>Co-volume,</td>
</tr>
<tr>
<td>b₁, b₂, b₃, b₄, b₅</td>
<td>Viscosity coefficients for Pedersen correlation</td>
</tr>
<tr>
<td>B₀</td>
<td>Oil formation volume factor</td>
</tr>
<tr>
<td>C</td>
<td>Solvent concentration (volume fraction).</td>
</tr>
<tr>
<td>C_D</td>
<td>Dimensionless solvent volume fraction</td>
</tr>
<tr>
<td>cᵢ</td>
<td>Volume shift value for component i</td>
</tr>
<tr>
<td>C_max</td>
<td>Maximum solvent concentration, volume fraction</td>
</tr>
<tr>
<td>C_min</td>
<td>Minimum solvent concentration, volume fraction</td>
</tr>
<tr>
<td>D</td>
<td>Mass diffusion coefficient, L²t⁻¹</td>
</tr>
<tr>
<td>d</td>
<td>Mass diffusion power</td>
</tr>
<tr>
<td>D_eff</td>
<td>Effective mass diffusion coefficient, L²t⁻¹</td>
</tr>
<tr>
<td>dᵢⱼ</td>
<td>Binary interaction coefficient.</td>
</tr>
<tr>
<td>D_num</td>
<td>Numerical dispersion coefficient, L²t⁻¹</td>
</tr>
<tr>
<td>Dₛ</td>
<td>Mass diffusion coefficient at Vapex interface, L²t⁻¹</td>
</tr>
<tr>
<td>F</td>
<td>Formation resistivity factor</td>
</tr>
<tr>
<td>fᵣ</td>
<td>Bitumen volume fraction in produced oil</td>
</tr>
<tr>
<td>fₒ, fₛ</td>
<td>Compositional parameter in Shu correlation</td>
</tr>
<tr>
<td>g</td>
<td>Gravitational acceleration, Lt²</td>
</tr>
<tr>
<td>H</td>
<td>Height, L</td>
</tr>
<tr>
<td>H'</td>
<td>Equilibrium liquid hold-up, L</td>
</tr>
<tr>
<td>K</td>
<td>Permeability, L²</td>
</tr>
<tr>
<td>L</td>
<td>Length of the horizontal well or width of the physical model, L</td>
</tr>
<tr>
<td>m</td>
<td>Kinematic viscosity power</td>
</tr>
<tr>
<td>n</td>
<td>HIM-Polynomial function degree</td>
</tr>
<tr>
<td>N</td>
<td>Original oil in place, L³</td>
</tr>
<tr>
<td>Nₛ</td>
<td>Vapex dimensionless number, Eq. 1-18</td>
</tr>
<tr>
<td>P</td>
<td>Pressure, ML⁻¹t⁻¹</td>
</tr>
<tr>
<td>P_c</td>
<td>Critical pressure, ML⁻¹t⁻¹</td>
</tr>
<tr>
<td>P_r</td>
<td>Reduced pressure</td>
</tr>
<tr>
<td>Pᵥ_propane</td>
<td>Propane vapor pressure, ML⁻¹t⁻¹</td>
</tr>
<tr>
<td>q</td>
<td>Vapex oil production rate per unite well-length, L²t⁻¹</td>
</tr>
<tr>
<td>Q</td>
<td>Dead oil rate, L² t⁻¹</td>
</tr>
<tr>
<td>qₒ</td>
<td>Vapex oil production rate per unit well length, L² t⁻¹</td>
</tr>
<tr>
<td>R</td>
<td>Recovery percent per day</td>
</tr>
<tr>
<td>S</td>
<td>Liquid saturation</td>
</tr>
<tr>
<td>S_d</td>
<td>Liquid saturation at the demarcator</td>
</tr>
<tr>
<td>S_G</td>
<td>Specific gravity</td>
</tr>
<tr>
<td>S;oᵢ</td>
<td>Initial oil saturation</td>
</tr>
</tbody>
</table>
$S_{or}$  Residual oil saturation  
$T$  Absolute temperature, $T$.  
$T_c$  Critical temperature, $T$  
$T_r$  Reduced temperature  
$U$  Velocity of vapour interface in Vapex, $L_t^{-1}$  
$u_x, u_y, u_z$  Fluid velocity in x, y and z direction, $L_t^{-1}$  
$V_{ci}$  Critical volume of component i. $L_t^{-3}M_t^{-1}$  
$v_s$  Flow velocity in the saturated region (EQ 2-18), $L_t^{-1}$  
$v_{ur}$  Flow velocity in the unsaturated region (EQ 2-19), $L_t^{-1}$  
$W$  Formation width, $L$  
$w_i$  Weight fraction of component i.  
$x$  Compositional parameter  
$x_{propane}$  Propane mole fraction  
$z_d$  Position of demarcator (EQ 2-26), $L$  

**Subscripts**  
**Definition**  
eff  Effective  
g  Gas  
i  Component i  
o  Oil  
s  Solvent-oil interface  
ss  Steady state  

**Greek Symbols**  
**Definition**  
$\beta$  Converging rate constant, EQ 2-33  
$\zeta$  Distance vertical from Vapex interface, $L$  
$\tau$  Tortuosity  
$\Delta \rho$  Density difference, $ML^{-3}$  
$\Omega$  Cementation factor  
$\alpha$  Viscosity correlation variable, Shu eq. and Pedersen.  
$\phi$  Porosity.  
$\gamma_{propane}$  Propane activity coefficient.  
$\mu$  Dynamic viscosity, $ML^{-1}t^{-1}$  
$\theta$  Vapex interface slope at each vertical location.  
$\nu$  Kinematic viscosity, $L^2t^{-1}$  
$\omega$  Acentric factor.
Chapter One: Introduction

1.1 Canadian Bitumen Resources

The heavy oil and bitumen reservoirs of Canada are one of the largest hydrocarbon resources in the world matching about 67% of the Saudi Arabia’s conventional oil reserves (Dunbar et al, 2003). As the remaining conventional oil reserves around the world declines, interest continues to grow toward the tar sands and heavy oil resources. The total remaining established Canadian reserves, which is defined based on the volumes that are likely to be recovered by mining or in-situ methods using established technology and under anticipated economic conditions, is 167.9 billion barrels as reported by ERCB in 2013. Only 5.3% of the initial established crude bitumen reserves has been produced since commercial production started in 1967 (ERCB annual report 2013). The Canadian oil sand resources are spread across more than 140,000 square kilometers of northern Alberta in the Western Canada Sedimentary Basin located in Athabasca, Cold Lake and Peace River as depicted in Figure 1-1 (Dunbar et al, 2003).

![Alberta’s Oil Sand Areas](image-url)

Figure 1-1: Alberta’s Oil Sand Areas (ERCB, 2013)
1.2 Recovery Methods

Extraction of heavy oil and bitumen is quite challenging due to high viscosity and immobility of these types of crudes in their original estate. Surface mining is an effective method applicable to shallow formations typically less than 70 m deep, which comprise only 10-15% of Canadian oil sand deposits. Production from the remaining reservoirs, which are buried too far below the surface, requires employing in-situ techniques. These in-situ techniques can be categorized as follows:

- **Primary recovery:**

  - CHOPS: Cold Heavy Oil Production with Sand

  CHOPS can be effective in thin unconsolidated oil sand formations where heavy oil viscosity at reservoir conditions is less than 50 Pa.s and the oil contains sufficient solution gas to promote the production through gas expansion mechanism. Producing sand along with heavy oil in CHOPS appears to increase oil rates by developing wormholes that can expand over distances in the reservoir and improve oil production.

- **Thermal recovery:**

  Thermal methods employ different strategies to reduce oil viscosity by means of providing heat to reservoir. These methods include:

  - CSS: Cyclic Steam Stimulation (CSS)

  CSS involves injecting steam into the reservoir through vertical wells at pressures high enough to create paths for the fluid flow. For each individual well, periods of steam injection are followed by periods of soaking and production (Dunbar et al, 2003).

  - THAI: Toe to Heal Air Injection

  THAI involves injecting air into the reservoir through a vertical injection well located near the toe of a horizontal well, which is placed near the base of the formation, to start ignition. In this process, a combustion front is created which burns the left behind oil residues, generates heat, reduces the oil viscosity and enables its flow by gravity to the horizontal producer.
SAGD: Steam Assisted Gravity Drainage

SAGD is based on injecting steam into the reservoir through a horizontal injection well, mobilizing the oil around the created steam chamber by means of heat conduction and producing it under the effect of gravity through a parallel horizontal producer located directly below the injector at a vertical distance of about five meters.

- Solvent-Based recovery:

Solvent injection based techniques can be employed to recover heavy oil and bitumen from reservoirs where thermal recovery methods fail to operate economically due to inevitable heat losses.

- VAPEX: Vapor Extraction

VAPEX is the solvent analogue of SAGD which is implemented under the same well configuration and involves injecting a solvent vapor, mobilizing the oil by means of dilution and producing it under gravity drive mechanism.

- Hybrid Thermal/Solvent Recovery

Hybrid recovery methods are emerging techniques which combine the benefit of higher recovery rates from thermal based methods with the benefits of less energy consumption and environmental impacts from the solvent based recovery methods. A wide range of hybrid recovery methods will be discussed in the next chapter.

1.3 Vapex Concept

Vapour Extraction (VAPEX) was invented by Dr. Roger Butler in 1989 by replacing the steam in SAGD with a light hydrocarbon solvent to mobilize oil by dilution and to produce it under the gravity effect. Vertical cross section of a typical reservoir under Vapex process is depicted in Figure 1-2.
In Vapex, solvent vapour is injected into the reservoir through a horizontal injector. The injected solvent vapor initially tends to rise up due to its lower density while the bitumen diluted by absorption of the solvent drains down towards the horizontal producer located vertically below the injector. Solvent rise up continues until the solvent reaches the top of the formation and a solvent chamber is built in the swept region around the injector. After reaching the top boundary of the formation, the chamber starts to spread laterally. At the solvent-bitumen interface of the created chamber, the vaporized solvent dissolves into the oil and solvent molecules penetrate into the bulk of bitumen. Therefore a mixing zone is formed along the boundaries of the chamber. A concentration gradient is built within the mixing zone with maximum concentration of the solvent at the interface. When the average concentration of the solvent within this mixing zone is sufficient for mobilization of the oil, the diluted oil flows toward the producer within a thin layer adjoining the chamber boundary under gravity (Karmaker et al 2003). Dissolution of the solvent in bitumen results in a drastic decrease in the oil viscosity as well as the interfacial tension between diluted oil and solvent vapour. As a result, interfacial forces holding up diluted oil will be overcome by gravitational forces and consequently residual oil saturation in the depleted zone is lowered. During the lateral growth of vapour chamber, oil drains at an almost stable rate. However, when the solvent chamber reaches the outer boundaries of the formation, the solvent-bitumen interface starts falling down. This results in decreased gravity head for drainage and
proportionately reduces the drainage rate and eventually the economic limit of the operation is reached.

Number of advantages can be listed for Vapex over other heavy oil and bitumen extraction methods. Vapex capital cost is comparatively low since it doesn’t require extensive surface facilities such as those for hot water or steam generation in thermal recovery methods and the subsequent wastewater treatment (Upreti, 2006). Huge energy consumption and high greenhouse emissions which are main concerns associated with thermal based recovery method are much less of concern in Vapex. In spite of high solvent consumption during Vapex process, laboratory results reveal that up to 90% of the injected solvent can be retrieved and recycled in this process at the end of the project (Butler et al, 1998). Vapex can be effectively applied to thin reservoirs, low porosity and high water content formations and in the presence of overlying gas caps and underlying aquifers where thermal based methods fail due to high chances of heat and fluid losses. The possibility of in-situ upgrading of bitumen as a result of asphaltene precipitation upon light hydrocarbon solvent injection in Vapex can be also noted as an advantage in relatively high permeable formations. In spite of these advantages, slow mixing of the solvent with bitumen and relatively slow rate of oil production makes economic viability of this process questionable.

1.4 Objectives of Current Thesis

Vapex has been experimentally and numerically studied since 1989 by many researchers, although the slow rate of oil recovery in this process has hindered its field implementation so far. Incorporating some heat injection along with the solvent injection appears to be the most viable option for improving the drainage rate. The obvious question then concerns the magnitude of temperature increase needed to make the drainage rate economical. Is a modest increase in temperature enough to make the process viable in Canadian heavy oil reservoirs? In the current study we have tried to answer this question by conducting Vapex experiments at elevated temperatures. This sidesteps the question of how the temperature will be raised in the actual field implementation of the process. The results of these tests will clarify the upside potential of rate enhancement by modest heating of the reservoir. The results will also reveal the applicability of Vapex to reservoirs in which the temperature is naturally much higher, such as in Venezuela.
The experimental phase of this research involves conducting Vapex experiments in physical sand-packed models using propane as the solvent to address the following objectives:

- Evaluating the effect of reservoir temperature on Vapex performance.
- Investigating the effect of solvent injection pressure at elevated temperatures on Vapex performance.
- Estimating the upper limit of oil recovery rates which can be achieved in N-Solv and warm Vapex processes.
- Evaluating the contribution of mass transfer and heat transfer on oil mobilization and production in hybrid heat-solvent recovery processes.
- Evaluating Vapex performance in reservoirs with originally mobile oil in place.

The basic premise in heated Vapex is that injection of heated solvent vapor will increase the reservoir temperature and the increased reservoir temperature will provide higher drainage rates. The maximum possible increase in drainage rate by such heating can be evaluated more reliably by decoupling the rate process involved in reservoir heating from the mass-transfer and gravity drainage aspects. By pre-heating the entire model to a target temperature, we eliminate the heat transfer complications and focus only on the mass-transfer and gravity drainage.

The numerical phase of this research is aimed at evaluating the capability of reservoir simulators to history match the obtained results of warm Vapex experiments at different operating conditions. The developed simulation model which is verified with experimental results can be used to extend the study to higher temperatures to evaluate Vapex performance under a wider range of operating temperature.
1.5 Thesis Structure

This dissertation is organized in six chapters.

The First Chapter is a brief introduction on the Canadian heavy oil and bitumen resources along with the current techniques to extract crudes from these resources. Vapex as the main process under current study is conceptually discussed and research objectives are presented. Structure of the dissertation is also described in this chapter.

Chapter Two provides an extensive literature review on various fundamental aspects of Vapex and discusses different mechanisms and parameters involved in this process. The developed analytical models of Vapex are also reviewed. A wide range of thermal-solvent based methods for heavy oil and bitumen extraction are outlined and compared. This chapter also reviews the reported studies on free fall gravity drainage mechanism as one of the significant mechanism involved in the current research.

Chapter Three provides a detailed description of the experimental Vapex set-up employed in the current study along with the experimental procedure.

Chapter Four includes the presentation and discussion of the experimental results of warm Vapex experiments conducted at elevated temperature and different injection pressures. Effect of operating condition (Temperature and Pressure) on thermo-physical properties of bitumen-solvent system and consequently on Vapex oil drainage rate is discussed. Experimental production rates from conducted warm Vapex tests are compared against earlier developed analytical models and empirical correlations.

Chapter Five discusses the numerical simulation studies conducted to history match the results of warm Vapex experiments. The method used to characterize the utilized Athabasca bitumen is described and the developed PVT model used in the numerical simulation model is thoroughly discussed. The verified simulation model is used to extend the observed experimental trend to a wider temperature range.

Chapter Six presents the concluding remarks on the performed experimental and numerical studies, which were aimed at evaluating the effect of incorporating modest reservoir heating on Vapex performance.
Chapter Two: Literature Survey

2.1 Vapor Extraction

2.1.1 History of Solvent Injection for Heavy Oil/Bitumen Extraction

Several solvent based recovery methods were proposed in 1970s to extract heavy oil and bitumen by injecting pure or mixture of solvents in liquid or vapor form into a single or multiple vertical wells. Allen (1974) proposed injection of low molecular paraffinic hydrocarbons as well as olefinic hydrocarbons, in mixtures or single component, at pressure higher than reservoir pressure into vertical wells. A cyclic injection and production pattern was suggested in which during the injection period heavy oil was absorbing the solvent and upon pressure decline during the production period, oil was produced through solution gas drive and volume expansion mechanisms. Allen and Redford (1976) proposed co-injection of a liquid solvent and a non-condensable gas to extract bitumen from tar sand deposits where the overburden is too thin for thermal flooding and too thick for strip mining. Authors discussed that injecting a solvent which remains entirely in the liquid phase at reservoir temperature and pressure results in formation of a diluted-bitumen bank moving toward the production well. At the far end of the bank, bitumen is almost immobile under the available pressure gradient and plugs the flow. Allen and Redford suggested that injecting a non-condensable gas along with the solvent can prevent the formation of viscous bitumen block by maintaining a two phase solvent-gas mixture flow. Nenninger (1979) suggested injection of a solvent (Carbon dioxide or ethane) at ambient reservoir temperature and at pressure equal or slightly less than the liquefaction pressure. According to Nenninger, after substantial viscosity reduction of oil by dilution, oil can be recovered under the action of a driving force such as gravity, differential gas pressure or a displacement medium. Hanna and Redford (1981) suggested that steam flood for heavy oil applications can be modified by simultaneous injection of a hydrocarbon diluent and a non-condensable gas under reservoir condition to mobilize and produce heavy oil. Naphtha as the diluent and carbon dioxide and oxygen as non-condensable gas were suggested for co-injection with steam to enhance the ultimate recovery.
These solvent based recovery methods appeared not to be economical in viscous heavy oil and bitumen reservoirs due to slow solvent penetration, ineffective oil mobilization and low rate production under vertical well configurations. After technological breakthrough in horizontal well drilling, Dunn et al (1989) proposed bitumen recovery by gravity drainage using low temperature soluble gas injection. The developed theory was an extension of thermally assisted gravity drainage theory and was investigated experimentally in a scaled test element using cold carbon dioxide and ethane with Athabasca bitumen. Although the experimental rates appeared to be higher than the predicted rates based on the developed mathematical model, the preliminary economic assessment of the process indicated that mass transfer assisted gravity drainage process with soluble gas would be more expensive than the steam versions. Solvent assisted gravity drainage was further investigated by Butler and Mokrys (1989) which lead to development of Vapor Extraction (Vapex).

2.1.2 Early Development of Vapex

Butler and Mokrys (1989) promoted the idea of Vapex as an alternative to SAGD to ameliorate some of the economic and environmental limitations associated with SAGD by replacing steam with a solvent. Solvent-leaching gravity drainage of bitumen with toluene was tested in a line source vertical Hele-Shaw cell, filled either with Athabasca or Suncor bitumen, at ambient pressure and temperature. The effect of different reservoir and fluid properties including gravity head, permeability, oil viscosity and solvent diffusivity on the rate of oil drainage were experimentally evaluated. Experimental oil drainage rates were found to be proportional to square root of gravity head and permeability. However oil drainage rates by solvent leaching appeared to be significantly lower than achievable rates from SAGD due to orders of magnitude difference between molecular and thermal diffusivities. Butler and Mokrys used the analogy between solvent assisted gravity drainage and SAGD to develop a theory of solvent chamber growth and bitumen drainage rate by substituting conductive heat transfer with diffusive mass transfer. The observed experimental trends were found to be aligned with the trends predicted from the developed theory.

In the next step, the liquid solvent was replaced with vapor solvent to enhance oil drainage rate by providing higher density difference between solvent and diluted oil and consequently higher
gravity driving force. For this purpose, Butler and Mokrys (1991) conducted hot Vapex experiments in a point source Hele-Shaw cell which involved simultaneous injection of hot water (40-70ºC) and vapor propane. This idea was re-evaluated in a visual cell packed with glass beads and saturated with Lloydminister (Tangleflags) heavy oil. Experimental drainage rates appeared to be notably higher than rates obtained with hot water injection alone. To investigate Vapex viability without hot water injection, Butler and Mokrys (1993) tested pure propane injection just below the dew point in the same packed model. The propane injection by itself resulted in rates comparable to those of the hot water and propane co-injection, and produced oil was observed to be upgraded due to deasphalting. Authors also assessed injection of ethane as solvent and concluded that the results were less promising and no deasphalting was observed.

Upon the observation of traces of deasphalting in earlier performed experiments, Mokrys and Butler (1993) investigated effect of injected gas-oil ratio, propane injection pressure and vapor chamber temperature on Vapex performance and quality of the produced oil. Oil drainage rate were found to be improved by increasing the operating pressure to very close to propane vapor pressure at the test temperature and produced oil appeared to be in-situ upgraded. Butler and Mokrys discussed possibility of extensive asphaltene precipitation, impairment of fluid flow and lower drainage rates in case of liquid propane injection.

Das and Butler (1994) carried out many Vapex experiments using butane as the solvent and Peace River /Lloydminster bitumen in Hele-Shaw cell as well as a packed cell of low to high permeability to assess how the production rates compare to each other. The obtained extraction rates in packed cell were found to be 3 to 5 times higher than rates predicted from experiments in Hele-Shaw cell using transmissibility scale-up method. Das and Butler (1994) also monitored rising, spreading and falling phases of chamber development by adjusting injection/production well configurations.

Das and Butler (1995) evaluated possibility of performing Vapex at higher operating pressures by injecting a non-condensable carrier gas along with a solvent. The suggested process was named Butex and involved injecting liquid butane along with high pressure nitrogen. Near the injector, the liquid solvent was evaporated and carried by the carrier gas to the bitumen interface. Injected solvent mobilized the bitumen under the same mechanisms as of Vapex while the
extracted region was occupied with non-condensable gas which maintained the operating pressure. The operating pressure appeared to have no significant effect on production rate. However, the production rates were found to depend on the solvent injection rate and a minimum amount of solvent was required to render the oil mobile. Very high injection rate of solvent was found to have a negative effect on oil recovery due to accumulation of liquid solvent in the chamber.

Das and Butler (1998) investigated the mechanisms leading to increased production rates observed during earlier Vapex experiments conducted in packed physical models. The authors attributed the unexpectedly higher rates to mass transfer enhancement in the presence of porous media. Different mechanisms including increased interfacial contact, increased effective diffusivity, and increased solubility at the curved vapor-liquid interface, capillary imbibition and surface renewal were conceptually discussed as contributing factors to mass transfer enhancement and consequently increased production rates in porous media.

Butler and Mokrys (1998) studied possibility of recovering and recycling the solvent during Vapex process by modifying the earlier Vapex apparatus and including a solvent recycling loop. The main element of this loop was a stripper which heated the recovered live oil, boiled it off and recycled the saturated vapor. According to that study, solvent consumption was reduced to 74% through closed-loop extraction.

During several years of research, different aspects of Vapex process have been investigated to understand the fluid flow and mass transfer mechanisms involved in the process as well as the role played by parameters that can be adjusted to improve the process. In the following sections the past research aimed at studying the effects of reservoir characteristics and solvent-oil system properties on Vapex performance will be briefly reviewed.

### 2.1.3 Effect of Reservoir Properties on Vapex Performance

During further development of the Vapex process, many researches have been performed to investigate effect of reservoir characteristics and well configurations on recovery rate. Some studies were aimed at evaluating the possibility of applying Vapex to problematic reservoirs
where thermal recovery methods fail to operate due to large heat and steam losses to the underlying aquifers or overlying gas caps. In this section these studied will be briefly discussed.

Jiang and Butler (1996) experimentally evaluated the effect of reservoir heterogeneity on Vapex performance by incorporating layers of alternating high and low permeability and discontinuous low permeability lenses in a sand-packed physical model. Vapex experiments involving sandpacks with low permeability layers resulted in lower production rates compared to the experiments involving uniformly high permeability packing. Rate of oil production in presence of discontinuous lenses of low permeability were higher compared to continuous ones since solvent vapor could extend around the lenses during rising phase of chamber and reach to the high permeability zones above the lenses while diluted oil was counter currently draining. The solvent chamber development through the less permeable layers was observed to be considerably slower since the ingress of vapor to fine pores was interfered by capillarity. Creating a vertical fracture between the injector and producer improved the Vapex performance in layered packing by facilitating the establishment of the vertical communication between the layers.

Butler and Mokrys (1998) discussed applicability of Vapex process to reservoirs with underlying aquifers where excessive vertical heat losses, water coning and steam condensation make thermal recovery inefficient. Experimental investigation by Butler and Mokrys (1998) revealed that presence of an underlying aquifer can accelerate establishment of communication between wells provided that horizontal wells were located close to oil-water contact separated by a lateral distance. Being almost insoluble in water, the injected vapor solvent was spread directly underneath the oil formation, overriding the water layer and simulating a planar well pattern. Under this configuration, the solvent-oil contact area was greatly increased. At the solvent-oil interface, solvent vapor formed rising fingers. At the boundaries of these penetrating fingers, solvent dissolved in bitumen and the diluted bitumen was drained counter currently by gravity. Oil production rate through upward leaching was found to be stable until the interface reached the top of the pay zone. Experimental results indicated that oil production rate in presence of the planar well and rising solvent fingers mainly depended on the draining area and was almost independent of the viscosity of original oil. Das and Butler (1996) also performed similar studies and concluded that countercurrent mode of extraction in presence of underlying aquifer in Vapex process resulted in 4-10 times higher rates compared to the stabilized rate observed during
sideway spreading of vapor chamber. The promising experimental results suggested that creating artificial water filled fractures underneath the formation can be beneficial to improve the Vapex performance.

Jiang and Butler (2000) investigated possible improvement to Vapex oil recovery by using widely spaced horizontal injectors and producers. Lateral distance between the wells delayed establishment of communication between the wells. However, the oil production rates in later stages were improved due to the increased interfacial area and enhanced mixing of solvent with the oil. Increasing the lateral well spacing was found beneficial to the process provided that oil viscosity and reservoir characteristics allow establishment of communication between the wells. Authors also tested bottom injection versus top injection and concluded that for conventional heavy oil, top injection results in higher rates both in heterogeneous and homogenous reservoirs (1996).

Karmaker and Maini (2003) evaluated Vapex performance in heavy oil reservoirs containing a gas cap. The experimental results revealed that presence of a small gas cap can be beneficial to recovery by providing a layer of high solvent concentration above the bitumen. This extended layer of gas accelerates the development of inter wall communication within the reservoir and increases the interfacial mass transfer rate considerably. Performing Vapex process in the presence of large gas caps was not recommended since it required considerably higher injection volumes to build up the solvent concentration within the gas cap.

According to Karmaker and Maini (2003), increasing the lateral spacing between injector and producer did not result in rate enhancement in reservoirs with overlying gas cap. The gas-oil contact area was less dependent on the lateral spacing in the presence of gas cap because of the rapid distribution of the solvent in the gas layer. Locating the horizontal injector near the gas cap zone was noted to be beneficial due to accelerating spread of the solvent over the gas cap in the early stages of the process.

Karmaker, Yazdani and Maini (2003, 2005) conducted several Vapex experiments in rectangular and cylindrical physical models of different heights to evaluate the effect of drainage height on production rate. Through these investigations, the transmissibility scale up methods involving proportionality of the rate to square-root of drainage height failed to correctly predict the oil
drainage rates in larger models. The oil rates were found to be proportional to height raised to a power higher than one. This larger effect of height was attributed to the increased convective dispersion that occurred due to the increase in effective height of solvent-oil interface at larger drainage heights. Authors also investigated the effect of sand grain size on magnitude of dispersion in sand pack experiments and concluded that permeability plays insignificant role on dispersive mixing within the permeability range tested in the experimental study.

Moghadam et al (2007) conducted Vapex experiments in sand-packed physical model over a wide range of permeability from ten to hundreds darcy. The experimental results indicated an increase in solvent content of produced diluted oil at lower permeabilities. Authors explained this observation through faster drainage of diluted oil at higher permeability which provided less contact time for the heavy oil to adsorb the solvent. Accordingly at lower permeability, heavy oil and solvent were in contact for a longer period due to lower velocity of draining diluted oil. In a theoretical study, Moghadam et al (2009) assumed a constant thickness for diffusion layer along the boundary of solvent chamber and developed a methodology to determine this thickness. Through this study, heavy oil production rate was derived for solvent chamber’s spreading and falling phases. The thickness of the diffusion layer was explicitly involved in the derived equations and was used as an adjustable parameter to minimize the discrepancy between theoretically calculated and experimentally measured production rates. Consequently, the fitted thickness involved in experiments conducted at different permeabilities were found to remain approximately constant during chamber growth and decreased with permeability.

2.1.4 Effect of Solvent-Oil Properties and Rock-Fluid Interactions on Vapex Performance

As it is evident from the Vapex concept, solvent dissolution into the oil under operating conditions, mass transfer between solvent and oil as well as fluid flow under gravity drainage are the main mechanisms controlling oil production in this process. A variety of experimental and theoretical studies have been performed over the last two decades to elucidate different aspects of these mechanisms, which will be reviewed in this section.

As it was discussed earlier, oil drainage rate in sand packed Vapex experiments appeared to be much higher than the predicted values from Hele-Shaw cells. This enhancement to recovery rate
has been studied by Das & Butler (1995), Boustani & Maini (2001), Oduntan et al (2001), Yazdani et al (2008), Alkindi et al (2008) and Cuthiell et al (2012) and was attributed to convective mixing of the fluid in the sand. Through these studies molecular diffusion was disproved to be the dominant mass transfer mechanism controlling the Vapex process and incorporation of dispersion effect was found to be essential to address the observed high recovery rates. The classic methods for direct determination of dispersion coefficient in porous media using linear core floods cannot be applied to Vapex case due to the complex flow geometry involved in the process. Moreover the drastic variation in the oil viscosity within a short distance perpendicular to the flow direction in Vapex can significantly influence the dispersion effect. To determine the involved dispersion coefficients in Vapex many direct and indirect experimental methods have been attempted. In direct methods composition of the extracted oil samples at different times during Vapex experiments was analyzed. In indirect methods, property variation of the system such as volume, pressures, position of the gas liquid interface or nuclear magnetic resonance were monitored and analyzed. The obtained experimental data were then used along with some simplified mathematical models to extract dispersion coefficients (Fisher et al, 2000; Karmaker and Maini, 2003; James and Chatzis, 2003; Wen and Kantzas, 2005; Kapadia et al, 2006; Upreti et al, 2006; Alkindi et al 2008; Abukhalifeh et al, 2009). Recognizing the contribution of dispersion effect into the total mass transfer involved in the Vapex facilitated history matching the Vapex experimental results in reservoir simulators as well as predicting Vapex behavior based on predictive models by incorporating appropriate dispersion coefficients.

In addition to the mass transfer studies, many studies have been focused on the role of rock and fluid properties in Vapex.

James and Chatzis (2004) used micromodels of pore networks etched on glass plates, which facilitated flow visualization, to elucidate the mechanisms of Vapex at the pore level. The pore invasion by vapor butane was observed to be affected by capillarity involving drainage and imbibition type displacement, coupled with film flow and gravity drainage near the vapor/bitumen interface. The penetration of the solvent into the bitumen was found to be limited to very few pores near the apparent interface. James and Chatzis (2004) also performed Vapex experiments in slabs of consolidated and unconsolidated glass beads saturated with heavy oil.
utilizing butane to evaluate effect of porous medium characteristics on Vapex mechanism. The results indicated that rate of interface advancement depends on permeability, diffusion distance, drainage flow path which all depend on particle size, pore size and aspect ratio. Vapor-oil interface was found to move linearly with time for a given cross-section. The velocity of advancement was observed to be essentially constant over the length of the model except in the top part where the interface was nearly flat.

Yang et al. (2005) investigated the effect of heavy oil/solvent interfacial tension on gravity drainage in the Vapex process through axisymmetric drop shape analysis (ADSA). Interfacial tensions between a heavy oil sample from Lloydminster and four different solvents including methane, ethane, propane and carbon dioxide over a range of pressure below their respective vapor pressure were measured. The measured interfacial tension was found to decrease almost linearly with pressure. The variation of the interfacial tension over the time was found to be very insignificant which indicated that heavy oil-solvent system interface can rapidly reach the equilibrium state, even though the solvent dissolution into the pendant oil drop continued for a long time. Authors calculated Bond number, which is defined as the ratio of gravity to capillary forces for the Vapex process in a physical model of sand pack for all the above mentioned solvents. According to the results, linear decrease in interfacial tension with pressure resulted in an order of magnitude increase in the Bond number. Although the observed increase in Bond number indicates relatively large effect of interfacial tension reduction on rate enhancement in Vapex in physical model, it is not expected to be very effective in typical heavy oil reservoirs because the capillary forces are already small compared to the available gravitational head.

Cuthiell et al. (2006) investigated role of capillary imbibition in Vapex processes recognizing the extended interfacial area of contact and imbibition to be partially responsible for the Vapex rate enhancement in porous medium. They have carried out a series of Vapex drainage experiments in low permeability sand (4.3 darcy) utilizing a CT scanner to monitor the mixing zone at the edge of the developed solvent chamber. According to the CT images along with the obtained PVT results, presence of gas phase in the diluted oil region was confirmed. Authors discussed this as a manifestation of capillary pressure which smears out sharp spatial variations in fluid saturation. Observed capillary cross flow between layers of low and high permeability in the CT images also confirmed this explanation. Authors also measured gas-oil (butane and Aberfeldy
oil) capillary pressure under the same condition as that of performed Vapex drainage experiments both under imbibition and drainage. In next step they numerically simulated the performed Vapex experiments by employing the obtained capillarity and PVT data. They could match the experimental butane uptake rates, diffusion layer thickness and fluid saturation with the developed simulation model without a need for embedding unrealistically large dispersivity values. They have recommended accounting for capillarity rather than utilizing increased dispersivity to justify Vapex enhanced rates in permeability ranges typical of heavy oil reservoirs.

Talbi and Maini (2003) evaluated effectiveness of carbon dioxide as the non-condensable carrier gas for the solvent in obtaining reasonable recovery rates in Vapex. Co-injection of CO₂ with solvent vapor (propane or butane) can improve economic viability of Vapex by reducing the solvent cost and reduce environmental impacts due to the possibility of CO₂ sequestration. CO₂ appeared to be a proper non-condensable gas for Vapex applications due to its higher solubility in oil and greater viscosity reduction compared to other alternatives, including methane. According to the experimental study performed at high operating pressures, injection of mixture of propane and CO₂ resulted in higher recovery rates compared to mixtures of propane and methane. However co-injection of propane and CO₂ can result in formation of multiple liquid phases under the reservoir pressure and temperature which may reduce oil drainage rate by introducing complex relative permeability effects. In further investigations, Badamchi Zadeh et al. (2008) conducted Vapex experiments in which CO₂–propane mixtures of different composition were injected to extract Athabasca bitumen. Accurate phase behavior studies on CO₂-propane-bitumen ternary system were earlier performed to adjust the injected gas compositions in Vapex to avoid multiple liquid phase formation. According to the Vapex results, the highest recovery rate was achieved with pure propane injection however viscosity reduction of bitumen in case of mixture injection was found to be promising.

Etminan et al. (2008) conducted number of Vapex experiments under different connate water saturations in a visual physical model and found connate water saturation to have a measurable effect on Vapex performance. Connate water was found to affect the process both in terms of the shape of the vapor chamber and rate of oil drainage. Vapor chamber was observed to grow faster in lateral direction in the presence of connate water saturation and thicker mixing zone was
observed compared to dry Vapex. This increase in the mixing zone thickness appeared to result from capillarity driven fingering phenomenon. From the recovery point of view, connate water increased the initial rate of oil production, however oil production rate was decreased subsequently.

Roopa et al. (2007) discussed the involved gravity drainage mechanisms in Vapex process based on observations from micromodels. According to this study, two types of gravity drainage flow are involved in Vapex process. First one is the bulk drainage which occurs at the boundary of vapor chamber where the diluted oil saturation is high so that effective permeability of diluted oil can be assumed close to the absolute permeability of the porous medium. Second one is the oil film flow drainage which occurs within the sand matrix of the vapor chamber and is very slow due to very small oil saturation (close to residual). Authors used micromodels to investigate flow within thin films within porous media saturated with three phases. By calculating the spreading coefficient, Roopa et al. (2007) figured water as the wetting phase and oil as the phase spreading on water. Based on the observations, thickness of the oil film decreased with time, oil film was thinner at the top and thicker at the bottom. Oil film was not always continuously connected throughout the model, however it was interconnected across few pore throats, and the connections were continuously forming and breaking. The broken gas bubbles were observed to be trapped in small pore throats by capillarity but as the drainage proceeded they could form continuous phase across a number of pores and move upwards. The average film drainage velocity was found to be approximately 1000 times slower than the average movement in a oil reservoir under viscous displacement.

Haghighat and Maini (2008) performed Vapex experiments in low permeability sand utilizing propane as solvent to investigate effects of in situ precipitation and deposition of asphaltenes on oil drainage rate and quality of the produced oil under different operating conditions. The focus of this study was to determine whether the beneficial effects of asphaltene precipitation will outweigh any formation damage in Vapex operations. According to the experimental results, the produced oil at higher injection pressures was substantially upgraded but the viscosity reduction by asphaltene precipitation did not improve oil drainage rate. In fact effect of viscosity reduction was negated by the accompanying damage to formation permeability. These results obtained from low permeability sand were not in agreement with the earlier observations from Vapex.
experiments in highly permeable medium as discussed by Butler and Mokrys (1991) and Das and Butler (1993). In their Vapex experiments in visual high permeability packed-cells with propane injection, asphaltene was observed to precipitate out in parallel ribbons leaving a record of the drainage process. Although produced oil was upgraded, no severe blockage and impairment of flow was reported.Das (1998) explained this through the fact that deasphalting occurs at the solvent-oil interface where the solvent concentration is the highest. He concluded that precipitates of Asphaltenes adhere to the sand surface in the extracted region as deasphalted oil flows downward and interface moves away. Since the asphaltene deposited sand matrix was already drained, permeability reduction seemed not to cause any impairment to the flow of diluted oil at the new position of the interface. Das mentioned that the occupied pore volume by deposits was less than 20%, which is low enough for the vapor to bypass the deposited asphaltenes and contact fresh bitumen in highly permeable medium.

Muhamad et al. (2012) evaluated possibility of rate enhancement in Vapex by pressure variation over the time. Experiments were conducted in two lab-scale physical models of different heights over a range of permeability and different strategies of propane injection versus time were tested. Injection strategies included: 1. injecting propane at constant pressure and 2. varying pressure by sudden release and re-injection of the vapor propane. In case of constant injection pressure, highest production rate was achieved when injection pressure was chosen very close to the propane dew point at the test temperature similar to the observations by Das et al (1997). In the case of pressure variation versus time, experimental results revealed that oil production could be improved by 20% by inducing long blips in injection pressure but short blips appeared to be more effective with low permeability medium and more viscous oils.

2.1.5 Vapex Mathematical Models

Butler and Mokrys (1989) developed a model to predict bitumen drainage rate and solvent chamber growth in a solvent-leaching gravity drainage process in a vertical Hele-Shaw cell. This proposed drainage rate equation for the solvent-leaching process was a direct analog of the rate from steam-assisted gravity drainage in which mass transfer by diffusion was incorporated rather than heat transfer by conduction. A sharp apparent interface along the solvent chamber was assumed in the theory and bitumen was assumed to be transported within a thin layer adjoining
to this interface (Figure 2-1). Within the diffusion layer, solvent concentration changes rapidly from equilibrium solubility to a minimum amount at which the oil is immobile. The oil production was assumed to occur under pseudo steady state condition implying that rate of solvent diffusion into bitumen was equal to rate of diluted oil drainage. Thus the interface advances uniformly at a constant velocity in the direction perpendicular to the interface and the concentration profile ahead of the interface does not change with time. An element of thickness $d\xi$ was assumed within diffusion layer and flow was assumed to be parallel to the interface and a one dimensional diffusion equation was written as follows:

$$-D \frac{dc_s}{d\xi} = Uc_s$$

EQ 2-1

The force balance equation within the diffusion layer along the flow direction was simplified to:

$$\Delta \rho g \sin(\theta) = \frac{\nu \mu}{k}$$

EQ 2-2

where $U$ is the velocity of the interface, $\nu$ is the average Darcy flow velocity, $\Delta \rho$ is the density difference between bitumen and solvent, $k$ is the permeability and $\theta$ is the inclination of the interface with horizon.

The total volume of drained oil as a result of solvent dissolution was obtained by integrating the flow over the entire thickness of the diffusion layer. The unspecified velocity of the interface movement was related to the amount of drained bitumen as depicted in the Figure 2-1.

Combining the equation for geometry of flow and total oil production equation within diffusion layer, steady state rate of dead oil drainage was derived by Butler and Mokrys (1989) to be:

$$Q = 2L\sqrt{2kg\phi \Delta S_c N_s H}$$

EQ 2-3

$$N_s = \int_{c_{min}}^{c_{max}} \frac{\Delta \rho D (1-c_s)}{\mu c_s} dc_s$$

EQ 2-4

Where $N_s$ is a dimensionless number consisting of the thermo physical properties of the solvent-bitumen system. $C_{min}$ and $C_{max}$ are the minimum and maximum solvent volume fractions along
the diffusion layer. Concentration dependency of \( \Delta \rho \) (density difference), \( D \) (diffusion coefficient) and \( \mu \) (viscosity) of diluted oil within diffusion layer needs to be properly considered in this model.

A comparison of the laboratory experiments conducted in sand-pack models and those conducted in Hele-Shaw cells revealed that obtained drainage rates in sand-packs are much higher compared to the rates in Hele-Shaw cells of same permeability (Das et al, 1998). Increased interfacial contact area, increased effective diffusivity and solubility in addition to capillary imbibition and surface renewal were explained as parameters responsible for enhancement of mass transfer and production rate in porous media by Das and Butler (1998). They modified the above mentioned model to account the porous media effect. Intrinsic diffusion coefficient was replaced with effective diffusion coefficient, defined for applications in porous media as:

\[
D_{\text{eff}} = D \phi^\alpha \quad \text{EQ 2-5}
\]

In which \( D_{\text{eff}} \) is the effective diffusion coefficient and \( \Omega \) is the cementation factor of porous media.

![Diagram of diffusion layer along the solvent-bitumen interface and extracted vapour chamber in Vapex (Butler & Mokrys model, 1989)](image)

**Figure 2-1:** Diffusion Layer along the solvent-bitumen interface and extracted vapour chamber in Vapex (Butler & Mokrys model, 1989)
Heidari (2008) modeled the mass transfer ahead of the interface in Vapex under unsteady state condition using Fick’s second law of diffusion and assuming a constant diffusion coefficient along the diffusion layer. In this approach, Vapex was considered a moving boundary problem and a new coordinate system perpendicular to the interface was defined to immobilize the front. The mass transfer ahead of the interface in the new coordinate system was:

$$\frac{\partial^2 c}{\partial \zeta^2} + \frac{U}{D} \frac{\partial c}{\partial \zeta} = \frac{1}{D} \frac{\partial c}{\partial t}$$

EQ 2-6

where $d\zeta$ is the element thickness within diffusion layer in the new coordinate system.

The Heat-Integral method was used to solve EQ 2-6 and a $n$-order polynomial was considered for the un-steady state distribution of concentration ahead of the interface.

Similar to the Butler et al. model, Darcy’s law was used to obtain fluid flow rate in an element of thickness $d\zeta$ parallel to the interface within diffusion layer. However, to capture the rapid viscosity changes of diluted oil as a result of concentration variation within diffusion layer, power law functionality was assumed:

$$\frac{\nu_o}{\nu_{mix}} = \left(\frac{c}{c_o}\right)^n$$

EQ 2-7

Where $\nu_o$ is kinematic viscosity of bitumen at the interface having concentration of $c_o$ and $\nu_{mix}$ is diluted oil kinematic viscosity at arbitrary concentration of $c$.

The fluid flow equation was integrated over the entire diluted zone thickness to obtain total flow rate. To find the velocity of the interface ($U$), same material balance equation relating drainage rate and interface movement, suggested by Butler and Mokrys was used.

The steady-state bitumen production rate in Heidari model was derived to be:

$$q_b = \sqrt{\frac{2 n k g H D S_o \phi}{\nu_o} \left[\frac{1}{mn + 1} - \frac{c_o}{nm + n + 1}\right]}$$

EQ 2-8

The Heidari model was more accurate in the sense of modeling concentration profile ahead of the interface compared to Butler et al. model; however ignoring the dependency of diffusion
coefficient on concentration within diffusion layer can be mentioned as one of shortcomings associated with this model.

Okazawa (2007) developed another model in which dependency of diffusion coefficient on solvent concentration was considered by assuming power law functionality between the two. This functionality imposed a non-linearity into the governing mass transfer equation (Fick’s second law of diffusion). Okazawa employed power law functionality to model viscosity behavior of diluted oil at different solvent concentration (EQ 2-7) and followed the same approach as that of Butler and Mokry to derive the following equation for steady state rate of dead oil production:

\[ q_B = \frac{2C_Bk g H \Delta S_o \phi}{\left( \frac{m}{d+1} \right) \mu_o^{1+\frac{d}{m}}} \]  

**EQ 2-9**

\[ \overline{C_B} = 1 - c \left( \frac{m + d}{m + d + 1} \right) \]  

**EQ 2-10**

Where \( m \) is the parameter in viscosity power law functionality to concentration, \( d \) is the parameter in diffusion power law functionality to concentration and \( \overline{C_B} \) is the average flowing fraction of bitumen.

Badamchi Zadeh (2012) considered Vapex a moving boundary problem and followed the Okazawa et al (2009) approach to account for dependency of mass transfer coefficient on solvent concentration due to huge viscosity reduction of diluted oil. Due to the inclusion of diffusion coefficient variation with solvent concentration in the Vapex process, this model provides an improved estimation of the drainage rate. In this model, a power law relationship between the diffusion coefficient and normalized solvent concentration within Vapex diffusion layer was assumed as follows:

\[ \frac{D}{D_s} = C_D^d \]  

**EQ 2-11**
To resolve the non-linearity in the governing mass balance equation as result of dependency of diffusion on concentration, Badamchi Zadeh defined pseudo concentration and pseudo time terms and applied “Heat Integral Method” to obtain concentration profile ahead of solvent-bitumen interface. By adopting the “average flow fraction of the heavy oil” concept from Okazawa, Badamchi Zadeh developed a correlation for steady state drainage rate of dead oil in Vapex as follows:

\[
q_b = \left( \frac{u_s}{u_{ss}} \right)^{-d} \frac{\phi \Delta S_s D_s}{m} \sqrt{f_{b,ss} \frac{2n(d+1)}{mn+1 + \phi \Delta S_s v_s D_s}}
\]

EQ 2-12

In this equation, \(u_{ss}\) and \(f_{b,ss}\) are the average kinematic viscosity and average bitumen volume fraction of produced live oil under steady state condition. \(u_s\) and \(D_s\) are kinematic viscosity and diffusion coefficient at the solvent-oil interface assuming that interface is at equilibrium with solvent vapour under operating pressure and temperature of the process. As discussed by Badamchi Zadeh, kinematic viscosity of solvent-diluted bitumen can be approximately correlated to normalized concentration as follows:

\[
\frac{u_s}{u} = C_D^m
\]

EQ 2-13

Different values of \(m\) can be obtained by correlating diluted oil viscosities to the normalized concentrations, depending on solubility of propane into the oil at test temperature and injection pressure.

According to this model, \(d\) values can be found by using the following equation with experimentally measured values of \(f_{b,ss}\).

\[
f_{b,ss} = 1 - C_s \left( \frac{mn + d + 1}{mn + n + d + 1} \right)
\]

EQ 2-14

With reliable experimental production data and measured properties of produced live oil samples from Vapex experiments in conjunction with accurate data on bitumen-solvent phase behavior, Badamchi Zadeh model can be utilized to correlate diffusion coefficient to solvent concentration and specifically to determine the solvent diffusivity into bitumen at the solvent interface.
Nenninger and Dunn (2009) developed a purely empirical correlation (EQ 2-15) to predict bitumen/heavy oil production rates for solvent based gravity drainage processes such as Vapex and N-Solv. The data set included 60 individual experimental rate measurements in the literature including 11 crude oils within viscosity range of 90 cp to 800000 cp, permeability range of 1.5 to 5400 Darcies in presence of ethane, propane, butane, CO$_2$ and C$_3$/C$_2$/C$_1$ mixture as solvent.

\[
\text{Mass Flux } \left( \frac{g}{m^2 \cdot hr} \right) = 43550 \left( \frac{k \phi}{\mu} \right)^{0.51}
\]  

EQ 2-15

Where \( k \) is the permeability in Darcies, \( \phi \) is the porosity and \( \mu \) is the viscosity of original dead oil in centipoise.

In Chapter three the above-mentioned mathematical and empirical models will be compared against the experimental results of conducted warm Vapex tests and some of the limitations associated with these developed models will be discussed.

### 2.1.6 Scale-up

Butler and Mokrys (1991) used two scaling conditions to scale up the laboratory experimental results to the field cases. The first scaling condition was obtained by considering the bitumen drainage flow in the reservoir:

\[
Q_b = \frac{V \phi \Delta S_o}{t} = \frac{k}{\mu} \frac{dP}{dH} A
\]  

EQ 2-16

Dividing the above equation by the area and replacing the pressure gradient by \( \rho g \), following equation was obtained:

\[
\frac{H \phi \Delta S_o}{t} = \frac{k \rho g}{\mu}
\]  

EQ 2-17

Where \( H \) is height of the cell or physical model.

Rearranging the equation and substituting \( \mu/\rho \) with kinematic viscosity (v) the following equation was obtained:

\[
kgt = H \phi \Delta S_o v \text{ or: } \left[ \frac{kt}{\phi \Delta S_o H} \right]_{\text{model}} = \left[ \frac{kt}{\phi \Delta S_o H} \right]_{\text{field}}
\]

EQ 2-18
The second scaling condition was obtained from the dimensionless ratio: \( \frac{Dt}{H^2} \) which lead to the following equation:

\[
\frac{Dt}{H^2}_{Model} = \frac{Dt}{H^2}_{Field} \quad \text{EQ 2-19}
\]

Assuming the diffusivity to be the same in the field and the model and inserting the porosity to represent the volume occupied by the rock pores, Butler and Mokrys came up with this equation:

\[
\frac{\phi t}{H^2}_{Model} = \frac{\phi t}{H^2}_{Field} \quad \text{EQ 2-20}
\]

Eliminating time ratios from equations 2-18 and 2-20, EQ 2-21 was obtained:

\[
k_{Model} = k_{Field} \frac{H_F \phi_M^2 (\Delta S_o)_M}{H_M \phi_F^2 (\Delta S_o)_F} \quad \text{EQ 2-21}
\]

For a given bitumen-solvent system at fixed temperature and pressure, assuming constant \( N_s \), the original drainage rate equation for Vapex can be simplified to the following form:

\[
\frac{Q_{Model}}{Q_{Field}} = \left( \frac{L \sqrt{kh \phi^m}}{L \sqrt{kh \phi^m}} \right)_{Model} \quad \text{EQ 2-22}
\]

Which \( L \) is the horizontal well length and \( \sqrt{kh} \) is named reservoir transmissibility factor.

The above scaling method is based primarily only on reservoir transmissibility factor. In the other word by assuming \( N_s \) parameter to be the same under field and model conditions, the effect of convective dispersion is completely neglected. Having conducted several experiments in a wide range of drainage height, Karmaker & Maini (2004) and Yazdani & Maini (2004) showed that transmissibility based scaling up methods seriously under-predicts the results at larger scales. Based on these investigations, Yazdani (2004, 2006) proposed empirical correlations for oil drainage rate in which the possible height dependency of dispersion coefficient, hidden in the dimensionless number \( N_s \), is embedded. To improve the previous scaling up methods, new empirical correlations were suggested to be applied for translation of the experimental results to the filed case:

\[
\frac{Q_{field}}{Q_{model}} = \left( \frac{h_{field}}{h_{model}} \right)^{n} \left( \frac{\sqrt{kh \phi^m}}{\sqrt{kh \phi^m}} \right)_{model} \quad (1.13 < n < 1.17) \quad \text{EQ 2-23}
\]
2.2 Solvent-Heat Hybrid Recovery Methods for Bitumen Extraction

During the years of search for efficient heavy oil/bitumen extraction methods, many efforts were made to assess and resolve the shortcomings inherent in solvent based recovery methods like Vapex as well as to overcome the drawbacks associated with thermal based methods like SAGD. The need for huge solvent supplies and losses due to solvent entrapment in the reservoir puts economic viability of Vapex in question. On the other hand, usage of large volumes of fresh water, costly post-production water handling and intensive energy consumption to generate steam, leading to high CO₂ emissions brings many economical and environmental concerns to SAGD. Considering these facts, many researchers have recently championed the idea of solvent/heat hybrid recovery methods to bypass these limitations. Solvent/heat hybrid recovery methods, which are transition methods between SAGD and VAPEX, cover the spectrum of light to full solvent concentration in injection stream. Through these methods, it is hoped to capture the benefit of higher oil recovery rate from thermal recovery based methods and the benefits of saving energy and reducing environmental impacts from the solvent based recovery methods. These transition methods are mostly implemented under same well configuration as of SAGD and use gravity drainage as the main drive mechanism to produce oil.

In the following sections, wet Vapex, SAS, ES-SAGD, SAP, SAVEX, LASER, warm Vapex and N-Solv will be reviewed as examples of hybrid recovery methods.

2.2.1 Vapex with Hot Water

In early stages of Vapex development, Butler and Mokrys (1991) tested simultaneous injection of hot water and vapor propane into Hele-Shaw cell and sand-packed physical models. In those experiments, the temperature and flow rate of water were chosen so that the reservoir temperature was raised to the range of 40 °C- 80 °C. At that point viability of Vapex without incorporation of heat was considered questionable. Hot water- solvent vapour injection allows internal recycling of solvent within the chamber. The hot water contacts the draining diluted oil at the bottom of the chamber, causing the dissolved propane to evaporate and rise up toward the top of the chamber. Therefore, during lateral development of the chamber, the propane acts like heat carrier and causes the heat to be distributed inside the chamber, which results in more
viscosity reduction than what would be possible by dilution only. The experiments revealed that a far higher recovery could be obtained compared to hot water alone. Moreover due to moderate temperature increase within the chamber, heat losses involved in this process were much lower than with conventional steaming.

2.2.2 Thermal Solvent Reflux

Thermal Solvent Reflux method is a hybrid method very similar to Vapex in which the well bore is heated by means of an electrical heater, steam or glycol loop. Heat may also serve to initiate communication between injector and producer. Similar to Vapex, solvent vapour moves to the oil interface at the edge of the vapour chamber and dissolves in the oil. The diluted oil is reduced in viscosity and flows down the edge of the vapour chamber to the production well. The provided heat vaporizes the injected solvent in the vicinity of the heated producer and the solvent, which is driven out of the oil, will return to the vapour chamber, where it will mobilize additional oil (Figure 2-2). Refluxing reduces the volume of the solvent that must be pumped to the surface with the oil, and also reduces the volume of the solvent to be separated, purified, stored and re-injected from the surface (Frauenfeld et al, 2005).

![Figure 2-2: Schematic of Vapor Chamber in Thermal Solvent Reflux Process (After Ivory et al, 2010)](image-url)
Frauenfeld et al. (2006, 2008, and 2010) conducted several scaled laboratory experiments using UTF bitumen/ Hillmond heavy oil and propane as solvent to evaluate thermal solvent reflux process. The producer was heated by means of electrical heater or steam injection to 180 °C and production rates were compared with the rates form a Vapex experiment of same pressure as the base case. According to the results, solvent reflux improved scaled oil production rate by 32% compared to Vapex. Production well heating by means of steam injection and electrical heater lead to similar recovery rates in solvent reflux process. Economic analysis by the authors indicated that the expected cost for solvent reflux method would be 27% lower per barrel of oil compared to Vapex however 35% higher cost per barrel of oil was estimated compared to SAGD.

2.2.3 Combined Steam and Vapour Extraction Process (SAVEX)

SAVEX is another solvent/heat hybrid method patented by Gutek et al. in 2001. In this method primarily steam is injected into the reservoir under same well configuration as SAGD to mobilize and recover at least a fraction of the reservoir hydrocarbons. SAVEX starts with SAGD phase to establish communication between the producer and injector over the entire length of the horizontal well. Starting the process with steam injection helps to create a chamber around the injector, while initial asphaltene precipitation takes place some distance away from the wellbores. Steam injection continues until the recovery rate of the hydrocarbon is approximately 25% to 75% of the peak predicted recovery rate using SAGD. With this rule of thumb, heat loss to overburden can be considerably reduced since the created steam chamber does not reach the top of formation. Moreover the created chamber will be large enough to sustain the reasonable solvent induced drainage rate. At this point, a viscosity reducing hydrocarbon solvent is injected below its vapour pressure to mobilize and recover additional hydrocarbon from the reservoir. Steam injection phase may continue for a while after start of solvent injection or stopped right after solvent injection start up. Although rate of oil production in SAVEX is lower than that of SAGD and bitumen/heavy oil recovery rate during solvent injection phase converges to the stand alone rate of Vapex, the convergence to the lower rate is delayed by the influence of the retained heat in the chamber from the steam injection phase (Figure 2-3). This thermal effect supplements
the viscosity reduction caused by oil dilution with solvent injection and improves oil recovery while reducing SOR at the same time.

![Figure 2-3: Liquid Production Rate in SAVEX in Comparison to SAGD (Gutek et al, 2001)](image)

### 2.2.4 Steam Alternating Solvent Process (SAS)

The SAS process involves alternating injection of steam and solvent, using a well configuration similar to that in SAGD. The basic idea behind this process is to replace a large amount of the required steam with a solvent to minimize energy input per unit of recovered oil (Zhao, 2004). The operating strategy within SAS process includes the following steps (Figure 2-4):

1. **Pure steam injection at start up**

2. **Interrupting steam injection and starting solvent injection as steam chamber grows to the top of pay zone and heat losses to the overburden become significant.**

3. **Interrupting solvent injection and starting steam injection as soon as the chamber temperature declines leading to an uneconomical recovery rate.**

4. **Repeating steam and solvent injection in alternating sequence.**

5. **Depleting the chamber pressure at the end to recover the retained solvent.**
Figure 2-4: Sketch of the SAS Process Injection Pattern (Zhao et al, 2004)

Although oil viscosity is higher in SAS in comparison to SAGD due to lower operating temperature, it can be partially compensated by dissolution of the solvent in the oil. The criterion for solvent selection in SAS is that the dew point temperature of the injected solvent should be between the initial reservoir temperature and the steam temperature under the operating pressure. Since in some regions away from the wells, temperature becomes lower than the dew point of the solvent at the injection pressure, a portion of the injected solvent condenses and this solvent condensation results in localized liquid-liquid mixing of solvent and oil. It is also claimed that mass and heat transfer at the chamber boundary would be enhanced in SAS due to the disturbance of solvent/steam distribution within the chamber, caused by alternating steam and solvent injection (Zhao et al., 2004). According to Zhao (2007) the relatively higher bitumen/solvent mixture viscosity in SAS can be also compensated to some extent by higher oil-phase relative permeability caused by higher oil saturation at the boundary of chamber.

PVT properties of the solvent-oil system under SAS, mixing of solvent vapour and steam in the gas chamber and diffusion/dispersion of solvent in oil phase were reported as parameters affecting heat and mass transfer mechanisms, and consequently the production rate. Retention of the solvent in the reservoir appeared to control the economic performance of SAS, which is dictated by the temperature and concentration distribution within the vapour chamber. SAS process efficiency needs to be optimized for a specific reservoir by designing an appropriate
injection pattern (varying the length of the steam-solvent injection periods in cycle) (Zhao, 2007).

An experimental study on SAS was performed by Zhao et al (2004) in a 2-D high pressure/high temperature apparatus with repetitive cyclic injection of a mixture of 98% propane and 2% methane for 50 minutes and steam for 10 minutes over 660 minutes at 218 ºC. The average recovery rate in SAS experiment was found to be lower compared to the rate obtained in SAGD base case at the same temperature. However the validated numerical simulation model developed by Zhao (2004) revealed that production rates in SAS process could be as high as that of SAGD at the field scale with 34% less energy consumption.

2.2.5 Expanding –Solvent -Steam-Assisted Gravity Drainage (ES-SAGD)

ES-SAGD is an enhancement of SAGD in which a hydrocarbon additive at low concentration is co-injected with steam into the reservoir in a gravity-dominated process. According to the experimental investigation reported by Nasr et al (2003), the hydrocarbon with closest saturation temperature to that of steam (±15 ºC) is the best additive candidate for ES-SAGD. Selecting the hydrocarbon additive in such a way that it would evaporate and condense at the same conditions as the water phase, helps in simultaneous condensation of the solvent with steam at the boundary of steam chamber. Condensed solvent around the interface of the steam chamber dilutes the oil and, in conjunction with the transferred latent heat, reduces the oil viscosity. Nasr et al (2003) reported that with the increase in the carbon number of the injected hydrocarbon and consequently increase in the vaporization temperature of the hydrocarbon additive, rate of oil drainage will be improved. Among the injected additives, hexane, having the closest condensation behaviour to steam, resulted in the maximum oil drainage rate (Figure 2-5).

Practically, the co-injected solvent will be a hydrocarbon mixture such as naphta or diluents (C₄-C₁₀) due to its availability and lower cost compared to pure hydrocarbons. ES-SAGD can potentially use less water and require smaller water handling and treatment facilities than that in SAGD in addition to the possibility of recycling and re-injecting the solvent. Designing an optimum solvent injection strategy (by adjusting the injection pressure and solvent fraction in the injected steam) for ES-SAGD operation can ensure high energy efficiency and solvent recovery (Gates and Chakrabarty, 2008).
Figure 2-5: Variation of Oil Drainage Rate in ES-SAGD with Carbon Number (After Nasr et al, 2002)

In an experimental study by Ayodele et al (2009) using a 2-D high pressure/temperature physical model, performance of low pressure ES-SAGD was compared with low pressure (500 kPag) and high pressure (2100 kPag) SAGD. In the low pressure ES-SAGD experiments, pure propane or a multi-component diluent at two different concentrations were co-injected with steam. According to the results, the low-pressure multi-components solvent ES-SAGD tests resulted in higher recovery rates compared to low pressure SAGD and low pressure propane-SAGD. The recovery of high concentration ES-SAGD test was higher than low concentration test at the early times of experiments however it was declining over the time. Production rate in high pressure SAGD appeared to be considerably higher than other conducted tests however energy consumed per oil recovered was estimated to be 30% higher compared to the other cases. No results on higher pressure ES-SAGD were reported.

Hosseininejad et al (2012) performed several SAGD and ES-SAGD experiments under different operating conditions in a 3-D scaled physical model co-injecting 1 mole% of hexane with steam. ES-SAGD and SAGD performance at low and high pressures (1500 kPa and 3500 kPa) were evaluated based on production rate and steam-oil-ratio profiles. Hosseininejad (2012) reported that hexane co-injection at low pressure resulted in 36% and 21% improvements to recovery and SOR compared to SAGD after 10 hours of experiment. 28% improvement to recovery and 30% improvement in recovery and SOR was reported comparing high pressure ES-SAGD to high
pressure SAGD. Rate enhancement by hexane was found to be higher at lower operating pressure however SOR improvement appeared to be slightly less. The obtained profiles of vapour chamber indicated that steam chamber growth was promoted by hexane at low pressure while it was slightly degraded at higher pressures. The post sand analysis revealed that hexane injection lowered the residual oil saturation from 2.8% to 0.4% in 250 Darcy sand.

2.2.6 Solvent Aided Process (SAP)

SAP is another hybrid process similar to ES-SAGD in which a small amount of lighter hydrocarbon solvent (such as propane, butane, or pentane etc) is added to steam in a steam assisted gravity drainage process for improving the performance of SAGD (Gupta et al, 2003). The added solvent to the steam in SAP remains in the vapour form under reservoir pressure and temperature and does not condense with steam at the boundary of the chamber although it causes additional viscosity reduction in oil through dissolution. SAP seems to be able to maintain economic rates for longer duration by reducing the heat loss to the overburden due to the lower ceiling temperature within the chamber and by increasing the oil drainage rate (Gupta et al, 2003). To capture the benefit of controlling the heat loss in SAP, timing of the solvent injection is the key consideration. Gupta and Gittins (2005) suggested that early solvent injection in SAP should be avoided since solvent vapour which rides over the steam saturated region acts like an insulating layer, leads to lower temperature and higher viscosities at the top of chamber and causes the chamber to develop in sideways more rapidly than vertically. To take advantage of the entire height of the reservoir and maximum driving force for the gravity drainage, steam injection alone can be more beneficial during rising chamber phase of the process. Once the chamber reaches top of the formation solvent co-injection can be started to promote faster lateral growth of chamber.

SAP has been pilot tested by Encana at Senlac thermal project and Christina Lake project. The field performance results from Christina Lake project during SAP and pre-SAP (SAGD) processes are illustrated in Figure 2-6 which confirm a significant improvement in oil production rate and steam-oil-ratio after switching from SAGD to SAP within a short period of six months (Gupta et al, 2005).
Gupta and Gittins (2006) investigated possible enhancement to SAP performance by solvent sequencing and cross flow methodology to recover solvent. Solvent sequencing was suggested to be performed by injecting successively reduced molecular weight solvents to recover the heavy solvents at the cost of lighter ones. Authors also suggested that cross flow of the natural gas between adjacent SAP patterns can efficiently recover most of the injected solvent during the process.

Figure 2-6: EnCana’s SAP Field Performance at Christina Lake Project (Gupta et al., 2005)
2.2.7 Liquid Addition to Steam for Enhancing Recovery (LASER)

LASER is a variation of CSS which involves injection of a liquid hydrocarbon (C5+) as steam additive in cyclic mode of operation. Application of the LASER process is preferable in later CSS cycles when the SOR is increasing because of the decline in the contribution of some existing drive mechanisms such as formation compaction, solution gas drive and steam flashing in oil mobilization. In late cycles of CSS when gravity drainage is believed to be the dominant recovery mechanism, solvent injection can be beneficial to improve the oil recovery rate (Leaute, 2002). Injected solvent under this circumstance behaves similar to the injected solvent in ES-SAGD process. Based on the promising results of LASER over the conventional CSS in initial laboratory tests by Leaute (2002), LASER was pilot tested in Cold Lake by Imperial Oil in 2002.

The field pilot test of LASER which involved co-injection of steam and 6% by volume of diluent resulted in 33% increase in oil-steam-ratio and 66% increase in recovery compared to CSS.

Figure 2-7 illustrates the improvement in bitumen that was achieved as a result of LASER intervention in cycle 7. According to this Figure, the result from LASER evidently surpassed that of reference CSS wells in the cycle. Leaute and Carey (2007) reported 80% recovery of the injected diluent from the LASER cycle and suggested dynamic refluxing of the solvent fractions as a plausible positive attribute of cyclically-operated thermal solvent recovery process, such as LASER.

![Figure 2-7: Comparison of LASER and CSS Pilot Performance (Leaute and Carey, 2007)](image-url)
2.2.8 Enhanced-Solvent Extraction (N-Solv) and Warm Vapex

The N-Solv process is a variation of Vapex in which a heated solvent vapour is injected at dew point into the reservoir. Solvent latent heat of condensation in association with its dissolution into the oil is used to reduce the bitumen viscosity and produce it via gravity drainage (Nenninger et al, US Patent, 2008). Due to higher solvent vapour pressure at higher temperatures, N-Solv can be applied to higher pressure reservoirs. N-Solv is also claimed to be less vulnerable to non-condensable gas contaminations (which may be co-injected with solvent or liberated from the live oil under reservoir condition) in comparison to Vapex. Accumulation of non-condensable gases within the chamber may interfere in mass and heat transfer mechanisms at the chamber boundary if the non-condensable gas is not properly vented. Condensation of the solvent at the oil/solvent interface during N-Solv, is said to provide the advantage of carrying most of the non-condensable gas to the production well and preventing its build up within the chamber (Nenninger et al, 2009). Nenninger et al (2009) discussed necessity of injecting pure solvent in N-Solv based on thermodynamic calculations that revealed significant effect of methane contamination on impairing the ability of the solvent to deliver latent heat to bitumen interface. Accordingly, injecting a high purity condensing solvent at moderate temperatures appeared to be 20 times more efficient in removing non-condensable gas from the chamber compared to SAGD.

Nenninger and Dunn (2008) proposed a purely empirical correlation (EQ 2-15) to predict solvent based gravity drainage rates and discussed that small amount of heat, which is provided by solvent condensation at the boundary of the chamber in N-Solv, can greatly accelerate the extraction rate by decreasing the oil viscosity by orders of magnitude. Based on this study, at 60 °C, N-Solv extraction rates were predicted to be three times of SAGD with energy savings of 75% relative to SAGD.

Warm VAPEX, is yet another variation of VAPEX process, which incorporates heat in VAPEX by injecting superheated solvent. The superheated vapour carries some sensible heat to the solvent–oil interface and also causes an additional mixing due to condensation of the solvent at the interface (Rezaei et al, 2010). Rezaei et al (2010) performed warm Vapex experiments by injecting superheated pentane into packed physical models of 220 and 830 Darcies. The models
were saturated with Cold Lake bitumen or Lloydminster heavy oil and the objective was to evaluate the effect of level of solvent superheat on the performance of the process. The packed model was inserted in an isothermal bath of circulating hot air (35 ºC) to lower the heat loss to the ambient. According to the results, higher rates were achieved in warm Vapex process with less solvent consumption compared to conventional Vapex process. The oil production rate reached a maximum at the midlevel solvent superheat while further increase in the degree of solvent superheat was found to cause a decrease in rate of production and solvent oil ratio at the same time. The asphaltene content analysis by Rezaei revealed that asphaltene precipitation was more pronounced in warm Vapex due to solvent condensation at the oil-vapor interface.

2.3 Free Fall Gravity Drainage

Oil drainage under the effect of gravity appears to be significant in certain volumetric oil fields where formation was thick or had tilted beds of high permeability without shale barriers. Lakeview Pool in Midway Sunset oil field is an example of strictly free-fall gravity drainage production since gravity head was the only essential energy left in the reservoir after extended blow out of the discovery well. 65% recovery was achieved by gravity drainage after 40 years of production in this field (Dykstra, 1978). The Hawkins Woodbines field is another example with large vertical permeability in which gravity drainage coupled with gas cap drive was able to recover over 80% of the original oil (King et al, 1970). Many researchers studied free fall gravity drainage as an effective drive mechanism theoretically and experimentally.

Katz (1942) investigated the gravity drainage phenomena according to the oil saturation data from Oklahoma City Wilcox formation as a case study and applied Leverette theory (1941) to discuss the involved interaction of capillary and gravitational forces. Following the study by Katz, Stahl et al (1942) carried out gravity drainage tests in vertical sand-packed tubes to evaluate drainage of water, heptane and Wilcox crude oil (API=38.2 at 60 ºF) from unconsolidated Wilcox sand (average porosity of 32% and average absolute permeability of 7.5 darcy). Authors presented the accumulative drainage curves (Figure 2-8) for the Wilcox crude oil at different temperatures and concluded that depletion of a reservoir through gravitational drainage involves unsteady flow of liquid.
By extracting the core samples along the tube height at different times of experiments, the transient and equilibrium liquid saturation distributions were generated for the case of Wilcox crude oil as depicted in Figure 2-9.

Cardwell and Parsons (1948) developed the original mathematical theory of gravity drainage to apply the laboratory data to the field problems. Theory was developed to estimate rate of gravity
drainage of a liquid out of a saturated vertical sand column where top and bottom of the column where open to the atmosphere. Liquid saturation distributions along the height of the column before and after reaching equilibrium are presented as curve B and A respectively in Figure 2-10. According to equilibrium curve A, there is a region of 100% liquid saturation at the lower end of the column above which liquid saturation gradually decreases along the height and on the very top reaches a zone of practically constant saturation.

Figure 2-10: Liquid Distribution in a Column of Porous Media Drained by Gravity (after Cardwell et al, 1948)

Any other distribution curves similar to B are unstable and would ultimately converge to the stable curve A. Cardwell and Parsons solved the governing equations of flow and continuity for the fully saturated zone and partially saturated zone separately. The fluid flow equation in its general form can be written for liquid drainage based on the Darcy’s law as follows:

\[
\nu = \frac{k}{\mu} \rho g - \frac{k}{\mu} \frac{\partial p}{\partial z}
\]

EQ 2-16
Where $k$ is permeability, $\rho$ is liquid density, $\mu$ is liquid viscosity, $p$ is local pressure and $z$ is the distance from the bottom of the column.

In the partially saturated region, pore space is partly saturated with liquid and the rest is filled with the air. Considering the vertical pressure gradient of the gas to be negligible, the liquid pressure gradient in this zone would be only a function of liquid saturation.

Obviously permeability in the above equation is also a variable depending on liquid saturation.

In the fully saturated region, there is no pressure gradient as a result of saturation gradient, however there must be a pressure gradient resulting in liquid retention. Cardwell and Parsons suggested that this pressure gradient in the liquid hold-up region after reaching the equilibrium can be calculated as follows:

\[
v = \frac{k}{\mu} (g\phi - \frac{\partial p}{\partial z}) = 0 \quad \Rightarrow \quad \frac{\partial p}{\partial z} = \rho g \quad \text{EQ 2-17}
\]

This pressure gradient arises from the pressure drop between liquid and gas across the boundary of partially and completely saturated regions (demarcator), which acts like a negative pressure at the interface. Substituting $H'$ (equilibrium liquid hold-up) in the above equation, the negative pressure must be of the magnitude $\rho g H'$. Considering this negative pressure ($\rho g H'$) which exists at the top of the demarcator, the flow equation for fully saturated zone can be written as:

\[
v_s = \frac{k}{\mu} (\rho g - \frac{\rho g H'}{h}) \quad \text{EQ 2-18}
\]

where $v_s$ is the flow velocity in the saturated region

Flow equation for the partially saturated zone is:

\[
v_u = \frac{k}{\mu} \rho g - \frac{k}{\mu} \frac{\partial p}{\partial z} = \frac{k}{\mu} \left( \rho g - \frac{\partial p}{\partial z} \cdot \frac{\partial S}{\partial z} \right) \quad \text{EQ 2-19}
\]

where $v_u$ is the flow velocity in the unsaturated region and $S$ is the liquid saturation.

To find the time variation of liquid saturation in the upper region, the following continuity equation is required:
\[ \frac{\partial v_u}{\partial z} = -\phi \frac{\partial S}{\partial t} \quad \text{EQ 2-20} \]

Assuming the dependency of the capillary pressure on saturation to be negligible at intermediate and low saturation ranges, EQ 2-19 and EQ 2-20 can be combined to the following differential equation:

\[ \frac{\partial S}{\partial t} = -\frac{\rho g}{\mu \phi} \frac{dk}{dS} \frac{\partial S}{\partial z} \quad \text{EQ 2-21} \]

Considering the column to be originally fully saturated with liquid, the solution of the above equation would be:

\[ z = \frac{\rho g}{\mu \phi} f(S)t \quad \text{EQ 2-22} \]

Where

\[ f(S) = \frac{dk}{dS} \quad \text{EQ 2-23} \]

And liquid saturation can be found from:

\[ S = f^{-1}\left(\frac{z \mu \phi}{\rho gt}\right) \quad \text{EQ 2-24} \]

which \( f^{-1} \) is the inverse of the \( f \) function.

With the capillary pressure term omitted in EQ 2-19, liquid flow velocity in the unsaturated zone can be calculated from:

\[ v_u = \frac{k}{\mu} \frac{\rho g}{\phi} \quad \text{EQ 2-25} \]

The velocity of the demarcator would be:

\[ \frac{dz_d}{dt} = \frac{v_s - v_u}{\phi(1 - S_d)} \quad \text{EQ 2-26} \]
where \( z_d \) is the position of the demarcator, \( v_u \) is the fluid velocity in the unsaturated region in an infinitesimal distance above the demarcator and \( S_d \) is the saturation at an infinitesimal distance above the demarcator. Substituting EQ 2-18 and EQ 2-25 into EQ 2-26, and considering \( L \) as the length of the draining column following equation was obtained:

\[
\frac{dz_d}{dt} = \frac{\rho g}{\mu \phi} \left( k_i \left( 1 - \frac{H}{L - z_d} \right) - k_d \right) \tag{EQ 2-27}
\]

where \( k_i \) is the absolute permeability and \( k_d \) is the effective permeability at the saturation at an infinitesimal distance above the demarcator.

Location of the demarcator can be found at any time by integrating EQ 2-27.

It is evident that the empirical permeability-saturation relationship and empirical liquid-hold up height (\( H \)) are required in this methodology to calculate the position of demarcator, saturation distribution along the column and rate of liquid drainage at any time. The authors reasonably verified recovery versus drainage time from this mathematical theory with earlier mentioned experimental data from Stahl et al (Wilcox crude at 130 °F). However the saturation history curves from the mathematical model appeared to be slightly different from actual saturation data due to neglecting capillary pressure gradients in the unsaturated region.

Dykstra (1978) generalized the approximate theory by Cardwell et al to account for immobile gas saturation at the start of gravity drainage and for oil relative permeability to be zero at residual oil saturation rather than at zero oil saturation which was missed in the theory by Cardwell et al. Incorporating these modifications, Dykstra derived the gravity driven recovery equation as follows:

\[
R = \frac{z_d}{L} \left[ \left( 1 - \frac{B_{oi} S_{oi}}{B_o S_o} \right) - \frac{B_{oi}}{S_{oi}} \left( 1 - S_{oi} \right) \left( A - 1 \right) \left( \frac{p z_d}{A t} \right)^{\frac{1}{A-1}} \right] \tag{EQ 2-28}
\]

Where
\[ p = \frac{\mu \phi (1 - S_{or})}{A k \rho g} \]  

EQ 2-29

Where \( A \) is the empirical exponent in relative permeability-saturation exponential function, \( B_o \) is oil-volume factor, \( z_d \) is the demarcator location from the top of the draining column, \( L \) is length of the draining column, \( S_{oi} \) is initial oil saturation, \( S_{ro} \) is residual oil saturation and \( k \) is absolute permeability.

By comparing the calculated recovery rates based on EQ 2-28 with seven sets of observed experimental recovery rates of gravity drainage from Stahl et al (1942) and King (1899), Dykstra showed that a shift along the time scale causes the curves to overlay. Accordingly, Dykstra suggested combining rock and fluid properties in the form of a “drainage modulus” which can be used to generate a universal recovery curve. For this purpose, he used the model from Elkins et al (1948) for oil downward flow in a rectangular, dipping formation (Figure 2-11) into a well under free-fall gravity drainage as follows:

\[ q_o = \frac{0.001127 k_o h W C (\rho_o - \rho_g) \sin \alpha}{\mu_o B_o} \]  

EQ 2-30

Where \( q_o \) is rate of oil flow (stock tank barrels of oil per day), \( k_o \) is effective permeability to oil (millidarcy), \( h \) is formation thickness (ft), \( W \) is formation width (ft), \( L \) is length of formation (ft), \( \mu_o \) is oil viscosity (cp), \( \rho_o/\rho_g \) are density gradient of oil/gas (psi per ft), \( B_o \) is formation volume factor of oil and \( \alpha \) is the angle of dip.

The constriction factor \( C \) which was not considered in the original model by Elkins et al (1949), was incorporated by Dykstra to include the constriction in flow because the area open to the flow in the well bore is not normal to the flow direction.

The original oil in place (bbl) at the start of drainage was calculated from:

\[ N = \frac{h W L S_{oi} \phi}{5.615 B_o} \]  

EQ 2-31
The rate of recovery (\% per day) was given by

\[
\frac{dR}{dt} = \frac{100 q_o}{N} = 0.274 \left[ \frac{k_o d_o C \sin \alpha}{\mu_o L S_o \phi} \right]
\]

\text{EQ 2-32}

Where \( d_o \) now is the oil density in gr/cm\(^3\).

The term in parentheses in above equation was called drainage modulus and has the unit of

\( \text{millidarcy} \cdot \text{gr/cm}^3 \) \( \frac{\text{cp}}{\text{ft}} \).

It was shown that the experimental recovery data versus drainage modulus for different systems of rock and fluid fits reasonably along the calculated curves. Dykstra suggested that Figure 2-12 can be applied as a universal dimensionless curve to any system where equilibrium height (H) is about 6\% of the length of the draining column and a simple exponential equation can be used to relate relative permeability to saturation. EQ 2-30 can be used to find the maximum possible gravity drainage rate obtainable from any formation.

\[\text{Figure 2-11: Dipping Formation with Flow Converging into the Wellbore} \quad (\text{Dykstra, 1978})\]

\[\text{Figure 2-12: Recovery as a Function of Drainage Modulus and Time} \quad (\text{Dykstra, 1978})\]
In another study Hagoort (1980) assumed gravity drainage as a displacement process in which gas displaces oil and theoretically discussed vertical displacement efficiencies of forced and free fall gravity drainage. According to Hagoort, reliable and accurate relative permeability data was essential to analyze downward displacement of oil by gas. Hagoort used the centrifuge method to measure oil relative permeability in sandstone and dolomite rocks and confirmed that gravity drainage can be a very effective recovery process in water-wet, connate water bearing reservoirs and can result in reasonably low remaining oil saturations. According to Hagoort, equilibrium oil saturation in the lifetime of the oil reservoir depends on the magnitude of the gravitational forces relative to the viscous forces, shape of the relative permeability curves and reservoir geometry and heterogeneity. He also derived a governing equation for the simple case of a stable, vertical, downward displacement of oil by gas in which capillary pressure was expressed through Leverette J function. The obtained saturation profile during free fall gravity drainage from this model was found to be identical to the saturation profile from Cardwell and Parsons model, provided that capillary pressure was neglected.

Li et al (2003) applied the empirical imbibition model developed by Arnofsky et al (1958) to gravity drainage process. Arnofsky et al model was originally proposed to characterize the spontaneous water imbibition in naturally fractured reservoirs. According to Li et al, free fall gravity drainage rate depends significantly on the properties of the porous media, fluids and their interaction which cannot be easily included in existing analytical models. They suggested matching the gravity drainage production data with an exponential empirical model expressed as follows:

\[
\text{Recovery} = \frac{1 - S_{wi} - S_{or}}{1 - S_{wi}} \left(1 - e^{-\beta t}\right)
\]

\[
\text{EQ 2-33}
\]

Where \(S_{wi}\) is initial water saturation, \(S_{or}\) is the average residual oil saturation after the gravity drainage and \(\beta\) is the converging rate constant. \(S_{or}\) and \(\beta\) can be obtained by a history match technique once the oil production data are available.

Li et al (2003) derived an equation for ultimate oil recovery by equating the capillary pressure force and gravity force at any position in the core/reservoir when the free fall gravity drainage
was completed. The capillary pressure was presented by Brook-Corey model in which capillary entry pressure and pore size distribution parameters were considered. Upon availability of free fall gravity drainage production data, the developed approach can be used to infer the capillary pressure curves by figuring entry capillary pressure and pore size distribution parameters specific to any reservoir and fluid system.

Free fall gravity drainage mechanism needs to be included in the current study due to the specific design of warm Vapex experiments and mobility of original oil in place. The oil becomes mobile enough to drain under the free fall mechanism when the temperature is increased and this drainage mechanism supplements the Vapex induced flow. To isolate the contribution of Vapex in the measured rates, it is essential to subtract the contribution of the free fall gravity drainage from the observed total drainage rate. Due to the large permeability of the utilized sand, capillarity effect can be neglected and free fall gravity drainage can be simply quantified by conducting a free fall drainage test along with a validated numerical simulation.
Chapter Three: Vapex Experimental Set up and Procedure

This chapter describes the experimental apparatus and the experimental methodology used during preparation and implementation of warm Vapex experiments.

3.1 Experimental Apparatus

A simplified schematic diagram of the experimental apparatus is depicted in Figure 3-1 and a photograph of the actual set-up is shown in Figure 3-2. The main component of the set-up was a high pressure physical model in cylindrical geometry that represented a cross section of the reservoir perpendicular to the well pair. Other important components of the apparatus included a metered solvent injection system, a high pressure windowed separator cell for collecting produced live oil and separating it from produced free gas at high pressure, a free gas production monitoring system, a system for withdrawing oil samples from the high pressure separator and flashing them to atmospheric pressure and measuring the released solution gas and an immersion heater (equipped with temperature controller) to adjust the test temperature. The components used in the Vapex apparatus will be individually described in the following sub-sections.

3.1.1 Physical Models

The main physical model used in this study is a vertical slice type model constructed in cylindrical geometry using the annular space between two large diameter co-axial pipes. The two coaxial stainless steel pipes were flanged to two stainless steel discs on top and bottom. The empty annular space between the two pipes was used as the housing of the sand-pack. A view of different pieces of the model is displayed in Figure 3-3 and the assembled physical model is displayed in Figure 3-4. The dimensions of the model are listed in Table 3-1.

Two sets of Viton O-rings were incorporated to provide sealing between the flanges and discs. Four openings for tubing connectors were located on the top lid. Three of the openings were equipped with 3/8 inches Swagelock® connectors with O-ring seals and served as the ports required for the packing and saturating processes. The fourth opening was used for inserting the injection well which was 1/4 inch Swagelock®-tube fitting covered with 200 mesh stainless steel
screen at the bottom end to prevent the sand from flowing out. A 1/8 inch Swagelock fitting was also incorporated on the top disc to install the thermocouple probe.

Figure 3-1: Schematic of the Vapex experimental Setup
Figure 3-2: Vapex Experimental Setup

Table 3-1: Cylindrical Model Dimensions

<table>
<thead>
<tr>
<th>Description</th>
<th>Measurement</th>
</tr>
</thead>
<tbody>
<tr>
<td>Height of the annular cavity</td>
<td>30.48 cm</td>
</tr>
<tr>
<td>Outside diameter of the annular cavity</td>
<td>30.73 cm</td>
</tr>
<tr>
<td>Total bulk volume of the annular cavity</td>
<td>4.9067 litre</td>
</tr>
<tr>
<td>Inside diameter of the annular cavity</td>
<td>27.20 cm</td>
</tr>
<tr>
<td>Width of the annular cavity</td>
<td>1.765 cm</td>
</tr>
</tbody>
</table>
The production well was located directly below the injection well. The production well was a short length (3 cm) of 1/4 inch perforated tubing. The perforated length was covered with 200 mesh stainless steel screen to prevent the sand from coming out of the well. Three other openings were located in the bottom disc directly below the top openings and these were fitted with three stainless steel mesh covered tube fittings.

The model was mounted on a stand via a steel clamp around the outer cylinder. Two mounting shafts were incorporated at the side of the model to facilitate rotating the model during packing or in cleaning phases.

The inner cylinder of the model was filled with water and an immersion heater was installed in it to maintain the water temperature at the desired test temperature. Insulation was wrapped around the outer cylinder to minimize the heat dissipation to the ambient. Considering these, heat distribution can be assumed to be homogeneous along the height and radius of the annular space. Temperature of the sand pack was monitored with a thermocouple, located in the sand-pack, during the Vapex experiments to confirm stable temperature.

![Figure 3-3: Upper View of the Cylindrical Model Components](image)
In experiment #8, a visible rectangular model was employed to be able to monitor the chamber development during this experiment. This visual model (Figure 3-5) was a 2-Dimensional medium pressure cell. This rectangular model consists of a main aluminum body, a transparent Plexiglass plate, acting like a window through which process can be observed, and a metal protector, which is bolted to the main body and the Plexiglass. Sealing is provided with O-rings. The cavity of the cell has the following dimensions:

**Table 3-2: Rectangular Visual Model Dimensions**

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Height of the rectangular cavity</td>
<td>15.2 cm</td>
</tr>
<tr>
<td>Length of the rectangular cavity</td>
<td>67.5 cm</td>
</tr>
<tr>
<td>Width of the rectangular cavity</td>
<td>3.1 cm</td>
</tr>
</tbody>
</table>

Three injection ports on the top and five production ports at the bottom and one port on each side of the rectangular model were incorporated which provided different possibilities for well configuration. In experiment #8, the middle injection port on top and the middle production port
at the bottom were used. With this configuration, a full pattern vapour chamber was created and monitored during solvent injection.

Figure 3-5: Front View of the Visual Rectangular Model

3.1.2 Gas Injection Facility

Propane with 99% purity, stored in a 5 litre tank, was injected as the solvent during the Vapex experiments. Injection pressure of propane was regulated by means of a Praxair 412 dual stage Brass Barstock regulator with the maximum working pressure of 21 MPa (3000 psig) which was equipped with two brass pressure gauges (outlet gauge: 0-400 psig, inlet gauge: 0-4000 psig. The propane tank was placed on a digital balance (Explorer Pro, Ohaus) with 0.001 kg accuracy and 23 kg capacity (Figure 3-6).

The decline in the weight of the propane tank was recorded during the course of experiments which provided a direct measure of volume of injected propane. A pressure gauge with the range of 300 psig was located in the gas injection line to measure and monitor the injection pressure.

In experiments #5 and #6, propane tank was warmed up by means of electrical heating tapes to be able to inject propane at higher pressures at the test temperature. The power applied to the heating tapes was manually adjusted to heat the propane tank and to be able to withdraw propane at the desired injection pressure.

In experiment #7, propane tank was replaced with a nitrogen Praxair tank to supply nitrogen during injection and the earlier pressure regulator was replaced with a nitrogen pressure regulator.
3.1.3 Gas Collection System

A simple system (Figure 3-8) was used to collect and measure the produced solution gas (liberated gas during live oil discharge to the oil collecting jar at ambient condition) and free gas (produced gas from the vapour chamber). This system includes two Plexiglas cylinders (Solution gas: ID=15.25 cm, Height = 90 cm and free gas: ID=12.7 cm, Height = 100 cm) with one end closed. These cylinders were placed vertically in a water tank with their open ends submerged in the water. Both were initially filled with water to the top with help of a vacuum pump connected to the top ends of the cylinders. The produced gas displaced water downwards and the volume of collected gas was monitored with help of graduations on the cylinder.
3.1.4 Oil-Gas Separator

A rectangular high pressure cell capable of holding up to 600 psi pressure was used to collect the drained diluted oil (Figure 3-7). This cell consists of 5 pieces: The main body is a 5 cm aluminum plate with a chamber in the middle, two Plexiglas plates of 2.5 cm thickness which act like the side windows of the separator and two plates of the steel (thickness=1.2 cm) with full-length windows cut out in the middle. A ruler was also installed along the window to measure the level of the accumulated oil inside the separator. Each centimeter on the ruler represents 13 cm$^3$ of the stored fluid inside the separator and the total capacity was measured to be around 700 cm$^3$. Viton O-rings were used for sealing between aluminum plate and the Plexiglas windows. Figure 3-7 presents a picture of the high pressure separator cell used in this work.

Live oil which is produced from the model flows toward the separator through a flexible line which was connected to the separator by means of a 3/8 inch size fitting on top of the separator. This fitting was drilled through and a 1/4 tube was inserted inside which provides the space for the flow of the live oil toward the separator. The produced free gas escaped toward the gas collection cylinder through the annular space between the 1/4 inch tube and the 3/8 inch outer fitting on the top of the separator.

In Vapex experiments at elevated temperatures, the steel plates of the separator were covered with heating tapes and then insulated to keep the stored live oil at the same temperature as of the test temperature. A thermocouple probe was installed into the separator and internal temperature was accurately monitored during the entire experiment.

3.1.5 Oil Collecting System

Produced live oil accumulated in the separator was withdrawn frequently in a glass jar which was held by means of a modified clamp acting as a jar holder (Figure 3-9). A rubber balloon was connected to the system for safety pressure relief. The balloon would burst before the jar in case of high pressure inside the jar. A 1/4 inch tubing connected the top of the oil sampler to conduct the released solution gas to the gas collection facility. Whenever the separator was discharged; the full jar was weighted and replaced with a tared new jar. Sealing was provided with O-rings and vacuum grease around the opening for the jar.
3.1.6 Immersion Heater

A ThermoScientific (Haake SC100) immersion heater including a temperature controller with ±0.02°C temperature stability (Minimum temperature: Ambient, Maximum temperature: 100 C) was installed in the water filled inner cylinder of the physical model to adjust and control the water temperature at the desired test temperature (Figure 3-10).

3.1.7 Piping and Fittings

1/4 inch diameter stainless steel tubes and same size Swagelok valves and fittings were used to assemble the Vapex apparatus. To reduce the number of fittings, the pressure gauge at the production end, the free gas regulating valve and the bypass line were all connected to a stainless steel manifold. A precise regulating valve was located in the free gas production line to adjust free gas flow rate during production.
Rate of free gas production was roughly adjusted to be 1 bubble per second. According to previous experience with similar equipment this flow rate is sufficient to ensure continuous two-phase flow in the production well. It is necessary to maintain two-phase flow in the production end of the model to prevent capillary hold-up of liquid in the model. Higher rate of free gas was required in experiments with high oil production rates to be able to stabilize the pressure in the separator.

Injection line, production line and valves located in the production line connecting the physical model outlet to the oil/gas separator were wrapped with heating tapes and maintained at the same temperature as the temperature in the model.

![Figure 3-10: Immersion Heater/ Thermo Controller](image)

### 3.2 Preparation Procedure

Prior to running each Vapex experiment, the sand-packed physical model was prepared through the following steps: Pressure testing, packing, vacuuming, water flooding and bitumen flooding. At the end of each experiment, physical model was opened, cleaned and reassembled.
3.2.1 Packing

As the first step, the physical model was assembled and left pressurized with nitrogen at 300 psig for a day to check the sealing O-rings in the top and bottom flanges, to detect possible leaks and to ensure that model can hold operating pressure of the experiments.

To prepare the sand-pack for current experimental study, 12-20 US mesh silica sand supplied by AGSCO, was packed into the model through three injection ports on the top lid. The physical model was vibrated during the packing process with a pneumatic vibrator. The vibrator employed during the packing process was a Martin NCT 29 rotary turbine vibrator which provides powerful vibratory force with air pressure range of 30-60 psig. Shaking and hammering the model helps in uniform settling of sand particles and ensures that no major empty space is left inside. The packing phase was stopped when the rate of sand intake with vibrating and tapping with a hammer became negligible. The sand intake of the model was measured to get a measure of the pore volume of the system by calculating the volume occupied by the sand grains through its density.

In the current experiments, it was tried to pack the model consistently with almost the same amount of sand to keep the permeability and porosity of the porous media approximately the same for the sake of comparisons. Figure 3-11 depicts the packing phase of sand-pack preparation.

It should be noted that permeability of the silica sand (12-20 US mesh) used for current experiments was determined by Tavallali (2012) in a different study and was reported to be around 250 Darcy. Tavallali used a sand-pack apparatus comprising a low rate pump, a differential pressure transducer, and a sand-pack holder to measure permeability of the same sand used in this study. By injecting water at specific rates and measuring the corresponding pressure difference along the sand pack, permeability was simply determined from Darcy’s law.
3.2.2 Evacuation of Sand-pack

In the next step, the meshed injection port on the top lid of physical model was connected to vacuum pump (Edward E-Lab2) to evacuate the pore volume. The vacuum pressure obtained with our vacuum pump was around 85 kPa after vacuuming for 24 hours, which was close the average atmospheric pressure of about 89 kPa in Calgary.

3.2.3 Water Flooding

After vacuuming, the model was fully saturated with degassed water. Water was injected from a water vessel into the model from a single injection point located at the bottom lid. The water transfer vessel was placed at a higher level than model to let the water imbibe into the vacuumed sand-pack under gravity head. Later the gas cap on top of water in transfer vessel was
pressurized with nitrogen to 15 psig to ensure the sand-pack is fully saturated with water. By measuring the weight of transfer vessel before and after water imbibition into the sand-pack, total pore volume was calculated. Figure 3-12 depicts the arrangement for water-flood phase of sand-pack preparation.

![Figure 3-12: Saturating the Vacuumed Sand-Pack with Water](image)

3.2.4 Oil Flooding

The model was flooded with bitumen to reach the irreducible water saturation. The bitumen used for saturating the sand-pack was Mackay River oil (13,440 mPa.s viscosity and 0.988 kg/m³ density at 50 °C and 1.03 MPa). Oil flooding process with such viscous oil takes relatively long time. To accelerate oil flooding, bitumen transfer vessel was wrapped with the heating tapes and heated to 50-60 °C. The transfer vessel was pressurized to 15 psig by nitrogen, which has very low solubility in oil at this low pressure, to push oil through the system. Physical model was preheated to 60 °C by means of heating tapes to speed up oil flow during the process.
Three injection ports on the top lid were used during the oil flood to displace water downward in a gravity stable manner. Near the end of the drainage process, the forth injection port directly above the production well was also used to uniformly displace water. Displaced water was allowed to flow in the production well and the volume of water displaced by the oil was measured to calculate connate water saturation. Oil injection was continued for more than one pore volume to assure proper and uniform oil saturation within the sand-pack.

As the model was under 15 psig pressure, after stopping the oil injection by closing all inlet valves, it was left for an hour with the outlet valve open to deplete the inside pressure. 24 hours were allowed for the saturated model to cool down and reach ambient temperature. A schematic of the process is shown in Figure 3-13 and Figure 3-14 displays the arrangement for oil-flood phase of sand-pack preparation. Characteristics of the prepared oil saturated sand packed media, prepared for eight conducted experiments are listed in Table 3-3.

![Figure 3-13: Schematic of Oil Saturating Process](image-url)
Figure 3-14: Oil-Flood Arrangement for Sand-Pack Preparation

Table 3-3: Characteristics of Prepared Oil Saturated Sand-Packs

<table>
<thead>
<tr>
<th>Test #</th>
<th>Pore Volume (cm$^3$)</th>
<th>Porosity</th>
<th>Oil Saturation</th>
<th>Connate Water Saturation</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1634.0</td>
<td>0.33</td>
<td>0.87</td>
<td>0.13</td>
</tr>
<tr>
<td>2</td>
<td>1686.5</td>
<td>0.33</td>
<td>0.89</td>
<td>0.11</td>
</tr>
<tr>
<td>3</td>
<td>1703.0</td>
<td>0.35</td>
<td>0.83</td>
<td>0.17</td>
</tr>
<tr>
<td>4</td>
<td>1629.3</td>
<td>0.33</td>
<td>0.85</td>
<td>0.15</td>
</tr>
<tr>
<td>5</td>
<td>1608.0</td>
<td>0.33</td>
<td>0.85</td>
<td>0.15</td>
</tr>
<tr>
<td>6</td>
<td>1594.6</td>
<td>0.32</td>
<td>0.84</td>
<td>0.16</td>
</tr>
<tr>
<td>7</td>
<td>1670.0</td>
<td>0.34</td>
<td>0.86</td>
<td>0.14</td>
</tr>
<tr>
<td>8</td>
<td>1120.0</td>
<td>0.34</td>
<td>0.88</td>
<td>0.12</td>
</tr>
</tbody>
</table>
3.3 Vapex Experiments

The Vapex apparatus was assembled by connecting the earlier mentioned components through stainless steel tubing network based on the flow diagram in Figure 3-1. The entire set-up was pressurized by nitrogen to 300 psig and leak tested. The oil-saturated sand-packed physical model was then installed in the experimental set-up.

Six different Vapex experiments were conducted in the cylindrical physical model under different operating temperature and pressure to evaluate how injecting warm propane into the pre-heated oil-saturated sand pack will affect the Vapex performance. The operating conditions of these experiments are tabulated in the Table 3-4. It should be noted that all the reported pressures in the following sections are absolute values.

<table>
<thead>
<tr>
<th>Experiment #</th>
<th>Injection Pressure (MPa)</th>
<th>Test Temperature (°C)</th>
<th>Injected Gas</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.817</td>
<td>22</td>
<td>Propane</td>
</tr>
<tr>
<td>2</td>
<td>0.817</td>
<td>40</td>
<td>Propane</td>
</tr>
<tr>
<td>3</td>
<td>0.817</td>
<td>50</td>
<td>Propane</td>
</tr>
<tr>
<td>4</td>
<td>0.817</td>
<td>60</td>
<td>Propane</td>
</tr>
<tr>
<td>5</td>
<td>1.23</td>
<td>40</td>
<td>Propane</td>
</tr>
<tr>
<td>6</td>
<td>1.55</td>
<td>50</td>
<td>Propane</td>
</tr>
<tr>
<td>7</td>
<td>1.55</td>
<td>50</td>
<td>Nitrogen</td>
</tr>
<tr>
<td>8</td>
<td>0.817</td>
<td>50</td>
<td>Propane</td>
</tr>
</tbody>
</table>

To warm up the sand-pack medium, the inner cylinder of the physical model was filled with water and water was heated with the immersion heater to reach the temperature set point of the experiments. The outer cylinder was insulated to avoid heat dissipation to the ambient. Experiment #1 was carried out at room temperature while propane was injected at 0.817 MPa. In experiments #2, #3 and #4, sand-pack was warmed up to 40°C, 50°C or 60°C respectively while injection pressure of propane was regulated to be 0.817 MPa. The separator and injection lines were warmed up with heating tapes. Covering the production lines and separator with insulation
minimized heat losses to the ambient and helps to keep the produced diluted oil at almost the same pressure and temperature of the vapour chamber. Through the thermocouple probes, inserted into the sand-pack and pressurized separator, temperature was monitored and adjusted.

In experiment #5 and #6, injection pressure of propane was increased to 90% of its vapor pressure at the test temperature (1.23-1.55 MPa, 165-212 psig) to investigate how solvent injection pressure affects oil production rate by altering solvent solubility into the oil.

Normally, in the Vapex process, the light solvent dissolves in the heavy oil/bitumen and diluted oil drains into the production well under gravitational force. There is little or no pressure gradient inside the vapor chamber and therefore the gravity drainage is the only flow mechanism. In current lab experiments, a pressure difference was initially applied between the injection and production wells to speed up the establishment of communication between the wells. For this purpose the pressure at the production end was kept atmospheric while propane was injected at 0.817 to 1.55 MPa depending on the operating pressure of experiment. The valve between the separator and free gas collection cylinder was closed to let the pressure in the separator build up. As soon as the pressure in the separator became almost equal to the injection pressure, free gas was allowed to flow from the pressurized separator toward the free gas collection cylinder. Free gas production rate was regulated by means of the regulating valve located after the separator in the free gas production line. Free gas production rate should be high enough to maintain continuous two phase flow in the production well, which ensures no liquid build-up develops in the model. However, too high a rate would be undesirable, since it would simply circulate the solvent through the model and would reduce the permeability to oil near the production well. The pressure at the production end was monitored with a digital pressure gauge to make sure no pressure gradient is externally applied to the vapor chamber. During oil production, free gas production was allowed and the bypass and solution gas lines were kept closed. The level of accumulated live oil in the graduated separator, level of free gas in the free gas collecting cylinder, injection and production pressures and weight loss of the propane tank were recorded at regular time intervals, extending from the time of starting to almost 3 days.

When level of accumulated live oil in the separator was sufficient to collect a sample, separator was discharged into a tared oil collection jar. For this purpose, the production was temporarily
interrupted by closing the outlet valve of the model, the free gas line was shut and solution gas line toward the collecting cylinder was opened. The valve between the separator and the oil collector was opened to pass the live oil from the pressurized separator to the glass jar at ambient pressure.

The dissolved gas in the live oil was liberated due to lower pressure after the separator and substantial volume of gas was collected in the solution gas cylinder during sampling. The oil production was resumed after withdrawing the sample. The accumulated oil in the jar was allowed to degas for some time before taking it off from the sample collection system. The weight of the collected oil sample was measured immediately after production and again after a week at atmospheric condition during which time the remaining solution gas is assumed to have escaped.

The solution gas cylinder was refilled with water 2-4 times during oil sample collecting by shutting the solution line toward the cylinder and opening the valve in the vacuum line. Free gas cylinder was refilled frequently during the experiment in the same way.

Each Vapex experiment with propane injection was run continuously for around 60 hours. Finally, the system pressure was depleted and the produced gas during blow-down was collected and measured. The blow-down phase took 4-5 hours.

In Experiment #7, instead of propane, nitrogen was injected into the physical model, which was warmed up to 50°C, to exclude the role of solvent dissolution on viscosity reduction and determine how much of the original mobile oil in the system can be drained by gravity without solvent dissolution. To mimic the earlier experiments, nitrogen was injected at 1.55 MPa while the production point pressure was atmospheric before the gas breakthrough.

In experiment #8, the cylindrical physical model was replaced with a rectangular visual model. The sand-pack for this experiment was prepared through the same procedure mentioned earlier. The rectangular physical model was warmed up by wrapping heating tapes and insulating all the faces of the model except the windowed side. Experiment #8 was operated at the same injection pressure and test temperature as of Experiment #3 to monitor chamber development during the process.
In this chapter, results of the heated Vapex experiments performed at different elevated temperatures and different propane injection pressures are presented, analyzed and discussed.

The objective of heated Vapex experiments was to figure out how mild increase in formation temperature will affect the oil production rate during the process. For this purpose, first four experiments were performed at propane injection pressure of 817 kPa while the sand-packed physical model was pre-heated from 22 °C to 40, 50 and 60 °C in experiments #1 to #4.

In experiments #5 and #6, propane injection pressure was increased to 90% of the corresponding vapor pressure at the test temperature to investigate how solvent injection pressure affects oil production rate by altering solvent solubility into the oil and diluted oil viscosity.

Effects of operating temperature and pressure on Vapex performance seem not to be obvious due to the complications involved in the process. As it can be realized from the original Vapex theory, the involved mechanisms in the process including mass transfer and fluid flow under gravity effect can be strongly influenced by altering the operating conditions. At elevated temperatures, oil viscosity declines while at the same time solubility of propane into the bitumen decreases. Less dissolution of propane into the oil at lower injection pressures but higher formation temperatures provides less driving force for mass transfer between solvent and oil. Moreover viscosity of diluted oil within the draining layer depends largely on the solvent concentration and the viscosity of original oil is influenced strongly by the temperature. Moreover dependency of diffusivity on temperature and solvent concentration also adds up to this complexity. This means that the complex relationship between the injection pressures, formation temperature and thermo physical properties of propane/bitumen system should be carefully understood to be able to analyze the mass transfer and fluid flow mechanisms in process. Results of the conducted experiments will be analyzed and discussed in this chapter to infer the overall effect of heat incorporation in Vapex process.

Results of experiment # 7 in which propane was replaced with nitrogen to eliminate the effect of solvent dissolution on oil mobilization and drainage will be also analyzed to quantify the contribution of free fall gravity drainage into total oil production. Results of this experiment can
be used to figure out how much oil can be produced just under the effect of heat in heat-solvent hybrid recovery methods.

Results of experiment #8, conducted in the visual rectangular model, will be also studied to examine development of the vapor chamber during Vapex experiment at elevated temperature.

The obtained experimental results can be helpful in quantifying the upper limit of oil rates achievable with heated solvent injection and can also be used to assess the applicability of Vapex to naturally warm reservoirs.

4.1 Thermo-physical Properties of Propane-Saturated Mackay River Bitumen

In order to understand the Vapex performance under different operating conditions, the effects of pressure and temperature on thermo-physical properties of the bitumen-propane system need to be understood. Being highly dependent on pressure and temperature, propane solubility in oil and potential viscosity reduction as a result of solvent dissolution and temperature increase should be considered together to evaluate and improve the process. Reduction of oil viscosity at higher temperatures may happen at the cost of lower solubility of propane into the oil and will not necessarily result in a dramatic improvement in oil recovery. Although viscosity of propane saturated bitumen under operating pressure and temperature of the experiments is just indicative of the diluted oil viscosity at the interface of the vapor chamber, mapping of the viscosity-pressure-temperature relationship can be helpful to comparatively evaluate the Vapex performance.

In this section, viscosity-propane solubility-temperature map of Athabasca Bitumen (MacKay River oil) is provided within the range of room temperature and critical temperature of propane (23-96.6 ºC) and within pressures range of 100 to 3500 kPa.

4.1.1 Viscosity of Dead Oil Mackay-River Oil

In the current study, the viscosity measurements of Mackay-River Bitumen, reported by Nourozieh et al (2011), were fitted to a double log correlation to obtain dead bitumen viscosity.
Viscosity data in Nourozieh et al study, within the pressure range of (1.034-13.790 MPa) and temperature range of (45-180 °C), were fitted to the following correlation:

\[
\ln\left[\ln\left(\mu_{\text{bitumen}}\right)\right] = 21.5 - 3.3352 \ln(T) + 0.00532 P_g \tag{EQ 4-1}
\]

where viscosity is in mPa.s, Temperature is in °K and Pressure is in MPa gauge.

Figure 4-1 depicts the correlated viscosity of Mackay River Bitumen versus temperature.

### 4.1.2 Solubility of Propane in Bitumen

Solubility of propane into the oil under different operating conditions (pressure and temperature) can be estimated through different equation of state models or activity coefficient models, verified by proper experimental measurements. The modified Raoult’s law which is based on equality of fugacities between the liquid and vapor phases is used in this section to correlate propane mole fraction to saturation pressure and propane vapor pressure at corresponding test temperature based on proper experimental data as follows:

\[
x_{\text{propane}} \cdot \gamma_{\text{propane}} \cdot P_{v_{\text{propane}}} = P \tag{EQ 4-2}
\]

where \(x_{\text{propane}}\) is propane mole fraction, \(P_{v_{\text{propane}}}\) is propane vapor pressure in kPa and \(\gamma_{\text{propane}}\) is the activity coefficient of propane in bitumen. Badamchi Zadeh (2009) measured saturation pressure of (5.2 to 26.1 weight %) propane in Athabasca bitumen mixtures over the temperature range of 10 to 50 °C and obtained the activity coefficient of 1.158 to fit the experimental saturation pressure data. Badamchi Zadeh also employed the Van LAr form of activity coefficient along the measured data and concluded that in low concentrating mixtures of propane and bitumen, compositional dependency of activity coefficient can be reasonably ignored. Figure 4-2 depicts the solubility of propane in Athabasca bitumen in Badamchi Zadeh study.

Since temperature and pressure ranges of current study fit well within the reported data by Badamchi Zadeh, the same semi-empirical model is used in this section to correlate propane solubility in Athabasca bitumen to pressure at different temperatures.
4.1.3 Viscosity of Propane-Bitumen Mixture

Viscosity of diluted bitumen can be generated through different empirical or semi empirical mixing rules including Shu, Bingham, Lobe and Pedersen (Yazdani et al, 2009). Yazdani and Maini (2007) tested different types of viscosity correlations to match measured viscosity of butane-heavy oil mixtures and suggested that generalized correlations have limited success in predicting the viscosity of heavy oil/light solvent mixtures and semi empirical correlation tuned against experimental viscosity data are required for accurate predictions.

In current study, modified Shu correlation (EQ 4-2) proposed by Badamchi Zadeh (2012) was used to find viscosity of propane-Mackay River bitumen mixtures. Shu correlation is a generalized correlation developed based on power-law mixing rule to calculate viscosity of heavy crudes diluted with light petroleum solvents. The correlation requires viscosity and density of the two blended components to predict the mixture viscosity at any composition.

Badamchi Zadeh adjusted the constant coefficients in the original Shu correlation to fit the measured viscosity data of 5.2 wt% to 26.1 wt % propane-bitumen mixtures within 10 to 90 °C temperature. Since pressure and temperature ranges, over which the correlation is valid, covers the pressure and temperature ranges of current study, same correlation is employed here to find viscosity of diluted bitumen.

\[
\mu = \mu_o \frac{f_o}{f_s} \mu_s \quad \text{EQ 4-3}
\]

\[
f_o = \frac{\lambda C_o}{\lambda C_o + C_s} \quad \text{EQ 4-4}
\]

\[
f_s + f_o = 1 \quad C_s + C_o = 1
\]

\[
\lambda = 0.000375 \Delta \rho^{8.56653} \rho_o^{4.567001} \rho_s^{-5.27094} \ln \left[ \frac{\mu_o}{\mu_s} \right] \quad \text{EQ 4-5}
\]

\[
\Delta \rho = \rho_o - \rho_s \quad \text{EQ 4-6}
\]
where \( \rho_o \) and \( \rho_s \) are specific gravities of oil and solvent respectively. \( \lambda \) is an empirical parameter defined as EQ 4-5. \( C_o \) and \( C_s \) are oil and solvent volume fractions, \( \mu_o \) and \( \mu_s \) are dead oil and solvent viscosities (mPa.s).

In the current study, \( \mu_o \) is calculated from EQ 4-1. Density and viscosity of liquid propane below its critical temperature is determined from the following correlations available in Handbook of Thermodynamic and Physical Properties of Chemical Compounds (Yaws, 2003):

\[
\log (\rho_{\text{propane}}) = \log (221.51) - \log (0.27744) \left( 1 - \frac{T}{369.82} \right)^{0.287} \quad \text{EQ 4-7}
\]

where \( \rho_{\text{propane}} \) is propane density in kg/m\(^3\) and \( T \) is temperature in \(^\circ\) K.

\[
\log (\mu_{\text{propane}}) = -3.1759 + \frac{297.12}{T} + 0.0095452 T - 0.000018781 T^2 \quad \text{EQ 4-8}
\]

where \( \mu_{\text{propane}} \) is propane viscosity in mPa.s and \( T \) is temperature in \(^\circ\) K.

Density of bitumen for viscosity calculations is obtained from EQ 4-9. This correlation was obtained by fitting a linear equation into the measured density values of Mackay River bitumen at different temperatures from Nourozieh et al (2011) which is depicted in Figure 4-3.

\[
\rho_o = -0.5983 T + 1180.4 \quad \text{EQ 4-9}
\]

where \( \rho_o \) is bitumen density in kg/m\(^3\) and \( T \) is temperature in \(^\circ\) K.

According to the experimental density measurements by Nourozieh et al (2011), bitumen density showed insignificant dependency on pressure within the pressure range of current study. Hence, the dependency of bitumen density on pressure is not considered in above correlation.

### 4.1.4 Viscosity-Solubility-Temperature Contour Maps of Mackay River Bitumen-Propane

To capture the overall effect of pressure and temperature on viscosity of diluted oil, propane solubility curves and diluted oil viscosity profiles within the pressure and temperature scopes of current research were combined and these are presented in Figure 4-4.
In Figure 4-4, at every injection pressure and temperature propane solubility in bitumen can be obtained from the constant solubility curves. At every injection pressure and temperature, viscosity of diluted oil can be figured from the Viscosity-Temperature-Pressure contours. Dependency of propane-diluted bitumen viscosity on propane concentration is implicitly considered through solubility values at different injection pressures. Based on phase behavior studies of propane-bitumen system, injecting liquid propane can result in formation of a second hydrocarbon liquid phase at the boundary of vapor chamber and also increases the chances of asphaltene precipitation. Therefore, these contours were generated for the pressure ranges lower than propane vapor pressures at each temperature to avoid vapour-liquid-liquid phase regimes.

Propane injection pressure and test temperatures corresponding to the conducted Vapex experiments can be examined against the contours of Figure 4-4 to determine how altering the operating conditions influences the viscosity of diluted oil.

Figure 4-1: Viscosity of Mackay River Oil versus Temperature
Figure 4-2: Solubility of Propane in Athabasca Bitumen (solid symbols: step-wise method, open symbols: continuous method, measured by Badamchi Zadeh, 2009)

Figure 4-3: Density of Mackay River Bitumen versus Temperature (Measurements by Nourozieh et al, 2011)
In experiments #1 to #4, the test temperature was increased from T= 22 °C to 40, 50 and 60 °C while propane was injected at 817 kPa in all four cases. Drawing a constant pressure line of 817 kPa and moving from T= 22 toward T=60, it can be realized that lower values of solubility are obtained along the line. Although at higher temperature lower viscosities are expected, negative effect of lower solubility surpasses the positive effect of temperature on viscosity reduction and finally slightly higher viscosities are achieved at higher temperatures.

In experiment #5, propane injection pressure was increased from 817 kPa to 1230 kPa while physical model was at 40 °C. As Figure 4-4 demonstrates, increasing injection pressure at constant temperature would result in more solvent dissolution and consequently lower diluted oil viscosities. The same trend can be observed for experiment #6 in which injection pressure of propane was increased from 817 kPa to 1550 kPa at test temperature of 50 °C.

Gates (2005) developed similar contour maps for viscosity of Athabasca Bitumen-hexane system to find optimum steam and solvent injection strategy in ES-SAGD process. According to that study, viscosity of diluted bitumen at the boundary of steam chamber at low pressures but high temperatures depends largely on temperature due to low solubility of hexane into the oil. However at higher pressure but low temperatures, effect of pressure was found to be more dominant on viscosity reduction. Gates suggested that the best operating strategy in ES-SAGD can be picked by finding the trajectory in the pressure-temperature space where the oil phase viscosity is the lowest. Thus the trajectory of the oil phase viscosity is the key, not the effects of solubility alone nor the effect of temperature alone.
Figure 4-4: Viscosity-Solubility-Temperature Contour Map of Athabasca Bitumen/Propane System
It should be pointed out that the viscosity information presented in Figure 4-4 is at equilibrium conditions, i.e. when the oil has absorbed as much solvent as the solubility limit will allow. In the Vapex process, the oil starts draining as soon as the viscosity becomes low enough to make it mobile, i.e. before it becomes fully saturated with the solvent vapor. Therefore, the equilibrium viscosity information from Figure 4-4 is not an accurate indicator of the actual drainage rate. The impact of oil viscosity on drainage rate in Vapex is rather complex due to the relationship between solvent diffusion coefficient and diluted oil viscosity. As the oil becomes diluted with the solvent, not only its viscosity decreases but the solvent diffusivity in the oil also increases. Thus both viscosity and diffusivity are concentration dependent. More accurate studies on concentration dependency of diffusivity and its impact on oil dilution and mobilization are required to fully assess the warm Vapex process. The current experimental study was performed to evaluate warm Vapex performance at different temperatures and examine the effect of elevating temperature on the underlying mass transfer and fluid flow mechanisms.

4.2 Production Results of Experiment #1 to #4

The injection/production profiles of first four conducted experiments are reported in this section. The performance of each experiment can be evaluated and compared with other experiments at elevated temperatures according to the obtained experimental injection and production data.

It should be noted that conducted Vapex experiments were started with atmospheric pressure in the separator to accelerate establishment of communication between wells. The communication between injection and production was established in less than half an hour in each case and very limited amount of oil was produced during displacement phase. After gas breakthrough occurs, pressure in the separator starts building up close to propane injection pressure and, from then on, there will be minimal pressure difference between injection and production points and gravity will be the only significant drive mechanism.

Since injection point was located right on top of the employed physical model, the typical raising chamber phase in Vapex oil production profiles is not observed in the current experimental results. As soon as gas breaks through in the production well and vapour chamber is formed between injection and production points, diluted oil starts to drain within diffusion layer at the
boundaries of vapour chamber. Obtained oil production curves during this period are the basis of current evaluations.

In this study, the first experiment was designed to be similar to classic Vapex process. The physical model was kept at room temperature and propane was injected at 817 kPa. As it can be seen in Table 3-3, the prepared sand-pack in this experiment had total pore volume of 1634 cm\(^3\) (33% porosity), was 87% saturated with bitumen and no mobile water was present within porous medium. Cumulative oil production and gas injection profiles of experiment #1 are presented in Figure 4-5. Gas production profiles including produced solution and free gas are presented in Figure 4-6. Total of 445 cm\(^3\) of dead oil was produced over 53 hours in this experiment.

In experiments #2, the physical model was warmed up to 40 ºC while propane was injected at the same injection pressure as of previous experiment (817 kPa). The pore volume, porosity and oil saturation of the prepared sand-pack for this experiment was 1686.5 cm\(^3\), 33% and 89% respectively. Experiment #2 was carried out for 60 hours and 678 cm\(^3\) of dead oil was produced in total. Oil production/gas injection profiles of this experiment are depicted in Figure 4-7 and Figure 4-8 depicts the gas production profile.

In experiments #3, the test temperature was increased to 50 ºC and propane was injected at 817 kPa similar to earlier two experiments. The pore volume, porosity and oil saturation of the prepared sand-pack for this experiment was 1703 cm\(^3\), 35% and 83% respectively. Experiment #3 was also carried out for 60 hours and 850 cm\(^3\) of dead oil was produced during the life of experiment. Oil production/gas injection profiles of this experiment are depicted in Figure 4-9 and Figure 4-10.

In current manuscript the volumetric gas injection and production rates are reported at standard condition SC (T= 273.15 ºK and P= 100 kPa).
Figure 4-5: Gas Injection and Oil Production Profiles of Experiment #1

Figure 4-6: Gas Production Profiles of Experiment #1
Figure 4-7: Gas Injection and Oil Production Profiles of Experiment #2

Figure 4-8: Gas Production Profiles of Experiment #2
Figure 4-9: Gas Injection and Oil Production Profiles of Experiment #3

Figure 4-10: Gas Production Profiles of Experiment #3
Figure 4-11: Gas Injection and Oil Production Profiles of Experiment #4

Figure 4-12: Gas Production Profiles of Experiment #4
In experiment #4, the physical model was warmed up to 60 °C and propane injection pressure was regulated at 817 kPa. The prepared sand pack for this experiment had 1629.3 cm³ pore volume, 33% porosity and was 85% oil saturated. Experiment #4 was run for 69 hours and a total of 1007.34 cm³ deal oil was produced during the course of experiment. Oil production/gas injection profiles of this experiment are depicted in Figure 4-11 and Figure 4-12.

4.3 Effect of Temperature Elevation on Vapex Performance

Cumulative dead oil production profiles of Vapex experiments at 22, 40, 50 and 60 °C can be compared in Figure 4-13. Recovery factor curves of these four experiments are depicted in Figure 4-14. As Figure 4-13 demonstrates, increasing the test temperature from 22 to 60 °C improved the rate of oil production at the early stage of experiments. However, after 20 hours of production, oil drainage rate appeared not to be significantly different in all four conducted tests.

![Figure 4-13: Comparison of Cumulative Dead Oil Production at Elevated Temperatures](image-url)
Getting back to Figure 4-4 and comparing the viscosities of propane saturated bitumen at T=22, 40, 50 and 60°C, it can be realized that viscosity of diluted oil at the solvent interface slightly increased at higher temperatures.

This increase in viscosity of diluted oil in spite of lower viscosity of dead oil is caused by less dissolution of propane in the oil and consequently less oil dilution at the interface in elevated temperatures. Figure 4-15 depicts the experimental solvent volume fractions in experiment #1 to #4 which can be indicative of the extent of solvent dissolution in oil in these experiments. Based on this graph, it can be realized that less propane was dissolved in bitumen when temperature was increased from 22 to 60°C while propane injection pressure was kept the same at 817 kPa. Similar viscosity of diluted oil at the solvent interface under operating condition of conducted warm Vapex tests can be a reason for similar rate of oil production at the later stages of production. However the reasons for observed initial improvement in Vapex performance seem not to be very obvious and it needs further investigation which will be addressed later.

Figure 4-14: Comparison of Recovery Factors at Elevated Temperatures
Based on Figure 4-13, it can be realized that oil production occurred at an almost stable rate in Vapex experiment at room temperature (22°C). However rate of oil production was gradually decreasing over the time in Vapex experiments at 40, 50 and 60 °C. Stable oil drainage rate during experiment #1 is consistent with oil drainage behavior expected from Vapex analytical models during pseudo steady state phase of chamber growth and before chamber reaches outer boundaries of the physical model.

Monotonic decrease in oil drainage rate in next three experiments at higher temperatures suggests that Vapex performs in a different manner at elevated temperatures.

In order to understand this difference, viscosity profile of Mackay oil versus temperature in (Figure 4-1) should be considered. According to the bitumen viscosity trend, it can be realized that oil viscosity decreases almost 10 times by 18 °C increase in temperature from 22 to 40 °C. This huge viscosity reduction may be sufficient to mobilize oil under the operating conditions of current experiments. If the oil within sand matrix is originally mobile, diluted oil drainage might not be the only flow mechanism contributing to oil production in conducted warm Vapex
experiments. Mobility of the original oil in place implies that oil might be produced not only by dilution at the boundary of vapor chamber but also from the un-contacted zones where gravity driving force was sufficient to overcome the viscous force resisting the oil drainage (Figure 4-16). The additional oil recovered this way causes the recovery profile to be different from that of classic Vapex with immobile original oil in place.

**Figure 4-16: Cross Section of Vapex Chamber in Case of Free Fall Gravity Drainage**

Drainage of oil in place under the effect of gravity is called “free fall gravity drainage” in the literature. This mechanism has been studied by many researchers and is discussed in more detail in Chapter 2. Cardwell et al (1948), Hagoort (1980), Li et al (2003) found capillary effect to be an important parameter controlling the oil production and oil saturation profiles in free fall gravity drainage. In the current set of experiments, capillary effect can be neglected due to relatively high permeability of the medium. Capillary hold up around the production well also seems not to be significant due to sufficient rate of free gas production.

At this point as a rough estimation, the maximum possible amount of oil that can be produced by free fall gravity drainage with open area to flow at the bottom of current physical model and in this set of experiments can be calculated by using Darcy equation as follows:

\[
\frac{Q}{A} = \frac{\Delta P \cdot k}{L \cdot \mu} \quad \text{EQ 4-10}
\]

Considering the gravity to be the only driving force to move the oil and neglecting capillarity, Darcy equation can be simplified to the following equation:
\[
Q = \frac{\rho \cdot g \cdot h \cdot k}{h \cdot \mu} = \frac{\rho \cdot g \cdot k}{\mu} \quad \text{EQ 4-11}
\]
\[
A = \frac{\pi (D_1^2 - D_2^2)}{4} = 160.52 \text{ cm}^2 \quad \text{EQ 4-12}
\]

Substituting the oil properties \((\rho, \mu)\) and cylindrical model’s dimension (Annular cross section area and height) to this equation, following table was generated:

<table>
<thead>
<tr>
<th>Temperature °C</th>
<th>Viscosity Pa.s</th>
<th>Density kg/m³</th>
<th>Open to Flow Area (cm²)</th>
<th>Permeability Darcy</th>
<th>Flow rate cm³/hr</th>
</tr>
</thead>
<tbody>
<tr>
<td>22</td>
<td>301.76</td>
<td>998</td>
<td>160.52</td>
<td>250</td>
<td>0.468</td>
</tr>
<tr>
<td>40</td>
<td>32</td>
<td>994</td>
<td>160.52</td>
<td>250</td>
<td>4.39</td>
</tr>
<tr>
<td>50</td>
<td>13.45</td>
<td>988</td>
<td>160.52</td>
<td>250</td>
<td>10.40</td>
</tr>
<tr>
<td>60</td>
<td>5.12</td>
<td>983</td>
<td>160.52</td>
<td>250</td>
<td>27.18</td>
</tr>
</tbody>
</table>

Above calculations indicate that contribution of free-fall gravity drainage (beyond vapour chamber) into oil recovery can be significant in these tests and needs to be excluded from the total rate of production to make it possible to study rate of oil production promoted by solvent dissolution individually. Free-fall gravity drainage rate in current warm Vapex experiments is lower than the above-calculated rate because of different geometry of flow. Moreover, it decreases over the time because of the decrease in the gravity force to viscous force ratio involved during oil flow toward the single production point. To be able to quantify oil production through free fall gravity drainage more accurately, Experiment #7 was designed and conducted which will be discussed later in this chapter.

It is obvious that performance of Vapex process in reservoirs with originally mobile oil in place is slightly different from that in reservoirs with immobile oil in place. This could explain the monotonic decrease in rate of oil production in conducted warm Vapex experiments. In these cases, Vapex in fact works as a parallel mechanism and not the only mechanism controlling the production. To be able to compare warm Vapex performance at different temperatures, free fall gravity drainage and diluted oil drainage should be evaluated alongside each other.
With this explanation, the higher initial rate of oil production in the warm Vapex tests at elevated temperature (40, 50 and 60 °C) can be partly due to higher rate of oil production through free fall gravity drainage. Contribution of free fall gravity drainage to the total oil production depends strongly on viscosity of dead oil in place which decreases significantly by elevating temperature. Lower viscosity of oil in Experiment #4 resulted in higher rate of oil production by free fall gravity drainage and finally higher initial rate of oil production in spite of less dissolution of propane in oil at elevated temperatures. The same explanation can be extended to Experiment #3 compared to #2 and #2 compared to #1.

To be able to quantify contribution of free fall gravity drainage at different temperatures, Experiment #7 was performed in which propane vapour was replaced with nitrogen to eliminate the effect of solvent dissolution on oil mobilization and just focus on dead oil gravity drainage. From this experiment, the volume of produced mobile oil by free-fall gravity drainage can be assessed. Results of experiment #7 will be discussed in section 4.6.

4.4 Production Results of Experiment #5 and #6

In experiments #5 and #6, injection pressure was increased to 90% of propane vapor pressure at the test temperature to investigate how solvent injection pressure affects oil production rate by altering solvent solubility into the oil.

In experiment #5, physical model was warmed up to 40 °C and propane was injected at 1230 kPa. The prepared sand pack for this experiment had 1608 cm³ pore volume, 33% porosity and was 85% oil saturated. Experiment #5 was run for 74 hours and total of 960 cm³ dead oil was produced during the experiment. Oil production/gas injection profiles of this experiment are depicted in Figure 4-17 and Figure 4-18.

In Experiment# 6, the temperature of physical model was increased to 50 °C and propane injection pressure was regulated at 1550 kPa. The characteristics of the prepared sand pack were: 1594.6 cm³ pore volume, 32% porosity and 84% oil saturation. Experiment #6 was continued for 62 hours and total of 1199 cm³ dead oil was produced during the course of experiment. Oil production/gas injection profiles of this experiment are depicted in Figure 4-19 and Figure 4-20.
Figure 4-17: Gas Injection and Oil Production Profiles of Experiment #5

Figure 4-18: Gas Production Profiles of Experiment #5
Figure 4-19: Gas Injection and Oil Production Profiles of Experiment #6

Figure 4-20: Gas Production Profiles of Experiment #6
4.5 Effect of Solvent Injection Pressure on Vapex Performance

Comparison of cumulative dead oil production at T= 40 °C with propane injection pressure of 817 and 1230 kPa is presented in Figure 4-21. Higher rate of oil production at 1230 kPa is due to higher solubility of propane in the oil and higher rate of solvent mass transfer due to larger concentration gradient and available driving force. Higher injection pressure of propane results in more pronounced viscosity reduction and consequently faster drainage rate. This explanation can be also confirmed through solvent volume fraction profile of the conducted experiments in Figure 4-22. The average volume fraction of solvent in produced oil increases from about 22% at 817 kPa to approximately 36% at 1230 kPa. Similar enhancement in drainage rate with increased pressure was observed at 50 °C. The rates of oil production at T= 50 °C with propane injection pressure of 817 and 1550 kPa are illustrated in Figure 4-23. Higher rate of oil production at 1550 kPa can be explained by more solvent dissolution at higher pressure as shown in Figure 4-24.

According to Figure 4-21 and Figure 4-23, injecting propane at higher pressures is able to keep the rate of oil production high for a longer time. In spite of gradual gravity head loss during the course of experiments (after 20 hours), oil was produced faster for a longer time at higher pressure compared to lower injection pressure. In fact at elevated temperatures but low injection pressures, rate of oil production starts declining faster due to insufficient amount of solvent to mobilize the oil and greater proportional contribution of free fall gravity drainage.
Figure 4-21: Comparison of Oil Production Profiles at $T=40 \, ^{\circ}C$ at Different Pressures

Figure 4-22: Comparison of Solvent Volume Fraction at $T=40 \, ^{\circ}C$ at Different Pressures
Figure 4-23: Comparison of Oil Production Profiles at T=50 °C at Different Pressures

Figure 4-24: Comparison of Solvent Volume Fraction at T=50 °C at Different Pressures
4.6 Production Results of Experiment #7: Nitrogen Injection

Experiment #7 was performed in the same physical model with similar sand pack properties as of the earlier experiments (Pore volume: 1670 cm$^3$, Porosity: 0.34 and Oil saturation: 0.86). The operating condition of this experiment was chosen to be the same as experiment #6. The physical model temperature was increased to be 50 °C and injection pressure was regulated at 1550 kPa. However in this experiment, nitrogen as a non-condensable gas which has negligible solubility in bitumen under operating condition of experiment was injected instead of propane. This experiment was aimed at measuring the volume of produced dead oil via free fall gravity drainage mechanism under the well configuration of Vapex process. A minimum rate of free gas production was maintained in this experiment to mimic the operating conditions of earlier warm Vapex experiments. Figure 4-25 depict the oil and gas production profiles of Experiment #7. According to the obtained results, 390 cm$^3$ of dead oil (27% RF) was produced during 72 hours of this experiment.

![Figure 4-25: Oil/Gas Production Profiles of Experiment#7](image_url)
Result of experiment #7 at T=50 °C with nitrogen was numerically simulated with Computer Modelling Group’s STARS simulator, using the same sand pack properties and physical model dimension as in the actual Experiments #7. Figure 4-26 shows numerically simulated rate of free fall gravity drainage, against the results of Experiment #7. According to the simulation study, oil production was found not to be sensitive to nitrogen injection pressure and rate of gas production. This observation is consistent with the gravity drainage theory in which the draining fluid and reservoir characteristics dictates rate of oil drainage in absence of capillarity.

![Graph showing experimental and simulated free fall gravity drainage rate at T=50 °C](image)

**Figure 4-26: Comparison of Experimental and Simulated Free Fall Gravity Drainage Rate at T=50 °C**

After history matching the simulation results, numerical results were extended to the temperatures of other Vapex experiments. Simulation results of free fall gravity drainage at T=40, 50 and 60 °C are compared in Figure 4-27. Being highly dependent on oil viscosity, free fall gravity drainage rate was higher at higher temperatures. The free fall gravity drainage rate
declines over the time due to increasing proportional impact of flow resistance in the horizontal direction for the oil to reach the production well.

Figure 4-27: Comparison of Simulated Rate of Free Fall Gravity Drainage Rate at Different Temperatures

These simulation results were used to exclude free fall gravity drainage rate from total recovery rates of conducted experiments. In Figure 4-28, magnitude of produced oil by free fall gravity drainage mechanism is quantified against total produced oil in warm Vapex experiments at T=40 °C, P= 817 kPa and T=40 °C, P=1230 kPa. In Experiments #2 and #5, free fall gravity drainage had 25% and 20% contribution into total recovery. Since both experiments were conducted at the same temperature of 40 °C, same magnitude of dead oil production by free fall gravity drainage is expected in these two cases. Thus the enhancement in oil production rate at higher pressure can be attributed to higher rate of oil production via solvent dissolution, oil dilution and oil drainage within diffusion layer at the boundaries of the propane chamber.
Figure 4-28: Contribution of Free Fall Gravity Drainage to Total Production in Experiment #2 and #5

Figure 4-29: Contribution of Free Fall Gravity Drainage to Total Production in Experiment #3 and #6
Figure 4-29 demonstrates cumulative dead oil production profiles of warm Vapex experiments at T=50 °C, P=817 kPa and P=1550 kPa which can be compared with dead oil production profile via free fall gravity drainage at this temperature. 30% of total produced oil in experiment #3 and 44% of total produced oil in experiment #6 were recovered by free fall gravity drainage mechanism.

Same evaluation was done for experiment #4 at 60 °C and P=817 kPa and results are presented in Figure 4-30. Contribution of free fall gravity drainage to total recovery increased to 55% in this experiment. Free fall gravity drainage rate is accelerated by increasing the formation temperature as a result of strong effect of temperature elevation on viscosity reduction of dead oil in place. By quantifying the extent of oil production through free fall gravity drainage, explicitly one can compare the effect of heat on oil mobilization and production in hybrid heat-solvent recovery methods. Although in real conditions of heated solvent injection a very limited zone of the reservoir can be warmed by heat penetration at the boundaries of chamber, these
results reflect the maximum oil production achievable by warming the formation to moderate temperatures. Excluding the contribution of free fall gravity drainage from total recovery in warm Vapex can be helpful in assessing the magnitude of oil which can be produced in hybrid recovery methods under the effect of oil dilution with solvent dissolution.

4.7 Production Results of Experiment #8

Experiment #8 was conducted at T=50 ºC in the visual physical model, as mentioned earlier, while propane was injected at 817 kPa. Visible model provided the option of monitoring early development of the solvent chamber in the presence of original mobile oil in place at elevated temperatures. The rate of low viscosity undiluted oil production through free fall gravity drainage (at elevated temperature) can be affected by the applied initial pressure difference between injection and production points. The gravity head pressure available for oil flow in the current physical model with Mackay River oil was calculated to be 2.75 kPa (0.4 psi). Applied pressure difference between the wells to establish communication at the beginning of the earlier performed experiments was in the range of 0.817-1.55 MPa depending on the case. This large driving force available for oil displacement at the early stage of the experiments could accelerate rate of oil production until the pressure in the separator builds up to become very close to the injection pressure. Producing oil at high initial rate may potentially leave the top part of the model partially depleted even before the vapor chamber is fully developed. Such early horizontal growth of the chamber on top of the model could result in faster loss of gravity head available for the diluted oil drainage and a more horizontal solvent-oil interface in Vapex. According to the oil production results of Experiment #8, very limited amount of oil was produced during this displacement phase of the process. Successive images of vapour chamber in this experiment reveals that oil displacement rate under initial applied pressure difference was not high enough to deplete top of the model due to very rapid establishment of communication between the wells. Figure 4-31 and Figure 4-32 depict oil/gas injection and production profiles of Experiment #8. Figure 4-33 depicts the solvent volume fractions for Experiment #8. Since Experiment #8 was operated under the same pressure and temperature conditions as of experiment #3, extent of solvent dissolution in this experiment was observed to be fairly close to that of Experiment #3.
Figure 4-31: Gas Injection and Oil Production Profiles of Experiment #8

Figure 4-32: Gas Injection and Gas Production Profiles of Experiment #8
In this section, experimental rates of diluted oil drainage from the warm Vapex experiments presented earlier are compared against predicted rates from available analytical Vapex models in the literature. For this purpose, the total experimental production rates were corrected by excluding the oil drainage rate by free fall gravity drainage for each conducted test to quantify the diluted oil drainage rate. The simulated rates of oil production via free fall gravity drainage from Figure 4-27 were subtracted from the total production rate at the time of reaching 20% oil recovery in all conducted Vapex experiments to calculate the rate of oil production via solvent dissolution. The contribution of free fall gravity drainage and diluted oil drainage to total recovery at the time of 20% recovery is tabulated in Table 4-2. Estimated rate of oil drainage under Vapex mechanism will be compared with predicted rates from Butler, Heidari and Nenninger models in the following section.
Table 4-2: Comparison of Total Production Rate versus Free Fall Gravity Drainage Rate and Diluted Oil Drainage Rate after 20% Recovery

<table>
<thead>
<tr>
<th>Experiment#</th>
<th>T (ºC)</th>
<th>P (kPa)</th>
<th>Total Rate of Oil Production (cm³/hr)</th>
<th>Rate of Free Fall Gravity Drainage (cm³/hr)</th>
<th>Rate of Diluted Oil Drainage (cm³/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>22</td>
<td>817</td>
<td>9</td>
<td>0</td>
<td>9</td>
</tr>
<tr>
<td>2</td>
<td>40</td>
<td>817</td>
<td>13</td>
<td>2.5</td>
<td>10.5</td>
</tr>
<tr>
<td>3</td>
<td>50</td>
<td>817</td>
<td>23</td>
<td>7.8</td>
<td>15.2</td>
</tr>
<tr>
<td>4</td>
<td>60</td>
<td>817</td>
<td>40</td>
<td>17</td>
<td>23</td>
</tr>
<tr>
<td>5</td>
<td>40</td>
<td>1230</td>
<td>25</td>
<td>3</td>
<td>22</td>
</tr>
<tr>
<td>6</td>
<td>50</td>
<td>1550</td>
<td>45</td>
<td>10</td>
<td>35</td>
</tr>
</tbody>
</table>

4.8.1 Butler and Mokrys Model

Butler and Mokrys model was discussed in more detail in Chapter two. According to Butler and Mokrys model (EQ 4-13), rate of oil production in Vapex depends on reservoir characteristics (permeability, porosity and dimensions) as well as a key dimensionless number $N_s$ which depends on thermo physical properties of solvent/oil system.

$$Q = 2L\sqrt{2\kappa \phi \Delta S_o N_s H}$$  \hspace{1cm} EQ 4-13

$$N_s = \int_{C_{s,\text{min}}}^{C_{s,\text{max}}} \frac{\Delta \rho D_s (1-c_s)}{\mu c_s} \, d c_s$$  \hspace{1cm} EQ 4-14

$N_s$ can be calculated by numerical integration however concentration dependency of viscosity, density and diffusivity within diffusion layer needs to be estimated.

In the current study, density of propane-diluted Mackay bitumen within diffusion layer is approximated by a linear mixing rule based on propane density (EQ 4-7) and bitumen density (EQ 4-9) and diluted oil viscosity is calculated from modified Shu correlation (EQ 4-3) at the corresponding operating pressure and temperature of each experiment.
The diffusivity of the propane in bitumen is calculated based on a correlation presented by Das and Butler (EQ 1-9), in which diffusion coefficient was back calculated from the results of Vapex experiments with propane as solvent and Peace River Bitumen in Hele-Shaw cell.

\[ D_s = 1.306 \times 10^{-9} \mu^{-0.46} \quad \text{EQ 4-15} \]

Das and Butler (1996) expressed diffusivity by the general form of Hayduk and Cheng equation (1971), which assumes diffusivity of a particular solute in a solvent to depend only on the solvent viscosity. The dependence of diffusivity on temperature and composition appears through the viscosity, which is a strong function of temperature and composition.

In Ns calculation, the lower limit of the integral \(c_{\text{min}}\) is the minimum solvent concentration within diffusion layer and can be assumed close to zero. The upper limit of the integral \(c_{\text{max}}\) is the maximum solvent concentration within diffusion layer and would be the equilibrium solubility of propane into the oil at the Vapex operating condition. Solubility of propane into Macky oil can be estimated from EQ 4-2.

Substituting all the required parameters in EQ (4-13), dead oil production rates are calculated and tabulated in Table 2-6 to figure out how altering the operating pressure and temperature affects Vapex performance. The trend predicted by Butler model for Vapex rates at elevated temperatures and constant pressure appeared to be opposite the trend observed in the experiments. However the discrepancy between the predicted and experimental rates was found to be far less pronounced at elevated temperature and elevated injection pressures.
Table 4-3: Predicted Oil Volumetric Rate Based on Butler Model

<table>
<thead>
<tr>
<th>$T$ ($^\circ$C)</th>
<th>$P$ (kPa)</th>
<th>$C_{s,max}$</th>
<th>$N_s$</th>
<th>$\phi$</th>
<th>$K$ (Darcy)</th>
<th>$H$ (m)</th>
<th>$Q_{Butler}$ (cm$^3$/hr)</th>
<th>$Q_{Exp}$ (cm$^3$/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>22</td>
<td>817</td>
<td>0.42</td>
<td>$1.09 \times 10^{-5}$</td>
<td>0.33</td>
<td>250</td>
<td>0.3048</td>
<td>8.39</td>
<td>9</td>
</tr>
<tr>
<td>40</td>
<td>817</td>
<td>0.17</td>
<td>$7.39 \times 10^{-5}$</td>
<td>0.33</td>
<td>250</td>
<td>0.3048</td>
<td>2.17</td>
<td>15.2</td>
</tr>
<tr>
<td>50</td>
<td>817</td>
<td>0.12</td>
<td>$8.77 \times 10^{-6}$</td>
<td>0.35</td>
<td>250</td>
<td>0.3048</td>
<td>2.42</td>
<td>15.2</td>
</tr>
<tr>
<td>60</td>
<td>817</td>
<td>0.10</td>
<td>$1.39 \times 10^{-4}$</td>
<td>0.33</td>
<td>250</td>
<td>0.3048</td>
<td>2.98</td>
<td>23</td>
</tr>
<tr>
<td>40</td>
<td>1230</td>
<td>0.39</td>
<td>$4.02 \times 10^{-5}$</td>
<td>0.33</td>
<td>250</td>
<td>0.3048</td>
<td>15.96</td>
<td>22</td>
</tr>
<tr>
<td>50</td>
<td>1550</td>
<td>0.41</td>
<td>$1.48 \times 10^{-4}$</td>
<td>0.32</td>
<td>250</td>
<td>0.3048</td>
<td>30.23</td>
<td>35</td>
</tr>
</tbody>
</table>

4.8.2 Heidari Model

Heidari Model (EQ 4-16) was developed by combining the mass transfer governing equation ahead of the solvent interface in Vapex with fluid flow equation and the mass transfer ahead of the interface was resolved by assuming a constant diffusion coefficient along the diffusion layer. Similar to Butler model, the interface between the solvent chamber and virgin oil was assumed to be sharp and no capillary effect was considered. A concentration profile in the form of 3rd order polynomial was assumed for the un-steady state distribution of concentration ahead of the interface and heat integral method was employed to solve the mass transfer differential equation. Viscosity variation with concentration was considered through a power type function with order of $m$ (EQ 4-17). In the current study, $m$ values were obtained by correlating diluted oil viscosity to concentration using earlier mentioned Shu correlation for each conducted experiment at corresponding operating pressure and temperature.

Predicted rates of oil production based on Heidari model are tabulated in Table 4-4. Predicted rates based on Heidari model appear to be slightly closer to experimental rates compared to the rates from Butler model. The discrepancy between predicted rates from Butler and Heidari model can be attributed to considering a more accurate concentration distribution ahead of the solvent interface in the later model (polynomial versus exponential). The discrepancy between the experimental results and predicted values from Heidari model can be partly attributed to the fact
that concentration dependency of diffusivity on concentration within diffusion layer was ignored in the model.

\[
Q = \sqrt{\frac{2 \times 3 \times k \times g \times H \times D \times \Delta S \times \phi}{v_o}} \left[ \frac{1}{n.m + 1} - \frac{c_o}{n.m + n + 1} \right]
\]

EQ 4-16

\[
\frac{v_o}{v_{mix}} = (\frac{C}{c_o})^m
\]

EQ 4-17

| Table 4-4: Predicted Oil Volumetric Rates Based on Heidari & Maini Model |
|-----------------|-----------------|-----------------|-----------------|-----------------|-----------------|-----------------|
| T (C)           | P (kPa)         | Kinematic Viscosity, \( v_0 \) (m²/s) | Diffusivity (m²/s) | \( c_o \) (Volume Fraction) | \( n \) | \( Q \) (cm³/hr) | \( Q_{Exp} \) (cm³/hr) |
| 22              | 817             | 4.95×10⁵          | 5.66×10⁹          | 0.42             | 4.11           | 3               | 9.37            | 9               |
| 40              | 817             | 5.2×10⁴           | 1.84×10⁹          | 0.17             | 2.37           | 3               | 2.42            | 10.5            |
| 50              | 817             | 6.78×10⁴          | 1.61×10⁹          | 0.12             | 1.63           | 3               | 2.51            | 15.2            |
| 60              | 817             | 6.86×10⁴          | 1.59×10⁹          | 0.10             | 1.33           | 3               | 2.78            | 23              |
| 40              | 1230            | 2.2×10⁵           | 8.32×10⁹          | 0.39             | 3.81           | 3               | 18.14           | 22              |
| 50              | 1550            | 9.1×10⁶           | 1.27×10⁸          | 0.41             | 3.67           | 3               | 34.37           | 35              |

4.8.3 Nenninger et al Model:

Nenninger’s empirical correlation (Nenninger et al, 2008) for predicting rates of solvent based gravity drainage mechanisms was earlier discussed in Chapter two. This correlation relates the production rate to reservoir height, permeability and original oil viscosity as follows:

\[
\text{Mass flux} \left( \frac{g}{m^2\text{hr}} \right) = 43550 \left( \frac{k\phi}{\mu} \right)^{0.51}
\]

EQ 4-18

Where mass flux is the production rate from half of the vapour chamber over the production area (height (m) times width (m) of the formation) and \( \mu \) is raw oil viscosity (centipoise) at chamber temperature, \( \phi \) is porosity and \( k \) is permeability (darcies).
Nenninger explained Vapex mechanism as a shallow solvent penetration and rapid surface renewal. According to Nenninger, most of the solvent driving force is consumed at the solvent-bitumen interface in a concentration shock. On solvent rich side of the shock, diffusion happens very fast, while on the raw bitumen side, diffusion is very slow. The rate limiting step in solvent extraction mechanisms was considered to be the initial penetration of the solvent into the raw bitumen which directly depends on viscosity of raw bitumen. The logic behind this correlation is that the extraction proceeds on a pore by pore basis and not within a diffusion layer as assumed in the original Vapex model by Butler. Based on Bulter model, oil extraction happens within a diffusion layer and oil drainage rate is controlled by the solvent/bitumen thermophysical properties. However, in this recent correlation extraction is explained through a pore scale phenomena like a concentration shock which depends only on the viscosity of raw bitumen. Consequently the extraction rate in Nenninger study was found to be insensitive to the choice of solvent, injection pressure and diluted oil properties. Although the effect of temperature is considered through viscosity reduction of raw bitumen, solubility and diffusivity of solvent in the bitumen were not used as parameters in the correlation.

In the current study, experimental oil production rates of conducted tests are compared with correlated rates based on EQ 4-18 in Table 2-8. Although Neinnger correlation seems to reasonably predict oil drainage rate in Vapex experiments with injection pressures close to the vapor pressure of propane (Exp#1,5 and 6), it is very optimistic at lower injection pressures. The reason for this overestimation can be the neglect of diluted oil viscosity within diffusion layer in the proposed correlation. It should be mentioned that the correlation was based on data from experiments most of which were run at pressures close to the vapour pressure of the solvents. Therefore, it is not surprising that it fails to predict the results of tests that are not at pressures close to the vapour pressure of the solvent.
Table 4-5: Predicted Oil Volumetric Rates Based on Nenninger Model

<table>
<thead>
<tr>
<th>$T\ (C)$</th>
<th>$P\ (kPa)$</th>
<th>Dead Oil Viscosity (Pa.s)</th>
<th>Permeability (Darcy)</th>
<th>Porosity</th>
<th>Nenninger et al Model (cm$^3$/hr)</th>
<th>$Q_{Exp}$ (cm$^3$/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>22</td>
<td>817</td>
<td>301.76</td>
<td>250</td>
<td>0.33</td>
<td>7.18</td>
<td>9</td>
</tr>
<tr>
<td>40</td>
<td>817</td>
<td>32</td>
<td>250</td>
<td>0.33</td>
<td>22.38</td>
<td>10.5</td>
</tr>
<tr>
<td>50</td>
<td>817</td>
<td>13.45</td>
<td>250</td>
<td>0.35</td>
<td>35.80</td>
<td>15.2</td>
</tr>
<tr>
<td>60</td>
<td>817</td>
<td>5.12</td>
<td>250</td>
<td>0.33</td>
<td>57.25</td>
<td>23</td>
</tr>
<tr>
<td>40</td>
<td>1230</td>
<td>32</td>
<td>250</td>
<td>0.33</td>
<td>22.34</td>
<td>22</td>
</tr>
<tr>
<td>50</td>
<td>1550</td>
<td>13.45</td>
<td>250</td>
<td>0.32</td>
<td>34.35</td>
<td>35</td>
</tr>
</tbody>
</table>

Figure 4-34: Comparison of Experimental Rates with Predicted Rates by Theoretical Models
Based on the two analytical models of Butler and Mokrys and of Heidari, elevating the temperature within the range of 40 to 60 °C (at constant pressure) is expected to result in lower rate of oil production compared to the rates at ambient temperature since less solvent dissolution and consequently higher viscosities are obtained at the solvent interface. However experimental results showed an opposite behavior at elevated temperatures revealing that the viscosity of diluted oil cannot be the controlling factor of the Vapex performance on its own. At elevated temperatures, the viscosity of original oil in place was reduced considerably which provided the chance for the solvent to penetrate further into the bitumen. Thus in the presence of less viscous virgin oil, diluted oil drained within a thicker penetration zone through which the average viscosity of diluted oil may be moderately lower compared to the case of highly viscous virgin oil at lower temperatures. On the other hand in Nenngier correlation, the rate of oil drainage is totally insensitive to the reduction of oil viscosity by solvent dissolution, which is obviously incorrect and leads to large errors when the pressure is not close to vapour pressure of the solvent. On the basis of these observations, it can be concluded that neither the diluted oil viscosity nor the dead oil viscosity can be individually used to predict Vapex drainage rate at elevated temperatures. It appears to depend on both. To verify this statement, experimental rate of oil production from conducted test #1 to 6 were regressed against live bitumen viscosity and dead bitumen viscosity to infer dependency of drainage rate on these two parameters. Other parameters including characteristics of the prepared sand-packs and dispersivity are assumed to be the same in all the cases.

The obtained correlation through regression can be expressed as:

\[
Q(\text{cm}^3/\text{hr}) = \frac{1442}{\mu_{\text{Live Oil}}^{0.19} \times \mu_{\text{Dead Oil}}^{0.35}}
\]

\[\text{EQ 4-19}\]

Where Q is volumetric rate in cm³/hr and µ is viscosity in centipoise.

The correlated rates of production are compared against the experimental rates in Figure 4-35. It is noteworthy to mention that this correlation was generated based on the production rates of warm Vapex experiments utilizing propane and Mackay River bitumen within temperature range of (22 to 60 °C) and pressure range of (817 to 1550 kPa). The thickness and height of the model
and sand permeability were the same in all experiments. The contributions of these factors are imbedded in the constant 1442.

Figure 4-35: Experimental Rates versus Rates from Empirical Correlation

Assuming that the height and permeability dependence of rate in Nenninger correlation is correct, the above correlation can be put in a more useful form by separating the effects of permeability, height and the thickness from the constant 1442 using the known values of these parameters in this study. The correlation then becomes,

\[ Q' = \frac{1.7 h \sqrt{k}}{\mu_{\text{Live Oil}}^{0.19} \times \mu_{\text{Dead Oil}}^{0.35}} \]

**EQ 4-20**

Where \( Q' \) is the drainage rate in cm³/hr per cm thickness of the model, \( h \) is the model height in cm and \( k \) is the permeability of sand in Darcy.
The correlated production rates from EQ 4-20 is also compared against Vapex/warm Vapex drainage rates from conducted experiments in partially scaled physical models by Yazdani and Maini (2007), Badamchi Zadeh et al (2012) and Rezaei et al (2010) in Figure 4-36.

Yazdani conducted several Vapex experiments using butane as solvent and three different types of oil (Penny with viscosity of 18000 cp at T=15 °C, Dina with viscosity of 18648 cp at T=9 °C and Elk Point with viscosity of 18656 cp at 21.6 °C) in physical models of different heights (7.5 to 100 cm) over a permeability range of 220 to 640 Darcy. Rezaei performed warm Vapex experiments in a physical model of 42 cm height over permeability range of 220 to 830 Darcy using Cold Lake bitumen (40500 cp at T=35 °C) and Lloyminster heavy oil (5400 cp at T=35 °C). Superheated pentane within the temperature range of 35 to 50 °C was used in the study by Rezaei as solvent. Badamchi Zadeh conducted a few Vapex experiments in the same physical model as that of current study using Athabasca bitumen with a viscosity of 116 cp at T=110 °C. Mixtures of carbon dioxide and propane were injected as solvent and permeability was reported to be 630 Darcy. Viscosity of diluted oil at the test temperature and injection pressures of the conducted experiments were provided for these studies by the authors. According to Figure 4-36, the empirical correlation from this study captures the overall trend of experimental data reasonably well.
Figure 4-36: Comparison of Experimental Rates against Current Correlated Rates

From this analysis, it can be concluded that rate of oil drainage in Vapex is controlled by the average viscosity within drainage layer and thickness of the drainage layer. This drainage layer can be assumed to be a zone of varying oil viscosity starting at the solvent-oil interface with lowest viscosity and ending with the minimum viscosity required for oil mobilization. Thus the thickness of the drainage layer depends on the distance from the interface where the solvent concentration is high enough to make the oil mobile. In the case of highly viscous original oil, higher concentration of the solvent is required to mobilize the oil which occurs at the shorter distance from the solvent interface. However when the original oil is less viscous the concentration of solvent needed to make the oil mobile becomes smaller and diluted oil at a larger distance from the interface can be mobilized by solvent penetration and drainage layer would be thicker. Viscosity of diluted oil as a key parameter defining the average oil viscosity within drainage layer and viscosity of dead oil in place as the key parameter controlling the drainage layer thickness need to be considered together to evaluate the rate of oil production under different operating conditions.
Chapter Five: Numerical Simulation

This chapter discusses the numerical simulation studies performed to history match the Vapex experiments presented earlier. Through simulation studies, the effects of different mechanisms during oil dilution and drainage in the heated Vapex process can be evaluated. A numerical simulation model verified with experimental results can be used to describe the link between operating conditions, thermo-physical properties of diluted oil and oil drainage rate in Vapex at elevated temperatures. Although the performed Vapex experiments were conducted within a limited ranges of temperature (22 – 60° C), the verified simulation model can be used to extend the study to higher temperatures to evaluate the Vapex performance under a wider range of operating conditions.

Due to the complexity of the process in terms of the compositional changes within diffusion layer, a compositional simulator is the best choice for simulating the VAPEX process. Compositional simulators also facilitate accurate modeling of viscosity behavior of solvent-diluted bitumen which is a key factor in Vapex.

In the current study, history matching was done using GEM (2011.12) from Computer Modeling Group (CMG) which is an advanced general equation-of-state compositional simulator.

Full characterization of Mackay River bitumen was performed in Oil Characterization module of VMGSim (6.5).

CMG’s commercial PVT package, WINPROP (2011-11), was used for pseudoization of bitumen, property calculations and matching the PVT data through regression.
5.1 Reservoir Fluid Model

Vapex concept is based on drainage of diluted oil within a mixing layer at the boundaries of a vapour chamber. Dissolution of the injected solvent into the heavy oil/bitumen, penetration of the dissolved molecules into the oil medium and viscosity reduction of diluted oil are the key factors governing oil mobilization and drainage. Simulation of the involved fluid flow mechanism in the process is not possible without proper modeling of thermo physical properties of the solvent-oil system. Interaction of the injected solvent with oil components and extent of solvent dissolution in oil should be properly modeled at the operating pressure and temperature of the process. A variety of PVT data are required to build an equation of state that can adequately represent solvent/oil interactions. In this section, the required procedure to model phase behavior of propane/Mackay River Oil system is discussed. A PVT equation of state model is tuned against reliably measured data in WinProp PVT Package. The adjusted equation of state model along with a tuned viscosity model was then imported to the compositional simulator GEM. In each time step and for each grid block in GEM, a two phase flash calculation is performed to determine the composition of each phase in the reservoir.

5.1.1 Bitumen Characterization

Phase behaviour studies of bitumen-solvent systems require identifying the compounds present in bitumen and retrieving their properties. The chemical structure and concentration of the thousands of individual compounds present in bitumen are unknown and this makes bitumen an undefined complex mixture. Having essentially a continuous boiling point distribution, bitumen mixture can be modeled as being composed of a number of pseudo components with different boiling points. Each pseudo compounds represents many chemical compounds of varying boiling point within a certain range and has its own identity in terms of physical properties. The properties of each pseudo-component can be determined from correlations based on the boiling point and other known properties, such as specific gravity and molecular weight. These pseudo components when lumped together must be able to approximately reproduce the phase behavior and properties of the original bitumen mixture (Mehrotra et al, 1984).
A complete crude assay data from atmospheric and vacuum distillation are the most accurate way of analyzing a reservoir fluid, especially when it is combined with measured specific gravities of each cut. Through these data, mole fraction of each Single Carbon Number (SCN) present in the crude, in addition to corresponding MW and SG of each SCN, can be estimated and the crude oil can be fully characterized. In case of bitumen, the full true boiling point distribution cannot be generated through atmospheric and vacuum distillations due to possible thermal cracking of the existing compounds below their boiling point. The TBP curves from distillation process for bitumen/heavy oil usually covers less than 70% (weight) distilled off. The remaining residue needs to be characterized to preferably more than one fraction to address the bitumen characteristics. Wilson et al (1978) suggested that naturally occurring hydrocarbon mixtures show a logarithmic relationship between the mole fraction and the corresponding carbon number. According to their method, an approximate composition of the TBP residue can be obtained from the analysis of the lighter fractions. The logarithmic functionality between the mole fraction and carbon number of the lighter subfractions could be determined by a least square fit to the TBP data and then same functionality can be used to extend the trend to heavier fractions in the residue. Some other approaches have also been tried to extrapolate the TBP to 100% distilled off by fitting polynomial functions to experimental TBP curves and using artificially generated data points (Pedersen et al, 1984).

In absence of full TBP data, when only bulk properties of the hydrocarbon plus fraction are available, different mathematical distribution functions have been suggested to describe the petroleum fluid properly. Exponential models with constant and variable coefficients (to be determined specifically for each mixture) were proposed by Lohrenz (1964), Katz (1983), Pedersen (1984) and Yarborough (1978) to describe how SCN groups of different property are distributed in a mixture. These models are mainly applicable to gas condensate systems or at most to volatile oils. In another approach by Whitson (1983), a probability density function was employed to describe the distribution of the molecular weight or boiling point of the existing compounds. Whitson used a three parameter gamma probability density function (EQ 5-1) to characterize the molar distribution of the heptane plus petroleum fractions. The key parameters of the gamma probability distribution function are \(\alpha=\)shape factor, \(\eta=\)minimum molecular weight found in distribution and \(M+=\) the average molecular weight of the fluid).
\[
p(x) = \frac{(x - \eta)^{\alpha-1} e^{-\frac{x-\eta}{\beta}}}{\beta^\alpha \Gamma(\alpha)} \quad \text{EQ 5-1}
\]

Where \( \Gamma \) is the gamma distribution function and \( \beta \) is a function of three other parameters:

\[
\beta = \frac{M^* - \eta}{\alpha} \quad \text{EQ 5-2}
\]

This model was found to be able to estimate, match or extend experimental molar distribution for a wide range of reservoir fluids from gas condensate to black oil. In another attempt, Riazi (1997) developed a general distribution model applicable to different types of petroleum mixtures especially heavy oils and residues. According to this method, a versatile correlation was proposed which was able to define various properties such as absolute boiling point, molecular weight, specific gravity or refractive index parameter versus weight, mole or volume fractions. The parameters of this general equation needs to be specified for each property and may vary from one sample to another. Numerical simulators mostly employ these mathematical techniques to characterize the plus fractions. Although mathematical approaches facilitate characterization procedure, lack of valid experimental data can result in unreliable thermodynamic behaviors. Experimental data, specific to each heavy oil/ bitumen under study, should be provided to the characterization packages to build mathematical models. Mathematical models developed based on real data on TBP, average SG and MW can reliably model hydrocarbon distributions in the heavy oil and bitumen.

5.1.2 HTSD Analysis for Macky River Bitumen

In current study, Mackay River Oil was characterized using an ASTM D-7169 simulated distillation curve. ASTM D-7169 is a High Temperature Gas Chromatography Simulated Distillation (HTSD) method which is used to determine the boiling point distribution and cut point intervals of crude oils and residue up to 720 °C.

The separation is accomplished using a gas chromatograph equipped with an oven and injector that can be temperature programmed. Carbon disulfide is used as solvent to dilute the samples. Helium as a carrier gas and a high boiling point polymer (polydimethylsiloxane) as the stationary
phase coating capillary columns are used to perform the gas chromatography analysis. Flame Ionization Detector is used to detect the eluted hydrocarbons. HTSD is in fact based on the experience that hydrocarbons elute from a nonpolar adsorbent or liquid phase in order of their boiling points (Workman, 2008). Since there is a well-defined relationship between the boiling point of normal paraffins and their elute time, calibration mixtures of normal paraffin hydrocarbons are used to develop curves of retention time versus boiling points. Comparing the chromatogram of the analyzed sample with the calibration curve, the boiling point temperature distribution versus percent distilled off can be generated for the sample under study. The HTSD analysis for the bitumen used in current study was performed by Catalysis for Bitumen Upgrading and Hydrogen Production Research Lab at University of Calgary.

5.1.3 Splitting of the Bitumen into Series of Pseudo Components

The TBP distribution curve from HTSD along with measured average molecular weight (512 g/mol) and bulk liquid density data (0.988 kg/m$^3$ density at 50 °C and 1.03 MPa) were used to characterize the Mackay River Oil Bitumen. The TBP curve, bulk SG and average MW of the bitumen, were supplied to “Oil Characterization” module from VMGSim process simulator to split the Mackay River Bitumen into a series of pseudo components. The experimental distillation curve from HTSD covers up to 80% weight distilled off bitumen. VMGSim extrapolates the experimental distillation curve over the residue fraction of the bitumen by estimating the final boiling point based on values recommended by Kaes (2000) for heavy oil. In the next step VMGSim discretizes the TBP curve into 30 cuts. The simulator estimates specific gravity of individual cuts in accordance to their boiling points and assuming nearly constant Watson-Murphy K factor for the whole bitumen. The pseudo component specific gravity distribution then will be adjusted to match the bulk specific gravity of the bitumen. VMGSim estimates the molecular weight of each pseudo component based on Hariu-Sage correlation using the corresponding normal boiling point and specific gravity of each cut. Finally molecular weight distribution is adjusted to match the average molecular weight of the bitumen.
Figure 5-1 shows the boiling point distribution versus % distilled off from HTSD analysis which can be compared with regenerated boiling point distribution of the bitumen after discretization into 30 cuts in VMGSim.

![Graph showing Mackay River Oil TBP Curve and Extended TBP Analysis from VMGSim](image)

**Figure 5-1: Mackay River Oil TBP Curve and Extended TBP Analysis from VMGSim**

5.1.4 **Lumping procedure**

The time and resources required for simulating phase behaviour of petroleum fluid increase considerably with the number of components used to characterize it. Hence, the grouping together of compounds in the mixture to fewer components is required to avoid excessive run time in compositional simulators. In general, lumping procedures are based on TBP distribution curves of the petroleum fluid considering that each lumped hypothetical component represents an exclusive range of boiling point. The lumping scheme should be able to regenerate the phase behaviour and properties of the original mixture. According to Whitson (1983), the number of
fractions used to describe the petroleum fluid does not play a significant role on accuracy of EOS predictions. On the other hand, the spectra of components, which each fraction represents, along with the lumping procedure considerably affect the accuracy of phase behaviour modeling.

In current study, after splitting the bitumen into series of pseudo components (30 cuts) by discretizing the TBP curve in VMGSim, WinProp module of CMG was used for lumping procedure. The full characterization scheme from VMGSim including 30 cuts with corresponding mole fractions, molecular weights, average boiling points and specific gravities were introduced into WinProp. Three distribution functions are available in WinProp: exponential, two-stage exponential and gamma distribution. In this case, a gamma function was matched to the molecular weight distribution versus mole fraction. In the next step Log (K) lumping method by Whitson (1983) was used to generate four pseudo components to describe the Mackay River Oil. Log (K) lumping method defines pseudo components as having similar volatility and similar ranges of Log (K) values. The mole based molecular weight approach is used to match the average molecular weight of bitumen with molecular weights of the pseudo components. The properties and composition of pseudo components after this lumping procedure are presented in Table 5-1 and these were used to characterize the Mackay River Oil in next steps.

Table 5-1: Mackay River Oil Characterization with 4-PC Scheme

<table>
<thead>
<tr>
<th></th>
<th>Mole fraction</th>
<th>MW (g/mol)</th>
<th>BP (°C)</th>
<th>SG (gr/cm³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>PC1</td>
<td>0.4266</td>
<td>291.96</td>
<td>310.21</td>
<td>0.91</td>
</tr>
<tr>
<td>PC2</td>
<td>0.3390</td>
<td>517.30</td>
<td>475.59</td>
<td>0.99</td>
</tr>
<tr>
<td>PC3</td>
<td>0.1544</td>
<td>771.83</td>
<td>643.65</td>
<td>1.06</td>
</tr>
<tr>
<td>PC4</td>
<td>0.0800</td>
<td>1162.60</td>
<td>812.95</td>
<td>1.14</td>
</tr>
</tbody>
</table>

5.1.5 Comparison of Phase Envelopes

To make sure the 4- PC characterization scheme adequately represents original bitumen, the pressure-temperature phase envelopes for full characterization scheme and 4-PC scheme should be compared. In the current study, the P-T envelopes were generated in WinProp using Peng-Robinson equation of estate. The required critical properties and acentric factors of bitumen
subfractions for EOS calculation were estimated from Lee-Kesler (1975) correlation. Critical properties are correlated to boiling point (°R) and SG in Lee-Kesler correlations, as follows:

\[
T_c (°R) = 341.7 + 811 \text{SG} + (0.4244 + 0.1174 \text{SG}) T_b + (0.4669 - 3.2623 \text{SG}) 10^3 / T_b
\]

\[
\ln P_c (\text{psi}) = 8.3634 - \frac{0.0566}{\text{SG}} - (0.24244 + \frac{2.2898}{\text{SG}} + \frac{0.11857}{\text{SG}^2}) 10^{-3} T_b
\]

\[
+ (1.4685 + \frac{3.648}{\text{SG}} + \frac{0.47277}{\text{SG}^2}) 10^{-7} (T_b)^2 - (0.42019 + \frac{1.6977}{\text{SG}^2}) 10^{-10} (T_b)^3
\]

EQ 5-3

For \( \theta > 0.8 \)

\[
\omega = -7.904 + 0.1352 K_c - 0.007465 K_c^2 + 8.3590 + (1.408 - 0.01063 K_c) / \theta
\]

For \( \theta < 0.8 \)

\[
\omega = \frac{- \ln (P / 14.7) - 5.92714 + 6.09648 / \theta + 1.28862 \ln (\theta - 0.169347 \theta^6)}{15.2518 - 15.6875 / \theta - 13.4721 \ln (\theta + 0.43577 \theta^3)}
\]

Where

\[
\theta = \frac{T_b}{T_c}
\]

\[
K_c = \sqrt[3]{T_b / \text{SG}}
\]

Mehrotra et al (1985) compared the critical properties predictions for Athabasca and Peace River Bitumen subfractions based on four different correlations available in the literature (Bergman-Cavett, Lee-Kesler, Huang-Daubert and Whitson). According to that study, predicted \( T_c, P_c \) and acentric factor from the four methods are very close for the lighter pseudo components while for the very heavy pseudo components the differences are more pronounced. Among the property correlations investigated by Mehrotra et al, Lee-Kesler was found to provide the best overall results when used in Peng-Robinson EOS and is also used in current study. Critical properties and acentric factors of the defined pseudo components are tabulated in Table 5-2.
Table 5-2: Critical Properties of Mackay River Oil Pseudo Components

<table>
<thead>
<tr>
<th></th>
<th>$T_c$</th>
<th>$P_c$</th>
<th>Acentric factor</th>
</tr>
</thead>
<tbody>
<tr>
<td>PC1</td>
<td>771.268</td>
<td>19.625</td>
<td>0.655</td>
</tr>
<tr>
<td>PC2</td>
<td>926.536</td>
<td>12.608</td>
<td>1.005</td>
</tr>
<tr>
<td>PC3</td>
<td>1069.620</td>
<td>8.007</td>
<td>1.333</td>
</tr>
<tr>
<td>PC4</td>
<td>1211.565</td>
<td>5.489</td>
<td>1.571</td>
</tr>
</tbody>
</table>

The phase envelopes generated for full characterization scheme and 4-PC scheme are compared in Figure 5-2. As it can be seen in this Figure, the 4-PC scheme appears to adequately represent the Mackay River Oil and reproduce its thermodynamic behavior.

Figure 5-2: Comparison of P-T Phase Envelope for Bitumen Before and After Lumping
5.1.6 Bitumen-Propane Phase Behavior

Modeling solvent dissolution in bitumen is the first crucial step in simulating Vapex process since it has a direct influence on oil dilution, mobilization and consequently overall performance of the process. Interaction of injected solvent and hydrocarbon components is specific for every oil/solvent system and verified experimental data for every specific pair are required for accurate solubility predictions. In the current study, solubility of propane in Mackay River Oil was modeled using the experimental solubility measurements for propane/Athabasca bitumen system by Badamchi Zadeh et al (2008). The molecular weight (552 gr/mole), specific gravity (1.01 at 15.56 °C and 101.325 Kpa) and the boiling point distribution (from SimDist procedure) of Athabasca bitumen in Badamchi Zadeh study were compared with that of current study (MW=512 gr/mole, SG=1.012 at 50 °C and 1.03 MPa) and were found to be fairly close.

Okazawa (2007) compared solubility of light hydrocarbon gases in different types of bitumen used in Vapex experiments and found that solubility values in mole fractions show little dependence on the kind of bitumen; however in mass fraction they exhibit a strong dependence. Considering this point, the mass fraction-based solubility values from Badamchi Zadeh (2008) at specific pressure and temperature were converted to mole fractions and mole-based solubility values were used in the current study. The corresponding temperatures (10 to 50 °C) of the PVT measurements by Badamchi Zadeh (2008) fit well within the test temperature range of the current Vapex experiments.

Defining bitumen as a mixture of 4 pseudo components, saturation pressure calculations were performed in WinProp regression subroutine to tune the modified Peng-Robinson EOS model. Peng-Robinson equation of estate (1978) is given as:

\[
P = \frac{RT}{v-b} - \frac{a}{v(v+b)+b(v-b)} \quad \text{EQ 5-4}
\]

The constants \(a\) and \(b\) are equation of state parameters and are related to the critical pressure and temperature, acentric factor and reduced temperature for pure components as follows:
\[
a_i = 0.45724 \frac{R^2 T_{ci}^2}{P_{ci}} \left[ 1 + c_i (1 - T_{ni}^{0.5}) \right]^b
\]

\[
b_i = 0.07780 \frac{RT_{ci}}{P_{ci}}
\]

Where: 
\[c_i = 0.3796 + 1.485 \omega_i - 0.1644 \omega_i^2 + 0.01667 \omega_i^3\]

\[a\) and \(b\) for mixture of components are defined as follows:

\[a = \sum_i \sum_j x_i x_j a_{ij}\]

\[b = \sum_i x_i b_i\]

\[a_{ij} = (1 - d_{ij}) (a_i a_j)^{0.5}\]

Where, \(d_{ij}\) is the binary interaction coefficient between component \(i\) and \(j\).

\(d_{ij}\) was theoretically introduced to account for the molecular interactions between dissimilar molecules and plays a significant role in the accuracy of phase behaviour calculations. The binary interaction coefficient between hydrocarbons can be determined from the proposed equation by Li, et al, 1985 as given below:

\[d_{ij} = 1 - \frac{2\sqrt{V_i^{1/3} V_j^{1/3}}}{V_i^{1/3} + V_j^{1/3}} \sigma \]  \textbf{EQ 5-5}

where \(V_{ci}\) is the critical volume of component \(i\), and \(\sigma\) is the hydrocarbon-hydrocarbon interaction coefficient exponent. The exponent for the binary interaction \((n)\) in above equation can be chosen as a regression parameter in WinProp to match the predicted saturation pressure with experimental data. In current study a binary interaction exponent of 0.62 was found to provide the best fit to the experimental data. Experimental saturation pressures and predicted values from the tuned Peng-Robinson EOS are compared in Figure 5-3. Tuned PR-EOS was also verified against the saturation pressure data set for Mackay-River Oil/Propane at \(T=24 \, ^\circ\text{C},\)
measured by Etminan (2013). Predicted values and experimental data from Etminan (2013) are compared in Figure 5-4.

In next step, the reported experimental density data for Athabasca bitumen/propane mixture from the same study were used to improve the prediction of the liquid density by PR-EOS in Winprop model by adjusting the volume translation parameters (Referred as volume shift in WinProp).

Volume translation technique was originally suggested by Peneloux et al. (1982) to modify volumetric predictions from Soave–Redlich–Kwong-EOS. Equations of states are in fact a generalized form of gas equation of state; therefore they predict gas properties accurately, while they have deficiency in predicting liquid densities. Peneloux et al. (1982) introduced a parameter, \( c \), into Soave–Redlich–Kwong-EOS to improve the molar volume of each component in the multicomponent mixture such that:

\[
\begin{align*}
\nu'_L &= \nu_L - \sum_i c_i x_i \\
\nu'_V &= \nu_V - \sum_i c_i y_i
\end{align*}
\]

where, \( \nu'_L \) and \( \nu'_V \) are respectively the translated molar volume of mixture in liquid and vapour phase, \( \nu_L \) and \( \nu_V \) are calculated molar volume from the cubic equation of state, \( c_i \) is the volume shift value for each component, \( x_i \) and \( y_i \) are mole fractions of each component in liquid and vapour phase. Peneloux et al demonstrated that the ratios of the original fugacities for the liquid and vapour phases remain unchanged during the volume translation technique, thus the original vapour-liquid phase equilibrium calculations are not affected.

Jhaveri and Youngren (1988) extended the Peneloux volume translation to be applied in PR equation. They defined a dimensionless shift parameter \( s_i \) for any component in the mixture as:

\[
s_i = \frac{c_i}{b_i}
\]

Where \( s_i \) is the dimensionless shift parameter for component \( i \), \( c_i \) is volume shift parameter for component \( i \), and \( b_i \) is the second parameter for component \( i \) in PR-EOS. For components heavier than heptane, \( s_i \) can be obtained from the following correlation:
\[ s_i = 1 - \frac{d'}{MW_i^e} \quad \text{EQ 5-8} \]

where \(d'\) and \(e\) are positive correlation coefficients varying for paraffins, napthenes and aromatics.

WinProp uses the correlation of Jhaveri and Youngren (1988) to calculate volume shift parameter for any components. Temperature-dependent volume shift option is also available in WinProp to consider thermal expansion of liquids. Liquid densities may be underpredicted when constant volume shift parameters are used. In WinProp, temperature dependency of volume shifts is considered through Rackett compressibility factor (ZRA) which varies with temperature.

Volume shift parameters can be chosen as regression parameter to match experimental data in WinProp. In the current study, PR-EOS was matched with experimental data by adjusting volume shift parameters for propane and each pseudo component as tuning parameter. Predicted density values after tuning the PR-EOS in WinProp are compared with experimental data in Figure 5-5.

Final values of volume shift for each pseudo component after regression in WinProp is tabulated in Table 5-3.

<table>
<thead>
<tr>
<th>(\text{C}_3\text{H}_8)</th>
<th>PC1</th>
<th>PC2</th>
<th>PC3</th>
<th>PC4</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volume Shift</td>
<td>-0.0383</td>
<td>0.1477</td>
<td>0.0565</td>
<td>-0.0413</td>
</tr>
</tbody>
</table>

**Table 5-3: Final Volume Shift Values after Regression**
Figure 5-3: Comparison of Measured (by Badamchi Zadeh, 2008) and Modeled Saturation Pressures for Athabasca Bitumen/Propane Mixtures

Figure 5-4: Comparison of Measured (by Etminan, 2013) and Modeled Saturation Pressures for Mackay River Oil/Propane Mixtures
5.1.7 **Viscosity of Bitumen/Propane Mixture**

Viscosity reduction of bitumen, as a result of dilution, affects mobility of oil and plays a vital role in the fluid flow mechanism during Vapex process. Viscosity of the solvent/bitumen system significantly varies with concentration of the solvent in the mixture, pressure and temperature. Viscosity of propane diluted bitumen can be modeled through different empirical or semi empirical mixing rules. Two types of viscosity correlations are available in WinProp: the Jossi, Stiel and Thodos (JST) correlation and the Pedersen corresponding states correlation. JST correlation is recommended for modeling the viscosity behavior of light oil/solvent systems and appeared to result in huge error if inaccurate phase density data is supplied.

Pedersen viscosity correlation (1987) seems to be a better correlation to model viscosity behavior of heavy oil/solvent systems. This correlation is based on the corresponding states principle (CSP) which was originally applied to viscosity calculations by Ely and Hanley (1977). The model is based on the idea that viscosity of a component or mixture can be related to the viscosity of a reference substance at the same reduced pressure and temperature. The extensive published experimental data on viscosity of methane at different pressures and temperatures made methane a convenient choice as reference substance. The classic CSP viscosity model worked well for mixtures of light hydrocarbons but required modifications for mixtures containing heavy hydrocarbons. Pedersen et al. (1984) modified the classic model to express the deviations from classical corresponding states principle (CSP) for heavier mixtures by embedding a parameter called, $\alpha$, to the original model as follows:

$$
\frac{\mu_{mix}(P, T)}{\mu_0(P_0, T_0)} = \left(\frac{T_{C, mix}}{T_{C0}}\right)^{-1/6} \left(\frac{P_{C, mix}}{P_{C0}}\right)^{2/3} \left(\frac{MW_{mix}}{MW_0}\right)^{1/2} \left(\frac{\alpha_{mix}}{\alpha_0}\right)
$$

EQ 5-9

where, $\mu$ is viscosity, $T_C$ and $P_C$ are critical temperature and pressure (kPa), $MW$ is molecular weight, and $\alpha$ is the correction parameter to CSP model. The subscript “mix” refers to the mixture property, and the subscript “0” refers to the methane property.

$MW_{mix}$ is calculated from:

$$MW_{mix} = b_1 \times (MW_w^{b_2} - MW_n^{b_2}) + MW_n
$$

EQ 5-10
Where \( MW_w \) is the weight fraction averaged molecular weight, and \( MW_n \) is the mole fraction averaged molecular weight and \( b_1 \) and \( b_2 \) are coefficients.

The CSP correction parameter is calculated as follows:

\[
\alpha = 1 + b_3 \times \rho_r^{b_4} \times MW_{\text{mix}}^{b_5}
\]

EQ 5-11

where, \( \rho_r \) is the reduced density of the reference substance and \( b_3, b_4, \) and \( b_5 \) are coefficients. Pedersen and Fredenslund (1987) stretched the performance of earlier Pedersen model into the region where methane is in solid form and could improve the results for heavy oil mixtures. Pedersen and Fredenslund (1987) correlation is available in WinProp and the corresponding coefficients can be adjusted through regression to match experimental viscosity data. Table 5-4 shows the final values of Pedersen (1987) viscosity model coefficients after tuning with Badamchi Zadeh measured data.

Figure 5-6 depicts the experimental data along with predicted viscosities of Athabasca Bitumen/Propane mixture from modified Pedersen correlation.

![Figure 5-5: Comparison of Predicted Liquid Densities from Tuned PR-EOS and Experimental Data from Badamchi Zadeh (2008)]
Table 5-4: Modified Pedersen Viscosity Correlation Coefficients after Regression

<table>
<thead>
<tr>
<th></th>
<th>b&lt;sub&gt;1&lt;/sub&gt;</th>
<th>b&lt;sub&gt;2&lt;/sub&gt;</th>
<th>b&lt;sub&gt;3&lt;/sub&gt;</th>
<th>b&lt;sub&gt;4&lt;/sub&gt;</th>
<th>b&lt;sub&gt;5&lt;/sub&gt;</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.00015789</td>
<td>2.923535700</td>
<td>0.005439400</td>
<td>3.765241800</td>
<td>0.169500000</td>
</tr>
</tbody>
</table>

Figure 5-6: Comparison of Measured Viscosities of Athabasca Bitumen/Propane Mixture (Badamchi Zadeh) and Predicted Values from Tuned Modified Pedersen Model

5.2 Reservoir Model Description

In this study, GEM (the equation of state (EOS) compositional module from CMG 2011) which is a finite difference simulator was used to history match the Vapex experiments. GEM solves the material balance equation (convection-diffusion), energy balance equation (convection-conduction) and phase equilibrium equations simultaneously for oil, gas and water components.
In order to simulate the current Vapex experiments, a symmetrical Cartesian model with the same dimensions as of the unrolled cylindrical physical model in the experiments was created. This Cartesian model (Figure 5-7) consisted of 84 grid blocks along the X-axis, 2 grid blocks along the Y-axis and 30 grid blocks along the K-axis. The grid block size is 1 cm in all directions. In Vapex simulation, it is essential to choose the grid sizes fine enough to be able to capture the mass transfer and fluid flow mechanisms involved within diffusion layer at the boundaries of vapor chamber. Building the simulation model with large grid blocks may result in unrealistic diffusion layer thicknesses and unreliable oil drainage rates (Yazdani et al, 2009).

The sand pack properties including absolute permeability, porosity and connate water saturation in each case of history match were defined in the simulation model to be homogeneous and honor the properties of the prepared sand pack in the physical model experiments (Table 3-4).

The model is assumed to contain two phases of bitumen and water with no solution gas. The bitumen is defined as four pseudo components discussed earlier.

Initial temperature of the reservoir model in each case was set to be the same as the test temperature of the Vapex experiment. Since heat loss to the ambient was avoided during the experiments and physical model temperature was controlled to be constant during the tests, no heat loss is considered in the simulation model.

5.3 Rock and Fluid Properties

Since there was no available experimental data on the rock-fluid properties in the current study, three phase relative permeability model of Corey along with tabulated typical end point relative permeability values from the literature (Table 5-5) were used to initially generate the oil-water and gas-oil relative permeability curves. The vertical and horizontal permeabilities were kept the same everywhere in the models with no heterogeneity. Due to the large permeability of the sand pack the capillary pressure was assumed to be insignificant. No mobile water content was assumed in all of the simulation models and the reservoir model in simulation was homogenously at connate water saturation to mimic the prepared sand packs for the conducted experiments.
Figure 5-7: Numerical Domain for Vapex Model (84×30×2)

Table 5-5: Rock and Fluid Properties

<table>
<thead>
<tr>
<th>Property</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>SWCON: Connate Water</td>
<td>0.15</td>
</tr>
<tr>
<td>SWCRIT: Critical Water</td>
<td>0.15</td>
</tr>
<tr>
<td>SOIRW: Irreducible Oil for Water-Oil</td>
<td>0.1</td>
</tr>
<tr>
<td>SORW: Residual Oil for Water-Oil Table</td>
<td>0.1</td>
</tr>
<tr>
<td>SOIRG: Irreducible Oil for Gas-Liquid Table</td>
<td>0.1</td>
</tr>
<tr>
<td>SORG: Residual Oil for Gas-Liquid Table</td>
<td>0.1</td>
</tr>
<tr>
<td>SGCON: Connate Gas</td>
<td>0.001</td>
</tr>
<tr>
<td>SGCRT: Critical Gas</td>
<td>0.005</td>
</tr>
<tr>
<td>KROC: Kro at Connate Water</td>
<td>1</td>
</tr>
<tr>
<td>KRWIRO: Krw at Irreducible Oil</td>
<td>1</td>
</tr>
<tr>
<td>KRGCL: Krg at Connate Liquid</td>
<td>1</td>
</tr>
<tr>
<td>( n_w, n_{ow}, n_{og}, n_g )</td>
<td>2</td>
</tr>
<tr>
<td>Rock Compressibility (1/kPa)</td>
<td>1.088E-7</td>
</tr>
<tr>
<td>Initial Reservoir Pressure (kPa)</td>
<td>101.5</td>
</tr>
</tbody>
</table>

5.4 Well Location and Constraints

Injection and production wells are located halfway along the model at the top and bottom layer of the simulation model which mimics the exact injection and production points of the physical model used in the experiments and results in full Vapex chambers. Well radii required modifications to fit in the fine grid blocks in the simulation model. In order to facilitate gas breakthrough in Vapex simulation model, a one grid size column of mobile water was defined.
between the injection and production wells which helps in rapid establishment of communication between wells. Since breakthrough occurred in less than half an hour in the physical experiments, early communication between wells in the simulation model helps to history match the results right after the gas breakthrough and helps to skip the uncertainties and complexities involved during early stages of the process. The well constraint for injection well in simulation model is defined as maximum bottom-hole pressure (0.817 MPa in first four experiments, 1.23 MPa in fifth experiment and 1.55 in sixth experiment) while the production well constraint was chosen to be at maximum surface gas rate (0.05-0.155 m3/day depending on the experiment) to imitate the exact operating conditions in conducted Vapex experiments.

Since the simulator does not differentiate between solution and free gas rate, the total produced gas rates of experiments are used as production well constraint. Increasing the produced gas rate will not affect the amount of dissolved gas in to the oil but will adjust rate of free gas production. Thus by manipulating the total gas production, free gas production can be adjusted while solution gas rate depends on solvent dissolution in to the oil under operating conditions of experiments.

5.5 History Matching Results-Validation of Simulation Model

The history matches were achieved by understanding the effect of different parameters on the simulation performance. The uncertain parameters including relative permeabilities and dispersivity coefficients which could not be directly measured in current experiments, besides viscosity coefficients and numerical control parameters were adjusted to obtain a reasonable match of the observed cumulative oil and gas production.

The most suitable combination of parameters for relative permeability, dispersivity and viscosity was searched to match the oil production and gas injection/production profiles of first four experiments at the same time. These parameters were obtained through trial and error procedure by matching each experiment in each step, moving to the next experiment, retuning the parameters and getting back to earlier experiments to re-match the results with modified parameters. In the following sections, this combination of parameters which was able to match the first four experiments is discussed.
5.5.1 Adjusting Numerical Constraints

Numerical constraints play a significant role on stability and accuracy of the numerical simulations. Among different constraints, the first time step size used after a well change, minimum and maximum sizes of time steps required modifications to minimize the evolved material balance error and avoid convergence failures in conducted simulation studies. Table 5-6 summarizes the adjusted numerical constraints in each case of history match.

<table>
<thead>
<tr>
<th>Experiment #</th>
<th>First Time Step Size after Well Change (day)</th>
<th>Minimum Time Step Size (day)</th>
<th>Maximum Time Step Size (day)</th>
<th>Material Balance Error (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>10^{-8}</td>
<td>10^{-12}</td>
<td>5 \times 10^{-5}</td>
<td>6</td>
</tr>
<tr>
<td>2</td>
<td>10^{-16}</td>
<td>10^{-16}</td>
<td>5 \times 10^{-4}</td>
<td>1.084</td>
</tr>
<tr>
<td>3</td>
<td>10^{-16}</td>
<td>10^{-16}</td>
<td>5 \times 10^{-4}</td>
<td>0.26</td>
</tr>
<tr>
<td>4</td>
<td>10^{-15}</td>
<td>10^{-15}</td>
<td>5 \times 10^{-4}</td>
<td>0.096</td>
</tr>
<tr>
<td>5</td>
<td>10^{-17}</td>
<td>10^{-13}</td>
<td>5 \times 10^{-4}</td>
<td>0.97</td>
</tr>
<tr>
<td>6</td>
<td>10^{-10}</td>
<td>10^{-12}</td>
<td>5 \times 10^{-5}</td>
<td>0.234</td>
</tr>
</tbody>
</table>

5.5.2 Adjusting Relative Permeability Parameters

Since sand pack was prepared consistently for all conducted experiments, not a considerable difference in sand pack properties is expected from case to case. Considering this, we tried to find the relative permeability parameters which can be consistently used in history matching of all conducted experiments. Figure 5-8 and Figure 5-9 display the relative permeability curves after adjustment and the final modified relative permeability parameters are tabulated in Table 5-7.

<table>
<thead>
<tr>
<th>KROCW: Kro at Connate Water</th>
<th>1</th>
<th>n_w</th>
<th>2</th>
</tr>
</thead>
<tbody>
<tr>
<td>KRWIRO: Krw at Irreducible Oil</td>
<td>1</td>
<td>n_{ow}</td>
<td>2</td>
</tr>
<tr>
<td>KRGCL: Krg at Connate Liquid</td>
<td>1</td>
<td>n_{og}</td>
<td>4</td>
</tr>
<tr>
<td></td>
<td></td>
<td>n_g</td>
<td>1.5</td>
</tr>
</tbody>
</table>
Figure 5-8: Oil/ Water Relative Permeability Curves after Adjustments

Figure 5-9: Liquid/ Gas Relative Permeability Curves after Adjustments
5.5.3 Adjusting Dispersivity Parameters

According to the original Vapex theory, mixing of solvent and oil within diffusion layer is controlled through molecular diffusion. Earlier studied by Das and Butler (1998), Boustani and Maini (2001), Oduntan et al (2001), Nghiem et al (2001), Yazdani and Maini (2007), Alkindi et al (2011) and Cuthiell et al (2012) revealed that diffusion coefficients much larger than physical diffusion coefficients of solvent/bitumen systems are required in Vapex analytical models and commercial simulators to address the extent of mass transfer involved in actual Vapex process. The observed enhanced mass transfer in Vapex process is attributed to convective dispersion in porous media. Bear (1972) defined convective dispersion as spreading or mixing of a component in a phase as a result of microscopic variations in the velocity field. According to Lake (1989), total dispersion depends on diffusion coefficient, local velocity gradients, locally heterogeneous streamline lengths, and mechanical mixing. Total dispersion is known to be comprised of two components of longitudinal and transverse dispersion. Longitudinal dispersion happens where both solute and solvent are flowing in the same direction as of the bulk flow. However transverse dispersion develops in the direction perpendicular to the bulk flow. According to Lake (1989), the total dispersion tensor can be expressed as follows:

\[
D_{ik} = \begin{bmatrix}
D_{ik}^{xx} & D_{ik}^{xy} & D_{ik}^{xz} \\
D_{ik}^{yx} & D_{ik}^{yy} & D_{ik}^{yz} \\
D_{ik}^{zx} & D_{ik}^{zy} & D_{ik}^{zz}
\end{bmatrix}
\]

EQ 5-12

where the tensor components are expressed as:

\[
D_{ik}^{xx} = \frac{D^*_{ik}}{\tau_k} + \frac{(\alpha_{kl} - \alpha_{kl})}{\phi S_k} u_{kx}^2 + \frac{\alpha_{kl} |u_k|}{\phi S_k}
\]

EQ 5-13

\[
D_{ik}^{yy} = \frac{\alpha_{kl} - \alpha_{kl}}{\phi S_k} u_{ky} u_{ky}
\]

EQ 5-14

\[
\tau_k = \frac{F \phi S_k}{u_{ky}}
\]

EQ 5-15
\[ |u_k| = \sqrt{(u_{kx}^2 + u_{ky}^2 + u_{kz}^2)} \]  \hspace{1cm} \text{EQ 5-16}

which \( D_{ik} \) is molecular diffusion of component \( i \) in phase \( k \), \( u_{kx} \) is the velocity vector component in direction of \( x \), \( \alpha_{il}, \alpha_{il} \) are longitudinal and transverse dispersivity parameters for phase \( k \), \( \tau_k \) is the tortuosity and \( F \) is formation resistivity factor.

GEM simulator, employs the above mentioned tensor to address total mass transfer flux including diffusion flux and dispersive flux. According to EQ 5-13 and EQ 5-14, contribution of dispersivity is incorporated into mass transfer by employing a constant multiplier to the phase velocity. Through velocity dependant dispersion tensor, dependency of dispersion coefficient on solvent concentration within solvent-bitumen mixing layer can be implicitly considered in Vapex simulation.

According to the geometry of flow in Vapex process, with solvent penetrating into the oil in the direction perpendicular to the bulk flow of diluted oil, longitudinal dispersivity seem not to play a vital role in oil-solvent mixing. Sensitivity analysis performed in current study confirmed that longitudinal dispersivity has almost no effect on solvent concentration within mixing layer and on oil drainage rate in Vapex simulation. However, larger transverse dispersivity values were found to result in thicker mixing layers and higher drainage rates.

The transverse dispersivity parameter \( (\alpha_{il}) \) was adjusted to \((0.009 \text{ m})\) to history match the results of first four conducted Vapex experiments (Experiments at \( T=22, 40, 50 \) and \( 60 \) °C at \( 815 \) kPa). The embedded value of transverse dispersion parameter resulted in total dispersion coefficients within the range of \( 10^{-5}-10^{-6} \text{ (m}^2/\text{s}) \) depending on the velocity of the draining oil.

The obtained dispersion values in current history matching were found to be comparable with dispersion values in earlier studies by Nghiem et al (2001), Yazdani et al (2009) and Das (2005).

Yazdani et al (2009) carried out an extensive analysis to study the hidden effect of numerical dispersion in Vapex numerical simulation. Numerical dispersion is defined as the truncation error evolved from discretization procedure applied to convection-diffusion differential equation. Yazdani (2009) used the general form of the numerical dispersion tensor (EQ 5-15), suggested by Fanchi (1983) for a fully implicit formulation in 2-D problems, to quantify the contribution of
numerical dispersion into the solution of Vapex problem. According to this equation, numerical dispersion depends on velocity components, grid sizes and time steps.

\[ D_{\text{num}} = \begin{bmatrix} \frac{u_x}{2} \left( \Delta y + u_x \frac{\Delta t}{\phi} \right) & u_x u_y \frac{\Delta t}{2\phi} \\ u_x u_y \frac{\Delta t}{2\phi} & \frac{u_y}{2} \left( \Delta y + u_y \frac{\Delta t}{\phi} \right) \end{bmatrix} \]  \text{EQ 5-17}

According to the Vapex experimental results by Yazdani and Maini (2007), dispersion coefficient was found to be scale dependant. Yazdnai history matched the experimental oil production profile in GEM simulator by adjusting dispersivity parameters. Depending on the grid size system, dispersivity coefficients in the range of 1e-9 to 1e-8 m^2/s was required to address the extent of mass transfer in the conducted experiments. The involved numerical dispersion was found to be in the range of 5.95e-8 to 1.16 e-6 m^2/s in that study which is very significant.

Through those grid sensitivity analyses, it was concluded that unrealistically small grid size is required in the simulators to overcome the numerical dispersion. In a numerical study in conjunction with experimental study by Das (2005), the GEM simulator was found to be able to capture the extent of mass transfer occurring in Vapex at macroscopic level. However simulator was not able to simulate the complicated mass transfer mechanism occurring mostly at the solvent-oil interface at the microscopic level. Das concluded that artificially large values of dispersion coefficients should be employed in the simulator to match the production oil rate while at the same time simulator failed to match the solvent saturation profile and resulted in unrealistic penetration depth.

In the current study, a grid sensitivity analysis was performed. The numerical dispersion values were calculated for each experiment and are presented in Table 5-8. Reducing the grid block size to less than 1 cm appeared to cause insignificant effect on the final results. Grid sensitivity analysis can be helpful in reducing the relative dominance of numerical dispersion. However the embodied dispersion coefficients are not indicative of the actual dispersion involved in diffusion layer in Vapex process.
To substantially reduce the effect of numerical dispersion in the simulation, order of magnitude smaller grid sizes were required which on the other hand caused instabilities in numerical simulation due to the very small time steps used in simulation.

### Table 5-8: Comparison of Numerical and Physical Dispersion

<table>
<thead>
<tr>
<th>Experiment#</th>
<th>Numerical Dispersion (m²/s)</th>
<th>Physical Dispersion (m²/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>$4.24 \times 10^{-6}$</td>
<td>$1.28 \times 10^{-5}$</td>
</tr>
<tr>
<td>2</td>
<td>$2.42 \times 10^{-6}$</td>
<td>$7.35 \times 10^{-6}$</td>
</tr>
<tr>
<td>3</td>
<td>$1.46 \times 10^{-6}$</td>
<td>$4.4 \times 10^{-6}$</td>
</tr>
<tr>
<td>4</td>
<td>$1.27 \times 10^{-6}$</td>
<td>$3.85 \times 10^{-6}$</td>
</tr>
<tr>
<td>5</td>
<td>$8.27 \times 10^{-6}$</td>
<td>$2.5 \times 10^{-5}$</td>
</tr>
<tr>
<td>6</td>
<td>$1.9 \times 10^{-5}$</td>
<td>$5.7 \times 10^{-5}$</td>
</tr>
</tbody>
</table>

#### 5.5.4 Adjusting Viscosity Coefficients

As it was earlier discussed, the viscosity model employed in current simulations was tuned against the experimental data from Badamchi Zadeh (2013). The original dead oil used in propane-diluted bitumen samples in Badamchi Zadeh measurements was found to be more viscous than Mackay-River Oil used in current Vapex experiments. Moreover, those measured data were limited to specific ranges of pressure and temperature. Viscosity of the original dead oil in place in current study has a strong effect on free fall gravity drainage which should be considered in evaluating the process. Thus tuned viscosity correlation in simulation model should be able to cover a broader range of viscosity from dead oil to diluted oil at least under specific pressure and temperatures corresponding to current experiments. For these reasons, the earlier developed viscosity model was modified during history matching to be able to obtain reasonable matches.

Figure 5-10 depicts the viscosity profile of propane-diluted bitumen at different temperatures and propane concentration (mole %) after re-adjusting the modified Pedersen viscosity coefficients.
Figure 5-10: Viscosity of Propane-Diluted Athabasca Bitumen after Adjusting Modified Pedersen Viscosity Correlation Coefficients

Figure 5-11 to Figure 5-14 depicts the oil production, gas injection and production profiles obtained by simulating the first four Vapex experiments. Through these figures, results of history matches can be compared with corresponding experimental profiles. The propane-content and viscosity maps of the vapour chamber after 24 hours are depicted in Figure 5-15 to Figure 5-18.
Figure 5-11: History Match of Vapex Experiment #1: P=817 kPa, T=22 °C

Figure 5-12: History Match of Vapex Experiment #2: P=817 kPa, T=40 °C
Figure 5-13: History Match of Vapex Experiment #3: P=817 kPa, T=50 ºC

Figure 5-14: History Match of Vapex Experiment #4: P=817 kPa, T=60 ºC
Figure 5-15: Simulated Propane-Content and Viscosity Maps within Vapour Chamber after 24 hours in Experiment #1: P=817 kPa, T=22 °C
Figure 5-16: Simulated Propane-Content and Viscosity Maps within Vapour Chamber after 24 hours in Experiment #2: P=817 kPa, T=40 °C after 24 hours
Figure 5-17: Simulated Propane-Content and Viscosity Maps within Vapour Chamber after 24 hours in Experiment #3: P=817 kPa, T=50 °C
Figure 5-18: Simulated Propane-Content and Viscosity Maps within Vapour Chamber after 24 hours in Experiment #4: P=817 kPa, T=60 °C
5.6 Discussion of Numerical Results

From the results of history match, it is evident that the developed numerical simulation model which was equipped with proper PVT and viscosity model could reasonably match the experimental data. The developed simulation model which was verified with the experimental results can be helpful in confirming the discussions provided in the preceding chapter regarding the performance of Vapex at elevated temperatures.

It is noteworthy to emphasize one more time on the difference between performance of Vapex at elevated formation temperatures with that of classic Vapex due to mobility of original oil in place in the former. As it was mentioned in an earlier chapter, even at moderate formation temperatures, dead oil within porous matrix was found to able to drain toward production well through free fall gravity drainage mechanism in conducted warm Vapex experiments. Accordingly, oil was produced not only by dilution at the boundary of vapor chamber but also from the un-contacted zones where gravity driving force was sufficient to overcome the viscous force resisting against oil flow toward the production point. Increasing the formation temperature dramatically increases the contribution of free fall gravity drainage to total oil production. According to the experimental results of test #2, 3 and 4 (40, 50 and 60 °C) in Figure 4-13, oil production at higher temperatures (same propane injection pressure) was accelerated at the earlier times of the experiments due to free fall gravity drainage.

The results of simulation study confirm that developed simulation model could properly reproduce the involved diluted oil drainage and free fall gravity drainage mechanisms in these four experiments. As it was expected from the experimental results, oil was initially produced at higher rate when formation temperature was increased in the simulation model due to faster free fall gravity drainage of dead oil in place. From the Viscosity maps in Figure 5-15 to Figure 5-18, it can be realized that viscosity of original oil in place was reduced from 302,000 to 32,000, 13,000 and 5100 cp by increasing the temperature from 22 to 40, 50 and 60 °C respectively. Reduction of dead oil viscosity at elevated temperatures resulted in faster oil production by free fall gravity drainage and faster lateral growth of the chamber on top of the simulation model.
As the simulation results show, viscosity of diluted oil at the solvent interface slightly increased due to less dissolution of propane into the oil at elevated test temperatures (22 ºC to 60 ºC). The calculated propane solubility and viscosities of diluted oil at the solvent interface by simulator at different temperatures are tabulated in Table 5-9. The calculated propane content in the production grid of the simulation model was compared with experimental solvent volume fractions and was found to be within 10% of each other.

In the simulation model, viscosity of oil at the interface slightly increased from 48 to 92, 156, 183 cp by increasing the temperature from 22 to 40, 50 and 60 ºC respectively. In spite of this slight increase, overall effect of elevating the temperature was confirmed in simulation studies to be helpful in increasing the overall rate of oil production.

This explanation can be also verified by evaluating the viscosity and solvent mole fraction maps, along with the shape of the vapour chamber, in cross section view of the simulation model at T = 40, 50 and 60 ºC in Figure 5-19. As it can be seen in this figure, at 60 ºC in spite of less solvent content and higher viscosity of diluted oil at the interface, lateral spread of vapour chamber on top of the model happens faster due to lower viscosity of dead oil in place. As the shape of the vapour chamber implies, the rapid depletion of the top part of the model by free fall gravity drainage at higher temperatures causes higher initial rate of oil production in these experiments. Faster lateral growth of the chamber on top of the formation at elevated temperatures could also affect rate of mass transfer by enhancing the area available to mass transfer, which accelerates oil production.

<table>
<thead>
<tr>
<th>Temperature (ºC)</th>
<th>( \mu_{\text{diluted oil at solvent interface (cp)}} )</th>
<th>Solvent mole fraction in simulation</th>
</tr>
</thead>
<tbody>
<tr>
<td>20</td>
<td>48</td>
<td>0.74</td>
</tr>
<tr>
<td>40</td>
<td>92</td>
<td>0.50</td>
</tr>
<tr>
<td>50</td>
<td>156</td>
<td>0.41</td>
</tr>
<tr>
<td>60</td>
<td>182</td>
<td>0.34</td>
</tr>
<tr>
<td>70</td>
<td>173</td>
<td>0.29</td>
</tr>
<tr>
<td>80</td>
<td>143</td>
<td>0.25</td>
</tr>
<tr>
<td>90</td>
<td>114</td>
<td>0.22</td>
</tr>
</tbody>
</table>

Table 5-9: Viscosity and Propane Content of Diluted Bitumen in Simulation Study
Figure 5-19: Viscosity and Propane-Content Maps within Vapor Chamber in Simulation Models at T=60, 50 and 40 °C, after 3 hours
5.7 Extending the simulation studies to higher temperatures

Once the simulation model including the tuned Peng-Robinson equation of state along with the adjusted viscosity correlation was validated through matching the simulation results of oil and gas production with obtained experimental production profiles, the model was used to extend the study to a wider range of temperature. The idea was to evaluate dependency of Vapex oil drainage rate to formation temperatures beyond 60 °C by understanding the link between oil production rate and thermo physical properties of bitumen/propane system at elevated temperatures.

In order to simulate the warm Vapex performance at different temperatures (70 to 90 °C), gas production rates were defined to be the same in all simulation models. Although solution gas rate is expected to be different at different temperatures, fixing the total gas rate will not inhibit solvent dissolution and production. In case more solution gas is released in production well, free gas production rate is adjusted by the simulator to keep the total gas production rate fixed in all simulation cases. Total gas production rate was reasonably set by trial and error to make sure in all cases free gas production rate was enough to maintain two phase gas and oil flow in the production well. The simulated results of oil production in warm Vapex at 70 to 90 °C are presented in Figure 5-20. From this graph, it can be realized that increasing the temperature beyond 60 °C will significantly improve rate of oil production with same amount of injected solvent. Viscosity and solvent mole fraction maps of the simulated warm Vapex cases at 70, 80 and 90 °C are depicted in cross section views in Figure 5-21 and are summarized in Table 5-9.

Solubility of propane into the bitumen is very sensitive to pressure. Injection of propane at same pressure, while increasing the test temperature results in less dissolution of the solvent in the oil. Although solvent dissolution declines at higher temperatures, viscosity of diluted oil within diffusion layer appeared to be decreasing due to positive effect of temperature increase on viscosity reduction. At higher temperatures, oil drainage rate is improved not only due to lower viscosity of draining diluted oil but also due to faster oil production by free fall gravity drainage at temperatures above 60 °C.
Figure 5-20: Predicted Cumulative Oil Production at T=70, 80 and 90 ºC from Simulation Study

The warm Vapex performance at 90 ºC in the current study, in which whole formation is warmed up to the target temperature, converges to the thermal-based end of the hybrid heat-solvent spectrum like ES-SAGD. On the other hand, results at milder temperatures, where solvent plays a more significant role in oil mobilization and drainage, can be helpful in quantifying the maximum possible rate obtainable in solvent-based end of the spectrum like N-Solv.
Figure 5-21: Viscosity and Propane-Content Maps within Vapor Chamber in Simulation Models at T=90, 80 and 70 ºC, after 3 hours
5.8 Chamber Growth: Simulation versus visible model

Experiment #8 was conducted in a visible rectangular model to monitor the lateral growth of vapour chamber during Vapex at elevated temperature. Operating condition of experiment #8 were chosen to be the same as experiment#3 (T=50 °C, P=817 kPa).

The reservoir simulation model created for this experiment is a rectangular model with 68×3×15 grid blocks of 1cm size in i, j and k direction to mimic the dimensions of the employed physical model. The earlier developed PVT model which was verified with results of earlier experiments was initially used in this case to re-evaluate the simulation model. To history match experiment #8, PVT model needed no modification and reasonable match could be obtained without adjusting dispersivity coefficient. However relative permeability parameters were slightly adjusted to be able to match dead oil production and gas/injection experimental profiles. Exponent for calculating $k_{\text{rog}}$ ($n_{\text{og}}$) was modified to 5, $n_w$ was modified to 3 and $k_{\text{ROCW}}$ was adjusted to 0.92 to match the experimental results. Figure 5-22 depicts the simulation results of experiment#8 after history match.

In earlier experiments, comparison of shape of the simulated chamber with developed chambers during experiments was not possible. Conducting experiment #8 in a visible model provided the chance of tracking the lateral growth of the chamber during the experiment. The chamber growth in numerical simulation and conducted experiment are compared in Figure 5-23. As it can be seen in Figure 5-23, simulation model could adequately represent the chamber expansion in the actual conducted warm Vapex experiment at T=50 °C.
Figure 5-22: History Match of Vapex Experiment #8 in visible model

Figure 5-23: Comparison of Chamber Development in Simulation Model and Visible Physical Model
5.9 Effect of Increased Injection Pressure on Vapex Performance

Results of experiment #5 and #6, in which injection pressures of propane were increased compared to the experiment# 2 and #3 at the same temperature, were discussed in the preceding chapter. In this section, the simulation results of the experiment #5 and #6 after history matching will be presented and compared with that of lower pressure experiments.

During history matching it was tried to keep the consistency between these two cases and four earlier history matched experiments. Therefore, relative permeability parameters and dispersivity values were kept the same as earlier verified simulation model. The porosity and residual connate water saturation for these two cases were chosen based on the characteristics of the prepared sand-pack as tabulated in Table 3-3.

The well constraint for injection well in the simulation model of experiment #5 was defined as maximum bottom-hole pressure (1.230 MPa) while the production well constraint was set at maximum surface gas rate within the range of (0.15 to 0.46 m$^3$/day).

The constraint for injection well in the simulation model of experiment #6 was defined as maximum bottom-hole pressure (1.550 MPa) while the production well constraint was set at maximum surface gas rate within the range of (0.1 to 0.4 m$^3$/day) to imitate the experimental operating conditions.

Same PVT model as of the earlier four experiments was used to model propane dissolution into bitumen to history match the injection and production results of these two later experiments. On the other hand the earlier developed viscosity correlation required some adjustment to be able to history match the results of experiment #5 and #6. Results of the simulation models appeared to be very sensitive to the embedded viscosity correlation. Since free fall gravity drainage played a significant role in performed warm Vapex experiments, it was essential to develop a viscosity model which was valid over a wide range from hundred thousands of cp (dead oil viscosity) to less than 10 cp (diluted oil viscosity). The modified Pedersen viscosity correlation which was addressed in section 5.1.7 was originally proposed by Pedersen and Fredenslund (1987) for a limited viscosity range of less than 10cp. Thus the tuned viscosity correlation used to history match experiment #1 to #4, in which viscosity was varied between 300,000 to 100 cp from dead
oil to diluted oil at the solvent interface, failed to predict viscosity of diluted oil under operating pressures of experiment #5 and #6, in which viscosity was lowered to less than 30 cp. Therefore, in simulation models of experiment #5 and #6, viscosity coefficients in modified Pedersen correlation were adjusted so that viscosity correlation can model viscosity behavior of dead oil as well as diluted oil under the corresponding operating pressure and temperatures. Tuned viscosity parameters leading to the final history match for these two experiments are tabulated in Table 5-10. The results of match to gas injection and oil/gas production profile for experiments #5 and #6 are presented in Figure 5-24 and Figure 5-25 respectively.

Table 5-10: Modified Pedersen Viscosity Correlation Coefficients after Regression

<table>
<thead>
<tr>
<th>Experiment #</th>
<th>b₁</th>
<th>b₂</th>
<th>b₃</th>
<th>b₄</th>
<th>b₅</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>0.0000961</td>
<td>2.58458</td>
<td>0.020989</td>
<td>5.3768</td>
<td>0.15834</td>
</tr>
<tr>
<td>6</td>
<td>0.0000834</td>
<td>2.38527</td>
<td>0.012146</td>
<td>2.6595</td>
<td>0.33107</td>
</tr>
</tbody>
</table>

Figure 5-24: History Match of Vapex Experiment #5: P=1230 kPa, T=40 °C
Figure 5-25: History Match of Vapex Experiment #6: P=817 kPa, T=50 °C

The viscosity and solubility distributions within the simulated vapor chamber after 25 hours can be compared between experiment #5 and experiment #2 in Figure 5-26.

Experiments #2 and #5 both were conducted at 40 °C while injection pressure was increased from 817 to 1230 kpa. Theoretically increasing pressure at the same temperature results in more dissolution of the solvent into bitumen, more viscosity reduction and consequently faster oil drainage rate. Considering the fact that rate of free fall gravity drainage beyond the vapor chamber depends on the operating temperature and is not significantly affected by injection pressure, same magnitude of oil production via gravity drainage is expected in these two experiments. Comparing the shape of developed chambers in Figure 5-27, it can be realized that increasing the pressure resulted in a more V shape chamber at elevated pressure compared to the more extended chamber on the very top part of the reservoir model in low pressure case. Upon this observation, one may conclude that ratio of the free fall gravity drainage rate, which depends on viscosity of dead oil in place, to the diluted oil drainage rate, which depends on thermo-
physical properties of solvent/oil system, can affect shape of vapour chamber in Vapex experiments with mobile virgin oil in place.

The same explanation can be made when simulated solvent chamber in experiments #3 and #6 are compared in Figure 5-27. Both experiments #3 and #6 were conducted at 50 °C while propane injection pressure was increased from 817 kPa in earlier experiment to 1550 kPa in the later one.

**Figure 5-26:** Comparison of Vapour Chamber in Warm Vapex Experiments at T=40°, with Low and High Injection Pressures after 25 hours
Figure 5-27: Comparison of Vapour Chamber in Warm Vapex experiments at T=50°, with Low and High Injection Pressures after 25 hours
The numerical simulation study of Vapex performance reveals that accurate information on thermo physical properties of oil-solvent system is crucial to achieve reasonable history matches with the commercial simulators. The numerical simulation models of Vapex are still at a stage where one can obtain a reasonable history match by tuning several parameters but there is no assurance that the tuned parameters are unique and reliable for making accurate predictions under vastly changed conditions. Therefore, there is a need for reliable experimental data and field observations.
Chapter Six: Conclusion

In the preceding chapters experimental and numerical study of a Vapex process at elevated temperatures was presented. In this chapter the main conclusions distilled from the preceding studies are summarized. Some recommendations for continuation of this research are also provided.

6.1 Summary and Conclusion

• Effect of temperature on Vapex performance was evaluated by conducting numbers of warm Vapex experiments in sand-packed physical models, saturated with Athabasca bitumen, over a temperature range of 22 to 60 °C and injecting vapour propane within pressure range of 817 to 1550 kPa.

• Elevating the reservoir temperature (22 to 60 °) at the same injection pressure of propane (817 kPa) appeared to accelerate oil drainage rate at the early stages although the increasing temperature caused less dissolution of propane in bitumen. Increasing the temperature to 50 °C without increasing the injection pressure, improved production rate by 70% and heating to 60 °C was required to achieve more impressive rate increase of nearly 200%.

• In reservoirs containing similar type of oil, dramatic improvement in Vapex performance by injecting superheated solvent vapors would require going to temperatures higher than 60 °C.

• Production results of conducted Vapex experiments in pre-heated physical models suggest that production profiles of Vapex in naturally warm reservoirs with original mobile oil in place differs from the classic Vapex profiles due to contribution of free fall gravity drainage mechanism to oil production.

• Increasing propane injection pressure is able to improve Vapex performance by causing additional viscosity reduction due to more solvent dissolution in bitumen. Oil drainage rate appeared to remain high for a longer time at higher pressure compared to lower injection pressure. In fact at elevated temperatures but low injection pressures, rate of oil
production starts declining faster due to limited solvent dissolution to mobilize the oil and faster spreading of the solvent vapor chamber near the top.

- The upper limit of oil recovery rate that can be achieved in N-Solv and warm Vapex processes was estimated.
- Contribution of free fall gravity drainage mechanism in total recovery during warm Vapex experiment at T=50 °C was quantified by conducting a free fall gravity drainage test using nitrogen under the same operating condition as that of the conducted warm Vapex experiment.
- A simulation model verified by the experimental results of the conducted free fall gravity test was employed to quantify free fall gravity drainage rate at other temperatures.
- Contribution of mass transfer and heat transfer mechanisms on oil mobilization and production in performed warm Vapex experiments was evaluated.
- Viscosity behavior of propane diluted Athabasca Bitumen within the operating pressure and temperature of corresponding Vapex experiments was modeled based on the available experimental data in the literature. The viscosity-temperature-solubility contour map of Athabasca bitumen-propane system was generated.
- Effect of viscosity of solvent-diluted bitumen at the solvent interface as well as effect of viscosity of dead oil in place on diffusion layer thickness and oil drainage was examined. Equilibrium viscosity value was found not to be an accurate indicator of the actual drainage rate in Vapex process and a weighted average viscosity within diffusion layer which depends on both dead oil and live oil viscosity was found to be the key parameter.
- Experimental oil productions of performed experiments were compared against predicted rates from Vapex analytical models in the literature (Butler and Mokrys (1989), Heidari (2008)) and N-Solv empirical correlation (Nenninger and Dunn (2009)).
- The N-Solv correlation from Nenninger and Dunn (2009) was modified based on the findings of the current experimental study by incorporating the effect of live oil viscosity into the original correlation. The newly proposed correlation was tested against series of the available experimental data from earlier Vapex and warm Vapex experiments in the literature.
• Mackay River Athabasca bitumen was characterized into four pseudo components based on the SimDist and true boiling point curves to build the reservoir fluid model for simulation study.

• Solubility, saturation pressure and density of propane-bitumen mixture were modeled by tuning the Peng-Robinson equation of state parameters against reported experimental data for Athabasca bitumen/propane system in the literature.

• Viscosity of propane diluted Mackay River Athabasca bitumen was modeled by adjusting the coefficients of modified Pedersen viscosity correlation against experimental data of a similar bitumen/propane system reported in the literature within the pressure and temperature ranges of current study.

• Production results of conducted warm Vapex experiments were history matched using CMG-GEM compositional simulator. Combination of parameters including relative permeability, dispersivity and viscosity parameters, which was able to match the experiments was searched, discussed and presented.

• The link between Vapex oil production rate and thermo physical properties of bitumen/propane system at elevated temperatures was addressed through the simulation study.

• Effect of free fall gravity drainage mechanism on Vapex performance which was observed in the conducted experiments was confirmed and addressed through the simulation study.

• The developed simulation model verified by the experimental results of performed warm Vapex experiments within temperature range of 22 to 60 °C was used to predict production rates up to 90 °C.

• The verified simulation model was successful in predicting the rates and vapour chamber growth of conducted Vapex experiment in visible physical model with minor adjustment to relative permeability parameters.

• Effect of dead oil and live oil viscosities on shape of the developed vapor chamber during Vapex process was visualized and discussed through the simulation study.
6.2 Recommendation

• The current experimental results suggest that viscosity of live oil at the solvent-oil interface along with the viscosity of dead oil in place affects penetration depth of the solvent in oil during Vapex process. Conducting numbers of Vapex experiments with different types of oil and solvents over wider operating ranges of pressure and temperature can reveal the impact of live oil/dead oil viscosities on penetration thickness and oil drainage rates involved in Vapex process. Experimental oil drainage rates versus widely distributed viscosity data can be used to improve the capability of the proposed empirical correlation to predict production rates of Vapex and N-Solv.

• An extensive numerical study with the focus on mass transfer aspect of warm Vapex under different operating conditions can be conducted to evaluate effect of dead oil/live oil viscosities on penetration thickness and oil drainage rates provided valid PVT data of reservoir fluid and injected solvent system are available.

• Economic optimum operating pressure and temperature of N-Solv and warm Vapex can be searched and correspondingly economic proficiencies can be assessed against that of thermal based solvent-heat hybrid recovery methods including ES-SAGD.
References


