Investigation of Interplay of Capillarity, Drainage Height, and Aqueous Phase Saturation on Mass Transfer Phenomena in Heavy Oil Recovery by Vapex Process

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Farid Ahmadloo, candidate for the degree of Doctor of Philosophy in Petroleum Systems Engineering, has presented a thesis titled, *Investigation of Interplay of Capillarity, Drainage Height, and Aqueous Phase Saturation on Mass Transfer Phenomena in Heavy Oil Recovery by Vapex Process*, in an oral examination held on July 18, 2012. The following committee members have found the thesis acceptable in form and content, and that the candidate demonstrated satisfactory knowledge of the subject material.

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ABSTRACT

The vapor extraction (Vapex) process has emerged as a promising recovery technique in low-pressure and shallow heavy oil and bitumen reservoirs where conventional recovery processes are not technically, environmentally, or economically feasible. The Vapex process utilizes horizontal wells for injection of a vaporized hydrocarbon solvent (e.g., propane and butane) at pressures close to their dew point pressures into the reservoir, dissolution of solvent in the oil, and production of in-situ upgraded oil.

The realistic approximation of diffusion and convective dispersion occurring on the edge of the vapor chamber is required for reliable prediction of production rates in this process.

This research was aimed at investigating the interplay of capillarity, drainage height, and aqueous phase saturation on efficiency of mass transfer rate in permeability range similar to those found in western Canadian heavy oil and bitumen reservoirs. For this purpose, a combination of experimental and numerical studies was conducted in this research.

Experimental study of this research included the design and manufacturing of a new experimental set up and approach for investigating the mass transfer phenomenon at the edge of vapor chamber realistically. The rock and fluid characteristics were investigated by measuring the capillary pressure and pore size distribution of the sand packs, as well as phase behavior measurements to collect adequate PVT data for analytical and numerical modeling of conducted experiments.

Numerical simulation of the laboratory experiments were carried out using a commercial simulator (i.e. CMG-STARS™) in order to history match the experimental results and determine the effect of studied parameters on mass transfer rate.
Conducted analytical and numerical modelings showed that the effective diffusion coefficient was in the range of $4.91 \times 10^{-8}$-$1.10 \times 10^{-5}$ cm$^2$/s. The effective diffusion of the butane vapor into the heavy oil saturated sand pack in absence of immobile water saturation was in the range of $4.91 \times 10^{-8}$-$7.89 \times 10^{-6}$ cm$^2$/s. In presence of immobile of aqueous phase saturation, the effective diffusion coefficient varied between $4.91 \times 10^{-8}$ and $7.89 \times 10^{-6}$ cm$^2$/s. The comparison between the calculated effective diffusion coefficients and reported molecular diffusion in literature by different researchers confirms that the velocity-dependent term in convective dispersion plays only a minor role at higher capillarities, i.e. lower permeability media. This means molecular diffusion becomes the major driver for mass transfer phenomena under these conditions. This study showed that the selection of a realistic mass transfer coefficient is required for the simulation of the performance of the Vapex process. On this basis, the proposed effective diffusion/dispersion coefficients up to 4 orders of magnitude higher that the molecular diffusion coefficients in the literature could not be used for simulation of the Vapex process at reservoir condition similar to western Canadian heavy oil reservoirs.
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<table>
<thead>
<tr>
<th>Chapter</th>
<th>Title</th>
<th>Pages</th>
</tr>
</thead>
<tbody>
<tr>
<td>ABSTRACT</td>
<td></td>
<td>II</td>
</tr>
<tr>
<td>ACKNOWLEDGEMENT</td>
<td></td>
<td>IV</td>
</tr>
<tr>
<td>NOMENCLATURE</td>
<td></td>
<td>XIV</td>
</tr>
<tr>
<td>CHAPTER 1: INTRODUCTION</td>
<td></td>
<td>1</td>
</tr>
<tr>
<td>1.1. OIL RESOURCES</td>
<td>1</td>
<td></td>
</tr>
<tr>
<td>1.2. IN-SITU RECOVERY TECHNIQUES OF OIL SANDS</td>
<td>2</td>
<td></td>
</tr>
<tr>
<td>1.2.1. Cyclic Steam Injection</td>
<td>3</td>
<td></td>
</tr>
<tr>
<td>1.2.2. Steam Assisted Gravity Drainage (SAGD)</td>
<td>4</td>
<td></td>
</tr>
<tr>
<td>1.2.3. In-Situ Combustion</td>
<td>5</td>
<td></td>
</tr>
<tr>
<td>1.2.4. Vapor Extraction Process (Vapex)</td>
<td>6</td>
<td></td>
</tr>
<tr>
<td>1.3. RESEARCH OBJECTIVES</td>
<td>9</td>
<td></td>
</tr>
<tr>
<td>1.4. STRUCTURE OF THE DISSERTATION</td>
<td>10</td>
<td></td>
</tr>
<tr>
<td>CHAPTER 2: BACKGROUND AND LITERATURE REVIEW</td>
<td>13</td>
<td></td>
</tr>
<tr>
<td>2.1. VAPEX PROCESS</td>
<td>13</td>
<td></td>
</tr>
<tr>
<td>2.2. SOLVENT SELECTION IN THE VAPEX PROCESS</td>
<td>15</td>
<td></td>
</tr>
<tr>
<td>2.3. HEAVY OIL AND BITUMEN VISCOSITY</td>
<td>18</td>
<td></td>
</tr>
<tr>
<td>2.4. EFFECT OF POROUS MEDIA CHARACTERISTICS ON THE VAPEX PROCESS</td>
<td>23</td>
<td></td>
</tr>
<tr>
<td>2.4.1. Permeability and Heterogeneity</td>
<td>23</td>
<td></td>
</tr>
<tr>
<td>2.4.2. Drainage Height</td>
<td>27</td>
<td></td>
</tr>
<tr>
<td>2.4.3. Water Saturation</td>
<td>28</td>
<td></td>
</tr>
<tr>
<td>2.4.4. Capillarity</td>
<td>29</td>
<td></td>
</tr>
<tr>
<td>2.4.5. Presence of Aquifer and Gas Cap</td>
<td>31</td>
<td></td>
</tr>
<tr>
<td>2.5. OIL RATE EQUATION</td>
<td>33</td>
<td></td>
</tr>
<tr>
<td>2.6. MASS TRANSFER PHENOMENON IN THE VAPEX PROCESS</td>
<td>38</td>
<td></td>
</tr>
<tr>
<td>2.6.1. Diffusion Process</td>
<td>39</td>
<td></td>
</tr>
<tr>
<td>2.6.2. Dispersion process</td>
<td>41</td>
<td></td>
</tr>
<tr>
<td>2.7. SCALE-UP FACTORS</td>
<td>49</td>
<td></td>
</tr>
<tr>
<td>CHAPTER 3: EXPERIMENTAL SETUPS &amp; PROCEDURES</td>
<td>53</td>
<td></td>
</tr>
<tr>
<td>3.1. PHASE BEHAVIOR STUDY</td>
<td>53</td>
<td></td>
</tr>
<tr>
<td>3.1.1. Molecular Weight</td>
<td>53</td>
<td></td>
</tr>
<tr>
<td>3.1.2. SARA Analysis</td>
<td>55</td>
<td></td>
</tr>
<tr>
<td>3.1.3. PVT Apparatus</td>
<td>56</td>
<td></td>
</tr>
<tr>
<td>3.2. POROUS MEDIA CHARACTERIZATION</td>
<td>58</td>
<td></td>
</tr>
<tr>
<td>3.2.1. Sand Pack Selection</td>
<td>58</td>
<td></td>
</tr>
<tr>
<td>3.2.2. Capillary Pressure</td>
<td>61</td>
<td></td>
</tr>
<tr>
<td>3.3. VAPEX EXPERIMENTS</td>
<td>69</td>
<td></td>
</tr>
<tr>
<td>3.3.1. Design of Physical Models</td>
<td>69</td>
<td></td>
</tr>
<tr>
<td>3.3.2. Experimental Procedure</td>
<td>71</td>
<td></td>
</tr>
<tr>
<td>3.3.3. Interface Tracking</td>
<td>74</td>
<td></td>
</tr>
</tbody>
</table>
CHAPTER 4: EXPERIMENTAL RESULTS .................................................................................. 75
4.1. PHASE BEHAVIOR EXPERIMENTS .................................................................................. 75
  4.1.1. Molecular Weight Measurement ............................................................................... 75
  4.1.2. SARA Analysis ........................................................................................................ 76
  4.1.3. PVT Experiments ..................................................................................................... 78
4.2. VAPEX EXPERIMENTS .................................................................................................. 83
  4.2.1. Results of Vapex Experimental Program ................................................................. 84
  4.2.2. Solvent Mass Balance .............................................................................................. 91
  4.2.3. Interface Advancement ............................................................................................ 95
4.3. DATA ANALYSIS ............................................................................................................. 100
  4.3.1. Effect of Permeability ............................................................................................. 100
  4.3.2. Effect of Drainage Height ....................................................................................... 103
  4.3.3. Effect of Capillarity .................................................................................................. 105
  4.3.4. Vapex Number ........................................................................................................ 107
  4.3.5. Solvent Penetration Depth ...................................................................................... 109

CHAPTER 5: ANALYTICAL & SEMI-ANALYTICAL MODELING STUDY OF EXPERIMENTAL PROGRAM .................................................................................................................. 112
5.1. PHASE BEHAVIOR MODELING ...................................................................................... 112
  5.1.1. PVT Model .............................................................................................................. 112
  5.1.2. VISCOSITY MODELS .............................................................................................. 116
5.2. ANALYTICAL MODELING OF THE VAPEX EXPERIMENTS .............................................. 121
  5.2.1. VAPEX EXPERIMENTS IN ABSENCE OF IMMOBILE WATER SATURATION ........... 123
  5.2.2. VAPEX EXPERIMENTS IN PRESENCE OF IMMOBILE WATER SATURATION .......... 127
5.3. RESULTS AND DISCUSSION .......................................................................................... 132

CHAPTER 6: NUMERICAL SIMULATION OF MASS TRANSFER PHENOMENA IN MIXING ZONE IN THE VAPEX PROCESS .................................................................................. 139
6.1. BOUNDARY AND INITIAL CONDITIONS OF SIMULATED CONTROL VOLUME ............. 139
6.2. IMPLEMENTATION OF BOUNDARY CONDITIONS .......................................................... 141
  6.2.1. Gridding and Model Definition .................................................................................. 142
  6.2.2. PVT and Viscosity Model ........................................................................................ 144
  6.2.3. Rock-Fluid Properties ............................................................................................. 144
  6.2.4. Diffusion and Dispersion Modeling in The Vapex Process ....................................... 147
  6.2.5. Simulation Constraints ............................................................................................ 150
  6.2.6. Film Flow in a Vertical Plane .................................................................................. 152
6.3. HISTORY MATCHING STUDY ....................................................................................... 154
6.4. NUMERICAL DISPERSION ............................................................................................. 162

CHAPTER 7: CONCLUSIONS AND FUTURE RESEARCH ......................................................... 165
7.1. CONCLUSIONS .............................................................................................................. 165
7.2. FUTURE WORKS ......................................................................................................... 168

REFERENCES ...................................................................................................................... 170
LIST OF TABLES

Table 1.1: Liquid petroleum classification (Schramm et al. 2010) ........................................ 1
Table 1.2: Canadian liquid petroleum reserves (Schramm et al. 2010) ............................ 2
Table 2.1: Comparison of correlations for estimation of liquid diffusivities (after Boustani and Maini 2001) ........................................................................................................ 41
Table 2.2: Summary of experimental data on diffusivity of gases in Athabasca bitumen 42
Table 3.1: Operational range of the PVT apparatus .......................................................... 59
Table 3.2: Average characteristics of sand packs used in this study ................................. 61
Table 3.3: Mercury injection test Sample#1 .................................................................... 64
Table 3.4: Mercury injection test Sample#2 .................................................................... 65
Table 3.5: Mercury injection test Sample#3 .................................................................... 65
Table 3.6: Dimensional characteristics of the physical models ........................................ 69
Table 4.1: SARA fractions of Plover Lake oil ................................................................. 77
Table 4.2: Properties of Plover Lake oil .......................................................................... 79
Table 4.3: Plover Lake/n-butane system at 21ºC and 200kPa ........................................... 81
Table 4.4: Plover Lake oil/n-butane (38.90 mol %) ........................................................ 81
Table 4.5: Properties of conducted Vapex experiments .................................................. 83
Table 4.6: Interface advancement rate in Sand#1 ............................................................ 96
Table 4.7: Interface advancement rate in Sand#2 ............................................................ 97
Table 4.8: Interface advancement rate in Sand#3 ............................................................ 97
Table 5.1: Error analysis of viscosity predictions of oil-butane mixtures by different approaches ......................................................................................................................... 120
Table 5.2: Back-calculated Vapex numbers and characteristics of Vapex experiments in absence of immobile water saturation ................................................................. 124
Table 5.3: Calculated $\alpha$ values in absence of immobile water saturation .................... 126
Table 5.4: Back-calculated Vapex numbers in presence of immobile water .................... 130
Table 5.5: Calculated $\alpha$ values in presence of immobile water saturation .................... 131
Table 6.1: Boundary and initial conditions of simulated control volume .......................... 142
Table 6.2: Interfacial tension and contact angles used for determination of capillary pressure curves from mercury injection data ................................................................. 145
Table 6.3: Summary of comparison between effective diffusion and calculated numerical dispersion based on proposed approach by Fanchi (1983) ............................................. 164
LIST OF FIGURES

Figure 1.1: Schematic of the Vapex process ................................................................. 8
Figure 2.1: The concept of the Vapex process ................................................................. 14
Figure 2.2: Concentration dependency of heavy oil and bitumen viscosity to butane concentration ......................................................................................................................... 20
Figure 2.3: Thermal dependency of the relative viscosity of maltenes recombined with various amounts of asphaltene (after Argillier et al., 2001) ................................................................. 22
Figure 2.4: Fluid flow and mass transfer take place in the boundary layer between the solvent vapor and immobile bitumen (after Das and Butler, 1998a) ................................................................. 34
Figure 2.5: Schematic of the vapor extraction in Hele-Shaw cell (after Das and Butler, 1998a) ............................................................................................................................... 36
Figure 2.6: Concentration dependency of viscosity and diffusion coefficient on normal butane concentration in Plover Lake Oil ............................................................................................................ 40
Figure 2.7: Effect of drainage height on back-calculated $N_{s,exp}$ values in reviewed studies (This study) ................................................................................................. 48
Figure 2.8: Effect of permeability on back-calculated $N_{s,exp}$ values in reviewed studies (This study) ................................................................................................. 49
Figure 2.9: Effect of permeability and solvent composition on back-calculated $N_{s,exp}$ values in reviewed studies (This study) ................................................................................................. 50
Figure 3.1: Schematic of vapor pressure osmometer setup (UIC Model 833 VPO) ........ 54
Figure 3.2: Schematic of PVT apparatus (after Freitag et al., 2005) ............................... 57
Figure 3.3: Grain size distribution of the selected sands in this study ......................... 60
Figure 3.4: Microscopic photographs of sand grains; (a) Sand#1, (b) Sand#2, and (c) Sand#3 ............................................................................................................................. 60
Figure 3.5: Pore size distribution Sand#1 ....................................................................... 66
Figure 3.6: Pore size distribution Sand#2 ....................................................................... 66
Figure 3.7: Pore size distribution Sand#3 ....................................................................... 67
Figure 3.8: Derived air-water capillary pressure Sand#1 ............................................... 67
Figure 3.9: Derived air-water capillary pressure Sand#2 ............................................... 68
Figure 3.10: Derived air-water capillary pressure Sand#3 ............................................ 68
Figure 3.11: Schematic of the physical models used for Vapex experiments; (a) front view, (b) side view of the physical model ................................................................. 71
Figure 3.12: Photographs of the fabricated physical models used for Vapex experiments; (a) solvent chamber (b) oil saturated porous media ........................................ 71
Figure 4.1: Comparison between the asphaltene content of dead oil and collected samples of produced oil ................................................................................................... 77
Figure 4.2: Carbon number distribution of Plover Lake oil ............................................. 79
Figure 4.3: Density variation of Plover Lake oil with pressure at 21 °C ............................ 80
Figure 4.4: Density dependency of Plover Lake oil to n-butane concentration ............ 82
Figure 4.5: Viscosity dependency of Plover Lake oil to n-butane concentration ............ 82
Figure 4.6: Oil drainage rate in Sand#1 at two drainage heights .................................... 85
Figure 4.7: Cumulative oil production in Sand#1 ......................................................... 85
Figure 4.8: Oil drainage rate in Sand#2 at two drainage heights .................................... 86
Figure 4.9: Cumulative oil production in Sand#2 ......................................................... 86
Figure 4.10: Oil drainage rate in Sand#3 at two drainage heights .................................. 87
Figure 4.11: Normalized cumulative oil production in Sand#3 ..................................... 87
Figure 4.12: Pressure and temperature profile in Experiments# 1 and 2 ..................... 88
Figure 4.13: Drainage rate comparison in absence of aqueous phase ......................... 90
Figure 4.14: Drainage rate comparison in absence and presence of aqueous phase ...... 90
Figure 4.15: Cumulative n-butane uptake in Sand#1 ................................................... 94
Figure 4.16: Cumulative n-butane uptake in Sand#2 ................................................... 94
Figure 4.17: Cumulative n-butane uptake in Sand#3 ................................................... 95
Figure 4.18: Oil-solvent interface position at different times in Sand#1 ....................... 98
Figure 4.19: Oil-solvent interface position at different times in Sand#2 ....................... 99
Figure 4.20: Oil-solvent interface position at different times in Sand#3 ....................... 99
Figure 4.21: Photographs of oil-solvent interface evolution in small physical model ... 100
Figure 4.22: Variation of stabilized drainage rate with permeability (S_w=0) ............... 102
Figure 4.23: Variation of stabilized drainage rate with permeability (S_w>0) ............... 102
Figure 4.24: Predicted stabilized drainage rates by Equation 2.40 (S_w=0.0) .............. 104
Figure 4.25: Predicted stabilized drainage rates by Equation 2.40 (S_w>0.0) .............. 104
Figure 4.26: Variation of stabilized drainage rate with specific pore surface area of porous medium .......................................................... 106
Figure 4.27: Effect of drainage height, capillarity, and connate water saturation on Vapex number.......................................................... 107
Figure 4.28: Interplay of permeability, drainage height, and water saturation on Vapex number in this study and reviewed studies ............................................. 108
Figure 4.29: Calculated n-butane penetration depth in Vapex experiments conducted in this study .............................................................. 111
Figure 5.1: Measured and modeled density of dead oil at studied pressures ........... 114
Figure 5.2: Measured and modeled density of oil/butane mixtures.......................... 115
Figure 5.3: Measured and modeled density of oil/butane mixture (C_4=38.90 mol%).... 115
Figure 5.4: Measured and modeled saturation pressures of oil-butane mixture ........ 116
Figure 5.5: Measured and modeled viscosity of dead oil ...................................... 119
Figure 5.6: Measured and modeled viscosity of oil-butane mixture (C_4=38.90 mol%). 119
Figure 5.7: Comparison between measured and predicted viscosity of oil-butane mixtures at different n-butane concentrations by different approaches ................................................. 120
Figure 5.8: Dependency of stabilized drainage rates on specific pore surface area of porous medium .................................................. 125
Figure 5.9: Dependency of \( \alpha \) parameter on drainage height and specific pore surface area of medium in absence of immobile water saturation ........................................ 126
Figure 5.10: Dependency of effective diffusion coefficient on investigated parameters in absence of water saturation ................................................. 128
Figure 5.11: Dependency of stabilized drainage rates with specific pore surface area of porous medium .................................................. 130
Figure 5.12: Dependency of \( \alpha \) parameter to drainage height, specific pore surface area of medium and oil-solvent mixture viscosity in presence of immobile water .......... 131
Figure 5.13: Dependency of effective diffusion coefficient on investigated parameters in presence of water saturation ......................................................... 132
Figure 5.14: Dependency of effective diffusion coefficient on drainage height, specific pore surface area, and immobile water saturation .............................................. 133
Figure 5.15: Comparison between the calculated effective diffusion coefficients and predicted effective diffusivity with proposed correlations for butane in literature ........ 134
Figure 5.16: Capillary dependency of the viscosity of produced oil samples in absence of immobile water saturation ............................................................................................... 135
Figure 5.17: Capillary dependency of the viscosity of produced oil samples in presence of immobile water saturation ............................................................................................... 136
Figure 5.18: Drainage rate dependency on immobile water saturation .................... 137
Figure 6.1: Control volume of the numerically simulated model ............................. 141
Figure 6.2: Relative permeability curves of Sand#3 in conducted simulation study ..... 146
Figure 6.3: Relative permeabilities in simulation (a) vapor chamber and (b) collection chamber .......................................................................................................................... 147
Figure 6.4: Comparison between the measured and predicted oil drainage rates in Experiment#5 in absence of immobile water saturation ................................................. 156
Figure 6.5: Comparison between the measured and predicted dissolved gas volume in Experiment#5 in absence of immobile water saturation ................................................. 156
Figure 6.6: Comparison between the measured and predicted oil drainage rates in Experiment#11 in presence of immobile water saturation ................................................. 157
Figure 6.7: Comparison between the measured and predicted dissolved gas volume in Experiment#11 in presence of immobile water saturation ................................................. 157
Figure 6.8: Oil Saturation change at two snapshots predicted by history matched model in Experiment#5: (a) 632 hours and (b) 1127 hours since start of the experiment ........ 158
Figure 6.9: Measured and predicted vapor-oil interface evolution rate in Experiment#5 ................................................................................................................................. 160
Figure 6.10: Measured and predicted vapor-oil interface evolution rate in Experiment#11 ................................................................................................................................. 161
Figure 6.11: Solvent-oil interface movement in Experiment#5 at four different times.. 161
NOMENCLATURE

$A$ area, cm$^2$

$A_p$ pore surface area, cm$^2$

$c$ solvent content, volume fraction

$C_s$ solvent content, mole fraction

$C_s$ solvent content, volume fraction

$D$ mutual diffusivity of solvent in bitumen, cm$^2$/s

$d_0$ Constant

$d_1$ Constant

$D_{\text{eff}}$ effective molecular diffusion coefficient, cm$^2$/s

$D_l$ total longitudinal dispersion coefficient, cm$^2$/s

$D_p$ pore diameter, cm

$d_p$ particle diameter, cm

$D_t$ total transverse dispersion coefficient, cm$^2$/s

$D_{ij}^{\text{phy}}$ $ij^{th}$ element of the dispersion tensor, cm$^2$/s

$f$ constant

$F$ formation resistivity factor, dimensionless

$f$ fractional flow, dimensionless

$g$ gravity acceleration, cm/s$^2$

$h$ drainage height, cm

$H$ height of interface at the no-flow boundary, cm

$k$ absolute permeability, cm$^2$

$K$ calibration factor, dimensionless

$k_{\text{ocw}}$ oil relative permeability at connate water saturation, fraction

$k_{\text{riro}}$ oil relative permeability at irreducible oil saturation, fraction

$k_{\text{gcl}}$ gas relative permeability at connate liquid saturation, fraction

$k_{\text{rogsg}}$ oil relative permeability at connate gas saturation, fraction

$L$ Length of horizontal well, cm

$MW$ molecular weight, g/gmole

$n$ constant
$N_s$ Vapex number, dimensionless

$N_{s,\text{exp}}$ experimental Vapex number, dimensionless

$N_w$ exponent for calculating the water isoperm from $k_{\text{rwi}}$

$N_{ow}$ exponent for calculating the oil isoperm in oil-water system from $k_{\text{rocw}}$

$N_{og}$ exponent for calculating the oil isoperm in oil-gas system from $k_{\text{rog}}$

$N_g$ exponent for calculating the gas isoperm from $k_{\text{rg}}$

$P$ pressure, mmHg

$P_c$ critical pressure, kPa

$P_{cl}$ capillary pressure, kPa

$PD$ pressure required for penetration into a cylindrical pore, kPa

$P_i$ partial pressure of the component $i$ in the mixture, kPa

$P_i^0$ vapor pressure of pure component $i$, kPa

$Q$ oil drainage rate per unit length of the production well, cm$^2$/s

$q$ oil production rate as per volume fraction per unit time, 1/s

$Q_d$ stabilized drainage rate ratios, dimensionless

$R$ gas constant, kJ/mol.K

$r_i$ pore entry radius, cm

$S$ saturation, fraction

$S_{\text{wcon}}$ connate water saturation, fraction

$S_{\text{wcrit}}$ critical water saturation, fraction

$S_j$ phase $j$ saturation, fraction

$S_{\text{oirw}}$ irreducible oil saturation for water-oil system, fraction

$S_{\text{orw}}$ residual oil saturation for water-oil system, fraction

$S_{\text{oirg}}$ irreducible oil saturation for gas-liquid system, fraction

$S_{\text{org}}$ residual oil saturation for gas-liquid system, fraction

$S_{\text{gcon}}$ connate gas saturation, fraction

$S_{\text{gcrir}}$ critical gas saturation, fraction

$S_m$ mercury saturation, volume fraction

$SOR$ solvent-to-oil ratio, cm$^3$/cm$^3$

$T$ temperature, ºC

$t$ time, s
$T_a$ absolute temperature, K

$T_c$ critical temperature, K

$U$ oil velocity in mixing zone, cm/s

$u$ velocity as a function of Hele-Shaw cell spacing, cm/s

$u$ Darcy velocity, cm/s

$v$ specific volume of the solvent liquid, cm$^3$/g

$V$ specific volume of the solvent, cm$^3$/g

$V_p$ pore volume, cm$^3$

$w$ variable distance within the Hele-Shaw cell spacing, cm

$x_i$ concentration of component $i$ in vapor phase, mole fraction

$y$ height at any point on the interface, cm

$i, j, k$ denote Cartesian coordinates with values ranging from 1 to N

$\Delta E$ activation energy, kJ

$\Delta H_v$ heat of vaporization, kJ/mol

$\Delta S_o$ change of oil saturation, dimensionless

$AV$ voltage change, $\mu$V

$\Delta p$ density difference between oil and solvent, g/cm$^3$

$\Delta t$ time step size, s

$x, y, z$ spatial coordinates

$\Delta x, \Delta y, \Delta z$ grid block lengths, cm

**Subscripts**

$cm$ mercury-air capillary pressure

$cw$ water-air capillary pressure

$f$ field

$L$ large

$m$ model

$max$ maximum

$min$ minimum

$mix$ solvent-oil mixture

$o$ heavy oil

$p$ porous media
$S$  small
$s$  solvent
$y$  at height $y$

**Greek Symbols**

$\sigma$  a measure of the inhomogeneity of the porous pack
$\varrho$  average interstitial velocity, cm/s
$\sigma_i$  interfacial tension, dyne/cm
$\phi$  porosity, dimensionless
$\alpha$  constant
$\beta$  constant
$\gamma$  constant
$\theta$  vapor chamber angle, degree
$\theta_i$  contact angle, degrees
$\kappa$  geometrical scaling factor
$\lambda$  fraction of solvent vaporized, cm$^3$/cm$^3$
$\mu$  solvent-oil mixture viscosity, g/cm.s
$\xi$  solvent penetration depth, cm
$\rho$  density, g/cm$^3$
$\Omega$  cementing factor, dimensionless
$\omega$  oil concentration, g/l
$\nabla$  gradient operator
$\kappa$  mechanical dispersivity, cm
CHAPTER 1: INTRODUCTION

1.1. Oil Resources

The world liquid petroleum (oil) resources are comprising over 600 billion m$^3$ (approximately 4 trillion barrels). Of this, approximately half, or 320 billion m$^3$, is represented by the bitumen contained in oil sands. The remainder is made up of conventional medium- and light-gravity crudes (Schramm et al. 2010). A practical approach for classification of these crudes is using the density and viscosity of these crudes as given in Table 1.1. According to this classification, the extra heavy crude oils and bitumen are essentially immobile at reservoir condition and require mobilization before exploitation. For instance, Athabasca bitumen has viscosity of 1 million mPa·s which is essentially immobile at reservoir condition.

Table 1.1: Liquid petroleum classification (Schramm et al. 2010)

<table>
<thead>
<tr>
<th>Petroleum Type</th>
<th>Density* (kg/m$^3$)</th>
<th>Viscosity* (mPa·s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Conventional light crude oil</td>
<td>&lt;934</td>
<td>&lt;100</td>
</tr>
<tr>
<td>Conventional heavy crude oil</td>
<td>890-1,000</td>
<td>100-10,000</td>
</tr>
<tr>
<td>Extra heavy crude oil</td>
<td>&gt;1,000</td>
<td>10,000-100,000</td>
</tr>
<tr>
<td>Bitumen</td>
<td>&gt;1,000</td>
<td>&gt;100,000</td>
</tr>
</tbody>
</table>

* At reservoir temperature

A significant fraction of the world’s conventional crude oil and most of the world’s bitumen is contained in the western Canadian sedimentary basin. Table 1.2 summarizes the Canadian liquid petroleum reserves (Schramm et al. 2010). According to the available statistics, bitumen comprises the majority of Canada’s liquid petroleum resources at
approximately 410 billion m³. These numbers have to be tempered with such factors as oil viscosity at reservoir temperature, availability of natural gas for upgrading, and proximity to the market.

Table 1.2: Canadian liquid petroleum reserves (Schramm et al. 2010)

<table>
<thead>
<tr>
<th>Petroleum Type</th>
<th>Resources (B bbl)</th>
<th>Reserves (B bbl)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Conventional light crude oil</td>
<td>175</td>
<td>2.7</td>
</tr>
<tr>
<td>Conventional heavy crude oil</td>
<td>50</td>
<td>1.4</td>
</tr>
<tr>
<td>Oil sands bitumen:</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mineable oil sands</td>
<td>200</td>
<td>31</td>
</tr>
<tr>
<td>Shallow in-situ oil sands</td>
<td>600</td>
<td>0</td>
</tr>
<tr>
<td>Deep in-situ oil sands</td>
<td>1,800</td>
<td>142</td>
</tr>
<tr>
<td>Total oil sands</td>
<td>2,600</td>
<td>173</td>
</tr>
</tbody>
</table>

Oil sands in Canada are an important part of energy supply of the world, and being exploited increasingly as the demand for petroleum increases. Currently the daily production of bitumen and synthetic crudes in Canada has risen to 180,000 m³ (1.1 million barrels) that represents approximately 40% of crude oil produced in Canada (CAPP 2007). According to estimations the Canadian bitumen production rate is going to triple by 2030, based on moderate growth case and taking into account the current economic slowdown (Schramm et al. 2010).

1.2. In-Situ Recovery Techniques of Oil Sands

There are various methods proposed for recovery of the heavy oil, however by far the more successful ones are based on injection of steam in some form. Other methods include in-situ combustion, electrical heating, geothermal heating, and solvent based
processes. These in-situ recovery techniques are feasible in deposits deeper than 75 m, which are too deep to mine. However, feasibility study of application of in-situ steam and solvent methods in depth range of 75-200 m showed that these recovery techniques are not suitable for extraction of oil sands in these deposits. The low efficiency of these recovery processes in these shallower deposits has been attributed to relatively low pressure and typically poor over/underlying confinement that make mobilization and extraction of the mobilized oil extremely challenging. Typical recovery factor of oil sands in depth range of 75-200 m is approximately 0%. However, in-situ recovery processes have shown promising results in deeper deposits (>200 m) and achieved typical recovery factors of 10% (Schramm et al. 2010).

1.2.1. Cyclic Steam Injection

One mature process for deep, thicker reserves, such as in Cold Lake and Peace River, involves cycling steam at single vertical injector/producer wells (this is sometimes referred to as "huff-and-puff"). In this process, a certain volume of steam is injected into a single well, over a period of several weeks, after which, following a short shut-in period, the well is put into production. The mobilized oil around the wellbore is driven into the wellbore by the existing reservoir drives, at first at a high rate, which declines fast. This process is repeated in several cycles, as long as the process is economic.

The cyclic steam injection has been extremely successful in heavy oil reservoirs, because of higher returns at the early stages of the production phase of this process. Typical steam-to-oil ratios, the major economic factor for steam-based processes, are 3:1 to 4:1 (Alberta Chamber of Resources, 2004).
Cyclic steam stimulation (often called CSS) has gained a considerable attention in Cold Lake, where the oil viscosity is around 100,000 mPa·s, and initial steam injectivity is extremely low. In this area, steam is typically injected at pressures higher than parting pressure of the formation (>10 MPa) to create vertical fractures by simultaneous injection of steam into many wells (e.g., 18 wells) at a time. Through this operation, the wells are linked and the subsequent operation proceeds in a manner akin to a sweep across the area by the hot zone (Vittoratos et al. 1990).

In Peace river oil sand deposits, a range of thermal recovery technologies have been tested over the years to overcome the difficulties associated with the a bottom high water saturation zone. Current operations use the multilateral horizontal CSS technique (Schramm et al. 2010). Despite the complexities related to steam injection interval and operational challenges, the production responses have been satisfactory (Farouq Ali, 2003).

1.2.2. Steam Assisted Gravity Drainage (SAGD)

Despite successful experience of steamflooding in other countries (e.g., Brazil, Indonesia), steam flooding in Cold Lake has not been commercially successful. This was attributed to the presence of fractures and communication pathways resulting from the preceding CCS or cold production techniques (Farouq Ali, 2003).

Advanced horizontal drilling technology has laid the foundation for steam assisted gravity drainage (SAGD) for application in reservoirs that would normally fail the so-called screening criteria developed based on previous industry experience on thermal processes. SAGD process was originally developed by Butler et al. (1981). In this process, typically two horizontal wells of approximately 800-1000 m are drilled within
approximately 5 m of each other in the same vertical plane. Steam is injected from the top well which rises to form a steam chamber with bitumen and condensate draining into the production well by gravity. Various well configurations have been considered to improve the efficiency of the process in presence of gas cap and aquifer (Jespersen and Fontain, 1993).

The SAGD process was first successfully tested in the Athabasca tar sands, with bitumen viscosity of millions of centipoises at reservoir condition. SAGD makes it possible to extract bitumen from thinner reserves, however good vertical permeability is essential. The Steam-to-oil ratio for SAGD ranges from 2.5:1, for high quality reservoirs, to 3.0:1 (National Energy Board, 2004). EnCana for example, claim they have achieved an industry leading steam-to-oil ratio, requiring only 2.5:1, and the company’s longer-term objective is further improvement to about 2:1 (EnCana, 2004).

In order to mitigate the environmental impact of the SAGD process (e.g., CO₂ emission and natural gas and water use), processes involving the addition of solvent to steam, often referred to as Hybrid thermal solvent processes are being developed. These processes are undergoing field-testing in the Athabasca and Cold Lake deposits. Examples of the processes with solvent additives include ES-SAGD (expanding solvent SAGD), SAP (solvent added process), and solvent (or liquid) addition to steam (Nasr and Isaacs, 2001; Gupta and Gittins, 2007; Leaute and Carey, 2007; Frauenfeld et al. 1997).

\textbf{1.2.3. In-Situ Combustion}

In-situ combustion is an oil recovery process in which heat is generated in the reservoir by reaction between oxygen and either original or modified fractions of crude oil. This process is initiated by heating the oil-bearing sand around an injection well with a gas
burner, and electric heater or by the injection of a hot fluid such as steam. An oxygen-containing gas is injected to ignite the oil and to propagate the combustion zone outward from the injector. The elevated temperature zone displaces reservoir fluids, which are advanced towards the production wells by mechanisms, such as vaporization, and distillation, steam displacement, hot and cold water displacement and gas flooding.

Over 200 in-situ combustion field tests and commercial operations have been carried out worldwide (Farouq Ali, 2003). According to available history of in-situ combustion in Canada, this process can be successfully applied under specific conditions. Despite high number of failures in implementation of this process, there are a few successful in-situ combustion projects in western Canada such as Battrum, Fosterton Northwest, Golden Lake Sparky, Marguerite Lake, Countess ‘B’, and Morgan project (Moore et al. 1995). Moore et al. (1995) have blamed the operating procedures used in many of unsuccessful in-situ combustion projects in Canada as the reason behind these failures. Miller et al. (2002) suggested the application of in-situ combustion for thin payzones in western Canadian oil sands after the cold production process. These authors believe that the regional heavy oil sands is likely too thin for follow up steam processes, and solvent-based processes may be impeded by solvent losses and slow mass transfer rates. On the whole, in situ combustion has been more successful in relatively mobile heavy oils than in very viscous oils, occurring in homogeneous formations.

1.2.4. Vapor Extraction Process (Vapex)

The vapor extraction process (i.e., Vapex) has emerged as a potentially new recovery technique in low-pressure and shallow heavy oil and bitumen reservoirs where conventional and/or thermal recovery processes are not technically, environmentally, or
The Vapex process is economically feasible. The Vapex process is an analogue to the steam assisted gravity drainage (SAGD) process. It utilizes pair of horizontal wells for injection of a vaporized hydrocarbon solvent (e.g., propane and butane) at pressures close to their dew point pressures into the reservoir, dissolution of solvent in the oil, and production of in-situ upgraded oil in the lower horizontal well.

The use of vaporized solvent produces a higher driving force for gravity drainage (as a consequence of a higher density difference between heavy oil and vaporized solvent), and it ensures a lower residual amount of solvent in the extracted reservoir.

In this process, the injected solvent in vapor phase slowly rises to form a vapor chamber, dissolving the heavy oil at the interface and diffusing into the bulk oil. The diluted oil which is mobilized as a result of dissolution of the solvent will drain under the effect of gravity in a thin layer near the production well. The efficiency of the Vapex process is controlled by the rate of mass transfer (i.e., molecular diffusion and convective dispersion) occurring between the solvent and oil at the edge of the vapor chamber. In the Vapex process, gravity is the only driving force for flow of mobilized oil as a result of dissolution of solvent in the oil phase, hence no forced displacement of oil phase occurs after establishing the connection between the injection and production wells. Schematic of the Vapex process is shown in Figure 1-1.

The Vapex process has a number of advantages over other heavy oil and bitumen recovery processes. The primary advantage of this process is the low capital costs, as the injection of vaporized solvents does not require extensive surface facilities, such as steam injection processes that require water treatment and heating facilities. Since solvent injection does not involve the energy losses that are characteristics of thermal recovery
processes, Vapex is relatively energy efficient. For a similar production rate, Vapex uses approximately 3% of the energy consumed by SAGD (Singhal et al. 1996). The application of solvent in the vapor state minimizes the solvent requirement in the implementation of the process and maximizes the driving forces for the displacement of the mobilized oil. Laboratory tests have shown that up to 90% of the injected solvent can be retrieved and recycled in this process (Singhal et al. 1996). The potential deasphalting that occurs due to dissolution of solvent into the oil phase can result in in-situ upgrading of the produced oil in this process. Moreover, implementation of the Vapex process can significantly reduce the green house emission. This process can also use CO₂ as the carrier gas to adjust the dew-point of the injected solvent according to the reservoir pressure and temperature. Therefore, this process can also sequester CO₂ into the reservoirs and reduce the CO₂ emissions (Talbi and Maini, 2003).

Figure 1.1: Schematic of the Vapex process
1.3. Research Objectives

It is known that the performance of the Vapex process is strongly dependent on the efficiency of the mass transfer rate between the solvent and heavy oil or bitumen. Therefore, realistic approximation of diffusion and convective dispersion occurring on the edge of the vapor chamber is required for reliable prediction of production rates in this process. Studies conducted at extremely high permeabilities (>200 D) have reported up to four orders of magnitude increase in the diffusion/dispersion rates between the solvent and heavy oil. Such increases could result in unrealistic overestimation of the performance of this process at real reservoir conditions (Kapadia et al. 2006). In order to address this shortcoming, this research was proposed to systematically investigate the interplay of capillarity, drainage height, and aqueous phase saturation on the efficiency of mass transfer rate in a permeability range similar to western Canadian heavy oil and bitumen reservoirs.

Based on a comprehensive literature review, in order to address some of the shortcomings of the available Vapex studies, the objectives of this study were defined as follows:

- To design and build a new experimental setup and develop a new operating procedure for investigating the mass transfer phenomenon in the Vapex process
- To conduct an experimental program to investigate the effects of drainage height, capillarity effect, and immobile water saturation on mass transfer rate in the Vapex Process
- To analytically and numerically model, simulate and history match the Vapex laboratory experiments in order to determine the range of molecular diffusion and
Convective dispersion occurring on the edge of the vapor chamber in these experiments

- To investigate the effect of each one of above studied parameters (i.e., drainage height, capillarity, and water saturation) on the value of effective diffusion/dispersion coefficient obtained from the history matching of the conducted Vapex experiments
- To develop a correlation for the estimation of effective diffusion/dispersion coefficient based on characteristics of solvent-oil system and porous media.

1.4. Structure of the Dissertation

This dissertation is organized in seven chapters. Chapter 1 provides a general description of heavy oil recovery processes, amount of oil sand resources in Canada, and research objectives. Chapter 1 also includes the organization of the dissertation and a brief description of remaining chapters.

Chapter 2 presents the conducted literature review. This chapter gives an overview on wide range of experimental and theoretical studies conducted on various aspects of the Vapex process since 1989. This chapter includes details on active recovery mechanisms, effect of porous media characteristics, diffusion and dispersion phenomena, scaling-up, and analytical and numerical modeling of the Vapex process.

Chapter 3 provides the details on experimental setups and procedures developed in this research. The experimental program in this section can be divided into three sections. This first section includes the details of experiments designed to measure the phase behavior properties of various oil-solvent mixtures. These measurements included: molecular weight, SARA analysis, and PVT measurements. The second series of
experiments included the measurements for characterization of porous media (e.g., sieve analysis, porosity and permeability, capillary pressure, and pore size distribution measurements of unconsolidated sands). The third section provides the details of experimental setup and experimental procedure utilized for conducting the designed Vapex experiments.

Chapter 4 presents the results of the conducted experimental study. The first section of this chapter provides the results of the phase behavior measurements on Plover Lake oil/n-butane mixtures. The second part of this chapter presents the results of conducted Vapex experiments. These Vapex experiments were conducted in three different sand packs at two drainage heights to investigate the effects of drainage height, immobile water saturation and capillarity on the performance of the Vapex process and the rate of mass transfer in the studied process.

Chapter 5 presents the conducted analytical and semi-analytical modeling study of experimental program. The first part of this chapter consists of semi-analytical modeling of the phase behavior for Plover Lake/n-butane system. This part of study was completed using CMG’s Winprop package. The second part of this chapter presents the analytical modeling to determine the effective diffusion/dispersion coefficients based on the Vapex experiments.

Chapter 6 presents the results of the numerical study to history match the conducted Vapex experiments. This chapter discusses various aspects of numerical simulation of the Vapex process. In parallel with the conducted analytical modeling, the history matched models are used to evaluate the role of various parameters on the molecular diffusion and convective dispersion.
Chapter 7 summarizes the contribution of this research on understanding the interplay of studied parameters on the rate of mass transfer in the Vapex process. This chapter also includes the areas for future work and recommendations.
CHAPTER 2: BACKGROUND AND LITERATURE REVIEW

2.1. Vapex Process

Vapex is a solvent-leaching gravity-drainage-based process that has emerged as a promising recovery technique in low-pressure heavy oil and bitumen reservoirs. In addition, this process is suitable for reservoirs with high water content, low thermal conductivity of formation, and active aquifer where conventional thermal recovery techniques are not feasible. Lower completion cost, higher energy efficiency, in-situ upgrading of the produced oil, and lower energy costs are considered as the main advantages of this process (Butler and Mokrys, 1991; Butler and Yee, 2002; Luhning et al., 2003; Roopa and Daw, 2007; Vargas-Vasquez and Romero-Zerón, 2007).

Vapex process utilizes horizontal wells for injection of vaporized hydrocarbon solvent (e.g., propane and butane) at pressures close to their dew point pressures at reservoir temperature, dissolution of the solvent into oil, and production of in-situ deasphalted oil. Butler and Mokrys (1989) introduced the concept of the Vapex process by investigating the efficiency of this process using Tangleflags North oil from Lloydminster area and Suncor coker feed bitumen. In this pioneering study, liquid toluene was used as the solvent in a line source Hele-Shaw cell. Since then, extensive experimental and numerical studies have been conducted to investigate various aspects of this and related solvent vapor extraction processes. In addition to these studies, there are reported cases of field trials of solvent aided processes in western Canada (Miller et al., 2003; Gupta and Gittins, 2005).

In conceptual design of the Vapex process, the injected solvent vapor forms a vapor chamber in the reservoir where the reservoir oil comes into contact with solvent at the
edge of solvent chamber. Upon dissolution of the solvent in the oil phase, a displacement front is formed by dilution, swelling, and deasphalting mechanisms that reduces adhesive forces between oil globules, films, and connate water (Beecher and Parkhurst, 1926; Welker and Dunlop, 1963; Simon, 1965; Das, 1995). The diluted oil drains under gravity toward the production well at the bottom of the vapor chamber. It is postulated that molecular diffusion; convective dispersion due to induced flow, gravity, density variation; and surface tension and capillary action in the porous media result in further mixing of the solvent and the oil phase (Das, 1995). This mixing is pronounced in the mixing zone surrounding the growing vapor chamber. Schematic of this process is illustrated in Figure 2.1. This figure shows the drainage paths forming on the edge of the solvent chamber at different stages of chamber expansion through the operating life of a Vapex process.

![Figure 2.1: The concept of the Vapex process](image-url)
2.2. Solvent Selection in the Vapex Process

Conducted experimental studies have shown that light alkanes are the most suitable solvents for the Vapex process. These studies have revealed that higher drainage rates and recovery factors were achieved when propane, butane, or mixture of propane-butane were utilized as solvent (Mokrys and Butler, 1991, 1993; Das and Butler, 1995; Talbi and Maini, 2008). Lim et al. (1994) provided evidence on successful application of ethane in cyclic solvent stimulation process for extraction of Cold Lake bitumen.

Das and Butler (1994a) conducted an experimental study using Peace River, Cold Lake, and Lloydminster bitumen samples in temperature range of 20-35 ºC in a Hele-Shaw cell. This study revealed deasphalting process occurs when the propane vapor was injected very close to the propane’s dew point pressure. They observed that the deasphalting process did not result in any permeability impairment. On the contrary, some enhancement in the drainage rates was observed that were attributed to viscosity reduction that occurs as a result of deasphalting process. Conducted experiments by Mokrys and Butler (1993) using liquid propane (i.e., 846-915 kPa and 20 ºC) showed a massive asphaltene deposition causing sever reduction in observed drainage rates. Qualitative experiments in a Hele-Shaw cell and packed beds showed that deasphalting process and in-situ upgrading occurred when butane was used as solvent instead of propane. In fact, butane vapor close to its dew point pressure did not cause deasphalting of Peace River bitumen at 25 ºC and it only caused smaller amount of asphaltene deposition of Lloydminster bitumen at 20 ºC. In contrast, extensive pattern of asphaltene deposition was observed in a Hele-Shaw cell using Peace River Bitumen at 20 ºC and
butane vapor. Measured stabilized drainage rates using propane as solvent were 2.5 time higher than those measured with butane.

In the Vapex process, noncondensable components (e.g., methane or carbon dioxide) are added to the solvent for adjusting the dew point of the solvent mixture. This eliminates the possibility of condensation of the solvent that could result in potential impairment of permeability because of asphaltene deposition and waste of solvent in the form of capillary trapping (Das and Butler, 1998; Talbi and Maini, 2003, 2008). Talbi and Maini (2008) used carbon dioxide as an alternative to methane as carrier gas because of its economic and environmental benefits in addition to its better solubility in the oil phase. Experiments conducted in packed beds revealed better performance of carbon dioxide-propane mixture at higher pressures (i.e., >4.13 MPa) compared to methane-propane mixture. However, appearance of a carbon dioxide-rich phase at higher pressures can negatively affect the mobility of the oil phase by reduction of relative permeability to the mobilized oil phase. In addition, the efficiency of the process can be adversely affected by lower solvent content of the injected solvent vapor. Experiments carried out at lower pressures (i.e., <4.13 MPa) showed insignificant effect of carrier gas type on the performance of the process.

Dong and Shirif (2004) developed a mathematical model to calculate the transient Rayleigh number in the mobile layer for the experiments conducted by Dunn et al. (1989). Calculated Rayleigh numbers confirmed the presence of natural convection induced in lower viscosities in cases where carbon dioxide was used as the solvent. On this basis, Dong and Shirif (2004) recommend replacing methane with carbon dioxide for enhancing the extraction rates in the Vapex process for lower viscosity
crudes. However, calculated Rayleigh numbers did not show the presence of convective instability at higher viscosities.

Cavallaro et al. (2005) used a complex solvent mixture (44.44% carbon dioxide, 2.84% ethane, 19.67% propane, 3.71% iso-butane, 8.19% normal butane, and 21.15% C5+) to perform a laboratory test for evaluation of the effect of solvent on performance of the Vapex process. This study was conducted in a 2D slab model with dimensions of 39 cm×38 cm×5 cm using Llancanelo oil from Argentina. This study was conducted at 9.4 MPa and 56 ºC to simulate the reservoir condition. Despite significant asphaltene deposition in the physical model, no permeability impairment or blockage was observed as a result of asphaltene precipitation. This study achieved recovery factor of 82.9% of the initial oil in the model.

Review of the performance of the Vapex process with respect to the type of solvent is not conclusive. Therefore, author believes the best policy for solvent selection is to determine the solvent composition theoretically, followed by experimental validation of the selected solvent in order to maximize the efficiency of the process. Experimental evaluation of phase behavior properties of oil-solvent system is necessary for evaluating solubility, viscosity reduction, and potential asphaltene deposition that could reduce the formation permeability at the reservoir condition.

Butler and Mokrys (1995) developed Equation 2.1 to determine the solvent-oil-ratio (SOR) in the Vapex process. Experimental studies using ethane, propane, and butane showed that practical values of the net cumulative SOR ranges from 0.12-0.3 m³ of the solvent per m³ of the oil. For SOR ratios higher than 1, the Vapex process may not be economically an attractive recovery option.
\[
SOR = \frac{1}{1 + \lambda \left( \frac{V}{v} - 1 \right)}
\]  

(2.1)

In N-Solv\textsuperscript{TM}, heated solvent vapour is injected at moderate pressures into the gravity drainage chamber. The vapour flows from the injection well to the colder perimeter of the chamber where it condenses, delivering heat and solvent directly to the bitumen extraction interface (Nenniger and Dunn, 2008; Nenniger and Gunnewiek, 2009). This process is operated based on condensing the solvent vapor below the bubble point of the solvent mixture compared to the Vapex process which is operated above the dew point pressure. Despite suggested advantages, efficiency of this process is very sensitive to buildup of noncondensable components in drainage chamber that negatively impacts the mass transfer rate to the extraction zone. In addition, potential asphaltene deposition occurring at higher solvent concentrations in N-Solv\textsuperscript{TM} on the edge of the gravity chamber can impair the permeability in the mixing zone and reduce the efficiency of the this process.

2.3. Heavy Oil and Bitumen Viscosity

High viscosity of heavy crudes is the major obstacle to exploit these resources. The viscosity of heavy oil or bitumen depends on its composition, temperature, and pressure. In the Vapex process, the main objective is the viscosity reduction of the oil through dissolving solvent into the oil phase.

The effect of solvent concentration on the crude viscosity is significant (Das and Butler, 1996; Mehrotra and Svrcek, 1986, 1988). For instance, conducted study by Mehrotra and Svrcek (1988) showed that the viscosity of carbon dioxide-saturated bitumen at 25 °C decreases from 50,000 mPa\textperiodcentered{}s to about 500 mPa\textperiodcentered{}s when the pressure is increased from
atmospheric pressure to about 5 MPa. Lederer (1934) developed a relation for determining the effect of dissolved solvent concentration on viscosity of heavy oil and bitumen (Equation 2.2).

\[ \mu_{\text{mix}} = \mu_o^{f_b} \mu_s^{f_s} \]  

(2.2)

where

\[ f_b = \frac{\gamma o}{\gamma o + c_s} \]  

(2.3)

and

\[ f_s = 1 - f_b \]  

(2.4)

Shu (1984) developed Equation 2.5 to determine the \( \gamma \) value in Equation 2.3.

\[ \gamma = \frac{17.04(\rho_o - \rho_s)^{0.5237} \rho_o^{3.2745} \rho_s^{1.6316}}{\ln\left(\frac{\mu_o}{\mu_s}\right)} \]  

(2.5)

Kendal and Monroe (1917) developed a simple approach to estimate the effect of solvent concentration dissolved in the oil phase on its viscosity (Equation 2.6).

\[ \mu_{\text{mix}} = \left(C_n \mu_s^{0.09} + (1 - C_n) \mu_o^{0.09}\right)^{\frac{1}{0.09}} \]  

(2.6)

In Equation 2.6, equivalent volume fraction of the solvent concentration can be determined by Equation 2.7.

\[ C_n = \frac{C_r \rho_s}{\frac{C_o \rho_o}{MW_o} + \frac{C_s \rho_s}{MW_s}} \]  

(2.7)

Figure 2.2 presents the dependency of the viscosity of heavy oil or bitumen on solvent concentration (i.e., butane) for three oils with different initial viscosities. This figure also
shows the comparison between the measured and predicted viscosity of Plover Lake oil-butane measured in this study.

Jin (1999) developed a correlation for estimating the effect of solvent concentration on bitumen based on study conducted on Peace River bitumen-butane system (Equation 2.8). The experimental data used in development of this correlation showed a viscosity reduction from 150 to 40 mPa·s at room temperature (23±2 °C) as butane mass fraction increased from 0.10 to 0.17.

\[
\mu_{\text{mix}} = 1.6609 \times 10^3 C_s^{-2.12}
\]  

(2.8)

Figure 2.2: Concentration dependency of heavy oil and bitumen viscosity to butane concentration

Investigation of the Vapex studies in physical models has established that the mixing of solvent and heavy oil causes asphaltene precipitation (Mokrys and Butler, 1993; Das and
Butler, 1994a; Lim et al.; 1994). Dissolution of solvent in the oil phase not only reduces its viscosity, it also could result in deasphalting of the oil at higher solvent concentrations. Nghiem et al. (2001) proposed a model for prediction of asphaltene precipitation (i.e., reversible and irreversible) that was evolved from an earlier model proposed by Kohse et al. (2000). This process can lead to further reduction of viscosity of the oil phase and in-situ upgrading of the oil (Mokrys and Butler, 1993; Cavallaro et al., 2005). Cavallaro et al. (2005) reported a strong correlation between concentrations of nickel and vanadium and asphaltene content of produced oil.

Experimental studies using mixtures of maltenes and asphaltene revealed the presence of a critical asphaltene content where dramatic increase in viscosity was observed above this concentration (Speight, 1980; Long, 1981; Svrcek and Mehrotra, 1989; Argillier et al., 2001, Luo and Gu, 2005). Argillier et al. (2001) concluded heavy oils can be described as entangled solvated asphaltene particles in a solvent constituted of the maltenes. High viscosities of this type of crudes are due to the enlargement of solvated asphaltene particles. Therefore, any process that can control or limit the asphaltene particles enlargement can reduce the viscosity. Figure 2.3 shows the thermal dependence of the relative viscosity of maltenes recombined with various amount of asphaltenes. In this figure, the relative viscosity is the ratio between viscosity of asphaltene-maltenes mixtures and viscosity of maltenes.

Singhal et al. (2002) studied the effect of aqueous phase presence on compositional change in the Vapex process using MRI imaging technique. They reported slower solvent chamber growth rates and no deasphalting of the produced oil in experiments conducted in absence of aqueous phase. They speculated that the facilitation of deasphalting by
water droplets is due to the increase of entrainment probability of droplets within the heavy oil. They concluded that the minimum required solvent content for deasphalting is less in porous media than the bulk oil.

Viscosity of the heavy is also a strong function of the temperature. Therefore, application of a hybrid process (i.e., warm Vapex) has been considered for improving the efficiency of the Vapex process (Butler and Mokrys, 1991; James et al., 2007). Seyer and Gyte (1989) reported up to five orders of magnitude viscosity reduction (i.e., 900,000 to 100 mPa·s) by temperature rise from 20 to 200 ºC for Athabasca bitumen. A two-parameter correlation developed by Das and Butler (1996) for viscosity prediction of Peace River bitumen at different temperatures is given in Equation 2.9.

\[
\log_{10} \log_{10} (\mu_c + 0.7) = 9.523535 - 3.57231 \log_{10} (T + 273.15)
\]  

(2.9)

Figure 2.3: Thermal dependency of the relative viscosity of maltenes recombined with various amounts of asphaltene (after Argillier et al., 2001)
2.4. Effect of Porous Media Characteristics on the Vapex Process

Since the pioneering study by Butler and Mokrys (1991), considerable research has been devoted to understand the effect of various characteristics of porous media that influence the performance of the Vapex process. This section assimilates and presents a review on the effect of various characteristics of porous media on the efficiency of the Vapex process.

2.4.1. Permeability and Heterogeneity

Butler and Mokrys (1989) carried out the first lab-scale study in a Hele-Shaw cell with Athabasca and Suncor Coker Feed bitumen with toluene as a solvent. On the basis of this study, they developed and validated a model for prediction of stabilized drainage rates in the Vapex process. In this model, the drainage rates were proportional to the square-root of the permeability. Oduntan et al. (2001) conducted a series of Vapex experiments in packed beds in permeability range of 25 to 192 D and different drainage heights. They concluded that there is a square root relationship between the observed stabilized drainage rates and permeability of the sand pack.

Butler and Mokrys (1989) observed that the drainage rates did not increase linearly with the square-root of permeability at higher permeabilities. In fact, for high permeabilities in the range of 220 to 640 D, the influence of permeability has been found to be insignificant (Karmaker and Maini, 2003a; Yazdani and Maini, 2005, 2008).

Moghadam et al. (2007) studied the effect of permeability on the efficiency of the Vapex process in the permeability range of 16-310 D. Measured solvent-oil ratio of the produced oil increased with reduction of permeability. A closer investigation of stabilized drainage rates revealed no dependency, unlike the square-root dependence on permeability as...
proposed by Butler and Mokrys (1989). Moghadam et al. (2007) concluded that higher solvent-oil mixing occurs at a low permeability because of increased contact time. This results in further in-situ upgrading of the oil and deviating from the square-root trend observed at higher permeabilities.

Formation heterogeneity due to presence of low- and high-permeability layers, shales, and fractures is the other parameter that can significantly affect the efficiency of the solvent mass transfer and consequently the performance of the Vapex process. Jiang and Butler (1996a) performed a series of experiments to investigate the effect of formation heterogeneity on the Vapex process. They simulated various heterogeneity scenarios using continuous and discontinuous low-permeability layers with horizontal layers of two different-sized sands to get different permeability sand packs (i.e., 20-30 mesh for 217 D and 30-50 mesh for 43.5 D). Conducted experiments showed that low-permeability layers yield a lower production rate than that with homogeneous model of high-permeability packing. In a layered model, the flow of solvent vapor from a high permeability region into a low permeability region could be limited by capillarity due to reduced pore size; the capillary pressure gradient obstructs the flow of gas. With bottom injection, oil cannot drain from a higher low-permeability until the height of diluted oil above the boundary is large enough to overcome the capillary pressure. Due to this effect, the oil production rate decreases when the vapor chamber reaches the lower boundary of low-permeability layer. Due to live oil potential gradients and different advancing velocities, low-permeability layers retained more oil than high-permeability layers. Capillarity effect in low-permeability layers limited the ingress of the vapor and resulted in a wider vapor chamber and consequently oil extraction from a larger area. The injection of solvent directly into
the low-permeability layer by a top injection well could speed the development of the vapor chamber and minimize the effect of capillarity due to the establishment of an imposed pressure gradient. The flow channels or fingers formed between the injector and producer by initial displacement result in the development of the vapor chamber in both layers and the oil drains from both low- and high-permeability regions. Jiang and Butler (1996a) also concluded that the presence of vertical fractures were beneficial in layered formations by establishing the communication between the layers.

Oduntan et al. (2001) investigated the effect of layering on the drainage rate in the Vapex process with layers of high- (192 D) and low- (85 D) permeability glass beads. In this study, seven different combinations of layers were examined. The layering and packing in the first six experiments were conducted in such a way to achieve an average permeability of 118 D. The last model had the highest permeability of 125 D. Comparison between the production rates from the first six models showed no significant differences. The highest production was for the last model with the highest permeability. It was observed that the production rate decreased slightly with an increase in the number of layers in the physical model. The production rate was found to be lower for a heterogeneous medium with the same average permeability. Residual oil saturation was observed to be high in low-permeability layers situated above high permeability layers.

Azin et al. (2008a) investigated the effect of solvent composition and pressure on the Vapex process in a fractured system. They concluded that high recovery factors could be achieved by operating the process closer to dew point of the solvent. They observed that presence of noncondensable component in the vapor phase negatively affect the recovery factors. Azin et al. (2008b) developed a mechanistic model in simulating the Vapex
process in fractured media. The model consisted of a 2D convective-diffusion equation for a constant variable system and the simplified 2D Brinkman equation for the flow of diluted oil. The two equations were coupled by the continuity equation. The Brinkman equation is an extension of Darcy law that includes the dissipation of the kinetic energy by viscous shear in porous media flow. The system of equations neglected the capillary effect and the relative permeability within the system. They used Perkins and Johnston correlation (1963) to calculate the dispersion coefficient in the studied process. However, they did not address the boundary condition of the system, the correlation used for viscosity reduction, or the molecular diffusion coefficient used in the simulation; nor did they mention the method used to distinguish the difference in physical properties between bitumen and dilute oil. The mechanistic model developed in this study was never used in a numerical simulation. Instead, a numerical simulation was simply conducted using STARS™. They verified the numerical model by comparing the simulation results to the production data of Butler and Jiang (2000), however, they did not address the advancement of the interface and the shape of the bitumen-solvent interface. Based on conducted simulation study of the Vapex process in fractured media, Azin et al. (2008b) concluded that the presence of fractures significantly increases the surface for solvent diffusion into the matrix. However, solvent breakthrough may occur earlier in the case of interconnected fractures surrounding the matrix blocks. Based on a simulation study of Vapex process in high-pressure fractured heavy oil reservoirs, they concluded that the dispersion mass transfer is dominant through the majority of process; however, diffusion is an important factor in the early stage of process. The simulation study showed at a constant pore volume injection of solvent oil production rates and
recoveries were less pronounced, and similar overall results for a range of solvent injection rates were obtained in laboratory experiments.

### 2.4.2. Drainage Height

Oduntan et al. (2001) used rectangular channels with different drainage heights (21-247 cm), but similar cross-sections to study the effect of drainage height on production rate in the Vapex process. The physical models were packed with glass beads to create a medium with permeability of 136 D and average porosity of 38%. Stabilized drainage rates were observed until 80-90% of the initial oil in the models had been produced. The measured stabilized drainage rates showed almost a square-root relationship with the drainage height as had been proposed in Butler and Mokrys model (1989). Equation 2.10 gives the correlation between the observed drainage rates and drainage height in study conducted by Oduntan et al. (2001). The experiments conducted in layered packed models with permeability between 86 D and 97 D also showed the same dependency to the drainage height despite lower drainage rates.

\[
Q = 3.9 \times 10^{-7} h^{0.55}
\]  

(2.10)

Karmaker and Maini (2003a) studied the effect of model height (i.e., 7.5-30 cm) and sand grain size (permeability range of 220 to 640 D) on drainage efficiency of the Vapex process. Later, Yazdani and Maini (2005, 2008) extended this study to the higher drainage heights (i.e., 7.5-100 cm) in the same permeability range. Three different oils with average viscosity of 18,451 mPa·s and density of 0.9831 g/cm³ were used. These studies showed that grain size distribution has no effect on the dispersive mixing in the studied permeability range. In contrast, the observed drainage rates displayed a significantly higher dependency on drainage height. In fact, this study revealed that the
stabilized rate was a function of drainage height to the power of 1.1 to 1.3. On this basis, Yazdani and Maini (2005) developed a new correlation to scale up the experimental data to the field condition (Equation 2.11).

\[
\frac{Q_f}{Q_w} = \left( \frac{h_f}{h_m} \right)^n \left( \frac{\sqrt{k\phi}}{\sqrt{k\phi}} \right)_m \quad 1.1 < n < 1.3
\]

Ramakrishnan (2003) studied the effect of dip angle on production rate using a homogeneous physical model with permeability of 156 D and dip angles of 45°, 75°, 80°, and 90°. The results showed that the production rate increased with the dip angle. The maximum production rate was along the vertical direction.

2.4.3. Water Saturation

There are Vapex studies in literature that have been conducted in absence of aqueous phase in the porous media. These studies were rationalized by the assumption of lack of any influence of immobile water saturation on oil phase isoperms and negligible solubility of the solvent vapor into the aqueous phase.

Fisher et al. (2000, 2002) carried out a set of MRI-based tests in a small physical model to investigate the effect of aqueous phase on the shape of the Vapex chamber. For this purpose, the study was conducted in absence and presence of immobile aqueous phase. Analysis of the results showed that the rate at which the vapor chamber expands was much faster when the experiment was conducted in the presence of immobile aqueous phase. It was found that the deasphalting mechanism seems to be promoted in the presence of connate water and this was evident through the so-called asphaltene streaks formed on the vapor chamber’s edges. Despite having semi-solid asphaltene streaks in the model, the flow was not seriously impeded.
Etminan et al. (2008) conducted an experimental program to investigate the role of aqueous phase in the Vapex process. This study was carried out in a low-permeability sand pack (10 D) and oil sample from the Frog Lake reservoir. In these experiments, the immobile aqueous phase saturation was in the range of 5.94-9.09%. Comparison between the dry and wet experiments revealed faster spreading of the vapor chamber in lateral direction in presence of immobile aqueous phase. The measured drainage rates showed that presence of aqueous phase increases the rates at the early stage of the drainage process that is followed by reduction in drainage rates compared to tests conducted in absence of aqueous phase. Presence of aqueous phase also caused appearance of more irregular viscous fingering patterns on the edge of the vapor chamber. Etminan et al. (2008) concluded that the presence of these fingering patterns resulted in a thicker saturation transition zone and consequently increase in surface for mass transfer between solvent vapor and oil phase.

2.4.4. Capillarity

Butler and Mokrys (1989) developed and validated their model on the basis of observed drainage rates in a Hele-Shaw cell. However, the subsequent experimental studies in porous media achieved considerably higher recovery rates. Most researchers have ascribed the increase in drainage rates to an increased rate of mass transfer. Das and Butler (1994b, 1998) used effective diffusion coefficients 3 to 10 times higher than the molecular diffusion coefficients to history match their measured oil drainage rates. Kapadia et al. (2006) used dispersion coefficients four to five orders of magnitude higher than the reported values of molecular diffusion of butane in heavy oil in order to history-match their experimental data for oil drainage rates. Dunn et al. (1989) had pointed out
the possibility of increased recovery due to dispersion. Although the introduction of the dispersion coefficient improved the rate of recovery of their analytical model by almost an order of magnitude, they were unable to address the 460-fold discrepancy with respect to some of their experimental results. Part of this discrepancy may have been due to the fact that they used a constant diffusion coefficient, independent of solvent concentration. Lim et al. (1994) used an effective diffusivity; two or three orders of magnitude higher than the molecular diffusivity, to history match their experimental results in sand packs. Boustani and Maini (2001) conducted a series of experiments in a Hele-Shaw cell with propane as the solvent and Lloydminster heavy oils (Dover and Penny) over the temperature range of 10.5 to 20.5 ºC. These tests were conducted in a Hele-Shaw cell with a drainage height of 7.6 cm and a permeability of 257 D. They concluded that Butler’s model tends to overestimate the molecular diffusion coefficient and underestimate the overall mass transfer in terms of a coupled diffusion-dispersion process in porous media due to neglecting the dispersion phenomenon. Therefore, including the effect of diffusion and dispersion could improve the quality of predictions of the Butler’s model.

Cuthiell et al. (2006) investigated the role of capillarity in the Vapex process using a sand pack with permeability of 4.3 D, similar to that of a typical heavy oil reservoir in western Canada. This study was carried out in a sand pack of dimensions 25 cm×30 cm×2.8 cm. In order to improve the quality of fluid distributions maps using CT scanner, these tests were conducted in absence of connate water. The oil used in this experimental study was heavy oil from Aberfeldy field in the Lloydminster area. The simulation study that was conducted to model the key mechanisms (i.e., solubility, viscosity reduction,
diffusion/dispersion, and capillary pressure) revealed that the convective dispersion from velocity-dependent effects play a minor role in mixing. In contrast, most of the mixing was explained by molecular diffusion, together with the capillarity effect. They also concluded that the effects of numerical and explicit dispersion could be separated by varying the grid size. Their estimated diffusion coefficient of $2.4 \times 10^{-5}$ cm$^2$/s was comparable with the value of $1.63 \times 10^{-5}$ cm$^2$/s predicted from the Das and Butler (1996) correlation. In other words, they concluded that dispersive mixing was consistent with molecular diffusion with little or no enhancement due to convective dispersion in drainage phenomena in presence of capillary forces. On this basis, they suggested that the occurrence of layered heterogeneity (i.e., different layers in a formation) would increase mixing and consequently drainage rates.

Ayub and Thuinuzzaman (2007) carried out experimental study and numerical modeling to investigate the capillarity effect in the Vapex process. They concluded capillary forces tend to retard gas production without affecting the overall recovery. Rostami et al. (2007) used a simulation approach and reported a delay in breakthrough time, wells communication, and gas production at the beginning of the solvent vapor injection because of higher capillary pressures. In addition, higher capillary pressures caused a thicker saturation zone around the edge of the Vapex chamber that lead to the enhancement of mass transfer.

### 2.4.5. Presence of Aquifer and Gas Cap

It is believed that when employing a saturated hydrocarbon vapor in conjunction with horizontal wells to mobilize and recover viscous oils and bitumen, the bottom water zone can serve as a beneficial tool for providing initial injectivity (Butler and Mokrys, 1998).
Since the vapor injected at reservoir temperature and it is essentially insoluble in water while strongly soluble in oil, there are no heat or material losses to the water layer. Furthermore, the mobile water layer will under ride the lighter oil and assist in moving it towards the producing well. In the presence of an aquifer, in situ mixing of solvent and oil may be further enhanced by the stirring action of water oozing through the porous media.

When a reservoir contains an overlying gas cap the primary recovery becomes even poorer due to gas conning problems. Steam assisted thermal methods are not economically viable because of significant steam loss to the gas cap. In situ combustion is also unsuitable due to the preferential migration of the injected air into the gas cap. The Vapex process can be successful in the presence of a gas cap because of its unique advantages over steam based thermal recovery methods (Karmaker and Maini, 2003). The migration of injected solvent into the gas cap can even accelerate the mass transfer and thereby improve the process performance. The horizontal injector might be advantageously completed near the gas cap zone. This will help the injected solvent to rapidly spread out over the gas zone and make a large fraction of the gas oil contact area effective for mass transfer by diffusion in the early stages of the process. Karmaker and Maini (2003b) experimentally evaluated the effectiveness the Vapex process in the presence of a small gas cap in contact with the oil zone. It was found that presence of a small gas cap was fully compatible with the application of the Vapex for heavy oil recovery. It was found higher solvent injection rates at the beginning of the process could be beneficial. Reduced lateral distance between the injector and producer could accelerate the breakthrough and consequently production phase.
Frauenfeld et al. (2004) used a visual physical model (drainage height of 30 cm) with average permeability of 400 D to study the effect of aquifer in the Vapex process. They showed that a bottom-water Vapex could combine the features of conventional Vapex (i.e., a rising vapor chamber above the injection well with a bottom-up extraction of oil at the water interface). All experiments produced substantial water first. This was necessary, as some water had to be displaced by the gas in order to contact the oil with solvent at the oil-water interface. This study highlighted the need for having the wells completed at the oil-water contact.

2.5. Oil Rate Equation

Butler and Mokrys (1989) developed the first mathematical model to predict the rate of bitumen extraction with liquid solvent in a Hele-Shaw cell. They assumed that oil flows along the interface in a thin diffusion boundary layer where drainage rate of the undiluted bitumen is negligible. Because of total miscibility of bitumen in toluene, there was no interfacial tension in the system studied.

Figure 2.4 shows a small part of the vapor-bitumen interface away from the drainage edge in a Hele-Shaw cell. In this diffusion layer, the fluid flow occurs nearly parallel to the interface. Assuming a pseudo-steady state, any position of the interface advances at a constant velocity, \( U \). In the cell thickness the solvent volume fraction, \( c_s \), and concentration dependent properties (e.g., viscosity, density, and diffusivity) can be assumed to be constant. On the basis of these assumptions, the material balance over a thin shell, the solvent concentration in the diffusion layer can be expressed by a 1D diffusion equation (Equation 2.12).
\[-D \frac{dc_s}{d\xi} = Uc_s\]  \hspace{1cm} (2.12)

In diffusion layer, the solvent concentration varies between \(c_{\text{max}}\) at the interface of oil-solvent and \(c_{\text{min}}\) that is the minimum solvent concentration required for mobilizing the oil phase.

By assumption of steady-state pressure gradient in the direction of the flow inside the diffusion layer, the balance of pressure gradient in direction of flow can be obtained by a simplified form of Brinkman equation (Bird et al., 2001) as given in Equation 2.13.

\[\Delta\rho g \sin \theta = u\mu/k + \mu d^2u/d\xi^2\]  \hspace{1cm} (2.13)

Figure 2.4: Fluid flow and mass transfer take place in the boundary layer between the solvent vapor and immobile bitumen (Das and Butler, 1998a)

In the diffusion layer, there is a capillary force normal to the interface due to interfacial tension between the solvent and the oil phase. Since the direction of flow is parallel to the interface, this term did not appear in the force balance given in Equation 2.13. Therefore, the drainage rate of diluted oil in the diffusion layer with solvent concentration of \(c_s\) can be determined by Equation 2.14.
\[ Q = \int_{\xi_{\text{min}}}^{\xi_{\text{max}}} u(1 - c_s) d\xi \]  \hspace{1cm} (2.14)

In Equation 2.13, Butler and Mokrys (1989) showed that the second term (i.e., second-order differential term due to change in velocity gradient) in a closely spaced plates of a Hele-Shaw cell is negligible. On this basis, Equation 2.13 becomes equivalent to Darcy’s equation. By substituting the related terms in Equation 2.14 from Equations 2.12 and 2.13 and using the proper limits of integration, Equation 2.15 can be derived to calculate the extraction rate in the diffusion layer.

\[ Q = \frac{k g \sin \theta}{U} \int_{c_{\text{min}}}^{c_{\text{max}}} \frac{\Delta \rho D(1-c_s)}{\mu c_s} dC_s = \frac{k g \sin \theta}{U} N_s \]  \hspace{1cm} (2.15)

Where

\[ N_s = \int_{c_{\text{min}}}^{c_{\text{max}}} \frac{\Delta \rho (1-c_s) D}{\mu c_s} dc_s \]  \hspace{1cm} (2.16)

The schematic of a typical Hele-Shaw experiment is presented in Figure 2.5. In this figure, the cross-hatched area presents the amount of oil drained through the section at ordinate, \( y \), during time, \( t \), from the beginning of the tests. This area is equivalent to the amount of bitumen drained through the section at the same time period. The can be expressed by Equation 2.17

\[ \int_0^t Q dt = \phi \Delta S_p \int_y^H x dy \]  \hspace{1cm} (2.17)

Differentiation with respect to \( t \) and \( y \) results in:

\[ \frac{\partial Q}{\partial y} = -\phi \Delta S_p \frac{\partial x}{\partial t} \]  \hspace{1cm} (2.18)

In vapor extraction in a Hele-Shaw cell, it is assumed through the growth of the vapor zone it advances solely in the \( x \)-direction. As a result, Equation 2.19 can determine the
geometric relationship between the velocity of the advancing interface and the advance of the boundary.

\[
U = \frac{\partial x}{\partial t} \sin \theta \tag{2.19}
\]

Combining Equations 2.15, 2.18, and 2.19 yield Equation 2.20.

\[
Q = \frac{-kg_\phi \Delta S_o N_s}{(\partial Q / \partial y)} \tag{2.20}
\]

![Figure 2.5: Schematic of the vapor extraction in Hele-Shaw cell (Das and Butler, 1998a)](image)

Finally, the drainage rate in diffusion layer at any ordinate, \(y\), can be determined by integration of Equation 2.20.

\[
Q_y = \sqrt{2kg_\phi \Delta S_o N_s (h - y)} \tag{2.21}
\]

The flow rate out of the cell (\(y=0\)) can be determined by Equation 2.22.
\[ Q_y = \sqrt{2k_g \phi \Delta S_o N_i h} \]  \hspace{1cm} (2.22)

Das and Butler (1998a) extended the applicability of developed mathematical model for Hel-Shaw cell (Equation 2.22) to porous media. In this study, the material balance of solvent in diffusion layer in porous media was expressed by Equation 2.23.

\[ -AD_p \frac{dc_s}{d\xi} = A\phi U c_s \]  \hspace{1cm} (2.23)

In Equation 2.23, the molecular diffusion of solvent was replaced with the apparent diffusion coefficient. The relationship between the apparent and intrinsic diffusion coefficient is expressed in Equation 2.24

\[ D_p = D\phi^\Omega \]  \hspace{1cm} (2.24)

Substitution of Equation 2.24 into Equation 2.23 results in Equation 2.25.

\[ -D \frac{dc_s}{d\xi} = \phi^{-\Omega} U c_s \]  \hspace{1cm} (2.25)

Using the same analysis presented for Hele-Shaw cell will result in Equation 2.26 for prediction of extraction rate in a porous medium in terms of drained area of vertical cross section.

\[ Q_y = \sqrt{2k_g \phi^\Omega \Delta S_o N_i h} \]  \hspace{1cm} (2.26)

For the case of unconsolidated rocks, it was proposed to use \( \Omega = 1.3 \) as a cementing factor, which was originally reported by Pirson (1958).

Equation 2.26 considers the flow of oil from one side of a Vapex chamber, which is symmetric against the vertical line, through the horizontal injector and producer. Therefore, to calculate the volumetric flow rates from both sides of a Vapex chamber in a horizontal well with the length of \( L \), Das and Butler (1998a) proposed Equation 2.27.
\[ Q = 2L \sqrt{2kg \phi^2 \Delta S_o \phi N_s} \]  

(2.27)

Dunn et al. (1989) also developed a theoretical model based on an experimental study in which ethane and carbon dioxide were injected at low temperatures to recover bitumen by gravity drainage. They adopted a similar mathematical methodology to the one developed by Butler et al. (1981) for describing the thermal effect and analyzing the SAGD process.

In another study, Okazawa (2009) developed a new analytical model to include cases of diffusion coefficients being dependent on solvent concentration. The developed model in this study confirmed the overall square-root dependence of drainage rates on most of the key reservoir parameters as indicated by earlier studies. The developed analytical model can be used for prediction of the total flow of the bitumen-solvent mixture and the flow of bitumen itself. Okazawa (2009) concluded that the developed model can quantify the vertical flow between simulation blocks better than the conventional finite-difference model.

### 2.6. Mass Transfer Phenomenon in the Vapex Process

Performance of the Vapex process is directly related to the amount of solvent dissolving into the heavy oil or bitumen (Boustani and Maini, 2001). Diffusion and convective dispersion are present almost everywhere in the model, but the remaining mechanisms prevailing only within the mixing zone. Particularly, interfacial tension and capillary forces become a material issue in the two-phase region only. Accurate estimation of mass transfer rate between solvent and oil is necessary to determine: i) the amount of flow rate of solvent vapor required throughout implementation of the Vapex process, ii) the extent
of heavy oil and bitumen reserves that undergo viscosity reduction, iii) the time required for optimum viscosity reduction of the heavy oil and bitumen necessary for mobilizing the oil, and iv) the rate of producing live oil from the formation (Upreti et al., 2007).

Despite the importance of diffusion and dispersion in the Vapex process, the knowledge about these mechanisms in vapor solvent-heavy oil systems is limited. In the following sections, the related work and literature review will be presented.

2.6.1. Diffusion Process

Molecular diffusion is the primary mixing mechanism responsible for solvent absorption by oil and bitumen resulting in dilution of the oil phase. There are different theories (e.g., hydrodynamic theory, kinetic theory, and absolute rate theory) for estimation of diffusion coefficients. Many of these theories were built around kinetic phenomena in liquids. However, none of these theories are quite satisfactory in prediction of realistic diffusivities for heavy oil and bitumen systems (Oballa and Butler, 1989). This is because of arbitrary assumptions made in treatment of kinetic forces involved in diffusion process.

In addition to these theories, there are different correlations for estimating the diffusion coefficient in heavy oil and bitumen systems. Correlations developed on the basis of molecular theory with inverse relationship between the diffusivity and viscosity are more popular for determination of diffusion coefficients in heavy oil and bitumen systems. Therefore, they make it possible to define the molecular definition as a function of solvent concentration. Oballa and Butler (1989) reported a strong concentration dependency on diffusivity in the bitumen-toluene system. Figure 2.6 shows the effect of solvent concentration on diffusion coefficient calculated by Equation 2.28 developed by
Das and Butler (1996) for Butane-Peace River system. In this figure, viscosities of Plover Lake Heavy oil at different solvent concentrations were determined by Equation 2.6. Table 2.1 summarizes some of popular correlations for estimating the molecular diffusion coefficient in heavy oil and bitumen systems.

\[ D = 9.907 \times 10^{-5} \mu_{mix}^{-0.46} \]  

(2.28)

![Figure 2.6: Concentration dependency of viscosity and diffusion coefficient on normal butane concentration in Plover Lake Oil](image)

Upreti and Mehrotra (2000,2002) used indirect non-intrusive pressure decay experimental method for measuring the diffusivity of CO\(_2\), CH\(_4\), C\(_2\)H\(_6\), and N\(_2\) gases in Athabasca bitumen in temperature range of 25-90 °C. The results of these measurements are summarized in Table 2.2. In addition, they developed a simple relationship between the diffusion coefficient and temperature (Equation 2.29).
\[ \ln D = d_0 + d_1(T + 273.15) \]  
(2.29)

The measured diffusion coefficients for various gases in heavy oil and bitumen using different techniques are in the order of $10^{-9}$ to $10^{-11}$ m$^2$/s (Riazi, 1996; Sveck and Mehrotra, 1982; Schmidt, 1989; Nguyen and Farouq-Ali, 1995; Zhang et al. 2000; Yang and Gu, 2006; Sheikha et al., 2006, James, 2009).

**Table 2.1: Comparison of correlations for estimation of liquid diffusivities (after Boustani and Maini 2001)**

<table>
<thead>
<tr>
<th>Correlation</th>
<th>Equation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Wilke and Chang (1955)</td>
<td>[ D = 7.4 \times 10^{-8}(\phi \Omega)^{0.5}T \mu_o^{-1}V_A^{0.6} ]</td>
</tr>
<tr>
<td>Hayduk and Cheng (1971)</td>
<td>[ D = \alpha \mu_o^{-\beta} ]</td>
</tr>
<tr>
<td>Hiss and Cussler (1973)</td>
<td>[ D = \alpha \mu_o^{-2/3} ]</td>
</tr>
<tr>
<td>Hayduk et al., (1973)</td>
<td>[ D = 0.0591 \times 10^{-9} \mu_o^{-0.545} ]</td>
</tr>
<tr>
<td>Hayduk and Minhas (1982)</td>
<td>[ D = 13.3 \times 10^{-8}T^{1.47}V_A^{0.71} \mu_o^{(10.2/V_c^{-0.791})} ]</td>
</tr>
<tr>
<td>Reid et al. (1987)</td>
<td>[ D = RT / 3\pi \mu_o d ]</td>
</tr>
<tr>
<td>Das and Butler (1996)</td>
<td>[ D = 1.306 \times 10^{-8} \mu_o^{-0.46} ]</td>
</tr>
<tr>
<td>Das and Butler (1996)</td>
<td>[ D = 4.131 \times 10^{-9} \mu_o^{-0.46} ]</td>
</tr>
</tbody>
</table>

**2.6.2. Dispersion process**

The proposed model in Equation 2.22 is for the prediction of oil drainage rates observed in the experiments conducted in a Hele-Shaw cell. However, the subsequent experiments in porous media resulted in much higher rates than predicted values using Equation 2.22.
Table 2.2: Summary of experimental data on diffusivity of gases in Athabasca bitumen

<table>
<thead>
<tr>
<th>Gas</th>
<th>Pressure MPa</th>
<th>Temperature °C</th>
<th>Diffusivity ( \times 10^9 ) (m(^2)/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>4</td>
<td>25</td>
<td>0.1335</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>50</td>
<td>0.2338</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>75</td>
<td>0.3739</td>
</tr>
<tr>
<td>CO(_2)</td>
<td>4</td>
<td>90</td>
<td>0.4280</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>50</td>
<td>0.3980</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>75</td>
<td>0.7436</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>90</td>
<td>0.9319</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>25</td>
<td>0.0810</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>75</td>
<td>0.2932</td>
</tr>
<tr>
<td>CH(_4)</td>
<td>4</td>
<td>90</td>
<td>0.4315</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>25</td>
<td>0.0582</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>75</td>
<td>0.1518</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>90</td>
<td>0.2029</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>25</td>
<td>0.2539</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>75</td>
<td>0.4203</td>
</tr>
<tr>
<td>C(_2)H(_6)</td>
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<td>90</td>
<td>0.6081</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>75</td>
<td>0.4916</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>90</td>
<td>0.6917</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>25</td>
<td>0.0180</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>50</td>
<td>0.0513</td>
</tr>
<tr>
<td></td>
<td>4</td>
<td>75</td>
<td>0.2335</td>
</tr>
<tr>
<td>N(_2)</td>
<td>4</td>
<td>90</td>
<td>0.4960</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>25</td>
<td>0.0555</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>50</td>
<td>0.1717</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>75</td>
<td>0.4649</td>
</tr>
<tr>
<td></td>
<td>8</td>
<td>90</td>
<td>0.7460</td>
</tr>
</tbody>
</table>
Das and Butler (1998) used effective diffusion coefficients of 3-10 times higher than the molecular diffusion coefficients to match their measured oil discharge rates. In study by Kapadia et al. (2006), effective diffusion/dispersion coefficients four/five orders of magnitude higher (order of $10^{-6}$ cm$^2$/s) than the reported values of molecular diffusion of butane in heavy oil (order of $10^{-11}$ cm$^2$/s) were used to history match the observed drainage rates. Dunn et al. (1989) had pointed out the possibility of increased recovery due to dispersion. In this study they used diffusion coefficients up to 460-fold the molecular diffusion to observed drainage rates match the theoretical model. This study showed that presence of aqueous phase reduces the effect of dispersion on the basis of comparison between the effective dispersion coefficients used to match the model.

Although, application of dispersion coefficient improved the rate of recovery of their analytical model by almost an order of magnitude, Dunn et al. (1989) were unable to address the 460 times discrepancy with respect to some of their experimental results. Part of this discrepancy might be due to the fact that they used a constant diffusion coefficient, independent of solvent concentration. Lim et al. (1994) used an effective diffusivity; two to three orders of magnitude higher than the molecular diffusivity to history match their experimental results in sand packs. They mentioned that the mechanism(s) relating to increase in diffusivity is not well understood, but they singled out physical dispersion as most likely cause of the enhanced mass transfer in porous media.

Most researchers have ascribed the increase in drainage rate to an increased rate of mixing, characterized by the parameter $N_s$ in Equation 2.16, between the solvent and the oil (Das and Butler, 1998; Oduntan et al., 2001; Boustani and Maini, 2001; Talbi and Maini, 2003; Kapadia et al., 2006). Although some researchers have proposed that the
rate has a different dependency on drainage height from that given in Equation 2.17 (Karmaker and Maini, 2003; Moghadam et al., 2007, Yazdani and Maini, 2005, 2008). Kohse et al. (2000) developed a model for prediction of asphaltene dropout and potential upgrading of bitumen by Vapex process. Later, Nghiem et al. (2001) developed a fully compositional model for numerical simulation of the Vapex process. Molecular diffusion and convective dispersion were incorporated into this model to handle the mass transfer phenomenon. They used constant dispersivity coefficient, as a direct multiplier to the phase velocity (rather than Peclet number). In this model, the mixing process is effectively controlled through a total dispersion coefficient. Conducted sensitivity analysis confirmed higher efficiency of mixing. However, this study did not shed any lights on selection of realistic diffusion and dispersion terms in modeling of the Vapex process.

**Longitudinal and transverse dispersion:** Dispersion mechanisms may be lumped into five categories: diffusion, fluids effects (gravity, density, viscosity contrast, saturation levels, and turbulence), medium effect (turtuosity, auto-correlation), fluid-medium interactions (adsorption, chemical reaction, hydrodynamics, and heterogeneity effects), and boundary/initial conditions. The dispersion phenomena in porous media comprise of two components: longitudinal and transverse dispersion.

Longitudinal dispersion develops along the direction of the bulk flow movement in systems where both solute and solvent are flowing in the same direction. Downward movement of diluted oil along the interface will cause longitudinal dispersion in the same direction and transverse dispersion, perpendicular to the direction of bulk flow and into the bitumen bulk. In the search for a mechanism for enhanced delivery of solvent into the
bitumen, transverse dispersion was earlier perceived to be the sole mechanism. However, it was realized that both longitudinal and transverse dispersion are actively present in 2D models of Vapex experiments (Boustani and Maini, 2001).

Perkins and Johnston (1963) presented the most popular empirical correlations for estimating overall mass transfer coefficients in presence of longitudinal and transverse dispersions, which are defined in Equations 2.30 and 2.31.

\[
\frac{D_l}{D} = \frac{1}{F\phi} + 0.5 \frac{V\sigma_d}{D} \tag{2.30}
\]

where \( \frac{V\sigma_d}{D} < 50 \)

\[
\frac{D_t}{D} = \frac{1}{F\phi} + 0.0157 \frac{V\sigma_d}{D} \tag{2.31}
\]

where \( \frac{V\sigma_d}{D} < 10^4 \)

The first term in Equations 2.30 and 2.31 corresponds to the molecular diffusion. The second term in these equations is the dispersive terms of mass transfer. The constants in these correlations must be determined experimentally. Delgado (2006) critically reviewed the available data and correlations in literature on dispersion in packed beds. This comprehensive literature review showed that majority of these studies used fluids with negligible density and viscosity contrast, which is not the case in the Vapex process.

It is likely that the application of these equations for simulation of the Vapex process is limited due to very different nature of experimental data used in developing them. For instance, in developing these relationships, forced convection is the major driving mechanism, while in Vapex, gravity is the only driving force. In addition, the flow pattern is parallel to the driving force in these equations, which is not the case in the
Vapex process. The other assumptions in the development of these equations are negligible density and viscosity differences between fluids in porous media. In reality, there are significant density and viscosity contrasts between the solvent and oil phase that cannot be neglected in the Vapex process. It is also known that there is drastic changes in the viscosity of heavy oil over a short distance in the Vapex process.

Several researchers have recently attempted to determine the effective diffusion and dispersion using imaging techniques (e.g., CT scan and MRI) or developing mathematical models (Fisher et al., 2000, Karmaker and Maini, 2003; Cuthiell et al., 2003; Wen and Kantzas, 2005; Kapadia et al., 2006; Okazawa, 2009). However, the understanding of mass transfer phenomenon in Vapex process using these approaches is limited to a specific problem and simplifying experimental or mathematical assumptions.

Yazdani and Maini (2009) developed a correlation for estimating effective diffusion/dispersion coefficient based on a comprehensive experimental program in a wide range of drainage heights and permeabilities (Equation 2.32). The developed correlation is based on analytical Vapex model by Das and Butler (1996) and proposed composition dependent diffusion correlation by Hayduk and Cheng (1971).

\[
D_{eff} = 1.495 \times 10^{-7} h^{1.55} \mu^{-0.46}
\]  
(2.32)

Despite the fact that numerous experimental and numerical studies have been conducted to understand the mass transfer mechanisms in this process, these mechanisms and the effect of porous media characteristics on solvent-oil mass transfer are still poorly understood.

Here, a review of the effect of some key parameters (i.e. drainage height, capillarity, permeability, and solvent type) on mass transfer phenomena in the Vapex process is
presented. Because of the existing quality of data, the geometry of the models, and various procedures used in different studies, Vapex number \((N_{s,\text{exp}})\) was chosen to investigate the effect of selected parameters on the rate of solvent-oil mixing and the subsequent performance of the Vapex process. The Vapex numbers for different experiments were back-calculated by Equation 2.33 which derived from Equation 2.26. In order to isolate the effect of each parameter on performance of the mass transfer between the solvent and oil phase, the \(N_{s,\text{exp}}\) values were limited to experiments conducted in homogeneous sand packs. As a result, a quantitative understanding was obtained on the effects of individual parameters and interplay between the studied parameters on the efficiency of the mixing process and performance of the process.

\[
N_{s,\text{exp}} = \frac{c_{\text{max}}^2}{c_{\text{max}}} \frac{\Delta \rho (1-c_s) D_{\text{eff}}}{\mu c_s} \int d\bar{c}_s = \frac{Q^2}{2kh\phi^3 g\Delta S_o}
\]  

(2.33)

Figure 2.7 represents the back-calculated \(N_{s,\text{exp}}\) values in the drainage height range of 7.5-247 cm. As seen in this figure, the \(N_{s,\text{exp}}\) values derived from the study by Oduntan et al. (2001) show virtually no dependency on drainage height, and are almost constant values. However, the \(N_{s,\text{exp}}\) values for other experiments show relatively similar dependency on the drainage height for specific permeability and capillarity.

The back-calculated \(N_{s,\text{exp}}\) values for experimental studies conducted at different permeabilities are presented in Figure 2.8. As shown, the \(N_{s,\text{exp}}\) values in higher permeabilities (>100 D) show almost no dependency on the permeability, and they are relatively constant values. However, in the lower permeabilities, the capillarity effect can be seen as a deviation from trend of the \(N_{s,\text{exp}}\) seen at higher permeabilities. The \(N_{s,\text{exp}}\) values decreased by almost three orders of magnitude as the permeability was reduced from 640 to 16 D.
The back-calculated $N_{s,exp}$ values for different studies in sand packs and Hele-Shaw cells using different solvents and oils are presented in Figure 2.9. As shown, the range of variation of $N_{s,exp}$ for different solvents-oil systems (i.e., propane, butane, carbon dioxide, carbon dioxide-propane, methane-propane, and complex solvent) are in the range of $8.00 \times 10^{-7}$ to $2.75 \times 10^{-2}$. This figure clearly shows the interplay between drainage height, capillary forces, permeability, and the solvent type on $N_{s,exp}$ values. As seen, the combination of these parameters can introduce changes in the Vapex dimensionless number that are up to five orders of magnitudes.

![Figure 2.7: Effect of drainage height on back-calculated $N_{s,exp}$ values in reviewed studies (This study)](image)

Because of the complexity of mass transfer phenomena in the Vapex process and the lack of full characterization of the porous media and phase behavior data, it is not possible to determine the effect of solvent type on the performance of the process accurately.
Figure 2.9 also compares the $N_{s,exp}$ values in Hele-Shaw cell and sand packs. As seen in this figure, the back-calculated $N_{s,exp}$ values in the Hele-Shaw model are lower than the observed values in the sand pack at higher permeabilities ($>100$ D). The $N_{s,exp}$ values in Hele-Shaw cell vary between $8.00 \times 10^{-7}$ and $2.95 \times 10^{-5}$.

![Figure 2.8: Effect of permeability on back-calculated $N_{s,exp}$ values in reviewed studies (This study)](image)

2.7. Scale-Up Factors

Scale up factors for Vapex production can be used for the projection of important production parameters from laboratory-scale physical models to field scale. These factors can be derived from the dimensionless analysis of fundamental equations governing the process. On the basis of mass and momentum balances, Lim et al. (1996) derived the
following scaling factors for geometry, time, and production rate in terms of pore volume fraction per unit time.

\[
\frac{\kappa_f}{\kappa_m} = \left[ \frac{\mu D}{\rho g K} \right]_f \left[ \frac{\mu D}{\rho g K} \right]_m^{-1}
\]  
(2.34)

\[
\frac{t_f}{t_m} = \left[ D\phi \left( \frac{\mu}{\rho g K} \right)^2 \right]_f \left[ D\phi \left( \frac{\mu}{\rho g K} \right)^2 \right]_m^{-1}
\]  
(2.35)

\[
\frac{q_f}{q_m} = \left[ \frac{1}{fD} \left( \frac{\rho g K}{\mu} \right)^2 \right]_f \left[ \frac{1}{fD} \left( \frac{\rho g K}{\mu} \right)^2 \right]_m^{-1}
\]  
(2.36)

Figure 2.9: Effect of permeability and solvent composition on back-calculated \( N_{v,exp} \) values in reviewed studies (This study)

Butler et al. (1998, 2000) derived and used the scaling factors with the following assumptions:
a. A physical model and the reservoir in the field have a similar geometry, vapor chamber, and well configuration.

b. The model and the reservoir have the same porosity and oil saturation.

c. The oil used in the model has the same properties as those in the reservoir.

d. The flow of oil without dissolved solvents is negligible, and the less-viscous oil drains by gravity.

These scaling factors are as follows:

Considering the cumulative oil production proportional to the reservoir volume,

\[
\frac{Q_m}{Q_f} = \frac{h_m^2 L_m}{h_f^2 L_f}
\]

(2.37)

For an equal value of dimensionless number \(Dt/h^2\) with the same diffusivity \(D\), the ratio of total production time of the field reservoir to the physical model is:

\[
\frac{t_f}{t_m} = \frac{h_f^2}{h_m^2}
\]

(2.38)

For different permeabilities \(K_m\) and \(K_f\) of the model and the field, Equation 2.26 yields the following relationship for the oil production rates per unit of well length:

\[
\frac{Q_f}{Q_m} = \sqrt{\frac{k_f h_f}{k_m h_m}}
\]

(2.39)

However, recent experimental studies suggested the following correlation for upscaling the experimental results to field-scale condition (Karmaker and Maini, 2003, Yazdani and Maini, 2008).
\[ \frac{Q_f}{Q_m} = \left( \frac{h_f}{h_m} \right)^n \sqrt[4]{\frac{k_f}{k_m}} \]  

(2.40)

where \( n \) is between 1.1 and 1.3 and incorporates the effect of increased dispersion induced by the increase in drainage height.
CHAPTER 3: EXPERIMENTAL SETUPS & PROCEDURES

The experimental part of this study is comprised of a combination of phase behavior, porous medium characterization, and Vapex experimental studies. This chapter provides a detailed description of experimental setups and procedures used to perform a comprehensive experimental program designed for investigating the effect of selected parameters (i.e., drainage height, capillarity, and presence of aqueous phase) on mass transfer rates in the Vapex process.

3.1. Phase Behavior Study

In this section, the experimental approaches for characterization of phase behavior properties of Plover Lake heavy oil-butane system are presented. The data were later used for developing phase behavior models required for analytical and numerical modeling of the experiments.

3.1.1. Molecular Weight

Molecular weight is required for characterization of heavy ends of the oil phase and developing a phase behavior model using EOS-based approach (Danesh, 1998). In this study, vapor pressure osmometer technique (U.I.C Inc. Model 833 VPO) was used in order to measure the molecular weight of the Plover-Lake dead oil sample. This technique is based on differential vapor pressure between a pure solvent and a solution that can be converted to a measurable temperate difference. vapor pressure osmometer technique can be used to measure the molecular weight in the range of 100 to 2500 g/mole.
The vapor pressure osmometer setup was comprised of a measuring chamber containing two thermistors, a solvent reservoir, and two wicks to saturate the chamber with solvent vapor (i.e., toluene). The two thermistors were connected electrically to a bridge circuit and the bridge unbalance was measured by a precision closed loop electronic control system. Figure 3.1 shows the schematic of the vapor pressure osmometer setup.

Figure 3.1: Schematic of vapor pressure osmometer setup (UIC Model 833 VPO)

The bridge was balanced with toluene on both thermistors. The toluene was then replaced by a small sample (0.25 ml) of Plover Lake oil-toluene solution at different concentrations (i.e., 1.0-6.5 g/l). Because of the lower vapor pressure of the solution, the result was a temperature difference that could be measured as a bridge unbalance. Through calibration of the setup with toluene, molecular weight was determined from the amount of the bridge unbalance signal.

The basic theories in molecular weight measurement using vapor pressure osmometer are Rault’s Law (Equation 3.1) and Clausius-Clapeyron equation (Equation 3.2).

\[ P_i = P_i^0 x_i \]  \hspace{1cm} (3.1)

\[ \frac{dP}{dT} = \frac{P \Delta H_v}{RT^2} \]  \hspace{1cm} (3.2)
Based on assumption of ideal gas behavior of the vapor phase in solvent chamber (at 0.1 MPa and 20 °C) and manipulation of Equations 3.1 and 3.2, the molecular weight of solute can be determined by Equation 3.3. In this Equation, the molecular weight of the solute should be calculated by regression analysis of the measured voltage changes for various concentrations of toluene-oil solutions. The calibration factor in this equation should be developed for a solvent (i.e., toluene) with known molecular weight.

\[
MW = \frac{K\omega}{\Delta V}
\]  
(3.3)

3.1.2. SARA Analysis

In a heavy oil (usually defined as oil with an API gravity below 20°, which is equivalent to a density greater than 934.9 kg/m³), the C\textsubscript{7}-plus fraction dominates. This fraction is not amenable to conventional simulated distillation, because a significant part of it boils at temperatures too high to be measured and decomposes too quickly from the heat. Theoreticians have developed correlations for predicting the boiling-point distribution into this temperature range, but even this approach falls short, because each carbon number is assumed to be a straight-chained paraffin (alkane). Paraffins comprise only a moderate fraction of a heavy oil (Freitag et al.2008).

SARA analysis is a method for characterization of heavy oils and bitumen based on fractionation, whereby a heavy oil sample is separated into fractions having different compositions (i.e., saturates, aromatics, resins, and asphaltenes). SARA characterization is based on adsorption fractionation chromatography. Each fraction consists of a solubility class containing a range of different molecular-weight species. Characterization of the individual SARA fractions can assist in providing a map of the crude oil and its behavior and a base for comparison between the crudes.
In order to determine the possibility of asphaltene deposition in conducted Vapex experiments in this study and better characterization of the oil phase, a SARA analysis was performed on a Plover Lake dead oil sample used in this study. The SARA fractions of Plover Lake oil sample was determined by SRC’s developed method (Freitag and Exelby 2006). In addition, numerous asphaltene content measurements were carried out during the course of experimental program to monitor the deposition of asphaltene in the models. In these measurements, normal pentane was used to determine the asphaltene content of the produced oil (Freitag and Exelby 2006). Comparison between asphaltene content of dead oil used for saturating the models and produced oil samples provided a quantitative understanding on the amount of the possible asphaltene deposition happening at various stages of conducted Vapex experiments.

3.1.3. PVT Apparatus

The primary feature of the PVT apparatus was the mixing mechanism to achieve vapor-liquid equilibrium. The mixing mechanism consisted of two high-pressure computer-controlled syringe pumps. In this apparatus, the pumps were slightly tilted to make sure that the gas phase was going to be completely transferred into the receiving cylinder before the liquid. The pumps were connected by a low diameter high pressure piping that brought the oil into contact with the vapor phase to reach the equilibrium. In preparations for the phase behavior tests, the content of the receiving cylinder was maintained at setpoint pressure while the delivery cylinder injected the fluid into the receiving pump at a constant rate. This process was reversed when there were few cubic centimeters of fluid left in the delivering pump. Due to reciprocal operation, the dead volume in the apparatus was negligible. The schematic of the PVT apparatus is shown in Figure 3.2.
There were four instrument modules in the apparatus that could measure the properties of the fluid during or after mixing. These modules were a viscometer loop, a capillary sight glass, a vibrating-tube densitometer, and a pair of filters, of which all the wetted parts were connected by a type-316 stainless steel. The design of the setup made it possible to measure all properties of a sample in one apparatus. This approach minimized the error due to random scatter and other inconsistencies associated with attempts to duplicate experimental conditions in separate equipments.

![Diagram of PVT apparatus](image)

**Figure 3.2: Schematic of PVT apparatus (after Freitag *et al.*, 2005)**

There were two filters placed in series upstream of the densitometer, which normally required that the flow to be in the same direction through the modules. These filters removed any potential organic precipitates (e.g., asphaltenes) before the fluid entered into the densitometer. Using this approach, the density of the fluid was measured accurately to monitor density changes and potential compositional change at different stages of mixing or detect the presence of any additional phases through the mixing process.

The viscosity measurement was conducted in a coiled tubing equipped with a differential pressure transducer. This module was calibrated against a commercial certified viscosity standard. A collection of tubes with different lengths and diameters were available to
achieve appropriate pressure drop required for accurate viscosity measurement in a wide range of viscosities.

The solubility (i.e., gas/oil ratio) of volatile solvents was measured by flashing of a sample of fluid to ambient conditions, and measuring both the volume of the flashed gas and the weight of the remaining liquid. The flashed gas composition could be determined by gas chromatography, if necessary.

The precipitated asphaltenes during mixing process were separated by forcing the equilibrated mixture through a 7-micron filter then a 2-micron filter. The precipitant could then be recovered with nearly quantitative accuracy by flushing the filter cake with solvent, and then dissolving the remaining asphaltenes in toluene. The asphaltene content was determined by evaporation of the toluene, and determination of the weight of the asphaltene.

Temperature of the PVT setup was controlled with precision of 0.1 ºC by hosting the pump cylinders and the measurement modules in a circulation-air bath, in which both heaters and a cooling coil had been installed. Pressures were measured within 0.1% of full scale with a pressure transducer with a maximum pressure rating of 3.44 MPa. The operating range of the PVT apparatus is summarized in Table 3.1.

3.2. Porous Media Characterization

3.2.1. Sand Pack Selection

In order to systematically study the capillarity effect on mass transfer phenomena in the Vapex process, it was required to find different grain size distributions producing sand packs with relatively similar permeabilities and porosities, but different capillarities. In addition, the produced sand packs should have characteristics similar to a heavy oil
formation to replicate the mass transfer phenomena in the Vapex process at reservoir condition.

**Table 3.1: Operational range of the PVT apparatus**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Min</th>
<th>Max</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sample volume, cm³</td>
<td>100</td>
<td>550</td>
</tr>
<tr>
<td>Pressure, MPa</td>
<td>0.1</td>
<td>34.5</td>
</tr>
<tr>
<td>Temperature, ºC</td>
<td>5</td>
<td>80</td>
</tr>
</tbody>
</table>

For this purpose, a sand sample from Aberfeldy field (Lloydminster area) was analyzed to determine the grain size distribution, porosity, and permeability. The measured grain size distribution for Aberfeldy sand was then used as a reference in searching for appropriate grain size distributions for the experimental program. Various sands from different suppliers were tested to determine their grain size distributions, porosities, and permeabilities. Finally, three grain size distributions with enough supply for completing the designed experimental program were selected. Figure 3.3 shows the grain size distribution of the selected sands. This figure also compares the grain size distributions of selected sands with Aberfeldy sand. Figure 3.4 compares the shape of the sand gains for the selected sands for experimental program. As seen in these figures, there are noticeable differences between the shape and sortness of the sand grains in the different sand packs. Further to qualitative nature of these measurements for characterization of capillary forces, no additional statistical or image analysis was conducted on microscopic photographs. Instead, robust data on pore size distribution and capillary pressures obtained from mercury injection tests carried out on selected sands.
In preliminary characterization of porous media, porosity and permeability of various sand samples including three selected sands were measured. For this purpose, the sand was packed in a lead sleeve with diameter of 2.54 cm and average length of 20 cm by attaching a vibrator to the sleeve. In order to replicate the characteristics of the sand pack within the physical models in Vapex experiments, the packing process was carried out in absence of liquid phase. Two 200 US mesh screens were placed at each end of the sand.
pack to avoid movement of sand particles during permeability measurement. Then, the sand pack was placed in a core holder.

The core holder was equipped with a differential pressure transducer to measure the pressure difference between the inlet and outlet of the sand pack. The overburden of core holder was pressurized to 1 MPa to ensure there is no space between the lead sleeve and sand pack that could introduce error in permeability measurements. Finally, distilled water was injected into the sand pack using a liquid metering pump. After reaching a stabilized pressure drop across the sand pack at a constant injection rate, the pressure differential was recorded. Permeability measurements were repeated at five different rates to ensure the measured permeabilities were repeatable. Darcy’s law was used to calculate the permeability of each sand pack based on measured pressure difference across the sand pack, injection rate of distilled water, and the geometry of each sandpack. Table 3.2 summarizes the average characteristics of sand packs used in experimental study of this project.

**Table 3.2: Average characteristics of sand packs used in this study**

<table>
<thead>
<tr>
<th>Sand</th>
<th>Permeability (D)</th>
<th>Porosity (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sand#1</td>
<td>6.46</td>
<td>37.1</td>
</tr>
<tr>
<td>Sand#2</td>
<td>5.19</td>
<td>30.3</td>
</tr>
<tr>
<td>Sand#3</td>
<td>5.62</td>
<td>35.6</td>
</tr>
</tbody>
</table>

**3.2.2. Capillary Pressure**

In order to measure the capillary pressure of sand packs, cleaned and dried sand samples were shipped to Weatherford laboratories in Calgary. In these measurements, the samples
were mounted with plastic shrinkable tubings and nylon screens on the ends. Each sand pack was then compacted by providing heat on the shrinkable tube to obtain samples with approximate 2.54 cm in diameter and 2.54 cm in length sand packs. This approach minimized the compaction of the prepared sand packs occurs in traditional approach using aluminum foil at higher pressure (> 6.5 MPa).

The prepared sand pack was then placed in a penetrometer and loaded into an automated mercury injection instrument (Micromeritics Autopore 9220) with a maximum mercury intrusion pressure of 414 MPa. This was equivalent to a pore size distribution curve with pore throat size to as low as 0.0036 microns. The sand pack was evacuated under vaccum. Mercury was then injected at multiple pre-determined pressure levels up to 414 MPa. At each equilibrium pressure, the volume of mercury intrusion was determined by the change in capacitance of the penetrometer reference cell.

Capillary pressure is a function of the fluid properties and of the saturation. For a given rock-fluid system and saturation history, the saturation at a given capillary pressure is function of pore geometry. If the porous media is conceived to be a bundle of capillary tubes of various radii, capillary pressure-saturation curve relates to the number and size of pores penetrated by the non-wetting fluid (i.e., mercury) at a give capillary pressure. Burdine et al. (1950) developed a method to determine pore size distribution of a porous media using mercury capillary pressure curves. In this method, a distribution function is defined as \( D(r_i) \), therefore, the total volume can be expressed by Equation 3.4.

\[
dV_p = D(r_i)dr
\]  

(3.4)

The distribution function can be computed from the mercury capillary pressure data by using Equations 3.5 and 3.6. On the basis of these equations, the distribution function can
be graphically evaluated by taking slopes of the capillary pressure curve at different values of mercury saturation, computing the pore radius from the capillary pressure corresponding to the point at which the slope was taken (Equation 3.5) and evaluating Equation 3.6 for the distribution function (i.e., $D(ri)$).

$$P_{ci}r_i = 2\sigma_i \cos \theta_i$$  \hspace{1cm} (3.5)

$$D(r_i) = \frac{P_{ci} \frac{dS_m}{r_i}}{dP_e}$$  \hspace{1cm} (3.6)

Washburn (1921) and Rootar and Prenzlow (1967) developed different approaches to compute pore surface area using mercury injection. Washburn (1921) used the bundle of tube concept and cylindrical geometry to determine the pore surface area in a porous medium. Rootar and Prenzlow (1967) used the concept of reversible work required to immerse a surface in mercury. Despite the difference between these approaches, two computational approaches resulted in identical relationships.

Washburn (1921) used the basic relationship describing the penetration of mercury into a cylindrical pore of diameter $D$ derived from equating the applied pressure to the resisting surface tension (Equation 3.7) to develop a relation for estimating the pore surface area.

$$PD = -4\sigma_i \cos \theta_i$$  \hspace{1cm} (3.7)

The relationship between wall area, diameter, and volume for a cylindrical shape can be expressed as Equation 3.8. Combining Equations 3.7 and 3.8 yields Equation 3.9.

$$A_p = \frac{4V_p}{D_p}$$  \hspace{1cm} (3.8)

$$A_p = -\frac{PV_p}{\sigma_i \cos \theta_i}$$  \hspace{1cm} (3.9)
This equation can be written in the form of Equation 3.10 to evaluate pore surface area of porous medium from pressure-volume mercury penetration data.

\[
\sum \Delta A_p = -\sum \frac{P \Delta V_p}{\sigma_i \cos \theta_i}
\]  

(3.10)

The details of mercury injection tests on each sand sample are summarized in Tables 3.3 to 3.5. Measured pore size distributions of sand samples are presented in Figures 3.5 to 3.7. The air-water capillary pressures of sand samples are shown in Figures 3.8 to 3.10. In these measurements, the air/mercury interfacial tension was 485 dynes/cm. The air/mercury contact angle was 130º. The air/water interfacial tension and contact angle were 72 dynes/cm and 0.0º. The air-water capillary pressures were calculated from mercury injection data by Equation 3.11 derived directly from Yung-Laplace equation.

\[
\frac{P_{cm}}{P_{cw}} = \frac{\sigma_w \cos \theta_{wa}}{\sigma_m \cos \theta_{ma}}
\]  

(3.11)

**Table 3.3: Mercury injection test Sample#1**

<table>
<thead>
<tr>
<th>Sample weight (g)</th>
<th>12.6</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sample porosity (fraction)</td>
<td>0.369</td>
</tr>
<tr>
<td>Grain density (g/cm³)</td>
<td>2.42</td>
</tr>
<tr>
<td>Total pore surface area (m²)</td>
<td>0.25</td>
</tr>
<tr>
<td>Median pore diameter (micron)</td>
<td>48.85</td>
</tr>
<tr>
<td>Threshold pressure (kPa)</td>
<td>3.7</td>
</tr>
<tr>
<td>Pore size distribution:</td>
<td></td>
</tr>
<tr>
<td>Macropores (&gt;3.0 micron)</td>
<td>100%</td>
</tr>
<tr>
<td>Mesopores (1.0-3.0 micron)</td>
<td>0.0</td>
</tr>
<tr>
<td>Micropores (&lt;1.0 micron)</td>
<td>0.0</td>
</tr>
</tbody>
</table>
**Table 3.4: Mercury injection test Sample#2**

<table>
<thead>
<tr>
<th>Sample weight (g)</th>
<th>12.7</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sample porosity (fraction)</td>
<td>0.369</td>
</tr>
<tr>
<td>Grain density (g/cm³)</td>
<td>2.66</td>
</tr>
<tr>
<td>Total pore surface area (m²)</td>
<td>31.07</td>
</tr>
<tr>
<td>Median pore diameter (micron)</td>
<td>44.14</td>
</tr>
<tr>
<td>Threshold pressure (kPa)</td>
<td>3.7</td>
</tr>
<tr>
<td>Pore size distribution:</td>
<td></td>
</tr>
<tr>
<td>Macropores (&gt;3.0 micron)</td>
<td>96.9%</td>
</tr>
<tr>
<td>Mesopores (1.0-3.0 micron)</td>
<td>0.9%</td>
</tr>
<tr>
<td>Micropores (&lt;1.0 micron)</td>
<td>2.2%</td>
</tr>
</tbody>
</table>

**Table 3.5: Mercury injection test Sample#3**

<table>
<thead>
<tr>
<th>Sample weight (g)</th>
<th>14.4</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sample porosity (fraction)</td>
<td>0.342</td>
</tr>
<tr>
<td>Grain density (g/cm³)</td>
<td>2.60</td>
</tr>
<tr>
<td>Total pore surface area (m²)</td>
<td>21.84</td>
</tr>
<tr>
<td>Median pore diameter (micron)</td>
<td>55.39</td>
</tr>
<tr>
<td>Threshold pressure (kPa)</td>
<td>3.6</td>
</tr>
<tr>
<td>Pore size distribution:</td>
<td></td>
</tr>
<tr>
<td>Macropores (&gt;3.0 micron)</td>
<td>97.2%</td>
</tr>
<tr>
<td>Mesopores (1.0-3.0 micron)</td>
<td>1.0%</td>
</tr>
<tr>
<td>Micropores (&lt;1.0 micron)</td>
<td>1.8%</td>
</tr>
</tbody>
</table>
Figure 3.5: Pore size distribution Sand#1

Figure 3.6: Pore size distribution Sand#2
Figure 3.7: Pore size distribution Sand#3

Figure 3.8: Derived air-water capillary pressure Sand#1
Figure 3.9: Derived air-water capillary pressure Sand#2

Figure 3.10: Derived air-water capillary pressure Sand#3
3.3. Vapex Experiments

3.3.1. Design of Physical Models

Two lab-scale slab-type physical models with different drainage heights (i.e. 24.5 and 47.5 cm) were assembled for this study. The models were rectangularly shaped 2D visual models with stainless steel frames. The specifications are given in Table 3.6. The front and back sides of these models were made of clear Plexiglas plates with a thickness of 5 cm. The main purpose of these transparent plates was to monitor the vapor-solvent interface movement in the model by using digital imaging techniques. To seal the apparatus, a pressure gasket was placed between the frame and each Plexiglas plate after which the Plexiglas plates were bolted to the steel frame. These models were designed to withstand pressures up to 1 MPa. To position the physical models at the different angles necessary for packing, cleaning, and saturating purposes, the physical models are mounted side by side on a portable stand. To monitor the pressure in each model independently and transfer the solvent vapor to the vapor chamber, pressure and injection ports for each model were located at the top of the model. The production ports were located at the bottom of the model and connected directly to the effluent collection system.

<table>
<thead>
<tr>
<th>Geometry</th>
<th>Small Model</th>
<th>Large Model</th>
</tr>
</thead>
<tbody>
<tr>
<td>Height, cm</td>
<td>24.5</td>
<td>47.5</td>
</tr>
<tr>
<td>Length, cm</td>
<td>20.0</td>
<td>38.0</td>
</tr>
<tr>
<td>Thickness, cm</td>
<td>5.0</td>
<td>5.0</td>
</tr>
<tr>
<td>Total volume, cm³</td>
<td>2450</td>
<td>9025</td>
</tr>
</tbody>
</table>
To experimentally simulate the mass transfer phenomenon in mixing zone on the edge of the vapor chamber in the Vapex process, the thickness (i.e., 5 cm) of each model was divided into two compartments. The first compartment had a thickness of 4.55 cm that accommodated the porous media that was saturated with the heavy oil. The second compartment was a solvent chamber with a thickness of 0.55 cm that was fully saturated with solvent vapor to simulate the vapor chamber in the Vapex process. A 120 US mesh sieve separated the two compartments without compromising the rate of solvent diffusion into the porous medium. To avoid collapsing of the porous media under its weight and/or potential change in the geometry of the compartments, the vapor chamber was packed with glass beads with an average diameter of 1 mm. This size of glass beads had a negligible capillary pressure and therefore, did not cause any significant change in the pattern of oil drainage over the course of each experiment (<20 days). A schematic and photographs of the models are shown in Figures 3.11 and 3.12, respectively. Figure 3.12 shows the porous media and solvent compartments of both physical models. As seen in this Figure, the design of the setup makes it possible to identify and track the oil-solvent interface by using transparent Plexiglas plates.

To eliminate any potential for pressure disturbance during collection of drained oil, effluent collection was conducted at the same pressure as the physical models into a graduated visual cell. To equalize the pressure in the collection system with the physical models, the collections cells were equipped with two ports. The first port was connected directly to the production port of the physical model. The second port connected to a bypass line tied to the solvent injection port at the top of each model. This approach eliminated the possibility of pressure disturbances affecting fluid production rates.
3.3.2. Experimental Procedure

Plover Lake oil from west-central Saskatchewan with a density of 0.983 g/cm³ and viscosity of 10,540 mPa·s was used in all experiments. Normal butane was utilized as solvent. Experiments were carried out at 200 kPa and 21 °C. The operating pressure was around 16 kPa below the dew point of butane to minimize the chance of injection of any liquid butane into the models.

Figure 3.11: Schematic of the physical models used for Vapex experiments; (a) front view, (b) side view of the physical model

Figure 3.12: Photographs of the fabricated physical models used for Vapex experiments; (a) solvent chamber (b) oil saturated porous media
In preparation for each experiment, the models and all the injection and production ports were cleaned with solvents (toluene and acetone) and dried with compressed air. In addition, the models were checked for pressure containment and their integrity. The model was then placed in a horizontal position. A 0.55 cm-thick plastic plate was placed on one side of the model to fill and secure the solvent chamber. A 120 U.S. mesh sieve that separated the compartments was placed behind the plastic sheet and the Plexiglas plate on the solvent side of the physical model was bolted to the frame. After the solvent chamber was secured, the model was rotated 180º to a horizontal position to fill the rest of depth of the model (i.e., 4.55 cm) with sand. The sand was packed under vibration using a vibrating pad. Finally, the other Plexiglas plate was sealed and bolted onto the frame to secure the sand pack.

To prepare the models for injection of oil into sand pack in experiments conducted in dry sand, the packed model was placed under vacuum for a few hours until an absolute pressure below 13 kPa was obtained. The oil was then injected into the model through the bottom production ports. During the period of saturating the models with oil, they were heated to 40 ºC in order to reduce the oil viscosity and achieve more uniform oil saturation across the model at lower injection rates. After uniform oil saturation was achieved, the models were cooled to the experimental temperature (21 ºC) over night. In the last stage of preparation process, the plastic sheet in the solvent chamber needed to be replaced with glass beads. For this purpose, the Plexiglas plate was removed to take out the plastic sheet. Then, the compartment was filled with glass beads and model was resealed.
In experiments conducted in presence of immobile water phase saturation, the model was saturated with brine (18,000 ppm NaCl solution) after vacuuming the model through the production ports at the bottom of the model at 21 ºC. Then the model was heated to 40 ºC overnight to reduce the viscosity of the oil and minimize the viscous fingering problem. Finally, the oil was injected through the top of the model to displace the water. The displacement process was conducted at low rates (0.5-5.0 cm³/min) to achieve a uniform displacement through the models.

The initial oil volume in the sand pack was determined from pump injection data and measurement of the effluent volume during saturating phase of each experiment. Subsequently, various connections were pressure tested to prepare for injection of solvent at 200 kPa. The effluent collection cells for each model were pressured up, using the bypass pressure line, to the desired experimental pressure and were connected to the models. Finally, the models were rotated and put into vertical position for starting of the test. The models were pressurized and checked for any potential leakage. As soon as the pressures of the models reached 200 kPa, the valves isolating the effluent collection system from the models were opened to collect any oil that drained from the porous media. The pressure and temperature of the models were monitored and recorded with LabView™. The volume of effluents from each model were also monitored and recorded. In addition to the volume of the produced oil, the solvent content of the oil was measured at various stages of tests through measurement of the gas-oil ratio (GOR) of the produced oil and the density of the dead oil. Due to the slow nature of the solvent-leaching gravity drainage mechanism and the amount of initial oil in these models, each of these experiments took between 16 to and 46 days to complete.
3.3.3. Interface Tracking

Interface tracking during the course of experiments was carried to monitor the growth of the solvent chamber in each experiment. For this purpose, the oil-solvent interface was traced using high resolution photographs taken from the models by a Sony Cyber-shot camera (i.e., 2592×1944 pixels) at different stages of the test. The collected photographs were adjusted and imported into Surfer 8.00 program to extract the interface locations. The interface advancement velocities were then determined from the digitized interface profiles, which were plotted against time at specific vertical locations.
CHAPTER 4: EXPERIMENTAL RESULTS

In this chapter, the results of phase behavior measurements and conducted Vapex experiments are presented, analyzed, and discussed. The phase behavior study included PVT measurements of Plover Lake oil/n-butane system at different solvent concentrations (i.e., 0-61.32 mol %), molecular weight measurement, and asphaltene and solvent content of the produced oil samples at different stages of the Vapex experiments. The Vapex experimental program included 18 experiments, which were performed in two 2D physical models with drainage heights of 24.5 and 47.5 cm. These experiments were conducted in three sands with different grain size distributions in permeability range of 5.19 to 6.46 D to investigate the interplay of drainage height, capillarity, and aqueous phase on efficiency of mass transfer in the Vapex process.

4.1. Phase Behavior Experiments

In order to gather required information for simulation of mass transfer phenomenon in the Vapex process a comprehensive phase behavior experiments was required. Therefore, a phase behavior study was designed and conducted, in parallel with the Vapex experiments, to measure: phase behavior of Plover Lake oil-n-butane, molecular weight of Plover Lake oil, SARA fractions, and carbon number distribution of the oil phase.

4.1.1. Molecular Weight Measurement

Molecular weight is an important parameter that is required for determining the effect of intermolecular interactions on physical properties of the oil phase. Vapor pressure osmometry technique was applied to measure the molecular weight of Plover Lake oil...
sample used in this study. Molecular weight measurement was conducted in Vapor Pressure Osmometer Model 833 using toluene as solvent.

In the performed measurements, the calibration curve was determined for sucrose octaacetate solution in toluene with molecular weight of 678.59 g/gmol based on measurements conducted at five different concentrations in the range of 1 to 5 g/l. The molecular weight of Plover Lake oil sample was measured using four diluted Plover Lake oil samples by toluene (i.e., 1.32, 2.33, 4.03, 6.49 g/l) and generated calibration curve at 50 °C. The measured molecular weight of Plover Lake oil used for saturating the physical models in experimental part of this study was 469.

4.1.2. SARA Analysis

In situ upgrading of the produced oil because of deasphalting process is important phenomena that can result in significant changes in the properties of the produced oil phase. Asphaltene deposition in porous media can also change the pore morphology, wettability, and permeability of the medium. In order to investigate the possibility and extent of asphaltene deposition in designed experimental program a complete SARA analysis was conducted on dead oil sample. In addition, numerous asphaltene content measurements were conducted on produced oil samples. The results of SARA analysis conducted on Plover Lake oil sample is presented in Table 4.1. The SARA fractions in this table were measured using Saskatchewan Research Council (SRC) analysis method.

To monitor the potential deposition of asphaltene in the Physical models during the course of Vapex experiments, different samples were collected and analyzed for asphaltene content. These samples were centrifuged at 1500 rpm for 30 min to remove any trace of brine and sand particles. The asphaltene content of collected oil samples
from different experiments are compared with the asphaltene content of the dead oil used for saturating the models in Figure 4.1. The samples in Figure 4.1 were collected at various stages of different Vapex experiments to provide a quantitative estimation of asphaltene deposition in Vapex experiments.

Table 4.1: SARA fractions of Plover Lake oil

<table>
<thead>
<tr>
<th></th>
<th>Saturates (wt %)</th>
<th>Aromatics (wt %)</th>
<th>Resins (wt %)</th>
<th>Asphaltenes (wt %)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>29.3</td>
<td>34.4</td>
<td>23.4</td>
<td>12.9</td>
</tr>
</tbody>
</table>

Figure 4.1: Comparison between the asphaltene content of dead oil and collected samples of produced oil

As shown in Figure 4.1, the measured asphaltene content of the oil samples fluctuates around the measured value in SARA analysis (±1 wt%). In addition to the asphaltene content measurements on collected oil samples, conducted phase behavior study on
Plover Lake oil/n-butane system at wide range of n-butane concentration (i.e., 0-61.3 mol%) did not show any indication of asphaltene deposition. These measurements provided strong evidence on absence of any asphaltene deposition in sand pack during the Vapex experiments.

4.1.3. PVT Experiments

The composition of Plover Lake oil was determined using high temperature simulated distillation (HTSD) technique that separates the components of a stable oil in the order of their boiling points. The separation is accomplished with a non-polar chromatography column using a gas chromatograph equipped with programmable temperature oven and a flame ionization detector (FID). The chromatograph elution times of the hydrocarbons were calibrated to the atmospheric equivalent boiling point of paraffins reference material. The measured carbon number distribution of the Plover Lake oil is shown in Figure 4.2. Conducted HTSD showed the Plover Lake oil did not have any fractions lighter than C_9.

To determine the phase behavior properties of Plover Lake oil/n-butane system, a separate phase behavior study was designed and conducted to measure different physical parameters (i.e., density, viscosity, and saturation pressure) at different solvent concentrations. These measurements were carried out at experimental temperature (21 °C) and pressure range of 200-3000 kPa.

Viscosity and density variation of Plover Lake oil with pressure is listed in Table 4.2. Isothermal compressibility coefficient of Plover lake oil was 5.936×10^{-7} kPa^{-1}. This parameter was derived from the density measurements conducted at various pressures. Figure 4.3 illustrates the effect of pressure on density of the oil phase.
Figure 4.2: Carbon number distribution of Plover Lake oil

Table 4.2: Properties of Plover Lake oil

<table>
<thead>
<tr>
<th>Pressure (kPa)</th>
<th>Temperature (°C)</th>
<th>Oil Density (g/cm³)</th>
<th>Viscosity (mPa·s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>200</td>
<td>21</td>
<td>0.9835</td>
<td>10541</td>
</tr>
<tr>
<td>500</td>
<td>21</td>
<td>0.9836</td>
<td>10571</td>
</tr>
<tr>
<td>1000</td>
<td>21</td>
<td>0.9839</td>
<td>10874</td>
</tr>
<tr>
<td>2000</td>
<td>21</td>
<td>0.9845</td>
<td>11519</td>
</tr>
<tr>
<td>3000</td>
<td>21</td>
<td>0.9851</td>
<td>12054</td>
</tr>
</tbody>
</table>

Phase behavior properties of Plover Lake oil at various concentrations of n-butane is listed in Tables 4.3. These measurements were performed at 21 °C and 200 kPa, which was similar to the operating condition of the Vapex experiments. Effect of pressure on
viscosity and density of oil phase is measured for live oil sample with n-butane concentration of 38.90 mol%. The results of these measurements are listed in Table 4.4.

![Graph showing density variation with pressure](image)

**Figure 4.3: Density variation of Plover Lake oil with pressure at 21 °C**

According to these measurements, the isothermal compressibility coefficient of the oil phase was $6.41 \times 10^{-7}$ kPa$^{-1}$. Figure 4.4 illustrates the effect n-butane concentration on oil/solvent viscosity. The measured viscosities are compared with the viscosity predictions using Kendal and Monroe (1917) correlation (Equation 2.6). The average absolute error of viscosity predictions using Equation 2.6 is 19.93%. The dependency of Plover Lake oil density on concentration of n-butane is shown in Figure 4.5. Figure 4.5 also includes the density predictions using a simple correlation developed on the basis of assumption of no volume change due to the mixing of solvent and oil and ignoring the
gas phase density with respect to the liquid phase (Equations 4.1 and 4.2). The average absolute error of predictions using Equation 4.2 was 7.46%.

\[
\rho_{\text{mix}} = \rho_o - C_s (\rho_o - \rho_s) \tag{4.1}
\]

\[
\rho_{\text{mix}} = 0.9835 - 0.4172C_s \tag{4.2}
\]

**Table 4.3: Plover Lake/n-butane system at 21°C and 200kPa**

<table>
<thead>
<tr>
<th>Butane Concentration (Mole fraction)</th>
<th>Density (g/cm³)</th>
<th>Viscosity (mPa·s)</th>
<th>Saturation Pressure (kPa)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.0000</td>
<td>0.98353</td>
<td>10541</td>
<td></td>
</tr>
<tr>
<td>0.2011</td>
<td>0.9657</td>
<td>2100</td>
<td>258</td>
</tr>
<tr>
<td>0.3890</td>
<td>0.9365</td>
<td>313</td>
<td>310</td>
</tr>
<tr>
<td>0.5797</td>
<td>0.8921</td>
<td>43</td>
<td>315</td>
</tr>
<tr>
<td>0.6134</td>
<td>0.8798</td>
<td>25</td>
<td>325</td>
</tr>
</tbody>
</table>

* In saturation pressure measurements of heavy oil-solvent systems at low pressures, SRC’s experience indicates substantial error in these measurements.

**Table 4.4: Plover Lake oil/n-butane (38.90 mol %)**

<table>
<thead>
<tr>
<th>Pressure (kPa)</th>
<th>Temperature (°C)</th>
<th>Oil Density (g/cm³)</th>
<th>Viscosity (mPa·s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>200</td>
<td>21</td>
<td>0.9365</td>
<td>313</td>
</tr>
<tr>
<td>500</td>
<td>21</td>
<td>0.9366</td>
<td>307</td>
</tr>
<tr>
<td>1000</td>
<td>21</td>
<td>0.9369</td>
<td>316</td>
</tr>
<tr>
<td>2000</td>
<td>21</td>
<td>0.9376</td>
<td>325</td>
</tr>
<tr>
<td>3000</td>
<td>21</td>
<td>0.93825</td>
<td>338</td>
</tr>
</tbody>
</table>
Figure 4.4: Density dependency of Plover Lake oil to n-butane concentration

Figure 4.5: Viscosity dependency of Plover Lake oil to n-butane concentration
4.2. Vapex Experiments

In order to investigate the effects of drainage height, capillarity, and aqueous phase on performance of the Vapex process, total of 18 experiments were designed and conducted. Among these experiments, 11 experiments could be successfully performed by achieving uniform oil saturation distributions in physical models. In addition to these experiments, 2 experiments were designed and conducted in both physical models (i.e., small and large) using Sand#1 to investigate the reproducibility of the measured drainage rates. The summary of the conducted experiments is given in Table 4.5. These experiments were carried out using Plover Lake heavy oil and instrument grade n-butane.

Table 4.5: Properties of conducted Vapex experiments

<table>
<thead>
<tr>
<th>Test No.</th>
<th>Sand No.</th>
<th>Pore Surface Area (m²/g)</th>
<th>Porosity (%)</th>
<th>Permeability (D)</th>
<th>Height (cm)</th>
<th>Sw (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1</td>
<td>0.019</td>
<td>37.1</td>
<td>6.46</td>
<td>24.5</td>
<td>0.00</td>
</tr>
<tr>
<td>2</td>
<td>1</td>
<td>0.019</td>
<td>37.1</td>
<td>6.46</td>
<td>47.5</td>
<td>0.00</td>
</tr>
<tr>
<td>3</td>
<td>2</td>
<td>2.446</td>
<td>30.3</td>
<td>5.19</td>
<td>24.5</td>
<td>0.00</td>
</tr>
<tr>
<td>4</td>
<td>2</td>
<td>2.446</td>
<td>30.3</td>
<td>5.19</td>
<td>47.5</td>
<td>0.00</td>
</tr>
<tr>
<td>5</td>
<td>3</td>
<td>1.517</td>
<td>35.6</td>
<td>5.62</td>
<td>24.5</td>
<td>0.00</td>
</tr>
<tr>
<td>6</td>
<td>3</td>
<td>1.517</td>
<td>35.6</td>
<td>5.62</td>
<td>47.5</td>
<td>0.00</td>
</tr>
<tr>
<td>7</td>
<td>1</td>
<td>0.019</td>
<td>37.1</td>
<td>6.46</td>
<td>24.5</td>
<td>12.92</td>
</tr>
<tr>
<td>8</td>
<td>1</td>
<td>0.019</td>
<td>37.1</td>
<td>6.46</td>
<td>47.5</td>
<td>12.62</td>
</tr>
<tr>
<td>9</td>
<td>2</td>
<td>2.446</td>
<td>30.3</td>
<td>5.19</td>
<td>24.5</td>
<td>13.42</td>
</tr>
<tr>
<td>10</td>
<td>2</td>
<td>2.446</td>
<td>30.3</td>
<td>5.19</td>
<td>47.5</td>
<td>13.18</td>
</tr>
<tr>
<td>11</td>
<td>3</td>
<td>1.517</td>
<td>35.6</td>
<td>5.62</td>
<td>24.5</td>
<td>13.15</td>
</tr>
</tbody>
</table>
Despite all the efforts for performing the Vapex experiment in Sand#3, a uniform saturation distribution of oil phase was not achieved in the large model. Lack of enough of supply for the sand specifications similar to Sand#3 made it impossible to investigate the effect of aqueous phase at drainage height of 47.5 cm.

4.2.1. Results of Vapex Experimental Program

The production histories of conducted experiments in different sand packs are shown in Figures 4.6 to 4.11. These figures show the effect of studied parameters on drainage rates and achieved recovery factors at various stages of each Vapex experiments.

In this experimental program, high drainage rates were observed at early stages of the experiments (~24 hours). These higher production rates could be attributed to the drainage of the oil on the surface of the sieve separating porous media and solvent chamber. Beyond this time, drainage rates started to stabilize for all the experiments.

The other characteristic of observed drainage rates was the fluctuation of measured rates. These rate fluctuations occurred around the average stabilized drainage rate. These small-scale rate fluctuations could be attributed to the temperature change during the course of experiments (±1 °C), potential small-scale non-uniformities in the sand pack, and short period shut downs of effluent collection system for replacement of the collection cells. The other possibility could be the induced pressure fluctuations of the physical models due to opening and closing of the pressure regulator to maintain the pressure around the setpoint pressure of 200 kPa by injection of n-butane into the solvent chamber. Similar behavior was observed in previous Vapex study conducted by different researchers (Ayub and Tuhinuzzaman, 2007; Yazdani and Maini, 2005, 2008). Figure 4.12 shows a sample of pressure and temperature profiles in Experiments#1 and #2.
Figure 4.6: Oil drainage rate in Sand#1 at two drainage heights

Figure 4.7: Cumulative oil production in Sand#1
Figure 4.8: Oil drainage rate in Sand#2 at two drainage heights

Figure 4.9: Cumulative oil production in Sand#2
Figure 4.10: Oil drainage rate in Sand#3 at two drainage heights

Figure 4.11: Normalized cumulative oil production in Sand#3
In addition to observed fluctuations in measured drainage rates, some anomalies in drainage rates were observed in the Vapex experiments that were conducted in the large physical model. The sizes of these anomalies were larger than the fluctuations around the average drainage rates. An extreme example of this behavior happened in Experiment#2 where a 232% increase in the drainage rates was observed at 227 hours from the start of the experiment. This period was lasted for 49 hours when it followed by a decrease to stabilized drainage rate. The measured production rates in Experiments 2, 4, and 6 were conducted with drainage height of 47.5 cm are compared in Figure 4.13. These experiments were performed in absence of aqueous phase in the sand packs. As shown, the size of the anomaly in production rate in Sand#1 is more prominent than the observed anomalies in experiments conducted in Sand#2 and Sand#3. No drainage rate anomalies were observed in Vapex experiments performed at drainage height of 24.5 cm. This could
be as a result of drainage rate noise that could mask the small scale drainage rate anomalies. Possibility of partial plugging by sand or asphaltene deposition in these experiments was rejected after close investigation of production ports, continuous monitoring of the oil level in the solvent chamber, and asphaltene content measurement of produced oil samples.

Similar behavior in drainage rates was observed in studies carried out to investigate the imbibition recovery in fractured reservoirs using experimentally simulated single matrix block (Saidi, 1987; Babadagli, 2003). Investigation for understanding of microscopic phenomena responsible for this behavior showed that the interplay of capillary and gravity forces could result in a counter-current drainage process in Vapex experiments. In experiments with higher drainage heights, the balance of gravity and capillary forces is affected. As a result of this imbalancement of forces, imbibition process could overcome the drainage process and reduce the drained oil from the top of the model. This phenomenon continues until the reduction of viscosity and interfacial tension of the heavy oil, as a result of further dissolution of n-butane, mobilizes the trapped oil. This explanation could describe the absence of drainage rate anomalies in experiments conducted in presence of water phase where this immobile phase occupies the smaller capilariries because of water-wet nature of the sand. Figure 4.14 shows the comparison between drainage rates in Experiments#2 and #8 in Sand#1. Because of experimental limitations, it was not possible to repeat the Expriment#2 to further investigate the extent of drainage rate anomalies. However, as both experimental models in Experiment#1 and #2 were connected to the same solvent feed line, there is no evidence on presence of potential pressure disturbances or liquid solvent injection in Expriment#2.
Figure 4.13: Drainage rate comparison in absence of aqueous phase

Figure 4.14: Drainage rate comparison in absence and presence of aqueous phase
Measured n-butane contents of the produced oil samples in experiment#2 showed a sharp increase in n-butane concentration. The average n-butane concentration of the produced oils before the anomaly in production rates was 43.53, which increased to 54.49 mol%. This increase in n-butane concentration translated to 66% decrease in viscosity of the oil phase. This also resulted in change in specific volume of the oil phase from 1.0795 to 1.1111 cm³/g. The details of n-butane uptake at various stages of the conducted Vapex experiments is provided in Section 4.2.2.

To the best of author’s knowledge, there are no reported cases of such anomalies in Vapex literature. First, this could be attributed to unrealistically high permeabilities (i.e., >100 D) of previous experimental models used to study various aspects of the Vapex process. These high permeability packed models could not realistically model the interplay between the gravity and capillary forces occurring in the Vapex process. Second, the traditional approach for experimental simulation of the Vapex process is based on a constant rate injection of solvent vapor by establishing a constant pressure drop between injection and production ports (Das and Butler, 1998; Talbi and Maini, 2008; Yazdani and Maini, 2008). Therefore, it is expected to see no anomalies in production rates in higher permeability models where presence of pressure drop between the injection and production ports could induce a forced displacement that could mask the gravity drainage phenomenon on the edge of the vapor chamber at reservoir condition.

4.2.2. Solvent Mass Balance

The performance of the Vapex process is a direct function of efficiency of mass transfer between the solvent and heavy oil or bitumen. Therefore, it is necessary to determine and monitor the solvent content of the oil phase during the course of the experiments. In order
to measure the solvent dissolution rate in the oil phase, which is required for any mathematical purposes, there are two different experimental approaches (Yazdani and Maini, 2008, and Ahmadloo et al. 2010).

The first approach is based on continuous circulation of solvent vapor into the physical model by establishing a pressure difference between the injection and production ports (Yazdani and Maini, 2008). In this approach, the solvent uptake for each experiment is determined by measuring the difference between the volumes of injected and produced solvent, which is measured by a Wet Test Meter (WTM) after separation process. However, there are two major shortcomings in this approach:

1) In order to continuously circulate the solvent between the injection and production ports, it is required to have a constant pressure difference between these ports. Considering the typically low drainage heights of the Vapex experiments (< 40 cm), this pressure difference could be much higher the pressure difference by density difference, which could result in forced displacement of the oil phase by vapor phase and further mixing of the solvent and oil phase.

2) Application of traditional separation technique by solely depressurization of oil-solvent mixture to atmospheric pressure is not an accurate approach for solvent content measurement in heavy oil and bitumens. In these systems, oversaturation of the oil phase could occur because of high viscosity of the oil phase. Analysis of the collected oil samples in this study after separation using solely depressurization to atmospheric pressure showed solvent contents up to 23.84 mol%.

The second approach is the analysis of the produced oil sample for the solvent content, which was developed in this study. This approach could address the shortcomings of the
traditional approach. However, this experimental study requires a complete phase behavior study of solvent-oil system at observed concentration range in Vapex experiments.

In this study, n-butane contents of the collected live oil samples from models were measured at two steps. At the first step, the effluent collection cell was connected to a separator, which was tied to a WTM. Then, the pressure of collection cell and sample was reduced to ambient pressure to separate the dissolved gas from the produced oil. In order to maximize the separation efficiency through this process, a combination of depressurization, heating, and agitation of the effluent cell was applied. Next, the oil was cooled down to 21 °C and examined for any further weight loss by vaporization of n-butane by mentoring the weight of effluent and separator cells. Finally, the density of the oil phase was measured by an Anoton Paar DMA 38 densitometer. On the basis of these measurements, n-butane content of the oil phase was determined by calibration curve developed for Plover Lake oil-n-butane system. Figures 4.15 to 4.17 show the cumulative solvent uptake of the collected produced oil samples at various stages of each test for conducted Vapex experiments.

As seen in Figures 4.15 to 4.17, there is a similar trend between the cumulative volume of the produced oil and the solvent uptake of the oil phase as a result of mass transfer between solvent and oil phase. The analysis of the results showed relatively stable solvent content of the oil phase during the Vapex experiments depending on drainage height and presence, sand pack characteristics, and absence of the aqueous phase saturation in the experiments. This explains the linear relationship between the n-butane uptake versus time in these experiments in time periods stabilized oil drainage rates were observed.
Figure 4.15: Cumulative n-butane uptake in Sand#1

Figure 4.16: Cumulative n-butane uptake in Sand#2
4.2.3. Interface Advancement

In order to obtain the interface advancement rates, images of physical model were collected and processed for determining the interface advancement rate for each experiment. Figures 4.18 to 4.20 show the oil-solvent interface position during the course of experiments.

Figure 4.18 compares the interface rate movement in Vapex experiments conducted using Sand#1. The calculated interface advancement rates for experiments #1, #2, #7, and #8 are listed in Table 4.6. These rates were calculated by regression analysis of the interface positions for first 300 hours of the experiments where constant drainage rates were observed. As shown in Tables 4.6, the calculated interface advancement rate is a strong function of drainage height of porous medium. In Sand#1, with increase of drainage height of the physical models from 24.5 to 47.5 cm, the rate of interface advancement
was increased between 38 to 171%. The observed advancement rates in experiments carried out in absence and presence of connate water saturation are in comparable range in Sand#1.

Table 4.6: Interface advancement rate in Sand#1

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Drainage Height (cm)</th>
<th>Water Saturation (%)</th>
<th>Interface Advancement Rate (cm/h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>24.5</td>
<td>0</td>
<td>0.0327</td>
</tr>
<tr>
<td>2</td>
<td>47.5</td>
<td>0</td>
<td>0.0887</td>
</tr>
<tr>
<td>7</td>
<td>24.5</td>
<td>12.92</td>
<td>0.0479</td>
</tr>
<tr>
<td>8</td>
<td>47.5</td>
<td>12.62</td>
<td>0.0663</td>
</tr>
</tbody>
</table>

The interface advancement rates for experiments conducted using Sand#2 are summarized in Table 4.7. These rates were also determined during the first 300 hours of experiments where constant drainage rates were observed in these four experiments. Considering similar operational procedure in this experimental study, selection of data within first 300 hours of experiments for calculation of the rates also made the comparison between the rates at different experiments more reliable.

In Sand#2, there is a strong correlation between the interface advancement rate and drainage height and presence of connate water saturation. The interface advancement rate increased by 6.21 and 9.18 times with increase of drainage height from 24.5 to 47.5 cm in the two sets of experiments. Repeating the experiments in presence of connate water saturation considerably increased the interface advancement rates. This was equivalent to 2.32 and 1.57 times increase in interface advancement rates in small and large model.
Table 4.7: Interface advancement rate in Sand#2

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Drainage Height (cm)</th>
<th>Water Saturation (%)</th>
<th>Interface Advancement Rate (cm/h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>3</td>
<td>24.5</td>
<td>0</td>
<td>0.0071</td>
</tr>
<tr>
<td>4</td>
<td>47.5</td>
<td>0</td>
<td>0.0652</td>
</tr>
<tr>
<td>9</td>
<td>24.5</td>
<td>13.42</td>
<td>0.0165</td>
</tr>
<tr>
<td>10</td>
<td>47.5</td>
<td>13.18</td>
<td>0.1025</td>
</tr>
</tbody>
</table>

The comparison between the interface advancement rates for experiments conducted in Sand#3 is provided in Table 4.8. In Sand#3, similar trends with Sand#2 in effects of drainage height and connate water saturation on interface advancement rate was observed. In these series of experiments, the increase in drainage height from 24.5 to 47.5 cm increase resulted in 2.9 times increase in interface advancement rate in absence of aqueous phase. Comparison between the interface advancement rates for experiments conducted in absence and presence of connate water saturation at the same drainage height (i.e., 24.5 cm) revealed that the connate water has a significant effect on the efficiency of mixing and performance of the Vapex process.

Table 4.8: Interface advancement rate in Sand#3

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Drainage Height (cm)</th>
<th>Water Saturation (%)</th>
<th>Interface Advancement Rate (cm/h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>24.5</td>
<td>0</td>
<td>0.0081</td>
</tr>
<tr>
<td>6</td>
<td>47.5</td>
<td>0</td>
<td>0.0316</td>
</tr>
<tr>
<td>11</td>
<td>24.5</td>
<td>13.15</td>
<td>0.0315</td>
</tr>
</tbody>
</table>
The calculated interface advancement rate using image processing and regression analysis provides a clear indication on interplay of capillarity, drainage height, and connate water saturation on the performance of the Vapex process. Presented results in this section are in agreement with solvent content, drainage rate, and achieved recovery factors discussed in previous sections of this chapter. However, presence of small-scale heterogeneity in the sand packs, wall effects, and change of wettability of the visual Plexiglas in experiments performed in presence of connate water saturation could introduce complexity in detection of interface and monitoring the movement of interface. For instance, presence of small-scale heterogeneity in the sand pack could result in uneven movement of interface during the course of experiments. On this basis, the presented interface advancement rates in this section could just provide a reasonable estimation of the interface advancement rate during the course of Vapex experiments.

![Figure 4.18: Oil-solvent interface position at different times in Sand#1](image)
Figure 4.19: Oil-solvent interface position at different times in Sand#2

Figure 4.20: Oil-solvent interface position at different times in Sand#3
Figure 4.21 shows samples of collected images of the small experimental model used in image processing in order to determine the location solvent-oil interface.

Figure 4.21: Photographs of oil-solvent interface evolution in small physical model

4.3. Data Analysis

This section provides a preliminary data analysis to investigate the effects of capillarity, drainage height, and aqueous phase on performance of the Vapex process. In the experimental program of this study a total of 11 experiments were conducted using three different sands at two drainage heights. The effect of studied parameters on stabilized drainage rates, Vapex number, and calculated solvent penetration depth are presented in the following sections.

4.3.1. Effect of Permeability

According to the proposed mathematical model by Das and Butler (1998) presented in Equation 2.26, the stabilized drainage rate should be proportional to the square-root of combined terms of permeability and porosity of the medium ($k\phi^{\alpha}$) at a constant drainage height. Conducted Vapex experiments at high permeability sand packs by different researchers have confirmed the square root dependency of the stabilized drainage rates on
permeability of the medium (Oduntan et al., 2001; Karmaker and Maini, 2003; Yazdani and Maini, 2005, 2008).

Figures 4.22 and 4.23 compare the measured stabilized drainage rates for conducted Vapex experiments in absence and presence of the immobile water saturation. As shown in these figures, the measured stabilized drainage rates in this study display a clear deviation from the trend that was observed in experiments conducted at unrealistically high permeabilities (>100 D). This difference could be attributed to the effect of capillarity, as the only variable in these experiments at two drainage heights, on the rate of solvent-oil mixing in porous media. In addition to this behavior, the measured recovery factors for the various experiments showed a strong correlation with the measured specific pore surface area of the sands used in this study (see Section 4.3.3).

Comparison between the drainage rates in absence and presence of connate water saturation also revealed the important role of connate water saturation on stabilized drainage rates. The effect of water saturation was more pronounced in sand packs with higher specific pore surface area. For instance, conducted experiment in Sand#2 in presence of connate water saturation of 13.42% showed two orders of magnitude increase in stabilized drainage rates at the same drainage height. Conducted experiments in lower specific pore surface area showed minor role of the water saturation on stabilized drainage rates. It was postulated that presence of immobile water saturation in porous media reduces the capillarity effect by occupying of smaller pores in the water-wet sand pack and enhances the mass transfer and flow of the oil phase in the medium.
Figure 4.22: Variation of stabilized drainage rate with permeability ($S_w=0$)

Figure 4.23: Variation of stabilized drainage rate with permeability ($S_w>0$)
4.3.2. Effect of Drainage Height

According to Butler’s model, as given in Equation 2.26, the stabilized drainage rate in the Vapex process has a square root dependency on drainage height at a constant permeability. Measured stabilized drainage rates in this study did not show the proportionality with the square root of drainage height. Instead, they presented proportionality with the drainage height to the power of 1.1 to 1.3 for experiments in absence of connate water saturation. This is in agreement with reported stabilized drainage rates by Yazdani and Maini (2008). Yazdani and Maini (2008) attributed this behavior to further enhancement of mass transfer by mechanical dispersion at higher drainage heights. On this basis, they introduced the concept of additional height to explain the further mixing happening at higher drainage rates (Equation 2.40).

Figure 4.24 and 4.25 compare the predicted stabilized drainage rates for the large physical model (h=47.5 cm) based on measured rates in the small physical model (h=24.5 cm) using Equation 2.40. Calculated drainage rates for experiments conducted in absence of connate water saturation by Equation 2.40 showed a good agreement with the measured drainage rates in this study. However, the calculated drainage rate by Equation 2.40 for the large physical model using Sand#1 was overestimated. These experiments showed lower dependency of rates to drainage heights in Sand#1 in presence of the immobile water saturation. In order to check the accuracy of the measurements, Experiment#8 was repeated and almost similar drainage rate were observed in this experiment. In contrast with Expriment#8, the calculated stabilized drainage rate using Equation 2.40 in Experiment#10 was in good agreement with the measured drainage rates in the small and large physical models.
Figure 4.24: Predicted stabilized drainage rates by Equation 2.40 ($S_w=0.0$)

Figure 4.25: Predicted stabilized drainage rates by Equation 2.40 ($S_w>0.0$)
4.3.3. Effect of Capillarity

The pore surface area of a porous medium can be determined using mercury injection data based on Equation 3.10. According to this equation, there is a direct relationship between the capillary pressure of a porous medium and total pore surface area in the same medium. Considering this direct relationship, specific pore surface area of each sand pack in this study was used to investigate a potential relationship between observed stabilized drainage rates and capillarity.

The relationship between the stabilized drainage rates with specific pore surface area of porous media for three sand packs at two different drainage heights for experiments conducted in absence and presence of connate water saturation is presented in Figure 4.26. As seen in this figure, the stabilized drainage rates showed a strong dependency on specific pore surface area of the porous media at the same drainage heights and relatively similar permeability range.

The conducted Vapex experiments in absence of connate water saturation showed a continuous reduction of drainage rates with increase of specific pore surface area of the medium. This behavior can be attributed the important role of capillary forces that limit the efficiency of mass transfer phenomenon as it was observed by n-butane content measurements of the produced oil from the physical models.

Repeating the Vapex experiments in presence of connate water saturation with similar sand pack specifications and boundary conditions revealed that presence of aqueous phase could measurably improve the performance of the Vapex process and efficiency of the mass transfer rate and consequently the drainage rate in the Vapex process. This hypothesis was proved by n-butane content measurements of the produced oil samples.
collected during the course of experiments. Comparison between the stabilized drainage rates clearly present the interplay of studied parameters on the rate of mass transfer between the oil phase and solvent vapor.

![Graph showing variation of stabilized drainage rate with specific pore surface area of porous medium](image)

**Figure 4.26: Variation of stabilized drainage rate with specific pore surface area of porous medium**

The comparison between the stabilized drainage rates and achieved recovery factors in these experiments clearly revealed the positive impact of immobile water saturation on the efficiency of the mass transfer and the performance of the Vapex process. This finding showed some similarities with the experimental study conducted by Schmidt *et al.* (1982), Fisher *et al.* (2000, 2002), and Etminan *et al.* (2008) who observed higher diffusion coefficients or drainage rates in presence of immobile water stauration in their experimental studies.
4.3.4. Vapex Number

The Vapex number is a dimensionless number that is consisted of physical properties of the solvent-oil system and the effective mass transfer coefficient. This parameter can be back-calculated using Equation 2.33 using the properties of the sand pack and measured stabilized drainage rates from the conducted experiments. Figure 4.27 shows the effects of drainage height, capillarity, and connate water saturation on Vapex number in this study.

Back-calculated Vapex number in Figure 4.27 shows that presence of the immobile water saturation increases the Vapex number by up to two orders of magnitude.

Figure 4.27: Effect of drainage height, capillarity, and connate water saturation on Vapex number

In Figure 4.27, the lowest Vapex numbers belong to experiments conducted in small model with drainage height of 24.5 cm in absence of immobile water saturation. In these
experiments, the Vapex numbers varied between $2.68 \times 10^{-9}$ and $1.28 \times 10^{-6}$. With increase in drainage height to 47.5 cm the Vapex numbers increased in the range of $1.51 \times 10^{-8}$ to $1.17 \times 10^{-6}$. In presence of aqueous phase, the Vapex number at drainage height of 47.5 cm was almost constant value in the range of $1.17 \times 10^{-6}$ to $1.28 \times 10^{-6}$. This means that the effect of capillarity in both sands were similar because of invasion of water into the smaller pores.

Figure 4.28 shows the comparison between the Vapex number in this study and back-calculated Vapex number in some of Vapex studies in literature conducted in homogeneous sand packs to investigate the effect of permeability, drainage height, and solvent type on performance of the Vapex process.

![Figure 4.28](image)

Figure 4.28: Interplay of permeability, drainage height, and water saturation on Vapex number in this study and reviewed studies
As seen in this figure, the $N_{s,exp}$ values in higher permeabilities (>220 D) had almost no dependency to the permeability and they are relatively constant values at a constant drainage height. However, the Vapex numbers in the reported permeability range that was more representative of western Canadian heavy oil reservoirs were not constant. This comparison also suggests that the higher mass transfer rates in high permeability porous media could be explained by significant increase in surface area for diffusion in absence of capillary forces, and higher achieved velocities that enhanced the mass transfer rate between the solvent and oil phase.

### 4.3.5. Solvent Penetration Depth

The penetration depth of the n-butane into the heavy oil-saturated sand pack is the parameter that can be approximated by applying Darcy’s law to the gravity-driven flow and calculated average velocity of the thin layer of the draining mixture. According to the Darcy’s law the average velocity in the mixing zone can be estimated using Equation 4.3.

$$u = \frac{Q}{A} = \frac{k \rho_{mix} g \sin \theta}{\mu_{mix}}$$  \hspace{1cm} (4.3)

In Equation 4.3, which was obtained by manipulation of Darcy’s law, the thickness of mixing zone can be estimated by Equation 4.4 with assumption of negligible drainage rate of dead oil.

$$\frac{Q}{A} = \frac{Q}{L \xi_{max}}$$  \hspace{1cm} (4.4)

In Equation 4.4, it is assumed that the vapor chamber forms a constant angle $\theta$ and the cross-sectional area is the product of the penetration depth and the width of the physical model. Further manipulation of Equations 4.3 and 4.4 results in Equation 4.5 that can be
used for approximation of the penetration depth of n-butane in conducted Vapex experiments.

\[
\xi_{\text{max}} = \frac{Q \mu_{\text{mix}}}{k \rho_{\text{mix}} g \sin \theta}
\]  

(4.5)

In order to estimate the contact angle \( \theta \) required for the calculation of penetration depths, this parameter was estimated for each Vapex experiment using the stabilized drainage rate, measured sweep efficiency of the process, and position of the solvent-heavy oil interface. The contact angle values were determined using the cumulative production for the first 300 hours of the Vapex experiments where relatively constant drainage rates were observed for all the experiments. Figure 2.4 shows the schematic of the technique used to estimate the contact angle values. In this approach, it was assumed that the interface was growing homogenously across the cross section of the model during the course of each experiment. The contact angle could be approximated by using volume balance method in which the volume of produced oil in the effluent collection cell at experimental condition (21°C and 200kPa) was similar to the volume expansion of the vapor chamber in the physical model. Using the collected data on position of interface, the contact angle was determined.

Figure 4.29 shows the penetration depths of n-butane in the Vapex experiments calculated from Equation 4.5. As seen in Figure 4.29, the calculated penetration depths increased with increase of absolute permeability or reduction of specific pore surface area of the porous medium. The calculated penetration depths also showed a similar relationship with the drainage height of porous medium in that higher penetration depths were observed in experiments conducted at the higher drainage height. The accuracy of these calculations is a sensitive function of the accuracy of the oil saturation
measurements conducted by liquid extraction of the oil from the samples collected from the physical models at the end of each Vapex experiment.

![Graph showing solvent penetration depth vs. permeability](image)

**Figure 4.29: Calculated n-butane penetration depth in Vapex experiments conducted in this study**

The calculated penetration depths are consistent with other measurements, such as measured drainage rates, and solvent content of produced oil. The calculated solvent penetration depths in this study were in the range of those reported by Yazdani and Maini (2009). However, the calculated penetration depths of the n-butane in this study were lower than those reported by Yazdani and Maini (2009) for Vapex experiments conducted at similar drainage heights but higher permeabilities (>220 D). The higher solvent penetration depths at higher drainage heights could be attributed to the higher hydrostatic pressure imposed by density difference in the Vapex experiments conducted in the larger physical model.
CHAPTER 5: ANALYTICAL & SEMI-ANALYTICAL MODELING STUDY OF EXPERIMENTAL PROGRAM

This chapter provides the details of conducted semi-analytical modeling for development of a thermodynamic model for characterization of Plover Lake/n-butane system required for analytical and numerical simulations of the conducted experiments. In addition, this chapter presents the results of analytical modelings to investigate the effects of studied parameters (i.e., drainage height, capillarity, immobile water saturation) on the rate of mass transfer between the solvent vapor and oil phase in porous media with similar characteristics as western Canadian heavy oil reservoirs. On the basis of this study, new solvent concentration dependent correlations were developed for prediction of effective diffusion coefficient of n-butane in the Vapex process. Conducted study provides strong evidence on better understanding of the effects of studied parameters on the efficiency of the mass transfer phenomenon in the Vapex process.

5.1. Phase Behavior Modeling

The phase behavior modeling of the Plover Lake oil/n-butane system included empirical and semi-analytical models. These models were investigated for prediction of phase behavior properties (e.g., density, and viscosity) of the studied oil-solvent mixtures at wide range of n-butane concentrations. The semi-analytical modeling in this section was conducted using CMG’s WINPROP™ package.

5.1.1. PVT Model

In order to develop a PVT model using equation of state (EOS) approach, the minimum requirements for characterization of heavy components are molecular weight, specific gravity, and normal boiling point. A minimum two of these three parameters are required
for calculation of critical pressure, critical temperature, and acentric factors of heavy components. The critical properties are calculated from specific gravity and normal boiling pint; molecular weight is used for determination of mass densities. Molecular weight and specific gravity of Plover Lake oil were experimentally measured as explained in chapter 4. The molecular weight of the oil in this study was 469 measured by vapor pressure osmometer. The specific gravity of the oil was 0.9835.

In development of PVT model, Peng-Robinson (1978) equation of state was used. In this regard, the heavy component (C_{9+}) was split and lumped into three pseudo components using two-stage exponential distribution function. The tuning of the equation of state was performed using the measured saturation pressures, and other collected phase behavior data on Plover Lake oil/n-butane system at observed range of n-butane concentrations. The regression analysis through the tuning process was performed by selection of combination of following regression parameters: critical properties, volume shifts, molecular weight, acentric factor, and binary interaction coefficients of the pseudo components.

Figures 5.1 to 5.4 show the quality of tuned PVT model predictions against the measured PVT properties. As shown in these figures, there is a very good agreement between the measured densities and predicted density values at different concentrations of n-butane and pressures. The conducted error analysis showed a maximum relative error of 1.61% between the measured and EOS predicted densities in this study. The predicted saturation pressures showed up to 12.3% deviation from the measured saturation pressures disregarding assignment of higher weights factors to saturation pressures throughout the regression analysis in tuning process.
The high errors in saturation pressure predictions could be partially result of supersaturation effect occurring at lower pressure range (200 kPa) of conducted PVT experiments. In addition, the compositional analysis of the vapor phase in equilibrium with the oil phase and the collected gas samples from the vessel containing the Plover lake oil revealed presence of trace of helium. Considering the immiscibility of helium in oil phase at studied pressure range, it could have a significant effect on measured saturation pressures. In order to include the effect of incondensable component in the vapor phase into regression analysis, the tuning of the PVT model was conducted by adding a trace of methane into the model to compensate for the effect of helium. The conducted tuning resulted in reduction of average absolute error in saturation pressure predictions to 8.47%.

![Figure 5.1: Measured and modeled density of dead oil at studied pressures](image_url)
Figure 5.2: Measured and modeled density of oil/butane mixtures

Figure 5.3: Measured and modeled density of oil/butane mixture (C₄=38.90 mol%)
Figure 5.4: Measured and modeled saturation pressures of oil-butane mixture

5.1.2. Viscosity Models

The viscosity of oil-solvent mixtures plays an important role in efficiency of mass transfer phenomena and performance of the gravity drainage process in the Vapex process. Commercial softwares are normally using two different types of viscosity correlations for prediction of vapor and oil viscosities. These common correlations are: the Jossi, Stiel, and Thodos (JST) as described by Reid et al. (1977) and Pedersen (1984) corresponding states correlation.

The quality of viscosity predictions by JST correlations very strongly depends on the critical volumes for viscosity (and to a lesser extent, the critical temperatures and pressures), and predicted density using equation of state. In general, JST is not recommended approach for viscosity prediction of heavy crudes because of the problems
with characterization of heavier components and accurate predictions of liquid phase densities.

Pedersen (1984) viscosity correlation uses the principle of corresponding states to calculate the viscosity of a component or mixture based on a reference substance at the same conditions of reduced pressure and temperature. In this approach, the deviation from simple corresponding states is accounted by a rotational coupling coefficient. Equation 5.1 presents the Pedersen correlation for viscosity estimation of a mixture.

\[
\frac{\mu_{\text{mix}}(P,T)}{\mu_o(P_o,T_o)} = \left(\frac{T_{c,\text{mix}}}{T_{c,o}}\right)^{1/6} \left(\frac{P_{c,\text{mix}}}{P_{c,o}}\right)^{2/3} \left(\frac{MW_{\text{mix}}}{MW_o}\right)^{1/2} \left(\frac{\alpha_{\text{mix}}}{\alpha_o}\right)
\]  

(5.1)

In Equation 5.1, the subscript “o” refers to the reference substance property. In Pedersen correlation this reference substance is methane. The critical properties of the mixture are calculated using mixing rules that are a function of critical temperatures and pressures, and mole fractions. The molecular weight of the mixture and rotational coupling coefficient are calculated by Equations 5.2 and 5.3.

\[
MW_{\text{mix}} = b_1(MW_w^{b_2} - MW_n^{b_2}) + MW_n
\]  

(5.2)

\[
\alpha = 1 + b_3\rho_r^{b_4}MW^{b_5}
\]  

(5.3)

The accuracy of viscosity predictions using this correlation is strongly dependent on the critical pressures, critical temperatures, molecular weight of the components, and the \(b_i\) values in Equations 5.2 and 5.3. Pedersen correlations (1984, 1987) were developed for crudes with viscosities less than 10 mPa·s, however, this correlation is more successful in viscosity prediction of heavy crudes by modification of default coefficients during the regression analysis to match the experimental viscosity data.
Figures 5.4 to 5.6 show the comparison between the predicted and measured viscosity of Plover lake oil/n-butane system. The average absolute error of viscosity prediction in investigate cases was 4.29% in the viscosity range of 25 to 10,541 mPa·s.

The quality of viscosity predictions using semi-analytical approach are compared against four common heavy oil viscosity correlations in Figure 5.6. The summary of error analysis of viscosity prediction for various correlations is summarized in Table 5.1. These viscosity correlations are:

1- Classical log type correlation

\[
\log \mu_{\text{mix}} = C_s \log \mu_s + C_o \log \mu_o
\]  
(5.4)

2- Bingham (1918) correlation

\[
\frac{1}{\mu_{\text{mix}}} = \frac{C_s}{\mu_s} + \frac{C_o}{\mu_o}
\]  
(5.5)

3- Kendall and Monroe (1917) correlation

\[
\mu_{\text{mix}} = \left(C_n \mu_s^{0.09} + (1 - C_n) \mu_o^{0.09}\right)^{\frac{1}{0.09}}
\]  
(5.6)

4- Shu (1984) correlation

\[
\mu_{\text{mix}} = \mu_o^x \mu_s^{x_o}
\]  
(5.7)

where

\[
x_o = \frac{\alpha'C_o}{\alpha'C_o + C_s}
\]  
(5.8)

\[
\alpha' = \frac{17.04 \Delta \rho^{0.5237} \rho_s^{1.2745} \rho_o^{1.6316}}{\ln \left( \frac{\mu_o}{\mu_s} \right)}
\]  
(5.9)

\[
\Delta \rho = \rho_s - \rho_o
\]  
(5.10)
Figure 5.5: Measured and modeled viscosity of dead oil

Figure 5.6: Measured and modeled viscosity of oil-butane mixture (C₄=38.90 mol%)
Figure 5.7: Comparison between measured and predicted viscosity of oil-butane mixtures at different n-butane concentrations by different approaches

Table 5.1: Error analysis of viscosity predictions of oil-butane mixtures by different approaches

<table>
<thead>
<tr>
<th>Correlation</th>
<th>Absolute Error (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Min</td>
</tr>
<tr>
<td>Pedersen (1984,1987)</td>
<td>3.25</td>
</tr>
<tr>
<td>Log type correlation</td>
<td>154.23</td>
</tr>
<tr>
<td>Bingham (1918)</td>
<td>99.87</td>
</tr>
<tr>
<td>Kendall and Monroe (1917)</td>
<td>11.53</td>
</tr>
<tr>
<td>Shu (1984)</td>
<td>33.40</td>
</tr>
</tbody>
</table>
5.2. Analytical Modeling of the Vapex Experiments

In order to investigate the role of studied parameters on effective mass transfer coefficient, modified form of original Butler’s model proposed by Das and Butler (1998) was used to analytically model the Vapex experiments. For this purpose, the first step was calculation of the Vapex numbers based on experimental data and characteristics of porous media. The Vapex numbers for each set of experiments were back-calculated by Equation 2.33. Then, the theoretical Vapex numbers were determined by numerical computation of integrant in Equation 2.16.

To determine the theoretical Vapex numbers, it was required to determine viscosity and density of the oil-solvent mixture in the observed range of concentration change between the upper and lower limits of the integral. For this purpose, viscosity predictions were performed by Kendall and Monroe (1917) correlation as presented in Equation 5.6. The density of solvent-oil mixtures were calculated by Equation 4.2.

In order to determine the theoretical Vapex number in Equation 2.16, Equation 5.11 was utilized to estimate the diffusion coefficient. Based on this equation, the molecular diffusivity of solvent is a concentration dependent parameter, which can be correlated with viscosity term. Figure 2.6 shows the solvent concentration dependency of viscosity and diffusivity of Plover Lake oil. In this figure, diffusivity coefficients were calculated using correlation developed for Peace River bitumen-butane system by Das and Butler (1996) in Equation 2.28. As it can be seen in Figure 2.6, the assumption of concentration dependency of effective diffusion has significant effect on the calculated diffusion coefficients. Three orders of magnitude change in diffusion coefficients was observed in the entire domain of the graph.
\[ D_o = \alpha \mu^\beta \]  
\[ \text{(5.11)} \]

To develop a correlation for estimating effective the diffusion/dispersion coefficient in the Vapex process, Equation 2.28 was utilized as starting point. In addition, it was assumed that similar diffusivity-viscosity relation for butane-Peace River Bitumen could be extended to butane-Plover Lake oil used in the experimental part of this study. On the basis of this assumption one of the constants in Equation 5.11 (i.e., \( \beta \)) was fixed at -0.46. The other coefficient in Equation 5.11 (i.e., \( \alpha \)) was used to history-match and develop the correlation based on back-calculated Vapex numbers using Equation 2.33. The value of \( \alpha \) for each test was calculated using Equation 5.12 that was derived from Equations 2.33 and 2.16. The value of the integrant in denominator of Equation 5.12 were computed by numerical integration technique and assumed relationship between the phase behavior properties of diluted oil through the Vapex process by dissolution of solvent.

\[
\alpha = \frac{N_v,\text{exp}}{\int_{c_m}^{c} \rho_{\text{mix}} \mu_{\text{mix}}^{-\left(1+\beta\right)} \left(1-\frac{C_s}{C_x}\right) dC_s}
\]
\[ \text{(5.12)} \]

In Equation 5.12, the lower limit of integral in the denominator \( (c_m) \), is the minimum n-butane concentration required for mobilizing the oil phase. This concentration should be fixed arbitrarily. It should be noted that this parameter cannot be zero because theoretically the integral becomes infinite. In numerical computation of the integral the limit of mobility was arbitrarily set to 9,400 mPa·s viscosity that is equivalent to n-butane concentration of 1.42 mol%. Sensitivity study of the selection of lower limit of integration showed that the choice of the lower limit of integral did not result in any significant changes in the numerically computed integrals as long as it is small and positive.
The upper limit of the integral in Equation 5.12 \((c_i)\) is the interfacial n-butane concentration i.e., the solubility of gaseous n-butane in oil phase at experimental pressure and temperature. This parameter should be obtained from vapor-liquid equilibrium data. This parameter was estimated using isothermal flash calculations for Plover Lake oil/n-butane system.

Once the integral limits and other various terms in Equation 5.16 were defined, the only unknowns are \(\alpha\) and \(\beta\). Assuming similarity between of diffusion-viscosity relationship for butane-Peace River bitumen developed by Das and Butler (1996), one of these parameters \((\beta=-0.46)\) was relaxed. On the basis of this assumption, Equation 5.12 was solved for \(\alpha\) for all the experimental points. After numerical computation of the integral for each experiment, relationships between calculated \(\alpha\) values and various combinations of drainage height, specific pore surface area of porous media, and oil-solvent viscosity were investigated. The details of calculations of \(\alpha\) for two series of experiments conducted in absence and presence of immobile water saturation are provided in the following sections.

5.2.1. Vapex Experiments in Absence of Immobile Water Saturation

In this part of study, six Vapex experiments were carried out at two drainage heights and three different sand packs. These experiments were performed in absence of immobile water saturation. The main objective of these experiments was investigation of effects of capillarity and drainage height on stabilized oil drainage rates and the investigation of effects of these parameters on solvent mass transfer rates and effective diffusion/dispersion coefficients.
The back-calculated Vapex numbers and characteristics of these experiments are reported in Table 5.2. In this table, reported microscopic sweep efficiency ($\Delta S_o$) for each experiment was determined by analysis of the oil content of the sand pack at the end of each experiment using solvent extraction technique. As seen in Table 5.2, capillarity and drainage height had significant effect on Vapex numbers and measurable effect on microscopic sweep efficiencies. For instance the comparison between the Vapex number in Experiment#1 and #3 showed two orders of magnitude decrease in the Vapex number, whereas these experiment were conducted at the same drainage heights and almost similar permeabilities.

**Table 5.2: Back-calculated Vapex numbers and characteristics of Vapex experiments in absence of immobile water saturation**

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Drainage height (Cm)</th>
<th>$\Delta S_o$</th>
<th>Oil rate ($\text{cm}^2/\text{s}$)</th>
<th>Vapex Number</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>24.5</td>
<td>0.877</td>
<td>1.77e-05</td>
<td>4.21e-07</td>
</tr>
<tr>
<td>2</td>
<td>47.5</td>
<td>0.897</td>
<td>3.44e-05</td>
<td>8.04e-07</td>
</tr>
<tr>
<td>3</td>
<td>24.5</td>
<td>0.592</td>
<td>9.85e-07</td>
<td>4.30e-09</td>
</tr>
<tr>
<td>4</td>
<td>47.5</td>
<td>0.625</td>
<td>3.03e-06</td>
<td>1.46e-08</td>
</tr>
<tr>
<td>5</td>
<td>24.5</td>
<td>0.612</td>
<td>2.29e-06</td>
<td>1.80e-08</td>
</tr>
<tr>
<td>6</td>
<td>47.5</td>
<td>0.717</td>
<td>9.30e-06</td>
<td>8.94e-08</td>
</tr>
</tbody>
</table>

The back-calculated Vapex numbers in Table 5.2 also indicate strong dependency on drainage heights of the physical models. Comparison between Vapex numbers showed 90.9 to 396% increase by increase of drainage height from 24.5 to 47.5cm. Figure 5.8 shows the relationship between the measured stabilized drainage rates and specific pore
surface area of the medium. As shown in this figure, there is a continuous reduction of the drainage rates and solvent content of the produced oil with increase of capillarity effect in the medium.

![Graph showing dependency of stabilized drainage rates on specific pore surface area of porous medium](image)

**Figure 5.8: Dependency of stabilized drainage rates on specific pore surface area of porous medium**

The calculated $\alpha$ values for the Vapex experiments conducted in absence of immobile water saturation are listed in Table 5.3. In order to investigate the possibility of developing a new model for estimation of the effective diffusion coefficient in the Vapex process, various combinations of studied parameters were tested to develop a correlation between the observed $\alpha$ values and selected parameters. Figure 5.9 shows the dependence of $\alpha$ values on drainage height and specific pore surface area of porous medium.
Table 5.3: Calculated $\alpha$ values in absence of immobile water saturation

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>$\alpha$ (Dimensionless)</th>
<th>Average $\alpha$ (Dimensionless)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>5.80e-05</td>
<td>8.45e-05</td>
</tr>
<tr>
<td>2</td>
<td>1.11e-04</td>
<td></td>
</tr>
<tr>
<td>3</td>
<td>5.92e-07</td>
<td>1.30e-06</td>
</tr>
<tr>
<td>4</td>
<td>2.01e-06</td>
<td></td>
</tr>
<tr>
<td>5</td>
<td>2.48e-06</td>
<td>7.39e-06</td>
</tr>
<tr>
<td>6</td>
<td>1.23e-05</td>
<td></td>
</tr>
</tbody>
</table>

Figure 5.9: Dependency of $\alpha$ parameter on drainage height and specific pore surface area of medium in absence of immobile water saturation

Based on the relationship between $\alpha$ values and studied parameter in this study, Equation 5.13 was developed to estimate effective diffusion/dispersion coefficient. The developed
correlation in Equation 5.13 accounts for drainage height and capillarity, which play important roles in studied permeability range in addition to the viscosity of oil-solvent mixture. In this correlation, the measured specific pore surface areas of the sand packs were used to represent the effect of capillary forces. The new correlation presents higher dependency on drainage height as it was proposed by Yazdani and Maini (2009) based on the experiments conducted in higher permeabilities and drainage heights up to 100 cm.

\[ D_{\text{eff}} = 1.357 \times 10^{-11} \times h^{3.235} \times A^{-0.154} \times \mu_{\text{mix}}^{-0.460} \]  

(5.13)

The dependency of effective diffusion/dispersion coefficients on n-butane concentration calculated using Equation 5.13 for experimentally simulated cases in this study is shown in Figure 5.10. Although, the solvent concentration varies between 0 and 1.0 in this figure, the calculated diffusion coefficients based on measured n-butane concentrations for each experiment is shown in Figure 5.10. As shown in this figure, the n-butane concentration of collected oil samples from the large model was consistently slightly higher than the oil samples from the small model. For instance, the n-butane concentration of the small model in experiment conducted in Sand#1 is 52 mol% compared 46 mole% in oils collected from the large model. This difference in n-butane concentration translates into reduction in viscosity from 121 to 49 mPa·s. This difference between the solvent content of the produced oils can be attributed to lower viscosities of the oil phase required at lower drainage heights to overcome the capillary forces for becoming mobile and drain under gravity effect.

5.2.2. Vapex Experiments in Presence of Immobile Water Saturation

In order to investigate the effect of immobile water saturation on performance of the Vapex process and efficiency of mass transfer phenomena, a series of Vapex experiments
were designed and conducted in presence of immobile water saturation. In this section, a total of five Vapex experiments were performed in water saturation range of 12.6 to 13.4%. Figure 5.11 shows the dependency of the measured stabilized drainage rates on specific pore surface area of the medium. As shown in this figure, there is a continuous reduction in stabilized drainage rates and specific pore surface area of the medium. This behavior is similar to behavior observed in experiments in absence of water saturation. However, significant increase in drainage rates was observed in these experiments. The experimental Vapex numbers for these experiments were calculated and listed in Table 5.4. The $\alpha$ values for these experiments were computed according to Equation 5.16 and listed in Table 5.5.

![Figure 5.10: Dependency of effective diffusion coefficient on investigated parameters in absence of water saturation](image)

Figure 5.10: Dependency of effective diffusion coefficient on investigated parameters in absence of water saturation
The dependency of \( \alpha \) values on combination of drainage height, specific pore surface area of the medium, and viscosity of solvent-oil mixture is presented in Figure 5.12. Comparison between Figures 5.8 and 5.12 shows the positive impact of immobile water saturation on the rate of mass transfer between solvent and the oil phase. The regression analysis shows significant reduction in the role of drainage height on effective diffusion coefficients.

Solvent content measurements of the produced oils from the experiments conducted in presence of aqueous phase were relatively similar in these experiments and considerably higher than the solvent content of oil produced in experiments in absence of immobile water saturation in Sand#2 and Sand#3. In these experiments the n-butane concentration varied between 48.1 to 50.1 mol\%. For instance, the n-butane concentration of the small model in experiments conducted in Sand#1 was 51.7 mol\% and compared 50.1 mol\% n-butane content of the oil collected from the large model. This is in contrast with measured n-butane concentration of collected oil samples in the Vapex experiments performed in absence of water saturation. The n-butane concentration in dry Vapex experiment varied between 15 and 53 mol\% where lower concentration were observed in Sand#2 and Sand#3.

On the basis of performed regression analysis on the derived \( \alpha \) values from the experimental study conducted in presence of immobile water saturation, Equation 5.14 was developed for estimation of effective diffusion coefficient of n-butane into the oil phase.

\[
D_{\text{eff}} = 3.572 \times 10^{-7} \times h^{0.677} \times A^{-0.211} \times \mu_{\text{mix}}^{-0.460}
\]  
\[(5.14)\]
Figure 5.11: Dependency of stabilized drainage rates with specific pore surface area of porous medium

Table 5.4: Back-calculated Vapex numbers in presence of immobile water

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Drainage height (cm)</th>
<th>$\Delta S_o$ (Dimensionless)</th>
<th>Oil rate ($cm^2/s$)</th>
<th>Vapex Number (Dimensionless)</th>
</tr>
</thead>
<tbody>
<tr>
<td>7</td>
<td>24.5</td>
<td>0.690</td>
<td>2.54e-05</td>
<td>8.80e-07</td>
</tr>
<tr>
<td>8</td>
<td>47.5</td>
<td>0.848</td>
<td>3.58e-05</td>
<td>1.26e-06</td>
</tr>
<tr>
<td>9</td>
<td>24.5</td>
<td>0.602</td>
<td>9.05e-06</td>
<td>2.61e-07</td>
</tr>
<tr>
<td>10</td>
<td>47.5</td>
<td>0.685</td>
<td>2.65e-05</td>
<td>9.64e-07</td>
</tr>
<tr>
<td>11</td>
<td>24.5</td>
<td>0.591</td>
<td>1.07e-05</td>
<td>2.78e-07</td>
</tr>
</tbody>
</table>
Table 5.5: Calculated $\alpha$ values in presence of immobile water saturation

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Water Saturation</th>
<th>$A$</th>
<th>Average $\alpha$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>%</td>
<td>Dimensionless</td>
<td>Dimensionless</td>
</tr>
<tr>
<td>7</td>
<td>12.92</td>
<td>6.61e-06</td>
<td>9.35e-05</td>
</tr>
<tr>
<td>8</td>
<td>12.62</td>
<td>1.21e-05</td>
<td></td>
</tr>
<tr>
<td>9</td>
<td>13.42</td>
<td>2.47e-06</td>
<td>3.05e-06</td>
</tr>
<tr>
<td>10</td>
<td>13.18</td>
<td>3.64e-06</td>
<td></td>
</tr>
<tr>
<td>11</td>
<td>13.15</td>
<td>3.29e-06</td>
<td>3.29e-06</td>
</tr>
</tbody>
</table>

Figure 5.12: Dependency of $\alpha$ parameter to drainage height, specific pore surface area of medium and oil-solvent mixture viscosity in presence of immobile water

According to the developed correlation in Equation 5.14, the dependency of the effective diffusion coefficients on porous media characteristics and the concentration of n-butane were significantly reduced in presence of aqueous phase. Comparison between the power
of drainage height and specific pore surface area terms in Equations 5.13 and 5.14 shows the positive impact of aqueous phase on effective diffusion confidents. The dependency of effective diffusion coefficients on n-butane concentration calculated using Equation 5.14 for experimentally simulated cases in this study are shown in Figure 5.13. Although, the solvent concentration varies between 0 and 1.0 in this figure, the measured n-butane concentrations for Vapex experiments are highlighted.

![Figure 5.13: Dependency of effective diffusion coefficient on investigated parameters in presence of water saturation](image)

5.3. Results and Discussion

The diffusion coefficients for all the Vapex experiments were calculated from history matching of these tests and are compared in Figure 5.13. This figure shows the effect of
immobile water saturation, capillarity, and drainage height on effective diffusion coefficient of n-butane in conducted experiments in relatively similar permeabilities.

Figure 5.14: Dependency of effective diffusion coefficient on drainage height, specific pore surface area, and immobile water saturation

The measured effective diffusion coefficients in absence of immobile water saturation varied between $4.91 \times 10^{-8}$ and $7.89 \times 10^{-6}$ cm$^2$/s. The range of variation in effective diffusion coefficients in presence of immobile water saturation was between $2.99 \times 10^{-6}$ and $1.10 \times 10^{-5}$ cm$^2$/s. The observed variation range of the effective diffusion coefficients in this study were either lower or in range of reported molecular diffusion coefficients in literature for carbon dioxide, and light alkanes (e.g., methane, ethane, propane, and butane) in heavy oils and bitumens (Schmidt et al., 1982; Zhang et al., 2000; Upreti and Mehrotra, 2002; Creux et al., 2005; Yang and Gu, 2006; Cuhicel et al., 2006). On this basis, it can be concluded that the convective dispersion or velocity-dependent effects are
playing a minor role in the mass transfer phenomena in the conducted Vapex experiments.

Figure 5.15 shows the comparison between the effective diffusion coefficients in this study and predicted diffusivities by correlations developed by Das and Butler (1996) and Yazdani and Maini (2009).

**Figure 5.15: Comparison between the calculated effective diffusion coefficients and predicted effective diffusivity with proposed correlations for butane in literature**

In Figure 5.15, the diffusivities were predicted based on measured solvent content of the produced oil in each set of experiments. As seen in this figure, the butane concentration in collected oil samples in experiments conducted in presence of immobile water saturation were in relatively similar range. The calculated effective diffusivities were in
similar range disregarding the specific pore surface area of the medium and in the range molecular diffusion coefficients predicted by Das and Butler (1996) correlation.

In contrast with experiments in presence of immobile water saturation, the measured diffusivity coefficients in experiments conducted in absence of water saturations had lower effective diffusivities. In these experiments there is a strong correlation between the butane content and viscosity of the produced oil with specific pore surface area of the medium as shown in Figure 5.16. This figure includes the viscosity of the oil phase at equilibrium condition at experimental pressure and temperature predicted by equation of state tuned using the conducted phase behavior study.

![Figure 5.16: Capillary dependency of the viscosity of produced oil samples in absence of immobile water saturation.](image)

Higher concentration of solvent (i.e., 50.12-52.58 mol%) were measured for oil samples from experiments performed in presence of immobile
water saturation. The solvent contents of oils in experiments performed in presence of aqueous phase are presented in Figure 5.17. As shown in this figure, the solvent content of the produced oil was almost similar, regardless of drainage height and specific pore surface area of the porous medium. In these experiments, it is believed that the immobile wetting phase occupies the smaller pore throats, thereby blocking them for flow of oil. This results in change in effective pore morphology experienced by the oil phase. Conducted experimental and numerical study by Verlaan et al. (1999) on effect of immobile wetting phase on diffusion and macroscopic dispersion in unconsolidated porous media showed 2 to 3 times higher dispersion coefficients in presence of water than in dry porous media.

![Figure 5.17: Capillary dependency of the viscosity of produced oil samples in presence of immobile water saturation.](image-url)
Comparison between the developed correlations in Equations 5.13 and 5.14 shows that the effective diffusion coefficients in dry sand experiments were governed by the drainage height and the viscosity of the oil-solvent mixture. This means that higher drainage heights or lower viscosities should be achieved to overcome the capillary forces and mobilized the oil phase. There is a drastic change in the rate of mass transfer in wet experiments where the impact of drainage height and oil-solvent mixture reduced considerably where the rate of mass transfer is proportional with the drainage height and inverse of the viscosity of oil-solvent mixture.

The effect of water saturation on drainage rates was also investigated by Equation 5.16. Figure 5.18 shows the trend of calculated drainage rate ratios versus the specific pore surface area of porous media.

![Graph showing drainage rate dependency on immobile water saturation](image)

**Figure 5.18: Drainage rate dependency on immobile water saturation**
This analysis shows that the presence of aqueous phase reduces the capillary forces required to be overcome for mobilizing the oil phase. For instance, conducted experiments in Sand#3 in presence of immobile water phase resulted in 9.19 times increase in the stabilized drainage rate compared to the similar test conducted in dry sand.

\[ Q_d = \frac{Q_{(sw>0)}}{Q_{(sw=0)}} \]  

(5.16)
CHAPTER 6: NUMERICAL SIMULATION OF MASS TRANSFER PHENOMENA IN MIXING ZONE IN THE VAPEX PROCESS

In this chapter, the results of a detailed numerical simulation study are presented and discussed. The main objective of the this simulation study was history matching of selected Vapex experiments in order to infer the range of molecular diffusion and potential physical dispersion coefficients controlling the mass transfer rate in experimentally simulated mixing zone on the edge of the vapor chamber. The commercial simulator used for this study was the 2010 version of STARS™ (Computer Modelling Group Ltd., 2010). In the following sections, the details of the simulated control volume, simulation model constraints, details of diffusion and dispersion modeling in the numerical simulator, and the results of history matching process are presented and discussed.

6.1. Boundary and Initial Conditions of Simulated Control Volume

In the Vapex process, a vaporized solvent comes into contact with the oil phase at the edge of the vapor chamber. The solvent dissolves and diffuses into the heavy oil and mobilizes the oil phase in the mixing zone. According to available experimental studies the thickness of this mixing zone is less than 14 mm (James et al., 2008; Yazdani and Maini, 2009; Moghadam et al., 2009; Ahmadloo et al., 2011). Considering relatively low pressure gradient between the injection and production wells in the Vapex process, gravity is the main driving force for displacement of the mobilized oil at reservoir condition. Therefore, it was required to conduct the simulations in absence of any forced
displacement in order to realistically simulate the mass transfer phenomena in the Vapex process.

Figure 6.1 shows a 2-D rectangular control volume used for simulation of the Vapex process based on the experimental setup in this research. This control volume was constructed of the following three major zones.

- **Zone I:** This zone represented the vapor chamber in experimental study that was packed with glass beads with average permeability of 830 D and saturated with n-butane vapor. The dimension of the modeled vapor chamber was 0.64 cm deep by 24.5 or 47.5 cm high depending on drainage height of the physical model across the width of the experimental models. The solvent vapor was injected at the top of this zone at a setpoint pressure (i.e., ~200 kPa).

- **Zone II:** This zone represented the reservoir condition at the edge of the vapor chamber which was packed with sand and saturated with the Plover Lake heavy oil. This zone also was 4.44 cm deep by 24.5 or 47.5 cm high depending on the size of the physical model used in the Vapex experiments. Zone II was separated from Zone I using a 120 U.S. mesh sieve without adding any resistance for diffusion of solvent into the sand pack.

- **Zone III:** This zone was located at the bottom of the physical models in order to collect all the drained oil from the Zone II under gravity drainage process. A production port which was located at the bottom of this zone collected drained oil from the sand pack. Since the pressure was equilibrated among all the zones and pressure disturbances were eliminated, the effect of n-butane pressure on the system was negligible.
In theory, at $t > 0$, n-butane begins diffusing into the oil phase through the separating sieve (Boundary 1) to mobilize the oil phase as it occurs in the Vapex process. Once the oil is mobilized, it drains into the collection zone and is produced through the production port (Boundary 4). The details of the boundary and initial conditions for the simulated control volume are summarized in Table 6.1. The “y” axis is vertical with gravity acting in the “-y” direction.

![Figure 6.1: Control volume of the numerically simulated model](image)

**Figure 6.1: Control volume of the numerically simulated model**

### 6.2. Implementation of Boundary Conditions

The 2010 version of CMG STARS™ simulator, was used to numerically simulate and history match the conducted Vapex experiments. The basic approach in building the model was to keep the model as simple as possible with sufficient complexity that the
simulator took into account of solubility, viscosity reduction, volume change because of solvent dissolution, diffusion/dispersion, and capillary pressure.

Table 6.1: Boundary and initial conditions of simulated control volume

<table>
<thead>
<tr>
<th>Boundary Conditions</th>
<th>Darcy’s equation</th>
<th>Mass transfer equation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boundary (1)</td>
<td>$P_m$</td>
<td>$C_s=0.76$</td>
</tr>
<tr>
<td>Boundary (2)</td>
<td>Symmetry/zero flux</td>
<td>Insulated/symmetry</td>
</tr>
<tr>
<td>Boundary (3)</td>
<td>Symmetry/zero flux</td>
<td>Insulated/symmetry</td>
</tr>
<tr>
<td>Boundary (4)</td>
<td>$P_m$</td>
<td>Insulated/symmetry</td>
</tr>
<tr>
<td>Boundary (5)</td>
<td>Symmetry/zero flux</td>
<td>Insulated/symmetry</td>
</tr>
<tr>
<td>Initial Condition</td>
<td>$P_o=P_m+\rho_ogh$</td>
<td>$C_s=0.0$</td>
</tr>
</tbody>
</table>

6.2.1. Gridding and Model Definition

The drainage layer of fluid in the mixing zone on the edge of vapor chamber was very thin. Through this thin zone, fluid flow and extreme viscosity reduction were occurred. Therefore, extremely fine grid blocks were required to calculate the concentration profile in the mixing zone and consequently capture the diffusion and dispersion in this zone. In addition, lack of considerable pressure differences between the injection and production ports in the physical models and slow saturation changes in the simulated experiments required extremely fine grids for avoiding the numerical instability during numerical simulation process.

In this regard, grid blocks with lateral dimensions as small as 2.5 mm in the sand pack and 3.2 mm in vapor chamber were used in history matching simulations. Typically, smaller grid size produced more accurate numerical solution; however, this approach
could significantly increase simulation times and memory requirement for running the simulation models. The grid block size of 2.5 mm was selected after comparison among different simulation models with different grid sizes based on computation times, accuracy of predictions, and material balance error analysis.

Considering the geometry of the experimental models, boundary conditions, and homogeneity of the sand packs, the simulations were conducted for a control volume with much smaller width than the real physical models instead of the full size model. The width of the simulated control volume was selected to be 2.5 mm to keep the aspect ratio of the grid blocks as one. This approach helped to optimize the CPU time without compromising the quality of numerical model predictions by using larger grid sizes than the observed mixing zone thickness in this study. The total drainage rates and n-butane production rates for each Vapex experiment could be determined by integration of the simulator results across whole width of physical models (i.e., 20 cm).

Since the physical models were packed by hand, it was not anticipated that there were high degree of heterogeneity of sand and associated properties of the sand packs, such as porosity and permeability. Therefore, constant values were used in the simulations based on measurements for each sand pack.

As explained in section 6.1, the simulated control volume was comprised of three zones with different characteristics, such as porosity, permeability, relative permeability, and capillary pressure curves. These different characteristics were required for representative implementation of experimentally simulated boundary conditions.
6.2.2. PVT and Viscosity Model

In simulation of the Vapex process, PVT and viscosity models play crucial roles. These parameters directly influence the amount of solvent dissolved into the oil phase and consequently the mobility of the oil phase in the simulated mixing zone. In order to accurately model the interaction of the oil phase components and the injected solvent, it was required to develop a PVT model. The developed PVT model was tuned against the experimental data for Plover Lake oil/n-butane system in observed range of n-butane concentrations. The details of the phase behavior measurements and PVT model development were provided in Chapters 4 and 5.

In order to optimize the computational times for extremely fine simulation models used in this study, the properties of the oil-solvent mixtures at each time step were calculated using equilibrium ratios (K-values) rather than using a fully compositional EOS-based simulation model. CMG’s WinProp™ PVT package (Computer Modelling Group Ltd., 2010) was used to generate the PVT input file for the simulator.

6.2.3. Rock-Fluid Properties

Mercury-air capillary pressures and pore size distributions of sand packs used in this study were measured in sand packs prepared similar to sand packs used in the Vapex experiments. The details are available in Section 3.2 of this dissertation.

In simulation study, it was required to derive the capillary pressure curves for gas-oil and oil-water systems. The measured capillary pressure measurements of air/mercury systems are shown in Figures 3.8 to 3.10. In order to determine the capillary pressures required for simulation of drainage Vapex experiments, Equation 3.11 was used for calculating the capillary pressure curves for n-butane/Plover Lake oil and brine/Plover Lake oil for each
sand pack. The parameters used for calculation of capillary pressures are listed in Table 6.2. Considering extremely high permeability of glass bead packs in Zone#1 and Zone#3, the capillary pressure in these sectors was set to be 0.

**Table 6.2: Interfacial tension and contact angles used for determination of capillary pressure curves from mercury injection data**

<table>
<thead>
<tr>
<th>System</th>
<th>IFT</th>
<th>Contact Angle</th>
</tr>
</thead>
<tbody>
<tr>
<td>Air/Mercury</td>
<td>485</td>
<td>130</td>
</tr>
<tr>
<td>Brine/Plover Lake Oil</td>
<td>10.6</td>
<td>0</td>
</tr>
<tr>
<td>n-Butane/ Plover Lake Oil</td>
<td>18.6</td>
<td>0</td>
</tr>
</tbody>
</table>

Equations 6.1 to 6.4 were used for calculation of relative permeability curves for oil-water and gas-oil systems in Zone#2. The exponents in these equations were assumed to be 2.0. Option of Stone’s Model II as modified by Aziz and Settari (1979) was applied to calculate three-phase relative permeabilities in case of presence of mobile water.

\[ k_{rw} = k_{rw_0} \left( \frac{S_w - S_{w_crit}}{1 - S_{w_crit} - S_{a_0}} \right)^{N_w} \]  \hspace{1cm} (6.1)

\[ k_{row} = k_{row_0} \left( \frac{S_o - S_{a_0}}{1 - S_{w_0} - S_{a_0}} \right)^{N_{ow}} \]  \hspace{1cm} (6.2)

\[ k_{rog} = k_{rog_0} \left( \frac{S_g - S_{a_0} - S_{w_0}}{1 - S_{g_0} - S_{a_0} - S_{w_0}} \right)^{N_{og}} \]  \hspace{1cm} (6.3)

\[ k_{rg} = k_{rg_0} \left( \frac{S_g - S_{g_crit}}{1 - S_{g_crit} - S_{a_0} - S_{w_0}} \right)^{N_g} \]  \hspace{1cm} (6.4)
STARS™ does not allow the option of only using gas-oil curves, even though those were the only two phases present in Vapex experiments conducted in dry sand. After adjusting the residual endpoints for oil and water to match the observed residual saturations in the experimental study, relative permeability was not adjusted further to achieve the history match. The relative permeability curves for selected experiments in Sand#3 for history matching are presented in Figure 6.2. Considering the lack of any forced displacement in these experiments to overcome the end effect, the relative permeability curves had lower values compared to other simulation studies conducted by other researchers (Cuthiell et al. 2006; Ayub and Tuhinuzzaman, 2007; Yazdani and Maini, 2009b).

![Figure 6.2: Relative permeability curves of Sand#3 in conducted simulation study](image)

The relative permeability curves in vapor chamber and collection zones of the physical model are shown in Figure 6.3. The oil and water relative permeability in vapor chamber was forced to be extremely low in order to minimize the flow of liquid phase into this zone as it was observed in experimental study. In contrary in the collection zone at the bottom of the model, the gas relative permeability was minimized in order to collect and produce all the drained liquid from the sand pack at the top of this zone. Application of this approach results in minimization of transmissibility of the liquid phase and gas phase.
into the vapor chamber and collection zone, respectively as it was experimentally simulated.

Figure 6.3: Relative permeabilities in simulation (a) vapor chamber and (b) collection chamber

6.2.4. Diffusion and Dispersion Modeling in The Vapex Process

The dispersive flux of components $i$ ($i=1,2,3,...,n_c$) in phase $k$ of a multiphase multi-component system in a porous media can be is expressed as Equation 6.5.

$$J_{ik} = -\phi D_{ik} \nabla C_{X_{ik}}$$

(6.5)

The porosity is corrected for the reduced cross-sectional area in porous media and the negative sign indicates that the dispersive flux is in the direction of decreasing
concentration. Equation 6.5 can be rewritten as Equation 6.6 to express the concentration in terms of mol fraction. Phase saturation, and molar density of the phase $k$.

$$ J_{ik} = -\frac{\phi S_k}{\phi S_k} D_{ik} \cdot \nabla_k (\rho_k X_{ik}) $$

(6.6)

The dispersion coefficient depends on the direction of flow and is a second order tensor. Molecular diffusion and mechanical dispersion, and later bulk phase convective velocity are contributing in dispersion coefficient. The elements of the dispersion tensor can are presented in Equation 6.7.

$$ D_{ik} = \begin{bmatrix} D_{ix} & D_{iy} & D_{iz} \\ D_{ix} & D_{iy} & D_{iz} \\ D_{ix} & D_{iy} & D_{iz} \end{bmatrix} $$

(6.7)

The elements of the dispersion tensor are given in Equation 6.8 (Bear, 1988; Fayers et al., 1989).

$$ D_{ik}^{mn} = \frac{D_{ok}}{F_k \phi S_k} \omega_{mn} + \alpha_{ik} \left[ \frac{u_k}{\phi S_k} \omega_{mn} + \frac{(\alpha_{kl} - \alpha_{kl})}{\phi S_k} u_{km} u_{kn} \right] $$

(6.8)

Where

$$ \omega_{mn} = \begin{cases} 1 & \text{for } m = n \\ 0 & \text{for } m \neq n \end{cases} $$

(6.9)

$$ |u_k| = \sqrt{u_{kx}^2 + u_{ky}^2 + u_{kz}^2} $$

(6.10)

The Darcy velocity in Equation 6.10 can be calculated by Equation 6.11.

$$ u_j = \left( \frac{k k_{j,i}}{\mu_j} \right) \left\{ \nabla P + \nabla P_e - \rho g \nabla d \right\} $$

(6.11)

Based on assumption of no adsorption and chemical reactions, the mass conservation equation for an arbitrary components $i$ in a control volume can be written in the form of Equation 6.12.
\[
\begin{aligned}
    &\oint_A \left( \sum_k \rho_k X_{ik} u_k \right) dA + \oint_A \left( \sum_k \phi \rho_k S_k D_{ik} \nabla X_{ik} \right) dA \\
    &+ \oint_V \left( \sum_k q_{ik} \right) dV = \frac{\partial}{\partial t} \oint_V \left( \sum_k \phi \rho_k S_k X_{ik} \right) dV 
\end{aligned}
\]  

(6.12)

With application of divergence theorem to the first two terms of the left-hand side of the Equation 6.12 and assumption of stationary control volume, Equation 6.13 can be derived (Shrivastava et al., 2002).

\[
\begin{aligned}
    &\oint_V \left( \sum_k \nabla \cdot (\rho_k X_{ik} u_k) \right) dV + \oint_V \left( \sum_k \nabla \cdot (\phi \rho_k S_k D_{ik} \nabla X_{ik}) \right) dV \\
    &+ \oint_V \left( \sum_k q_{ik} \right) dV = \frac{\partial}{\partial t} \oint_V \left( \sum_k \phi \rho_k S_k X_{ik} \right) dV 
\end{aligned}
\]  

(6.13)

According to assumption of arbitrary control volume, the integrands on both sides of Equation 6.13 must be equal. This results in Equation 6.14, which is a molar conservation equation.

\[
\begin{aligned}
    &- \sum_k \nabla \cdot (\rho_k X_{ik} u_k) dV + \sum_k \nabla \cdot (\phi \rho_k S_k D_{ik} \nabla X_{ik}) dV + \sum_k q_{ik} dV = \frac{\partial N_i}{\partial t} 
\end{aligned}
\]  

(6.14)

where

\[
N_i = \sum_k (\phi \rho_k S_k X_{ik}) 
\]  

(6.15)

In Equation 6.12, the integral over the grid block volume bounded by external area of \( A_s \) can be approximated by Equation 6.16.

\[
\begin{aligned}
    &- \sum_k \sum_s \left( A_s \rho_k X_{ik} u_k \right) + \sum_k \sum_s \left( A_s \phi \rho_k S_k D_{ik} \nabla X_{ik} \right) + \sum_k q_{ik} \\
    &= \frac{(VN_i)^{n+1} - (VN_i)^n}{\Delta t^n} 
\end{aligned}
\]  

(6.16)

where \( n \) and \( n+1 \) refer to previous and current time levels.
In Equation 6.16, the first term on the left-hand side represents the convection flux, the second term the dispersive flux, and the third term the source/sink. The only term on the right-hand side represents the accumulation.

In CMG’s STARS™, the total dispersion coefficient, which represents the combined diffusion and convective dispersion effects, has the following forms for isotropic dispersivity cases (Equation 6.17) as proposed by Nghiem et al. (2001).

\[ D_{ik} = \frac{D_{ik}}{F} + \alpha \mu_k \]  

(6.17)

With regard to the issue of longitudinal versus transverse dispersivity, it is assumed implicitly that the predominant flow is in one grid direction throughout the simulation. In tensor terms, the dispersion tensor is diagonal. The longitudinal value can be assigned to the predominant flow direction, and transverse value can be assigned to the other two directions.

The accuracy, stability, and consistency of a numerical solution strongly depends on the grid sizes, time steps, and numerical method. Further to extremely small pressure difference across the experimental setup due to gravity (i.e., 2.35 and 4.56 kPa), a system of extremely fine grids blocks (2.5 mm×2.5 mm×2.5 mm) and small time steps in the order of $10^{-7}$ to$10^{-3}$ hours were utilized to improve the accuracy of predictions. All the grid blocks were treated in a fully implicit manner to avoid the instability of the numerical model and minimize material balance errors.

### 6.2.5. Simulation Constraints

In simulation phase, the models were explicitly initialized by defining saturation and composition of each phase at the start of the simulation runs within each single grid
block. In addition to composition, the initial pressure and temperature were set to initial condition of the physical models in order to let the simulator calculate the phase behavior properties of each phase and fluid in place at the start of the simulation. The fluid in place in each simulated model was checked against the experimental values to ensure accurate simulations.

Inhibition of movement of the drained oil from the sand pack into the solvent chamber was one of the major constraints of these simulations. This constraint should be defined in the numerical model without blocking the surface for diffusion of the solvent into the sand pack. For this purpose, the transmissibilities of the blocks on the edge of the solvent chamber were modified to stop the liquid phase from entering into this chamber. Transmissibility, as defined by Equation 6.19, depends on permeability, relative permeability, molar densities, viscosities, and grid block sizes. In this simulation study, the relative permeabilities to the liquid phases were forced to be extremely low values to make the edge grids to behave as a membrane with selective permeability to a specific phase.

\[
T_k = \left( \frac{kA}{\Delta l} \right) \left( \frac{k_r \rho_k}{\mu_k} \right) \tag{6.19}
\]

where

\[
kA = \begin{cases} 
\frac{\Delta y \Delta z}{\Delta x} & \text{in the } x \text{-direction} \\
\frac{\Delta z \Delta x}{\Delta y} & \text{in the } y \text{-direction} \\
\frac{\Delta x \Delta y}{\Delta z} & \text{in the } z \text{-direction}
\end{cases} \tag{6.20}
\]
Similar to the conducted experiments, the simulated vapor chamber was equipped with an injection well with injection pressure constraint of 200 kPa with capability of injecting butane vapor into this chamber at variable rates. Similarly, the production well was set at the bottom of the model to capture and produce all the drained oil from the sand pack under the effect of gravity. In order to simulate the boundary conditions of the Vapex experiments, except pressure constraints no other constraints were put on injection and production ports in the simulation models.

The numerical simulation work was performed using two PC machines. The first desktop computer had dual core 3.00 GHz processors and 4.00 GB RAM with a 32-bit operating system. The second PC had quad core 2.33 GHz processors with 6.00 GB RAM on a 64-bit Windows™ operating system. Each single simulation run took between 24 to 250 hours to complete due to very fine grid blocks of the model and extremely tight numerical constraints, which were designed to minimize the material balance and numerical errors. These constraints were defined and refined during the course of simulation study based on analysis of output files and acceptable error range required for simulation of smooth evolution of vapor chamber into the oil-saturated sand pack.

6.2.6. Film Flow in a Vertical Plane

A liquid film with a constant thickness attached to a smooth wall start to flow under its own gravity. Under unsteady state condition, the film thickness varies with height and time. The flow under the above conditions, ignoring the inertia and surface tension effects, can be expressed by Equation 6.21 derived from making material balance over a small elements $z$ in the vertical direction (Saidi, 1987).
As \( \Delta z \) and \( \Delta t \) approach zero, Equation 6.21 can be written as Equation 6.22. Equation 6.22 is also known as continuity equation.

\[
- \frac{\partial TV_z}{\partial z} = \frac{\partial T}{\partial t}
\]

(6.22)

The initial and boundary conditions are given in Equation 6.23.

\[
\begin{align*}
&at \quad t = 0 \quad T = T_i \quad for \quad 0 \leq z \leq L \\
&at \quad t > 0 \quad T = 0 \quad at \quad z = o
\end{align*}
\]

(6.23)

with substitution of the average developed for steady state condition, Equation 6.22 can be written as Equation 6.24 (Saidi, 1987).

\[
- \frac{T^2 \rho g}{\mu} \frac{\partial T}{\partial z} = \frac{\partial T}{\partial t}
\]

(6.24)

Equation 6.25 is the solution for differential Equation 6.24. Equation 6.25 gives the thickness of the film at different time.

\[
T^2 = \frac{\mu \xi}{\rho g t}
\]

(6.25)

The volume of the film from the top to the depth \( z \) can be calculated by integration of Equation 6.25 over the width of the vertical plate.

\[
V = \omega \int_0^z Tdz
\]

(6.26)

Substituting \( T \) from Equation 6.25 into Equation 6.26 gives the total volume of film, per unit width of the wall.
\[ V = \frac{2}{3} \alpha \left( \frac{\mu \varepsilon^3}{\rho g t} \right)^{0.5} \]  

(6.27)

The flow rate can be calculated by differentiating Equation 6.27 with respect to time can be determined by Equation 6.28.

\[ Q = -\frac{dV}{dt} = \alpha \left( \frac{\mu \varepsilon^3}{\rho g t^3} \right)^{0.5} \]  

(6.28)

### 6.3. History Matching Study

Conducted analytical modeling based on proposed model by Das and Butler (1998) showed that history matched effective diffusion coefficients for observed stabilized drainage rates were in the range of reported molecular diffusion coefficients of alkane-heavy oil systems in the literature. Considering relatively similar behavior of the production trends in experiments conducted in different sands in this study and available computational power, the history matching study focused on experiments, which were conducted in Sand#3. The grid size of the simulation models were 2.5 mm with aspect ratio of one that was selected after conducting a sensitivity study of grid size on quality of predictions, computation times, and material balance of simulations. On this basis, the total number of grid blocks in simulation model of the small experimental model was 1960 (20×1×98). This model was built to simulate the control volume with the thickness of 2.5 mm out of total width of the experimental models, as explained in Sections 6.1 and 6.2.

Based on the adopted experimental approach, the major focus in history matching process was matching the observed oil drainage rates and measured solvent uptake of the oil phase, as back calculated from the analysis of the collected produced oil samples during
the Vapex experiments. In addition, it was tried to match the measured cumulative oil and gas production values during the course of experiments.

Figures 6.4 to 6.7 show the best achieved history matching for the selected experiments in Sand#3 (i.e., Experiment#5 and #11). The conducted simulation study investigated the effect of capillary forces and presence of immobile water saturation on the rate of mass transfer between the solvent and oil phase. These figures present the comparison between the measured and simulated drainage rates and n-butane content of the produced oil during the course of experiments. Figure 6.8 shows the contour map of oil saturation change at two different stages of these experiments.

As it can be seen in Figures 6.4 to 6.7, there is a satisfactory history match for simulated experiments. The simulated oil drainage rates are in agreement with observed drainage rates at the pseudo-steady state regime with a relatively stable and constant drainage rate. In these figures, the simulation results showed some deviation from measured drainage rates at the early stage of experiments (e.g., first 200 hours of experiment in Experiment#5).

This deviation can be attributed to film flow which is happening because of presence of a small volume of the oil at the interface of oil saturated porous media and solvent saturated chamber that could flow in absence of capillary forces. Modeling of early stages of oil production by film flow theory for Experiment#5 is presented in Figure 6.4, as shown, predicted drainage rates based on film flow model closely matches the early time production.
Figure 6.4: Comparison between the measured and predicted oil drainage rates in Experiment#5 in absence of immobile water saturation

Figure 6.5: Comparison between the measured and predicted dissolved gas volume in Experiment#5 in absence of immobile water saturation
Figure 6.6: Comparison between the measured and predicted oil drainage rates in Experiment#11 in presence of immobile water saturation

Figure 6.7: Comparison between the measured and predicted dissolved gas volume in Experiment#11 in presence of immobile water saturation
Figure 6.8: Oil Saturation change at two snapshots predicted by history matched model in Experiment#5: (a) 632 hours and (b) 1127 hours since start of the experiment
Presence of high drainage rates at the early stage of experiments observed in experimental study, which was followed by stabilization of the drainage rates was the common characteristic of all the experiments conducted in this research. Despite the negligible error in predictions at the early stages of these simulations, numerical model could accurately capture the physics of the mass transfer phenomena in the mixing layer and the comparison between the trend of the experimental and numerical values represented a good match.

The spikes in simulated oil drainage rates in Figures 6.4 and 6.7 are due to STARSTM using automatic time step control depending on the maximum saturation and pressure changes occurring within the simulation model.

Figure 6.4 integrates the comparison between observed drainage rates at the early stage of Experiment#5 and calculated film flow rates using Equation 6.28. As it can be seen in this figure, there is strong correlation between the predicted film flow rates and observed rates for the first 72 hours of experiment. After 72 hours, the film flow slows down and the oil drainage from the sand pack takes over as n-butane diffuses into the oil phase in the sand pack and mobilizes the oil phase. At this stage, the solvent-vapor interface moves into the porous media and the drainage of the diluted oil by solvent occurs in porous media. In the transition period which is occurring between 72 and 200 hours, combination of drainage rates as film flow and numerically simulated drainage rates are in the range measured drainage rates in Experiment#5.

Figures 6.9 to 6.10 compare the simulated and observed position of the vapor-oil interface at three different periods of the experiments. Experimentally the interface position in these experiments were determined by image proceeding of the photos.
captured during the experiments from the visual side of the model (Figure 11). In these figures, the interface position in the simulation model was determined based on the criterion of 5% gas saturation in the last grid block at the end of the simulated sand pack. As it can be seen in these figures, simulated interface positions are in good agreement with the observed interface position. It should be mentioned that 5% saturation change limit was arbitrary and based on comparison between the numerical simulation and captured vapor-oil interface images. In these experiments, non-intrusive measurement of gas saturation in drained sand pack was not possible; therefore, this is a qualitative match. Based on the achieved history match for above experiments the effective diffusion coefficients were in the range of $1.4 \times 10^{-7}$ and $3.5 \times 10^{-6}$ cm$^2$/s that are similar to values observed from conducted analytical modeling based on proposed Butler’s model (1998).

![Figure 6.9: Measured and predicted vapor-oil interface evolution rate in Experiment#5](image)

**Figure 6.9: Measured and predicted vapor-oil interface evolution rate in Experiment#5**
Figure 6.10: Measured and predicted vapor-oil interface evolution rate in Experiment#11

Figure 6.11: Solvent-oil interface movement in Experiment#5 at four different times
The comparison between effective molecular diffusion coefficients for experiments in presence and absence of aqueous phase saturation shows the positive impact of aqueous phase on mass transfer rate of solvent into the oil phase. The presence of immobile aqueous phase saturation resulted in 110% increase in solvent content of the oil phase despite similar boundary condition and operating conditions in Experiment#5 and #11. The results of achieved history match provides a clear indication that the mechanical dispersion is not an important mechanism controlling the mass transfer rate between the solvent and the oil phase on the edge of the vapor chamber.

6.4. Numerical Dispersion

Most of reservoir simulators obtain solutions to fluid flow equations (i.e., nonlinear PDEs) by replacing derivatives with finite-difference approximations (Aziz and Settari, 1979; Abou-Kassem et al., 2006). These approximations were derived by manipulation of Taylor’s series that could introduce error into approximations known as truncation error. The ranges of error of approximations are very small in wide range of problems that makes these models sufficiently accurate for engineering problems. However, the truncation error can result in significant solution inaccuracies for certain types of problems, such as miscible floods and immiscible floods in which viscous forces are much larger than the capillary forces (Peaceman, 1977).

In simulation of the Vapex process where diffusion and dispersion are playing important roles in the efficiency of the process, approximation of the range of numerical dispersion is required for accurate estimation of effective diffusion coefficient using history matching of the conducted experiments.
In order to approximate the range of numerical dispersion, proposed analytical expressions for multidimensional numerical dispersion by Fanchi (1983) were used. These expressions were derived by performing a truncation error analysis on a 3D convection-dispersion equation similar to Equation (6.20). Fanchi et al. (1982) presented the details of the truncation error analysis in Appendix A of this reference. For a 2-D problem and an implicit backward difference in space and implicit-in-time scheme the numerical dispersion tensor is defined by velocity component, grid sizes, and size of time steps (Equation 6.21).

\[
\phi \frac{\partial S}{\partial t} = D_x \frac{\partial^2 S}{\partial x^2} + D_y \frac{\partial^2 S}{\partial y^2} + D_z \frac{\partial^2 S}{\partial z^2} - u_x \frac{\partial S}{\partial x} - u_y \frac{\partial S}{\partial y} - u_z \frac{\partial S}{\partial z} \tag{6.20}
\]

\[
D_{\text{num}} = \begin{bmatrix}
\frac{v_x}{2} \left( \Delta x + \frac{\Delta t}{\phi} \right) & v_x v_z \frac{\Delta t}{\phi} \\
\frac{v_y}{2} \left( \Delta y + \frac{\Delta t}{\phi} \right) & \frac{v_y v_z}{2} \left( \Delta z + \frac{\Delta t}{\phi} \right)
\end{bmatrix}
\tag{6.21}
\]

Statistical summary of grid size, effective diffusion coefficient, average time step size, and average velocities were obtained from simulation model. Calculated numerical dispersion for history matched simulation runs is presented in Table 6.3. For simplification, it was assumed that the x-direction is oriented in the flow direction \((v=v_x)\) and \(v_z=0\).

In Table 6.3, calculated numerical diffusion using average time step and oil phase velocities calculated are three orders of magnitude smaller than the effective diffusion coefficient derived from history match models. This shows that the estimated effective diffusion through history matching process is a representative estimation of the rate of solvent dissolution into the oil phase which was not masked by numerical dispersion.
based on average time steps, grid block size used by the simulator to calculate the velocities.

**Table 6.3: Summary of comparison between effective diffusion and calculated numerical dispersion based on proposed approach by Fanchi (1983)**

<table>
<thead>
<tr>
<th>Experiment No.</th>
<th>Time Step (s)</th>
<th>Velocity (Cm/s)</th>
<th>$D_{\text{eff}}$ (cm$^2$/s)</th>
<th>$D_{\text{num}}$ (cm$^2$/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5</td>
<td>$1.0 \times 10^{-4}$</td>
<td>$4.57 \times 10^{-7}$</td>
<td>$1.00 \times 10^{-7}$</td>
<td>$2.94 \times 10^{-10}$</td>
</tr>
<tr>
<td>6</td>
<td>$1.0 \times 10^{-4}$</td>
<td>$1.51 \times 10^{-6}$</td>
<td>$3.50 \times 10^{-6}$</td>
<td>$3.20 \times 10^{-9}$</td>
</tr>
</tbody>
</table>
CHAPTER 7: CONCLUSIONS AND FUTURE RESEARCH

This dissertation presented the results of a combination of experimental, analytical, and numerical investigations of interplay of drainage height, capillary forces, and immobile water saturation on the rate of solvent mass transfer and performance of the Vapex process. The following conclusions and recommendations were drawn from this study.

7.1. Conclusions

The findings in this study have shown that the penetration depth of the solvent into the oil phase is greatly controlled by the viscosity of the oil phase, capillary force that is a function of pore structure and morphology of the medium, and the drainage height. Additionally, it was observed that the effective diffusivity of the tested solvent in presence of capillarity forces is similar to the molecular diffusion coefficients reported in literature, therefore, unrealistically high diffusion coefficients typically used, in simulation studies, to match the experimental results could lead to unrealistic production rate predictions.

Based on the results obtained from this study, the following conclusions are made:

- A new experimental approach was designed and successfully implemented for realistic simulation of the mass transfer phenomenon in the mixing zone on the edge of the Vapor chamber in the Vapex process. This was achieved by elimination of forced displacement occurring in traditional experimental approach. The newly developed experimental approach made it possible to conduct the Vapex experiments in absence of any pressure disturbances, which could have caused additional mixing of the solvent and the oil phase and resulted
in unrealistically higher oil drainage rate.

- The comparison between the observed drainage rates in Vapex experiments conducted at higher permeabilities (>220 D) and measured stabilized drainage rates in this study clearly showed that permeability, capillary forces, and presence of immobile aqueous phase saturation were playing important roles in the Vapex process by controlling the rate of mass transfer between the solvent and oil phase.

- The stabilized drainage rates and solvent content of the produced oil samples were higher for experiments conducted in presence of an immobile water phase. Solvent content measurement of produced oil samples revealed that that the solvent content of the oil phase was always lower than the equilibrium values predicted by a phase behavior model tuned using conducted phase behavior study in this research. The solvent content of the oil phase in experiments conducted in dry sand varied between 15 and 51 mol%. In experiments conducted in wet sand, the solvent content of oil phase was in range of 48-53 mol%.

- Conducted analytical and numerical modeling showed that the effective diffusion coefficient was in the range of $4.91 \times 10^{-8} - 1.10 \times 10^{-5}$ cm$^2$/s. On this basis, it can be concluded that molecular diffusion is the major driver for the mass transfer phenomena at higher capillary pressures (i.e. lower permeability formations similar to the western Canadian heavy oil reservoirs).

- The effective diffusion of the butane vapor into the heavy oil saturated sand pack in absence of immobile water saturation was in the range of $4.91 \times 10^{-8} - 7.89 \times 10^{-6}$ cm$^2$/s. The back calculated values for effective diffusion coefficients (i.e., analytical and numerical studies) showed dependency on the drainage height and
capillary pressure of the sand pack. In other words, with increase of drainage height and reduction of the effect of capillary forces, the effective diffusion coefficient of the butane vapor was improved up to two orders of magnitude.

- The effective diffusion coefficients in experiments conducted in the presence of immobile water saturation were in the range of $2.96 \times 10^{-6} - 1.1 \times 10^{-5}$ cm$^2$/s. The back calculated effective diffusion coefficients in presence of immobile water saturation were higher than the values observed in experiments conducted in the similar sand packs in absence of water saturation. The back calculated effective diffusion coefficient in these experiments were in narrower range compared to experiments conducted in dry sand.

- A multi-variable correlation was developed for the prediction of effective diffusion coefficients. This correlation is a concentration-dependent relationship that accounts for drainage height and the capillarity effect. In order to estimate the effective diffusion coefficient, this correlation employs a combination of solvent solubility, specific pore surface area of medium measured by mercury injection data, and the drainage height of the medium.

- A comprehensive simulation study was conducted in order to history match the measured drainage rates and solvent content of the produced oil from the experiments conducted in Sand#3. This study revealed that effective diffusion coefficients (i.e. $1.00 \times 10^{-7}$ and $3.50 \times 10^{-6}$ cm$^2$/s) were in the range of back-calculated values using Das and Butler model.

- The numerical simulation in this study showed that very fine grid blocks and small time steps were required to capture the complexity of the mass transfer in
the mixing zone in the Vapex process. The comparison between the material balance errors, small time step sizes, and convergence rate of simulations showed discretization of the mixing zone using the minimum of two grid blocks could provide a better simulation of the process and minimized the numerical dispersion.

- The calculated numerical dispersions, based on the results of history matching exercise, were in the range of $2.94 \times 10^{-10} - 3.20 \times 10^{-9}$ cm$^2$/s. On this basis, it could be concluded that the derived effective diffusion coefficients were relatively accurate and numerical dispersion did not mask the physical diffusion occurring during the Vapex process.

- This study showed that selection of a realistic mass transfer coefficient was required for simulation of the performance of the Vapex process. This means that the proposed effective diffusion/dispersion coefficients up 4 orders of magnitude higher that the molecular diffusion coefficients in the literature could not be used for simulation of the Vapex process at reservoir condition.

7.2. **Future Works**

- Deasphalting precipitation is an important phenomenon that can affect the rate of mass transfer and consequently the performance of the Vapex process. This phenomenon can significantly impact the viscosity of oil phase, pore morphology and wettability of the porous media. Conducted research showed the strong dependency of destabilized drainage rates on specific pore surface area of the medium. Therefore, quantifying the effect of asphaltene deposition on the rate of mass transfer in lower permeability range representing heavy oil and bitumen
reservoirs has merit and need be further analyzed and explored.

- Using finite difference model and commercial simulators showed the limitations of these simulators for capturing the complexity of the mass transfer phenomena in the Vapex process. It is recommended to investigate the application of other numerical techniques for discretization of the space and time (e.g., finite element and control volume). These numerical approaches can minimize numerical dispersion using higher order approximations and better handling of accuracy, convergence, and stability.

- This study showed that the efficiency of the Vapex process is controlled by the diffusion rate of the solvent into the heavy oil. Considering the positive impact of pressure pulsation on the mass transfer rate between the solvent and heavy oil, it is recommended to investigate the effect of this technique on the rate of solvent mass transfer in the Vapex process.

- Accurate simulation of multiphase flow in porous media requires application of realistic relative permeability curves. The conducted literature review in heavy oil literature showed that measurement of the relative permeability curves in experimentally simulated cases in this study has merit and needs to be further investigated.

- This study was conducted in absence of carrier gas (e.g., N₂, CO₂ and methane). Understanding the effect of potential impurities on the rate of mass transfer in the Vapex process can be beneficial for successful simulation and implementation of the solvent-based process in heavy oil and bitumen reservoirs.
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