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UNIVERSITY OF CALGARY

A Numerical Simulation Study of the $N\text{-}Solv^{TM}$ Process

by

Ke Cao

A THESIS

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Abstract

Given the search for recovery processes that are more efficient with lower water consumption and greenhouse gas emissions than Steam-Assisted Gravity Drainage (SAGD), solvent based recovery processes are a potential alternative. Vapor Extraction (VAPEX) and N-SolvTM are two potentially competing solvent recovery processes that offer water-free operation.

Although VAPEX has demonstrated success in laboratory experiments, it has not been demonstrated to be technically or commercially successful in the field due to the slow oil production rate. To improve oil recovery rate over that of VAPEX, the N-SolvTM process injects heated solvent into the reservoir to provide additional energy to the reservoir which enhances the reduction of the oil phase viscosity beyond that of solvent alone. The heated solvent is injected at pressures under its vapor pressure (at injection temperature) to ensure it is vapor. At the chamber interface, the solvent condenses releasing latent heat to the oil sand. Condensed solvent diffuses into bitumen and dilutes it reducing its viscosity to achieve mobility and drainage of the oil to the production well under the action of gravity.

In the research documented here, the mechanisms and effectiveness of the N-SolvTM process in a McMurray oil sand reservoir is investigated by fine grid (8 cm by 8 cm grid blocks in the cross well pair plane) thermal reservoir simulation. A sensitivity analysis of the N-SolvTM process was also done to explore possible ways to enhance oil production and reducing solvent retention in the depleted reservoir. The results reveal that dynamics at the edge of the chamber – including phase change, oil viscosity reduction by elevated temperature and solvent diffusion, and the interaction between heat and mass transfer are complex. Also, in a dead oil reservoir, the injection of hot propane into oil sands reservoirs can yield comparable oil production rates to that of SAGD. Furthermore, solvent loss, which is a major concern, can be reduced by injecting at superheated conditions and relatively low pressures. Lastly, in the presence of solution gas and high connate water saturation in the reservoir, the performance of N-SolvTM suffers significantly.

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To my dear parents

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List of Symbols, Abbreviations and Nomenclature

Symbols

Cs	solvent concentration, volume fraction
$C5_+$	alkanes having more than 5 carbons
d _p	average particle diameter of the medium, m
D	diffusivity, m ² /s
Do	molecular diffusivity, m ² /s
D _B	intrinsic diffusivity of bitumen into toluene, m ² /s
D _S	intrinsic diffusivity of toluene into bitumen, m ² /s
D _{BS}	overall/mutual diffusivity in toluene-bitumen system, m ² /s
D _p	apparent diffusivity, m ² /s
F	formation electrical resistivity factor
h	drainage height, m
J	mass flux in Fick's law, mol/m ² -s
k	permeability, mD
k _{ro}	oil phase relative permeability, mD
k _{rw}	water phase relative permeability, mD
k _{rg}	gas phase relative permeability, mD
K _L	longitudinal dispersion coefficient, m ² /s
K _T	transverse dispersion coefficient, m ² /s
m	mass flux rate, m^3/d
N _s	dimensionless number
Q	volumetric flow rate per unit length of a production well, m^2/d
T _{sat}	saturation temperature, °C
U	superficial fluid velocity, m/s
x	diffusion distance, m
ΔS_{o}	change in oil saturation of region swept by solvent

Δρ	density difference between diluted bitumen and solvent, $\mbox{kg/m}^3$
α	characteristic coefficient of the solute-solvent pair
β	characteristic coefficient of the solute-solvent pair
φ	porosity
μ	viscosity, cP or mPa.s
ρ	solvent-bitumen mixture density, kg/m ³
θ	dip angle of the drainage front
Ω	cementation factor

Abbreviations

2D	two-Dimensional
ARC	Alberta Research Council
bbl	barrel
BPT	Bubble Point Temperature
CHOPS	Cold Heavy Oil Production with Sands
cm	centimeter
cP	centipoise
CSS	Cyclic Steam Stimulation
EOR	Enhanced Oil Recovery
ES-SAGD	Expanding Solvent-SAGD
GHG	Green House Gas
GOC	Gas Oil Contact
kPa	kilopascal
IFT	Interfacial Tension
LASER	Liquid Addition to Steam for Enhanced Recovery
mD	millidarcy
mPa.s	millipascal-second
NCG	Non-Condensable Gas
OSR	Oil Steam Ratio

SAGD	Steam-Assisted Gravity Drainage
SAP	Solvent Additive Processes
SA-SAGD	Solvent-Assisted SAGD
SAS	Steam Alternating Solvent
SOR	Steam Oil Ratio/Solvent Oil Ratio
UTF	Underground Test Facility
VAPEX	Vapor Extraction
WOC	Water Oil Contact
wt	weight

Chapter 1: Introduction

1.1 Background

Every year the Oil and Gas Journal updates the proven oil reserves worldwide (reserves are the fraction of an oil deposit which can be commercially produced with existing recovery technology). On Jan 1st 2014, the total world oil reserves were estimated at 1.64 trillion bbl. Venezuela has the biggest oil reserves with 297.7 billion bbl, followed by Saudi Arabia with 265.8 billion bbl. Canada is the third largest oil reserve country, possessing 173.2 billion bbl (Oil and Gas Journal, 2013), of which 98% is in the form of heavy oil and bitumen, which fall under the unconventional resources category.

Unconventional resources exist in petroleum accumulations that are pervasive throughout a large area. They are oils that aren't produced by drilling conventional, vertical wells and using the reservoir's pressure along with secondary oil recovery techniques to extract the oil. Examples of unconventional resources include shale gas, tar sands, heavy oil, bitumen, shale oil, gas to oil and biofuels. Unlike conventional resources, unconventional resources require specialized extraction technology to produce. Enhanced oil recovery (EOR) needs to be employed, for example, steam and/or solvents are used to mobilize bitumen for in-situ recovery. Moreover, the extracted petroleum may require significant processing prior to sale (e.g., bitumen upgraders). Figure 1.1.1 shows the composition of total world oil reserves by type. The total amount of unconventional oil resources in the world considerably exceeds the

1

amount of conventional oil reserves, but is much more difficult and expensive to develop. Note that this plot is made in 2006, not accounting for the large amount of shale gas that has been discovered and exploited in the United States in recent years.



Total World Oil Reserves

Figure 1.1.1: Total world oil reserves by type (Alboudwarej et al., 2006)

As oil demand continues to rise and conventional oil is being produced, the world is turning towards unconventional oil. With about 1.7 trillion barrels of heavy oil and oil sands resources, of which 10% are currently recoverable, Canada has a key role to play in the global petroleum fossil fuel market.

EOR methods that use steam to recover unconventional oil consume large volumes of water and generates large amount of greenhouse gases (GHG). Solvent injection based techniques can be employed to recover unconventional oil and it offers an energy efficient and environmental friendly alternative method to steam injection based techniques. This research work attempts to explore possibilities of solvent based techniques.

1.2 Heavy Oil and Bitumen

Heavy oil and bitumen are the two most popular types of unconventional oil found in Canada. Oil is commonly categorized based on its API gravity, which is a measurement of how heavy/light it is compared to water (if it's greater than 10, the oil is lighter and floats on water; if it's less than 10, the oil is denser and sinks in water). An API gravity lower than 10° and viscosity higher than 10,000 centipoise is considered as bitumen, whereas oils with API gravities between 10° and 22° and viscosities between 100 and 10,000 centipoise are considered heavy oils (USGS, 2007). There is an inverse correlation between oil's API gravity and oil viscosity – the lower the API gravity of oil, typically the higher is its viscosity. There are exceptions to this rule, with in-situ conditions also playing an important role in oil viscosity. Heavy oils and bitumen found in Canada have viscosities ranging from several thousand to several million centipoise, which is similar to the viscosity of caulk.

The chemical and physical properties of heavy oils vary greatly from those of conventional light oils. Beyond the higher viscosity, bitumen has high asphaltene (much higher than that of light oils), high sulfur (4% and higher whereas light oils are typically between 1 and 4%), and heavy metals (nickel, vanadium, and iron content). Asphaltene can pose a problem in recovery process when asphaltene precipitation occurs. Heavy metals are considered

environmental contaminants and cause coking and catalyst poisoning in refineries. Bitumen also consists of more of high molecular weight hydrocarbons and less of lighter ends than conventional crude, such as gasoline and diesel that generate more revenue. In terms of composition this leads to a deficiency in hydrogen compared to lighter oils (the average light oil is 14% hydrogen compared to 10% in heavy oils (NEB, 2000)), which makes it necessary to upgrade the oil to synthetic crude oil for refining and pipelining. All of these factors contribute to energy intensive recovery process with waste products and lower the value of the oil.

1.3 Canada's Heavy Oil and Bitumen

Canada's oil reserves are mostly found in western Canada, mainly in Alberta and Saskatchewan, as shown in Figure 1.3.1. The three main bitumen deposits are the Athabasca, Cold Lake and Peace River oil sands. Heavy oil and bitumen deposits account for 98% of Canada's current oil reserves (EIA, 2012) and for 66% of Canada's daily oil production totaling approximately 2.1 million bbl/day (CAPP, 2012). Heavy oil is found near the border of Alberta and Saskatchewan with approximately 1.3 billion barrels of reserves (NEB, 2011). Although not as prolific, the heavy oil deposits are attractive due to their low capital costs and operating expenses and are often sufficiently mobile to be produced under primary recovery process such as CHOPS (cold heavy oil production with sands). A summary of physical properties of three Canadian major oil deposits is listed in table 1.3.1.



Figure 1.3.1: Heavy oil and oil sands deposits in Alberta (Peacock, 2010)

Conventional oil has a low viscosity (< 100 mPa.s) and is typically found in deep, high pressure reservoirs where primary recovery is due to the pressure difference between the reservoir and well. In contrast, the heavy oil and bitumen in western Canada are generally shallower (current producing fields are found at the surface to a maximum depth of 800-1000 m and have low reservoir pressures), the oil viscosities are much higher (often immobile at reservoir conditions), and found in unconsolidated sandstone (porosity ranges from 26-32% and the permeability from 1200- 7500 mD) and fractured carbonates with average of 20 m pay zone.

Location	Viscosity at 12 °C,	API Gravity	Physical composition (wt/wt%)		
	cP		asphaltenes	resins	oils
Athabasca	2,000,000	8	15.7	34.1	49
Cold Lake	60,000	10	16	28.7	55.6
Lloydminster	5,000	15	13	38.4	48.7

Table 1.3.1: Oil properties of three major Canadian oil deposits (modified from Speight, 1991)

1.4 Recovery Techniques

The high viscosity and shallow depth (low pressure) of the heavy oil and bitumen make it act as solids under reservoir conditions and therefore very difficult to produce economically using conventional pressure-drive production techniques. Two major techniques to recover heavy oil and bitumen are surface mining and in-situ production (EOR) techniques.

Surface mining involves digging up the oil sand then transporting it to a treatment facility where hot water extraction is used to separate the bitumen from the sand. This technique can recover up to 75% of the oil; however it is only economical to produce oil that is less than 75 m below surface and only 20% of Canada's oil is found at these depths (NEB, 2000). In addition, two tons of sand must be mined to produce one barrel of oil, leaving a huge volume of sand containing the remaining 25% oil to be disposed of. Surface mining also leaves a large footprint requiring extensive land reclamation projects.

Improving the flow of oil by changing the properties of the crude oil in the reservoir is called in-situ enhanced oil recovery (EOR) technique. By reducing the crude oil's viscosity or changing the surface chemistry of reservoir fluids, the oil may begin to flow, allowing for production. The use of EOR techniques can bring oil recovery to over 60% of original oil in place (RWE, 2004). Common stimulation method to make the oil sufficiently mobile is adding heat to the reservoir to lower the oil viscosity. Oil viscosity is a strong function of temperature, as shown in Figure 1.4.1. Thermal methods increase the oil production rate by increasing the reservoir temperature with the addition of heat, thereby decreasing the viscosity. Heat is added in the form of steam, hot water or in-situ combustion. Currently popular thermal methods of bitumen production in Canada are steam assisted gravity drainage (SAGD) and cyclic steam stimulation (CSS). They are displayed in Figure 1.4.2 and 1.4.3 respectively.



Figure 1.4.1: Oil viscosity as a function of temperature (Raicar and Proctor, 1984)



Figure 1.4.2: Illustration of the SAGD process (Schlumberger, 2002)



Figure 1.4.3: Illustration of the cyclic steam stimulation process (CAPP, 2012)

The use of steam has some disadvantages. The industry uses a parameter SOR (steam-to-oil ratio) to measure the energy efficiency of thermal method by steam injection. Usually production of one barrel of oil requires two to four barrels of steam. Steam is generated by burning natural gas, therefore the cost of the process is sensitive to the price of natural gas. Burning natural gas also results in a large production of carbon dioxide (CO₂). Steam production also requires a large source of water, usually a lake or aquifer. After the steam condenses in the reservoir, both oil and water will be produced, thus separators are needed to remove and reclaim the injected water. Steam generation facilities account for about 30% of capital cost in SAGD (Das, 1995). A significant amount of surface equipment is required to produce steam and separate the produced oil-water mixture. Heat loss is a major concern in thermal methods. Despite heavy insulation, heat is lost in lines between the steam generators and the well bore, through the well casing, to the overburden and to ground water below the reservoir. Formation damage can also occur in reservoirs which contain more than 10% clay. When the steam condenses into water, clay will absorb the water, resulting in clay swelling. Clay swelling can lead to pore blockage, reducing permeability and limiting the oil production. For SAGD, operating at high temperature and pressure can also cause chamber confinement problem, this includes loss of steam into thief zones or lean zones and potential loss of overburden integrity. Another issue is the blowback of steam from new pads into previously depleted chambers, blowback could greatly increase SOR. In conclusion, thermal methods are only suitable for reservoirs having thick pay zones and no bottom aquifers, and reservoirs that have low clay content.

Another alternative for in-situ EOR technique is solvent addition to the reservoir to reduce heavy oil and bitumen viscosity. In this case the viscosity is reduced by the mass transfer of a light hydrocarbon solvent (normally propane, butane, and pentane) into the heavy oil and bitumen, to achieve low viscosity and mobility by dilution. Oil viscosity is also a strong function of the amount of hydrocarbon solvent in diluted bitumen. Figure 1.4.4 shows this relationship between solvent concentration and diluted bitumen viscosity.



Figure 1.4.4: Reduction of Elk Point heavy oil viscosity using butane as a solvent at 22 °C (Yazdani, 2007)

Two examples of solvent technique are Vapor Extraction (VAPEX) and N-SolvTM. Both of them use similar well configurations as SAGD. VAPEX injects vapor solvent into the reservoir below the solvent dew point pressure to avoid solvent condensation in the depleted chamber. Due to the limited vapor pressures of the light hydrocarbons, non-condensable

gases (such as methane, nitrogen) are injected together with the solvent to achieve the reservoir pressure and ensure that the solvent remains in its vapor phase. N-SolvTM also uses vaporized solvent, however it has some fundamental difference from VAPEX. First, solvent is heated to the temperature that is higher than that of reservoir, allowing solvent to condense and release its latent heat to the colder bitumen. Therefore N-SolvTM is a thermal-solvent technique, except that the thermal energy is provided by heated solvent, not steam. Second, instead of adding methane to the injected gases, N-SolvTM tries to eliminate methane from the reservoir. In order to achieve this, the injection temperature and pressure are maintained at bubble point conditions to ensure removing methane (methane as a solution gas is common in heavy oil and bitumen, it can be released from the bitumen as production proceeds) from the reservoir by condensing and producing it together with the diluted oil.

Compared to thermal oil recovery techniques solvent recovery techniques have a lower capital cost, requires less energy (VAPEX has only 3% of the energy required by steam processes (Das, 1998)), produces less pollution (without the need for steam production, it is estimated that VAPEX process produces 80% less CO₂ emissions than SAGD (Das, 2002)) and may lead to in-situ upgrading of bitumen, thus reduces the surface facilities for upgrading. Solvents are less likely to escape through the cap rock, they are insoluble in water, hence there is no loss of solvent through top or bottom water zones. After oil production has ceased, the majority of the solvent held in the reservoir can be recovered through a "blowdown" procedure (lower down reservoir pressure so the condensed retained solvent can be produced as gas).

The cost of producing oil from heavy oil and bitumen is much higher than production from conventional resources. Additional energy is required to get bitumen to the surface since artificial lift and EOR techniques are required. Once bitumen has reached the surface, substantial refining is required to produce the more useful lighter ends. Additional costs are needed to treat and dispose large amounts of waste products. Canada is the world leader in developing EOR techniques for in-situ heavy oil production, and Canadian researchers play an important role in reducing the cost and environmental impact through their dedication to innovation and technology.

1.5 Objectives and Thesis Outline

The objective of this thesis is to investigate the feasibility of N-SolvTM process through numerical simulation, and explore possible ways of optimizing this process. The thesis consists of five chapters as summarized below:

Chapter 2 presents a review of the literature on several in-situ EOR techniques, including VAPEX, solvent additive-SAGD, and N-SolvTM. The review also includes processes mechanisms, experimental studies and field pilot projects.

In Chapter 3, the oil extraction mechanisms involved in N-SolvTM, and its oil production performance compared to SAGD is explored by using fine-grid thermal reservoir simulation.

Chapter 4 describes a sensitivity analysis of the N-SolvTM process from an injection condition point of view. The impacts of solution gas are also studied.

Chapter 5 summarizes the conclusions of this research study, and makes recommendations for future research work.

Chapter 2: Literature Review

2.1 Abstract

Two ways of reducing the high viscosity of Alberta's oil sands are to either provide heat or dilute the oil with a solvent within the reservoir. The common heat supplier used in the industry is steam, known for its use in oil recovery technologies such as Steam-Assisted Gravity Drainage (SAGD), Cyclic Steam Stimulation (CSS), and steam flooding etc. The solvent method uses a hydrocarbon solvent. Hydrocarbon solvent can dissolve into bitumen, make it swell and diluted, thus reduce the viscosity. A well-known technology using solvent is Vapor Extraction (VAPEX).

Researches on solvent recovery initially attracted people's attention because of its inherent advantages in solvent leaching that are absent in steam recovery processes. Although steam recovery has proven its success in commercial utilization over 20 years, it is an energy intensive process, requiring burning huge amount of natural gas. The facility requirement is also big as a result. This high natural gas requirement results in large amount of CO₂ emission, which contributes the global warming. Furthermore, due to its high operating pressure and temperature, chamber confinement can be a problem; Heat efficiency is also an issue because of heat loss to overburden and underburden. As a result, steam recovery is not suitable for reservoirs that have large aquifer below it, or those that have thin pay zones. Solvent recovery technology, on the other hand, doesn't have the above mentioned issues. energy efficiency, which also means low foot imprint, low green house gas (GHG) emissions, therefore environmental friendly. However, no single recovery method is perfect; the problem with solvent lies in the slow solvent diffusivity in bitumen. Considered as a non-thermal process, solvent recovery is inherently slower due to its orders of magnitude lower mass diffusivity compares to heat diffusivity in thermal (steam) processes. Therefore it cannot generate commercial oil rate. Besides, solvent is very expensive and its loss in the reservoir is a concern. Whether it will be economical is a question.

Both of these two methods have their advantages and disadvantages, hence researchers have been trying to combine them, take advantages of both, and at the same time avoid their disadvantages. Several steam-solvent hybrid recovery methods were proposed, and some of them have been tested in the field and showed promising results. These methods include but not limited to, Expanding Solvent-SAGD (ES-SAGD), Solvent Additive Processes (SAP), Liquid Addition to Steam for Enhanced Recovery (LASER), Steam Alternating Solvent (SAS), etc. The use of solvent in SAGD is also called Solvent-Assisted SAGD (SA-SAGD).

N-SolvTM, is also a heat-solvent combined method, however, it does not uses steam but heated solvent as the heat supply. N-SolvTM has just started its pilot project. In this chapter, VAPEX, SA-SAGD, and N-SolvTM will be reviewed in details.

2.2 VAPEX

2.2.1. VAPEX Drainage Mechanisms (Macroscopic)

Reservoir fluids are subject to three types of force, gravitational force, viscous force and capillary force. Oil in western Canadian oil sands reservoirs is so viscous that it is hard to flow in its natural state. Once the viscosity has been reduced, gravity force can overcome the viscous force and capillary force, and the oil can flow under gravity. Assisted gravity drainage is the drainage mechanism that harnesses this feature of reservoir fluids. In such processes, two horizontal wells (one injector, one producer, usually 1000 meters long) are drilled in the reservoir, with the injector being placed several meters above the producer. VAPEX is one of the oil recovery methods that utilize this gravity drainage mechanism.

Figure 2.2.1.1 shows a typical well configuration of gravity drainage for VAPEX. It shows the cross section of two horizontal wells, A and B. Well A is the injector, and well B is the producer. In oil sands reservoirs the injector is normally located 5 meters above the producer. The injector and producer can be both in the same vertical plane, or they can be staggered in reservoirs with lower oil viscosity (Das, 1998). Vaporized hydrocarbon solvent (usually of light molecular weight) is injected into the reservoir in well A, contacts and dissolves in the bitumen in the reservoir. The dissolved solvent will further diffuse into the bulk bitumen and dilutes it. Bitumen therefore swells and become less viscous, then flows under gravity to the producer. If the solvent concentration in bitumen is high enough, it can leave the heaviest

molecules (asphaltene) behind in the extracted sand matrix and reduce the viscosity even more, only producing the lighter part of bitumen, the de-asphalted oil.

Once the bitumen between two wells is extracted by solvent, communication is achieved between the injector and producer. After this, vaporized solvent will rise up further in the reservoir and form a vapor solvent chamber (The pores drained of oil become filled with solvent known as the solvent chamber) in the extracted sand matrix above the injector. In this rising vapor chamber stage, the solvent contacts the bitumen at the vapor chamber and bitumen interface, repeats the dilution mechanism and gravity drainage process, until the vapor chamber reaches the caprock (horizontal boundaries) at the top of the reservoir. Once the chamber reaches the caprock, it starts to spread laterally until the vertical boundaries of the reservoir are reached. In this sideway spreading phase, gravity head will continue to decrease until the oil draining rate becomes so slow (gravity head decreases to a point that it becomes harder to overcome the capillary force and viscous force) that it's not economical to continue the process.



Fig. 2.2.1.1: The concept of the VAPEX process (Das and Butler, 1997)

Gravity head is directly proportional to the fluid height. As the height increases, the available gravity head increases proportionally but this effect is counterbalanced by the longer fluid travel length from the mixing point to the producer. Nevertheless, the height plays an important role in VAPEX and there are more physical phenomena involved in this effect.

In VAPEX vapor chamber height and its numerical relationship to the drainage rate has been investigated by many researchers. Butler's first analytical model (Butler and Mokrys, 1989) on solvent recovery process states that the flow rate is proportional to the square root of chamber height. Yazdani and Maini further investigated this relationship and concluded that flow rate has a stronger dependency on chamber height; their correlation indicates the height dependency exponent is between 1.126 and 1.172 (Yazdani and Maini, 2006). They attribute this stronger dependency to the higher convective dispersion caused by the height effect. Specifically, the penetration depth increases in proportion to the square root of time. Therefore the thickness of solvent-bitumen mixing zone is increased as the drainage height increases due to the longer path and exposure time. Convective dispersion is consequently larger due to this longer mixing time, which in turn enhances the oil rate.

Cuthiell and Edmunds (Cuthiell and Edmunds, 2012) did numerical simulations based on Yazdani and Maini's experiments (Yazdani and Maini, 2009a) and argued that the height dependent convective dispersion perspective might not be valid since the experiments showed an increasing solvent fraction in the produced oil phase as height increases. And this increased solvent fraction in the produced oil has a positive effect on production rates. They artificially removed this solvent mixing effect from their numerical simulations by changing the K values, and found out the remaining height dependence exponent is only 0.86 instead of 1.126 to 1.172. They pointed out that convective dispersion is velocity dependent, and there is no obvious dependence on height. Thicker mixing zone and longer residence time don't necessary mean better mixing (higher convective dispersion coefficient).

Darcy's law to describe flow rate in porous media states that the flow rate is proportional to the permeability. Butler's analytical model (Butler and Mokrys, 1989) to predict flow rate in solvent recovery process indicates that the flow rate is proportional to the square root of permeability. This has been confirmed by laboratory experiments with varying scales (Bulter and Mokrys, 1989; Yazdani and Maini, 2006) and numerical simulations (Cuthiell and Edmunds, 2012). It has also been observed that permeability has an effect on the thickness of the solvent-bitumen mixing zone. Moghadam (Moghadam et al. 2009) found that, the lower the permeability, the thicker the mixing zone. This is easy to understand, lower permeability leads to slower oil drainage rate, which means longer residence time for diluted oil to stay in the reservoir. This gives more time for solvent to diffuse further into bulk bitumen, hence increases the thickness of the mixing zone.

2.2.2 VAPEX Microscopic Mechanism

Researches on the microscopic mechanism of VAPEX process focus on what happens at the solvent-bitumen interface. At the interface, solvent has to find its way to dissolve into

bitumen and then mix with it in order to make it less viscous. Several mechanisms happen in this process including, but not limited to, solvent dissolution in raw bitumen, molecular diffusion and convective dispersion, interfacial tension, and capillary force driven imbibition of oil into gas zone, surface renewal and transient mass flux at the interface. To illustrate them, we can refer to Figure 2.2.2.1 and 2.2.2.2.



Figure 2.2.2.1: VAPEX mechanisms in microscope (Cuthiell and Edmunds, 2012)



Figure 2.2.2.2: VAPEX process in and around capillary mixing zone (Boustani and Maini, 2001)

Solvent contacts and dissolves in the bitumen after being injected into the reservoir. The dissolved solvent will further diffuse into the bulk bitumen and dilutes it. During this process, there are three zones that can be formed at the interface. These three zones are, capillary mixing zone (two phase zone), mobile zone (single phase zone), and viscosity reduced immobile zone (single phase zone).

As shown in above Figures, the solvent-bitumen interface is not smooth. Once solvent dissolves into bitumen, bitumen viscosity can be reduced to the extent that IFT can wick it into the nearby pore throats. This capillary rise generates a capillary mixing zone, sometimes called the two-phase zone, where gas (vapor solvent) and liquid (diluted oil) both exist. Ahead of it is the mobile zone where there is enough amount of solvent dissolved in the bitumen for it to be diluted enough to move under gravity. It is believed that it is in this zone where the convective dispersion takes place and the majority of the diluted oil is drained to the producer (Das and Butler, 1998). Solvent concentration is high in both the capillary mixing zone and in the mobile zone. Ahead of the mobile zone, is the zone that has solvent dissolved in the bitumen, but solvent concentration is not high enough to mobilize bitumen. Further ahead of this zone is the raw bitumen zone that has not yet been contacted by the solvent.

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2.2.2.1 Dissolution in Porous Media

Solvent dissolves both in the capillary mixing zone and the mobile zone. In the capillary mixing zone, where gas and liquid coexist, solvent vapor and the diluted oil reach equilibrium and as a result the solvent concentration in the oil is controlled by its solubility at the local temperature and pressure. The VAPEX extraction rate depends on how much solvent can dissolve into bitumen (the solubility). Normally solubility is a function of ambient temperature and pressure, but in porous media, due to the capillary condensation effect, a decrease in vapor pressure at the curved vapor-liquid interface in fine capillaries will cause the condensation of the solvent vapor for pressures less than the vapor pressure at a flat interface. This can increase the solubility of gaseous solvent at the interface.

VAPEX is considered as a non-thermal process. However when gas dissolves into the liquid phase, there is some amount of heat released (the solution effect). Nevertheless this amount of heat is so small that the temperature increment because of it can be ignored (Dunn and Nenniger, 1989).

2.2.2.2 Molecular Diffusion and Convective Dispersion

Once solvent dissolves at the bitumen interface, it will diffuse further into the bulk bitumen, during this process molecular diffusion and convective dispersion play significant roles.

Perkins and Johnston (1963) pointed out that, if fluids are flowing through a porous medium, mixing due to dispersion may be greater than that due to diffusion alone (Perkins and Johnston, 1963). This can be interpreted to imply that the total dispersion is the summation of dispersion that is caused by molecular diffusion in porous media and the dispersion that is caused by various interactions between fluid and pores in porous media (convection). They further divided this total dispersion into two parts, longitudinal dispersion (dispersion in the flow direction) and transverse dispersion (dispersion that is perpendicular to the flow direction), their correlations for the two dispersion coefficients are given as follows (Perkins and Johnston (1963)):

$$K_L = \left(\frac{D_0}{F\phi}\right) + 1.75Ud_p$$
$$K_T = \left(\frac{D_0}{F\phi}\right) + 0.055Ud_p$$

Where K_L is the longitudinal dispersion coefficient, and K_T is the transverse dispersion coefficient. F is the formation electrical resistivity factor, ϕ is porosity, U is superficial fluid velocity and d_p is the average particle diameter of the medium. D_o is molecular diffusivity without any porous medium. When fluids flow through porous media, the value of the diffusion coefficient is smaller than that without porous media due to the tortuosity in porous media. The first term on the right hand side is called apparent molecular diffusivity; it is the molecular diffusivity that includes the tortuosity effect. The second term on the right hand side is the convective dispersion coefficient. Molecular diffusion describes how fluids mix on a molecular level, and convective dispersion represents how fluid mix on a bulk fluid level. Although K_L is larger than K_T , K_T interests VAPEX researchers more because the flow directions of solvent and bitumen are perpendicular to each other. K_T determines how fast solvent can mix with bitumen.

Perkins and Johnston (1963) also discussed about concentration-dependent molecular diffusion coefficient in miscible fluids, and gave a correlation of calculating molecular diffusivity at a specific concentration.

In the context of solvent-bitumen two components system, there are two types of diffusivity. One is the intrinsic diffusivity of one component into the other, and the other is the mutual diffusivity of the two components. Experimentally it is hard to separate and measure the intrinsic diffusivity alone. Once solvent dissolves into bitumen, theoretically bitumen can counter diffuse into the diluted oil, which is a mutual diffusion process. Some researchers have done measurement of mutual diffusivities between solvent and bitumen (Hossein et al., 2013; Oballa and Butler, 1989; Guerrero-Aconcha et al., 2008).

Oballa and Butler (1989) did experiments using liquid toluene and bitumen in a vertical cell at 20°C and 1 atm to investigate the molecular diffusivities without convection, specifically, D_B (intrinsic diffusivity of bitumen), D_S (intrinsic diffusivity of toluene), and D_{BS} (overall/mutual diffusivity in toluene-bitumen system). They experimentally measured the mutual diffusivity D_{BS} , and then calculated the other two intrinsic diffusivities D_B and D_S . They concluded that these three coefficients are functions of bitumen (or solvent) concentration (volume fraction in their experiments) in the mixture of these two. Their conclusions are, 1) the overall diffusivity reaches a maximum at an intermediate concentration; 2) the maximum binary diffusivity of toluene bitumen system is 4.8×10^{-10} m²/s with 54 vol % of toluene in bitumen; 3) permeability does not affect molecular diffusivity.

It has been widely accepted that diffusivity is inversely proportional to viscosity. For an ideal mixture, many authors assume the product of viscosity and diffusivity is a constant. For a non-ideal mixture, some correlations take the form $D = \alpha u^{-\beta}$. Coefficients of α and β depend on the characteristics of the solute-solvent pair (Hayduk et al., 1973; Hiss and Cussler, 1973, Das and Butler, 1996). Other famous correlations between diffusivity and viscosity are, Wilke-Chang correlation (Wilke and Chang, 1955) and Stocks-Einstein equation. However these correlations have their own applicability restrictions. Stocks-Einstein equation is valid for estimating diffusivity of large spherical molecules in a continuum, Wilke-Chang correlation was proposed for estimating diffusivity of smaller molecules. However its estimation is 2 to 3 orders lower than experimental results (Oballa and Butler, 1989). Das and Butler (1996) designed an experiment in Hele-Shaw cell using Peace River bitumen for bitumen-gaseous solvent system to estimate solvent diffusivity into bitumen as a function of bitumen's viscosity, their correlations take the form: $D_s = 1.306 \times 10^{-9} \mu^{-0.46}$ for propane, and $D_S = 4.13 \times 10^{-10} \mu^{-0.46}$ for butane, where D_S is the diffusivity of solvent in bitumen, μ is viscosity of solvent-bitumen mixture.

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The diffusivity for liquid solvent-bitumen system is different than gaseous solvent-bitumen system. Liquid solvent diffusing into bitumen is a miscible process. In condensing solvent recovery process, bitumen can counter diffuse into liquid solvent at the initial contact. Using Wilke-Chang correlation, at 15°C, the diffusivity of propane into bitumen is estimated to be $13 \times 10^{-11} \text{ cm}^2/\text{s}$ yet at 50°C it is estimated to be $5.2 \times 10^{-9} \text{ cm}^2/\text{s}$. Increasing the temperature increases the diffusion of solvent into the bitumen by two orders of magnitude. Using the Stokes-Einstein equation to predict the diffusivity at infinite dilution of bitumen into propane, the diffusion coefficients at 15°C and 50°C are estimated to be 6.1×10^{-5} and $6.8 \times 10^{-5} \text{ cm}^2/\text{s}$ respectively. At a reservoir temperature of 15° C, the condensing solvent at the interface will create a counter-diffusion process where it is initially six orders of magnitude easier to move the bitumen into the solvent than the reverse. Without condensation the bitumen cannot initially diffuse into the solvent until a sufficient solvent diffuses into the heavy oil to create a live oil zone for reverse diffusion (James, 2009).

Convective dispersion were observed experimentally both in Hele-Shaw cell (Butler et al, 1985; Butler and Mokrys, 1989) and porous media (James, 2009). Its magnitude compared to molecular diffusivity have drawn a lot of attention. Referring to Perkins and Johnston's equation for transverse dispersion coefficient, the apparent molecular diffusivity is a function of solvent-bitumen mixture viscosity (μ), formation resistivity (F) and porosity (ϕ). The convective dispersion coefficient is a function of the superficial velocity (U) and particle diameter (d_p), which are functions of permeability (k), oil phase relative permeability (k_{ro}), solvent-bitumen mixture viscosity (μ), solvent-bitumen mixture density (ρ), porosity (ϕ) and dip angle of the drainage front (θ). Based on this, Cuthiell and Edmunds estimated the relative magnitude of convective dispersion coefficient compared to the apparent molecular diffusivity. Their calculations indicated that for k= 220 Darcy, μ = 39 cP, convective dispersion coefficient is roughly 1.7 times the size of apparent diffusivity. When μ = 8 cP, convective dispersion coefficient is roughly 4 times the size of apparent diffusivity. However when the permeability is reduced to a typical field permeability, k= 4 Darcy, keeping μ = 8 cP, convective dispersion coefficient is roughly only 0.01 times the size of apparent diffusivity. They concluded that under the field conditions, molecular diffusivity completely dominates convective dispersion coefficient (Cuthiell and Edmunds, 2012). A similar conclusion was reached, based on low permeability laboratory experiments, by several other investigators (Ahmadloo et al., 2011; Cuthiell et al., 2006). In condensing solvent process, asphaltene precipitation is en masse, which means the permeability could be damaged even more.

For the diffusion coefficient to be considered basically constant there are three key factors that have to be achieved: 1) the molecular diameter and shape of the molecules involved in the diffusion process should be similar. 2) The inter-molecular forces should be negligible within the diffusing mixture. 3) There should be a non-reacting environment (no reactions between the diffusion pair or environment and non-association between the components) (Guerrero-Aconcha et al., 2008).

The solvent-heavy oil system fulfils the third condition (for the systems studied here) but there are some cases where asphaltene deposition can occur and then none of the above conditions would be reached. In spite of the fact that the solvent-heavy oil pair can not be treated as a system with constant diffusivity, most of the work in the literature accepts it (Guerrero-Aconcha et al., 2008). In the present work, a constant number for total diffusivity is used in the simulator, with the value of convective dispersion coefficient being neglected. Therefore the total dispersion coefficient is equal to the mutual molecular diffusivity.

Some researchers (Guerrero-Aconcha et al., 2008) have investigated the mutual diffusivity between n-alkane and heavy oil. Using the formula they provided in the literature, and assuming the solvent volume fraction in diluted bitumen is 0.5, the mutual diffusivity between liquid hexane, liquid heptane and liquid octane and heavy oil are calculated to be $3.64 \times 10^{-10} \text{ m}^2/\text{s}$, $3.06 \times 10^{-10} \text{ m}^2/\text{s}$ and $1.84 \times 10^{-10} \text{ m}^2/\text{s}$, respectively. Researchers have also found the diffusivity increases for solvent-heavy oil system with the decrease in n-alkane carbon numbers (Guerrero-Aconcha et al., 2008; Wen and Kantzas, 2005; Salama and Kantzas, 2005). Based on this, in this study (chapter 3 and 4), the mutual diffusivity of liquid propane and bitumen at $60 \,^{\circ}$ C is set to be $4 \times 10^{-10} \text{ m}^2/\text{s}$ (The viscosity of bitumen at $60 \,^{\circ}$ C is about 5,000 cP, which can be treated as heavy oil. This number is just an approximation). To determine the mutual diffusivity of liquid propane and bitumen at $26 \,^{\circ}$ C, the above reference can't be used. In our model, the initial viscosity at $26 \,^{\circ}$ C is 1.2×10^{-6} cP, which is beyond the viscosity of heavy oil. We used Butler's correlation (Das and Butler, 1996) to calculate the intrinsic diffusivity of propane into bitumen, and it gives us $1.14 \times 10^{-10} \text{ m}^2/\text{s}$. This value is used in the numerical model as an approximation.

2.2.2.3 Interfacial Tension (IFT), Capillary Pressure and Imbibition

Discussed above, the capillary mixing zone contains both gas and liquid phases. This is because the gaseous solvent and bitumen, are not miscible with each other. As the solvent dissolves into bitumen, the interfacial tension becomes smaller. Still, there will be substantial IFT because solvent has limited solubility in bitumen. How much it can dissolve depends on the solvent solubility, which in turn, depends on the type of solvent, and the operating pressure and temperature.

This IFT combined with the adhesive forces between the diluted oil and rock surface, will overcome the viscous force and gravitational force, wicking the diluted bitumen into the nearby tiny pores. Such capillarity driven migration of the wetting phase to displace the non-wetting phase is called spontaneous imbibition.

Understanding the role of this capillary mixing zone in Vapex is not complete. Uncertainties still exist, to name a few, the thickness of this zone, how much the capillary pressure is, etc. A prevalent belief is that the drainage occurs in the mobile zone. However this has been questioned by some researchers. In their recent paper, Cuthiell and Edmunds (2012) designed simulations to show the roles of capillary pressure in VAPEX. They found that capillary

force is about one-sixth of the gravity force. Given how high it can be, the capillary mixing zone can be a lot thicker than thought before. A thicker mixing zone implies more interfacial contact area, and more oil can be drained in this zone. They further investigated its effects on oil rate and found that the oil rate more than doubled when they made the capillary pressure 5 times higher. In the end they concluded that when a realistic magnitude of capillary pressure is included in the model, much of the drainage takes place in the capillary mixing zone rather than the mobile zone ahead of it.

2.2.2.4. VAPEX Analytical Model

Butler and Mokrys (Butler and Mokrys, 1989) derived the first VAPEX drainage rate analytical model based on their laboratory experiments using liquid solvent in a Hele-Shaw cell. Their first model resulted in the following equations for predicting the volumetric flow rate of oil produced per unit length of a production well:

$$\mathbf{Q} = \sqrt{2} \mathbf{k} g \mathbf{\phi} \Delta \mathbf{S}_{\mathrm{o}} \mathbf{N}_{\mathrm{s}} \mathbf{h}$$

Where, N_s is the dimensionless number given by:

$$N_{s} = \int_{c_{min}}^{c_{max}} \frac{\Delta \rho D_{s} (1 - c_{s})}{\mu c_{s}} dc_{s}$$

Later on, Das (1995) modified Butler and Mokrys' model for its application for porous media. He introduced an apparent diffusion coefficient D_p and related it to the intrinsic diffusivity of solvent in bitumen D_o , cementation factor Ω and system porosity ϕ . The modified equation is as follows:

$$D_p = D_o \phi^{\Omega}$$

and the modified analytical expression for estimation of production rates in a porous medium has become:

$$\mathbf{Q} = \sqrt{2} \mathbf{k} g \phi^{\mathbf{\Omega}} \Delta \mathbf{S}_{\mathrm{o}} \mathbf{h} \mathbf{N}_{\mathrm{s}}$$

To verify the predictions of the above expression, Das and Butler carried out VAPEX experiments in sand pack with Peace River and Athabasca bitumen using butane as solvent. They found that the measured production rates were considerably higher than the predictions. They attributed this observation to the properties associated with porous media, such as extended interfacial area, increased effective diffusivity, increased solubility due to capillary condensation, transient interfacial mass transfer aided by capillary imbibition as well as surface renewal, and enhancement during the rising of solvent chamber (counter-current flow).

Many efforts have been tried out to find the mechanisms behind this discrepancy between analytical predictions and experimental results. Maini and his colleagues (Karmaker and Maini, 2003; Yazdani and Maini, 2006, 2009a; Boustani and Maini, 2001) investigated the role of chamber height and convective dispersion, and found that the dependency of oil rate on chamber height is more than square root indicated in Butler's model. They also suggested that convective dispersion, which was missing in Butler's model, plays an important role in the draining process and therefore should be included in the analytical model. However their opinions have been challenged by some researchers (Alkindi et al., 2011; Cuthiell and Edmunds, 2012) who thought that convective dispersion was not the reason for the high oil rate, and there must be something else missing in Butler's analytical model that has caused the discrepancy.

Capillary effects was examined and thought of as a factor to cause the discrepancy, which was also missing in Butler's model (Ayub and Tuhinuzzaman, 2007; Cuthiell and McCarthy et al, 2006). More work should be done to justify the above postulated factors, and improve the original model by incorporating them in it.

The above analytical model is applicable to the solvent sideways expanding phase. Butler and his colleagues also developed analytical models for the solvent rising phase in VAPEX. One model (Mokrys and Butler, 1993) was developed based on liquid solvent upwards leaching in Hele-Shaw cell, later on they developed another model (Das and Butler, 1996) based on gaseous solvent rising up in packed bed. They found out the solvent-bitumen countercurrent flow phase produced 4 to 10 times more oil than the solvent sideways expanding phase.

2.2.3. Asphaltene Precipitation in Solvent Recovery Process

If solvent concentration in bitumen is high enough, de-asphalting will happen. The extent of de-asphalting depends on the amount of solvent dissolved in bitumen. The solubility of a vaporized solvent is at the maximum near its dew point pressure. Hence, the solvent pressure should be as close as possible to its vapor pressure at the reservoir temperature to cause de-asphalting.

The minimum solvent concentration in bitumen to cause de-asphalting depends on the solvent as well as bitumen. Bray and Bahlke (1938) studied the effectiveness of various solvents in de-asphalting, and found that among all those solvents that were tested, ethane gives the maximum asphaltene precipitation. In addition, the precipitation decreases in the order of propane, butane, pentane, and hexane. Das and Butler (1994) used propane on Cold Lake bitumen and Lloydminster heavy oil and found that the minimum concentration required to trigger de-asphalting were 20% and 30% by weight, respectively.

On one hand, de-asphalting can reduce the produced oil viscosity even more and promote the oil flow rates, as well as enhance the oil quality and therefore reduce many downstream problems. On the other hand, it can also potentially cause formation damage problem, in other words, reduce the permeability and hinder the flow. This has been the concern since the beginning of VAPEX. Mokrys and Butler (1993) did several experiments in a 2D cell with permeability of 81,030 Darcy, using propane to recover Lloydminster type oil (viscosity around 10,000 mPa.s) to examine whether if de-asphalting can slow down the oil rate. Their conclusion was that, when injecting at lower pressures, there was no asphaltene precipitation, oil viscosity was reduced to about 9000 mPa.s, oil rate was moderate; when injecting at saturation pressure, there was some asphaltene precipitation, oil viscosity was reduced to about 9000 mPa.s, oil rate was moderate; when injecting at saturation pressure in its liquid phase), the asphaltene precipitation was "en masse", oil viscosity was reduced to about 1000 mPa.s, oil rate was the lowest. They attributed this low oil rate to the flow blockage caused by asphaltene precipitation.

Other researchers have different opinions on the formation damage by asphaltene precipitation. They found out that asphaltene will precipitate at the trailing edge of the interface where it has the highest concentration of solvent, adhere to sand grains, be deposited at where it was formed without moving with oil, therefore will not cause permeability damage and reduce the flow rates (Das, 1998; Nenniger and Dunn, 2008; Rezaei and Chatzis, 2008; James et al. 2007).

Haghighat did experiments with 3 Darcy permeability (field permeability) sand pack, using propane as solvent, and observed severe damage to permeability and flow blockage. It was concluded that the reason for it was that asphaltene can actually migrate with oil and therefore block the flow (Haghighat and Maini, 2008).

2.2.4. VAPEX Optimization and Applicability

Right after Butler developed his first analytical model to predict oil rate, he compared the order of magnitude of the analogous terms for VAPEX and SAGD and concluded that VAPEX can proceed at a rate only about 1% of that of SAGD. Then he tried to look for potential means of improving solvent recovery process. These were, use vapor instead of liquid solvent to increase the density difference $\Delta \rho$; increase solvent solubility in bitumen and leave asphaltenes in the formation to decrease viscosity μ ; increase temperature to decrease viscosity μ and increase diffusivity D_s ; Use high permeability reservoirs as targets for

VAPEX; use heavy oil (e.g. Lloydminster type oil), not bitumen, to have lower viscosity μ and higher diffusivity $D_{s.}$

Following his suggestions, less viscous oil reservoirs are better candidates than more viscous ones. Karmaker and Maini (2003) did experiments and concluded that oil rates more than doubled when the original oil viscosity was 15 fold lower.

Permeability wise, high permeability reservoirs will perform better than low permeability ones. In addition to higher flow rates (Darcy's law), higher permeability is also favorable due to larger convective dispersion as a result (Butler and Mokrys, 1989; Cuthiell and Edmunds, 2012). This again gives Lloyminster heavy oil reservoirs (higher permeability) more credits than Athabasca oil sands reservoirs. Permeability heterogeneity proved to be not favorable in laboratory experiments (Jiang and Butler, 1996; Oduntan, 2001) by giving lower production rates and higher residue oil saturations.

As discussed earlier, vapor chamber height plays a positive role in oil rates. This makes thicker pay zone reservoirs more desirable. However VAPEX works on thin pay zone reservoirs as well which are not suitable for SAGD due to the heat loss.

From a solvent selection point of view, lighter solvents, such as propane and butane, were proposed (Butler and Mokrys, 1989) because of their lighter densities and higher diffusivities compared to heavier n-alkanes. Lighter densities of solvents give higher density difference between solvent and diluted bitumen, hence higher driving force in gravity drainage processes. Propane was considered to be better than butane, because of its higher diffusivity in bitumen and the likelihood of more upgraded bitumen (propane causes more de-asphalting than butane). However, other researchers found the scenario otherwise. It has been reported that butane has a higher solubility than propane and when injected at near its vapor pressure, can recover more dead oil than propane at similar conditions (Haghighat, 2008).

From an injection strategy point of view, injecting solvents close to its vapor pressure at reservoir temperature gives the highest production rate (Mokrys and Butler, 1993).

Injecting at a particular location in reservoirs might also bring some benefits. Karmaker and Maini (2003) did experiments and concluded that completing the injectors at the gas oil contact (GOC) and injecting solvent at GOC in gas-cap reservoirs were beneficial. The mechanisms behind this could be, a fast communication between the producer and gas cap achieved by allowing gas coning into the production well, more oil produced under this gas cap drive, high free gas production rates help with pushing the accumulated diluted oil held by capillary force near the bottom of the pay zone into the producers.

Das and Butler (1996) proposed to place the injector and producer at the water oil contact (WOC) in reservoirs underlain by water. Illustrated in Figure 2.5, solvent is injected into water and spread along the WOC zone of the reservoir and contact bitumen through a vast

area resembling a planar well. This can considerably increase the solvent bitumen contact area than the traditional two horizontal wells placement described in section 2.2.4.1.



Figure 2.2.4.1: The concept of counter current extraction (Das and Butler, 1996)

The benefits of this well placement is accentuated even more by two factors, first, the injectivity of the water layer is higher than that in the bitumen zone. Secondly, lab experiments show that the counter current extraction rate is 2 to 3 times higher than the corresponding stabilized sideways spreading rate for a similar chamber height (Das and Butler, 1997).

Injecting solvents with hot water and steam in VAPEX have also been studied by many researchers. The results showed better recovery performance by incorporating heat and solvent together (Butler and Mokrys, 1991; Mokrys and Butler, 1993).

Finally, pressure cycling has been studied by some researchers and it has shown some merits. It has been reported that a 20% to 40% increase in oil rates by pressure cycling than constant pressure injection were observed (Mokrys and Butler, 1993; Muhamad et al. 2012; Singhal et al. 1996;). The reasons are thought to be: 1) less formation permeability damage due to de-asphalting. 2) The solution gas drive effect. Some dissolved solvent gas will come out of bitumen when pressure is lowered and pushes some oil out into the gas chamber. 3) As a result of point 2, convective dispersion is enhanced.

Last but not least, being economical is a major consideration for VAPEX process. Solvents are priced as fuels, therefore it's desirable to recover them all at the end of operations. However some solvent retention in the reservoirs can easily happen and is a big concern. The most straightforward way to reduce solvent retention in the reservoirs is to ensure that they are in their gaseous forms instead of liquid forms as much as possible throughout the operations. In this regard, condensing solvent process (N-SolvTM) will be likely to cause more solvent retention than non-condensing solvent process (VAPEX).

2.3 Solvent-Assisted SAGD (SA-SAGD) Process

2.3.1 Overview

Steam-solvent hybrid process involves injecting both steam and solvent in the reservoir to recover bitumen. This combination has been explored by researchers since the 1970s when SAGD was not invented yet. At that time, various types of solvents with different concentrations have been examined to increase the performance of displacement mode thermal processes (Farouq Ali and Abad. 1976: Farouq All, 1976; Hernandez and All, 1972: Junnak and Soni. 1987; Redford.1982; Redford and McKay. 1980: Sarma et al., 1996; Shu and Hartman. 1988). In their work, different solvents were used to improve the efficiency of steamflooding and Cyclic Steam Stimulation (CSS) using vertical wells. Although the potential use of various hydrocarbon solvents have been studied and reported in the literature, the field applications have been rather limited due to unfavorable economics. The consensus from the studies with hydrocarbon additives for displacement model processes is that the recovery efficiency is closely linked with the nature of the solvent co-injected with steam; heavy, medium and light (Ardali et al., 2012). According to Shu and Hartman (1988), lighter solvents yield faster recovery with lesser loss of solvent: while, medium solvents give best oil recovery but at the expense of higher loss of the solvent in the reservoir. Heavier solvents are not recommended (Sarma et al., 1996).

Ever since SAGD was successfully implemented in the field, researches have turned their attention on improving the performance by solvent addition to SAGD. In most cases, this hybrid process takes the same well configuration as traditional SAGD and VAPEX processes; where steam and solvent are co-injected through the horizontal injection well, and diluted bitumen, water and solvent flow down to the horizontal production well by gravity. This steam-solvent hybrid SAGD process is known by different names based on variations in its form, such as, Expanding Solvent SAGD (ES-SAGD) (Nasr et al, 2003), Solvent Additive Process (SAP) (Gupta et al., 2003), Liquid Addition to Steam to Enhance Recovery (LASER) (Leaute, 2002), Steam Alternating Solvent (SAS) (Zhao, 2004), etc. Many laboratory experiments, numerical simulations and analytical studies have been done to understand this process. Some of these processes have gone through field tests and shown encouraging results. The observed benefits of adding solvent to SAGD include: reduced SOR, increased well productivity, reduced capital intensity in startup, improved recovery via reduced SOR, and increased recovery via higher (economic) volumetric sweep (Edmunds, Moini and Peterson, 2010). These processes and their pilot tests are discussed below.

2.3.2 ES-SAGD

The concept of ES-SAGD was first proposed by Nasr et al. (2003), who conducted several experiments at Alberta Research Council (ARC) to examine different types of solvents to improve SAGD process. They tested a wide range of hydrocarbon solvents, from ethane to octane. In all cases oil drainage rate was enhanced by all of the hydrocarbon solvent tested. It

was found that hexane and a diluent (mixer from butane to octane), co-injected with steam, gave the best results (almost 50% higher oil rate than steam injection alone). It was observed that these two additives have similar phase behavior characteristics with steam. In the end they proposed an "ES-SAGD" process, which uses low concentration solvent additives co-injected with steam, and the solvent is chosen in a way that it evaporates and condenses at nearly the same conditions as steam (Nasr et al, 2003). Later on, Ayodele et al., carried out further experiments where the details of propane, hexane, and multi-component diluent co-injection with steam were explained. These experiments were conducted at pressures of 500, 1500, and 2100 kPa. They concluded that low pressure multi-component ES-SAGD at low concentration is fairly competitive with SAGD at a high pressure. The propane-SAGD test gave very low recovery. The energy inputs measured in terms of the heat of vaporization of injected fluids were reduced by 30-40% in all experiments (Ayodele et al 2008, 2009, 2010).

Field test wise, Nexen did an ES-SAGD pilot project in Long Lake for 3 months in 2006. A diluent which contained mainly heavier components (C_7 to C_{12}) was injected at the operating pressure of 1400 kPa. Initial diluent volume was 10% and it was reduced to 5% later. The oil rate was increased by 6% and SOR was reduced by 7% (Orr, 2009). The results were not that impressive compared to other steam-solvent hybrid field tests. It was thought that relatively heavier hydrocarbons did not remain in the vapor phase in the steam chamber; instead, they condensed before reaching to the steam chamber interface and potentially reduced diffusion into bitumen. Another field test by Suncor utilized naphtha for co-injection with steam in a

SAGD project in Firebag area. There was no increase in bitumen production and naphtha was thought to not improve the well's performance. The reason was interpreted to be the same as in Nexen's field test, which was the relatively heavy nature of naphtha (average carbon number was 8). Long term solvent recovery was approximately 70% (Orr, 2009).

In general, in ES-SAGD, an effective solvent should be fully vaporized in the injection well and steam chamber so that it could be transported to the edge of the steam chamber. Preferably, solvent condenses at the same conditions as steam.

2.3.3 SAP

Gupta et al. (2003) has pioneered the study and testing of butane addition to SAGD, termed Solvent Additive Process or SAP. In SAP, a lighter solvent such as butane is injected with steam, solvent vapor accumulates ahead of the steam front, forming a warm "vapor blanket" where it mobilizes and drains oil from regions that may be considerably cooler than the steam zone (Edmunds, 2013).

It has been through several field tests, from Senlac, Saskatchewan to Christina Lake, Alberta, both were done by EnCana (Gupta et al., 2002; Gupta and Gittins, 2006). EnCana has been developing SAP since 1996 and piloted the process first at its Senlac Thermal Project in 2002. Butane was employed as the solvent in the field for co-injection with steam. Published results show that after 3 months, oil production increased by more than 30%. SOR reduced from 2.6 to 1.6 rn³/m³ and energy intensity was reduced by 30% (Gupta and Gittins. 2006). The solvent recovery factor was not reported but is expected to be around 70-90%. Cenovus is currently running a SAP project at Christina Lake. It is planning to use SAP at its Narrows Lake SAGD project that is currently under construction.

Edmunds published a paper analyzing the transport mechanisms for SAP (Edmunds, 2013). He asked a number of questions in the paper: what controls the speed of the front (speed-up vs. SAGD)? What mechanisms cause the SOR reduction? What considerations affect the optimization of choice of solvent and dosage? The answers to these questions were discussed throughout the paper. Edmunds used an example where 5% by mass butane (8.1% by volume, 1.6 mole%) was co-injected with steam. Using this example, he discussed the followings aspects in the paper: Enthalpy profile and phase behavior of water-butane system, phase behavior of water-solvent-bitumen system, temperature profile ahead of moving steam front, controlling factors on solvent-oil drainage front (dosage controlled and solvent transport controlled), and an approximate correlation for the transport limit. Most of his important findings are summarized as follows: 1) Solvent additives in fractional concentrations do not materially alter the condensation enthalpy or temperature of the carrying steam. An identifiable steam front exists, and the injection rate (at constant pressure) is coupled to its rate of advance. 2) Since heat transfer is little altered compared to pure steam injection, the temperature-distance relationship ahead of the steam front can be estimated as for SAGD. 3) Whenever solvent is present in 3-phase conditions, the oil phase properties are fully determined if only temperature and pressure are known. 4) Increasing the ratio of available

solvent to bitumen, progressively decreases the system bubble-point temperature (at which the vapor phase vanishes). The BPT (bubble point temperature) represents the coldest reservoir temperature to which solvent can penetrate as vapor. 5) The effect of increasing the solvent to bitumen ratios on system BPT is asymptotic, as BPT approaches the saturation temperature of the pure solvent. 6) Proper selection of solvent for reservoir oil and pressure, supported by detailed PVT modeling, can ensure that oil phase viscosity remains low at all temperatures between steam and solvent saturation temperatures. 7) This in turn means that oil will substantially drain away, before the steam arrives, thus creating a "vapor blanket" of a certain thickness, which is like a VAPEX vapor chamber, but with a very large temperature gradient. 8) If distance is transformed to temperature via 2), it follows that the thickness of the vapor blanket is controlled by the amount of available solvent. However, beyond a certain point, massive increases in solvent injection will have little incremental effect on drainage, because the temperature asymptote in 5. is transformed to a maximum possible thickness of the vapor blanket. 9) SOR reduction relative to SAGD is limited by the fact that the solvent front is tied to the steam front by the limited range of thermal diffusion. To move a solvent front quickly, the steam zone has to keep up: and if carried to depletion, the greater part of the drainage volume will be heated to steam temperature. 10) The Butler-Mokrys VAPEX theory does not seem to correctly predict two cases of particular interest in SAGD engineering: how lab data scales to reservoir height, and the rate of VAPEX drainage when $T > T_{sat}(P)$ of the solvent. 11) The square-root dependence of B-M is rooted in the basic physical formulation. A measured exponent of 1 or more, indicates that the fundamental assumption of an

accumulating diffusive layer, is not correct. Butler Mokrys model cannot logically be corrected by treating the height exponent as an empirical fitting parameter.

In SAP, the heat loss to overburden is reduced by cooling down of the top of the formation to solvent saturation temperature. Also, the condensation of water before the condensation of solvent can lead to more segregated drainage of water and oil. This may improve the relative permeability effects. With proper design, SAP with butane co-injection can be promising.

2.3.4 Other SA-SAGD Processes (LASER and SAS)

Liquid Addition to Steam for Enhancing Recovery (LASER) was tested in laboratory in a 3D large physical model using Cold Lake bitumen (Leaute, 2002). It involves the injection of a liquid hydrocarbon ($C5_+$) as steam additive in CSS mode of operations. The experiments results show that addition of 6% by volume of diluent can give up to 40% increase in bitumen productivity. Numerical simulations results are consistent with those from physical model experiments (Leaute, 2002).

Encouraged by this result, the first field LASER process was tested for a single cycle in Imperial Oil's CSS operations in Cold Lake. The results of the pilot have been reported by Leaute and Corey (2005; 2007). Following the previous research concept (Leaute, 2002), the pilot added a small 6% volume fraction of $C5_+$ condensate (diluent) into 8 wells during CSS cycle 7, the last cycle. Later cycles in CSS are believed to have low OSR (oil steam ratio), and therefore LASER is preferably implemented in later cycles. The early CSS recovery mechanisms- formation compaction, solution gas drive, no longer contributes substantially in mobilizing the bitumen. Gravity drainage is the dominant driving mechanism. This pilot test lasted 2 years and bitumen production increased by 35% and solvent recovery was estimated to be around 70%. Based on the success of the pilot test results, a large scale application of LASER was implemented by Imperial Oil at Cold Lake that encompassed a total of 240 wells operated from 10 separate pads. Imperial Oil has been operating 189 wells for LASER process on a commercial basis since 2007.

The field measured retention is less than 0.02 m³ solvent per m³ of produced oil. The relevance of a CSS test to SAGD modifications is twofold. Firstly, there is good reason to believe that the basic SAGD mechanism is responsible for the majority of production in the late cycle CSS recovery of bitumen. Thus, the effect of solvent additives is probably similar as well. Secondly, the cyclic nature of LASER can be seen as a strategy for optimizing between production rate and retention; each cycle scavenges its own solvent, and can be extended as required. As in the case of steam, having an injector-producer pair does not rule out cyclic operation, i.e. intermittent steaming (Edmunds et al., 2010).

Zhao (2004) proposed a Steam Alternating Solvent (SAS) process which involves injecting steam and solvent alternatingly. The basic well configurations are the same as in the SAGD process. He first did numerical simulations and concluded that alternating steam with solvent could lead to higher oil rate relative to SAGD, and the energy input is 18% less than SAGD process. He later did laboratory experiments (Zhao et al., 2005) where steam and a mixture of propane and methane were injected alternatingly in a 2D model with Burnt Lake samples. Experiments were conducted at medium pressure (2200 kPa). However, the experimental results were not encouraging in terms of oil recovery compared to SAGD process. The main problem with SAS process is the cooling of the system in solvent injection cycles. In addition, a portion of propane and methane stays in the steam chamber and subsequently reduces the thermal efficiency of the process in steam injection cycles.

2.3.5 Solvent Retention

The principal challenge in hydrocarbon solvent addition is the cost of solvent retained in the reservoir. Edmunds showed a cost comparison among some candidate solvents (Edmunds et al., 2010) and concluded that there should be a strong preference for the use of propane, as the most cost effective diluent. However researches have shown propane tends to actually reduce the oil rate when added to steam, presumably because it is too volatile to condense anywhere near steam temperature and therefore acts more like a non-condensable gas.

Although dynamic retention can be significant when injecting a constant solvent rate, the economic impact is lessened by a shorter pattern life, before final scavenging of the solvent begins. In SAP field tests, oil rate and hence oil steam ratio increased by more than 30%, at constant steam rate. This response was very close to that predicted by simulation, which is very significant from an engineering viewpoint. It suggests that solvent transport is not a

limiting mechanism for the amounts that were injected, and that the physics responsible for this transport are adequately represented in the simulator (Edmunds et al., 2010).

2.3.6 SA-SAGD Optimization

From all the physical experiments and numerical simulation results, it appears that Hexane gives the best oil production performance in SA-SAGD process (Nasr et al., 2003; Li et al., 2011). This is due to its similar phase behavior to steam. Hexane and steam will condense at almost the same conditions, unlike light hydrocarbons (methane, ethane and propane), it will not build up in front of steam chamber and retard the heat transfer. However hexane is more expensive compare to lighter solvents. Butane, is less expensive, and is not as volatile as lighter hydrocarbon solvent. Therefore while preventing heat loss and reducing steam oil ratio, it also condenses easier than lighter hydrocarbon solvents under the same operation conditions. In SAP with co-injected butane, the butane blanket at the front resembles N-SolvTM process.

Hosseininejad et al. (2009; 2010) performed several numerical simulations on addition of butane and hexane to steam and reached the following conclusions: For butane, higher injection pressure leads to lower oil production relative to SAGD, higher solvent retention, and less oil recovered for one unit of butane retained. For hexane, higher injection pressure leads to more retention, less oil recovered for each unit of hexane retained; lower injection pressure gives more incremental oil. They suggested use of relatively low injection pressure, such as 1500 kPa, instead of 1900 kPa. They also tested these two solvents on three major Canadian reservoirs and concluded that hexane is more effective in high viscosity oil (Athabasca reservoir). Butane shows more benefits when injected at high concentration in Athabasca reservoir, but shows no benefits in less viscous reservoirs; low concentration of butane reduces oil production rate, high concentration only makes a small increase in oil rate; steam oil ratio improves only at higher butane concentration.

A recent concept is to vary solvent compositions with time (Gupta and Gittins, 2007b; Gates and Gutek, 2008). The idea is to go from heavier to lighter solvents with time. The heavy solvents are used early, because they are most effective at high temperatures and rates. Lighter solvents are used later on. A lighter solvent helps recover the more expensive heavy solvent, and is then easier to recover and less costly to leave behind (Edmunds et al., 2010).

2.4 N-SolvTM

Nenniger (2008) has described a thermal solvent process, N-SolvTM, which does not use steam. Instead it uses heated pure solvent vapor at pressures that are elevated with respect to the saturation pressure of the solvent at the initial reservoir temperature. This means the vaporized solvent will condense in the reservoir and release the latent heat to warm up the reservoir, as the front expands.

As described by Nenniger (2008), N-SolvTM is a solvent based gravity drainage process that can be used to recover bitumen from oil sands reservoirs. The well configuration is similar to SAGD and VAPEX; two horizontal wells with the injector being several meters above the producer. The elevated temperature in the reservoir because of latent heat can reduce bitumen viscosity to a level that makes it easier for the solvent to diffuse into raw bitumen. The condensed solvent will then diffuse into the viscosity reduced bitumen, dilute the bitumen, and reduce its viscosity even more. The diluted bitumen and solvent mixture (live oil) subsequently flow down to the production well under gravity.

N-SolvTM operates at a moderate temperature and pressure. The flow chart in Figure 2.4.1 shows the process.



Figure 2.4.1: N-SolvTM operation flow chart (www.n-solv.com)

Alberta Research Council Inc. did comprehensive N-SolvTM experiments in the laboratory in 2005. The experiments showed promising results. N-SolvTM pilot project started injection in February 2014. It is located at the Suncor Dover lease in Athabasca, near Fort McMurray, Alberta. It uses butane as the injected solvent, injection temperature is 90 °C, initial reservoir pressure is about 600 kPa, and reported oil rate is about 500 bbl/day (www.n-solv.com).

Two observations that have inspired Nenniger to propose the N-SolvTM process are: 1) the rate limiting step is the high bitumen viscosity because diffusivity is inversely related to bitumen viscosity; 2) the solvent concentration shock phenomenon across the solvent bitumen interface.

Based on these two observations, Nenniger (2008) proposed an empirical correlation to calculate the mass flux rate (oil extraction rate): $m = 43,550 \text{ x} (k\phi/\mu)^{0.51}$, where, k is the permeability in Darcy, ϕ is the porosity, μ is the initial oil viscosity in centipoise. The parameter $(k\phi)^{0.51}$ has units of length and has been previously identified by Yazdani and Maini (2004) as the "grain size" parameter. Figure 2.4.2 shows this correlation.



Figure 2.4.2: Solvent based gravity drainage correlation (Nenniger, 2008)

This correlation includes 11 different crude oils, 4 different solvents, more than 4 orders of magnitude variation in dead oil viscosity and more than 3 orders of magnitude variation in permeability. This correlation includes both VAPEX and N-SolvTM, and both Hele-Shaw experiments and packed bed experiments. It successfully predicts oil production rates over a range spanning almost 5 orders of magnitude. Based on this correlation, Nenniger has shown that the elevated temperatures greatly accelerate drainage rates compared with VAPEX, even to the point of exceeding SAGD rates.

To elaborate on the two observations that have inspired Nenniger, Figure 2.4.3a to 2.4.3d show concentration shock phenomenon from experiments that Nenniger has performed.



1Figure 2.4.3a: "concentration shock" in liquid Hexane and bitumen in Hele-Shaw cell experiment (Nenniger, 2008)



Figure 2.4.3b: "concentration shock" in sandpack bed experiment (Nenniger, 2008)



Figure 2.4.3c: "concentration shock" in sandpack bed experiment (Nenniger, 2008)



Figure 2.4.3d: "concentration shock" in sandpack bed experiment (Nenniger, 2008)

The bitumen interface shown in the above pictures is very clear, sharp and distinct. In the first picture, the surrounding region of hexane diluted bitumen is remarkably uniform and shows very little evidence of any concentration gradient across the 6.5 mm gap (Nenniger, 2008).

The pictures in sandpack bed experiments also show clear extraction interface. These pictures present either fully solvent diluted bitumen or raw bitumen with nothing in between. Solvent concentration gradient can hardly be seen in the photos.

In the last picture, the extracted sand is quite clean with asphaltene residue within one sand grain distance (i.e. pore length) of the raw bitumen. Butler found that solvent concentrations above 50 vol% were required to trigger de-asphalting (Mokrys and Butler, 1993), the last picture suggests that the solvent concentrations exceeded 50 vol% within 100 microns of the raw bitumen. This photomicrograph provides strong support for the hypothesis that solvent extraction in packed beds proceeds via a concentration shock mechanism (Nenniger, 2008).

Both Hele-Shaw and sandpack experiments provide evidence that most of the concentration gradient occurs in an extremely narrow zone, in the magnitude of microns, located right at the bitumen interface. This also means that most of the solvent "driving force" is consumed within a few microns of the raw bitumen interface, a "concentration shock", to get the solvent penetrate into the raw bitumen (Nenniger, 2008).

Nenniger further pointed out that it is extraordinarily difficult to get the solvent to penetrate into the raw bitumen, and the difficulty is due to the high bitumen viscosity. Both Stokes-Einstein and Wilke-Chang empirical correlations indicate that diffusion coefficient is inversely related to viscosity. Since most of the solvent "driving force" is consumed within an extremely narrow zone to overcome this difficulty, it is important to reduce the raw bitumen viscosity first. Therefore the main obstacle for solvent to diffuse into bitumen is the viscosity of the bitumen.

N-SolvTM is a condensing solvent process. Bitumen viscosity is reduced before the start of mass diffusion due to the elevated temperature from condensation latent heat. When liquid solvent contacts bitumen, bitumen can diffuse into solvent at the beginning of the diffusion process due to the miscibility between liquid solvent and bitumen. Therefore this mutual diffusion nature of condensing solvent process makes the diffusivity higher than that of VAPEX where solvent has to dissolve into bitumen to create a live oil layer for bitumen to counter-diffuse. As a consequence of this miscibility, high concentration in bitumen is expected. This in turn reduces the viscosity (and therefore enhances the mutual diffusivity) even more.

Once the obstacle, high bitumen viscosity, is overcome, the mutual diffusivity will then be enhanced and the solvent will mix with bitumen even faster due to the reduced viscosity. Figure 2.4.4 shows, at the same temperature, adding solvent can decrease the viscosity significantly. When volume fraction of propane exceeds 50%, its viscosity can be reduced to below 1 cP level, the viscosity of light oil. Nenniger's laboratory experiments show little evidence of permeability damage due to asphaltene precipitation. This very much diluted oil will flow fast as a result.

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Figure 2.4.4 Effect of temperature and propane volume fraction on the viscosity of Athabasca (UTF) bitumen (Nenniger, 2008)

Diluted oil flowing fast means its retention time is small. Therefore a new bitumen surface can be exposed frequently. Followed by more incoming solvent condensing on it, a new extraction process begins- this process repeats itself, at a very fast rate (fast surface renewal).

A well-known formula to calculate mass flux is Fick's law: $J = -D \frac{\partial \phi}{\partial x}$, the first term on the right hand side is the diffusivity (D), the second term on the right hand side is concentration gradient. ϕ is concentration, x is position, and J is the mass flux. In N-SolvTM, both D and concentration gradient is bigger compared to VAPEX, combined with fast surface renewal, a high mass flux is possible. James (2009) did experiments using butane as solvent to extract bitumen at the room temperature. She observed that the interface propagation speed is four times faster in condensing solvent process than non-condensing solvent process. She concluded some benefits from heating solvents: 1) vapor pressure increases giving higher injection pressure, 2) viscosity reduction due to increased temperature and latent heat of
condensation, 3) concentration difference across the diffusion distance is initially 1, whereas in non-condensing solvent process (VAPEX), the solvent concentration difference is initially the solubility of solvent at the corresponding temperature and pressure, which is much less than 1.

Another important suggestion Nenniger has for N-SolvTM is the injection purity. Figure 2.4.5 shows a binary phase diagram of propane-methane system at 8 °C. On the left side of the dotted black line is the single vapor phase, and under the blue line is the single liquid phase. Suppose operating VAPEX at its dew point at reservoir conditions of 8 °C and 1000 kPa, these conditions require the composition of 65 mol% propane and 35 mol% methane. However when the vapor dissolves in bitumen, the condensed liquid composition is 97 mol % propane and 3 mol% methane. As time goes, methane will accumulate more and more. When the methane builds up greater than 35 mol% it will no longer condense. As a result, it is important to have as little non-condensable gas as possible. Patents are available to mitigate methane contamination (Nenniger and Nenniger, 2001; 2002).



2Figure 2.4.5: Dew point and bubble point for methane-propane mixture versus pressure at 8 °C (Nenniger, 2008)

Solvent can be retained in the reservoir either dynamically as a running inventory, or ultimately after final scavenging. The biggest disadvantage of N-SolvTM is the high dynamic retention of solvent that follows from using condensing solvent to raise and maintain the temperature. Condensing solvent will deliver a great amount of solvent into bitumen, thus the residual oil in the reservoir contains a great deal of condensed solvent. Also, heat losses will ensure that condensing conditions will persist in most of the chamber until the final blowdown, so that liquid solvent saturations in the depleted zone remain elevated, above the oil phase residual (Edmunds, 2010). This solvent condensing in the depleted chamber before reaching to the interface has been observed by some researchers when investigating heated VAPEX experiments (Rezaei et al., 2008). Like VAPEX, N-SolvTM is solvent recover process. Any de-asphalting induced formation damage that affects VAPEX can affect N-SolvTM as well.

Solvent-addition to SAGD has some advantages in terms of solvent retention. It might be thought of as N-SolvTM with added steam. Having a steam front behind the solvent condensation front, maintains temperatures in the depleted zone high enough to prevent accumulation of liquid solvent. Furthermore the solvent oil ratio may be kept low because solvent condensation is not required to heat the reservoir (Edmunds, 2010).

Chapter 3: Investigation of N-SolvTM Mechanisms and Oil Production Rates by Numerical Simulations

3.1 Abstract

Given the search for recovery processes that are more efficient with lower water consumption and greenhouse gas emissions than SAGD, solvent-only processes are a potential alternative. N-SolvTM process uses pure heated solvent to recover heavy oil and bitumen in oil sands reservoirs.

Although Vapor Extraction (VAPEX) has demonstrated success in laboratory experiments, it has not been demonstrated to be technically or commercially successful in the field. N-SolvTM process uses the same well configuration as the VAPEX process – two parallel horizontal wells are placed in the reservoir, the injection well is positioned several meters above the production well. In N-SolvTM process, pure heated solvent is injected into the reservoir to provide additional energy to the reservoir which enhances the reduction of the oil phase viscosity beyond that of solvent alone. The heated solvent is injected at pressures under its vapor pressure (at the injection temperature) to ensure it is vapor. At the edge of the chamber, the solvent condenses releasing latent heat to the oil sand. In this study, the effectiveness of N-SolvTM process in a McMurray oil-sand reservoir is investigated by fine grid (8 cm by 8 cm grid blocks in the cross well pair plane) reservoir simulation. The results from the simulation reveal that dynamics at the edge of the chamber – including phase change, oil viscosity reduction by elevated temperature and solvent diffusion/dispersion, and the interaction between heat and mass transfer are complex. Also, the injection of hot pure propane into oil sand reservoirs yields competitive oil production rate than SAGD both with and without solution gas in the reservoir. However solvent losses and permeability damage are concerns that must be addressed through process design. Lastly, in the presence of solution gas and high connate water saturation in the reservoir, the performance of N-SolvTM suffers significantly.

3.2 Introduction

Steam-assisted gravity drainage (SAGD) has achieved commercial success in western Canadian oil sands reservoirs. However the high energy intensity and GHG emissions associated with SAGD have motivated researchers to seek a low energy input and environmental friendly alternative. Vapor Extraction (VAPEX) was proposed in this context. Both SAGD and VAPEX aim to overcome the same challenge: the high oil viscosity of the oil sands. SAGD uses heat to reduce the viscosity, and VAPEX uses hydrocarbon solvent to dilute bitumen and reduce its viscosity. SAGD has proven to be effective commercially, but VAPEX is extremely slow and therefore hasn't been commercially successful.

It is believed that it's the small mass diffusivity in VAPEX that causes the slow production rate. Nenniger (2008) did experiments and observed that the characteristic length scales of

mass transfer are orders of magnitude smaller than those of heat transfer (the solvent concentration gradient is steep). He also pointed out that the initial high viscosity of bitumen is the reason why solvent diffusivity (mass transfer) is slow, and even a small temperature increase can drop the viscosity to a significant lower level. Based on these observations, he concluded that mass flux can be quite high if moderate amount of heat is provided along with solvent.

Nenniger proposed an oil recovery technology, N-SolvTM, based on the above observations. N-SolvTM uses pure heated solvent, where both sensible heat and latent heat from heated solvent are the sources of heat supply. Based on Nenniger's experiments N-SolvTM can achieve commercial oil rate with a low capital input and is a promising substitute for SAGD. So far there hasn't been numerical work done to justify this method, this study presents some results from numerical simulations of N-SolvTM using commercial simulator CMG STARSTM. In this chapter, N-SolvTM is compared with SAGD in terms of oil extraction mechanisms and oil rates.

3.3 Models Descriptions

Two cases, Case N-SolvTM and Case SAGD (without solution gas), were run to compare N-SolvTM and SAGD. The project life is from January 1st 2006 to August 9th 2007. The first four months is the preheat stage. The reservoir model was based on a highly heterogeneous McMurray oil sand reservoir, the geological model was built in Petrel and then imported into

CMG BUILDERTM. Overburden and underburden heat loss features are incorporated in the model. The simulator used is CMG STARSTM (Computer Modeling Group Limited, 2013), a fully implicit and compositional thermal simulator. It is a 2D model where only the cross well pair plane is studied. The model properties are summarized in Table 3.1. Asphaltene precipitation is not included in the models. The determination of the value of total dispersion coefficient was explained in Chapter 2.

Item	Description	Simulation input
Reservoir properties	Dimension of reservoir (m)	100 x 500 x 36
	Dimension of grid blocks (m)	0.08 x 500 x 0.08
	Number of grid blocks	1250 x 1 x 450
	Average porosity (%)	29
	Average permeability (D)	Vertical: 2.3; Horizontal: 4.6
	Average oil saturation (%)	0.79
Initial conditions	Reference depth (m)	-235
	Pressures (kPa)	800
	Methane dissolved in oil phase (mole %)	4
	Temperature (°C)	12
components	Names	C_3H_8 , Bitumen, H_2O (N-Solv TM), Bitumen H_2O (SAGD)
	K Values, K=(K _{v1} /P) x exp (K _{v4} /(T- K _{v5})) K _{v1} (kPa), K _{v4} (°C), K _{v5} (°C)	$\begin{array}{c} C_{3}H_{8}: K_{v1} = 9x10^{5}, K_{v4} = -1.87x10^{3}, \\ K_{v5} = -2.48x10^{2} \\ H_{2}O:K_{v1} = 1.19x10^{7}, K_{v4} = -3.82x10^{3}, \\ K_{v5} = -2.27x10^{2} \end{array}$
	Bitumen viscosity (cP) at 12 °C	$1.2 \ge 10^{6}$
	Oil phase viscosity vs temperature profile	See Figure 3.3.1
	Oil phase viscosity vs pressure profile	See Figure 3.3.2
Rock/Fluid properties	Rock wettability	Water wet
	Model for evaluating 3-phase kro	Stone's Model 2

Table 3.3.1: Model properties for Case N-SolvTM and Case SAGD (without solution gas)

Item	Description	Simulation input	
Rock/Fluid	Pelative permeability and points	krw=0.2, krow=0.9, krg=0.95,	
properties	Relative permeability end points	krog=0.2	
	Propane total dispersion coefficient in	1.14×10^{-10}	
	oil phase (isotropic), m ² /s	1.14 X 10	
Well	Injection pressure (kPa)	969	
constraints	injection pressure (Kr a)	202	
	Injection temperature (°C)	26 (N-Solv TM), 178.4 (SAGD)	
	Steam/solvent trap (m ³ /d)	10	
	Steam quality (SAGD only)	0.95	



Figure 3.3.1: Oil phase (live oil) viscosity versus temperature profile



Figure 3.3.2: Oil phase (live oil) viscosity versus pressure profile

3.4 Results and Discussion

The oil rates and cumulative oil of the two cases, Case N-SolvTM and Case SAGD, are shown in Figure 3.4.1. At low temperature (26 °C), N-SolvTM gives a competitive oil rate compared to SAGD. SAGD chamber built up slow but increased faster than N-SolvTM later on. As a result, SAGD cumulative oil is exceeding that of N-SolvTM from early August. Both cases oil rates are slowly increasing as the chambers are growing bigger, the difference in the oil rate is not significant. Another two cases, Case N-SolvTM (Methane) and Case SAGD (Methane) (with solution gas), are tested to investigate how N-SolvTM performs in the presence of solution gas methane. Table 3.2 summarizes well constraints of the two cases. All other model properties are the same as those in Case N-SolvTM and Case SAGD. The projects are set to be 2 years.

Two cases with solution gas	Case SAGD (Methane)	Case N-Solv TM (Methane)	
Initial reservoir pressure, kPa	1000		
Maximum injection pressure, kPa	2000		
Injection temperature, °C	212	60	
Steam/solvent trap, m ³ /d	10		
Components	Bitumen, H ₂ O, Methane	Bitumen, H ₂ O, Propane, Methane	
Propane total dispersion coefficient, m ² /s		4 x 10 ⁻¹⁰	
Solution gas content (methane oil mole		4	
Components Propane total dispersion coefficient, m ² /s Solution gas content (methane oil mole fraction), %	Bitumen, H ₂ O, Methane	Meth 4 x 1	

Table 3.4.1: Summary of two cases (with solution gas)

Figure 3.4.2 shows oil rate (left) and cumulative oil of Case N-SolvTM (methane) and Case SAGD (methane). It is obvious that in the presence of solution gas methane, N-SolvTM gives higher oil production than SAGD at a moderate injection temperature (60 °C). Figure 3.4.3 shows methane gas production rate (left) and cumulative methane production of Case N-SolvTM (methane) and Case SAGD (methane). It can be seen that N-SolvTM produces more methane from the reservoir than SAGD. In N-SolvTM, methane condensation temperature becomes higher in the presence of propane. In SAGD, methane and water are not miscible;

therefore their condensation temperatures are not affected by each other, which means methane condensation temperature is lower than that of N-SolvTM. In other words, it's easier for methane to condense in N-SolvTM process. Given that methane impedes oil production in both N-SolvTM and SAGD, it's good to know that N-SolvTM has a better ability to remove methane than SAGD. This observation is consistent with Nenniger's experiments.



Figure 3.4.1: Oil rate (left) and cumulative oil comparisons between Case N-SolvTM and Case SAGD (both cases without solution gas, N-SolvTM injection temperature 26 °C)



Figure 3.4.2: Oil rate (left) and cumulative oil comparisons between Case N-SolvTM (methane) and Case SAGD (methane) (both cases with solution gas, N-SolvTM injection temperature 60 °C)



Figure 3.4.3: Solution gas production rate (left) and cumulative solution gas produced for Case N-SolvTM (methane) and Case SAGD (methane) (both cases with solution gas, N-SolvTM injection temperature 60 $^{\circ}$ C)

The N-SolvTM mechanisms inferred from experiments were discussed in chapter 2, the following discussion explores the mechanisms from a numerical simulation perspective. Some insights can be gained by examining the property profiles. Figure 3.4.4 shows the depleted chamber development for Case N-SolvTM and Case SAGD. The SAGD chamber has the upside down triangle shape, and N-SolvTM takes an umbrella shape. However in the experiments Nenniger has performed, N-SolvTM has the traditional S-shape (www.n-solv.com). The reasons for this are the different oil extraction mechanisms between N-SolvTM and SAGD under the background of the current geological heterogeneity, which are explained in details as follows.



Figure 3.4.4: Chamber developments for SAGD (top three) and N-SolvTM (both cases without solution gas)



Figure 3.4.5: Initial water saturation

Figure 3.4.5 displays the initial water saturation. There is a high water saturation layer (green strip) between the injector and producer. The high water saturation layer is not symmetrical around the wellbore. On the left side of the wellbore, high water saturation layer ended not far away from the wellbore, whereas on the right side of the wellbore, the high water saturation layer extends all the way to the reservoir boundary.

Water has a relatively higher mobility than bitumen, therefore it will flow relatively faster once the production begins. This means the high water saturation zones will be depleted sooner than the low water saturation zones. As a result, the chamber will develop faster in the high water saturation zones. This explains why the N-SolvTM chamber has two "wings" around the wellbore. As chamber develops, the horizontal development of the left wing slows

down when it hits the end of high water saturation layer. The right wing kept growing horizontally at the same speed, and after two years, the right wing is longer than the left wing.

The effect of connate water on solvent process has been investigated by researchers before. Etminan et al. (2007) has shown that connate water can accelerate the spreading of vapor chamber in the lateral direction and tend to increase the thickness of the mixing zone. This increase in the mixing zone thickness appears to result from capillarity driven fingering phenomenon. The mixing zone had a distinct uneven appearance that was similar to patterns generated by frontal instabilities in miscible displacements. Connate water can also affect the drainage rate of diluted oil. The effect of connate water on the drainage rate was an increase in the initial rate, but a reduction in the rate, subsequently. The presence of mobile water speeds up the communication between the two wells and leads to even faster spreading of the vapor chamber (Etminan et al., 2007).

The above findings are consistent with our simulation results. Figure 3.4.6 displays the enlarged chamber of gas phase mole fraction of propane of Case N-SolvTM. Several phenomena can be observed that are consistent with Etminan's conclusions: 1) the mobile water does make chamber expands faster in the lateral direction, and that's how the "wings" were formed, 2) the mixing zone on the tip of the right wing is thicker than the mixing zones of the other parts of the reservoir, 3) the presence of higher water saturation zone did make communication between the injector and the producer achieve faster (communication achieved first on the right side of the wellbore where it has higher connate water saturation).



Figure 3.4.6: Development of communication between two wells of N-SolvTM (gas phase mole fraction of propane shown, without solution gas)

It should also be noted that since connate water saturation can change the shape of chamber from S-shape to the umbrella shape, it could make the permeability damage more severe. To be specific, if the chamber ends up with being an umbrella (or some other irregular shape), diluted oil will flow from top to bottom, instead of flowing along the chamber edge. This means oil will not be able to bypass the damaged zones as it will be in S-shape chambers. The subsequent effects are complicated in terms of the mass flux rate (related to the thicker mixing zone) and oil flow rate (taking permeability damage into consideration).

As discussed in Chapter 2, two important mechanisms of N-SolvTM are: first, the extremely short diffusion distance (steep concentration gradient), and second, the higher diffusion coefficient (which enables easier initial penetration) due to the reduced viscosity by elevated temperature. In the simulations, both of these mechanisms can be visualized in absence of solution gas, as shown below. Case N-SolvTM and Case SAGD are used as examples to visualize the mechanisms, with Case N-SolvTM, the dispersion coefficient was changed to 1 x 10⁻⁹ m²/s to save the run time. It doesn't affect the distributions of the properties, only makes the depleted chamber sizes bigger.



Figure 3.4.7: N-SolvTM enlarged chamber edges (oil saturation distributions shown, without solution gas)

Figure 3.4.7 shows N-SolvTM enlarged chamber edges (oil saturation distributions shown) for depleted grid block (left) and not yet depleted grid block. N-SolvTM chamber edges from above pictures are showing the same feature as Nenniger's experiments: clear, sharp and distinct. In the depleted grid block it is the fully diluted bitumen, and the grid block next to it has the initial oil saturation, with no oil saturation transition in between, indicating a concentration shock phenomenon.

Figure 3.4.8 shows enlarged temperature distributions (top two) and viscosity profiles for SAGD, chamber edge to raw bitumen is shown here. From vertical grid block number 140 to 55, with each grid block being 8 cm, the distance shown in Figure 3.4.8 is (140-55) x 8 cm=6.8 m. Specifically, temperature reduces from depleted chamber temperature (177.3 °C) to 25 °C, and viscosity increases from residue oil viscosity (15.5 cP) to around 160,000 cP in 6.8 m. Figure 3.4.9 shows enlarged temperature distributions (top two) and viscosity profiles for N-Solv[™]. From vertical grid block number 156 to 153, temperature reduces from depleted chamber temperature (26.34 °C) to 24.59 °C, and viscosity increases from residue oil viscosity increases from residue oil viscosity (0.16 cP) to around 160,000 cP in 0.24 m (grid block number 153 to 156). Temperature has small change (1.75 °C) in N-Solv[™] compared to SAGD, where temperature has to rise about 152 °C to reduce the viscosity to the same extent as solvent does. With small temperature change (little heat) compared to SAGD, N-Solv[™] reduces oil viscosity to the same extent as SAGD in a distance that is 30 times shorter than SAGD, indicating solvent is effective at reducing the viscosity of oil.



Figure 3.4.8: SAGD enlarged temperature distributions (top two) and viscosity distributions (without solution gas)



Figure 3.4.9: N-SolvTM enlarged temperature distributions (top two) and viscosity distributions (without solution gas)

Figure 3.4.10 shows enlarged distributions of oil phase mole fraction of propane. Oil phase mole fraction of solvent changes from 0.4% to 98% in 0.24 m, implying that a large amount of solvent has dissolved in bitumen in this short distance. A simple calculation can convert solvent oil mole fraction of 45% and 6% to volume fraction of 11% and 1%. Referring to Figure 2.4.4, these numbers correspond to live oil viscosities of 250 cP and 65,000 cP approximately, which are consistent with the simulation results.

From Figure 3.4.7 to 3.4.10, we can see that simulation results provided evidence that the diffusion distance, also called the live oil layer in some literature, is very short (in the this case, it's about 2 to 3 grid blocks long, 0.24 m). Solvent oil mole fraction reduces from 98% to 0.4% over a distance of 0.24 m, meaning large amount of solvent mixes with bitumen in short distance. This is another evidence of the steep concentration gradient, which is consistent with Nenniger's experiment.



Figure 3.4.10: N-SolvTM enlarged distributions of oil phase mole fraction of propane (without solution gas)

Nenniger has provided experimental evidence that the diffusion distance is related to the length of grain size. Some literature (James, 2009; James and Chatzis, 2004; Moghadam et al., 2009) conclude that live oil layer is about a few pores deep, and the mobile live oil layer, i.e., the part of live oil layer that is diluted enough to be mobilized by gravity, is on average one pore deep. If solvent molecules proceed in a pore to pore manner, then simulation results are only valid if each grid block acts like a sand grain in the simulator. Looking at our simulations, solvent oil mole fraction reduces from 98% to 46% in one grid block (the mobile live oil layer), Although these distances are much larger than pore sizes, the presence of very sharp gradients in the simulation results proves that our simulation is a reasonable approximation.

Figure 3.4.11 shows distributions of temperature (top two) and oil viscosity for SAGD (left) and N-SolvTM in which depleted chambers are shown. Solvent addition to a 26 °C elevated reservoir temperature can reduce viscosity to a value that's 75 times lower than steam alone (178.4 °C). Much lower viscosity oil flows faster than higher viscosity oil in the reservoir. Oil relative permeability is also bigger in N-SolvTM than in SAGD due to single phase flow (immobile connate water can be ignored). These are the factors that make possible the relatively large oil drainage rates of N-SolvTM. Note that the Case N-SolvTM simulation was re-run with higher dispersion coefficient (1 x 10⁻⁹ m²/s) to make this plot. It has affected the chamber size but does not change the temperature and oil viscosity distribution.



Figure 3.4.11: Distributions of temperature (top two) and oil viscosity for N-SolvTM (left two) and SAGD (without solution gas)

Figure 3.4.12 shows enlarged chamber edges (oil saturation distributions shown) for SAGD (left) and N-SolvTM. Compared to N-SolvTM, SAGD chamber edge is more blurry, a draining layer (yellow layer) can be seen. In N-SolvTM, oil flow rate is higher than in SAGD due to its lower diluted oil viscosity and the relative permeability effect. At the same time, the speed of mass transfer is much lower than the speed of thermal conductivity. Therefore, compared to N-SolvTM, the speed difference between heat transfer and resulting oil flow rate in SAGD is higher. This means speed of oil drainage cannot keep up with the speed of thermal conduction. Therefore the surface renewal in SAGD is slower, and the viscosity reduced oil has a longer retention time in the reservoir. That's why in SAGD a much thicker draining layer exists. This again proves that N-SolvTM has a faster surface renewal compared to SAGD, which is another point that agrees with Nenniger's observation.

Figure 3.4.13 shows distributions of oil saturation for SAGD (left) and N-SolvTM; depleted chambers are shown. Another benefit of miscible recovery process is that the residual oil saturation is smaller than non-miscible process. N-SolvTM residual oil saturation is about half of SAGD's. It means that for the same swept volume, N-SolvTM can recover more oil than SAGD. The recovery process is more efficient in N-SolvTM. Note that the Case N-SolvTM simulation was re-run with higher dispersion coefficient (1 x 10⁻⁹ m²/s) to make this plot. It has affected the chamber size but does not change the oil saturation distribution.



Figure 3.4.12: Enlarged chamber edges for SAGD (left) and N-SolvTM (without solution gas)



Figure 3.4.13: Distributions of oil saturation for N-SolvTM (left) and SAGD (depleted chambers shown, without solution gas)

Lastly, a simple calculation is shown to prove that vapor propane in our simulation has actually condensed. Figure 3.4.14 shows distributions of oil phase mole fraction of propane in depleted chamber. Residual oil mole fraction of propane is on average 0.97. This value in the produced oil is 0.9646. Now convert the mole fraction to volume fraction. Assuming bitumen molecular weight is 500 g/mol, the API gravity of bitumen in our model is 8.6, Nenniger's experiments have shown that the API gravity can increase about 5 degrees, this gives us about API gravity of 13.6 for the mobilized dead oil, which corresponds to density of 975.2 kg/m³. The density of liquid propane is 493 kg/m³. Hence the volume ratio of propane to dead oil is, (0.9646 x 44/493)/ (0.0354 x 500/975.2)=4.74:1. In non-condensing VAPEX, propane volume fraction has been measured to be maximum 40% (Boustani and Maini, 2001), only condensing solvent process can deliver such large amount of solvent in bitumen. Similar value of oil mole fraction of propane was observed in cases with solution gas, indicating propane has also condensed in those cases.



Figure 3.4.14: Distribution of oil phase mole fraction of propane in depleted chamber of N-SolvTM (without solution gas)

3.5 Conclusions

N-Solv[™] extraction mechanisms have been confirmed by numerical simulations. The simulation results— the oil volume fraction of propane, its corresponding diluted oil viscosity, and the thickness of live and mobile live oil layer are consistent with results reported in the literature. Without considering asphaltene precipitation induced permeability damage, the results support the concepts of steep concentration gradient (high driving force), low oil viscosity (high oil flow rate), fast surface renewal, and better sweep efficiency can make a faster oil production rate than SAGD possible. If high water saturation zones are present in reservoirs, the chamber shape will deviate from the traditional S-shape. The subsequent effects are complicated in terms of oil extraction rate (related to the thicker mixing zone) and oil flow rate (taking permeability damage into consideration).

Chapter 4: A Sensitivity Analysis of the N-SolvTM Process

4.1 Abstract

It has been shown experimentally and numerically that N-SolvTM has potential for comparable oil extraction rates compared to SAGD. However it also has limitations, such as, solvent retention and potential permeability damage from precipitation of asphaltenes. Given solvent retention, the N-SolvTM process, for field application, must be evaluated further from economic perspectives. The optimization goal is to enhance the production at a lower cost. Here, a sensitivity analysis is done to suggest improvements to the recovery process operation.

The effects of temperature and pressure on oil production volume and the Solvent-to-Oil Ratio, denoted by SOR, are explored. SOR is defined by: $SOR = \frac{\text{total liquid solvent injected - total liquid solvent recovered}}{\text{total bitumen recovered}}.$

The results show that injecting at superheated conditions can improve SOR without affecting cumulative oil production. Injecting at high pressure leads to high SOR with high oil production.

4.2 Introduction

One big disadvantage of condensing solvent processes is that the solvent will potentially be retained in the reservoir more than non-condensing process (VAPEX). This has been shown from the distribution of oil phase mole fraction of propane (higher than 95%) in the depleted chamber in the simulation cases shown in Chapter 3. Solvents are priced as fuels and thus solvent retention in the reservoir can make the recovery process uneconomical.

Solvent can remain in the reservoir dynamically as a running inventory as the process evolves or ultimately after final scavenging as would be the case in a blow down operation after the oil recovery process is stopped. In this study, solvent retention as a running inventory is examined.

The density of vapor propane at 15°C 1.91 kg/m³ and the density of liquid propane at 15°C 493 kg/m³, are used in the definition of SOR to convert the volumes of vapor propane to the equivalent volume of liquid propane. The SOR measures how much propane is retained in the reservoir per volume of dead oil produced. The effects of injecting superheated propane, high injectivity (at superheated conditions) and high pressure (at saturated conditions) on solvent retention are examined.

It is widely accepted that non-condensable gas (NCG), e.g. methane, co-injection can potentially have a detrimental effect on solvent process (Friedrich, 2005; Nenniger 2009) and SAGD process (Yuan et al., 2010; Sharma et al., 2012). In some SAGD cases (low initial oil viscosity, low concentration) NCG accumulated at the top of the formation might reduce the heat losses to the overburden (Hosseininejad and Maini, 2009; 2010) and have a positive effect. However, co-injected NCG would not be beneficial to a pure solvent process since heat losses are not a key control on the process efficiency. NCG practically does not condense and tends to accumulate at the top of the chamber, due to its high volatility. At the depletion chamber edge, it forms a blanket between the injected condensable gas and the raw bitumen which impedes heat and mass transfer. Referring to the phase behavior diagram (Figure 2.4.5), adding a more volatile component (NCG) to a single component solvent (propane) system will make the less volatile component (propane) more volatile and consequently more difficult to condense. In N-SolvTM, 1 mole percent concentration of methane reduces the propane condensation temperature by about 5°C (Nenniger, 2008). Thus, it only takes a few mol% methane contamination in the vapor phase to significantly impair heat transfer to the interface.

4.3 Injecting at Superheated Conditions

To understand how temperature and pressure affect the N-SolvTM process in terms of oil production rate and the SOR, eight superheated N-SolvTM cases were tested. The parameters of these models are kept the same as those in Chapter 3, except for the well constraints and components (methane is added as a solution gas). A summary of well constraints is listed in Table 4.3.1. At 2000 kPa, Propane's saturation temperature is 58.5 °C, therefore these eight
cases are superheated, with temperatures being 1.5, 11.5, 21.5 and 31.5 °C higher than the saturation temperature. The propane total dispersion coefficient was set to be 1 x 10^{-9} m²/s to shorten the run time without affecting the comparative results trend (only the absolute oil rates will increase).

N-Solv TM									
eight	Case 1	Case 3	Case 5	Case 7	Case 2	Case 4	Case 6	Case 8	
superheated									
cases									
Initial									
reservoir	1000				2000				
pressure,									
kPa									
Maximum									
injection	2000								
pressure,	2000								
kPa		r	r	r	[r	ſ	r	
Injection									
temperature,	60	70	80	90	60	70	80	90	
°C									
Maximum									
bottom hole									
gas	10								
production									
rate, m ³ /d									
Components	Bitumen, Propane, Water, Methane								
Propane									
total	1 x 10 ⁻⁹								
dispersion									
coefficient,									
m²/s									
Solution gas									
content (oil									
mole	4								
fraction of									
methane), %									

Table 4.3.1: Summary of N-SolvTM eight superheated cases

The results of cumulative oil, oil rate and solvent oil ratio of the eight superheated cases are shown in Figures 4.3.1, 4.3.2 and 4.3.3, respectively. We have drawn the following conclusions from the results:

Cumulative oil:

Cumulative oil decreases slightly as injection temperature increases, but increases significantly as injectivity increases.



Figure 4.3.1: Cumulative oil volume produced for Cases 1 to 8



Figure 4.3.2: Oil rates profiles for Cases 1 to 8



Figure 4.3.3: SOR profiles for Cases 1 to 8

SOR:

As a running inventory, the volume of propane retained per volume of dead oil produced ranges from 0.5 to $0.3 \text{ m}^3/\text{m}^3$. The SOR decreases as the process evolves, and as the injection temperature increases. Also, the SOR rises as the injectivity increases.

To explain these results, the high injectivity cases (Cases 1, 3, 5 and 7) are used as examples to explain the results. The profiles of oil, gas, and water saturation, temperature, pressure, gas mole fraction of propane, oil mole fraction of propane and bitumen are explored in detail. The spatial distributions of reservoir properties are shown after three years of injection (date equal to 07/01/2009). Only depleted chambers are shown, and color scales are set to be the same so as to compare the four cases.

Figure 4.3.4 shows the oil saturation distributions, top two pictures are Cases 1 (left) and 3, bottom two pictures are Case 5 (left) and 7 (all the property distributions shown after follow this order). Between the cases, the sizes of the depleted chambers and oil saturation distributions are about the same, indicating that the oil productions are about the same. The oil saturation distributions near the wellbores are quite different, implying that there is a small oil production difference among the four cases.

Figure 4.3.5 shows the temperature distributions. The results show that it appears that the extra heat carried by the superheated propane cannot reach the chamber edge and that it is largely consumed directly around the wellbore. To take a closer look at the chamber

temperature, the color scales were adjusted further from 57.5 °C to 60 °C and the results are displayed in Figure 4.3.6. The saturation temperature of propane at 2000 kPa is equal to 58.45 °C, Figure 4.3.6 shows that only the heat of saturated propane can be delivered to the edge and that the superheated propane only heats up the near wellbore region. The reason for this could be due to the limited injection rate set by the injection pressure (2000 kPa for all cases). Injection rate is not high enough to heat up the entire reservoir for higher temperature cases. Also the results in this figure reveal that the chamber interface temperatures (where mass and heat transfer take place) are almost identical among the four cases.



Figure 4.3.4: Oil saturation distributions for Cases 1, 3, 5, and 7



Figure 4.3.5: Temperature distributions for Cases 1, 3, 5, and 7



Figure 4.3.6: Temperature distributions at a closer look for Cases 1, 3, 5, and 7

Figure 4.3.7 shows the pressure distributions. Similar to the results for temperature, the distribution of pressure are also very similar among the four cases. It can be concluded that temperature and pressure should not be responsible for the difference in SOR and cumulative oil production. The near wellbore superheated zone is the region where it shows different property distributions among the four cases, hence the reasons for the difference in SOR and cumulative oil production can be found here.

Figure 4.3.8 displays distributions of the gas saturation. The superheated region has higher gas saturation than the rest of the reservoir. The higher the injection temperature, the larger is the high gas saturation zone. This is because when condensed propane flows down to the production well, it is re-vaporized due to the higher temperature region around the wellbores. The higher the temperature is, the greater the amount of condensed propane that is and the larger is the high gas saturation zone.



Figure 4.3.7: Pressure distributions for Cases 1, 3, 5, and 7



Figure 4.3.8: Gas saturation distributions for Cases 1, 3, 5, and 7

Figure 4.3.9 shows the distributions of oil phase mole fraction of propane, it decreases as injection temperature increases. As shown in Figure 4.3.8, the higher the temperature, the more condensed propane will be re-vaporized into vapor. Therefore less of it will be left in the live oil, in other words, the smaller the oil phase mole fraction of propane will be.

Figure 4.3.10 presents the distributions of the bitumen mole fraction in the oil phase. The images show that the mole fraction rises as the injection temperature increases. As live oil drains down to the wellbore, the higher the injection temperature, the greater the amount of propane that is re-vaporized and, the larger the bitumen that is left in the reservoir as irreducible oil. Hence, less bitumen will be produced. This explains why cumulative oil decreases (slightly) as injection temperature increases.



Figure 4.3.9: Distributions of propane mole fraction in oil phase for Cases 1, 3, 5, and 7



Figure 4.3.10: Distributions of bitumen mole fraction in oil phase for Cases 1, 3, 5, and 7

Figure 4.3.11 shows the distributions of the global mole fraction of water. It has the same value at the superheated region for all cases. As shown, in the superheated area, high gas saturation zone is bigger as the superheated area temperature is higher (injection temperature is higher), and the water saturations are the same with different injection temperature, these explains why oil saturation is lower as the superheated area temperature becomes higher.

From the above figures, we have concluded that, at the superheated region, both oil saturation and oil phase mole fraction of propane are lower when the injection temperature is higher. This means propane retention (the amount of propane that equals to oil saturation times oil phase mole fraction of propane) in the reservoir is lower, which in turn explains why SOR decreases as injection temperature increases.

Figure 4.3.12 displays the distributions of the oil saturation for Cases 1 (right) and 2. The initial reservoir pressures are 1000 kPa and 2000 kPa for Case 1 and 2, respectively, and the injection pressure is the same for both cases (2000 kPa). Depleted chamber is noticeably bigger when the injectivity is higher. It can be concluded that high injectivity gives a lot more oil production; its effect on oil production is more pronounced than temperature. However, the benefit of having a higher injectivity is reduced by the fact that it also leads to higher SOR.

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Figure 4.3.11: Distributions of global mole fraction of water for Cases 1, 3, 5, and 7



Figure 4.3.12: Oil saturation distributions for Case 1 (right) and 2

To summarize, the effects of injecting superheated propane and injectivity on SOR and oil production were studied. When injecting at superheated conditions, SOR is improved. The more superheated, the less propane is retained, the more improved is SOR. Injecting at more superheated conditions can lead to a slightly smaller cumulative oil production than less superheated conditions at the same injection pressure. However the difference is very small. It can be concluded that different temperatures at superheated conditions do not significantly affect cumulative oil production. When injection pressures are the same, the one that has higher injectivity can give significantly more oil production. However it also leads to higher SOR.

4.4 Solution Gas Effects

Figure 4.4.1 and 4.4.2 show mole fractions of methane distributions in oil and gas phases, respectively. Oil phase mole fraction of methane ranges from 0 to 0.07 from the depleted chamber to the edges of the chamber. It is a small fraction compare to propane and bitumen, showing limited solubility of methane in the oil. We can tell from the color scale, gaseous methane only accumulates at the edge of the chamber, as in the oil mole fraction profile. In the gas phase near the chamber edge, methane dominates the composition, with more than 99% gas phase mole fraction. From both pictures methane is accumulated at the top edge of the chamber and its fingers can be seen due to the high injectivity. Such high concentration of methane accumulating at the edge of chamber blocks the contact of propane with bitumen, whereby it not only hinders the mass transfer, but also impedes the heat transfer.



Figure 4.4.1: Distributions of oil phase mole fraction of solution gas methane for Cases 1, 3, 5, and 7



Figure 4.4.2: Distributions of gas phase mole fractions of solution gas methane for Cases 1, 3, 5, and 7

To see how solution gas methane impacts on oil rates quantitatively, two cases, Case M1 and Case M2 were tested. These two cases are the same models as the previous eight superheated cases, with the only changes being the initial conditions, well constraints, and components, which are listed in Table 4.4.1. Again propane total dispersion coefficient was set to be 1 x 10^{-9} m²/s to save run time, it doesn't affect the trend. Project life is two years.

N-Solv TM two saturated cases	Case M1 (no solution gas)	Case M2 (with solution gas)			
Initial reservoir pressure,	800				
kPa					
Maximum injection	060				
pressure, kPa	909				
Injection temperature, °C	26				
Production well					
constraints, maximum	10				
BHG, m ³ /d					
Components	Bitumen, H ₂ O, Propane	Bitumen, H ₂ O, Propane, Methane			
Propane total dispersion	1 x 10 ⁻⁹				
coefficient, m ² /s					
Solution gas content					
(methane mole fraction in	4				
oil phase), %					

Table 4.4.1: Summary of N-SolvTM two saturated cases (with and without solution gas)

Figure 4.4.3 displays the oil rate (left) and cumulative oil comparison between Cases M1 and M2. Case M1 that does not suffer from solution gas methane contamination gives more oil production.

Figure 4.4.4 shows oil saturation distributions of Cases M1 (left) and M2. As of 01/01/2008,

the depleted chamber of the reservoir that does not contain solution gas is much bigger than the depleted chamber of the reservoir that has solution gas, indicating more oil production. Methane as a solution gas can affect the chamber shape, the chamber, in the presence of methane, is shorter. Due to methane accumulation at the top of the reservoir, it can slow down mass and heat transfer and therefore make chamber development in the upward direction slower.



Figure 4.4.3: Oil rate (left) and cumulative oil comparison between Cases M1 and M2



Figure 4.4.4: Oil saturation distributions of Cases M1 (left) and M2

Figure 4.4.5 shows the oil rate profiles for the eight superheated cases (Case 1 to 8) and Case M1. With solution gas methane, there is a hill after the initial spike in each oil rate curve for Case 1 to 8. In these cases oil rate follows the trend of increasing first and then decreasing. As production goes, more and more methane will accumulate at the top edge of the reservoir. In the early stage when not much of methane has been accumulated to impede the mass and heat transfer, oil rate will keep increasing due to the bigger contact area as chamber grows. However when it hits the point where it has accumulated enough to start blocking the contact between propane and bitumen, the oil rate will start decreasing. This trend does not exist in the non-solution gas case (Case M1) where the oil rate keeps increasing, following the prediction of Butler's analytical model.



Figure 4.4.5: Oil rate profiles for the Cases 1 to 8 (with solution gas) and Case M1 (without solution gas)



Figure 4.4.6: Distributions of gas phase mole fraction of solution gas methane at different times for Case 1

Figure 4.4.6 presents gas phase mole fraction of methane distributions at the top edge of the chamber at different times for the eight superheated cases, Case 1 is shown here as an example. Methane accumulates more and more over time at the top edge of the reservoir. It started to accumulate after August, which is the date when oil rate starts to decrease. This is consistent with the oil rate curves in Figure 4.4.5.

To summarize, solution gas has a detrimental effect on the N-SolvTM process. It can considerably decrease the oil rate. It also affects the chamber shape, making it shorter and wider. Its amount should be minimized in the reservoir by proper process design.

4.5 Injecting at Saturated Conditions

In this section, five saturated N-SolvTM cases are studied (with solution gas). Propane is injected at the saturation temperatures corresponding to the injection pressures. A summary of these five cases is listed in Table 4.5.1. Case S1 is also Case M2. Again model properties are kept the same except for the well constraints listed in table 4.5.1. The propane total dispersion coefficient is again set to be $1 \times 10^{-9} \text{ m}^2/\text{s}$ to save run time. Project length is two years.

N-Solv TM five saturated cases	Case S1	Case S2	Case S3	Case S4	Case S5	
Initial reservoir pressure, kPa	1000			2000		
Maximum injection pressure, kPa	969	1300	2000	2000	2980	
Injection temperature, °C	26	39	58.5	58.5	80	
Propane total dispersion coefficient, m ² /s	1 x 10 ⁻⁹					
Maximum bottom hole gas, m^3/d	10					
Solution gas content (methane mole fraction in oil phase), %	4					

Table 4.5.1: Summary of N-SolvTM five saturated cases

Figure 4.5.1 shows oil rate (left) and cumulative oil, and Figure 4.5.2 shows SOR for the five saturated cases, Cases S1, S2, S3, S4, and S5. The results show that high temperature (high pressure) will give higher oil production, SOR is higher as well.



Figure 4.5.1: Oil rate (left) and cumulative oil profiles for Cases S1, S2, S3, S4, and S5 (with solution gas)



Figure 4.5.2: SOR profiles for Cases S1, S2, S3, S4, and S5 (with solution gas)

At high temperatures, although the latent heat of propane is small, sensible heat plays a role as well by heat conduction through reservoir rock. The temperature profiles of our simulations show that, chamber temperatures are the injection temperatures, and decrease gradually from chamber edges. Higher injection temperature will decrease the bitumen viscosity to a lower level than lower injection temperature, accelerating the viscosity reduction process and giving a higher oil rate as a consequence. This result also indicates the effect of sensible heat over that of the smaller latent heat from a higher injection temperature. Some researchers also have shown that, when bitumen is heated, the viscosity reduction due to high temperature surpasses the lowered solubility due to high temperature (Haghighat and Maini, 2012). Although in this study there is no solubility limitation (miscible process), it is a point worth noting. Overall, at the saturated conditions, high temperature/high pressure will give more oil production.

SOR follows the same trend as in the eight superheated cases: higher injectivity causes higher SOR. In this section, the higher injectivity is due to the higher injection pressure (initial reservoir pressures are the same among all cases).

To summarize, at saturated injection conditions, high temperature/high pressure give higher oil production as well as higher SOR. In order to have high oil production with low cost, economic analysis is needed.

4.6 Conclusions

Injecting at superheated conditions improves SOR. The more superheated, the less propane is retained, the more improved is SOR. When injection pressures are the same, cases that have higher injectivity give significantly more oil production. However it also leads to higher SOR.

Solution gas has a detrimental effect on the N-SolvTM process. It can considerably decrease the oil rate. It also affects the chamber shape, making it shorter and wider. Its amount should be minimized in the reservoir by proper process design.

When injecting at saturated injection conditions, high temperature/high pressure give higher oil production as well as higher SOR. In order to have high oil production with low cost, economic analysis is needed.

Chapter 5: Conclusions and Recommendations

5.1 Conclusions and Recommendations for Future Research

5.1.1 The N-SolvTM Process Mechanisms

The N-SolvTM process extraction mechanisms have been confirmed by numerical simulations. Our simulation results—the oil volume fraction of propane, its corresponding diluted oil viscosity, and the thickness of live and mobile live oil layer—are consistent with the reported experimental results.

The steep concentration gradient (high driving force), low oil viscosity (high oil flow rate), fast surface renewal, and better sweep efficiency all together make a competitive oil production rate possible compared to SAGD.

5.1.2 N-SolvTM Oil Production Performance

Without considering asphaltene precipitation induced formation permeability damage, and geological heterogeneity, such as high water saturation zones and shale zones, N-SolvTM can give a competitive or higher oil rate than SAGD depending on the injection temperatures.

High water saturation zones in the reservoir can make N-SolvTM chamber shape deviate from the traditional S-shape, thus amplify the effects of permeability damage. Literature also suggests high water saturation zone can cause oil rate of VAPEX decrease later on.

In future, more work should be done on the following:

- 1. Effects of high water saturation zone on N-SolvTM oil rate and solvent retention.
- Effects of permeability damage on heterogeneous reservoirs, since geological heterogeneity can affect the chamber shape.
- 3. Butane has been reported leaving less asphaltene in the reservoir, therefore should cause less permeability damage. The performance of injecting heated butane can be investigated.
- At what reservoir conditions asphaltene will be able to move should also be investigated.

5.1.3 Effects of Operation Temperature and Pressure on N-SolvTM Process

When injecting at superheated conditions, SOR is improved compared to saturated conditions. The more superheated the solvent, the less propane is retained, the more improved is SOR. Injecting at superheated conditions can give a slightly smaller cumulative oil production than saturated conditions at the same injection pressure. However the difference is very small. It can be concluded that different temperatures at superheated conditions do not affect cumulative oil production. When injection pressures are the same, the one that has higher injectivity can give significant amount of more oil production. However it also leads to higher SOR.

When injecting at saturated injection conditions, high temperature/high pressure give higher oil production, as well as higher SOR.

In future, more work should be done on the following:

- Varying injection pressure has been reported enhancing oil production by 20 to 30% in VAPEX, it is worth trying the same in N-SolvTM in terms of oil production and solvent retention.
- Increasing injection temperature while keeping pressure constant might be a good way to reduce heat loss in the chamber as chamber grows bigger. It might also reduce energy input. This can be investigated more.
- 3. Butane's performance on SOR can also be studied.

5.1.4 Effects of Solution Gas on N-SolvTM Process

Solution gas has a detrimental effect on N-SolvTM process. It can considerably decrease the oil rate. It also affects the chamber shape, making it shorter and wider. Its amount should be minimized in the reservoir by proper process design.

N-SolvTM pilot project purifies the produced fluid by flashing off methane before re-injection. Temperature and pressure have an effect on the amount of methane condensation at the extraction interface. Based on this, patent was invented to mitigate methane poisoning. Maintaining the operation temperature or pressure at bubble point conditions can remove methane from the reservoir and ensure continuous solvent condensation on the extraction interface.

5.2 Operational Recommendations

Based on the above findings, some operational recommendations can be made:

- Low permeability, low connate water saturation, and less shale zones reservoirs are preferred target reservoirs.
- 2. Low initial pressure reservoirs are preferred target reservoirs, for several reasons: 1) Heat loss in the depleted chamber is a concern, it is desired to keep a high injection rate to supply sufficient heat to the chamber edge. When injecting at a certain pressure, lower pressure reservoirs can give higher injectivities (higher injection rate). 2) Propane critical point is 96.6 °C, 4250 kPa. Injecting at conditions close to this point will not deliver much latent heat to the reservoir. To obtain more latent heat from condensing solvent process, a moderate temperature and pressure is desired. 3) From solvent retention point of view, high injection pressure is not economical, hence low pressure reservoir is preferred. 4) Asphaltene will be able to flow with oil when
temperature is higher than certain threshold value. Operation temperature should therefore not be higher than this temperature.

- 3. It is better to inject at superheated conditions to avoid solvent condensation (solvent retention) in depleted chamber due to heat loss.
- 4. For higher operation temperature, considering that asphaltene will become mobile as temperature increases, it might be necessary to periodically clean the production well to ensure less blockage of flow. Also, due to this reason, low viscosity reservoirs are better candidates because low operation temperature will be good enough to achieve desired oil rate.

In conclusion, with proper process design, and the assumptions used in the simulations conducted in this thesis, N-SolvTM can be a potential alternative for reservoirs that are not suitable for SAGD (thin pay zone, bottom aquifer, high clay content, etc.), it has low energy input, it is environmental friendly in terms of low GHG emission and low foot imprint, and it offers upgraded high quality bitumen, which also reduces the cost for upgrading. However, more research is required since there are still uncertainties associated with the process.

The biggest risk factor in N-SolvTM is the loss of solvent by migration into untargeted zones or thief zones. The process involves high rate injection of large volumes of solvent (many times the volume of oil produced) and it is essential that a very large fraction of this injected solvent be recovered with the produced oil. Unfortunately the loss of solvent to untargeted zones can only be evaluated by field trials. Therefore, the optimistic conclusions drawn from this simulation study need to be verified in field trials before the process can be considered economically viable.

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