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Comparative Simulation Study and Economic Analysis of Thermal Recovery Processes in Athabasca Reservoirs

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by

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ABSTRACT

Simulation studies of three thermal recovery processes used in Athabasca reservoirs have been carried out for a 10-year production period. The recovery processes studied are Steam-Assisted Gravity Drainage (SAGD), Fast-SAGD, and Expanding Solvent-SAGD (ES-SAGD). Normal pentane (n-C_5) was the solvent of choice used in ES-SAGD simulations with its molar concentration varied from 2% to 5.9%. The main objective of this study is to conduct an economic analysis of the three recovery processes with the goal of determining the most economically viable process. The economic indicator that will be assessed to ascertain the most viable recovery process is their Net Present Value (NPV.)

2D simulation studies based on homogeneous Athabasca reservoirs have been performed. Results obtained show that of the three recovery processes, Fast-SAGD had the lowest cumulative oil produced, followed by SAGD and ES-SAGD, the highest. The cumulative oil produced also increased with increasing molar concentration of n-C_5. Furthermore, it was shown that as expected, the CSOR of ES-SAGD was the lowest of them while that of Fast-SAGD was the highest. The CSOR of the ES-SAGD processes reduced as the concentration of the n-C_5 increased.

The economic analysis showed that of the three recovery processes, ES-SAGD is the most economically viable process. Furthermore, the effect of solvent on the viability of ES-SAGD over the other recovery processes is dependent on the price regime of pentane. In this analysis, two extreme price regimes were chosen and the result showed that for a low price regime, varying the molar ratios of n-C_5 had a significant effect on the NPV up to a point before its effect diminishes. In fact, increasing the molar concentration of n-C_5 from 2% to 3.76% significantly increased the NPV while further increasing it from 3.76% to 4% and thereafter to 5.9% had no noticeable effect. However, it seems that increasing it from 3.76% to 5.9% had a diminishing effect especially after the 3-year period.

Nevertheless, the significant NPV improvement ES-SAGD has over SAGD and Fast-SAGD diminishes once the price regime of pentane is more than 3 times that of oil. In fact, this high price regime showed that 5.9% molar concentration of $n\text{-}C_5$ is no longer more viable than the SAGD counterpart. There is still some benefit up till about 4% molar concentration of $n\text{-}C_5$ but this benefit is greatly diminished.

In conclusion, ES-SAGD has been shown to be the best recovery process for Athabasca reservoirs based on economics but further research is needed to evaluate the molar concentration that will provide the most economic benefit for a real Athabasca reservoir.

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DEDICATION

This thesis is dedicated to the memory of my mother who passed away December 2014.

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1. INTRODUCTION

Despite the clamour for alternative sources of energy, the driving force for fossil fuels exploitation is the high cost of producing alternative sources of energy. This will likely be the prevailing scenario in the next decades. Consequently, due to the high demand for energy worldwide, fossil fuels will likely dominate the energy source landscape in the near future.

However, as conventional sources of fossil fuels are becoming depleted worldwide, there is an increasing need for unconventional sources to sustain the world's energy needs. This has led to the shale oil and gas boom in the United States of America as well as heavy investments in the bitumen deposits of Canada.

Note that Canada, Venezuela, and USA accounts for about 80 % of the world's bitumen and heavy oil and Canada accounts for over 95 % bitumen deposits in North America. Most of Canada's oil sands are located in Northern Alberta. In fact, the oil sands of Northern Alberta are the largest bitumen deposits in the world and cover an area exceeding 140,000 square kilometres (Li, 2010). It is estimated to have about 270 million cubic meters (about 1.7 trillion barrels) initial volume in place with about 300 million barrels that can be ultimately recovered (Hein and Marsh, 2008; Beach and Purdy, 1997.) This makes Canada recoverable oil reserves to be second only to Saudi Arabia and it seems these immense reserves of bitumen will dominate the energy landscape of Canada for many decades.

Recovery of these bitumen resources has been a major focus of the oil and gas sector and academics for so many years. This has led to several recovery schemes. Broadly speaking, the main recovery methodologies used to recover bitumen from Alberta oil sands are surface mining and in-situ methods. When oil sands reservoirs become too deep for mining (over 80 m), other extraction methods have to be employed such as in-situ thermal extraction techniques to recover bitumen from deep oil sands

mines. Over 80 % of Alberta's oil sands are far too deep to be recovered by surface mining and, therefore, require in-situ technologies (Shin and Polikar, 2007). The in-situ technologies, which are mainly thermal driven methods, are favoured over primary recovery methods due to the viscous nature of Canada oil sands. In fact, Athabasca oil sands sometimes can have viscosities of about 2 million centipoise, which require immense energy input to make it mobile and, therefore, recoverable.

Several in-situ thermal methods have been developed in the last two decades. The most prominent one is Steam-Assisted Gravity Drainage (SAGD) developed by Butler in 1994 (Butler, 1994). SAGD has been a commercial success (Al-Gosayir and Babadagli, 1996) as it is used mostly in extracting very viscous bitumen in the order of 1 million centipoise such as those found in Athabasca.

SAGD generally employs two horizontal wells with one well as the injector and the other, a producer. Normally, the producer well is about 5 meters below the injector well. High-temperature and high-pressure steam is injected through the injector well which heats up the bitumen. Heating a 1 million centipoise bitumen to about 250 ˚C reduces its viscosity to about 4 cP enabling the bitumen to become mobile. The mobile bitumen falls under gravity to the producer, which is now produced with the help of pumps.

The major drawback to SAGD is its high capital and operating costs as well as the high carbon footprints relative to the extraction methodologies employed for conventional sources of energy. Consequently, there have been several derivatives of SAGD developed primarily to reduce these costs as well as minimize the carbon footprints. Vapour extraction, commonly known as VAPEX (vapour extraction) developed by Butler and Mokrys in 1990 (Butler and Mokrys, 1990) is one of SAGD variants. The only difference between SAGD and VAPEX is the replacement of steam with solvent in VAPEX. Despite this change, VAPEX has not become a big commercial success as SAGD. Other

variants are Steam-Alternating Solvent (SAS), which alternates solvent and steam during injection. This also has not achieved commercial success as SAGD.

One other variant of SAGD is Expanding Solvent Steam-Assisted Gravity Drainage (ES-SAGD). This is also very similar to SAGD but instead of injecting a pure fluid, a mixture of steam and solvent is injected through the injected well. This has been proven to have some commercial success (Leaute, 2002; Leaute and Carey, 2005.)

If the viscosity of the oil sands deposit is in the order of 10 thousand centipoise such as the bitumen from the Cold Lake formation, SAGD might not be a very efficient and cost-effective recovery method. Instead, the most preferred method is Cyclic Steam Stimulation (CSS). For this recovery method, instead of utilizing two wells for an injection and a producer, only one well is utilized, as both the injector and producer. This system is operated cyclically in this order: injection for some days followed by soak for a few days and lastly production for an extended period until production becomes very low. This cycle is repeated until the reservoir is depleted.

One last method that has received some attention is Fast-SAGD. This is a combination of SAGD and CSS. This method utilizes a CSS well in between two SAGD well pairs. Normally, to achieve better recovery, injection in the CSS wells starts after the steam chambers of the SAGD well pairs have reached the top of the reservoir (Nguyen et al, 2013).

1.1. Objectives of Study

The main objective of this thesis is to compare three recovery methodologies on the basis of economics to determine the most viable recovery method. The three recovery methods that will be undertaken are: SAGD, ES-SAGD, and FAST-SAGD. The economic indicator that will be assessed to ascertain the

most viable recovery process is their Net Present Value (NPV). The process that will be employed in this study to achieve the above stated objective is summarized below:

1. Explore SAGD, ES-SAGD, and Fast-SAGD recovery processes by varying operational parameters such as injector bottom hole pressure, producer bottom hole pressure and a maximum steam rate;
2. Use CMG CMOST to optimize each recovery process;
3. Compare the optimized cases on the basis of ultimate cumulative oil produced, a recovery factor, and a cumulative steam-oil ratio (CSOR);
4. Finally, compare the NPV of the various processes to ascertain the most viable option economically.

To reduce the complexity of this study, a homogeneous reservoir model was modelled in CMG STARS for the comparisons. Furthermore, WINPROP was used to determine fluids PVT profiles.

2. LITERATURE REVIEW

2.1. Steam-Assisted Gravity Drainage (SAGD)

Steam-Assisted Gravity Drainage (SAGD) was first developed by Butler in the 1990s and has become a commercially successful process. In fact, the SAGD process has proven to be the most successful method for commercial in situ thermal recovery of heavy oil and bitumen (Al-Gosayir and Babadagli, 1996). A high recovery factor of about 50-60 % has been achieved with the use of SAGD in Alberta (Al-Gosayir and Babadagli, 1996; Shin, 2012). In fact, Das (2005) stipulated that except for a few projects that involve surface mining, about 2.6 million bpd of bitumen out of the total 3.5 million bpd will be recovered by SAGD. This encapsulates the importance of SAGD in Alberta.

Al-Bahlani and Babadagli (2009) undertook a comprehensive review of the SAGD process. Despite its commercial success, there have been several research efforts to make it more economically and environmentally viable compared to recovery methods used for conventional sources of energy. Consequently, to achieve this objective, it became apparent that a better understanding of the SAGD process be gleaned.

2.1.1. Challenges of SAGD

Farouq-Ali (1997) listed some fundamental problems and limitations of the SAGD process. These are sand control, hot effluent/high water-cut production, frequent changes in operating regimes which requires close monitoring of the temperature and pressure changes, deterioration of production at late stages, and high operating costs. These were further expanded on by Singhal et al. (1998). Edmunds and Chhina (2001) listed the dependence on natural gas for producing steam, the substantial carbon footprint, and the significant water requirement as the challenges of the SAGD process. Butler and Yee (2002) opined that in the SAGD process, steam is usually wasted during injection and production as the

entire part of the reservoir that is depleted gets heated to steam temperature, whereas, to avoid steam coning, only the reservoir near the production well needs to be heated. There is also significant loss of heat to the overburden as pointed out by Butler (1997). Deng (2005) enumerated the downsides of the SAGD process such as the significant energy requirements, excessive CO_2 emissions, and costly water treatments.

2.1.2. Economics of SAGD

Consequently, due to the high operating cost incurred in SAGD operation, which ultimately correlates with a high carbon footprint, it is pertinent to make the economics of SAGD attractive to investors. One critical economic indicator of SAGD is the cumulative steam-oil ratio (CSOR). Gates and Chakrabarty (2005) captured this sentiment by showing that the economic efficiency of the SAGD process is mostly dictated by energy cost (that is, natural gas cost used in heating up the steam) and secondly by water recycling and treatment costs.

In fact, Edmunds and Chhina (2001) enumerated SAGD performance indicators. They listed a steam-to-oil ratio (SOR) as the most commonly used performance indicator for SAGD efficiency. They, however, mentioned that other performance indicators are bitumen production rates and recovery factors.

Scott (2002) and Collins (2007) contended with using CSOR as the appropriate energy efficiency indicator. Scott opined that an energy efficiency indicator should be based on the quantity of external gas required to produce 1 m^3 of bitumen. Collins on the other hand suggested using energy consumed as an energy efficiency indicator. This includes the energy recovered.

2.1.3. Factors that Affect the Success of SAGD

Singhal et al. (1998) enumerated factors that need to be considered for a successful SAGD project in bitumen of moderate viscosity (i.e., less than 35,000 cP) and high viscosity greater than 65,000 cP. These factors are:

1. Ensuring that a steam chamber is vertically confined to prevent it from heating the formation top. This involves locating projects in formations with fining upward geology or injecting a non-condensable gas with steam;
2. Creating a large steam chamber within a short period of time;
3. Ensuring that a pressure drop is minimal along a horizontal section of a well, especially injectors;
4. Choosing cost-effective well spacing between injectors - producers and injectors – injectors;
5. Choosing the most economic oil lifting options;
6. Oil saturation and pressure have to be high for the project to be attractive; and
7. Recirculating waste heat of effluent to warm surface facilities.

2.1.4. Effects of Reservoir Properties on SAGD Performance

Llaguno et al. (2002) in their analytical study of SAGD showed that porosity, oil pay thickness, and oil saturation affects SAGD performance more than permeability, viscosity and reservoir pressure. Oil pay thickness, in particular, was found to directly correlate with SAGD performance (Sasaki et al. 2001; Chan et al. 1997; Shin and Polikar, 2007; Singhal et al. 1998; Edmunds and Chhina, 2001; McCormack, 2001). That is, as net pay thickness increases, oil production also increases accordingly. In fact, Edmunds and Chhina (2001) stipulated that net pay thickness that is less than 15 m would be uneconomic because of the significant heat loss to the overburden and underburden for a thin reservoir,

which causes an increase in the CSOR. Bharatha et al. (2005) concluded that the presence of dissolved gas in bitumen causes a reduction in a bitumen production rate. That is, gas saturation reduces bitumen production.

Permeability, on the other hand, has been found to have a varied effect on bitumen production. Kisman and Yeung (1995) found that decreasing vertical permeability, K_v initially reduces the calendar day oil rate (CDOR) and oil-steam ratio (OSR) significantly. However, CDOR and OSR increased at later stages of production. This is in agreement with the findings of Nasr et al. (1996) that showed a decrease in OSR as permeability decreases. Shin and Polikar (2007) corroborated this finding by concluding that higher permeability correlates to higher ultimate bitumen recovery as well as lower CSOR.

Das (2007), Isaacs et al. (2001), and Yuan et al. (2001) have studied the effect of wettability. Das (2007) whose SAGD work was on carbonate reservoirs reported lower oil recovery with oil-wet reservoirs compared to their water-wet counterparts. This is in stark contrast to the findings of Isaacs et al. (2001) and Yuan et al. (2002) that showed higher oil recovery for oil-wet reservoirs.

2.1.5. Optimization of Operating Conditions and Well Placement Key to the Success of SAGD Projects

Operating conditions affect the amount of bitumen produced. The important operating parameters that have been observed to affect bitumen production are steam quality, subcool temperature or steam trap, injection pressure and temperature. Furthermore, well placement is critical to bitumen production - that is, having an optimum injector-to-producer spacing as well as injector-to-injector spacing is key to having a successful SAGD operation.

Gates and Chakrabarty (2005) emphasized using steam quality that is as high as possible at the sandface. This is because any condensate formed in the injected fluid has no significant thermal effect as it falls under gravity downwards from an injector to a producer. Chung and Butler (1989) in their experimental work found no significant difference on emulsified water content in the fluid samples with wet or dry steam.

Doan et al. (1999) stipulated that steam trap control should be used to reduce the amount of steam produced or wasted in the producer. Das (2005) enumerated three distinct advantages of steam trap control. They are: (1) energy conservation and reduction of SOR, (2) reduction of high vapor flow, which limits the lifting capacity of the well and surface facilities, and (3) reduction of sand movement through the liner, which increase the likelihood of erosion. Ito and Suzuki (1996) showed that the optimal steam trap control should be between 30 – 40 ºC. That is, the producer should be set to a temperature between 30 – 40 ºC lower than the saturation temperature of the injected steam. Edmunds (2000) on the other hand found that steam trap control of 20 - 30 ºC was optimum.

Operating SAGD at low or high pressure has led to different outcome for different researchers. However, it is worth noting that the first determinant of injection pressure is the bottom hole pressure (BHP). Injecting below BHP will essentially guarantee that there may be no injectivity. Consequently, the starting injection pressure is the formation pressure. In fact, it has been shown categorically that injecting steam at high pressure increases oil production (Kisman and Yeung, 1995; Robinson et al., 2005). However, one other complexity of the injection process is the pressure schemes maintained throughout the lifetime of the wells. Das (2005) and Butler and Yee (2002) showed that injecting steam at low pressures at the beginning of a SAGD process is highly uneconomical as it takes significant time for a steam chamber to grow to a reasonable size. However, reducing the injection pressure periodically as production continues has been found to be more beneficial. For example, Butler and Yee (2002) started injecting steam at 5 MPa but injection pressure was kept at 1 – 2 MPa two years later using

periodic steaming. They observed that CSOR decreased over time. This is in agreement with Das (2005) who started injection at 3.5 MPa for 2 years before reducing the injection pressure to 2 MPa and subsequently to 1.5 MPa. Das (2005) found a 15 % and 20 % reduction in SOR as the injection pressure was reduced to 2 MPa and 1.5 MPa, respectively.

2.2. Fast Steam-Assisted Gravity Drainage (Fast-SAGD)

Polikar et al. (2000) initiated the Fast-SAGD recovery process. This process combines SAGD and cyclic steam stimulation (CSS). The difference between Fast-SAGD and SAGD is that Fast-SAGD employs an additional single offset horizontal well in between and parallel to SAGD well pairs. The offset well is at the same depth with the SAGD producers. The offset wells are operated at significantly higher pressure than those used in SAGD. While it has been shown to have a promising future, there seems to be very few open literatures dealing with this recovery process.

The handful of literatures dealing with Fast-SAGD have consistently shown an improvement in bitumen recovery compared to a conventional SAGD. Polikar et al. (2000) first started a SAGD well pair process before employing an offset well 3 years into SAGD operations. They found that Fast-SAGD led to a dramatic increase in ultimate oil recovery, energy saving, and a significant reduction in SOR compared to a traditional SAGD. Nguyen et al. (2013) on the other hand waited until the steam chamber from SAGD well pairs reached the overburden before injecting steam into the offset wells. Nguyen et al. (2013) equally observed that Fast-SAGD led to significant incremental bitumen recovery and better energy efficiency compared to a traditional SAGD with the same number of wells. These results are in agreement with Gong et al. (2002), Shin and Polikar (2006, 2007), and Coskuner (2009). However, Coskuner (2009), Kamari et al. (2015) and Shin and Polikar (2007) found that CSOR was higher for Fast-SAGD processes compared to SAGD.

2.2.1. Fast-SAGD Configuration

Fast-SAGD combines SAGD wells with Cyclic Steam Stimulation (CSS) offset wells. It is generally recommended to drill offset wells in-between two SAGD wells after utilizing the SAGD wells for some time. In fact, Nguyen et al. (2013) recommended that before injecting steam through the offset wells, the steam chamber from the SAGD wells should have made its way to the top of the reservoir. Consequently, Fast-SAGD initially starts out with just SAGD wells before introducing the offset wells in order to access some of the oil that may have been bypassed by SAGD.

There are three major phases involved in the offset wells, namely, an injection phase, a soak phase, and a production phase. These phases constitute a cycle. The injection phase comprises of the period where steam is injected into the offset wells. The injection phase typically lasts for several days or even months depending on the conditions. The soak phase follows the injection phase and it would normally not exceed two weeks. While sufficient time is required for efficient heat transfer from the steam to the bitumen, too much time will result in the steam losing its latent heat and thereby will not have any effect on viscosity reduction and oil production. The last phase of the cycle is the production phase, which typically lasts for several months. Typically, it is one year minus the number of days used for the injection and soak phases.

2.3. Expanding Solvent Steam-Assisted Gravity Drainage (ES-SAGD)

The goal of improving oil recovery while at the same time minimizing carbon footprints led to the discovery of ES-SAGD. It became apparent that co-injecting solvent with steam could further reduce bitumen viscosity beyond what is usually obtained with just steam injection (SAGD). This has the

added benefit of minimizing energy requirement due to the lower temperature required for ES-SAGD operations compared to SAGD (Gates and Chakrabarty, 2008). ES-SAGD involves co-injecting a single hydrocarbon solvent or a combination of hydrocarbon solvents with steam to recover more bitumen. The solvent, which is injected in the vapour phase, expands along with steam and mixes with bitumen at the steam chamber edge. This further causes a significant drop in bitumen viscosity due to the presence of the solvent.

ES-SAGD was initiated by Nasr and Isaacs in 1999 while working at the then Alberta Research Council (Nasr and Isaacs, 2001). However, it was Allen (1973), Brown et al. (1977), and Nenniger (1979) who first obtained patents that showed adding of solvent will be beneficial to extracting viscous oil. In selecting the appropriate solvent, critical issues need to be resolved. For example, the selected solvent has to evaporate and condense at nearly the same conditions as the water phase. Based on this criterion, only high carbon hydrocarbon solvents (C6 and C7) will suffice. Nevertheless, others have found that using heavier solvents (above C7) improves oil rates (see Redford and McKay, 1980; Li et al., 2011b). Still others have seen better results with lighter solvents (Govind et al. 2008; Ardali et al., 2010). In addition, the phase of the solvent, injected in the vapour phase, must be maintained at the sandface. Moreover, Nasr and Isaacs (2001) stated that the hydrocarbon solvent selected should have its evaporation temperature within a maximum range of about ±150 ºC of the steam temperature.

Govind et al. (2008) considered the impact of several key variables on bitumen recovery and SOR. Among the variables were a solvent fraction, a solvent type, a preheating period, injection pressure, and an injection strategy. They found that butane facilitated better solvent dilution in bitumen which consequently led to a more lowering of bitumen viscosity. This is because butane is more soluble in bitumen at higher pressures compared to the hydrocarbons tested. As a result, butane co-injection produced more oil and lower SOR compared to other co-injections. However, some other researchers

have found that an oil rate increases as the solvent become heavier than butane (Redford and McKay, 1980; Li et al., 2011b).

Nasr and Isaacs (2001), Nasr et al. (2003), and Gates (2007) all agreed that ES-SAGD requires less amount of steam to recover the same amount of bitumen compared to SAGD. This, therefore, suggests that ES-SAGD will bring a net benefit in the reduction of carbon footprints. In fact, Kershavarz et al. (2014) found that for ES-SAGD, solvent accumulation in the gaseous phase resulted in a chamber-edge temperature that was significantly lower than that of SAGD.

One other advantage of ES-SAGD over SAGD is the reduction of the residual oil saturation in ES-SAGD compared to SAGD. Nasr and Ayodele (2006) found that residual oil saturations in their ES-SAGD experiments were lower than in SAGD. Although Nasr and Ayodele (2006) used Cold-Lake live oil bitumen, Deng et al. (2010) and Kershavarz et al. (2014) obtained the same finding with Athabasca bitumen. Yazdani et al. (2011) also concurred with the above findings but in addition proffered the reasoning behind it. Yazdani et al. (2011) stipulated that the reason there is a lowering of the residual oil saturation in ES-SAGD is due to a reduction in the interfacial tension between phases during steam-solvent co-injection. However, Jha et al. (2012) proffered a different explanation. That is, ES-SAGD reduces residual oil saturation because the condensed solvent that was initially mixed with the bitumen partially evaporates.

However, Nexen conducted an ES-SAGD pilot project at Long Lake. The solvent used was named Jet B, which is a mixture of C7 to C12. They observed minimal improvements of 6 % in the oil rate and a 7 % reduction in SOR according to Nexen (2007) and Orr (2009). These improvements were less than those obtained in the pilot tests undertaken by EnCana (SAP) and Imperial Oil (LASER). In addition, Nasr et al. (2003), Gates, 2007, Gates and Chakrabarty (2008), Ivory et al. (2008), Li et al. (2011a, b),

Yazdani et al. (2011), and Kershavarz et al. (2014) all obtained impressive oil rates from ES-SAGD compared to SAGD.

2.4. Economic Analysis of SAGD, ES-SAGD, and Fast-SAGD

Many literatures have dealt with the economic analysis of SAGD. Some of these studies have detailed investigations of SAGD bitumen supply costs while others are limited economic analyses. The detailed supply cost normally considers the supply costs of bitumen produced by an entire SAGD plant with many well pads (CERI, 2017; Giacchetta et al., 2015). The plant normally could be assumed to produce up to 30,000 bpd or more. On the other hand, a limited economic analysis normally involves calculating the NPV of a subset of the entire plant such as assessing production from one injector and one producer (Edmunds and Chhina, 2001; Deng, 2005; Keshavarz et al., 2014; Lawal, 2014.)

For many years, the gold standard for a detailed supply cost analysis of SAGD has been the Alberta Energy Regulators (AER) and Canadian Energy Research Institute (CERI). Every year these organizations engage in detailed analyses of the bitumen supply costs. CERI publishes their finding every year and their recent supply cost study was published this year (CERI, 2017a,b.)

While there are many literatures that have dealt with the economic analysis of SAGD, only a few literatures have evaluated that of ES-SAGD such as Deng (2005), Gupta and Gitins (2009), and Keshavarz et al. (2014). The few available literatures dealing with the NPV of ES-SAGD have only done a very limited economic analysis. In fact, most of the economic assumptions made today are still based on the work of Deng (2005). All the studies that did a comparative economic analysis of SAGD and ES-SAGD uniformly concluded that ES-SAGD is significantly economically more viable than SAGD except Deng (2005). Deng found that SAGD and ES-SAGD had comparable supply costs.

There is paucity of literature that has dealt with the economic analysis of Fast-SAGD in comparison with SAGD. Kamari et al. (2015) is the only recent available literature that gave an in-depth economic analysis, albeit their work focussed on naturally fractured reservoirs. In fact, they used net present value (NPV) as an economic indicator to assess the relative economic viability of the SAGD process with that of Fast-SAGD. In their analysis, they found that Fast-SAGD was economically more viable than their SAGD counterpart. However, there is no open literature that has dealt with the economic analysis of Fast-SAGD in non-naturally fractured reservoirs except the work of Shin and Polikar (2006). This work is very dated and was performed on SAGD and Fast-SAGD processes that were not optimized, thereby making the economic comparisons not very reliable. Furthermore, their analysis which formulated a STEP (Simple Thermal Efficiency Parameter) program as alternative to NPV relied mostly on very old and currently unreliable supply cost regime.

3. RESERVOIR SIMULATION STUDY

3.1. Athabasca Bitumen

Athabasca bitumen has been widely studied and its properties are well documented. It is regarded as extra heavy because of its highly viscous nature and has been shown to contain about 20% asphaltenes and 80% maltenes (Khan et al., 1984). Its average molecular weight also varies from about 500 – 600 kg/kmol. For this study, the value of 572.5 kg/kmol obtained by Mehrotra and Svrcek (1986) was used.

For simulation purposes, one of the most important property is its viscosity. Consequently, there have been several literatures devoted to this particular area, especially the effect of temperature and pressure on the viscosity. Jacobs et al. (1980) studied the viscosity of gas-saturated Athabasca bitumen and came up with several relationships between viscosity and temperature of carbon-dioxide-, methane-, and nitrogen-saturated bitumen. Khan et al. (1984), on the other hand, evaluated viscosity models for gas-free Athabasca bitumen experimentally. They assessed several viscosity models and came up with a linear viscosity model that works almost as well as a nonlinear model. However, one of the most cited work in this area is that of Mehrotra and Svrcek (1986) who empirically studied the correlation of the viscosity of compressed Athabasca Bitumen with temperature and pressure. They came up with a natural logarithmic correlation that has been used for very significant work involving dead bitumen (C_D) and live bitumen. In this context, C_D is bitumen that has less than 0.04 mole fraction of C_1. In fact, the Computer Modelling Group (CMG) uses it in their STARS simulator. This viscosity model is represented in Equation (3.1):

$$lnln(\mu_B) = 22.8515 - 3.5784lnT + 0.00511938P$$

(3.1)

where μ_B is the viscosity of bitumen, P is the gauge pressure in MPa and T is the temperature in K. For this particular study, it was assumed that the composition of Athabasca bitumen comprises of 0.04 molar fraction of C_1 with the remainder being C_D. Using Equation (3.1), the viscosity of the live bitumen was calculated with respect to temperature and pressure. It was assumed that the bottom of the formation (243 m) sits at 1800 kPa. This implies that P used in Equation (3.1) is 1.8MPa. Figure 1 shows the variation of viscosity of the bitumen used in this simulation with temperature.

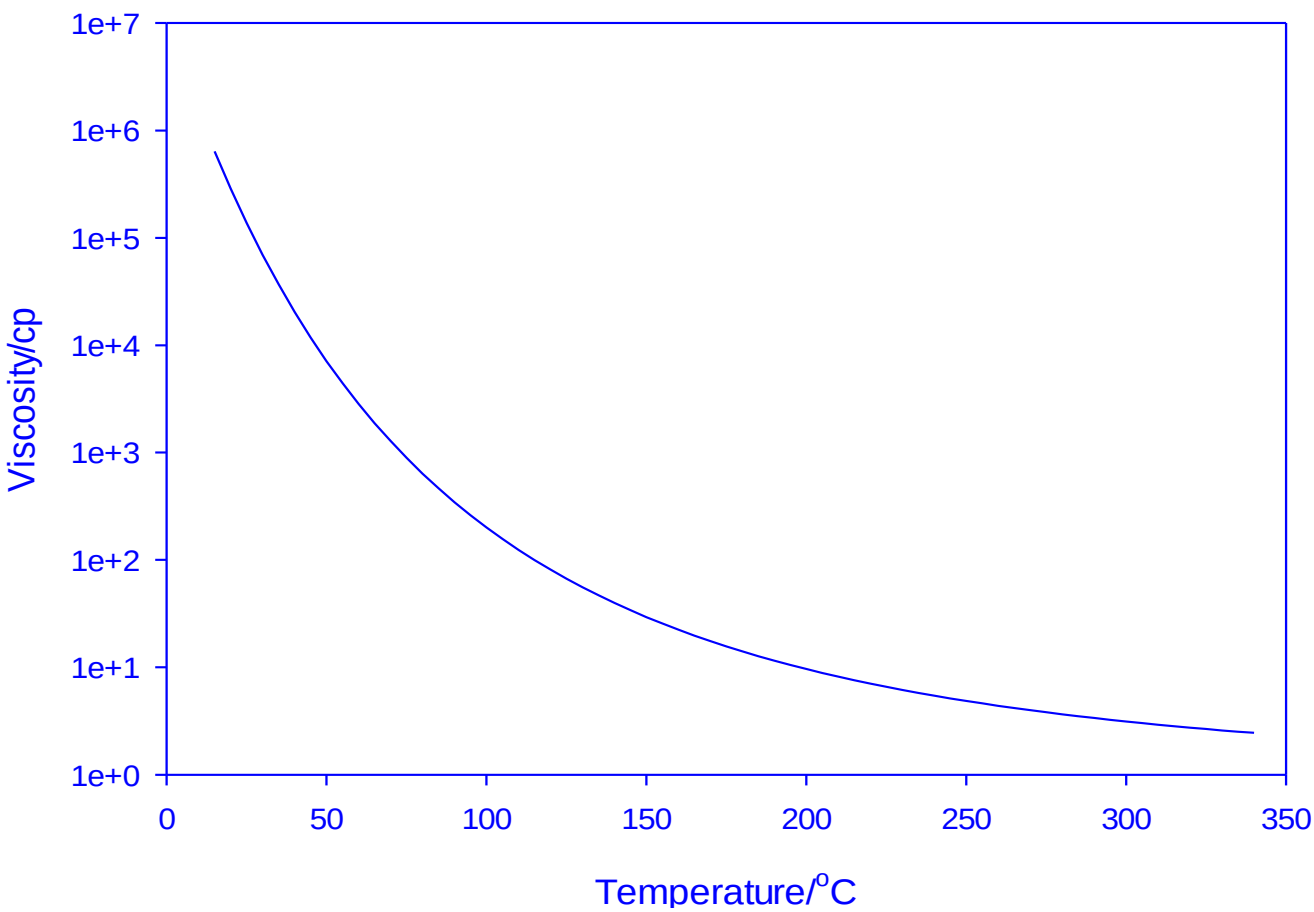

Figure 1: Variation of the viscosity of Athabasca bitumen with temperature

3.1.1. Viscosity of Bitumen Mixed with Solvent

Since this study encompasses expanding solvent-steam-assisted gravity drainage, pentane (*n*-C_5) was chosen as the co-injected solvent for the ES-SAGD simulation study. This choice was informed by the superior drainage rate of pentane that was observed by Keshavarz et al. (2014) compared to the corresponding SAGD case. For ES-SAGD applications, it is required that we are able to calculate the viscosity of a mixture of bitumen with pentane at various temperature. However, before the viscosity of the mixtures is determined, the viscosity of pentane with temperature is required. For this purpose, a correlation used by CMG's STARS library (CMG, 2015) for solvents viscosity was employed. This correlation is shown in Equation (3.2):

$$\mu_s = A_{visc} \times Exp\left(\frac{B_{visc}}{T}\right)$$

(3.2)

where $A_{visc} = 0.019104$, $B_{visc} = 722.23$, μ_s is the solvent viscosity, and T is the temperature in K. Figure 2 shows the variation of *n*-C_5 with temperature.

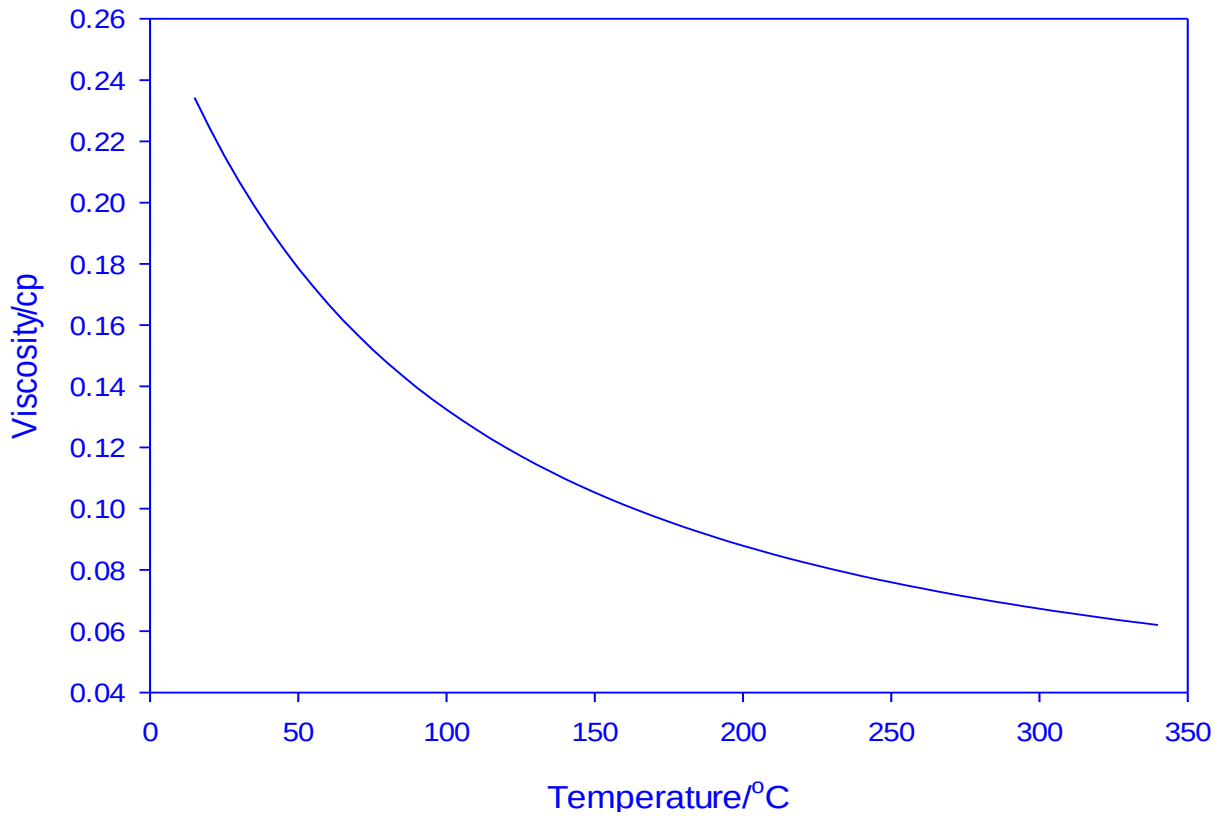

Figure 2: Variation of the viscosity of *n*-C_5 with temperature

To obtain the viscosity of the blend of *n*-C_5 with bitumen, a linear correlation proposed by Cragoe (1933) was used. Cragoe's correlation is shown in Equation (3.3):

$$\frac{1}{ln(2000\mu_m)} = \frac{w_s}{ln(2000\mu_s)} + \frac{1-w_s}{ln(2000\mu_B)} \tag{3.3}$$

where μ_m is the viscosity of the blend and w_s is the weight fraction of *n*-C_5. The weight fraction of solvent was obtained by making the assumption that 0.2 molar fraction of *n*-C_5 mixes with 0.8 molar fraction of bitumen. This assumption is consistent with CMG STARS and also verified by Keshavarz et al. (2014) who showed that there is no significant change when you assume a larger mole fraction of 0.5 for the solvent. Figure 3 shows the viscosity of the blend.

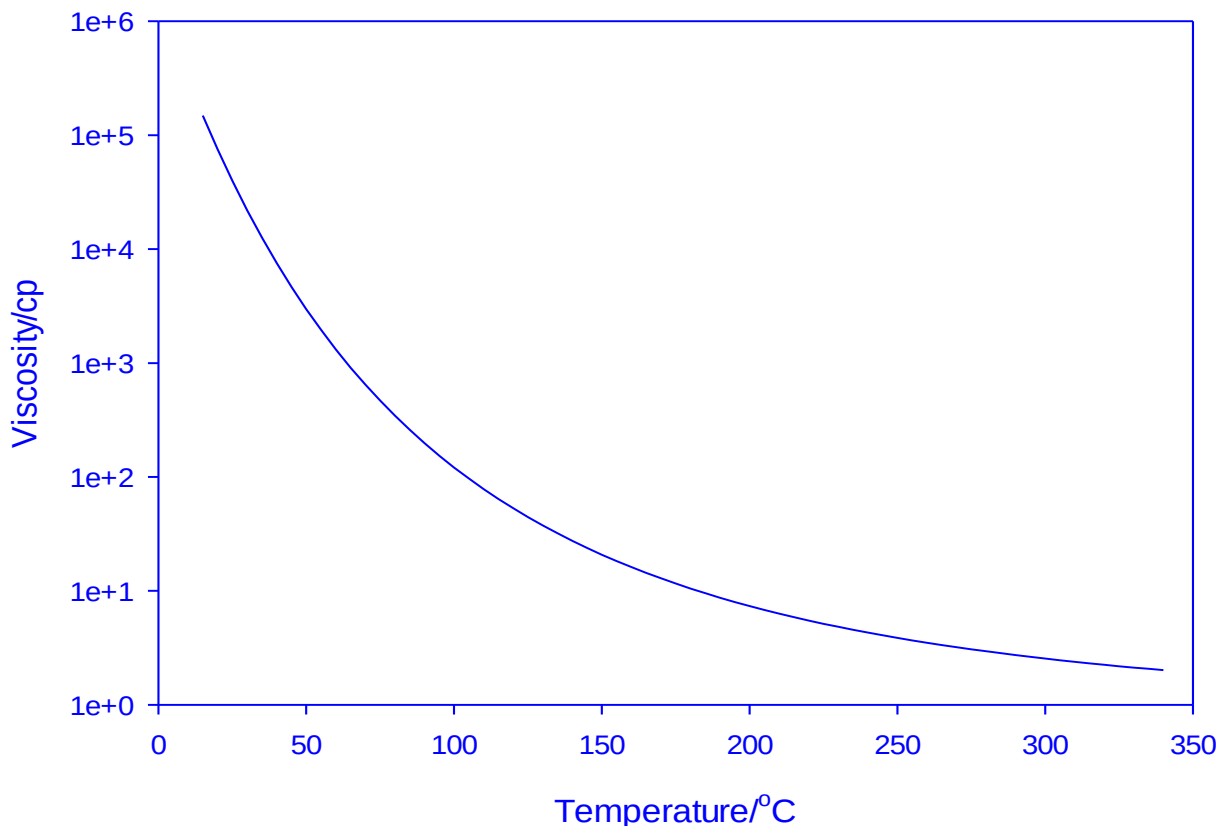

Figure 3: Variation of the viscosity of Athabasca bitumen mixed with nC_5 versus temperature

3.2. CMG WinProp Simulation

The fluid *K*-value correlation coefficients and pressure/volume/temperature (PVT) tables needed for this study were obtained using CMG WinProp and thereafter the generated *K*-value coefficients and PVT tables were now exported into CMG STARS for simulation. The desired input into WinProp are the viscosities of the solvent, blend of bitumen with solvent, and that of dead bitumen versus temperature. For the blend viscosities, an assumption that 0.2 molar fraction of n-C_5 mixes with 0.8 molar fraction of bitumen at the chamber edge was made.

One set of generated *K*-value coefficients and PVT tables were used for all the different recovery processes undertaken in this research. However, the injected fluid was changed accordingly in STARS

Wells and Recurrent section depending on the process that was studied. The *K*-values and other PVT data that were used in this study are shown in Appendix A.

3.3. Reservoir Simulation Model

An Athabasca reservoir with a horizontal length of 900 m (J-direction), width 151 m (I-direction), and depth of 33 m (k-direction) was chosen for this study. The top and bottom of the pay zone is, respectively, 210 m and 243 m TVD. The reservoir was assumed homogeneous. Consequently, assuming no pressure drop and flow resistance along the horizontal wellbore, a 2D model is suitable for studying SAGD, ES-SAGD, and Fast-SAGD recovery processes since the steam chamber growth will be the same in any J-plane cut. Figure 4 shows the 2D configuration employed for the SAGD and ES-SAGD studies while Figure 5 shows the configuration of the Fast-SAGD. Note that this Fast-SAGD configuration was chosen in order to keep the number of wells uniform in all the recovery processes thereby making the capital costs similar. The horizontal length is only represented by a block of 50 m; therefore, to account for the total production and injection from the reservoir, a factor of 18 was applied to all recovery processes.

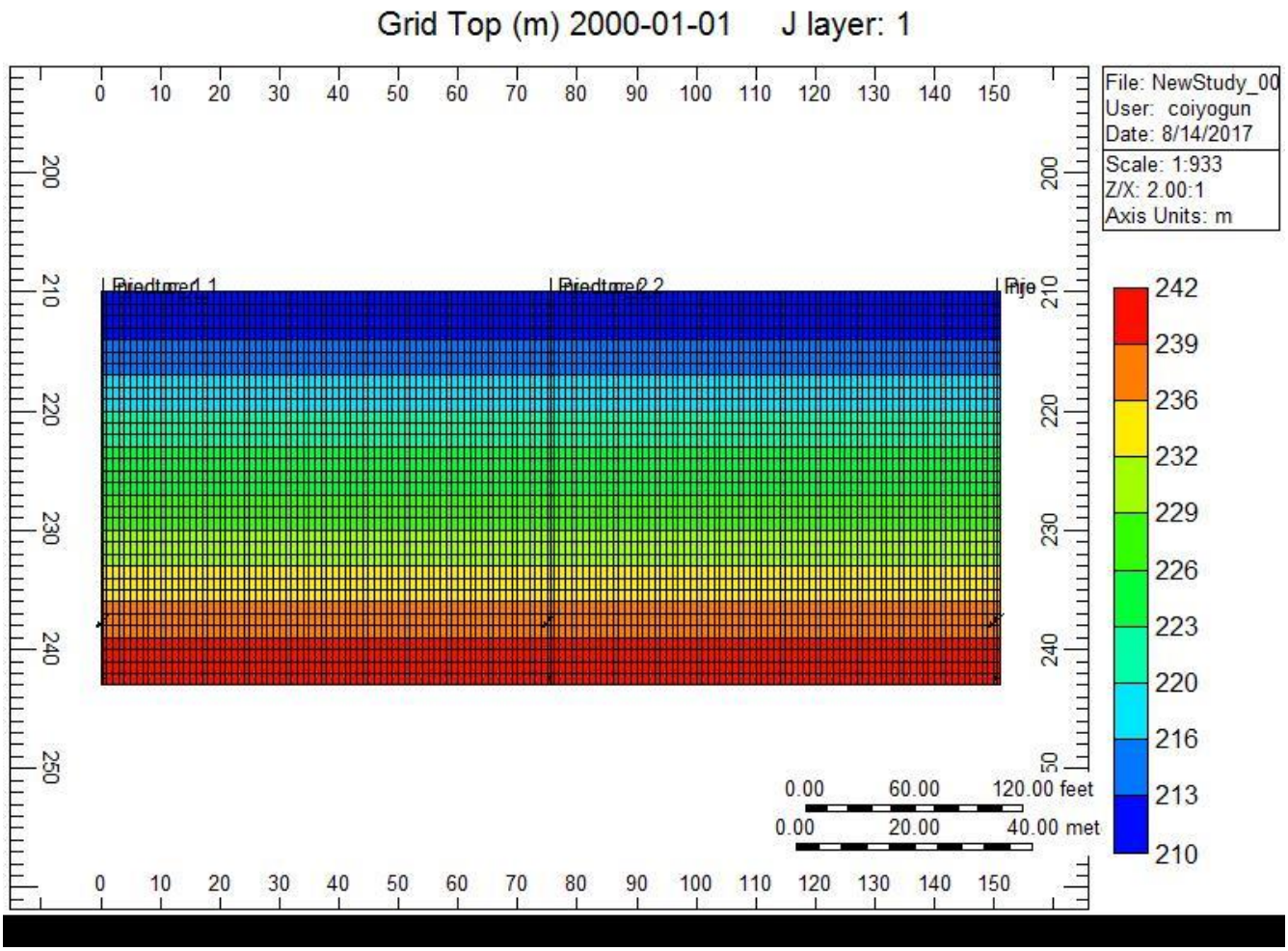

Figure 4: 2D grid configuration used for SAGD and ES-SAGD

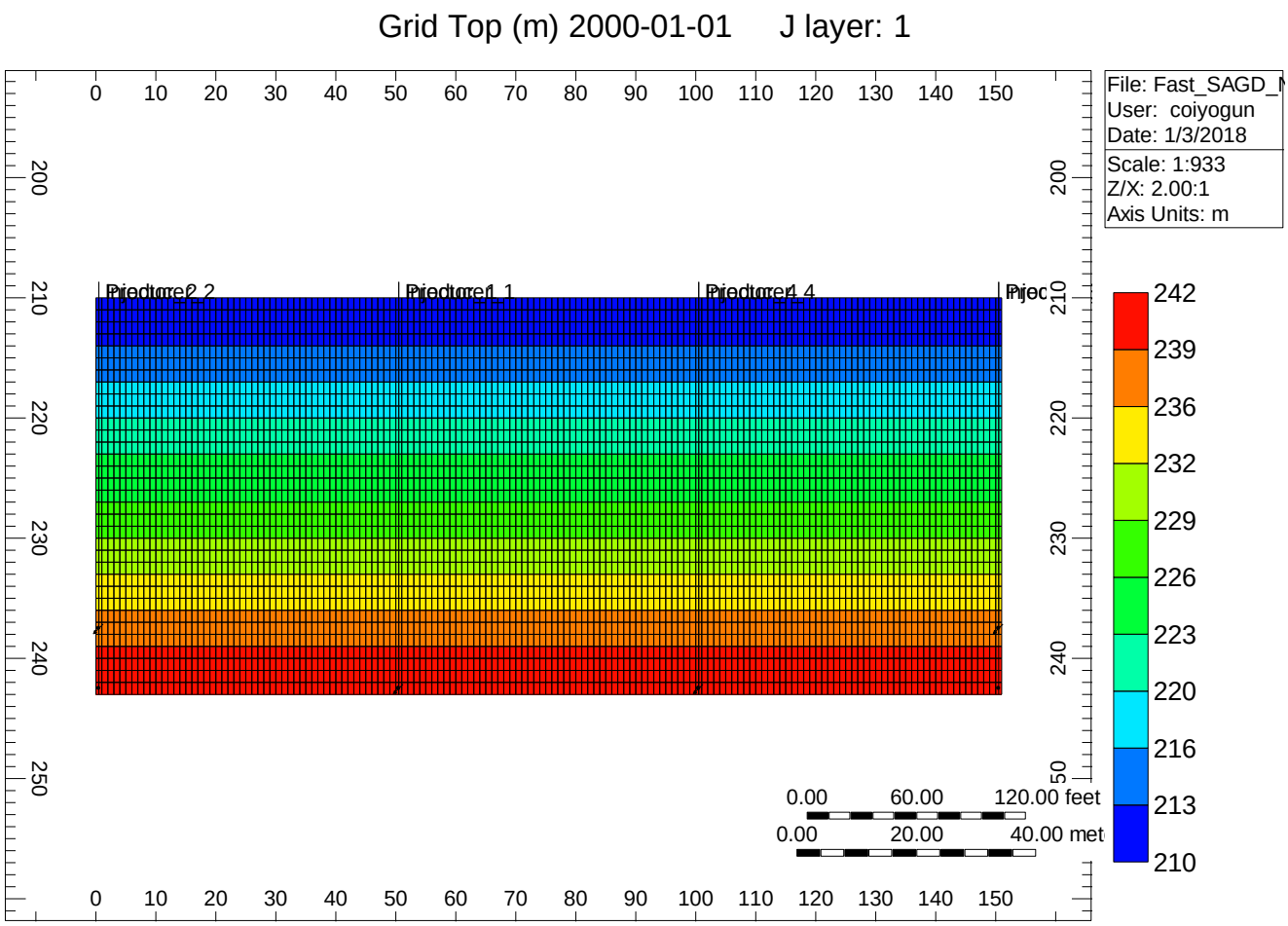

Figure 5: 2D grid configuration used for Fast-SAGD

The producer is 0.5 m from the bottom of the reservoir pay zone and a constant injector/producer vertical spacing of 5 m was maintained for all simulations. The 2D grid system includes 33 blocks in the vertical direction (K-direction) with 1 m spacing between blocks, 151 blocks in the lateral direction (I-direction) with 1 m spacing between blocks, and 1 block of 50 m grid size along the length of the reservoir. A Cartesian grid type was used.

For the SAGD and ES-SAGD simulation models, there are three well pairs with the following block locations: Injector 1 – J layer: 1, I layer: 1, K Layer: 28; Injector 2 – J Layer: 1, I Layer: 76, K Layer: 28, and Injector 3 – J Layer: 1, I Layer: 151, K Layer: 28. The corresponding producers were all

located in block K Layer: 33 (That is, 5 m below the injector). The edge wells (Injector 1 and Producer 1 as well as Injector 3 and Producer 3) had a fraction of 0.5. This fraction dictates that only 50% of well index and flow rates will be applied to the edge wells.

However, for the Fast-SAGD simulation model, the locations of the wells are different. Figure 5 shows that there are 4 locations of the wells in the I-direction as opposed to three locations in the SAGD/ES-SAGD models. Nevertheless, the total number of wells are the same since Fast-SAGD alternates between CSS wells and SAGD wells. In Figure 5, it can be seen that the central wells are the CSS types while the edge wells are two SAGD well pairs. Note that while the SAGD wells in each location are in pairs (injector and producer), the CSS wells are only one as it serves as both the injector and producer. To this end, there are four wells for each recovery process since the edge wells are counted as half-wells for each well. That is, the SAGD well pairs at each edge counts as one well making a total of two SAGD wells and two CSS wells.

Properties of the reservoir used in the simulation are summarized in Table 3.1:

Table 3.1: Reservoir properties used in the simulation model

Properties	Values
Porosity	0.35
Oil mole fraction (CH_4)	0.04
Oil mole fraction (Heavy)	0.96
Vertical permeability (md)	2500
Horizontal permeability (md)	5000
Reservoir pressure at bottom of reservoir (243 m TVD) (kPa)	1800
Initial reservoir temperature (ºC)	11
Initial oil saturation	0.87
Initial water saturation	0.13
Rock heat capacity (kJ/m^3 ºC)	2350
Formation compressibility (1/kPa)	1.8E-5
Rock thermal conductivity ((kJ/m day ºC)	660
Over/Underburden thermal conductivity (kJ/m day ºC)	150
Over/Underburden heat capacity (kJ/m^3 ºC)	2350
Bitumen thermal conductivity (kJ/m day ºC)	11.5
Gas thermal conductivity (kJ/m day ºC)	2.89
Water thermal conductivity (kJ/m day ºC)	53.5

3.3.1. Relative Permeability Model Used

The rock-fluid relative permeability data was picked from CMG STARS for a typical Athabasca oil sands reservoir. Figure 6 shows the relative permeability of oil and water with respect to water saturation while Fig. 7 represents the gas and oil counterparts.

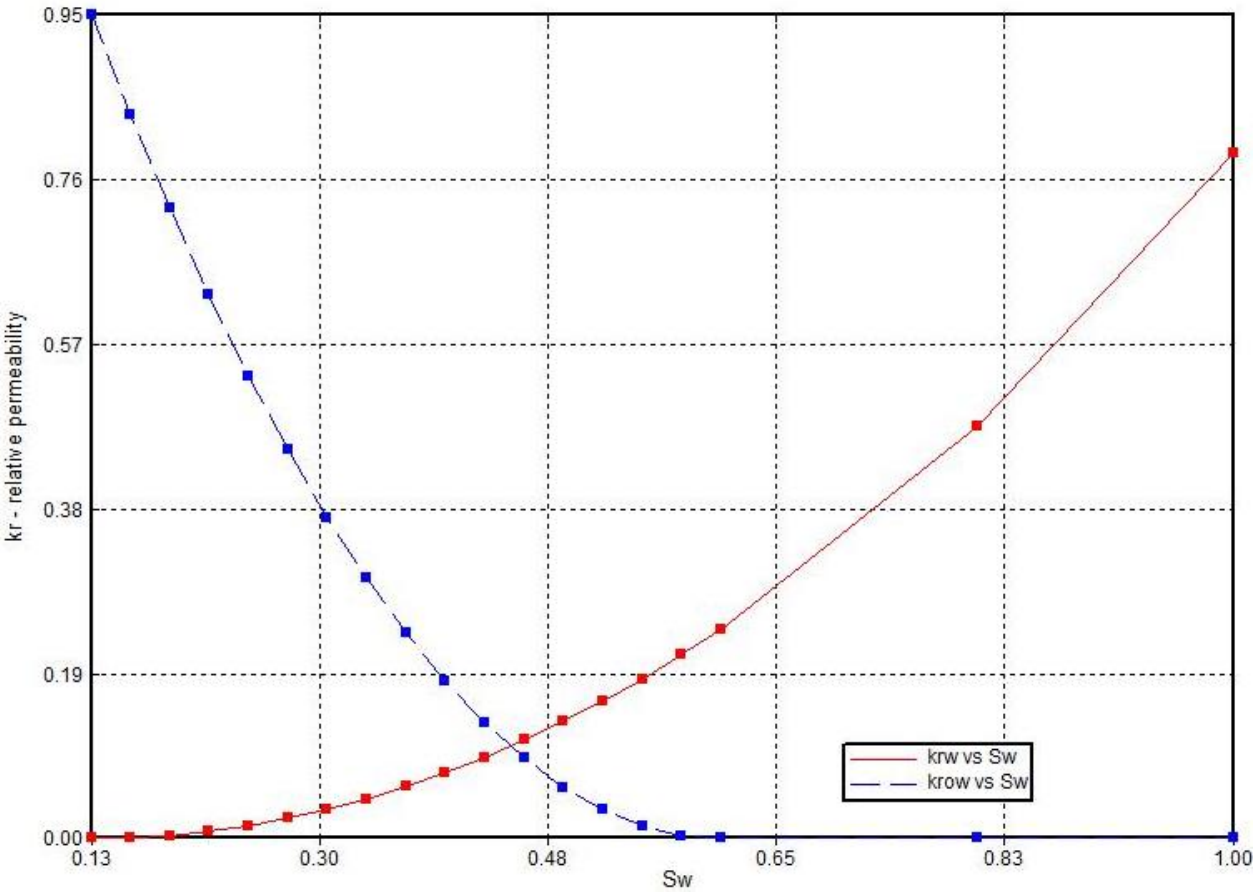

Figure 6: The rock-fluid relative permeability curve for water and oil

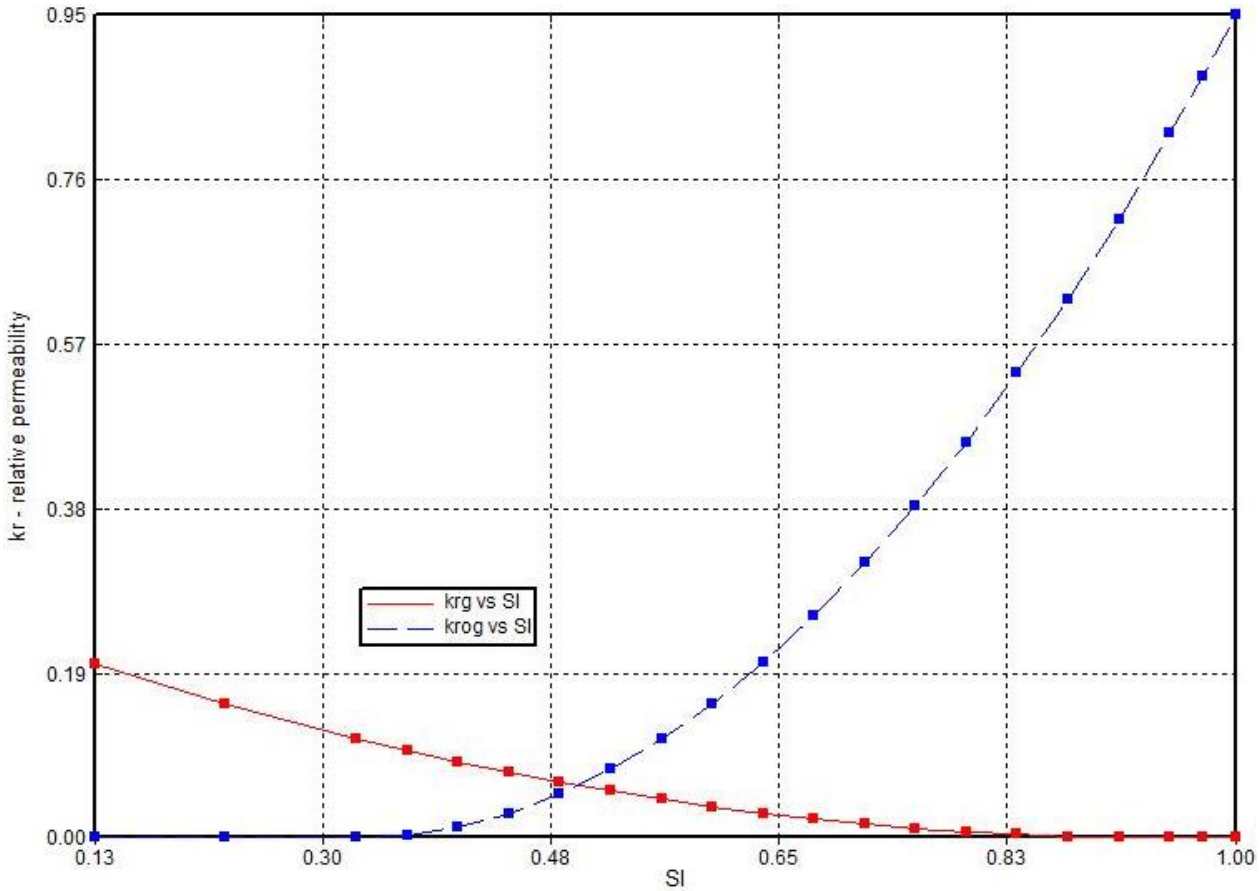

Figure 7: The rock-fluid relative permeability curve for oil and gas

3.3.2. Pre-Heating Period, Simulation Period, and Steam Quality

In the field, the reservoir usually has a pre-heating period of about 2 months or more. During this period, the reservoir is heated up by injecting high quality steam or a steam-solvent mixture from both the injector and producer until thermal communication is established between the producer and injector. Once communication is achieved, the wells are opened and injection of steam or steam-solvent mixtures continues from the injector while bitumen and steam migrate to the producer for production.

However, in this particular simulation research, the pre-heating of the reservoir is achieved with a line heater system. This is just a space heater, which is used to model the real field steam injection process during the pre-heating period. For all simulations carried out in this research, the pre-heating period

was 4 months after which the wells were now opened for injection of steam or steam-solvent and production of oil and water. The simulation performance is for a 10-year period of which the first 4 months was designated for pre-heating.

Steam quality is required to be very high for SAGD processes but could be lower for CSS and ES-SAGD processes. However, for this research, the steam used for all the recovery processes was 0.9 except at the CSS wells which had a quality of 0.8.

4. RESULTS AND DISCUSSIONS

This chapter will cover the presentations of the results and will conclude by evaluating the Net Present Value (NPV) of the different processes.

4.1. Operating Constraints

For SAGD and ES-SAGD simulation, a base case was chosen arbitrarily but with input from the work of Nguyen et al. (2013). Consequently, the base case starts out with very high injection pressure compared to any optimized case that will be presented below. This essentially reduces the impact of optimization as only minimal improvement in recovery factor can be achieved as a result of starting out with high injection pressure. The base case for SAGD and ES-SAGD has the following operating constraints. The injectors had a maximum bottom hole pressure (BHP) and a maximum surface total phase rate (STF) of 4500 kPa and 1000 m^3/day, respectively. On the other hand, the producers had three distinct constraints, namely: minimum BHP of 2500 kPa, a maximum steam rate of 5 m^3/day, and a maximum surface oil rate (STO) of 150 m^3/day.

For Fast-SAGD, the base case was treated differently. First, it was necessary to determine the time it took for the steam chamber to get to the reservoir overburden. This piece of information came from the SAGD counterpart. Once this time was known, the two CSS injectors are opened to either coincide with this date or after. In this particular study, it was found that the most optimal date was 22 months after the commencement of the SAGD processes in the two SAGD wells. The two SAGD wells for the Fast-SAGD process had the same operating constraints as the SAGD base case. The two CSS injectors/producers had the following operating constraints: the injectors had a maximum bottom hole pressure and maximum STF of 11500 kPa and 1000 m^3/day, respectively, while the producers had a

minimum BHP of 2500 kPa and a maximum steam rate of 5 m^3/day. Furthermore, two groups were created that combined each CSS well with each group being controlled by a cycling group control. The cycling group control was divided into different phases with one phase being the injection phase followed by a soak phase and lastly a production phase. Each cycle comprising of these three phases was for a one year duration. Three cycles were done after which the CSS wells were left opened for the remainder of the 10-year production period.

Figure 8 shows the Fast-SAGD base case with the start of the CSS wells indicated (i.e., the edge wells). Note that the SAGD wells for this case commenced operations January 1, 2000 while this figure also shows the initial commencement of CSS wells on January 1, 2002.

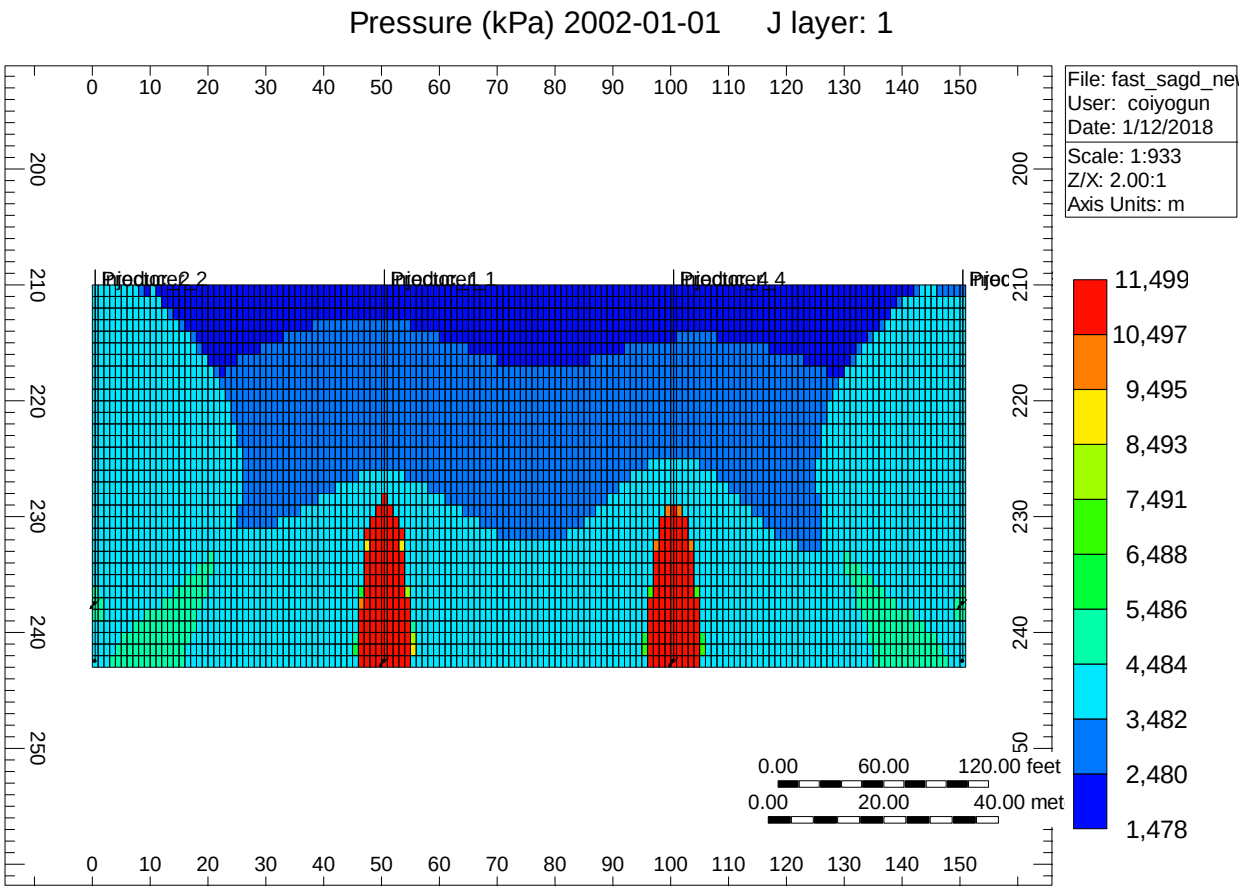

Figure 8: Start of CSS wells in a Fast-SAGD process

4.2. Base Case Compared with the Optimized case

Normally, a base case is chosen and simulated for each of the recovery process after which it is optimized to get the best operating conditions that maximizes Net Present Value (NPV). Table 4.1 shows the optimized constraints and operating conditions with their CMOST range used for all the recovery processes. Note that the wells are named from left to right (i.e. injection1 for instance, is the leftmost well while injection4 is the rightmost well when applicable.) The optimization was done using CMG CMOST and the objective function is to maximize Net Present Value (NPV). The engine setting in CMOST used for the optimization is the CMG Designed Evolution, Controlled exploration (DECE) algorithm and between 150 – 350 runs were simulated for each recovery process before arriving at an optimized case. Consequently, over 1000 runs were carried out in this study.

Thermal Process	Constraints (CMOST range)	Injector 1	Injector 2	Injector3	Injector 4	Constraints (CMOST range)	Producer1	Producer2	Producer3	Producer4
SAGD	Max. BHP (3375-5625 kPa)	4680	4668.75	4871.25	n/a	Min. BHP (1875-3125 kPa)	2537.5	2925	2537.5	n/a
	Max. STF (7500-1250 m^3/d)	11025	8075	9200	n/a	Max. Steam Rate (1-6.25 m^3/d)	1.13125	1.105	1.39375	n/a
						Max. STO (112.5-250 m^3/d)	138.25	234.2	224.5	n/a
ES-SAGD										
2% nC5	Max. BHP (3375-5625 kPa)	4927.5	4927.5	4927.5	n/a	Min. BHP (1875-3125 kPa)	2631.25	2525	3075	n/a
	Max. STF (7500-1250 m^3/d)	10150	8975	12150	n/a	Max. Steam Rate (1-6.25 m^3/d)	1.13125	1.07875	1.13125	n/a
						Max. STO (112.5-250 m^3/d)	232.8125	144.125	186.0625	n/a
3.76% nC5	Max. BHP (3375-5625 kPa)	5625	5625	5625	n/a	Min. BHP (1875-3125 kPa)	1875	2212.5	2668.75	n/a
	Max. STF (7500-1250 m^3/d)	7500	9400	12500	n/a	Max. Steam Rate (1-6.25 m^3/d)	1	1	1	n/a
						Max. STO (112.5-250 m^3/d)	250	189.5	250	n/a
4% nC5	Max. BHP (3375-5625 kPa)	5625	5625	5625	n/a	Min. BHP (1875-3125 kPa)	1875	1875	3125	
	Max. STF (7500-1250 m^3/d)	12500	12150	7500	n/a	Max. Steam Rate (1-6.25 m^3/d)	1	1	1	
						Max. STO (112.5-250 m^3/d)	112.5	122.125	244.5	
5.9% nC5	Max. BHP (3375-5625 kPa)	5625	5625	5625	n/a	Min. BHP (1875-3125 kPa)	2618.75	1900	2937.5	n/a
	Max. STF (7500-1250 m^3/d)	10700	12400	10725	n/a	Max. Steam Rate (1-6.25 m^3/d)	1	1	1	n/a
						Max. STO (112.5-250 m^3/d)	123.5	115.25	177.8125	n/a
Fast-SAGD										
SAGD wells	Max. BHP (3375-5625 kPa)	4995	n/a	n/a	4725	Min. BHP (1875-3125 kPa)	2318.75	n/a	n/a	2868.75
						Max. Steam Rate (1-6.25 m^3/d)	4.57	n/a	n/a	2.155

						Max. STO (112.5-250 m^3/d)	154.6875	n/a	n/a	127.5
CSS Wells	Max. BHP (8625-12000 kPa)	n/a	11831.25	11814.375	n/a	Min. BHP (1875-3125 kPa)	n/a	2362.5	2518.75	n/a
	Injection time (35-270 days)	n/a	176	214.775	n/a	Max. Steam Rate (1-6.25 m^3/d)	n/a	3.4675	2.18125	n/a
	Soak time (6-14 days)	n/a	10.48	8.4	n/a	Production rate (90-300 m^3/d)	n/a	161.4	119.4	n/a
						Production time (*day)	n/a	178.52	141.825	n/a

*Production time was based on formula, that is, Production time = 365 – (injection time +soak time)

0 Table 4.1: Constraints used for all optimized cases (wells are numbered from left to right.)

The cumulative steam-oil ratio (CSOR), oil recovery factor (RF), and cumulative oil produced are key parameters used in evaluating the economic performance of a steam or steam/solvent injection process. Loosely speaking, a lower CSOR and higher RF and cumulative oil produced generally translates to a better economic outcome. Figures 9 to 11 compare the CSOR, RF, and cumulative oil produced of the base cases with the CMOST optimized cases. The figures are, respectively, for SAGD, ES-SAGD (i.e., 2% molar concentration of n-C_5 and 98% molar concentration of water), and Fast-SAGD. These figures consistently showed an improvement in all the parameters used in assessing the economic performance of these recovery processes except for Fast-SAGD which had a slightly higher CSOR for the optimized case but significantly more oil produced (1030178 m^3 compared to 986153 m^3.) For the SAGD process, RF increased from 71.6% to 72.8%, CSOR reduces from 3.7 to 3.4 and cumulative oil produced increases from 967679 m^3 to 984200 m^3. The same trend is generally seen for both Fast-SAGD and 2% molar concentration of n-C_5. Note that the ES-SAGD cumulative oil produced and RF reported in this work are all based on net oil produced. That is, the cumulative oil produced, for instance, is calculated based on total oil produced less the combined n-C_5 produced. A typical ES-SAGD process recovers about 70 - 90 % of the solvent injected. Figure 12 shows the amount of solvent injected and produced for the 3.76% molar concentration of n-C_5. In this case, the ultimate solvent recovered was about 84.5% of the solvent injected.

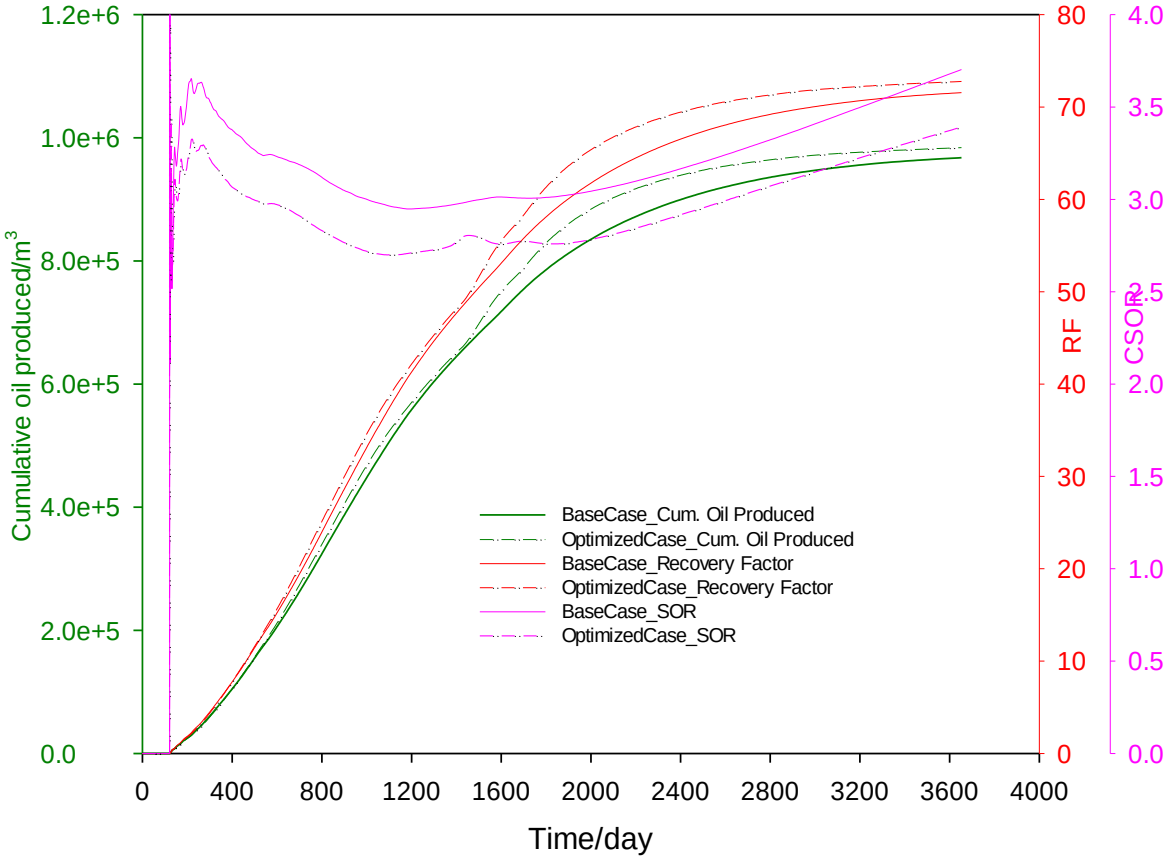

Figure 9: Comparison of the cumulative oil produced from SAGD base case versus its optimized case

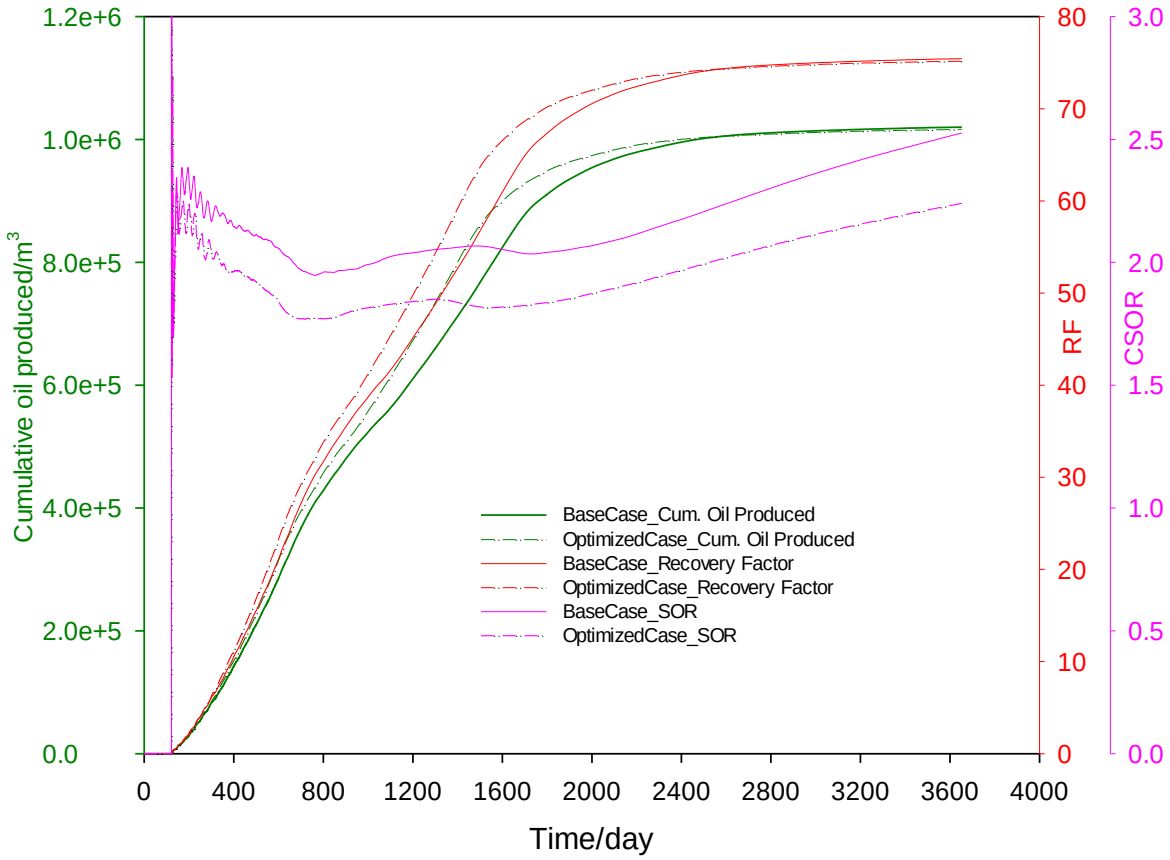

Figure 10: Comparison of the cumulative oil produced from 2% molar *n*-C_5 base case versus its optimized case

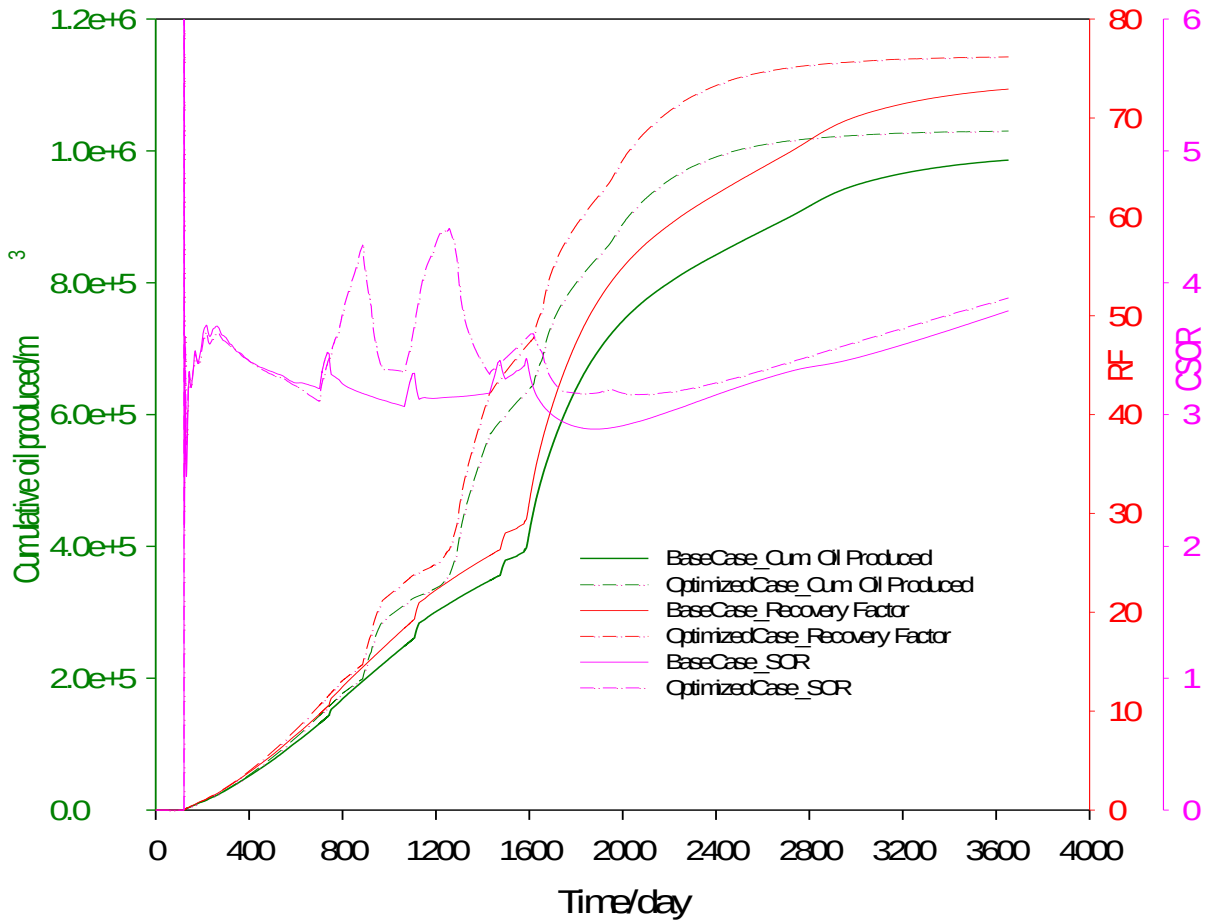

Figure 11: Comparison of the cumulative oil produced from Fast-SAGD base case versus its optimized case

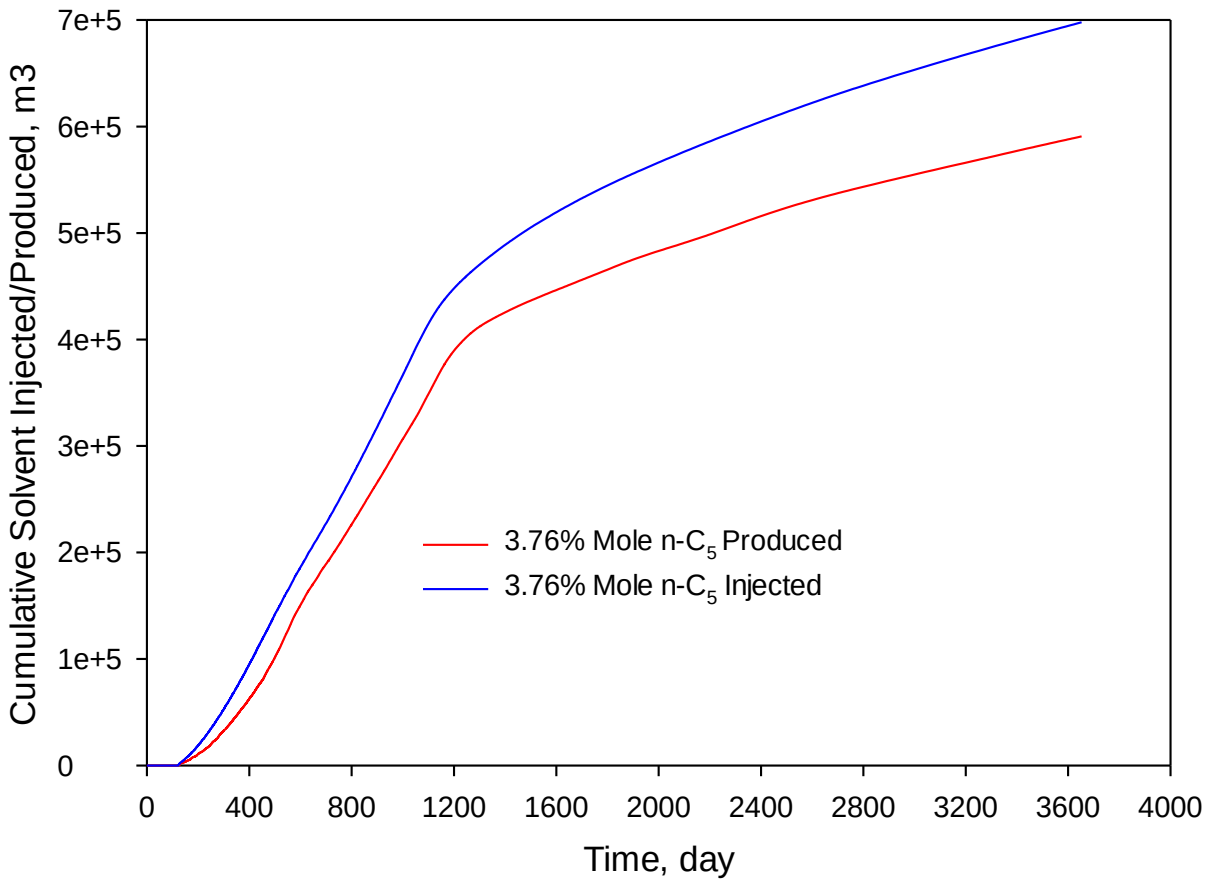

Figure 12: A typical profile showing cumulative solvent injected/produced for the 3.76% *n-C₅* process.

4.3. Comparing the Optimized Cases for all Three Recovery Processes

The following results presentations are all based on the optimized cases for each recovery process. Figure 13 compares the cumulative oil produced and CSOR of SAGD, Fast-SAGD, and ES-SAGD (2% molar concentration of n-C_5). It shows that at the early stages of production, the 2% molar ES-SAGD process produces a higher drainage rate compared to the other two processes. In fact, for the 2% molar concentration of n-C_5 there is a 44% increase in cumulative oil produced compared to the SAGD counterpart and about 183% higher than the Fast-SAGD at about the 365 days into the production period. Meanwhile, Fast-SAGD initially had the lowest cumulative oil produced as it employs fewer wells during this period. Once the CSS wells were opened, its drainage rate greatly improved but it took almost 7 years before it produced more oil than the two other processes.

The figure on the other hand shows that the 2% molar concentration of solvent has the lowest CSOR followed by SAGD with Fast-SAGD reaching almost 3.9 m^3/m^3.

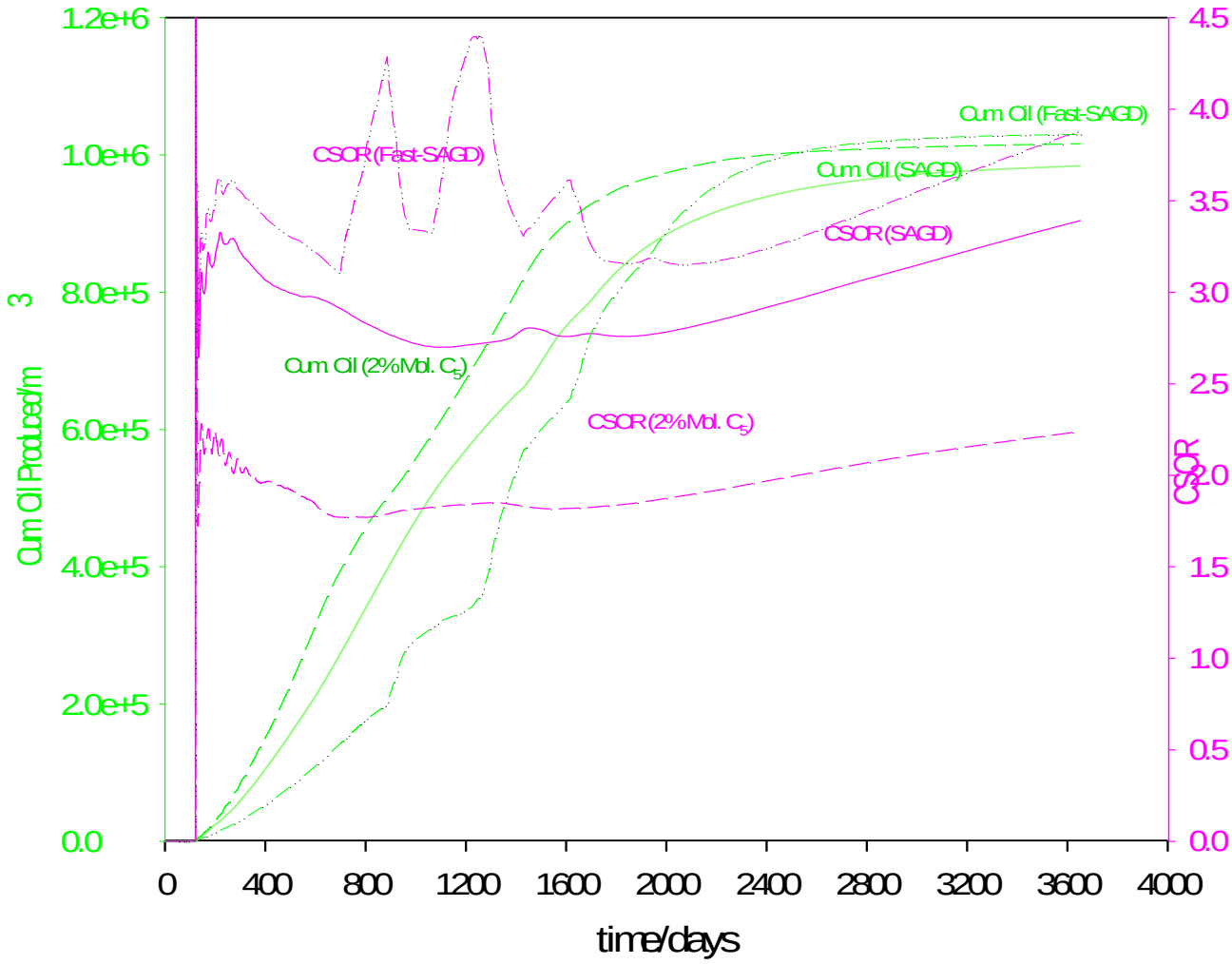

Figure 13: Comparison of SAGD, Fast-SAGD, and 2% molar *n*-C_5 CSOR and cumulative oil produced

Figure 14 compares the cumulative oil produced from the ES-SAGD process as the molar composition of the solvent changes while Figs. 15 and 16 show the corresponding RF and CSOR, respectively, of the different composition. What these figures clearly reveal is that during the early production period, there is a significant jump in the drainage rate when the solvent composition increases from 2% mole to 3.76% mole. In fact, about 77% more oil/year was recovered at about the 3rd year (about 1095 days)

when 3.76% mole solvent is used compared to the 2% mole counterpart. However, there is only a mild increase when the mixing ratio is increased from 3.76% to 4% and furthermore to 5.9%. The 3rd year volume of oil produced from 3.76% mole solvent increased by about 2% when solvent molar concentration is about 4% but slightly increases to about 8% when the mixing ratio increases to 5.9%. The RF follows the same trend as the cumulative oil produced. There is, however, a corresponding reduction in CSOR as the mixing ratio increases. It reduces to 1.43 m^3/m^3 for the 5.9% molar ratio from 2.24 m^3/m^3 for the 2% molar ratio.

Several deductions can be made from these figures. First, it seems that when pentane is used as solvent, the optimal molar ratio for n-C_5 needs to be more than 2%. In fact, based on this work between 3.76% and 5.9% is the ideal range. Second, it seems there is no great economic benefit when the molar ratio changes from 3.76% to 5.9%. However, the molar ratio that is economically more beneficial can only be deduced from an economic analysis, which will be carried out later in this study. What is also shown by these figures is that right around 1200 days into the production period, the cumulative oil produced remains almost flat especially for 3.76% to 5.9% molar concentration of solvent. This is because most of the bitumen had been assessed by these processes compared to the 2% molar mixture.

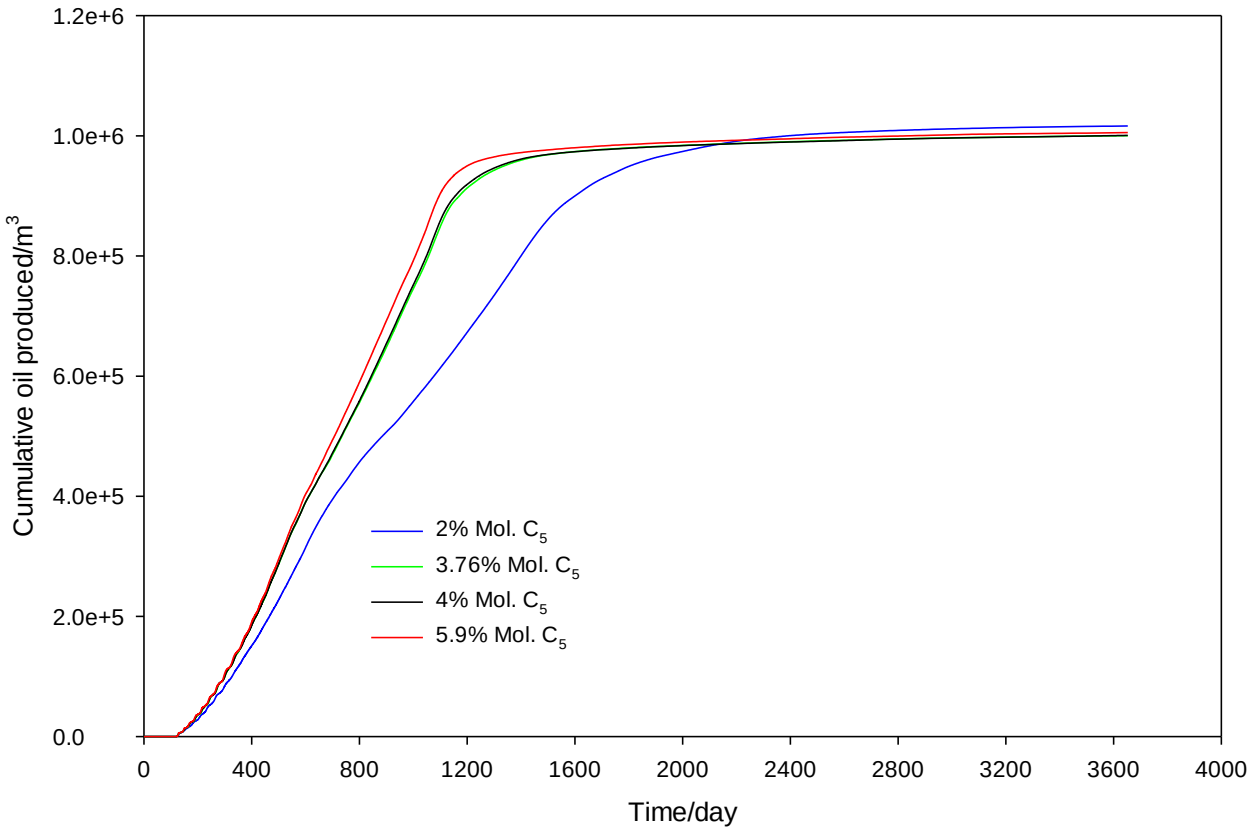

Figure 14: Comparison of the cumulative oil produced for different molar ratios of *n*-C_5

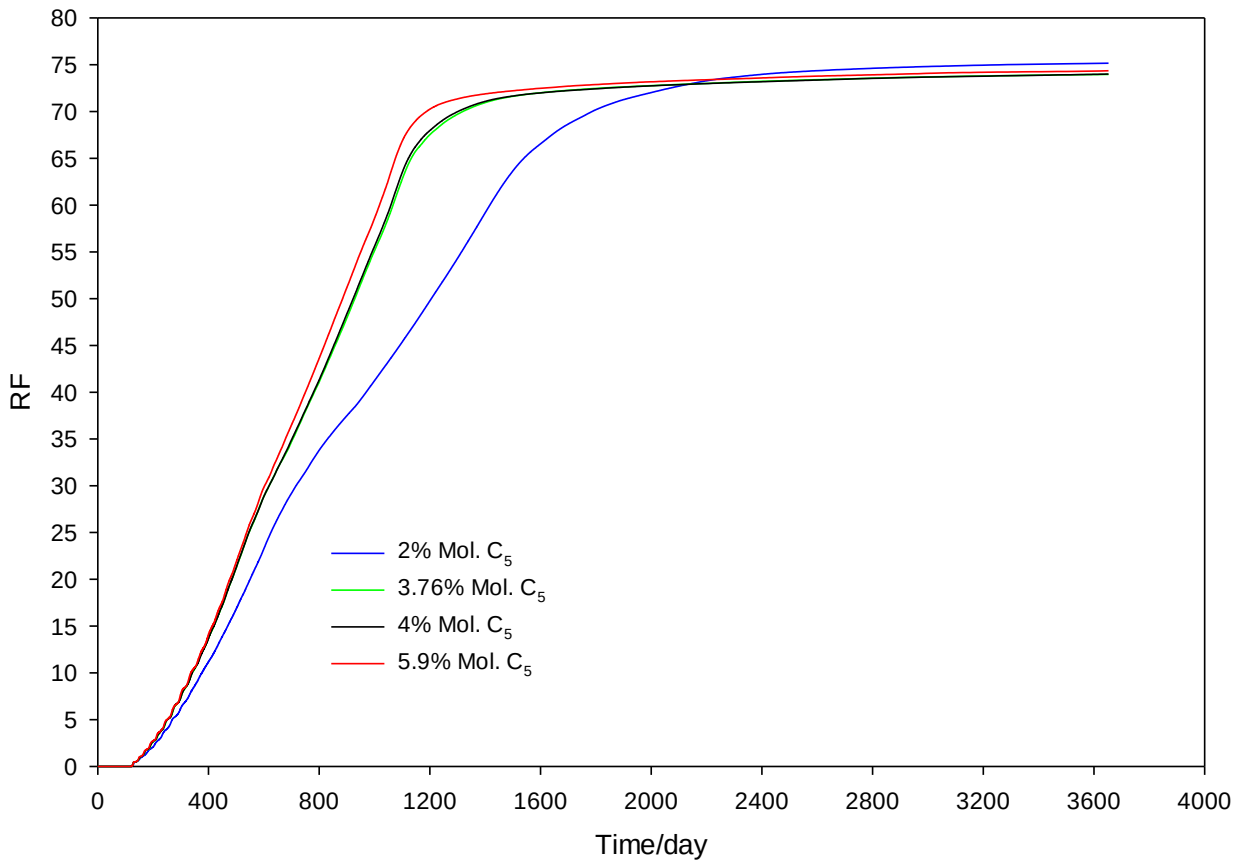

Figure 15: Comparison of the RF for different molar ratios of *n*-C_5

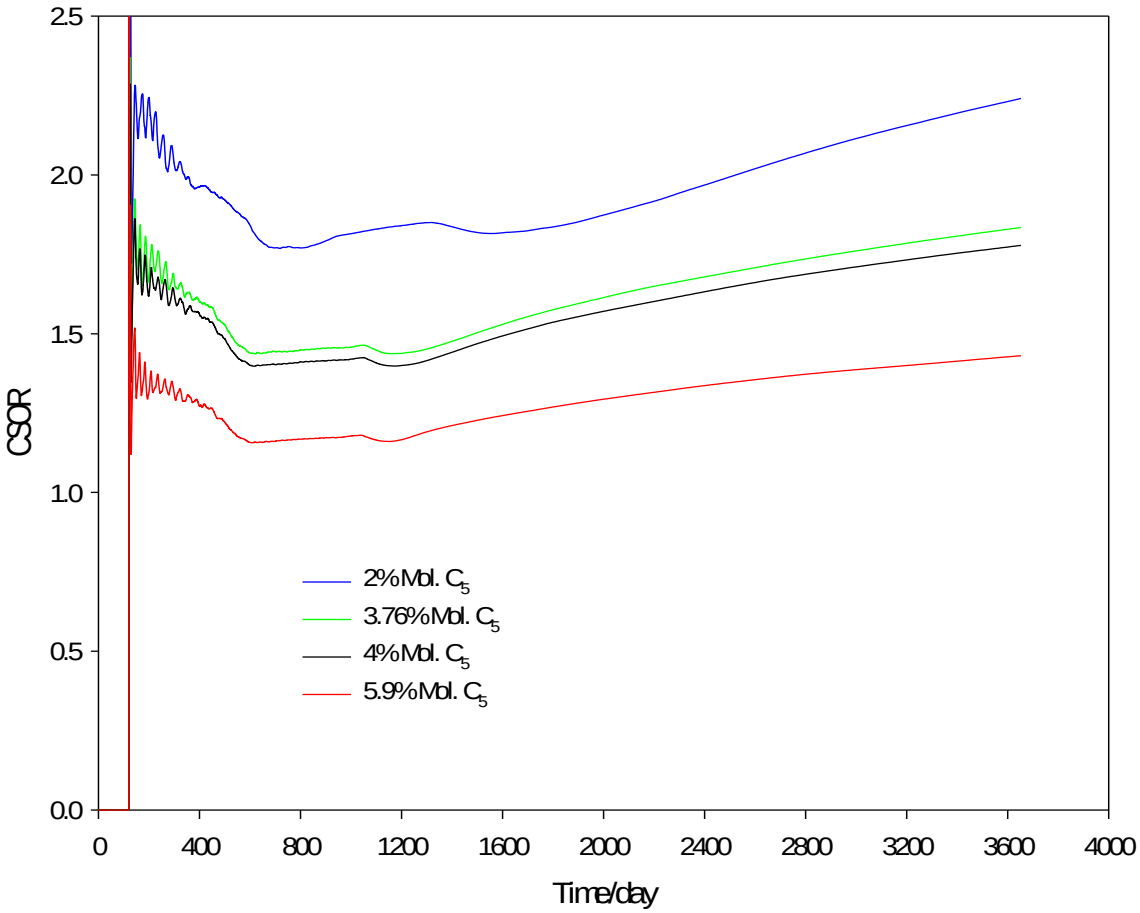

Figure 16: Comparison of the cumulative Steam Oil Ratios for different molar ratios of *n*-C_5

Figure 17 compares the daily oil rate of SAGD, Fast-SAGD, and 2% molar *n*-C_5. This figure shows clearly that during the 10-year production period, three sub-production stages can be identified. The period of the stages is also dependent on the thermal process. For SAGD, the first stage (Stage 1) is from 0 to about 721 days while it is faster for the 2% molar concentration of *n*-C_5. In stage 1, the oil production rate increases up to a maximum value until the steam chamber reaches the overburden. However, in Stage 2 (which varies from 721 to 1461 days for the SAGD process), the oil rate initially decreases before reaching a quasi-stable rate until the steam chamber touches the reservoir sides. Stage 3 on the other hand encompasses the period during which the oil rate continuously decreases until the

production period elapses. This stage also corresponds to the period where the oil in the reservoir is severely diminished which explains why the rate is decreasing.

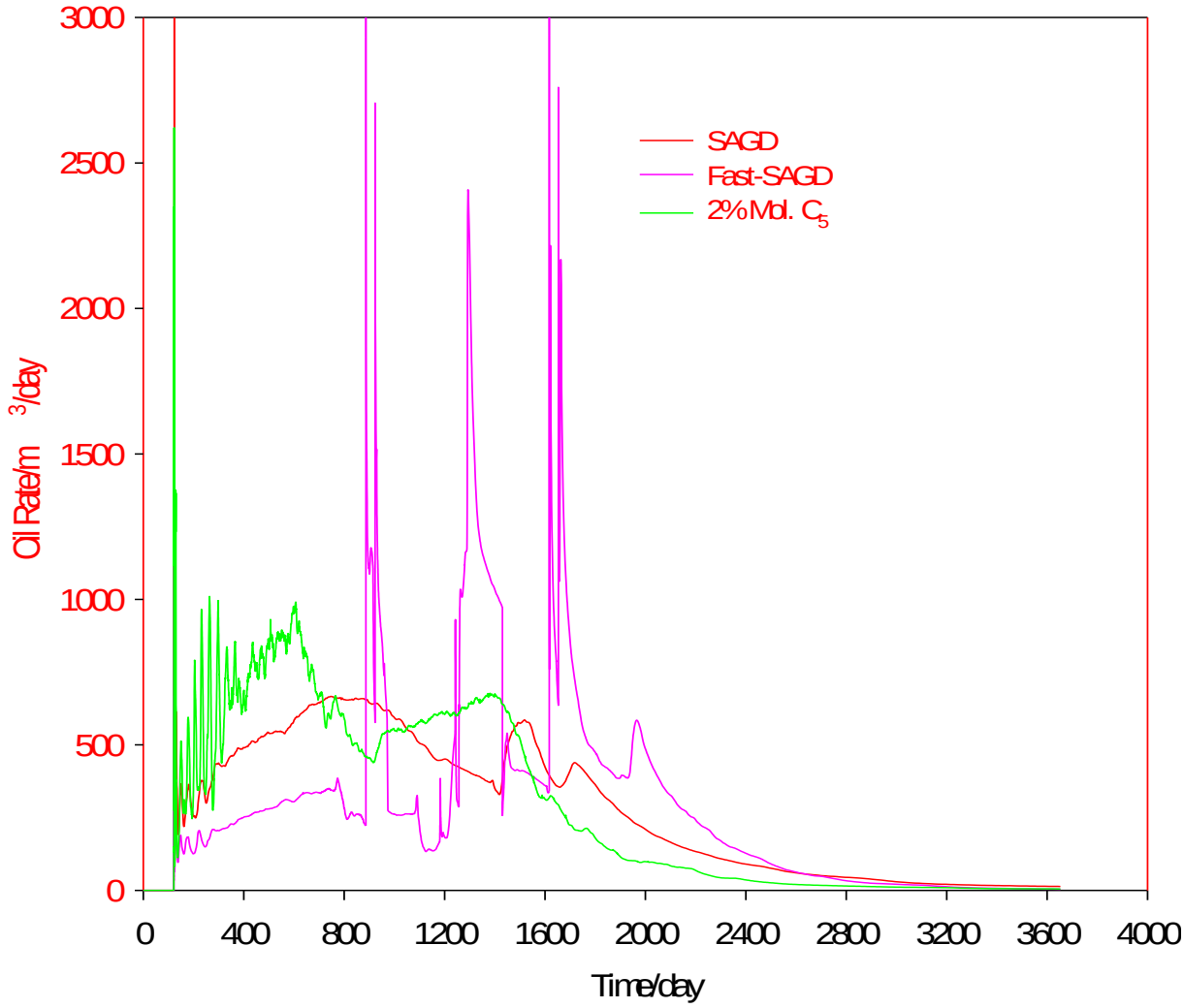

Figure 17: Comparison of SAGD, Fast-SAGD, and 2% molar concentration of *n*-C_5 oil rate

Furthermore, this figure shows that in Stage 1, the 2% molar concentration of *n*-C_5 has the highest daily oil rate while the Fast-SAGD is slightly lower than SAGD due to more wells being used in SAGD during this period. However, Stage 2 sees a more initial sudden drop of the oil rate from 2% molar concentration of *n*-C_5 process while it is somewhat gradual in SAGD as solvent assisted drainage is seen to progress upward at a much rapid pace compared to the SAGD counterpart. However, it is all

over the place for Fast-SAGD since this stage corresponds to the start of the CSS wells, which is why there are noticeable spikes in an oil rate due to the contributions from the CSS wells.

Figure 18 compares the oil rate as a function of molar concentration of solvent. This figure shows clearly that as the molar concentration of solvent increases from 2% to 5.9%, the oil rate increases accordingly. It also shows that the period of Stage 1 becomes narrower as molar concentration of solvent increases. This can also be explained by the fact that as the molar concentration of solvent increases, the rate of upward drainage of the oil accelerates. Consequently, it seems that increasing the solvent concentration has a net positive effect on the oil produced.

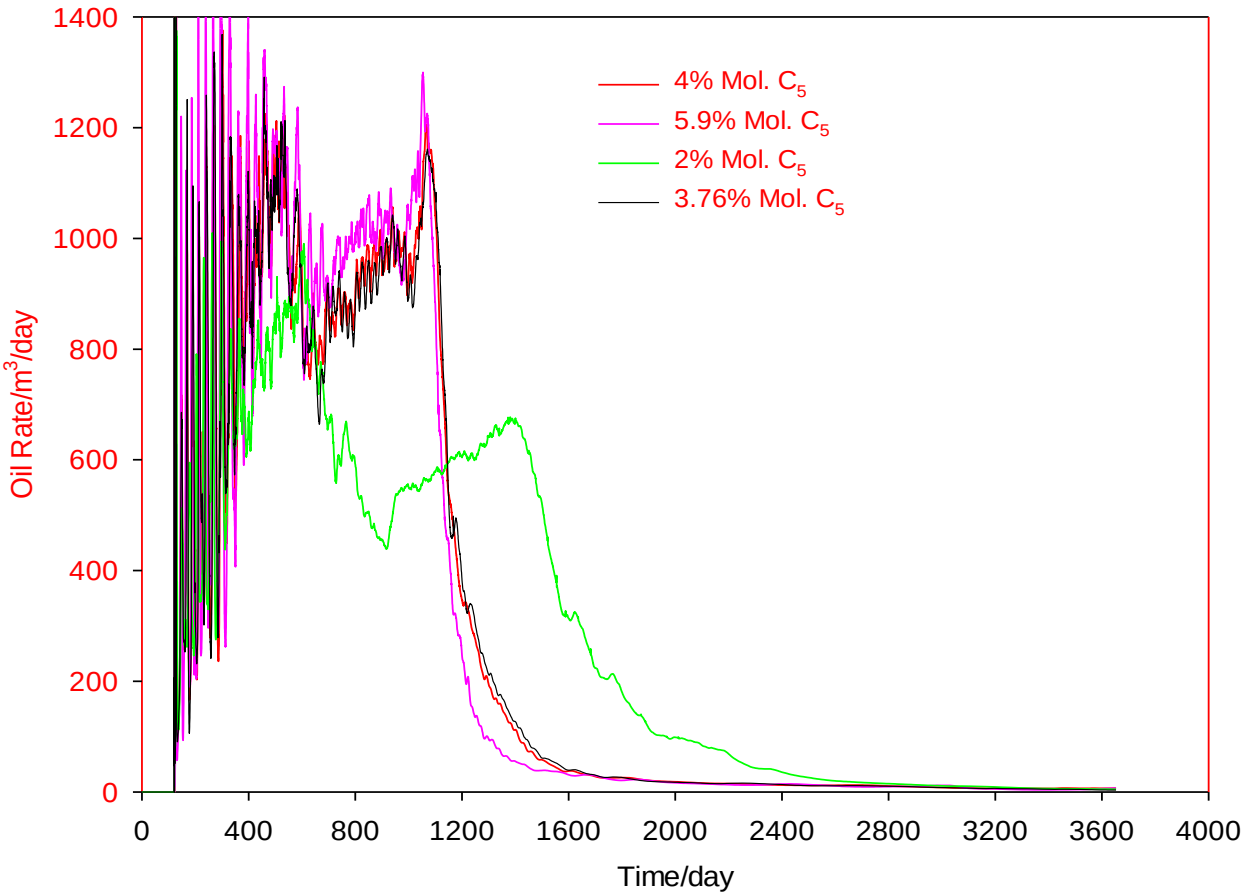

Figure 18: Comparison of the oil rate for different molar ratios of *n*-C_5

4.4. Economic Analysis

As a result of the harsh economic climate brought about by low oil prices, it has become increasingly more important to assess the economic vitality of thermal recovery processes before embarking on one. Furthermore, it has also become critical to evaluate or compare different recovery processes in order to select the most viable option. It is, therefore, under this premise that three thermal recovery processes were studied in this research. However, having raw numbers showing higher cumulative oil produced may not sufficiently indicate that one thermal recovery process is economically more viable than another is. Consequently, one of the most trusted economic indicator used to discriminate one process over another is the net present value (NPV) (Cenovous Energy, 2014). The NPV brings future earnings and expenses into the present by considering the "time value of money."

There have been several economic analyses of SAGD recovery processes as outlined in the literature review. Nevertheless, there are paucity of information regarding the economic analysis of ES-SAGD and largely no thorough economic analysis of Fast-SAGD process. Furthermore, even the many economic analyses of SAGD that are available in the open literature have been found to be outdated due to how dated their analyses have been. Note that in the last three years there has been a significant drop in oil prices, which have also affected associated costs involved in analysing the economic performance of SAGD projects. Consequently, since most SAGD economic analyses have relied on data that have been more than three years, their conclusions may also be misleading or not very reliable for today.

The present economic analysis is mostly based on very recent data and, therefore, more reliable and recent. Most of the data is from the report of Canadian Energy Research Institute (CERI, 2017a report) which utilized very recent costs incurred by operators utilizing the SAGD process for oil sands exploitation.

The NPV in the CERI report (CERI, 2017a) was based on a 10% annual rate of return (ROR) and a 2.0% inflation rate for an effective annual rate of return of 12.0% (nominal). The following supply costs were analysed with an NPV goal of zero:

1. Capital costs which include initial capital costs and sustaining capital costs;
2. Operating working capital costs;
3. Fuel costs;
4. Other operating costs including electricity purchased;
5. Royalties;
6. Income taxes;
7. Emissions compliance costs; and
8. Abandonment costs

Capital costs include the costs of drilling and completion, production pumps, well pads, gathering lines/pipelines, central processing and water treatment facility, and non-process buildings (see CERI, 2017a&b). The details of the other costs can be found in their report.

CERI (2017a) used the following design assumptions in calculating the supply costs (all costs reported in this work are in Canadian dollars):

1. The facility produces 30,000 barrel per day with a production life of 30 years.
2. Initial capital expenditure came to about $39,760 per barrel of capacity, sustaining capital came to an annual average of $43,800,000 and operating working capital was about 45 days payment.
3. Natural gas requirement of 35,910 GJ/day and electricity needed was about 300 MWh/day.

4. Royalties and taxes were calculated differently. The federal and provincial taxes were based on the current rates and held constant for the production life of the facility while the royalties were calculated based on Western Texas Intermediate (WTI) price projections.
5. Carbon tax was based on the current carbon tax policy instituted by the Alberta Government but there was a 2.0% annual inflation rate included throughout the life of the facility.
6. Non-energy operating cost came to about $70,600,000.
7. Abandonment and reclamation was about 2% of the total capital cost.

With their analysis, CERI obtained the following supply costs presented in Table 4.2:

Table 4.2: SAGD supply costs obtained by CERI (2017a)

Cost Categories	**Supply Cost ($/bbl) Based on 10% ROR**
Capital (initial and sustaining)	19.25
Operating working capital	0.40
Fuel (Natural gas)	5.87
Other operating costs (including electricity)	7.53
Royalties	7.14
Income taxes	2.82
Emissions and compliance costs	0.27
Abandonment costs	0.03
Total	**$43.31/bbl**

The $43.31/bbl is the cost of bitumen at the plant gate. In the present research, it was determined to wait for this analysis before completing the comparative economic analysis of SAGD, ES-SAGD, and Fast-SAGD since CERI used more recent data in determining their supply costs. Furthermore, it was

decided to reduce the supply costs by 10% for the current year since the discount rate of 10% will be added to the NPV calculation in the current study.

The following assumptions were made in the current NPV study:

1. The SAGD economic analysis utilized the following supply costs shown in Table 4.3 – 90% of the costs used by CERI (2017a):

Table 4.3: SAGD supply costs used in the current study

Cost Categories	**Supply Cost ($/bbl)**
Capital (initial and sustaining)	17.325
Operating working capital	0.36
Fuel (Natural gas)	5.283
Other operating costs (including electricity)	6.777
Royalties	6.426
Income taxes	2.538
Emissions and compliance costs	0.243
Abandonment costs	0.027
Total	**$38.979/bbl**

5. The SAGD capital cost was regarded as the base cost for both ES-SAGD and Fast-SAGD except some additional capital costs were incurred in ES-SAGD such as solvent distribution line costs. However, Fast-SAGD used the same capital cost for SAGD since the number of wells is equal in both processes.

6. For SAGD analysis, the total cost of initial and sustaining cost was amortized for ten years which corresponds to each year of production. This was the same treatment for ES-SAGD except that solvent distribution line cost was an additional cost. As for Fast-SAGD, the total cost of initial and sustaining cost was also amortized for ten years but treated slightly different. Since only two SAGD wells were initially used for two years with the Fast-SAGD process compared to four SAGD wells in the SAGD counterpart, the fraction of the costs for two wells were used for the Fast-SAGD process for the first two years, after which the remainder of the total cost were amortized equally for the remaining eight years of production.
7. The reference amount of natural gas required in ES-SAGD and Fast-SAGD is the same as that from SAGD. However, the final fuel costs for ES-SAGD and Fast-SAGD were adjusted based on findings from the literature. For ES-SAGD, CERI (2017b) reported that the natural gas requirement of ES-SAGD reduces by about 35% of the amount required by SAGD. Consequently, this was used as a benchmark. Furthermore, for Fast-SAGD, since the CSOR was more than that of SAGD a correction was applied based on the recommendation of Giacchetta et al. (2015). Giacchetta et al. (2015) proposed increasing energy requirement by 0.16 GJ/bbl for every half-point increase in SOR. This same correction was applied to SAGD when the CSOR is above 2.8 though the value used in Giacchetta et al. (2015) was SOR of 2.7.
8. SAGD emissions and compliance costs were used as reference for both the SAGD and Fast-SAGD. In fact, for Fast-SAGD the supply costs for emissions and compliance were identical to those of SAGD. This was presumed to be the case despite the higher CSOR for Fast-SAGD processes. Note that for Fast-SAGD processes, fewer wells were used at the beginning of production for almost 2 years before the CSS wells, which contributed to the higher CSOR, were opened. However, a reduction of about 15% was used in calculating the emission and compliance supply costs for ES-SAGD. This value was obtained from the CERI (2017a) report

that saw a reduction in GHG emissions of between 15-20% for ES-SAGD processes compared to SAGD. The lower value was taken in this analysis.

9. For ES-SAGD, the annual cost of nC_5 for the 10-year production period is calculated from this formula:

$$Total\ annual\ cost\ of\ nC_5 = cost\ of\ 1\ bbl\ of\ nC_5 \times (cumulative\ annual\ nC_5\ injected\ in\ bbl - cumulative\ annual\ nC_5\ produced\ in\ bbl) \quad (4.1)$$

10. The reference cost of pentane was very difficult to get. As a result, two different scenarios were explored. First, it was assumed that using a readily available cost of pentane plus may give a fair representation of pentane being used. Note that the composition of pentane plus is mostly C5+. Therefore, using pentane plus may have the same effect as using pentane as solvent. For this case, 1 m^3 of Pentane Plus for the month of December 2017 (latest month reported) according to Alberta Energy Regulator (AER, 2018) is about $406.74. Second scenario is using available price data from companies that supply pentane in bulk. In this case, the cost of 1 m^3 pentane was found to be $937.50 according to Right Price Chemicals (2018) and Alibaba.com website. These two scenarios will give a good representation of the effect of solvent cost on the economic analysis.

11. Apart from the fuel cost adjustment made for Fast-SAGD as shown in 4 above, all the supply costs treated in SAGD were similar to those in Fast-SAGD.

12. ES-SAGD had other additional costs such as solvent recompression costs and solvent handling costs. The solvent handling costs were about $20,000/year while the solvent recompression costs were about $0.017/Std m^3.

4.4.1. First Method of Economic Analysis

Based on these assumptions, a comparative NPV analysis of the three recovery processes was undertaken. Annual net cash flow (NCF) for SAGD and Fast-SAGD is given by:

$$NCF = R_v - (C_f + C_{ec} + C_{ow} + C_{oo} + C_r + C_t + C_a)$$

(4.2)

where R_v is the annual gross revenue ($/year) from sale of oil, which was pegged at $50/bbl at the plant gate in this study. C_f, C_{ec}, C_{ow}, C_{oo}, C_r, C_t, and C_a are the annual fuel cost ($/year), annual emissions and compliance cost ($/year), annual operating working capital cost ($/year), annual other operating cost (including electricity) ($/year), annual royalty cost ($/year), annual tax cost ($/year), and annual abandonment cost ($/year), respectively.

The annual net cash flow (NCF) for ES-SAGD is given by:

$$NCF = R_v - (C_f + C_{ec} + C_{ow} + C_{oo} + C_r + C_t + C_a + C_s + C_{sh} + C_{sr})$$

(4.3)

where C_s is the annual solvent cost calculated in accordance with assumption 5 above, C_{sh} is the annual solvent handling cost ($/year) based on assumption 7 above, and C_{sr} is the annual solvent recompression cost ($/year) also based on assumption 7 above.

$$NPV = \sum_{n=1}^{10} \left\{ -C_{tc} + \frac{NCF_i}{(1+i)^n} \right\}$$

(4.4)

where i is the discount rate (which is taken to be 10% here), n is the year counter (from 1 to 10), and C_{tc} is the total annual capital cost ($/year) including both initial and sustaining capital costs.

4.4.1.1.Using Cost of Pentane Plus for Economic Analysis

Using the cost regime of pentane plus as the first case scenario, Figures 19 compares the NPV of SAGD, Fast-SAGD, and 2% molar *n*-C_5 processes. This figure clearly shows the NPV of *n*-C_5 (about \$74.7 million in the 5th year) to be the highest amongst the three especially from the start of production to the end of production. At the 2-year mark, the NPV of the *n*-C_5 is almost double that of SAGD. Generally, the Fast-SAGD process recorded the lowest NPV during this period except at the 6th year production period where it caught up to that of SAGD. The Fast-SAGD suffers from this fate because of the deployment of fewer wells at the start of production until two years afterwards before the two additional CSS wells were opened. Consequently, cumulative oil produced was significantly lower than both of the other processes especially at the start of production. It took almost 6 years before Fast-SAGD caught up to the SAGD counterpart. Furthermore, this figure shows that for the SAGD and 2% molar *n*-C_5 processes, there is a decline in NPV after the 5th year while this decline occurs at the 6th year for the Fast-SAGD process. This is due to the fact that most of the oil was produced within the 5-year window after which very little oil was left to be produced. Consequently, the cost of producing a barrel of oil significantly increases for all the recovery processes after this period, with the *n*-C_5 process significantly higher.

There is a critical deduction to be made from this figure. It seems that it is better to switch from using solvent/steam injection and steam only to probably other means of injection after the 5-year mark as most of the oil has been produced at this time by the solvent/steam and steam processes. This explains why there is a significant drop in cumulative NPV after the 5-year mark as the amount spent buying solvent and producing steam far outweighs its benefit.

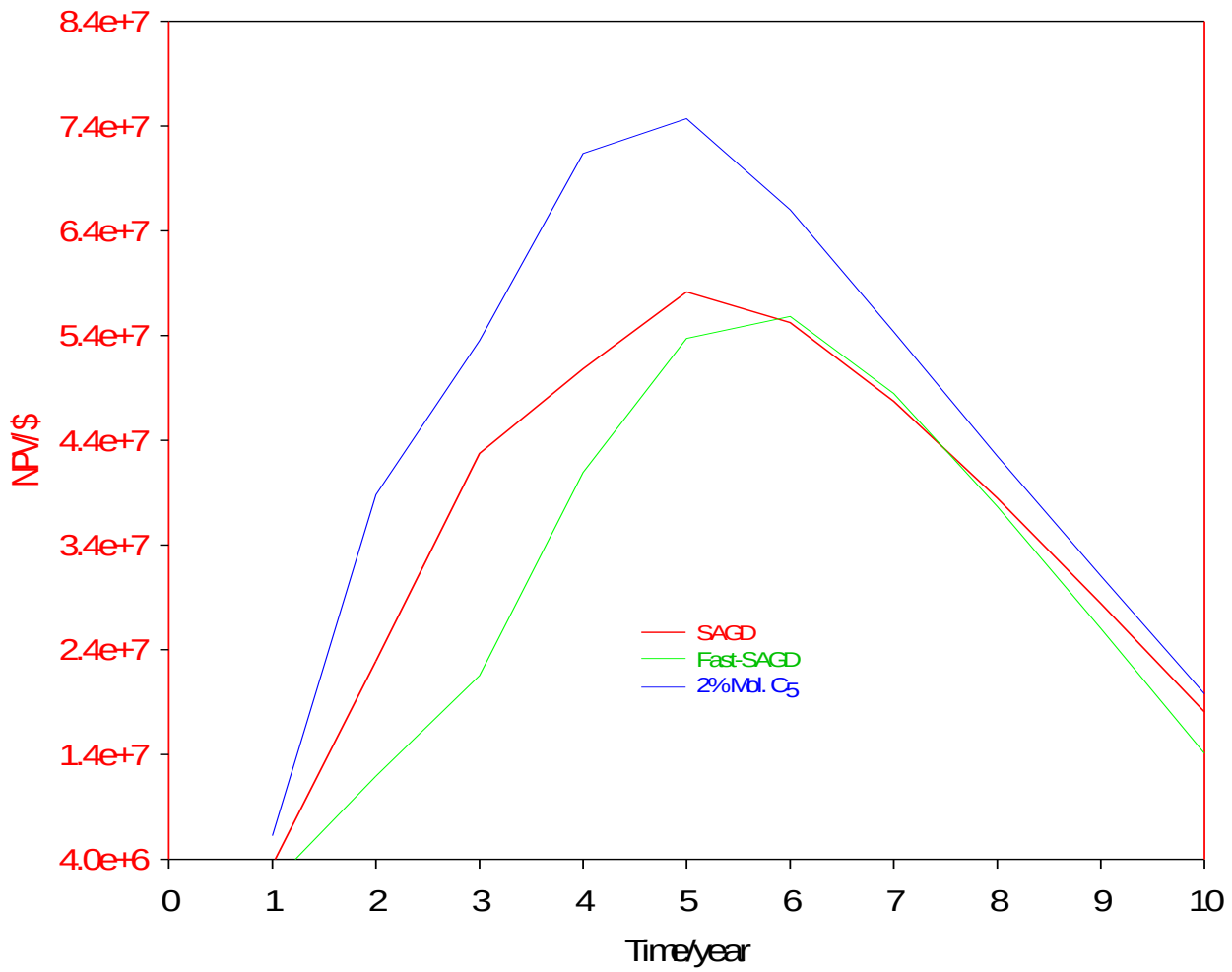

Figure 19: Comparisons of the NPV for SAGD, Fast-SAGD, and 2% molar concentrations of *n*-C_5 using pentane plus price as reference price for solvent

Still using the cost regime of pentane plus as the first case scenario, Figure 20 compares the NPV of the various molar concentrations of *n*-C_5 processes. This figure shows a corresponding increase in the NPV as the molar concentration of *n*-C_5 increases during the 4-year window or the 3-year window for the higher molar *n*-C_5 process. This is a result of a very significant jump in oil produced during this period,

as the oil rate is much higher for the higher molar concentration of *n*-C_5. Nevertheless, after this period there is again a precipitous drop in NPV, which is more prominent as the molar concentration of *n*-C_5 increases. There is also no noticeable improvement in NPV as the molar concentration of *n*-C_5 increase from 3.76% to 4% and subsequently to 5%. In fact, it could be argued the NPV rather reduced showing a diminishing return. All these are a reflection of the costs involved in producing a barrel of oil increasing without any corresponding increase in oil produced since most of the oil had been produced during the 3-year window.

The foregoing is an indication of the importance of solvent extraction. That is, with solvent there is a significant increase in oil produced especially at the start of production until the OIP is mostly depleted. The above NPV numbers show that it is more economical to use the solvent-steam extraction method in the early stages of a thermal process but there needs to be a migration to a much cheaper cost of production or proceed to well abandonment earlier since most of the OIP would have been recovered using a solvent-steam mixture.

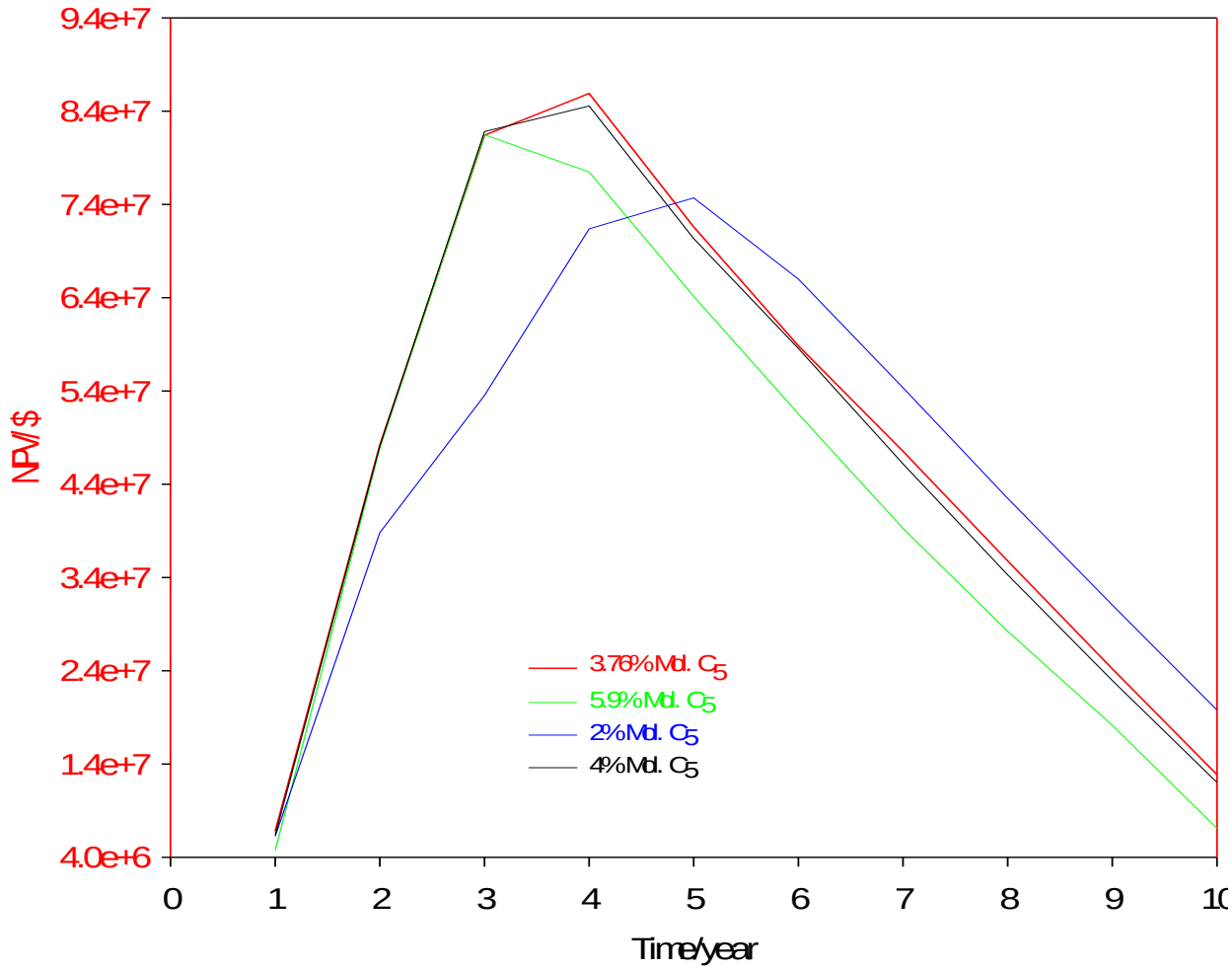

Figure 20: Comparison of the NPV for different molar ratios of *n*-C_5 using pentane plus price as reference price for solvent

4.4.1.2.Using cost of Pentane ($937.5 for 1 m^3) obtained from Companies

Figure 21 shows that as the cost of 1 m^3 of solvent increases from $406.74 to $937.5 (compare Figure 19 to Figure 21,) there is no noticeable difference in NPV between SAGD and 2% molar concentration of *n*-C_5. They are almost similar until about the 5th year after which the decline in NPV for the 2% molar concentration of *n*-C_5 is more drastic indicating that the cost of producing a barrel of oil has considerably increased. This figure also shows that the NPV of Fast-SAGD is the lowest especially at the start of production till about the 5th year when it caught up with 2% molar concentration of *n*-C_5 and 6th year for SAGD.

This figure gives a clear indication of the limitations of ES-SAGD. That is, at this molar concentration and based on this cost regime, SAGD is equally as good as 2% molar concentration of pentane.

Figure 22 compares the NPV of the various molar concentrations of n-C_5 processes based on this cost regime. This figure clearly shows no positive correlation between NPV and molar concentration. In fact, it shows that the lowest NPV is that of the 5.9% molar concentration of n-C_5. The highest NPV belongs to the 3.76% molar concentration especially after the first year of production.

The most critical deduction to make from this figure is that despite the considerably high price of pentane in this analysis, it is still more beneficial to use solvent compared to SAGD (compare Figure 21 to Figure 20). That is, solvent molar concentration between 3.76% and 4%. However, since most of the OIP has been depleted by the 4th year of production, it is highly recommended to switch to other means of production since there is a steep decline in NPV after the 4th year especially for the 3.76% and 4% molar concentrations of n-C_5.

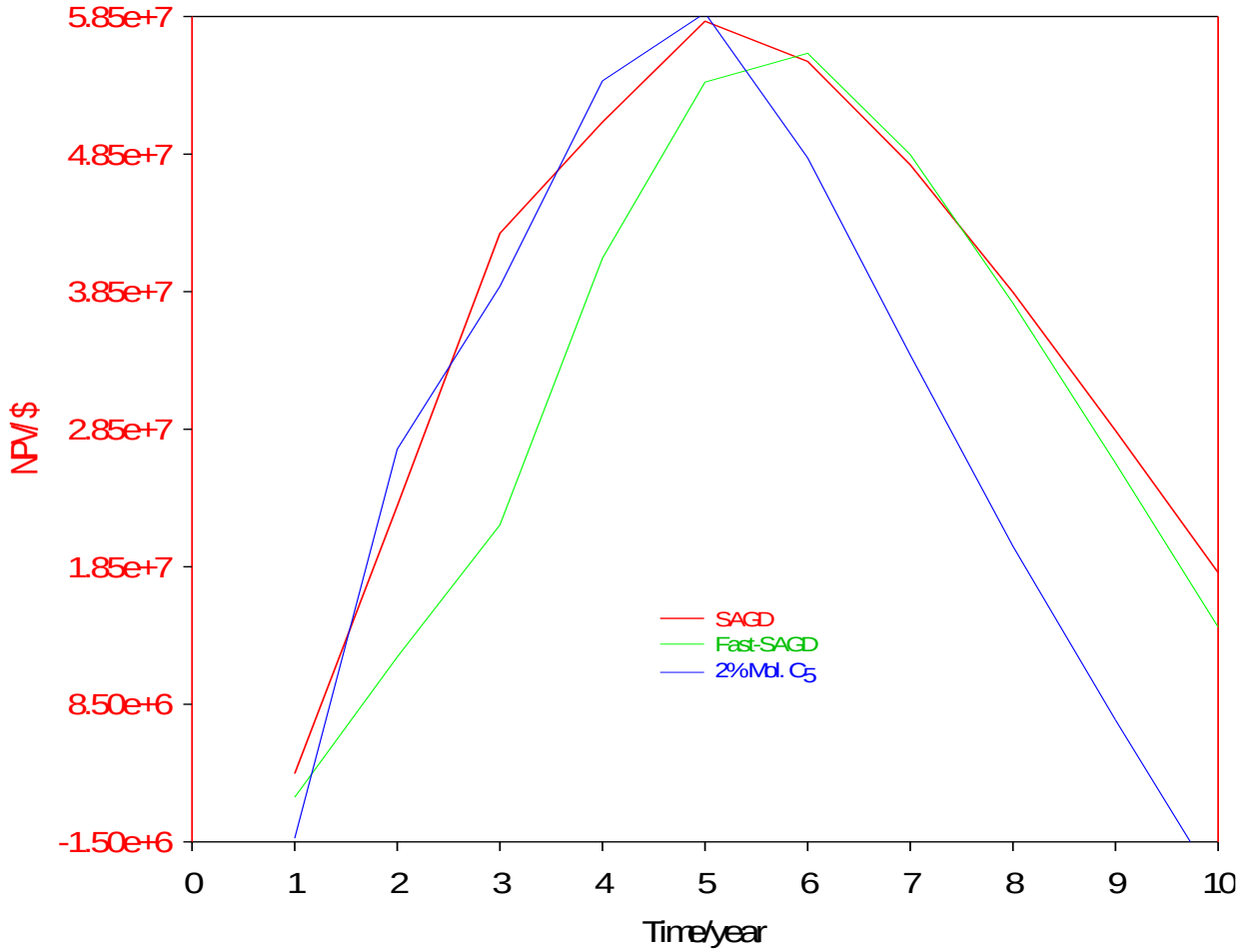

Figure 21: Comparisons of the NPV for SAGD, Fast-SAGD, and 2% molar concentrations of *n*-C_5 using pentane price obtained from company as reference price for solvent

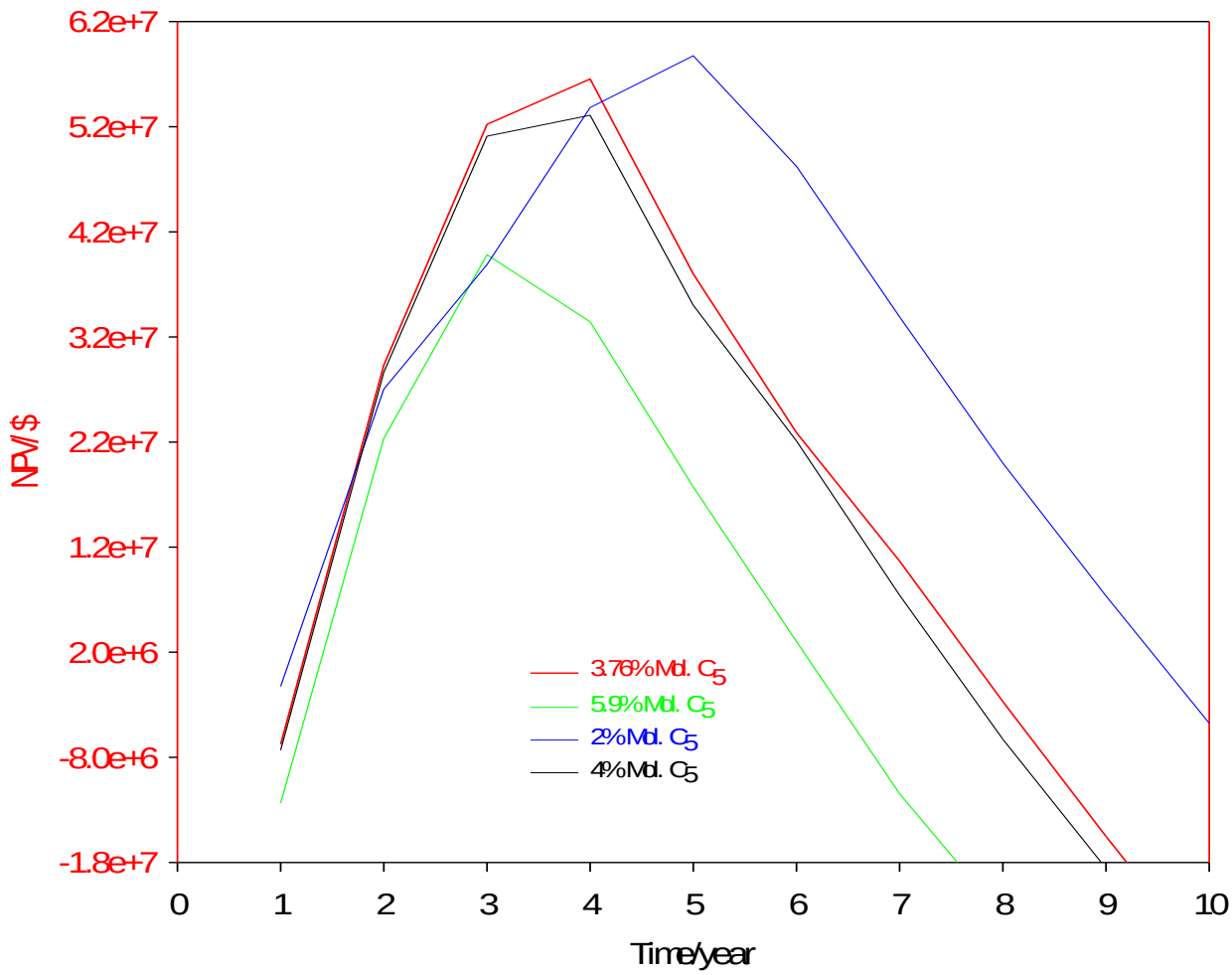

Figure 22: Comparison of the NPV for different molar ratios of *n*-C_5 using pentane price obtained from company as reference price for solvent

4.4.2. Giacchetta Method of Economic Analysis

This method has traditionally been employed for simple SAGD or ES-SAGD projects that mostly involve a single well pair and with a facility that has a production facility much less than the facility treated by CERI above. Nevertheless, this method has also been used for large-scale facility such as that employed by Giacchetta et al. (2015). In the following sections, I will outline the work of Giacchetta et al. (2015) on which this current economic analysis is based. This is a discounted cash flow (DCF) analysis with the following equations:

$$C_t = C_{tf} + C_{OM} \tag{4.5}$$

where C_t is the total annual cost (\$/year), C_{tf} is the total annual fuel cost (\$/year), and C_{OM} is the total annual operation and management cost (\$/year).

$$OI = R_t - C_t \tag{4.6}$$

where OI is the operating income and R_t is the total annual revenue generated (\$/year), which is simply the revenue generated from selling oil produced annually:

$$A_k = C_c / P_{amm} \tag{4.7}$$

where A_k is the amortization, C_c is the total capital and sustaining cost, and P_{amm} is the amortization period. This work assumes an amortization period of 10 years which is the duration of operation of the wells.

$$TO = OI - A_k \tag{4.8}$$

where TO is the total operating income.

$$Tax = TO \times tax\ rate \tag{4.9}$$

$$NI = OI - Tax \tag{4.10}$$

where NI is the net income. The discounted cash flow is

$$DCF_k = \frac{NI_k}{(1+i)^k} \tag{4.11}$$

where i represents the discount rate.

$$NPV = -C_c + \sum_{k=1}^{n} \frac{NI_k}{(1+i)^n} \tag{4.12}$$

where k varies from 1 to 10 years.

4.4.2.1.Gianchetta Method Based on Pentane Plus Price

Using the analysis above and based on the cost of 1 m^3 of Pentane Plus at $406.74, Figure 23 compares the NPV results of SAGD, Fast-SAGD, and 2% molar $n\text{-}C_5$. This figure clearly shows that the breakeven is much faster for the 2% molar concentration of $n\text{-}C_5$ compared to the other two processes. The breakeven point was reached before the 2nd year of production for the 2% molar concentration of $n\text{-}C_5$ while it took SAGD process about 3 years to reach the breakeven point and 4 years for the Fast-SAGD counterpart. The reason it took Fast-SAGD such a long time to reach breakeven point is because fewer oil producing wells were initially deployed in Fast-SAGD. The figure also reveals the NPV of 2% molar concentration of $n\text{-}C_5$ to be the highest of the three processes throughout the production period while the Fast-SAGD was the lowest.

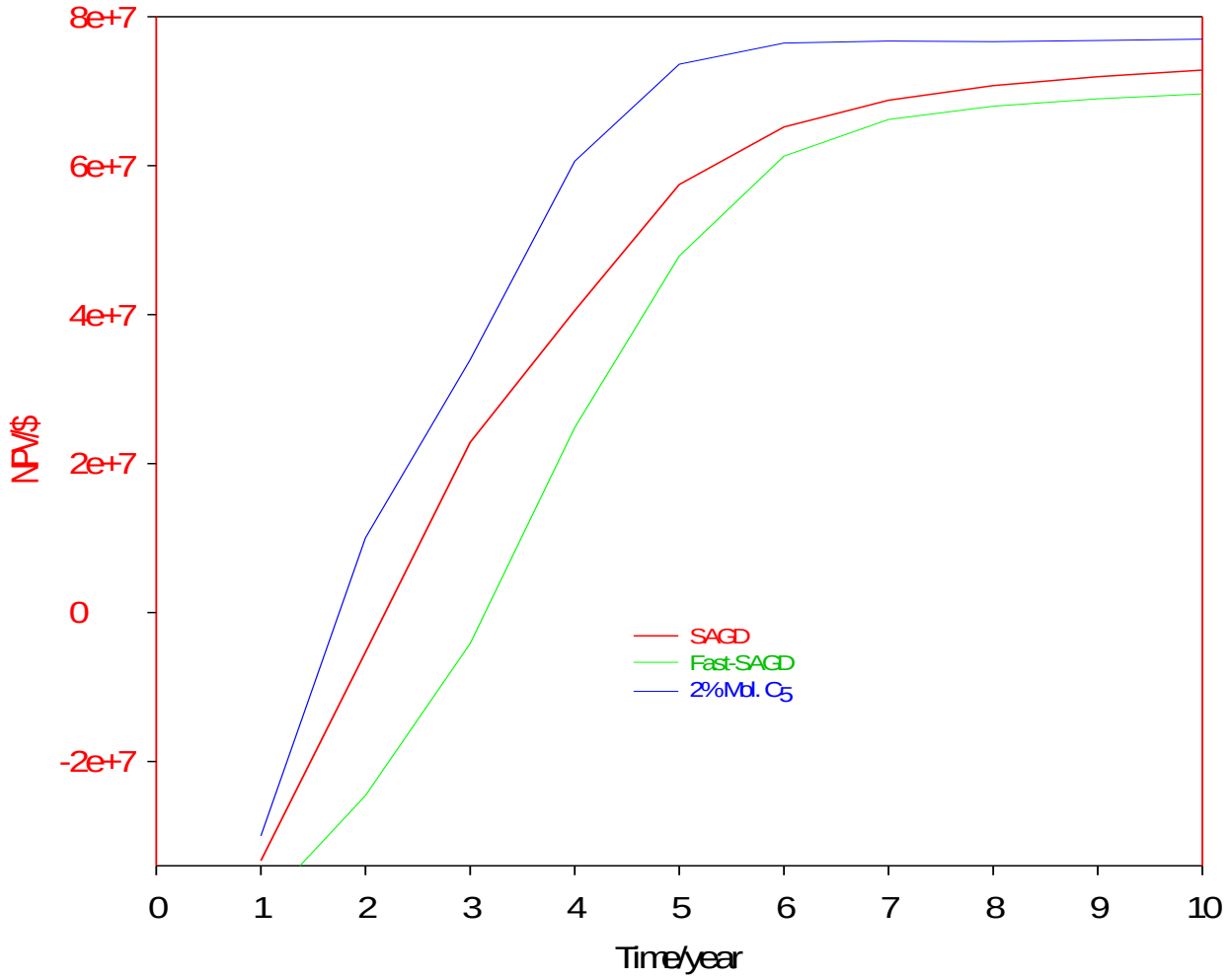

Figure 23: Comparisons of the NPV for SAGD, Fast-SAGD, and 2% molar *n*-C_5 using Giacchetta et al. (2015) method and pentane plus price as reference price for solvent

Figure 24 on the other hand compares the NPV of various molar concentration of *n*-C_5. It shows a significant increase in NPV as the molar concentration increases from 2% to 3.76%. However, no further noteworthy improvement is seen as the molar concentration increases from 3.76% to 4% and finally to 5% - only minor improvement. It also shows that the higher molar concentration of *n*-C_5 reached the breakeven point faster than the 2% molar concentration of *n*-C_5 Furthermore, the figure shows that after 4 years, a peak NPV was achieved for molar concentration of 3.76% to 5.9% *n*-C_5. In

fact, the NPV of 5.9% molar concentration of *n*-C_5, which was initially the highest of them all, became lower than the 3.76% and 4% molar concentration of *n*-C_5 after the 3-year period while the NPV of the 2% molar concentration of *n*-C_5 became the highest at about the 5-year period.

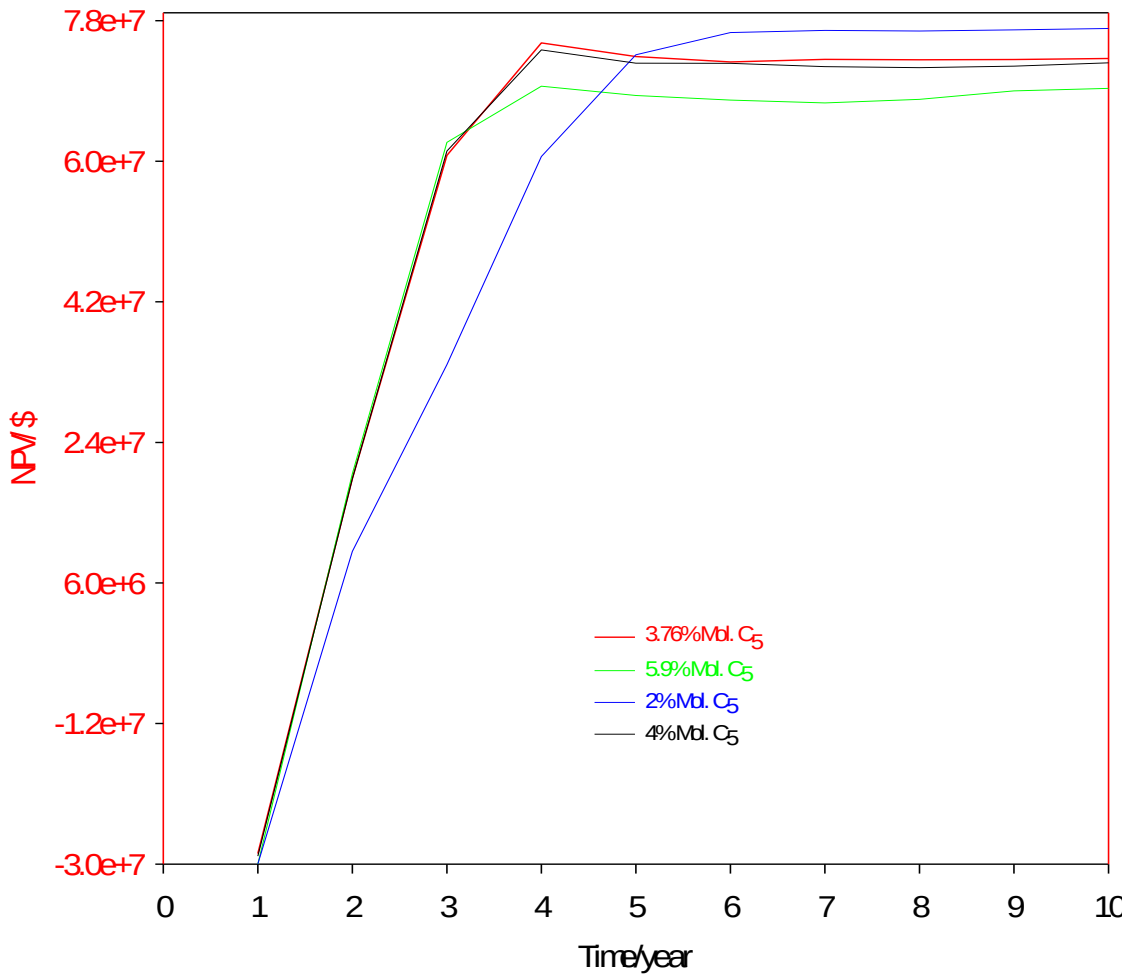

Figure 24: Comparison of the NPV for different molar ratios of C_5 using Giacchetta et al. (2015) method and pentane plus price as reference price for solvent

4.4.2.2.Gianchetta Method Based on Pentane Price obtained from Companies

Using the analysis above and based on the cost of 1 m^3 of Pentane at $937.5, Figure 25 compares the NPV results of SAGD, Fast-SAGD, and 2% molar *n*-C_5. The figure clearly shows that it took the Fast-SAGD process about 4 years to reach the breakeven point while it took SAGD less than 3 years and 2%

molar concentration of n-C_5 less than 2 years. It shows that the 2% molar concentration of n-C_5 has the highest NPV from the 2nd year to about the 6th year after which the SAGD and the Fast-SAGD caught up and overtook it. This figure however, also shows that the Fast-SAGD is significantly less than that the two other processes and therefore not as viable. Consequently, it appears the 2% molar concentration of n-C_5 is the most viable process especially at the early stages of production until about the 6th year. Thereafter, switching to a more viable recovery process will be advisable.

Figure 26 on the other hand compares the NPV of various molar concentration of n-C_5. This figure again shows no positive correlation between NPV and molar concentration. In fact, it shows that the lowest NPV for the most part is that of the 5.9% molar concentration of n-C_5. The highest NPV belongs to the 3.76% molar concentration especially after the first year of production until the 4th year. It also shows that the 2% molar concentration of n-C_5 overtakes the NPV of the 3.76% and 4% molar concentration of n-C_5 after the 4th year. Based on these observations, it can be stated that as the cost of pentane becomes prohibitive or at least 4 times the cost of oil, the advantages of solvent/steam injection greatly diminishes and utilizing higher concentrations of n-C_5 may have an adverse effect as shown in this figure. Nevertheless, comparing this figure to Figure 25 shows that solvent/steam injection is still more economically viable than steam-only injection.

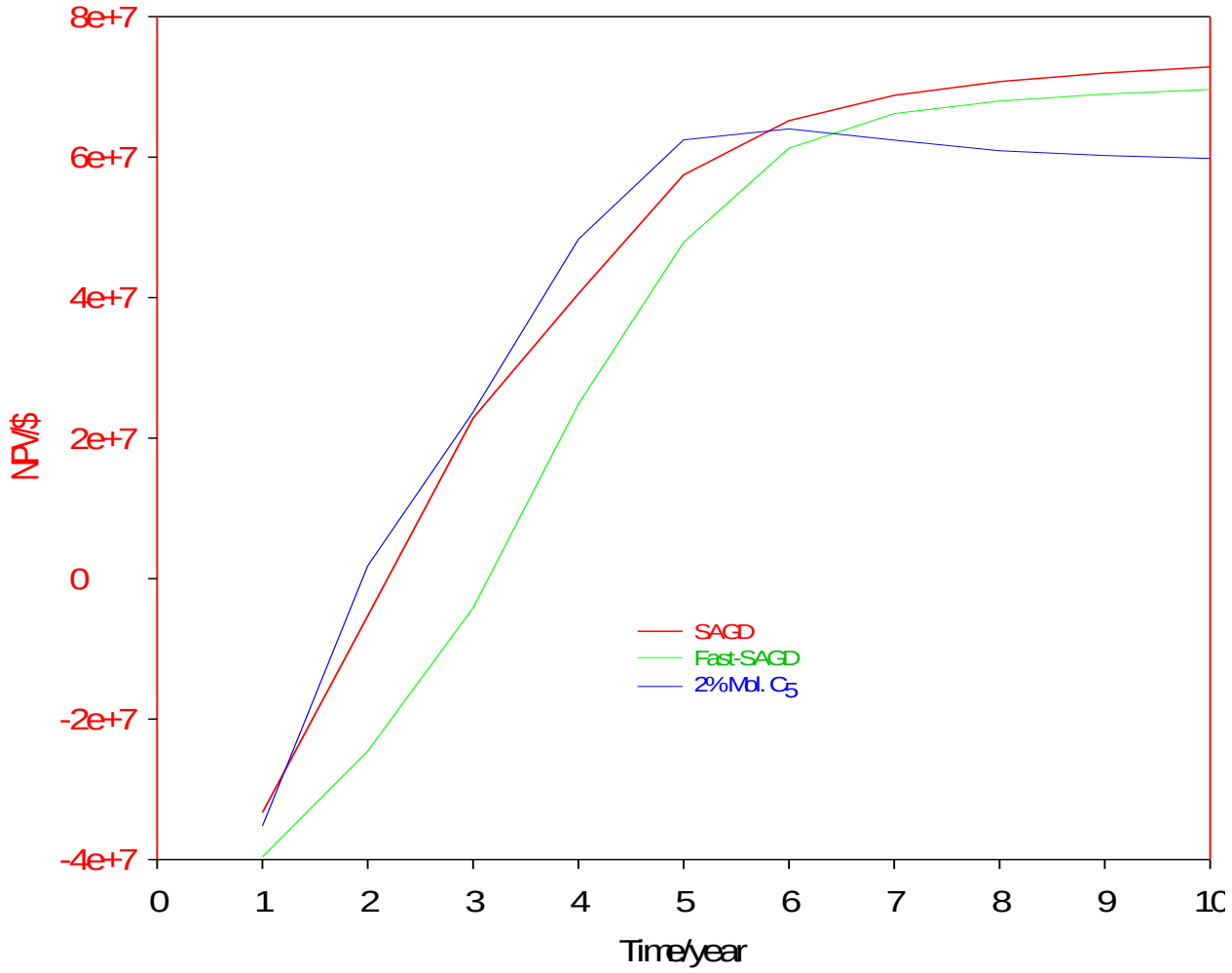

Figure 25: Comparisons of the NPV for SAGD, Fast-SAGD, and 2% molar *n*-C_5 using Giacchetta et al. (2015) method and pentane price obtained from company as reference price for solvent

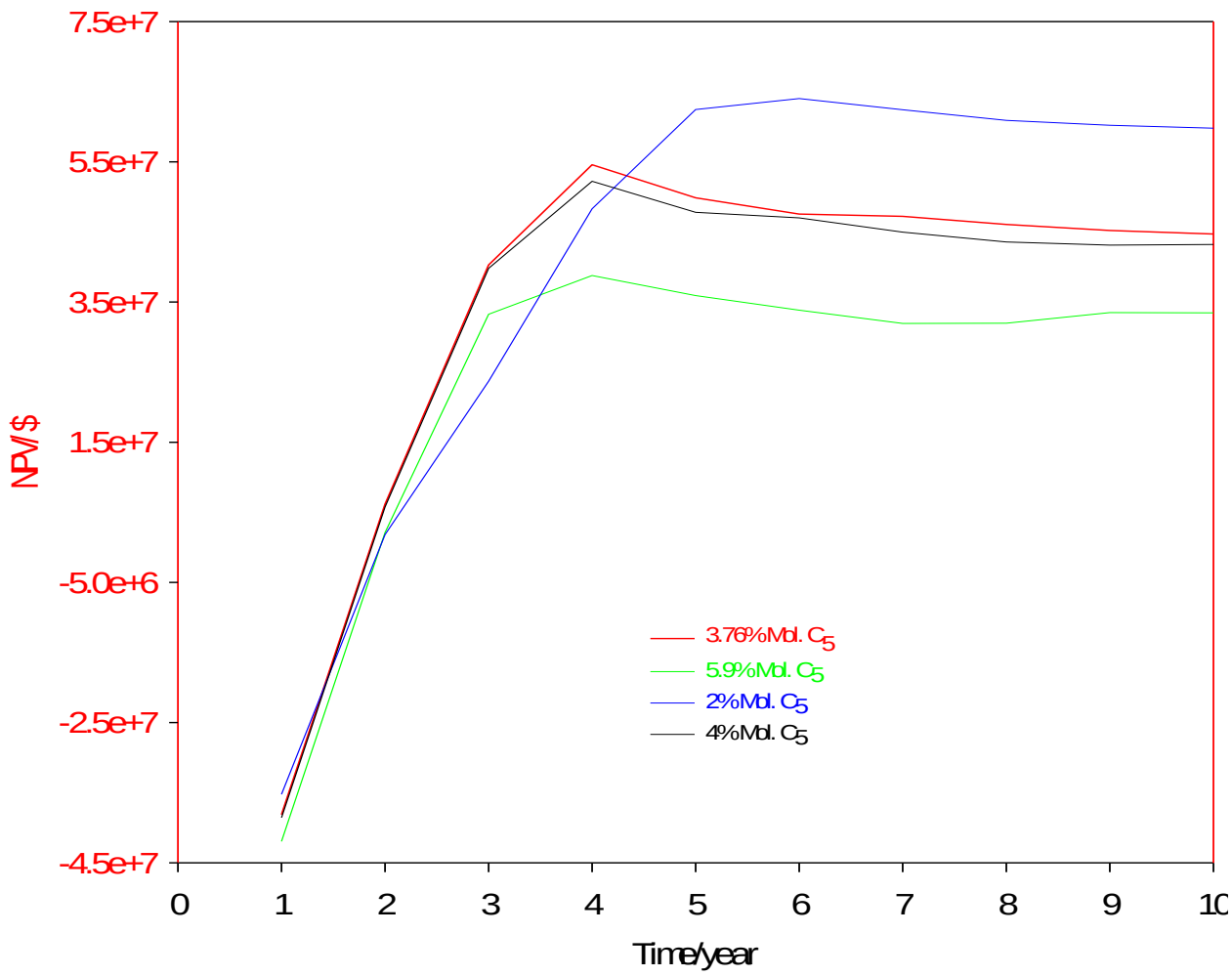

Figure 26: Comparison of the NPV for different molar ratios of C_5 using Giacchetta et al. (2015) method and pentane price obtained from company as reference price for solvent

5. CONCLUSIONS AND RECOMMENDATIONS

5.1. Conclusions

Simulations were carried out for three thermal recovery processes, namely, SAGD, Fast-SAGD, and ES-SAGD. Pentane (n-C_5) was used as the solvent in ES-SAGD and the molar concentration of n-C_5 was also varied. The purpose of this work is to use NPV as the discriminating economic indicator to determine the most economically viable thermal recovery process to be used in an Athabasca bitumen reservoir. The following main conclusions were reached from this simulation study:

1. ES-SAGD is a better thermal recovery process compared to SAGD and Fast-SAGD from the standpoint of NPV.
2. While the ultimate recovery of Fast-SAGD was the highest as it was able to access oil that would be likely bypassed by other thermal recovery processes, its NPV was mostly the lowest of the three recovery processes. This is due to the fact that fewer SAGD wells were initially deployed for production before CSS wells were later added.
3. Using pentane plus price as a representative cost for pentane, there seems to be a significant improvement in NPV as the molar concentration of n-C_5 increases from 2% to 3.76% but no noteworthy increase when the molar concentration increases from 3.76% to 5.9%. This may indicate that while increasing the molar concentration increases NPV, there is a point of diminishing return.
4. However, utilizing a very high cost of pentane, NPV showed no positive correlation with molar concentration of n-C_5. In fact, it was shown that for the most part, the 5.9% molar concentration of n-C_5 had the lowest NPV while the 3.76% had the highest until later in the production stage before it was overtaken by the 2% molar concentration of n-C_5. This shows that as the cost of

solvent becomes too prohibitive, the advantage of ES-SAGD over solvent-only injection process reduces.

5.2.Recommendations for Further Studies

1. This study should be extended to a heterogeneous reservoir. In fact, a real reservoir with production data can be co-opted into a future study.
2. It is also important to do a whole assemble of molar concentration of n-C_5 to see the point of diminishing return.
3. There also needs to be a more accurate and up to date description of costs associated with ES-SAGD processes since most of the costs used in this study are not very current.

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APPENDIX A: CMG PVT PROPERTIES EXPORTED FROM WINPROP

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*****************************************************************************
** THE FOLLOWING KEYWORDS CAN BE USED IN THE INITIALIZATION SECTION IN
STARS
*****************************************************************************
** MFRAC_OIL 'CH4' CON  3.2000E-02
** MFRAC_OIL 'NC5' CON  2.0000E-01
** MFRAC_OIL 'Heavy' CON  7.6800E-01
*****************************************************************************
** THE FOLLOWING SECTION CAN BE USED FOR THE COMPONENT PROPERTY INPUT
INTO STARS
*****************************************************************************
** PVT UNITS CONSISTENT WITH *INUNIT *SI
** Model and number of components
MODEL 4 4 4 1
COMPNAME 'WATER' 'CH4' 'NC5' 'Heavy'
**          --------   --------   --------   --------
CMM
0 0.016 0.0722 0.5725
PCRIT
0 4600.15 3375.7 1072.72
TCRIT
0.00 -82.55 196.82 888.16
** low/high pressure; low/high temperature
KVTABLIM 100 11600 10 460
**   60.000
**  110.000
**  160.000
**  210.000
**  260.000
**  310.000
**  360.000
**  410.000
**  460.000
** Gas-liquid K Value tables
KVTABLE 'CH4'
**
     320.55    54.509     30.34    21.285    16.547    13.636    11.667    10.249    9.1792    8.3445
7.6756     7.128     6.672    6.2865    5.9567    5.6715    5.4225    5.2035    5.0093     4.836    4.6804
4.54    4.4126    4.2965
        411    69.347      38.4    26.791     20.71     16.97    14.438     12.61    11.229    10.149    9.2815
8.5697    7.9753    7.4714     7.039    6.6639    6.3355    6.0455    5.7877    5.5569    5.3492    5.1612
4.9902    4.8341
     482.07    79.392    43.911    30.573    23.574    19.263    16.341    14.229    12.632    11.383
10.378    9.5528     8.863    8.2777     7.775    7.3386     6.956     6.618    6.3172    6.0478     5.805
5.5852    5.3852    5.2024
```

527.32 85.379 46.796 32.554 25.09 20.486 17.361 15.1 13.387 12.046 10.966 10.079 9.3368 8.7067 8.165 7.6944 7.2818 6.9171 6.5924 6.3014 6.0391 5.8016 5.5854 5.3878
550.16 89.394 48.166 33.177 25.486 20.787 17.608 15.31 13.57 12.206 11.108 10.204 9.4474 8.8047 8.252 7.7715 7.35 6.9772 6.6452 6.3475 6.0792 5.8361 5.6148 5.4125
555.22 90.826 48.835 33.315 25.362 20.569 17.369 15.078 13.353 12.005 10.922 10.032 9.2873 8.6548 8.1108 7.6379 7.2229 6.8559 6.5288 6.2357 5.9713 5.7317 5.5135 5.3141
546.42 89.861 48.456 33.025 25.031 20.186 16.958 14.663 12.95 11.621 10.56 9.6917 8.9679 8.3548 7.8286 7.3717 6.9712 6.6172 6.302 6.0195 5.7649 5.5341 5.324 5.132
526.99 87.008 47.065 32.134 24.358 19.612 16.432 14.165 12.472 11.163 10.123 9.276 8.5731 7.9802 7.473 7.034 6.6501 6.3115 6.0106 5.7412 5.4986 5.279 5.0792 4.8966
499.52 82.72 44.863 30.695 23.299 18.77 15.722 13.539 11.904 10.637 9.6296 8.8102 8.1316 7.5606 7.0737 6.6534 6.287 5.9646 5.6786 5.4232 5.1936 4.9861 4.7975 4.6255
465.98 77.349 42.041 28.818 21.907 17.668 14.809 12.755 11.213 10.014 9.059 8.281 7.6363 7.0939 6.6318 6.2336 5.8869 5.5825 5.313 5.0727 4.8571 4.6626 4.4862 4.3254
** 60.000
** 110.000
** 160.000
** 210.000
** 260.000
** 310.000
** 360.000
** 410.000
** 460.000
** Gas-liquid K Value tables
KVTABLE 'NC5'
**
0.81117 0.15222 0.093594 0.072572 0.062393 0.056891 0.053891 0.052433 0.052035 0.052436 0.053484 0.055091 0.057208 0.05981 0.062887 0.066441 0.070484 0.075032 0.080107 0.085737 0.091951 0.098782 0.10626 0.11444
3.9386 0.71214 0.42165 0.31452 0.25992 0.22766 0.20701 0.19322 0.18384 0.17751 0.17339 0.17096 0.16985 0.16983 0.17072 0.17239 0.17476 0.17775 0.18131 0.1854 0.19 0.19507 0.2006 0.20657
11.758 2.0959 1.2193 0.89125 0.72103 0.61792 0.5496 0.50167 0.46673 0.4406 0.42075 0.40554 0.39386 0.38495 0.37827 0.3734 0.37003 0.36795 0.36695 0.3669 0.36768 0.36919 0.37136 0.37412
25.461 4.5115 2.6116 1.8934 1.5158 1.2838 1.1278 1.0162 0.93312 0.86924 0.819 0.7788 0.7462 0.71949 0.69746 0.67921 0.66404 0.65143 0.64098 0.63235 0.62529 0.61958 0.61504 0.61153
44.69 7.8406 4.5038 3.253 2.5936 2.1851 1.9074 1.7066 1.5552 1.4373 1.3432 1.2666 1.2033 1.1503 1.1055 1.0672 1.0343 1.0058 0.98109 0.95949 0.94056 0.92394 0.90931 0.89641

67.755 11.774 6.6974 4.8005 3.8085 3.1967 2.7803 2.478 2.2486 2.0686 1.9237 1.8047 1.7054 1.6215 1.5496 1.4876 1.4335 1.3861 1.3443 1.3071 1.2739 1.2442 1.2175 1.1934
92.295 15.914 8.9807 6.3871 5.0323 4.2004 3.6371 3.2296 2.9206 2.678 2.4822 2.3209 2.1857 2.0708 1.972 1.8861 1.8109 1.7444 1.6854 1.6326 1.5852 1.5424 1.5035 1.4682
115.92 19.87 11.145 7.8782 6.1704 5.1217 4.4126 3.9012 3.5146 3.2118 2.968 2.7673 2.5991 2.456 2.3328 2.2255 2.1313 2.0479 1.9736 1.907 1.8469 1.7925 1.7431 1.6979
136.61 23.312 13.017 9.1595 7.1417 5.9018 5.0633 4.4587 4.0022 3.6453 3.3585 3.1229 2.9258 2.7584 2.6144 2.4891 2.3791 2.2818 2.195 2.1171 2.0469 1.9832 1.9252 1.8722
152.88 26 14.469 10.148 7.8861 6.4958 5.5551 4.8767 4.3646 3.9643 3.643 3.3793 3.159 2.9723 2.8118 2.6725 2.5503 2.4424 2.3462 2.26 2.1822 2.1118 2.0477 1.989
** 60.000
** 110.000
** 160.000
** 210.000
** 260.000
** 310.000
** 360.000
** 410.000
** 460.000
** Comparison of WinProp (W) and STARS K-value (S) phase split calculations
** A = Aqueous, L = Liquid, V = Vapor
** Pressure, kPa
** T, deg C 1.0000E+02 6.0000E+02 1.1000E+03 1.6000E+03 2.1000E+03 2.6000E+03 3.1000E+03 3.6000E+03 4.1000E+03 4.6000E+03 5.1000E+03 5.6000E+03 6.1000E+03 6.6000E+03 7.1000E+03 7.6000E+03 8.1000E+03 8.6000E+03 9.1000E+03 9.6000E+03 1.0100E+04 1.0600E+04 1.1100E+04 1.1600E+04
** 10.000 <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 60.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 110.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 160.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 210.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L

> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 260.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 310.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 360.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 410.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** 460.000 <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: LV,S: LV> <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L > <W: L ,S: L >
** Gas-liquid K Value tables
KVTABLE 'Heavy'
**
1e-016 2.0447e-016
1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1e-016 1.9997e-016 3.3406e-016 5.592e-016 9.3709e-016 1.5707e-015 2.6315e-015 4.4037e-015 7.357e-015 1.2264e-014 2.0392e-014 3.3802e-014 5.5843e-014
7.4244e-014 2.8021e-014 2.4735e-014 2.6805e-014 3.1933e-014 4.014e-014 5.2202e-014 6.9471e-014 9.3975e-014 1.2864e-013 1.7763e-013 2.4689e-013 3.4478e-013 4.8314e-013 6.7865e-013 9.5473e-013 1.3442e-012 1.893e-012 2.665e-012 3.7493e-012 5.2688e-012 7.3933e-012 1.0357e-011 1.4478e-011
6.3042e-011 2.543e-011 2.3646e-011 2.4151e-011 2.6353e-011 3.0036e-011 3.5256e-011 4.2249e-011 5.1391e-011 6.3209e-011 7.8394e-011 9.7841e-011 1.2269e-010 1.544e-010 1.9481e-010 2.4625e-010 3.1163e-010 3.9462e-010 4.9981e-010 6.3293e-010 8.0109e-010 1.0131e-009 1.28e-009 1.615e-009
1.0599e-008 3.6682e-009 3.6875e-009 3.9979e-009 4.3523e-009 4.7895e-009 5.3418e-009 6.0309e-009 6.8782e-009 7.9087e-009 9.1523e-009 1.0646e-008 1.2432e-008 1.4562e-008 1.7097e-008 2.0107e-008 2.3676e-008 2.79e-008 3.289e-008 3.8776e-008 4.5707e-008 5.3855e-008 6.3416e-008 7.4615e-008
5.7265e-007 1.6855e-007 1.563e-007 1.7142e-007 1.9329e-007 2.1659e-007 2.408e-007 2.6678e-007 2.9546e-007 3.2765e-007 3.6403e-007 4.0525e-007 4.5196e-007 5.0485e-007 5.6467e-007 6.3221e-007 7.0835e-007 7.9407e-007 8.9042e-007 9.9853e-007 1.1197e-006 1.2552e-006 1.4066e-006 1.5755e-006
1.3565e-005 3.5243e-006 2.9437e-006 3.0176e-006 3.3104e-006 3.7011e-006 4.1345e-006 4.5864e-006 5.0512e-006 5.5321e-006 6.0357e-006 6.5691e-006 7.1389e-006 7.751e-006

```
8.411e-006  9.1242e-006  9.8954e-006   1.073e-005  1.1632e-005  1.2608e-005  1.3661e-005  1.4799e-
005  1.6026e-005  1.7347e-005
     0.00017358  4.1094e-005  3.1485e-005  2.9965e-005  3.1038e-005   3.337e-005  3.6412e-005
3.9856e-005  4.3516e-005  4.7289e-005  5.1129e-005   5.503e-005  5.9005e-005  6.3077e-005
6.7273e-005  7.1618e-005  7.6137e-005  8.0851e-005  8.5781e-005  9.0944e-005  9.6359e-005
0.00010204     0.000108   0.00011426
      0.0013875   0.00030639   0.00021964   0.00019655   0.00019277    0.0001979   0.00020809
0.00022139   0.00023665   0.00025308   0.00027015   0.00028755   0.00030509   0.00032269
0.00034034   0.00035809   0.00037598   0.00039411   0.00041251   0.00043127   0.00045044
0.00047009   0.00049024   0.00051097
      0.0076921    0.0016117    0.0010983   0.00093683   0.00087879   0.00086659   0.00087939
0.0009075   0.00094555   0.00099011    0.0010388      0.00109    0.0011426    0.0011956    0.0012488
0.0013018    0.0013545     0.001407    0.0014593    0.0015117    0.0015641    0.0016169    0.0016702
0.001724
** reference pressure,     corresponding to the density
PRSR 1800
** reference temperature,  corresponding to the density
TEMR 11
** pressure at surface,    for reporting well rates, etc.
PSURF 101.325
** temperature at surface, for reporting well rates, etc.
TSURF 15.556
** Surface conditions
SURFLASH KVALUE
K_SURF 'CH4' 327.39
K_SURF 'NC5' 0.98516
K_SURF 'Heavy' 1e-016
MOLDEN
0 23250 9120 1768
CP
0 2.838e-006 1.068e-006 1.748e-007
CT1
0 0.001272 0.0001848 2.831e-005
CT2
0 3.189e-006 2.28e-006 4.071e-007
CPT
0 2.8e-007 2.126e-009 1.688e-010
** T, deg C        'WATER'       'CH4'       'NC5'     'Heavy'
**              --------    --------    --------    --------
VISCTABLE
  *ATPRES 100
**      temp
         10         0   0.09434   0.37822  1.8285e+008
         60         0   0.60976   0.93236        84620
        110         0   0.66923   0.89163       1258.8
        160         0   0.61159   0.72432       95.325
        210         0   0.55754   0.60508        18.05
        260         0    0.5242   0.53415       5.9027
        310         0   0.50224   0.48895       2.7143
```

```
        360        0   0.48124   0.45343       1.5539
 *ATPRES 1814.29
**      temp
         10        0  0.075481   0.38865  1.9286e+008
         60        0    0.1141   0.22834       123290
        110        0   0.15946   0.15205       1887.4
        160        0   0.37371   0.42727       103.83
        210        0   0.39296   0.41641       19.731
        260        0   0.41694   0.42008       6.4656
        310        0   0.43442   0.42203       2.9807
        360        0   0.43838    0.4144       1.7132
 *ATPRES 3528.57
**      temp
         10        0  0.077319   0.39905  2.0328e+008
         60        0   0.11676   0.23569       131630
        110        0   0.16149   0.15867       2029.7
        160        0   0.20553   0.10965       164.11
        210        0   0.37259   0.39982       21.488
        260        0   0.38664   0.39356       7.0511
        310        0   0.40729   0.39882       3.2561
        360        0    0.4173   0.39755       1.8769
 *ATPRES 5242.86
**      temp
         10        0  0.079162   0.40948  2.1412e+008
         60        0   0.11943   0.24297       140350
        110        0   0.16416   0.16494       2178.8
        160        0   0.20779   0.11696       175.73
        210        0   0.37766   0.40888       23.301
        260        0   0.39262   0.40371       7.6532
        310        0   0.40833   0.40428       3.5383
        360        0   0.41505   0.39964        2.044
 *ATPRES 6957.14
**      temp
         10        0  0.081005   0.41994  2.2541e+008
         60        0   0.12208   0.25016       149450
        110        0   0.16691   0.17097       2334.1
        160        0   0.21042   0.12335       188.04
        210        0   0.38389   0.41876       25.153
        260        0   0.39992   0.41472       8.2663
        310        0   0.41791   0.41758       3.8252
        360        0   0.42731   0.41543       2.2136
 *ATPRES 8671.43
**      temp
         10        0  0.082846   0.43042  2.3715e+008
         60        0    0.1247   0.25729       158910
        110        0   0.16972   0.17681       2494.8
        160        0    0.2134    0.1292       200.83
        210        0   0.39122   0.42943       27.032
        260        0   0.40855   0.42663        8.886
```

```
        310        0   0.42873   0.43166       4.1149
        360        0   0.44055   0.43178       2.3848
  *ATPRES 10385.7
**      temp
         10        0  0.084679    0.4409  2.4933e+008
         60        0   0.12729   0.26434       168710
        110        0   0.17259   0.18249       2660.2
        160        0    0.2167   0.13467       213.97
        210        0   0.39948   0.44081       28.925
        260        0   0.41835   0.43939       9.5088
        310        0   0.44069   0.44651       4.4061
        360        0   0.45474   0.44871        2.557
  *ATPRES 12100
**      temp
         10        0  0.086501   0.45138  2.6197e+008
         60        0   0.12984   0.27134       178850
        110        0   0.17547   0.18803       2829.8
        160        0   0.22025   0.13986       227.37
        210        0   0.40849   0.45276       30.827
        260        0   0.42912   0.45288       10.133
        310        0   0.45363   0.46205       4.6979
        360        0   0.46977   0.46618       2.7297
```

** The following is the complete WinProp fluid model description.

```
WINPROP *TITLE1     'Live Oil and Solvent Model for STARS'
WINPROP *TITLE2     ' '
WINPROP *TITLE3     ' '
WINPROP *INUNIT *SI
WINPROP *MODEL   *PR   *1978
WINPROP *NC        3      3
WINPROP *TRANSLATION 1
WINPROP *PVC3  1.8002835E+00
WINPROP *COMPNAME
WINPROP 'CH4     ' 'NC5     ' 'Heavy   '
WINPROP *HCFLAG
WINPROP   1  1  1
WINPROP *SG
WINPROP   3.0000000E-01  6.3100000E-01  1.0120000E+00
WINPROP *TB
WINPROP  -1.6145000E+02  3.6050000E+01  6.0886030E+02
WINPROP *PCRIT
WINPROP   4.5400000E+01  3.3315603E+01  1.0586893E+01
WINPROP *VCRIT
WINPROP   9.9000000E-02  3.0400000E-01  1.6372964E+00
WINPROP *TCRIT
WINPROP   1.9060000E+02  4.6996879E+02  1.1613062E+03
WINPROP *AC
WINPROP   8.0000000E-03  2.5100000E-01  1.2228739E+00
```

```
WINPROP *MW
WINPROP   1.6043000E+01  7.2151000E+01  5.7250000E+02
WINPROP *VSHIFT
WINPROP   0.0000000E+00 -4.1004487E-02  2.1380997E-01
WINPROP *VSHIF1
WINPROP   0.0000000E+00  0.0000000E+00  0.0000000E+00
WINPROP *TREFVS
WINPROP   1.5556000E+01  1.5556000E+01  1.5556000E+01
WINPROP *ZRA
WINPROP   2.8760000E-01  2.6850000E-01  2.3704990E-01
WINPROP *VISVC
WINPROP   9.9000000E-02  3.0400000E-01  1.6372964E+00
WINPROP *VISCOR *MODPEDERSEN
WINPROP *VISCOEFF
WINPROP   1.3040000E-04  2.3030000E+00  6.9064120E-03  2.4713244E+00  4.0285012E-01
WINPROP *OMEGA
WINPROP   4.5723553E-01  4.5723553E-01  4.5723553E-01
WINPROP *OMEGB
WINPROP   7.7796074E-02  7.7796074E-02  7.7796074E-02
WINPROP *PCHOR
WINPROP   7.7000000E+01  2.3150000E+02  1.1167113E+03
WINPROP *HREFCOR *HARVEY
WINPROP *IGHCOEF
WINPROP  -2.8385700E+00  5.3828500E-01 -2.1140900E-04  3.3927600E-07 -1.1643220E-10
1.3896120E-14 -6.0286900E-01
WINPROP   9.0420900E+00  1.1182900E-01  2.2851500E-04  8.6331000E-08 -5.4464900E-11
8.1845000E-15  1.8318900E-01
WINPROP   0.0000000E+00 -2.4160419E-02  3.8400171E-04 -5.6574156E-08  0.0000000E+00
0.0000000E+00  0.0000000E+00
WINPROP *HEATING_VALUES
WINPROP   8.4429001E+02  3.3536600E+03  0.0000000E+00
WINPROP *COMPOSITION *PRIMARY
WINPROP   4.0000000E-02  0.0000000E+00  9.6000000E-01
WINPROP *COMPOSITION *SECOND
WINPROP   0.0000000E+00  1.0000000E+00  0.0000000E+00
```