Techno-economic assessment of solvent-based bitumen extraction technologies including in-situ electromagnetic heating

By

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A thesis submitted in partial fulfillment of the requirements for the degree of

Master of Science

In

Engineering Management

Department of Mechanical Engineering

University of Alberta

Abstract

The oil sands are a vast fossil fuel resource that supports the worldwide energy supply. The bitumen found in fossil deposits is too viscous to flow under reservoir conditions. For this reason, steam-based processes such as steam-assisted gravity drainage (SAGD) are used to increase the reservoir temperature and allow the bitumen to flow. These processes have been successfully implemented. However, they are energy- and greenhouse gas (GHG) emission-intensive because a large amount of fossil fuels is burned to generate the steam. Methods of extraction using solvents are more promising technologies to lower the environmental impact of oil sands extraction. Solvent-based bitumen extraction technology as well effective solvent extraction incorporating electromagnetic heating (ESEIEH) technologies have been successfully tested on laboratory and pilot scale however there is very limited information on the comprehensive techno-economic assessment of bitumen produced from these technologies. This study conducts a techno-economic assessment of these technologies to understand the economic viability of these new processes.

In this study, process models of solvent-based bitumen extraction and ESEIEH technologies were developed to evaluate their equipment size and energy requirement. The solvent purification unit was optimized with a high-temperature distillation column. The techno-economic models for solvent-based bitumen extraction and ESEIEH evaluate the supply cost of dilbit produced by these technologies. Considering the uncertainty in the results, the supply costs range from C$48.20/bbl to C$63.70/bbl and from C$55.20/bbl to C$64.40/bbl for the solvent-based extraction and ESEIEH technologies, respectively. The solvent loss in the reservoir is the parameter that most affects solvent-based bitumen extraction technology cost. For the ESEIEH process, the diluent and transportation costs were the major cost contributors. The application of shallow and deep reservoirs was also investigated. For both the bitumen extraction technologies, the use of butane for shallow reservoirs increased the supply cost slightly. The solvent purification unit in the solvent-based bitumen extraction process was modeled through two additional pathways. Pathway I uses a dehydration and refrigeration system to lower the solvent losses in the processing facility. Pathway II uses high-pressure separators to reduce the plant size and lower the impurities in the solvent. The supply costs increased by 5.8% and 2.9% from the base case in pathways I and II, respectively. The developed scale factors of the solvent-based extraction process and ESEIEH technologies are 0.72 and 0.85, respectively. These results suggest that at larger plant capacities, there is a cost benefit due to economies of scale. The results also show that the solvent-based extraction and ESEIEH processes are cost-competitive for oil prices above US$50/barrel. The findings of this study can assist policymakers and industry in decision-making regarding these new bitumen extraction technologies.

Preface

This thesis is original work by Isabel Toro under the supervision of Dr. Amit Kumar. Chapter 2 and chapter 3 will be submitted to peer-reviewed journals as "The development of a techno-economic model for the assessment of Nsolv technology" and "The development of a techno-economic model for the assessment of the Effective Solvent Extraction Incorporating Electromagnetic Heating Technology" by Isabel Toro, Abayomi Olufemi Oni, Amit Kumar, respectively. I was responsible for concept formulation, data collection and analysis, model development, and manuscript composition. Dr. A.O. Oni contributed by providing guidance to conduct the research, assisting in the model development, and reviewing the manuscript. Dr. E. Gemechu contributed by reviewing the manuscript. Dr. A. Kumar was the supervisory author and was involved with the concept formation, model development, development of input data and assumptions, result assessment, and manuscript composition and review.

Acknowledgements

I am very grateful to have the opportunity to work with my supervisor, Dr. Amit Kumar. I thank him for his support and guidance during my research. I greatly admire his professionalism, critical thinking, hard work, and knowledge. I am deeply thankful to Dr. Amit Kumar for trusting me and my abilities.

I would also like to thank NSERC/Cenovus/Alberta Innovates Associate Industrial Research Chair in Energy and Environmental Systems Engineering and the Cenovus Energy Endowed Chair in Environmental Engineering for providing financial support for this project. As a part of the University of Alberta's Future Energy Systems (FES) research initiative, this research was made possible in part thanks to funding from the Canada First Research Excellence Fund (CFREF).

I thank Dr. Abayomi Olufemi Oni and Dr. Eskinder Gemechu for their supervision and guidance through my research. I especially thank Dr. Oni for always encouraging me to improve my work skills and for his unconditional support during my master's studies. I thank Astrid Blodgett for her editorial assistance with my research papers and thesis. I also thank my colleagues from the research group, Giovanni Di Lullo, Miguel Baritto, and Karina Anaya, for their support and feedback on my research.

I am greatly thankful for the unconditional love and support of my parents. Thanks to all my friends and colleagues who have made my experience at the University of Alberta very enjoyable.

Nomenclature

API	American Petroleum Institute gravity
bbl	Barrel
BPD	Barrel per day
CCS	Cyclic steam stimulation
cEOR	Cumulative electricity-to-oil ratio
CH_4	Methane
CHSI	Cyclic hot solvent injection
CO_2	Carbon dioxide
CO_2e	Carbon dioxide equivalents
COM	Cost of manufacturing
CPF	Central processing facility
CSI	Cyclic solvent injection
CSS	Cyclic steam stimulation
DC	Distillation column
DCFA	Discounted cash flow analysis
Dilbit	Diluted bitumen
EH	Emulsion heater
ES-SAGD	Expanding solvent-SAGD
ESEIEH	Effective Solvent Extraction Incorporating Electromagnetic Heating
ESP	Electrical submersible pump
FCI	Fixed capital investment
FH	Fired heater
FWKO	Free-water knockout

GHG	Greenhouse gas
GOR	Gas-to-oil ratio
HCR	Hydrocarbon recovery column
HPS	High-pressure separator
HX	Heat exchanger
IGF	Induced gas flotation
LHV	Low heating value
LPS	Low-pressure separator
NCG	Non-condensable gases
NGCC	Natural gas combined cycle
OT	Oil treater separator
OTSG	Once-through steam generator
OTU	Oil treatment unit
RFD	Radio frequency device
RFT	Radio frequency transmitter
RUST	Regression, Uncertainty, and Sensitivity Tool
SAGD	Steam-assisted gravity drainage
SAP	Solvent-aided process
SAS	Steam-alternating solvent
SLT	Slop tank
SOR	Solvent-to-oil ratio
SOR_{Steam}	Steam-to-oil-ratio
SPC	Solvent purification column
SPU	Solvent purification unit
ST	Skim tank

SVI	Solvent vaporization and injection unit
VAPEX	Solvent vapor extraction
VRU	Vapor recovery unit
WCS	Western Canadian Select
WLS	Warm lime softener
WOR	Water-to-oil ratio
WTI	West Texas Intermediate
WTU	Water treatment unit

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Chapter 1

Introduction

1.1. Background

The global economy is growing significantly due to population growth and the industrialization of developing countries, which inevitably will increase global energy demand [1]. A 19% increase in world energy demand is expected by 2040 from 2019 levels [2]. Although economic growth is a good indicator of social welfare, the consequent increase in energy requirements challenges environmental sustainability as majority of energy comes from fossil fuels. In 2018, 54.3% of the world's primary energy was supplied by oil and gas and only 18.8% by cleaner sources such as nuclear and renewable energy [3]. The overall greenhouse gas (GHG) emissions could be reduced by increasing the share of cleaner energy sources while reducing fossil fuel use. However, studies have forecasted that by 2050 almost 49% of the energy demand will still rely on oil and gas supply [4]. Thus, it is crucial to search for cleaner methods of crude oil extraction to reduce GHG emissions.

Because conventional oil reservoirs are being depleted, oil production now relies on unconventional oil deposits such as oil sand reservoirs. Canada is the fourth-largest producer of crude oil and its oil sands are the third-largest proven oil reserve in the world [5]. The oil sands are a mixture of bitumen, sand, water, and clay. Oil sands can be extracted by surface mining from shallow deposits, but around 80% are located in deep reservoirs and must be extracted by in situ recovery [6]. At reservoir conditions, the bitumen is too viscous to flow and requires other methods of extraction than conventional oil. Thermal processes such as steam-assisted gravity drainage (SAGD) and cyclic steam stimulation (CSS) are the most used in situ recovery methods of extracting oil sands [7]. These processes lower bitumen viscosity by increasing the reservoir temperature (via thermal conduction) through the injection of heated steam. This allows the bitumen to flow easily to the producer well to be pumped to the surface. Although these extraction methods have been successfully implemented, they are energy- and GHG emission-intensive since they require large amounts of fossil fuels and water to generate steam.

To reduce GHG emissions, the oil sands industry has focused on improving the extraction performance of steam-based methods. Several methods of co-injecting solvents with steam have been proposed to reduce the water use and energy consumption of SAGD [8]. These hybrid steam-solvent processes, such as expanding solvent-SAGD (ES-SAGD), solvent-aided process (SAP), and steam-alternating solvent (SAS), use heat and mass transfer to reduce bitumen viscosity [9]. The pilot tests of these technologies reported an increase in oil production rate and a reduction in the steam-to-oil ratio (SOR_{steam}) compared to conventional SAGD [10]. However, the difficulty of sustaining the solvent in the vapor phase at steam chamber conditions eventually reduced solvent diffusion in the bitumen and disrupted the extraction process [11]. This obstacle was resolved by exclusively using solvents in what are known as solvent-based methods. These methods eliminate the use of water while addressing the thermodynamic equilibrium concern of the hybrid processes.

Cold solvent injection processes such as solvent vapor extraction (VAPEX) were proposed next [12]. During cold solvent injection extraction, a light hydrocarbon, typically propane or butane, is injected into the reservoir. Reducing oil viscosity relies on the mass transfer of the solvent into the bitumen; however, the cold nature of oil sands reservoirs makes the process too slow to be viable [13]. Oil production rates, as well as oil recovery, were significantly lower in VAPEX than SAGD [14]. To address the low production rates, technologies that inject heated solvents were developed.

Effective solvent extraction incorporating electromagnetic heating (ESEIEH) and Nsolv technology are promising processes that tackle the main drawbacks of previous approaches [15]. ESEIEH technology pre-heats the reservoir through electromagnetic heating and subsequently injects a heated solvent. Nsolv only injects a heated solvent, relying on the latent heat of condensation to increase the bitumen temperature [16]. Since the Nsolv process depends on the condensation of the solvent at reservoir conditions, the solvent poisoning by non-condensable gases disrupts the extraction process [17]. Therefore, Nsolv requires a higher degree of solvent purity while ESEIEH, which relies on electromagnetic heating, is more flexible on the solvent purity requirements. Pilot plants for these two technologies have reported higher oil production rates and lower GHG emissions than SAGD [18] [19].

Economic challenges arise from the bitumen produced by the steam-based and hybrid steam-solvent processes. For instance, the bitumen produced by steam-based processes is viscous and extra heavy, requiring expensive diluents for transportation and complex facilities for refining. Since these expenses are not covered by the oil producers and because of market conditions, Canadian oil sands are commonly sold at lower prices than light oil [20]. Additionally, steam generation increases costs compared to conventional oil production. In SAGD, the cost of natural gas, which is mainly used for steam generation, accounts for 60% of the total operating cost [21]. Likewise, the emission-intensive nature of steam generation increases the expense to cover carbon taxes imposed by the government. With strict environmental regulation, the government makes the steam-based process more expensive and encourages investments in less emission-intensive technologies [22]. Although Nsolv and ESEIEH have arisen as potential technologies to lower GHG emissions, their economic feasibility must be ascertained before they can be commercialized. This thesis aims at addressing this gap in knowledge.

In this research, the techno-economic performances of ESEIEH and Nsolv technology were evaluated. To assess their potential, ESEIEH and Nsolv processes were modeled to capture their equipment and energy requirements. Subsequently, techno-economic assessment models were developed for both technologies. The supply cost of diluted bitumen (dilbit) produced by these technologies was calculated and compared to current oil prices. The results assist the decision-making of those in industry and policymakers regarding the deployment of ESEIEH and Nsolv technology.

1.2. Literature review

Most of the studies on solvent-based methods of extraction have focused on the thermodynamic equilibrium of the extraction process. For instance, Azinfar et al. [23] studied the phase behavior of the Athabasca bitumen in the presence of propane. The authors developed a correlation of propane solubility at different reservoir conditions and concluded that at a higher pressure there is a faster solubility of propane into the bitumen. Haddadnia et al. [24] conducted liquid-vapor equilibrium experiments to assess the solubility and viscosity reduction of bitumen when in contact with light hydrocarbons. Nourozeieh et al. [25] studied the solubility of vapor butane into bitumen and concluded that the effect of solubility on reducing viscosity is higher at high pressure and low temperature. Unlike the previous studies, Sadeghi et al. [26] conducted experimental analyses on the liquid-liquid equilibrium of propane, butane, and bitumen systems at isobaric conditions at different temperatures.

Other studies have explored more specific aspects of Nsolv and ESEIEH technologies. Irani and Gates [27] investigated liquid pool development in an Nsolv pilot plant. The authors concluded that operating at greater temperatures could improve the stability of the Nsolv process. Zhang et al. [28] compared oil production and performance of Nsolv technology and cyclic hot solvent injection (CHSI). Although Nsolv resulted in lower production rates than CHSI, solvent use was more efficient in Nsolv than in CHSI. For ESEIEH, the radio frequency power that allows the continuous growth of the desiccates zone was investigated by Irani et al. [29]. The efficiency of the antenna and reservoir mobility were identified as key parameters to maintain the growth of the desiccated zone. Others have conducted numerical modeling of the reservoir heating through radio frequency in the presence of solvents [30] [31]. Despande et al. [32] studied the well completion required to install the radio frequency device. Wang et al. [33] and Saeedfar et al. [34] focused on the antenna layout to optimize the extraction process. Although technical aspects and thermodynamic equilibrium of Nsolv and ESEIEH technologies have been intensively investigated, there are few studies on the GHG emissions and economic performance of these solvent-based methods of extraction.

Soiket et al. [35] and Safaei et al. [36] developed process models to assess the energy and GHG emissions of Nsolv and ESEIEH technologies, respectively. The studies concluded that Nsolv and ESEIEH have the potential to lower GHG emissions of bitumen extraction compared to steam-based processes. Both studies used energy-intensive units to purify the solvent, reducing the ability to capture the potential of Nsolv and ESEIEH to lower energy consumption. Emissions reduction Alberta (ERA) supported field demonstrations of both technologies to assess their oil production rate, energy, and emission intensity. The results showed around 50% reduction in GHG emissions at competitive oil production rates compared to SAGD [18] [19]. These pilot tests do not present a clear description of the processing facility, and the small-scale production might not exhibit the specific energy requirement of the technologies at a commercial scale. Thus, there is no comprehensive modeling of Nsolv and ESEIEH technologies in which the solvent purification and recovery are optimized in a large capacity plant.

Another knowledge gap in the literature on Nsolv and ESEIEH technologies is related to economic performance and scale factor. Scale factor is defined as the parameter which helps in determining the capital cost of the extraction plant at various capacities. An anticipated Nsolv commercial plant in Alberta was expected to cost 50% less than SAGD; however, the project was not developed, and so the economics could not be explored [18]. Spence et al. [37] conducted a high-level evaluation of the supply cost of bitumen produced from the ESEIEH process. The study found that the bitumen is cost-competitive compared to steam-based methods. However, the assessment provided no details on the processing facility, nor bitumen and solvent recovery, which are important aspects of the process. Process parameters that might change the profitability of ESEIEH, such as operating conditions, were not specified, nor was their cost impact investigated. This thesis, therefore, addresses these gaps by developing process models of Nsolv and ESEIEH technologies to assess their economic performance. The process models allow us to determine mass and energy balance as well as equipment size. The supply costs of dilbit produced by Nsolv and ESEIEH technologies were calculated through development of data-intensive discounted cash flow analyses. For comparison purposes, the Nsolv technology was also modeled using a refrigeration system and high-pressure separators for the solvent purification process. The application of these technologies in shallow reservoirs and the scale factor were also investigated. Additionally, sensitivity and uncertainty analyses were performed to identify the key input parameters and improve the accuracy of the results.

1.3. Research gaps

This thesis is aimed to address the following key knowledge gaps:

- Current studies of Nsolv and ESEIEH technologies use energy- and cost-intensive units to purify the solvent. These studies use conventional units used in natural gas processing plants to remove impurities from the solvent. However, they do not consider other units that could improve the energy performance of the technologies. Thus, a detailed process model was developed which focuses on a more energy- and cost-efficient ways to recycle solvent.
- The type of solvent used in Nsolv and ESEIEH technologies will depend on the reservoir conditions. Changing the solvent would change the energy consumption and cost. However, studies assessing the energy and cost of Nsolv and ESEIEH using different solvents were not found in the literature. To address this gap, process models for Nsolv and ESEIEH were developed for shallow and reservoir conditions. The results help to identify the advantages and disadvantages of each application.

- There are very limited information in the literature on the economic performance of Nsolv and ESEIEH technologies. The studies evaluate small-scale plants yet provide no details on the processing facility, particularly bitumen and solvent recovery. Given the potential for benefits from economies of scale, assessing small-scale plants might not capture the cost competitiveness of the technologies in the commercial stage. To address this issue, both technology plants were modeled at a commercial scale and the separation and purification processes are described in detail for the reader.
- No scale factor for Nsolv and ESEIEH technologies has been developed. The scale factor helps to identify the cost benefit at different plant capacities. This study presents the scale factors for Nsolv and ESEIEH technologies. The results will help investors to estimate capital cost requirements for greenfield projects using these solvent-based technologies.
- Most studies do not consider the uncertainty in their results. Thus, sensitivity and uncertainty analyses were performed to identify the parameters that most impact the results. The accuracy of the results of the cost performance of Nsolv and ESEIEH are improved by considering the differences in input values.

1.4. Objectives

This research aims to develop simulation models to assess the techno-economic performance of ESEIEH and Nsolv technologies. The main objectives are accomplished by carrying out the following specific objectives:

- Developing process simulation models for ESEIEH and Nsolv technologies to process 25,000 barrels per day.
- Developing techno-economic models to determine the supply cost of dilbit produced by ESEIEH and Nsolv technologies.
- Evaluating the economic feasibility of the dilbit produced by ESEIEH and Nsolv technologies.
- Identifying the key economic parameters that influence the supply cost of dilbit produced by ESEIEH and Nsolv technologies through comprehensive sensitivity and uncertainty analyzes.
- Developing a scale factor for the ESEIEH and Nsolv processes to determine the cost of production of bitumen at various capacities.

1.5. Scope and limitations

Few studies on the economic performance of Nsolv and ESEIEH technologies have been conducted. This study aims to understand the cost competitiveness of the Nsolv and ESEIEH processes. In this work, process models of both technologies were developed to obtain input data for equipment and energy calculations. Techno-economic models were then developed to evaluate the supply cost of dilbit produced by these technologies. The scale factors of the technologies were determined for evaluation of investment costs. Sensitivity and uncertainty analyses were performed to capture the variability of the input data to refine and improve model results.

The major limitation of the present study is the lack of information on oil specifications produced by Nsolv and ESEIEH. Although this research accounts for improvements in density, the grade of in situ upgrading will depend on reservoir conditions and raw bitumen properties. Since oil quality is expected to improve when solvents are used, the economic benefit will differ from conventional bitumen produced by steam-based methods. Thus, knowing the equivalence of the produced oil will allow more accurate assessment of the economic viability of these processes.

1.6. Organization of the thesis

This is a paper-based thesis comprised of four chapters that can be read independently. The chapters share assumptions and data required for the elaboration of the models. Thus, information repetition is unavoidable. The thesis chapters are described as follow:

Chapter 1 (Introduction) presents the knowledge gaps, research motivation, and general outline of the study.

Chapter 2 describes the techno-economic evaluation of the Nsolv process, a solvent-based bitumen extraction process. The supply cost of the dilbit produced by Nsolv was calculated through development of techno-economic model. The process model was developed to gather equipment and energy information for the cost analysis. For comparison purposes, three pathways based on the solvent purification unit were developed for Nsolv technology. The pathways were used to assess solvent purification capability and solvent losses, and to identify the more energy-efficient solvent recovery process. The three pathways were compared from an economic perspective using the less energy-intensive pathway as the base case. The scale factor and application of Nsolv technology to deep and shallow reservoirs were studied. The uncertainty and sensitivity were also included to identify the probable range of the model outputs. Finally, the supply cost of dilbit was discussed under current and future oil prices scenario to assess its cost competitiveness.

Chapter 3 presents the techno-economic assessment used to evaluate the supply cost of the dilbit produced by ESEIEH technology. Since the ESEIEH process is more flexible with respect to the solvent purity requirement, the model was developed using the less energy-intensive pathway described in chapter 2. The supply cost was calculated using the techno-economic model. The major cost components of dilbit production were identified and discussed. The scale factor of the ESEIEH technology was also calculated to identify the cost-benefit at different capacity plants. Sensitivity and uncertainty analyses were carried out to identify the most sensitive input parameters and their impact on supply cost. The cost of the ESEIEH process was compared to current and future oil prices to assess the economic viability of this technology.

Chapter 4 summarizes the key findings and new contributions of this research. The main conclusions are compiled and recommendations for future work are proposed.

Chapter 2

Techno-economic assessment of solvent-based bitumen extraction technology

2.1. Introduction

Traditional methods of in situ oil sands extraction use high-pressure steam to mobilize bitumen deep beneath the ground known as steam-assisted gravity drainage (SAGD) [38]. To produce high-pressure steam, however, requires burning fossil fuels, thus generating greenhouse gas (GHG) emissions [39]. Because of the tight regulations on GHG emissions, the oil sands industry has committed over $2.3 billion in technology funding to reduce fossil fuels [40]. Additionally, oil sands production is more challenging and cost-intensive than conventional oil extraction. For steam-based methods, the addition of diluent to reduce its viscosity during transport and the low quality of the bitumen reduce its selling price down compared to conventional crude oils [41]. Over the last 10 years, the Western Canadian Select (WCS), a benchmark for Canadian diluted bitumen (or dilbit), fell by US$ 17/bbl from the West Texas Intermediate (WTI), the most traded oil benchmark in the world [42]. For a profitable steam-based process, an equivalent WTI price in the range of 45.9-52.8 US$/bbl is required [43]. However, at lower oil prices, steam-based processes might not be profitable, and more cost-efficient methods of extraction will be needed.

Among the proposed technologies, the use of solvents has aroused a lot of interest from the industry since it eliminates the use of large amounts of steam. This method has been successfully tested at laboratory and pilot scales, and considerable reductions in GHG emissions have been reported [10]. However, the operation of this technology is different from traditional steam-based methods. There is very limited information available on the economic viability of these methods in the public domain. Thus, its economic feasibility must be ascertained before it is taken to the commercial stage [44]. This thesis is aimed at addressing this gap in knowledge.

The use of hydrocarbon solvents for the in situ extraction of oil sands began with co-injecting light hydrocarbons with steam [10] to reduce the steam requirement while increasing oil production rates [45]. Although this method improved the steam-based performance, water use and energy consumption were still high and further improvements were needed. To further decrease GHG emissions, solvent injection methods such as solvent vapour extraction (VAPEX) and cyclic solvent injection (CSI) were proposed. These methods are known as isothermal or cold solvent injection, in which oil viscosity is reduced by the molecular diffusion of the solvent into the bitumen [46]. The energy requirement to vaporize the solvent was about 3% of the energy required to generate the steam [47]. However, solvent diffusion is significantly slower than heat diffusion at reservoir conditions, and therefore these methods have low production rates compared to conventional steam-based methods [48]. Thermal solvent injection processes such as warm VAPEX and Nsolv were proposed to tackle the low production rates in the cold-solvent injection approach. Thermal solvent injection processes use superheated solvents that transfer the heat of condensation to the bitumen [49]. The heat transfer increases the oil temperature while the solvent condenses and dilutes the bitumen. Warm VAPEX was tested successfully on a laboratory scale; however, this technology does not consider the effects of non-condensable gases (NCG) in the process [50]. Some studies have shown how NCG accumulation in the solvent chamber impedes solvent condensation and eventually disrupts the extraction process [17] [51]. Unlike warm VAPEX, Nsolv technology ensures the injection of a purified solvent, preventing NCG contamination and allowing extraction stability [15].

The Nsolv process has been successfully tested on laboratory and pilot plant scales. Several studies carried out experimental analyses of the phase behavior and equilibrium of Nsolv technology [23] [26]. Others explored technical aspects such as sub-cool control to ensure the balance of the Nsolv process [52] [53]. Comparative analysis of the production rate performances of Nsolv and cyclic hot solvent injection (CHSI), another solvent-based extraction method, has been conducted [28]. This study claims that Nsolv's solvent use is more efficient and achieves similar oil production rates to CHSI. Other studies have investigated the precipitation of asphaltene, a heavy fraction of the bitumen, present in solvent-based methods such as Nsolv [54] [51]. The GHG emissions and energy consumption of the Nsolv process have also been studied earlier. Soiket et al. [35] developed a process model of solvent-based bitumen extraction technology to assess its energy requirement and GHG emissions. The authors found that solvent-based bitumen extraction reduces energy consumption by 68-87% from steam-based methods of extraction. This reduction is in accordance with the energy requirement reported by a pilot plant. The pilot plant showed a 54% reduction of GHG emissions compared to SAGD, high oil production rates, and improved oil quality [18]. A commercial Nsolv plant was expected to cut almost 50% in capital costs and be economically viable at reasonable oil prices; however, the project did not reach commercial scale and so this could not be shown yet [18]. No other commercial-scale plants have been developed; therefore, there is little or no information on the economic sustainability of solvent-based bitumen extraction technology at a commercial scale.

Understanding the cost performance of new, cleaner, and more cost-efficient technologies is fundamental to their commercialization. Currently, there are no studies on the economic feasibility of the solvent-based bitumen extraction process. Such information is important for decision-making. This study is aimed at filling this knowledge gap. We developed a techno-economic model to assess the economic feasibility of the solvent-based bitumen extraction technology. A process simulation model of solvent-based bitumen extraction similar to Nsolv's processing facility was developed using a high-temperature distillation column (base case) to purify the solvent. Since achieving high solvent purity is crucial in solvent-based bitumen extraction technology, we investigated two alternative pathways with different solvent purification units. Then, we evaluated the supply cost of the dilbit produced by solvent-based extraction texhnology and compared it with market oil prices. We investigated the cost impacts of plant capacity and solvent selection. Scale factors were developed for all three scenarios. The developed scale factors can help industry make initial estimates required for the capital investment of a project. Furthermore, we carried out sensitivity and uncertainty analysis to evaluate the variability in supply costs due to uncertainties in inputs parameters. The present work was conducted with the following specific objectives:

- To develop process simulation models for the solvent-based bitumen extraction process.
- To evaluate the economic feasibility of the dilbit produced by this process.
- To identify the key parameters that influence the supply cost of dilbit from solvent-based bitumen extraction technology.
- To develop a scale factor for this extraction technology.

2.2. Process description

Solvent-based bitumen extraction technology is based on the gravity drainage concept used in SAGD. Thus, it uses the same structure of the two horizontal wells located one above the other. The upper and lower wells are known as injector and producer wells, respectively. In the solvent-based bitumen extraction process, a purified solvent is injected into the reservoir through the injector well. The vaporized solvent, commonly propane or butane, delivers the condensation heat to the bitumen. Once the solvent condenses, it dilutes the bitumen, and an emulsion of solvent, bitumen, and connate water is formed. The emulsion flows to the producer well assisted by gravity, where it is pumped to the surface through an electric submergible pump (ESP). The emulsion is transported through the well pads and gathering lines to the central processing facility (CPF).

Once at the surface, the emulsion enters the oil treatment unit (OTU) where the water, bitumen, and solvent are separated. The emulsion is heated in the emulsion heater (EH) and sent to a high-pressure separator (HPS-1) where gases (solvent, NCG, and water) are released from the liquid emulsion. The emulsion continues to the free-water knockout (FKWO) and oil treater (OT) vessels that separate the oil from the connate water and off-gases. The bitumen is mixed with diluent, and the blend (dilbit) is shipped through the pipelines to terminals or upgrader facilities. The gas from the HPS-1 is liquified by reducing its temperature and sent to another high-pressure separator (HPS-2). In the HPS-2, NCG is released from the top while the liquid solvent is sent to a hydrocarbon recovery column (HCR). In the HCR, the bitumen trapped in the solvent is recovered as the bottom product. Then, the solvent is directed to the solvent purification unit (SPU). In the SPU, the liquid solvent from the HCR is supplied at saturation conditions into a high-temperature distillation column, known as the solvent purification column (SPC). Finally, the NCG and purified solvent are produced as top and bottom products of the SPC, respectively.

The connate water separated in the OTU is pumped to the water treatment unit (WTU) where the remaining oil droplets and solid particles are removed in the skim tank (ST), induced gas flotation (IGF), and slop tank (SLT). The water is then processed through warm lime softening (WLS), which reduces the amount of dissolved minerals, and can be disposed. Finally, the off-gases recovered in the CPF are directed to the fuel gas system to be used as a source of energy in the plant. Figure 2.1 outlines the above-mentioned process.

2.3. Process modeling and simulation

The solvent-based bitumen extraction processing facility similar to the proposed facility by Nsolv process was simulated using Aspen HYSYS Version 10. The Peng-Robinson fluid package was used to model the fluid components since it adequately represents the phase behavior of hydrocarbon mixtures. The CPF was designed to process 25,000 BPD of bitumen mixed with 701,825 stdft3/day of propane (solvent) and 14,037 stdft3/day of water which is based on a standard size of the SAGD unit. Additionally, the CPF was simulated using butane to understand the impact of solvent selection and reservoir conditions on the economics of solvent-based extraction bitumen extraction process. Table 2.1 shows the bitumen and diluent properties as well as production rates used for the simulation.

Table 2.1: Bitumen properties and production rates

Parameter	Value	Comments/Remarks
Raw bitumen viscosity [cP]	2,377,340	Assumed
Bitumen density [API]	12	[18]
Diluent density [API]	60	[55]
Solvent-to-oil-ratio [m3/m3]	5	[18]
Solvent hold-up in the reservoir [%]	20	[18]
Gas-to-oil-ratio [m3/m3]	5	[56]
Water-to-oil ratio [m3/m3]	0.1	[56]
Average plant capacity	95%	[57]

2.4. Supply cost analysis

2.4.1. Capital cost

The capital cost comprises the fixed capital investment (FCI) and the working capital. The FCI estimate includes the investment of greenfield CPF construction for solvent-based bitumen extraction technology, well pairs, well pads, and gathering lines. The working capital is the liquidity required to start the operation of the project and was assumed to be 15% of the FCI [57].

The equipment sizes and CPF construction cost were calculated through the cost estimator in Aspen Icarus. The investment of the CPF was then re-evaluated by adjusting the material selection of the equipment based on their operating conditions. The CPF construction cost was converted to 2021 Canadian dollars (CAD). The well pair, well pad, and gathering line costs were taken from industry reports for SAGD projects given that the well configuration is similar to Nsolv technology. These costs were inflated to 2021 and are shown in Table 2.2. The number of well pairs was calculated based on the Nenniger correlation and assumed reservoir conditions shown in Table A1 (Appendix) [58].

Table 2.2: Capital cost assumptions

Parameter	Value	Comments/Remarks
Drilling and completion cost per well pair [million CAD]	3.5	[35]
Well pairs per well pad	8.0	[35]
Cost per well pad [million CAD]	28.5	[35]
Cost of gathering lines [million CAD]	95.8	[35]

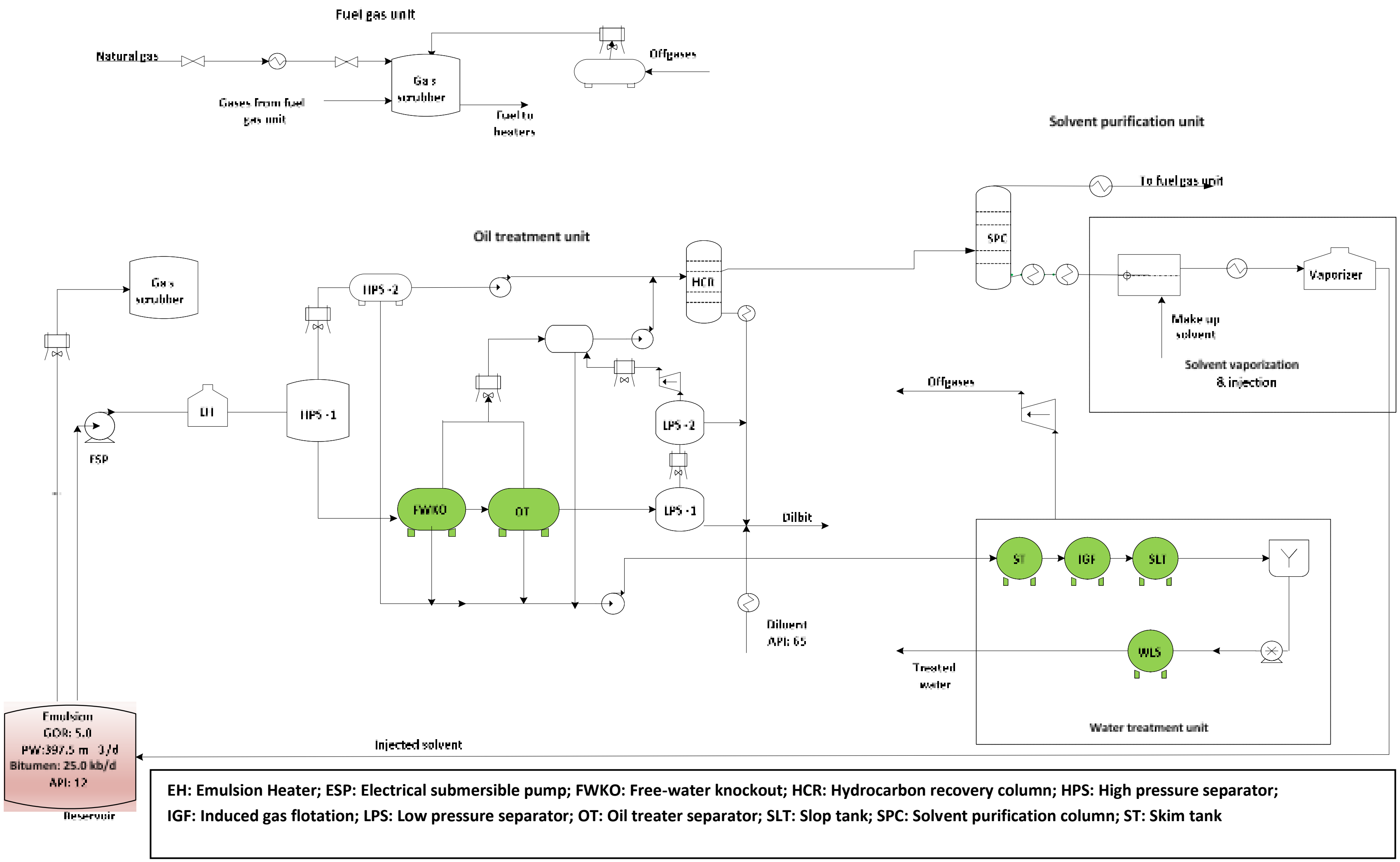


Figure 2.1: Process flow diagram for the solvent-based bitumen extraction technology (base case)

2.4.2. Operating cost

The operating cost refers to ongoing operation expenses of a project. These expenses in this study are the direct and indirect costs incurred while extracting, producing, processing, diluent and transporting bitumen. The study uses the data developed by Aman et al. [59] to calculate the transportation cost of the dilbit from Edmonton to Cushing, Oklahoma (US) with no diluent return. The utility requirements were divided in the natural gas and electricity consumption of the CPF equipment and the ESPs in the field. Fan cooling power was calculated through the developed simulation model.

Energy requirement and utility costs are evaluated with equations 1-3. The utility prices of electricity and natural gas were taken from Canada Energy Regulator data for the province of Alberta [60]. The input information of the operating cost (including equipment efficiencies) and assumptions are shown in Table 2.3 and Table A2 (Appendix), respectively.

Pump:

$$Power\ required_{real} = \frac{(P_{out} - P_{in}) * m_{liquid}}{Liquid\ density * \eta_p} \tag{1}$$

Centrifugal compressor:

$$Power\ required_{real} = \frac{(P_{out} - P_{in}) * m_{gas}}{\eta_c} \tag{2}$$

where m is the flow rate, η is the efficiency of the equipment, and P_{out} and P_{in} represent the inlet and outlet pressures, respectively.

Fired heaters and reboilers:

$$Natural\ gas\ \left[\frac{kg}{h}\right] = \frac{Heat_{required}[kW]}{LHV\left[\frac{kWh}{kg}\right] * \eta_b} \quad (3)$$

where LHV is the low heating value of natural gas and η_b corresponds to the boiler efficiency.

Table 2.3: Energy and operating cost input data

Parameter	Value	Reference
Electricity [C$/GJ]	16.5	[60]
Natural gas [C$/GJ]	2.6	[60]
Propane [C$/gal]	0.58	[61]
Butane [C$/gal]	0.71	[62]
Diluent cost [C$/bbl of diluent]	63	[55]
Operating labor [C$/h]	38	[63]
Prime efficiency [%]	75	[64]
Boiler efficiency [%]	80	[65]
LHV natural gas [MJ/kg]	46	[66]

2.4.3. Bitumen production cost

A discounted cash flow analysis (DCFA) was used to estimate the supply cost of the dilbit produced by solvent-based bitumen extraction process. The supply cost determines the average selling price of the dilbit in which all the expenses are covered, and a specific rate of return is assumed throughout a project's duration. The DCFA is a valuation method that uses the theory of the time value of money to calculate the present worth of an investment from its future cash flows [67]. The present value of the net income in year k (DCF_k) is given by the equation A1-A2 (Appendix):

2.4.4. Scale factor

The scale factor helps to develop cost estimates for different plant capacities [68]. The production rates and plant capacity were varied to determine the plant's economies of scale. The cost-to-capacity method was used to calculate the scale factor using the following equation (from Tribe and Alpine [69]):

$$\frac{C_2}{C_1} = \left(\frac{Q_2}{Q_1}\right)^x \tag{4}$$

where C_1 is the known capital investment at Q_1 plant capacity, C_2 is the capital investment to be estimated for Q_2 plant capacity, and x is the scale factor.

Oil production per well pair was fixed and thus the total capital investment for well pair and well pad construction changed when production rates were varied. Additionally, the supply cost was calculated at different capacities to see the impact of economies of scale on bitumen production cost.

2.4.5. Sensitivity and uncertainty analyses

The present study conducts a techno-economic assessment based on fixed values of the process and economic parameters. However, these input values may vary. The variability in input parameters brings uncertainty to the supply cost. To identify the input values that influence the supply cost, sensitivity analysis was performed. The DCFA model was incorporated in the RUST tool, developed by Di Lullo et al. [70], which assess the sensitivity and uncertainty of the model outputs. The RUST tool performs sensitivity analysis through the Morris method. The Morris method evaluates the supply cost by changing the input values within their range of variability. The input values were populated in a uniform distribution within the range given to each parameter. The Morris method recognizes the most sensitive parameters of the model through the mean and standard deviation of the supply cost. Once the most sensitive parameters were identified, uncertainty analysis was carried out for those parameters. To evaluate the uncertainty in the results, we ran a Monte Carlo simulation in the RUST tool. The Monte Carlo simulation randomly designates values to the parameters and recalculates supply costs. Those values are assigned from the probable ranges of economic parameters in the market and found the literature. Table 2.4 shows the input information for the sensitivity and uncertainty analyses.

Table 2.4: Input ranges for RUST tool

Input	Range
Capital investment [MCAD]	405 – 766.5
Discount rate [%]	9 – 15
$ Electricity [CAD/GJ]	15.85 - 43.41
$ Natural gas [CAD/GJ]	1.87 - 9.43

\$ Propane [CAD/m^3]	130- 176
\$ Diluent [CAD/bbl]	47.25 - 78.75
\$ Transportation [CAD/bbl]	6.75 - 11.25
\$ Operating labor [CAD/hour]	28.5 - 47.5

2.5. Results of economic assessment

2.5.1. Supply cost

Table 2.5 presents the economic breakdown of the dilbit produced using the solvent extraction process. The supply cost of dilbit is estimated to be C$62.2/bbl. Solvent make-up takes the largest share of the supply cost, contributing 34%. The need for solvent make-up is a result of solvent losses in the reservoir and CPF. The losses in the reservoir account for 96% of the overall solvent loss. For the base case scenario, we assumed solvent loss in the reservoir to be 20% of the injected value [18]. This value is based on Nsolv Corporation's bitumen extraction project field test results, and it may vary from one reservoir to another [18]. Reservoirs with the potential for high solvent loss are a major economic concern with the solvent-based process. Solvents such as propane or butane are expensive, and high solvent loss in the reservoir will significantly impact the production cost of dilbit. A reservoir solvent lost of 5% (instead of 20%), for example, will lower the cost of dilbit to about C$45/bbl, which is economically more attractive than the base case (C$62.2/bbl). Increasing solvent loss in reservoirs is not beneficial. Methods such as blowdown and injection of non-valuable gases for solvent recovery can be implemented to reduce solvent loss in reservoirs. This method could reduce reservoir solvent loss to about 15% [18]. Solvent losses in the CPF are due to unrecovered solvents in the OTU and SPU. Most of these losses (62%) occur in the SPC, leaving at the top of the column with non-condensable gases. The remaining 38% is unrecoverable from the bitumen mixture. However, they help to lower the amount of diluent needed to meet pipeline specifications.

Diluent and transportation costs also contribute a significant share to the supply cost, about 30%. This cost includes the purchase of diluent and the pipeline tariffs to transport the dilbit to Cushing, Oklahoma from Edmonton, Alberta. Since the bitumen produced is a partially upgraded oil [18], it requires less diluent for transportation than the bitumen produced from steam-based extraction. For example, producing a typical SAGD bitumen increases the diluent requirement by 13% vol. and the supply cost by 15%. A partially upgraded bitumen also increases the transportation capacity of bitumen in pipelines. Thus, the partial upgrading of the solvent-based methods represents not just a potential increase in the selling price of bitumen but also a cost reduction in the diluent and transportation cost. It is important to mention that although upgrading improves oil quality, the precipitated asphaltenes can plug the permeable channels and lower oil production [71]. Special attention must be paid to asphaltene precipitation management to ensure successful oil extraction.

The capital investment makes up 10.6% of the supply cost. The construction of well pairs, well pads, and gathering lines have the highest share at 84% of the fixed capital investment (FCI). Although these investments are comparable with a typical SAGD project, the CPF investment differs since the process requires a solvent purification unit (SPU), a solvent vaporization and injection unit (SVI), and some modification in the oil treatment unit (otu). Compared to SAGD, solvent-based bitumen extraction CPF is less expensive. Solvent-based extraction benefits from a lower CPF cost because it does not require expensive steam generators and has a smaller plant size. Steam generators are the heart of steam-based extraction processes like SAGD. More importantly, the steam-based process uses a large amount of water, which increases throughput and consequently the size of SAGD's CPF. Solvent-based bitumen extraction's CPF investment can be C$170 million less than SAGD's. Other operating cost parameters include depreciation, research and development, maintenance and repairs, plant overhead, administration cost, laboratory charges, and clerical labor. They account for 23% of the supply cost. Depreciation and maintenance and repairs are the major component with a 50% share of these "other" operating costs. Utilities have a smaller contribution to the Nsolv supply cost.

Table 2.5: Distribution of supply cost (Base case)

Parameter	**CAD per bbl**
Fixed capital investment	5.6
Electricity (grid)	0.3
Natural gas	1.1
Solvent make-up cost	21.3
Diluent and transportation	18.7
Other operating cost	14.3

Sensitivity and uncertainty analyses

The Morris plot in Figure 2.2 identifies sensitive parameters that may impact the supply cost of dilbit produced from solvent-based bitumen extraction process. The supply cost is sensitive to solvent loss in the reservoir, capital cost, transportation cost, discount rate, propane (solvent) price, diluent price, and natural gas price. These parameters are sensitive and thus show relatively large mean and standard deviations on the Morris plot. Electricity price and operating labor are less sensitive; their mean and standard deviations are close to zero. With a large mean value, solvent loss in the reservoir is the most sensitive parameter. To understand how the change in input parameters impacts the supply cost, we performed uncertainty analysis. The result show uncertainty values from C$48.2/bbl to C$63.7/bbl. As expected, solvent loss has significant impact on the supply cost of dilbit. Solvent loss contributes about 76.1% to the total variability in the supply cost.

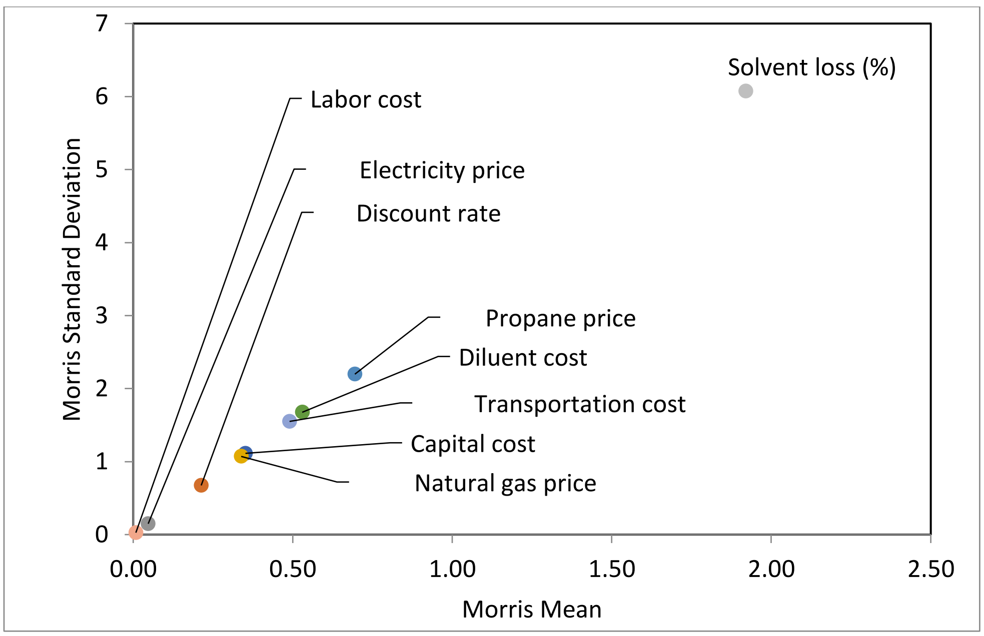


Figure 2.2: Morris sensitivity plot for solvent-based bitumen extraction technology

2.5.2. Solvent purification unit using a refrigeration system and high-pressure separators

Reducing solvent loss in the CPF can also improve the economic performance of Nsolv technology. As mentioned earlier, the largest share of solvent loss in the CPF is in the SPU. The SPU is also the second most capital-intensive unit in the CPF. To address these issues, we investigated two alternative solvent purification pathways developed by Soiket et al. [35] and Safai et al. [72]. The first, pathway I, sets out to reduce solvent loss in the CPF using a refrigeration system to condense solvent while the non-condensable gases separate. The second, pathway II, reduces the solvent purification unit size. It uses high-pressure separators to remove solvent impurities. The impurities are mainly non-condensable gases mixed with solvents in the vapor phase. In both pathways, the impurities are used as fuel in heaters. Figure 2.3 shows the process flow diagram of both pathways. In pathways I and II, the supply cost increased by 5.8% (C$65.8/bbl) and 2.9% (C$64.0/bbl), respectively, from the base case. Pathway I reduces solvent loss in the CPF significantly (by 50%) compared to the base case. However, the required capital and operating investment to achieve this reduction is high. Its fixed cost investment (FCI) increased by 13.0% (59 million CAD) and electricity consumption by 15%. On the other hand, pathway II had a low capital investment, 15% less than the base case. It does not require an expensive purification unit. However, solvent loss in the CPF is four times the base case value. This value is high because high-pressure separators are not as effective as using a distillation column, which is used in both the base case and pathway I.

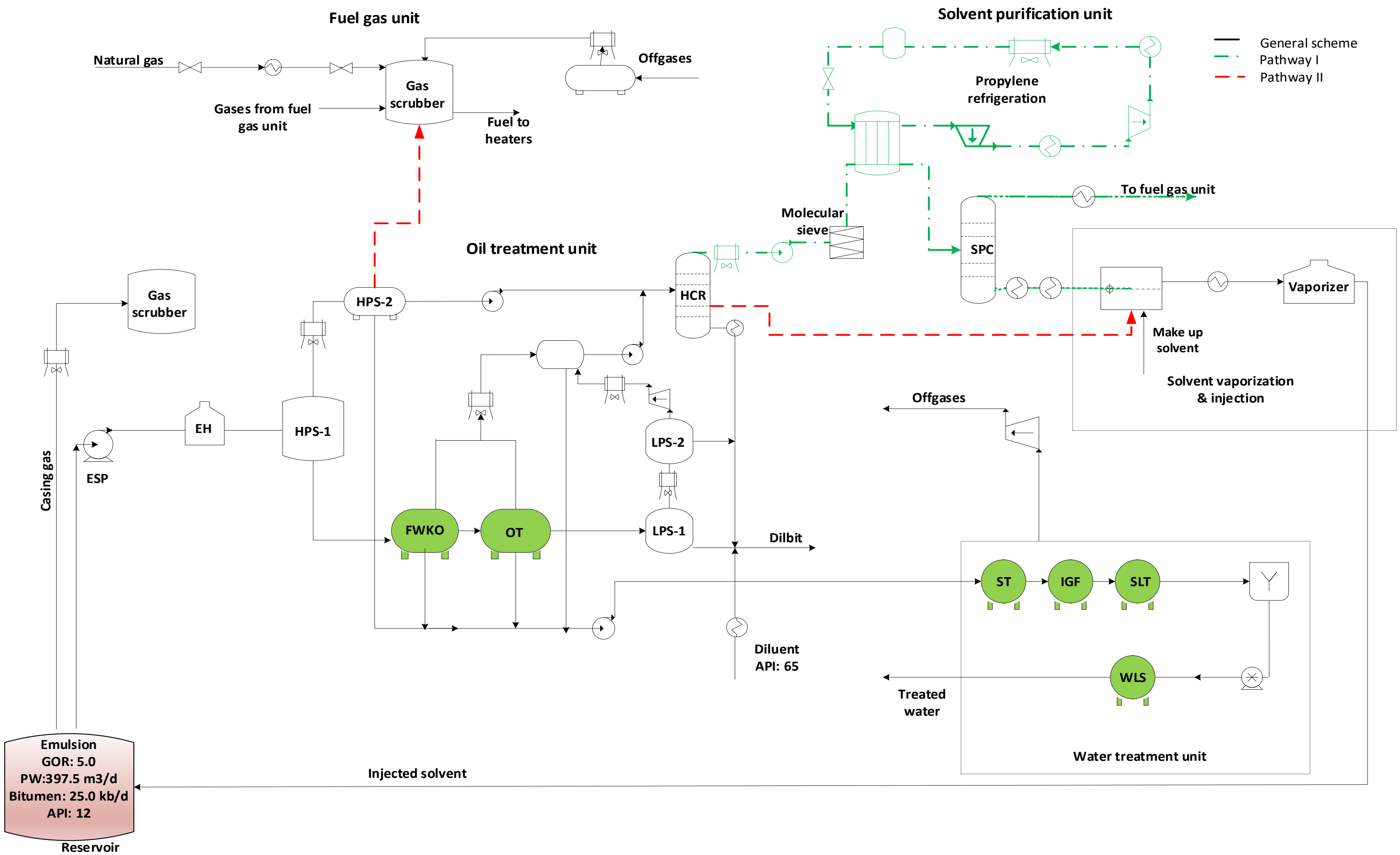


Figure 2.3: Process flow diagram for pathways I and II

2.5.3. Effects of solvent selection

Solvent choice depends on reservoir conditions. Butane is preferred for shallow reservoirs where the solvent can condense at lower pressure and temperature. For this case, the supply cost was evaluated by modeling the CPF using butane as the solvent. The production rates were the same as with propane, including the SOR, produced gases, and connate water. Some of the operating conditions were modified to optimize the separation and purification of butane. The supply cost increased by 28% with butane. The solvent make-up cost had the greatest effect on cost in the solvent's application. The amount of solvent lost in the CPF was similar for both propane and butane. Propane loss in the CPF was 3.7% (6.52 stdm^3/h) and butane loss was 3.8% (6.45 stdm^3/h). However, butane is more costly than propane and thus increases supply costs. Other significant cost differences between propane and butane use were in the natural gas consumption and the capital investment. The natural gas cost increased by 20% with butane. Butane has a higher molecular weight and requires more energy to be heated. Thus, the vaporizer and SPC's reboiler needed, respectively, 62% and 37% more heating energy when butane was used.

The total capital investment increased by 11% when butane was used. The CPF required larger equipment such as fired heaters and distillation columns compared to the propane case. Other expenses such as electricity and "other" operating costs had insignificant cost implications for the shallow reservoir application. The only cost reduction observed with butane was in the diluent and transportation costs. The amount of butane dissolved in the bitumen was slightly higher than in the propane case; therefore, the bitumen blend required less diluent. Thus, for the butane application the diluent and transportation costs were favorable and lowered by 3%. For comparative purposes, pathways I and II were modeled with butane. Similar trends were observed for the three CPF designs using butane, including similar solvent losses and increments in capital investment and energy consumption. Figure 2.4 shows the cost breakdown of the supply cost using propane and butane as solvent for the three designs.

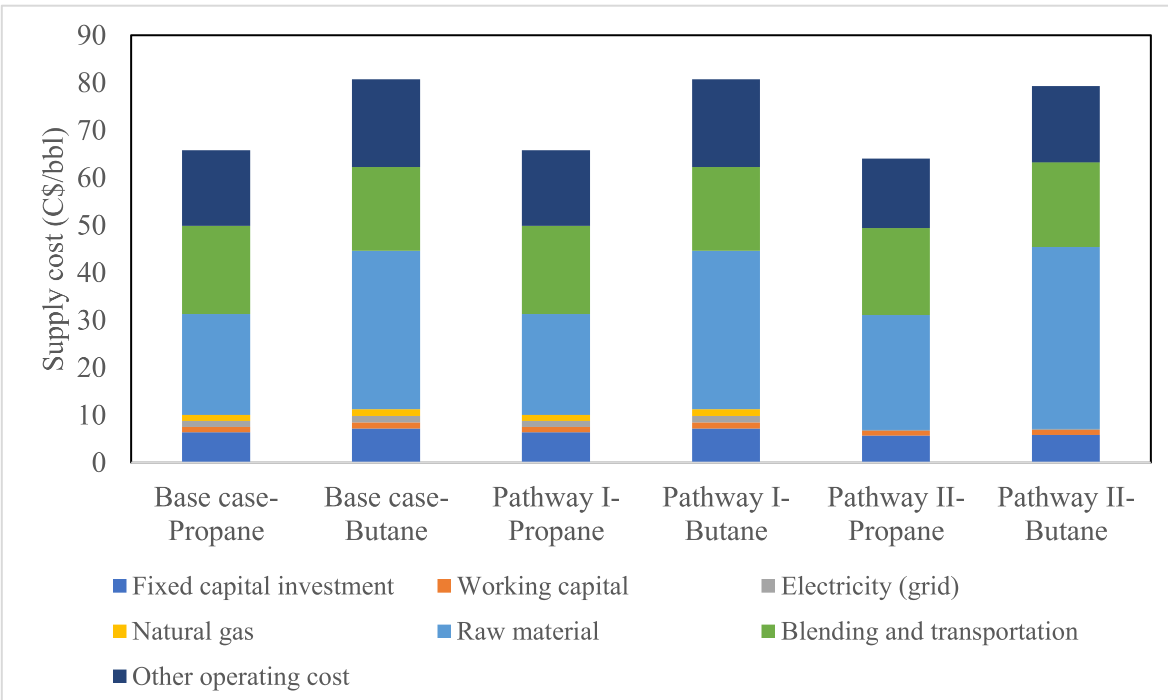


Figure 2.4: Supply cost for the shallow reservoir application

2.5.4. Effect of plant capacity on the Nsolv supply cost and economies of scale

The economies of scale were investigated by calculating the specific capital cost investment when the plant capacity changes. For comparative purposes, pathways I and II were included in the analysis. As illustrated in Figure 2.5, we calculated scale factors of 0.72, 0.74, and 0.67 for the base case and pathways I and II, respectively. Since these scale factors are less than 1.0, the capital investment per barrel of dilbit from solvent-based bitumen extraction process decreases when plant capacity increases.

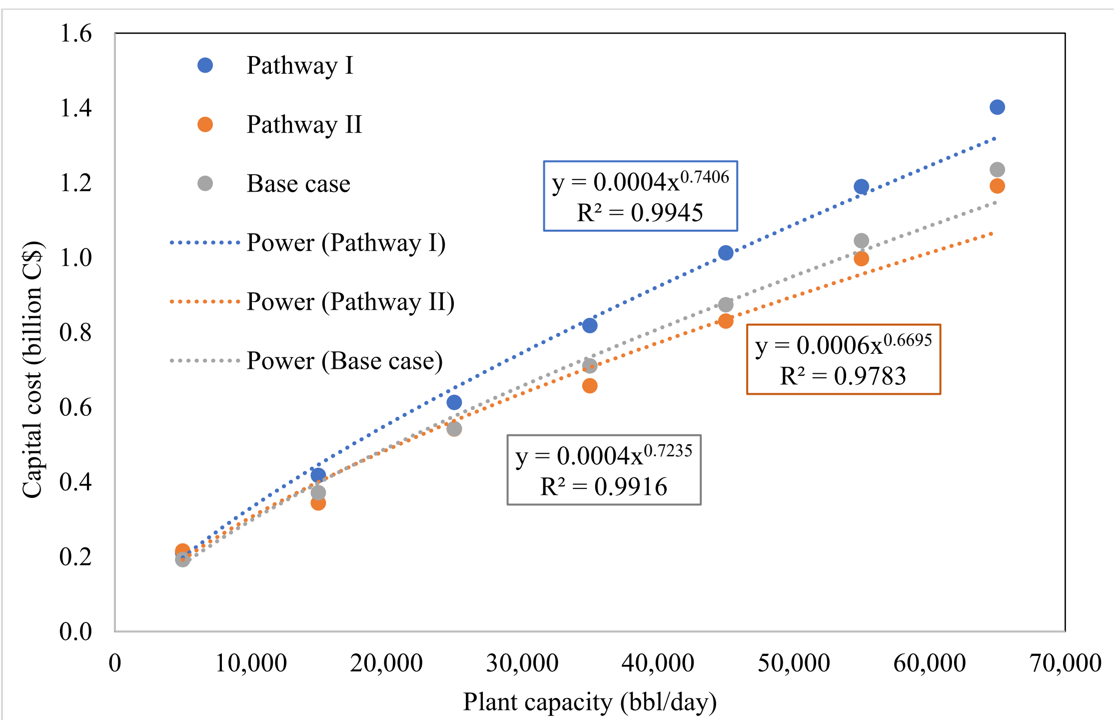


Figure 2.5: Plot of capital cost versus plant capacity

The effect of the plant capacity on dilbit supply cost was also investigated. The results are shown in Figure 2.6. As shown in the graph, the supply cost of solvent-based bitumen extraction process can be C$99/bbl and C$57/bbl when plant capacity is increased to 10,000 bpd and 65,000 bpd, respectively. The supply cost reduction with increased plant capacity is due to economies of scale, wherein the fixed costs are spread across more barrels of dilbit produced, and the reduced specific capital cost investment. This result suggests that bigger oil sands projects have an economic advantage over small plants. For instance, a plant of 12,000 bpd such as the Hangingstone (Athabasca Oil Corporation) and a plant of 60,000 bpd such as the Sunrise (Cenovus) would have supply costs of around C$93/bbl and C$59/bbl, respectively.

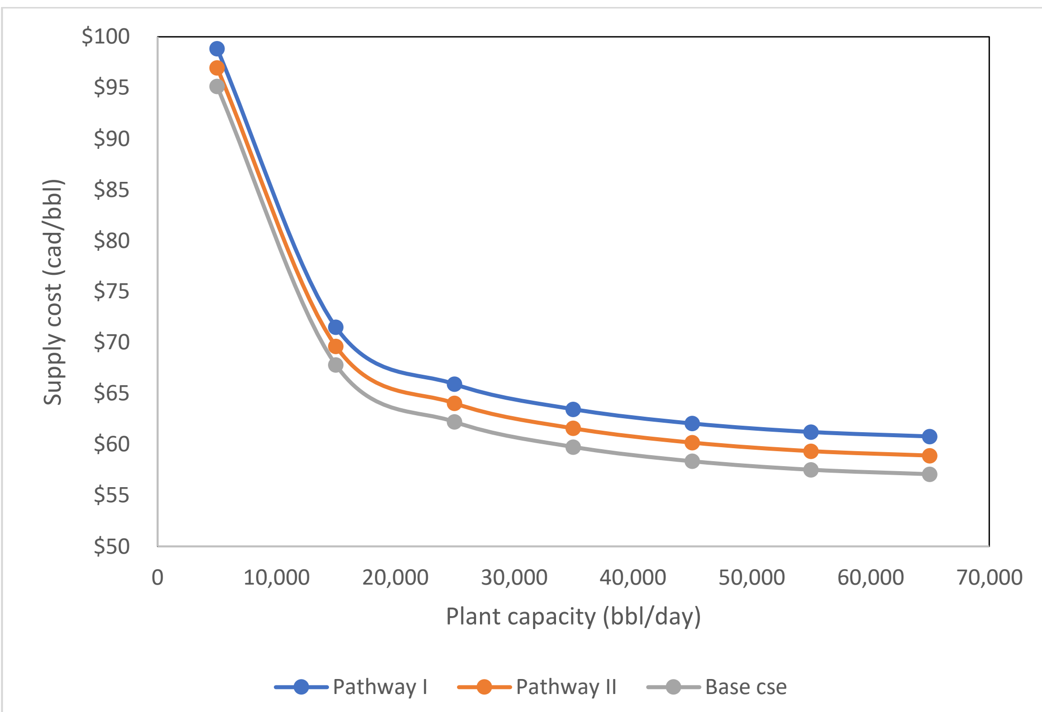


Figure 2.6: Effects of plant capacity on supply cost

2.5.5. The performance of solvent-based bitumen extraction technology

The solvent-based bitumen extraction process is a solvent-based method that changes the bitumen yield. As mentioned before, the properties of bitumen from this process improve because of the in situ partial upgrading of the crude. For instance, the bitumen produced by SAGD is equivalent to 0.9 barrels of transportation fuel, while this process yields 1.1 barrels [18]. For this reason, a straightforward comparison cannot be made between the dilbit produced through different methods of extraction. To address the differences in bitumen properties, the supply cost is linked to an equivalent West Texas Intermediate (WTI) price. The WTI is the oil benchmark widely used as a reference for buying and selling crude oils worldwide [73]. The expected selling price of the dilbit produced by solvent-based bitumen extraction can be estimated with a discount from the WTI price. The discounted price will depend on the quality of the crude as well as the market at a specific geographic location. The oil benchmark commonly used for diluted bitumen in Canada is the Western Canada Select (WCS). The WCS is a heavy oil blend of about 20 crude streams that incorporates bitumen, diluent, and synthetic crude oil (SCO) [21]. In the last decade, WCS has been discounted on an average of 26% from the WTI price [42]. A reasonable approach is to assume that the dilbit from solvent-based bitumen extraction could be sold at around the same discount as WCS. However, dilbit from solvent-based bitumen extraction differs in sulphur content and its supply cost was calculated in this study at a different geographic location. WCS is traded at Hardisty, Alberta, Canada, and the supply cost of bitumen from solvent-based extraction process includes transportation to Cushing, Oklahoma, USA. In addition, the sulphur content of dilbit solvent-based bitumen extraction is expected to be lower than for WCS. For this reason, the discount is expected to be lower for dilbit from solvent-based bitumen extraction process. A discount of 20-26% was assumed to assess the economic viability of dilbit from solvent-based bitumen extraction based on the forecasted WTI prices. The US Energy Information Administration (EIA) forecasted an average WTI price of US $58.89/bbl for 2021 [74]. Assuming a WTI price of about US$60/bbl for the entire period of a project, the selling price for dilbit from solvent-based bitumen extraction process would range from C$55.5/bbl to C$60/bbl. This range overlaps with the probable range of the supply cost of the dilbit produced by solvent-based bitumen extraction. Thus, based on expected future oil prices, solvent-based bitumen extraction technology is expected to be economically viable.

2.6. Conclusions

In this study, we developed a process simulation model of solvent-based bitumen extraction technology to assess its economic performance. The supply cost of dilbit from solvent-based bitumen extraction was estimated. A production capacity of 25,000 barrels per day was considered. Solvent recovery and purification are important unit operations of the solvent-based bitumen extraction process. The purification process was modeled using a high-temperature distillation column. Because of the variabilities in the input parameters, uncertainty in supply cost was evaluated. The uncertainties in supply cost range from 48.2-63.7 C$/bbl. Solvent loss in the reservoir makes up the largest share of the solvent-based bitumen extraction cost. Reducing the solvent losses to 5% could decrease the supply cost of dilbit to C$45/bbl. Thus, solvent recovery is an important factor for the economic competitiveness of solvent-based bitumen extraction technology. Diluent and transportation costs are reduced since the diluent requirement is 13% lower than SAGD's. In situ partial upgrading reduces the diluent requirement and potentially increases the selling price of dilbit. The results also show a cost benefit to solvent-based bitumen extraction technology for high plant capacities because of economies of scale. The scale factor is 0.72. Compared to current and short-term forecasted crude oil prices, dilbit from solvent-based bitumen extraction supply costs are economically viable. The findings of this study are useful for policy and investment decision-making on solvent-based bitumen extraction technology.

Chapter 3

The development of a techno-economic models for the assessment of the Effective Solvent Extraction Incorporating Electromagnetic Heating technology

3.1. Introduction

Oil sands are abundant in Canada and vital to its economy [75]. About 80% of the oil sands in Canada is found deep beneath the earth and can be recovered by in situ methods [75]. Thermal processes such as steam-assisted gravity drainage (SAGD) and cyclic steam stimulation are the most commonly used in situ recovery methods for oil sands extraction [76]. These processes inject heated steam into the reservoir. The reservoir temperature rises, reducing the oil viscosity and improving the mobility of the oil into the producer well. Although these extraction methods have been successfully implemented, the cost of producing bitumen is high compared to conventional crude oil production. These thermal processes are also greenhouse gas (GHG) emissions-intensive because steam production requires a high amount of fossil fuel, i.e. natural gas. The GHG emissions, together with wastewater disposal, are key environmental challenges for these processes [77], [78]. From an economic perspective, oil sands projects require large upfront investments with long payback periods [79]. According to Coylar et al. [55], on average, Athabasca dilbit is sold at 27% below West Texas Intermediate (WTI). Assuming a WTI price of US$50 per barrel and considering the cost of blending and transportation, bitumen producers receive around $26 per barrel. Higher rate of return is required to cover oil sands production costs [55]. An earlier study concluded that to obtain a rate of return of 10% in a greenfield SAGD project, the price of WTI must be around US$53 per barrel [43]. Thus, the current low oil prices along with environmental concerns discourage investments in oil sands projects. In response, oil sands industries are seeking new extraction methods that are economically and environmentally efficient. Some proposed methods include SAGD hybrids, in situ combustion, solvent-based injection, and electromagnetic heating [80].

Studies on solvent-based extraction methods show them to be promising because they significantly lower the onsite combustion of fossil fuels and water use and treatment [81], [82]. These methods make use of hydrocarbons such as propane, butane, or pentane that reduce oil viscosity. Using these solvents causes partial upgrading in situ, thus improving bitumen quality [83]. Heated solvent-based methods such as N-solv use the latent heat of solvent condensation to heat the bitumen [15]. The limitations of this concept are that it requires a large amount of solvent and a high degree of solvent purity, and it has a low production rate [19]. To address these drawbacks, Harris Corporation proposed Effective Solvent Extraction Incorporating Electromagnetic Heating (ESEIEH) [15]. ESEIEH uses an electromagnetic device that preheats the reservoir, thus eliminating the dependence on condensation heat and therefore the strict requirement of solvent purity. The preheating facilitates the diffusion of the solvent in the oil, producing an emulsion of solvent, bitumen, and water at a better production rate than other solvent-based methods [84]. The emulsion is sent to surface facilities where bitumen is recovered. Since hydrocarbon solvents are expensive, they are recovered and purified using a technology other than SAGD.

Many studies have investigated various aspects of the ESEIEH process. In 2015, Harris Corporation in collaboration with Devon Canada, Suncor Energy Inc., and Nexen Energy ULC, commissioned a pilot plant using ESEIEH and reported that the ESEIEH process can reduce almost 50% of GHG emissions when the electric grid is used and more than 65% using cogeneration compared to SAGD [19]. Mohsen et al. [72] developed a process model to evaluate the GHG emissions of ESEIEH. Their results showed that the ESEIEH process has the potential to reduce the GHG emissions in bitumen extraction. Other studies investigated technical aspects such as well completion [85], processing facility [86], antenna layout optimization [33], reservoir heating through an electromagnetic device [30], phase behavior and growth of the solvent chamber [29], and asphaltene precipitation [87]. These studies suggest the viability of ESEIEH for bitumen extraction.

Although various aspects of the ESEIEH process have been explored, there are very few studies on its economic feasibility. Spence et al [37] evaluated the supply cost of a small-scale ESEIEH process. Their study shows that the bitumen from the ESEIEH process is cost-competitive with the currently used oil sands methods. However, they conducted a high-level analysis that provided no details on economic feasibility. Besides, a small-scale plant can undermine the ability of a technology at its full economic potential. Plant operating conditions, scale factor, and other economic parameters that might impact the economics of the ESEIEH process were not investigated. Understanding these parameters can provide information for effective decision-making. This thesis research aims at addressing these gaps.

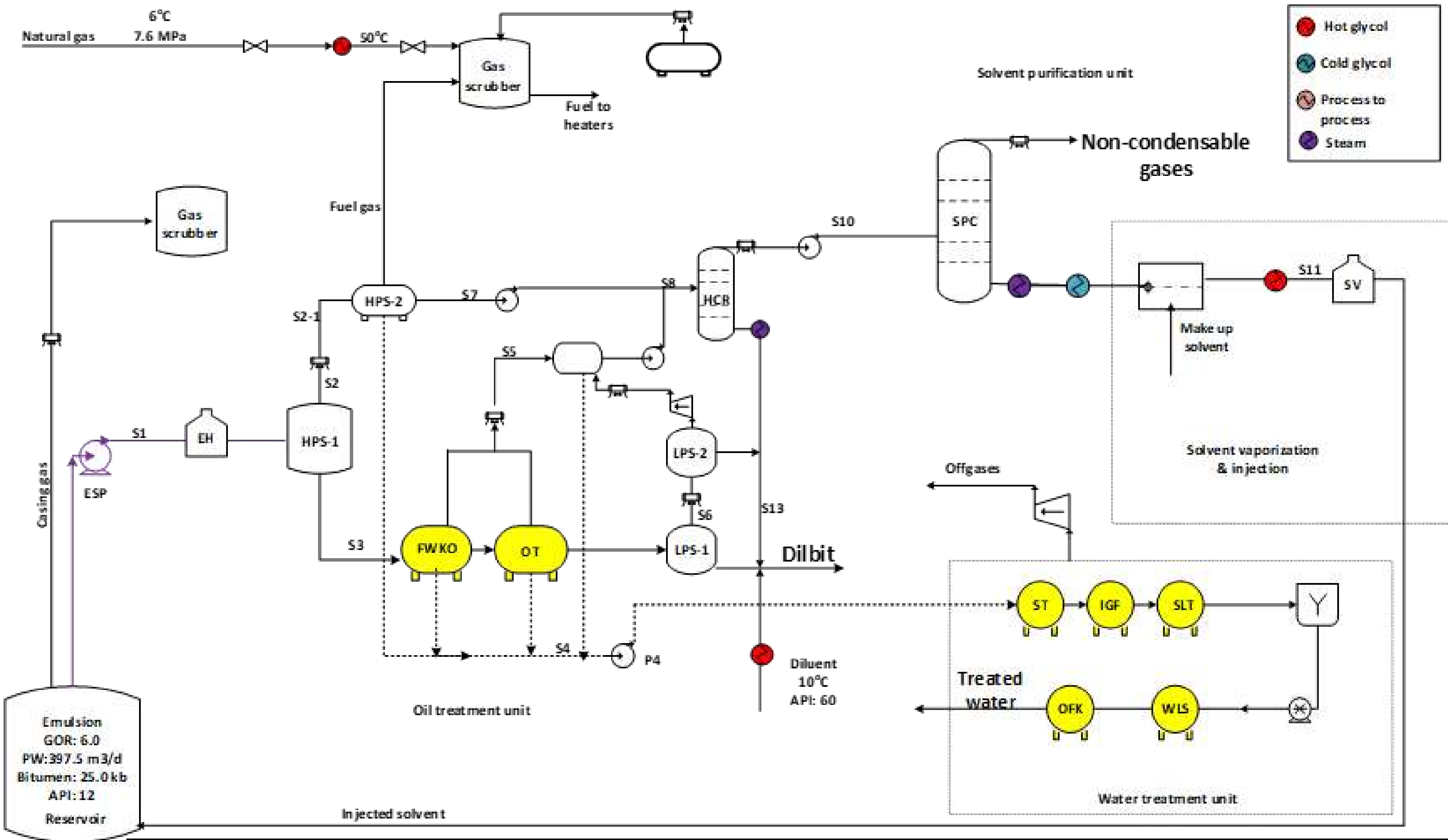
The overall objective of this study is to develop a techno-economic model to evaluate the economic feasibility of the dilbit produced from the ESEIEH process. A process simulation model was developed to determine material input and output, energy consumption, and equipment size. The supply cost of dilbit, as well as the transportation cost and rate of return, was estimated through discounted cash flow analysis. The scale factor was also determined. Sensitivity and uncertainty analyses were conducted to identify key parameters that may impact the process cost. The specific objective was accomplished through the following specific objectives:

- Developing a process simulation model for the ESEIEH technology.
- Determining the supply cost of dilbit produced by the ESEIEH technology.
- Identifying the key economic parameters that influence supply cost through comprehensive sensitivity and uncertainty analysis
- Developing a scale factor for the ESEIEH process.

3.2. Process Description

3.2.1. Extraction

Figure 3.1 is a schematic diagram of the ESEIEH process. ESEIEH is an extraction method that uses solvents assisted by electromagnetic heating and gravity. Because the process is gravity-assisted, the configuration of the wells resembles SAGD [15], [84]. This configuration consists of two horizontal wells drilled one above the other about 5.0 m apart [19]. The upper well is the injector well where electromagnetic heating is transferred to the reservoir. The electromagnetic heating is generated and delivered by a radio frequency device (RFD). The RFD consists of the radio frequency transmitter (RFT), transmissions lines, and radio frequency antennas. The RFT generates the radio frequency current that is sent to the antennas through the transmission lines. The antennas are located inside the injector wells. In the first stage of the process, the antenna creates an electromagnetic field that pre-heats the connate water and the bitumen in the injector well. The pre-heating stage significantly reduces the viscosity of the bitumen and facilitates the subsequent diffusivity of the solvent. Once the bitumen reaches a temperature of 40-70 °C (depending on reservoir conditions) [84], the antenna frequency is reduced and a vapor solvent is injected into the injector well [19], [37]. The injected solvent creates a vapor chamber in the reservoir. The temperature in the solvent chamber is slightly higher than the saturation temperature at reservoir conditions in order to maintain the solvent in a vapor state but also maximizes diffusivity in the pre-heated bitumen [88]. Thus, the reservoir conditions determine the most appropriate solvent to use. For deep reservoirs, propane is recommended, and in shallow reservoirs, butane [18]. The base case considers propane; however, the shallow reservoir is investigated as an independent scenario to assess the impact on the supply cost. Both organic solvents can dilute the bitumen, further decreasing its viscosity and allowing it to flow.



EH: Emulsion Heater; ESP: Electrical submersible pump; FWKO: Free-water knockout; HCR: Hydrocarbon recovery column; HPS: High pressure separator; IGF: Induced gas flotation; LPS: Low pressure separator; OT: Oil treater separator; SLT: Slop tank; SPC: Solvent purification column; ST: Skim tank; WLS: Warm line softener

Figure 3.1: A proposed process model for the ESEIEH process

The emulsion of bitumen, connate water, and solvent is drained into the lower well assisted by gravity. The lower well is the producer well from which an electro submersible pump (ESP) pumps the emulsion to the surface. The gases are produced through the casing of the well. These gases are known as casing gases and are mainly composed of methane (CH4) [89].

3.2.2. Separation process

The emulsion, once on the surface, is taken to the oil treatment unit where multiphase separation is carried out. The temperature of the emulsion is increased before it enters a flash tank with the help of a fired heater, also called an emulsion heater (EH). The flash tank reduces the pressure of the emulsion and releases the solvent, light hydrocarbons, and non-condensable gases su83ch as CH4 and carbon dioxide (CO2) dissolved in the emulsion. These gases are collected at the top of the flash tank and the liquid emulsion is sent to the free-water knockout (FWKO) tank, a three-phase separator that separates gases and water from the bitumen. The bitumen is subsequently sent to the oil treater separator (OT) for further separation. The bitumen is then stored and blended with diluent (dilbit) in the dilbit storage tank before being sent through the pipeline. The gases recovered in the flash tank are cooled and sent to a hydrocarbon recovery (HCR) column. The HCR column separates hydrocarbons heavier than the solvent used. Thus, the solvent and non-condensable gases are produced as distillate, while the heavy hydrocarbons are produced as the bottom outlet stream. The heavy hydrocarbons are sent to the dilbit storage tank to be mixed with the bitumen. The solvent and non-condensable gases are sent to the solvent purification unit [86].

The water from the bottom of the FWKO is sent to the de-oiling unit, where the water-oil emulsion is broken, and the recovered oil is sent back to the oil treatment unit. In the de-oiling unit, the emulsion is pumped to the skim tank that uses gravity and retention time to separate the oil droplets from the water phase. The water recovered from the bottom of the skim tank is then treated in the induced gas flotation (IGF) unit where bubbles of gas are injected to help separate the solids and remaining oil droplets. It is subsequently sent to the slop tank for final separation. The water is then pumped to the water treatment where warm lime softening (WLS) reduces the amount of dissolved minerals and removes the remaining oil droplets to meet the quality specifications.

The casing gases from the well are sent to the fuel gas system that cleans and dries the gases to be sequentially used as a source of energy in the plant. The off-gases from the de-oiling unit are sent to the vapor recovery unit (VRU), where the retained liquid is separated from the light hydrocarbons in the vapor phase. The liquid is sent back to the oil treatment unit and the gases are sent to the fuel gas system.

3.2.3. Solvent purification

To recycle the solvent, non-condensable gases are removed in the solvent purification unit. The impure solvent is compressed, cooled, and refined in a solvent purification column (SPC) to remove the non-condensable gases as distillates. The bottom product, liquid solvent, is sent to a storage tank where make-up solvent is added to meet the required SOR. The composition and purity of the solvent required for this process depend on reservoir conditions and solvent [18]. Finally, the solvent is heated in a solvent vaporizer (SV) and reinjected to the reservoir to continue the cycle.

3.3. Process simulation

We developed a process simulation model in Aspen HYSYS for the ESEIEH process. The Peng-Robinson equation of state was used as the fluid package. The Peng-Robinson equation of state allows an adequate representation of the phase equilibrium behavior of hydrocarbon mixtures [90].

3.3.1. Material balance

In this work, petroleum naphtha with a density of 45.4 lb/ft3 was used as diluent for the extracted bitumen to meet pipeline specifications [55]. The dilbit density for pipeline specification was taken as 19 API [55]. The mass and energy balances allow the sizing and cost estimation of the equipment required in the processing facility. The ESEIEH plant is modeled to handle an emulsion of 25,000 bbl/day of bitumen, 9,142 std.m3/d of solvent, 2500 bbl/day of water, and 0.70 MMScf of non-condensable gases. The oil production rate was evaluated using the equation proposed by Nenniger et al. 2008 [58] for solvent-based extraction methods assisted by gravity. Equation 1, from Hyne et al., was used to determine the well-pair production rate and the number of wells required [28].

$$m = 43{,}550 \left(\frac{k\emptyset}{\mu}\right)^{0.51} \qquad (5)$$

Here, m is the oil mass flux, ∅ is the porosity of the reservoir, k is the permeability and μ is the oil viscosity in centipoise. Table 3.1 shows the production parameters used in the model. The bitumen density is assumed to be 12 API (984.3 kg/m3). The emulsion was produced at a pressure of 2.40 MPa and a temperature between 45°C and 60°C.

Table 3.1: Production parameters

Parameter	**Unit**	**Value**	**Reference**
Solvent-to-oil-ratio	m^3/m^3	2.3	[14]
Solvent hold-up in the reservoir	%	20.0	[24]
Solvent loss in the reservoir	%	5.0	[72], [35]
Gas-to-oil-ratio	m^3/m^3	5.0	[29]
Water-to-oil ratio	m^3/m^3	0.1	[30]
Raw bitumen viscosity, μ,	cP	2,377,340	-

The aquathermolysis reaction occurs in thermal processes such as SAGD. The reservoir oil reacts with steam at high temperature (200^0C - 300^0C) to produce reservoir gases. The reservoir gases produced through aquathermolysis are CH4, CO2, H2S, etc. [91, 92]. Since ESEIEH operates at a relatively low temperature (40^0C -70^0C), the aquathermolysis reaction is expected to be significantly reduced. For this reason, the reservoir gases were excluded in this study and only the gas dissolved in the bitumen was considered. This gas is known as solution gas and its composition was assumed to be methane.

3.3.2. Energy analyses

The energy requirements of the entire extraction, separation, and re-injection process are divided into requirements for electricity and natural gas. The pumps, compressors, cooling fans, and the radio frequency transmitter use electricity to operate.

The equations used for calculating the electric power requirement are shown below:

Centrifugal compressor:

$$p_{compressor} = \frac{(P_{out} - P_{in}) * m_{gas}}{\eta_c} \tag{6}$$

Pump:

$$p_{pump} = \frac{(P_{out} - P_{in}) * m_{liquid}}{\rho * \eta_p} \tag{7}$$

Here, m is the flow rate, ρ is the liquid density (kg/m^3), η is the efficiency of the equipment (%), and P_{out} (kPa) and P_{in} (kPa) represent the inlet and outlet pressures, respectively.

Radio frequency transmitter:

$$p_{RFT} = cEOR * bbl \tag{8}$$

Here, cEOR is the cumulative electricity-to-oil ratio and bbl is barrels of oil produced in the period to be analyzed.

Other equipment such as reboilers and heaters require steam generated through the burning of natural gas. The amount of natural gas required ($\frac{kg}{h}$) is given by:

$$Natural\ gas\ = \frac{Heat_{required}[kW]}{LHV * \eta} \tag{9}$$

where LHV is the lower heating value of natural gas ($\frac{kWh}{kg}$) and η is the boiler efficiency (%). The power required by fans was estimated using the simulation model. Table 3.2 presents the input data of the energy requirement.

Table 3.2: Input data for energy balance

Parameter	**Unit**	**Value**
Solvent input temperature	°C	60
Solvent input pressure	kPa	2,400
Re-injection pressure	kPa	1,700-2,500
cEOR	GJ/bbl	0.1
Radio frequency transmitter efficiency	%	72

Prime efficiency	%	75
Boiler efficiency	%	80
LHV natural gas	MJ/kg	46

3.3.3. Electricity source

The ESEIEH process relies significantly on electricity [72]. For the base case, electricity from the power grid was assumed. However, it is common for conventional plants to generate their own electricity through onsite cogeneration plants. Mohsen et al. [72] reported that the use of cogeneration electricity could reduce GHG emissions by 25%. For this reason, a cogeneration plant scenario was investigated to observe the impact of cogeneration on the supply cost of ESEIEH. A natural gas combined cycle (NGCC) plant was added to the economic model. The combined cycle reuses the exhaust gas for further steam generation and thus increases the overall efficiency of the electricity generation plant [93]. Based on a 702 MW plant, a capital cost of C$1,310/kW, fixed operating cost of C$14.7/kW-year, and variable operating cost of C$4.7/MWh were taken from capital cost estimates by the U.S. Energy Information Administration [94].

3.4. Techno-economic Assessment

3.4.1. Capital cost

For the initial investment estimate of the ESEIEH process, the equipment was sized and quoted. The total capital cost is the fixed capital investment (FCI) in addition to the working capital. For the FCI, the construction of a greenfield central processing facility (CPF), the drilling and completion of the wells, well pad construction, and gathering lines were considered. The assumed values of these parameters are presented in Table 3.3.

Table 3.3: Capital cost input parameters (million Canadian dollar, MCAD)

Parameter	Value	Reference /comment
Drilling and completion cost per well-pair	3.5	[35]
Radio frequency device cost per well-pair	9.4	[36]
Cost per well pad	25	[35]
Cost of gathering lines	84	[35]

The CPF equipment size and cost were evaluated using developed process model. The cost reported by Aspen ICARUS includes direct, indirect, and working capital costs [95]. Working capital is assumed to be 15% of the total capital cost and, for this reason, working capital was recalculated to include the costs incurred in the construction of wells, well pads, and gathering lines [57].

To calculate the total capital investment of well-pair and well pad construction, the number of production wells is needed. As mentioned earlier, Equation 1 was used to calculate oil production per well-pair and thus the number of well-pairs needed to produce 25,000 bbl/day of bitumen. Table 3.4 shows the reservoir and well conditions considered.

Table 3.4: Reservoir conditions

Parameter	Unit	Value	Comments/Remarks
Reservoir pressure	kPa	1,500	[96]
Reservoir initial temperature	°C	10	[96]
Porosity	%	30	[96]
Permeability	D	1	[96]
Well length	m	800	[19]

3.4.2. Operating costs

Operating costs are the costs incurred on a day-to-day basis. These costs are known as the cost of manufacturing (COM). The COM includes utilities, maintenance, operating labor, raw material, operating supplies, and other direct costs. Indirect costs of the process – administration and plant overhead – are included. Electricity and natural gas prices were taken from the Canada Energy Regulator forecast for the industrial sector of the province of Alberta, Canada [60]. The blending cost and transportation cost were also included. The model developed by Aman et al. [59] was used to evaluate the transportation cost. The dilbit was transported 4,124 km from Edmonton (Alberta, Canada) to Cushing (Oklahoma, USA). Table 3.5 and Table 3.6 list the input information and the assumptions made in the COM calculation, respectively.

Table 3.5 Input information for COM calculation

Parameter	Unit	Value	Reference
Electricity from the grid	C$/GJ	16.8	[37]
Natural gas	C$/GJ	2.8	[37]
Propane	C$/gal	0.58	[38, 39]
Butane	C$/gal	0.71	[40]
Blending cost	C$/bbl of diluent	63	[6, 41]
Transportation cost	C$/bbl dilbit	9	[59]
Operating labor	C$/h	38	[43]

Table 3.6: Cost of manufacturing calculation

Direct supervisory/Clerical labor	18% of Operating Labor	Direct Cost
Maintenance and repairs	6% of FCI	Direct Cost
Operating supplies	0.9% of FCI	Direct Cost

Laboratory charges	15% of Operating Labor	Direct Cost
Depreciation	10% of FCI	Fixed Cost
Plant overhead costs	70.8% of operating Labor + 3.6% of FCI	Fixed Cost

3.4.3. Supply cost of dilbit

To evaluate the economic performance of ESEIEH, the supply cost was calculated. The supply cost is defined as the average price required to cover the operating cost and capital expenses while earning a specific rate of return during the lifespan of the project [43]. The supply cost is calculated through discounted cash flow analysis (DCFA). DCFA is used to calculate the present value of an investment based on future cash flows. The main equations are shown in Equation B1-B2 (Appendix) from Williams [67], assuming 2% inflation (as per the Bank of Canada) [97].

We calculated the supply cost for the different scenarios through the DCFA. The supply cost is calculated at a nominal discount rate of 12%. The average capacity of the plant is assumed to be 95%.

3.4.4. Sensitivity and uncertainty analysis

Several assumptions were made during the techno-economic assessment. We assumed a fixed value of the model inputs. The variabilities in the inputs can impact the output results, creating a level of uncertainty. The sensitivity analysis and uncertainty analysis are important parts of the modeling processes since they evaluate the uncertainty of the model outputs. The sensitivity analysis allows us to identify the input parameters that contribute the most to the uncertainty of the outputs. The Morris method was used to perform a global sensitivity analysis, wherein the mean and standard deviation of the samples were evaluated. The Morris method allows us to evaluate the model outputs by changing the values of the input parameters within a given range one step at a time [98]. The samples were assessed at specific points selected based on a uniform distribution of the input values. Subsequently, the uncertainty analysis was performed using a Monte Carlo simulation that quantified the uncertainty of the results according to the most sensitive input parameter given in the sensitivity analysis. The Monte Carlo simulation assigns random values of the model inputs and recalculates the model outputs to give the most probable range (in this case, of the supply cost) [70]. The sensitivity and uncertainty analyses were performed using RUST [70] integrated with the DCFA model. For both analyses, the economic inputs were considered within their most probable ranges found in the literature and from market data provided in Table 3.7.

Table 3.7: Input parameters for sensitivity and uncertainty analyses

Input	Range	Reference
Capital investment [MCAD]	757 - 1,262	-
Discount rate [%]	9 – 15	-
Electricity [CAD/GJ]	15.85 - 43.41	[37]

Input	Range	Reference
Natural gas [CAD/GJ]	1.87 - 9.43	[37]
Propane [CAD/m^3]	95.22 - 371.7	[38, 39]
Diluent [CAD/bbl]	47.25 - 78.75	[6, 41]
Transportation [CAD/bbl]	6.75 - 11.25	[59]
Operating labor [CAD/hour]	28.5 - 47.5	[43]

3.5. Results and Discussion

3.5.1. Material and energy balance results

The amount of diluent required was 728.8 std m^3/d for a final dilbit production of 4,660 std m^3/d (29,311 bbl/day). The diluent accounted for 15.6% of the dilbit volume shipped through the pipeline. Its addition to bitumen affects the cost and transportation capacity in the pipeline.

At the reservoir pressure of 1,500 kPa, the solvent reached optimum diffusion conditions at 40°C. We assumed that the antenna maintains this temperature in the reservoir. At these conditions, the viscosity of raw bitumen decreased from 2,377,340 cP to 1144 cP, thus keeping oil production per well at 92 m^3/day and the required number of well-pairs to produce 25,000 bbl/day of bitumen at 44.

Tracking the amount of solvent in the process is necessary to identify areas of solvent loss and required make-up. Table 3.8 shows the material balance of the solvent in the ESEIEH process. In the oil treatment unit, most of the solvent was recovered in the flash tank with 79% of the propane sent to the solvent purification unit and 21% recovered in the HCR unit. About 552 kg/h of propane cannot be recovered from the bitumen stream. 313 kg/h was produced as a bottom product in the HCR; this represents 26% of the solvent losses in the CPF. In the solvent purification unit (SPC), the propane removed together with the non-condensable gases accounted for the remaining 74% of the losses in the CPF. These gases were used as fuel. The total solvent make-up requirement, including the solvent requirement for the assumed losses in the reservoir, was 12,633 kg/h of propane (597 std m^3/d), and the final solvent purity was 99.95%.

Table 3.8 Solvent mass balance

Parameters	**Reservoir**	**Oil treatment unit**	**Solvent purification unit**
Propane mass input [kg/h]	193382.0	183712.9	182848.3
Propane mass output [kg/h]	183712.9	182848.3	180749.1
Propane losses [kg/h]	9669.1	865.0	2099.2
Propane purity [%]	-	98.9%	99.9%

Figure 3.2 shows the energy consumption breakdown in the ESEIEH process. The process required 198 GJ/h in heating duty and 114 GJ/h in electric power. The off-gases accounted for 66% of the total fuel requirement with an LHV of 47 MJ/kg. The remaining 34% was supplied by purchased natural gas. The SV (solvent vaporizer) accounted for the highest amount of heat energy (75 GJ/h). The EH consumed 56 GJ/h. For electricity, the RFT required 91% of the total electric power in the ESEIEH process, the ESP 5% and the CPF 4%. The operating conditions of main equipment are shown in Table 3.9.

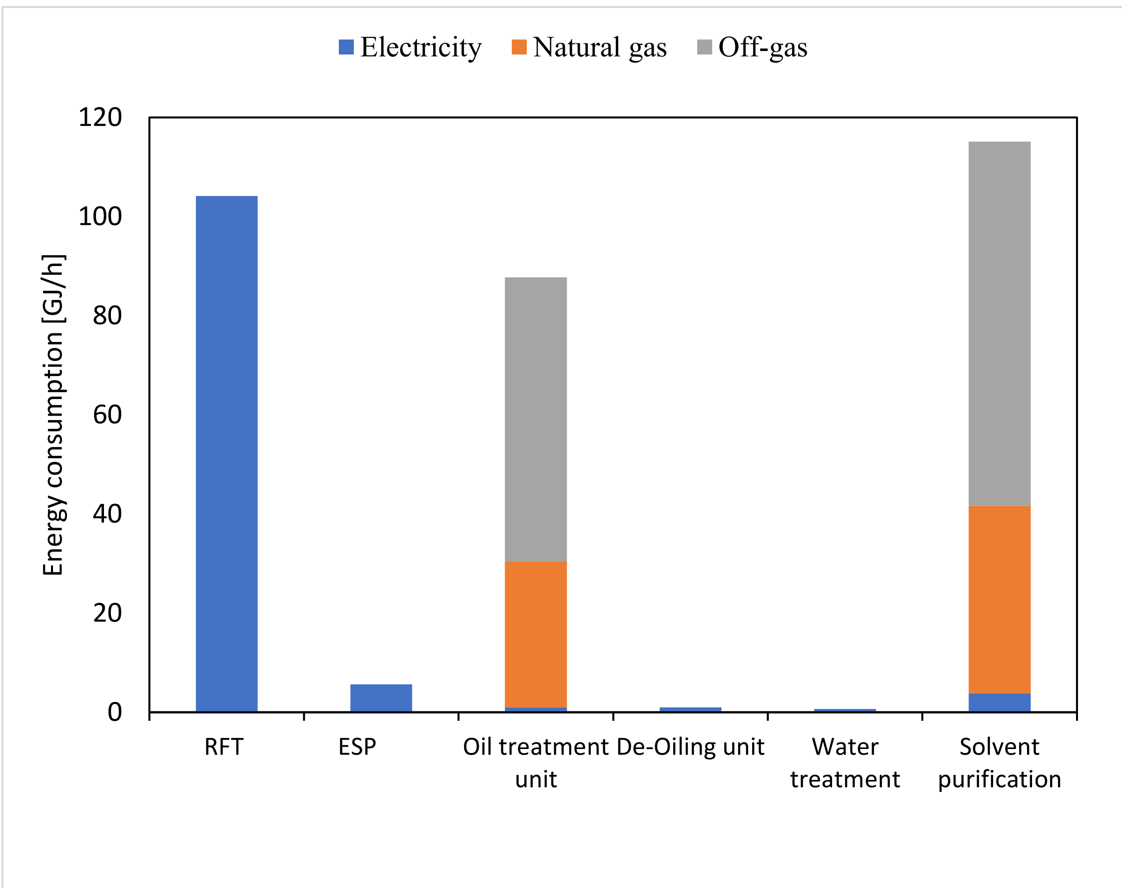


Figure 3.2: Energy consumption of the ESEIEH process.

Table 3.9: Operating conditions of principal equipment

Parameter	Value
FWKO temperature [°C]	87
FWKO pressure [kPa]	1,100
OT temperature [°C]	86
OT Pressure [kPa]	1,020
Distillation Column (HCR)	
Inlet temperature [°C]	40
Inlet pressure [kPa]	1,495
Number of trays	22
Compression System (Solvent purification unit)	
Inlet pressure [°C]	1,355
Outlet pressure [kPa]	3,550
Distillation Column (SPC)	
Inlet temperature [°C]	46
Inlet pressure [kPa]	3,500
Number of trays	22

To understand the impact of process parameters the energy consumption, we performed a sensitivity analysis using the developed process model. Process parameters, i.e., the SOR, GOR, WOR, cEOR, boiler efficiency, and pump efficiency, were changed from their base case values, as shown in Table 3.10. The most sensitive parameter was the cEOR, followed by boiler efficiency and the SOR. Increasing the cEOR from 0.1 to 0.18 increases electrical power consumption by 73%. For the boilers, reducing their efficiency to 60% increases fuel demand by 33%. For the SOR, increasing their efficiency from 2.3 to 2.9 increases the total energy consumption by 23%. Other parameters like the GOR, WOR, and prime mover efficiency do not impact the total energy significantly.

Table 3.10: Process parameters ranges in process model in the case study

Parameter	Base value	Start value	End value	Step size
WOR	0.1	0.08	0.15	0.03
cEOR	0.1	0.08	0.18	0.04
SOR	2.3	1.73	2.9	0.39
GOR	5	3.75	6.25	0.83
Pump efficiency [%]	75	56	94	12.67
Furnace efficiency [%]	80	60	100	13.33

3.5.2. Supply cost

For the base case, the price of dilbit should be minimum C$56.84/bbl for the ESEIEH project to be economically viable. The total capital investment was C$1,009 million and the COM was C$456 million/year. The COM was further divided into energy cost, blending and transportation cost, and other operating costs to understand how they impact the supply cost.

Table 3.11 presents the breakdown of the supply cost. The capital investment, utilities (natural gas and electricity), blending and transportation cost, and other operating costs account for 21.6%, 3.0%, 33.0%, and 42.4% of the supply cost, respectively.

Table 3.11 Supply cost distribution (including the rate of return).

Parameter	**Cost (C$/bbl)**
Capital investment	12.25
Electricity (grid)	1.56
Natural gas	0.15
Blending and transportation	18.76
Other operating costs	24.12

3.5.3. Operating cost

3.5.3.1. Utility cost

The electricity and natural gas make up the lowest share of the supply cost. The total energy cost was C$17.5 million per year. The electricity requirement represents more than 60% of the energy consumption and 91% of the total energy cost. The RFD is the electricity-intensive unit responsible for 91.3% (C$14.5 million/year) of the electricity cost, followed by ESPs at 4.9% (C$0.8 million/year) and the solvent purification unit at 3.4% (C$0.5 million/year). The other units in the CPF such as the oil treatment unit, de-oiling unit, and water treatment account for only 0.4% (C$0.06 million/year) of the total electricity cost. With respect to natural gas, the amount of purchased natural gas was reduced by close to 35% of the initial natural gas required by using casing gas and off-gas from the plant. The small natural gas requirement in the ESEIEH process along with the lower natural gas purchased amounted to a contribution of 9% (C$1.6 million/year) of the total energy cost and less than 0.3% of the supply cost.

3.5.3.2. Blending and transportation cost

The blending and transportation costs are directly related to the pipeline specification for bitumen. The cost of blending and transporting bitumen can be reduced if the viscosity of the produced bitumen favours pipeline specifications, thus reducing the diluent requirements and increasing the capacity of the transported bitumen in the pipeline. In solvent-based extraction, bitumen viscosity can be improved by precipitating a considerable amount of the asphaltene content using the solvent. Other bitumen properties such as the API and sulfur content are improved alongside, thus bitumen's market value is also increased. To assess the impact of bitumen properties on the supply cost, changes in the quality of the produced bitumen were investigated. The asphaltene content was lowered for the bitumen such that its density was reduced from an API of 12.0 to an API of 19, and the need for diluent was eliminated. Bitumen's heaviest fraction was increased to change the density from an API of 12 to an API of 8.

Bitumen with an API of 19 lowers the supply cost by 7%, while bitumen with an API of 8 increases the supply cost by 6% from the base case. When the density is reduced, a higher amount of propane is trapped in the oil, increasing the flow rate and the heat duty required by the EH and the HCR by 42.6 GJ/h. However, eliminating the diluent cost offsets the increment in energy cost and reduces the supply cost. Bitumen with an API of 8 requires 28% of diluent volume in the dilbit stream, 12.4% more than in the base case. Thus, the cost of diluent will increase by C$127.7 million/year.

The diluent requirement and supply cost have a linear relationship, suggesting that the differences in energy consumption explained above have an insignificant impact on the supply cost. As mentioned earlier, the contribution of energy consumption to the supply cost is small, unlike the blending and transportation cost. Therefore, small differences in energy required to change the quality of the produced oil cannot be seen in the general trend. The linearity represents the fixed price of diluent and transport cost without considering changes in pipeline tariffs or diluent price. Additionally, the total capital investment is not affected when a different oil is produced.

3.5.3.3. Other operating cost

The other operating cost was estimated to be C$247 million per year and it impacts the supply cost significantly. The depreciation, maintenance, and raw material costs were the major cost components driving the operating cost. The depreciation contributed to 35% (C$85 million/year), maintenance to 21% (C$52 million/year) and the cost of raw material to 13% (C$32 million/year) of the operating cost. The depreciation and maintenance cost depend on the FCI, and variation in the capital investment can impact the operating cost. Solvent make-up is an important raw material. The solvent, propane, is expensive, and its losses negatively impact the economy of the process. While minimizing solvent losses in the CPF is possible, losses in the reservoir are sometimes inevitable. Several factors contribute to solvent loss in the reservoir, such as fluid saturation, reservoir pressure and temperature, geological trapping, and reservoir mineral composition [99]. These factors affect the amount of solvent that is dissolved in unproduced oil and reservoir water or trapped in the reservoir due to interfacial forces and solvent that is lost to other formations [99].

3.5.4. Capital investment

The total capital investment was estimated to be C$1.0 billion. Figure 3.3 shows the distribution of the FCI, which represents 85% of the total capital investment. The RFD is the most capital-intensive equipment; it accounts for 48% (C$413 million) of the FCI. The RFD cost is followed by the drilling and completion of the well-pairs at C$154 million and well-pad construction at C$150 million. These upfront investments depend on the number of well-pairs required. Since the project fixes the capacity of bitumen production, the oil production per well-pair is an important factor in this economic analysis. For instance, assuming an increase of 20% of oil production per well-pair, the capital cost and supply cost could decrease by up to C$874 million and C$55.2/bbl, respectively. Other investments that do not depend on the number of well-pairs, such as CPF and propane start-up, only contributed to 6.7% of the FCI.

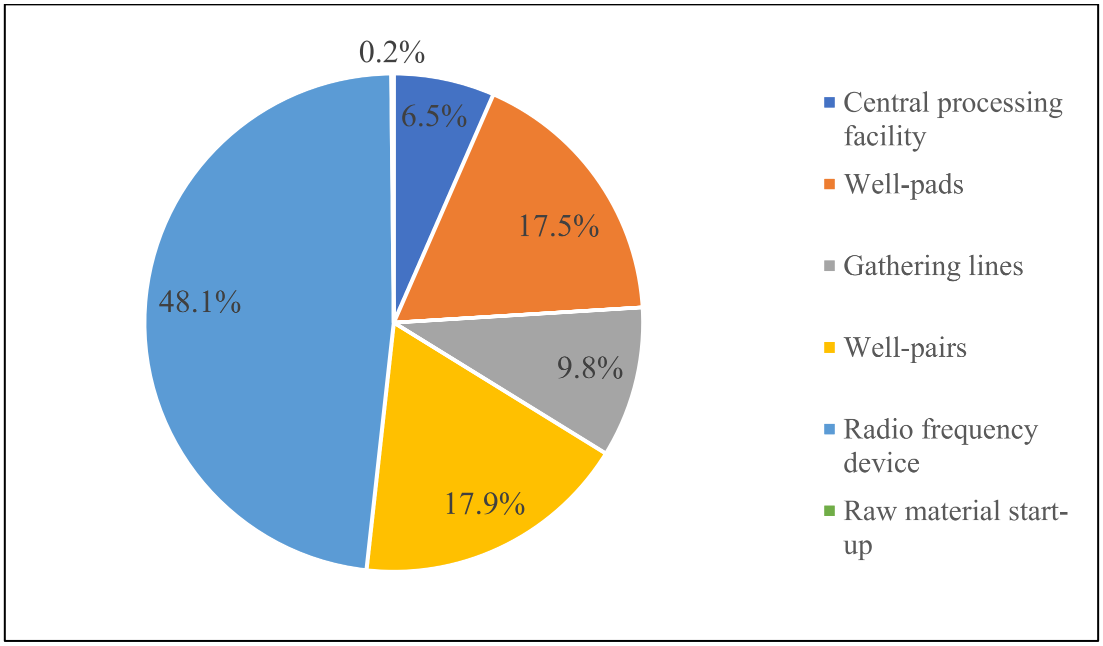


Figure 3.3: Distribution of fixed capital investment

3.5.5. Scale factor

The scale factor provides the relationship between investment cost and plant capacity. These relationships can guide decision-making on plant capacity appropriate for profitability. The ESEIEH plant shows economies of scale with a scale factor of 0.85. This value compares reasonably with 0.78 reported for SAGD by Sapkota et al [100]. The ESEIEH scale factor incentives investments on large-scale plants, especially above 25000 bbl/day. Figure 3.4 shows the effect of plant capacity on supply cost. Increasing plant capacity reduces the supply cost of dilbit because the scale factor of the ESEIEH process is less than 1. The results also suggest that a large plant capacity can drive the supply cost to a favorable competitive oil price in the global market.

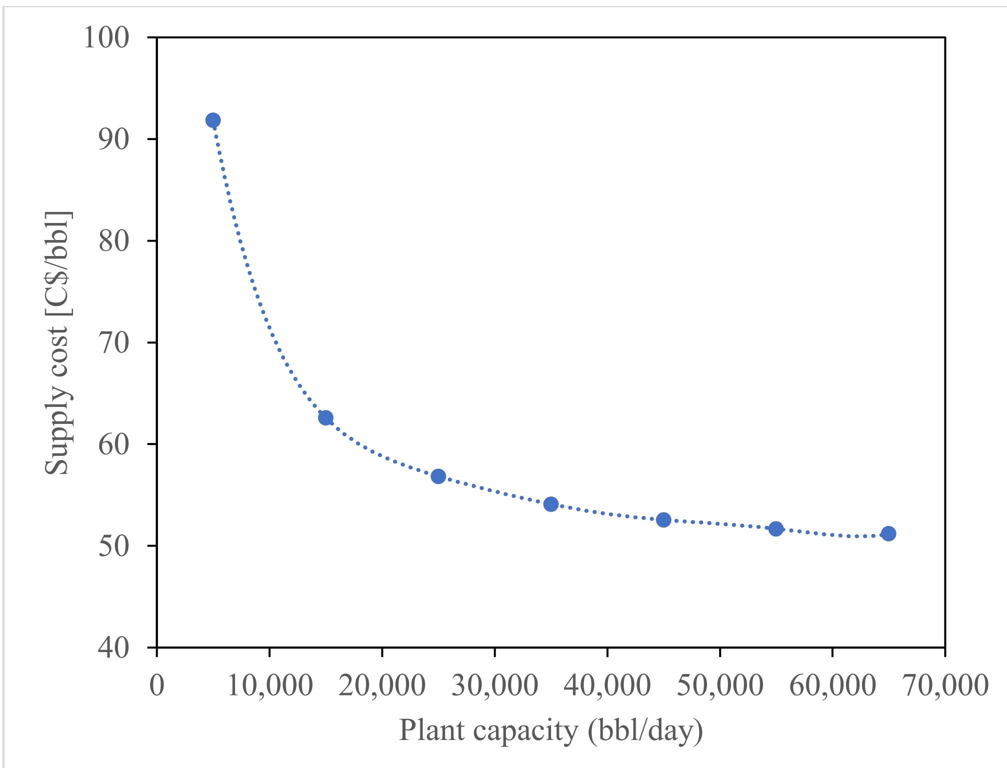


Figure 3.4: Supply cost of dilbit vs plant capacity

3.5.6. The application of ESEIEH in a shallow reservoir

To evaluate the supply cost of dilbit when ESEIEH is applied in a shallow reservoir, we developed a process simulation model using butane as solvent. Figure 3.5 shows the supply costs for deep and shallow reservoirs. For shallow reservoir application, the supply cost is C$1.5/bbl higher than in a deep reservoir (the base case). The solvent make-up is the main difference in the supply cost. Both scenarios have similar solvent make-up requirements; however, as shown in Table 5, butane costs more than propane. This difference in solvent price led to an increment of C$7 million per year in the COM. Additionally, the separation of butane from the emulsion requires more fuel and larger equipment. Although the increment in energy consumption did not significantly impact the COM, the larger equipment size added C$8 million to the capital investment for the shallow reservoir. The increase in capital investment and higher solvent price for a shallow reservoir application made the supply cost higher than for the base case.

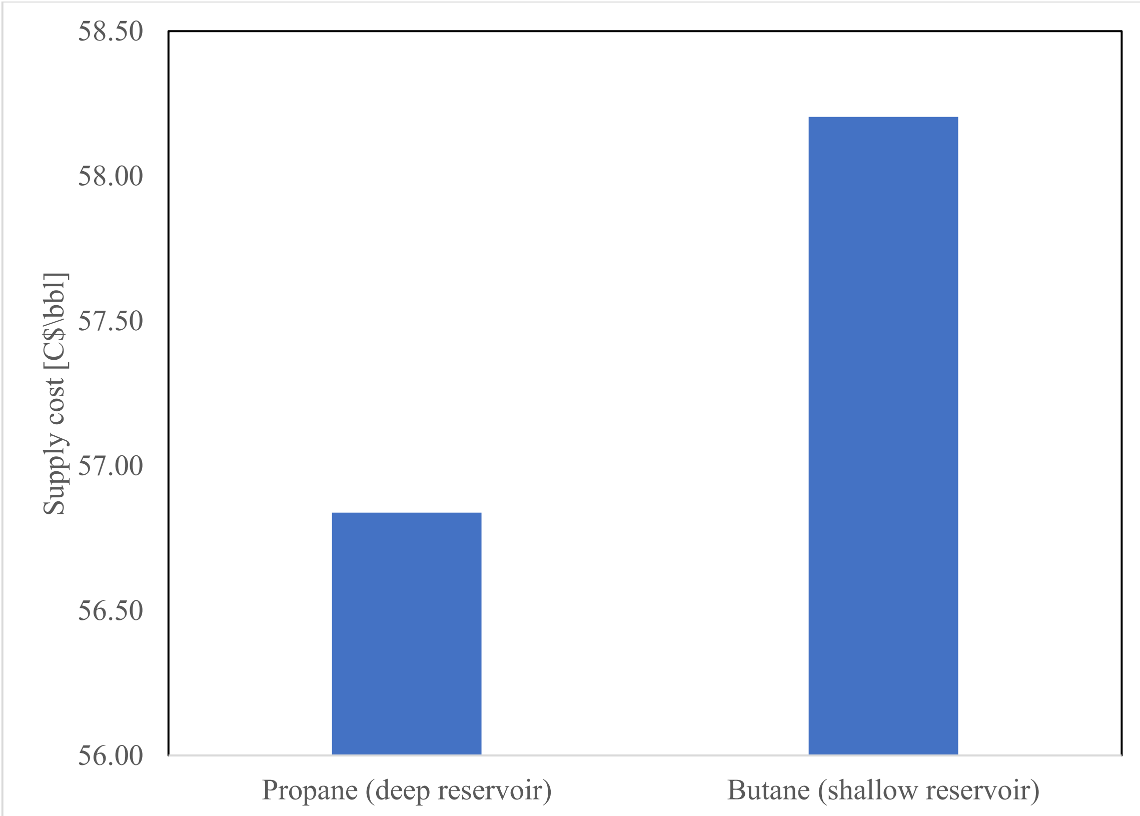


Figure 3.5: Solvent selection impact on supply cost

3.5.7. Electricity supply from cogeneration plant

A 32 MW plant is required to supply electric power to the ESEIEH process. In this plant, total capital cost will increase by 5% (C$50 million) while utilities and operating cost will fall by 51% (C$9 million/year) in utilities and 2% (C$8 million/year), respectively. However, the final supply cost will decrease by 20 cents/bbl from the base case. The small contribution of electricity consumption in the final supply cost significantly reduces the cost benefit of adding a power plant to the CPF.

3.5.8. Sensitivity and uncertainty analyses

Figure 3.6 shows the results of the Morris global sensitivity analysis. The supply cost is sensitive to the price of propane, discount rate, capital investment, diluent and transportation cost, and electricity price. The propane price does not contribute more than 5% of the supply cost; however, its price is influenced by volatile commodities such as crude oil and natural gas [101]. The discount rate is the return to the investors and the inflation for the lifetime of the project. It impacts supply cost depending on the minimum acceptable rate of return of the project. For instance, if investors accept a rate of return below 10%, the project is more likely to meet their expectations and they will be able to execute it. The capital investment is also sensitive, as 21% of the supply cost is required to cover the capital expenses. Several factors can affect the capital investment, such as the cost of the RFD, the geographical location of the project, changes in production parameters that change equipment size, and so on. The cost of diluent, transportation, and electricity also impact the supply cost. Variations in their mean values significantly affect the supply cost. Supply cost is insensitive to the parameters on the lower left side of the graph in Figure 6 (operating labor and natural gas price). These parameters lead to a small change in the supply cost. The insensitive parameters were excluded from the uncertainty analysis since their change will not impact the supply cost significantly.

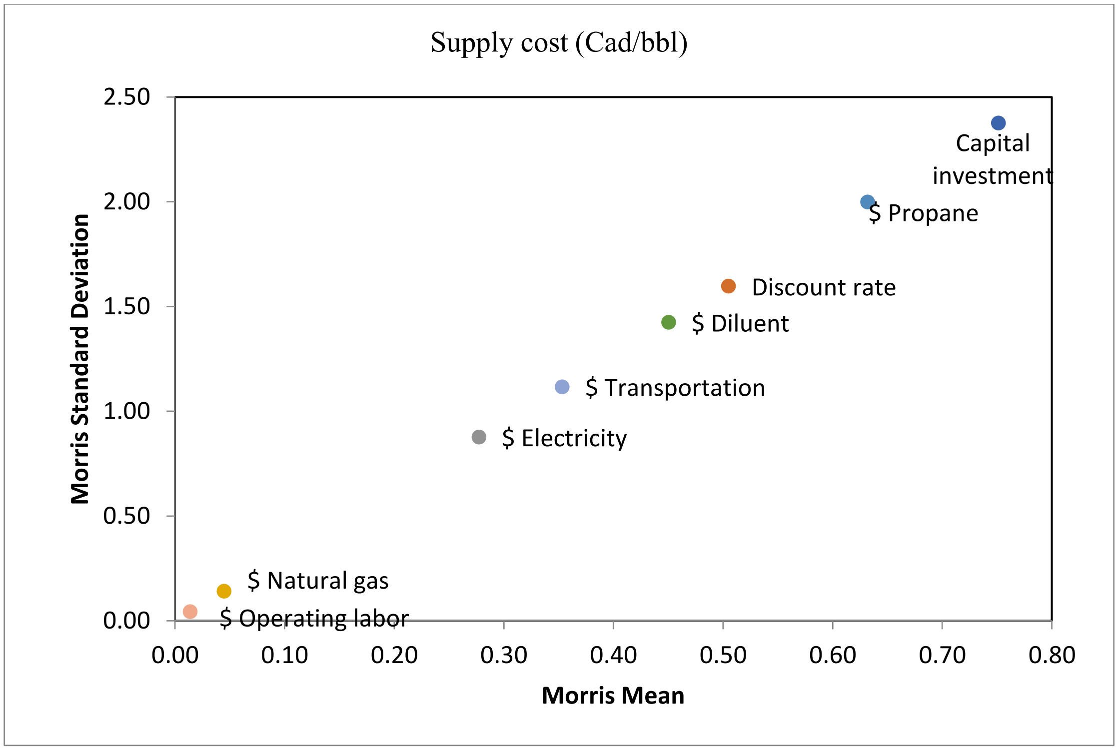


Figure 3.6: Sensitivity analysis of economic parameters

Using the sensitive inputs, we performed uncertainty analysis through a Monte Carlo simulation. Figure 3.7 gives the results of the uncertainty analysis by showing the frequency distribution (percentiles) of the supply cost values in a box diagram. The bottom error bar represents the 10^{th} percentiles, bottom box side the 25^{th} percentiles, the middle of the box the median, the top box side the 75^{th} percentiles, and the top error bar the 90^{th} percentiles. The discount rate and propane and diluent prices led to higher variations in the supply cost when their values were changed within a given range (see Table 3.7). The overall supply cost ranged from C$55.2/bbl to C$64.4/bbl with a median and mean value of C$60.2/bbl and C$59.9/bbl, respectively. It is important to mention that the analysis presents a 90% confidence interval for each variable, and supply cost values outside this range are not probable. This range gives a more realistic insight of the ESEIEH cost performance, allowing the assessment of economic viability based on forecasted oil prices and comparison with other technologies. The range in supply cost by changing each sensitive parameter can be seen in Figure 3.8.

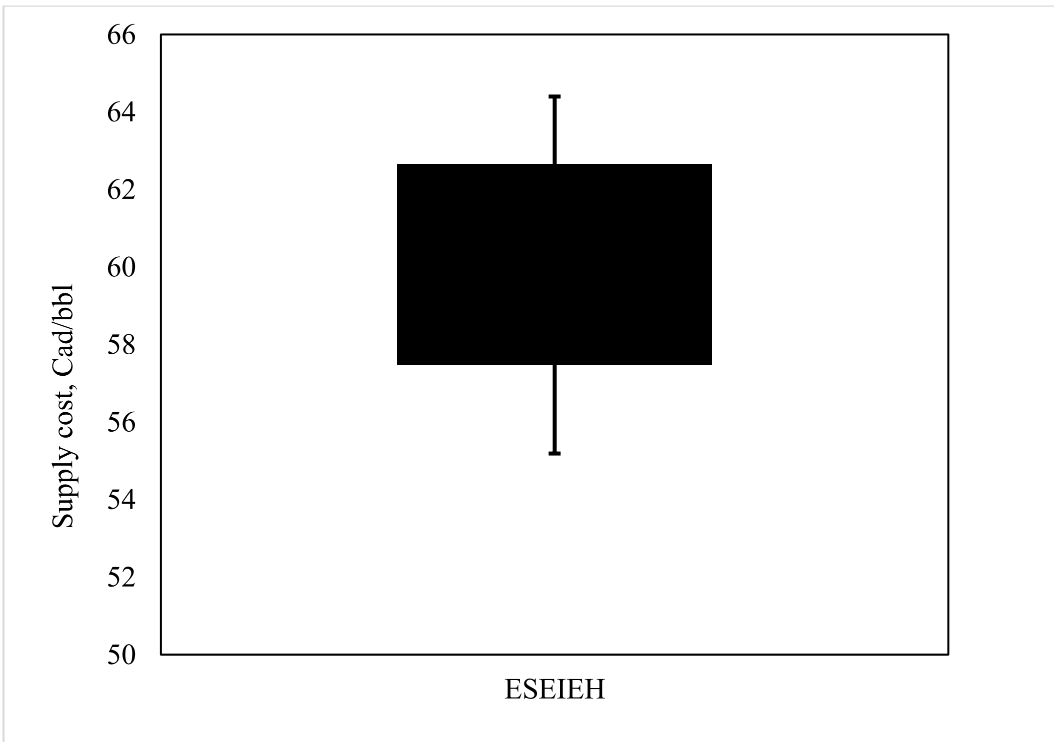


Figure 3.7: Results of the Monte Carlo simulation

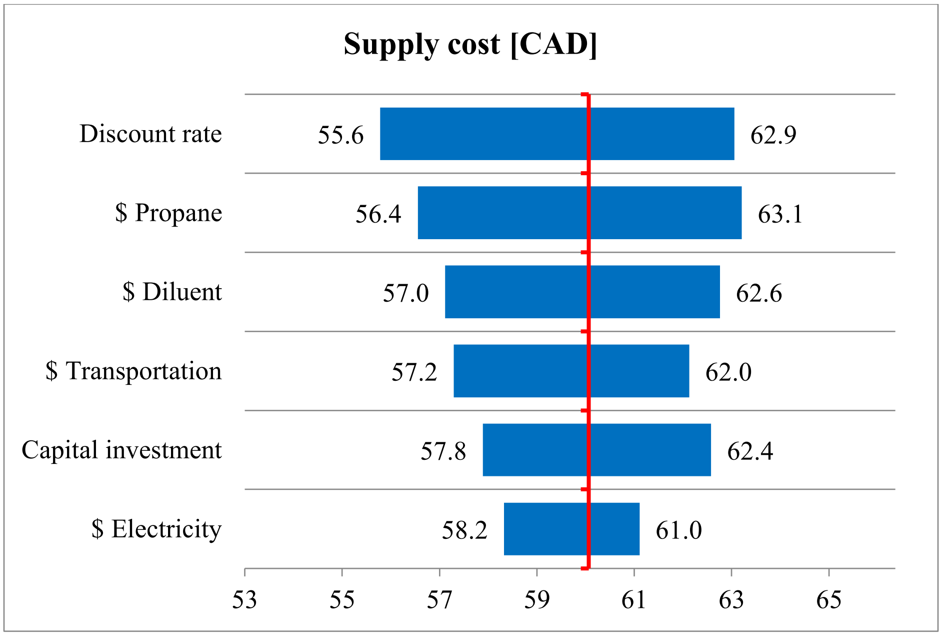


Figure 3.8: Results of uncertainty analysis: Tornado plot

3.5.9. Discussion

Oil price fluctuates over time and depends on factors such as quality, demand, and supply. The market uses benchmark crude oils with specific properties that provide a reference for pricing different types of crudes at specific locations. In Canada, the benchmark commonly used for diluted bitumen is the Western Canadian Select (WCS) [41]. GLJ forecasted an average WCS price of C$56.82 per barrel for the period from 2021 to 2030 [102]. ESEIEH dilbit is economically viable if sold in the range of the forecasted WCS prices. The range of the ESEIEH supply cost covers the WCS selling price and although the mean value (C$59.9/bbl) is above the expected WCS price, several factors could boost ESEIEH cost performance. For instance, the sulfur and metal content of the oil produced by ESEIEH is lower than bitumen produced by SAGD, due to the in situ upgrading. Thus, ESEIEH produces a dilbit of higher quality than WCS and therefore a higher market value. Additionally, the WCS pricing point is at Hardisty (Alberta, Canada) while the transportation cost in this study was assumed to Cushing (Oklahoma, US), the most important oil trade city in USA. The location difference increases the transportation distance for ESEIEH dilbit, leading to additional transportation cost that is not considered in the WCS selling price. It is expected that future construction and expansion of pipelines will reduce the transportation cost and thus improve the ESEIEH supply cost.

Since oil quality is enhanced through ESEIEH, a different approach can be made to calculate the market selling price. The WCS price is set based on a quality discount from the West Texas Intermediate (WTI), the most common oil benchmark in the world. The ESEIEH dilbit price is then calculated with a lower discount from the WTI than the WCS. As mentioned earlier, the WCS is discounted approximately 27% below the WTI. Assuming a discount of 20% for ESEIEH dilbit, the average WTI price for the project to meet economic viability is US$51.80-$58.80/bbl based on the probable ESEIEH supply cost range. Depending on the discount, the market conditions to make ESEIEH technology cost competitive can be more accurately known. Currently, the price of WTI hovers around US$40 per barrel [103]. However, oil price forecasts have estimated an upward trend in future WTI prices because of the increase in oil demand by emerging economies like China's and India's [104]. For the next decade (2021-2030), GLJ predicts a WTI average price of US$58.40 per barrel [102]. Thus, ESEIEH technology is economically viable at expected market conditions.

As described above, the ESEIEH process has the potential to thrive in the oil and gas industry. However, conventional extraction methods such as SAGD have become cost efficient because of deflation in the supply chain cost and reengineering [105]. This cost-efficiency of the conventional methods and technological maturity can negatively impact the ESEIEH attractiveness in the industry. For instance, Sapkota et al. [100] estimated production costs of C$21.60/bbl and C$20.80/bbl for a SAGD plant with and without cogeneration, respectively. Although these costs do not include diluent and drilling cost, they indicate the low production cost of the SAGD process. Millington et al. [43] reported an equivalent WTI price for a 40,000 bbl/day greenfield SAGD project of around US$50 per barrel of dilbit [43]. For comparison purposes, corporate taxes, royalties, and emission costs were included in the economic model and the ESEIEH supply cost was recalculated. The results showed an increase of 12.5% in the supply cost from the base case, reaching C$63.90/bbl and an equivalent WTI price of US$59 per barrel. Although ESEIEH dilbit requires higher oil prices to be profitable compared to SAGD dilbit, the technology aligns with the industry's long-term environmental objectives while being profitable. Likewise, other advantages were identified during this study, in particular the reduced size of the CPF and avoidance of the high capital-intensive equipment used for steam generation such as the once-through steam generator (OTSG).

3.6. Conclusions

In this study, techno-economic models were developed to estimate the supply cost of dilbit produced through ESEIEH. A process simulation model was developed for a capacity of 25,000 bbl/day of bitumen (29,300 bbl/d of dilbit). The supply cost of dilbit was evaluated using the techno-economic model. For 29,300 bbl/day of dilbit, the supply cost is C$56.80/bbl. The supply cost ranges from C$55.20/bbl to C$64.40/bbl when uncertainties in the input parameters are considered. The improvement in bitumen quality through ESEIEH technology reduces the diluent by 12.5 vol% compared to SAGD dilbit. There is potential to decrease the supply cost by up to 7% when the diluent requirement is eliminated. The ESEIEH application for shallow reservoirs increases the supply cost to C$58.20/bbl when butane is used. These results give an insight of the importance of the in situ upgrading of the bitumen and solvent prices in the ESEIEH supply cost. The solvent, diluent, transportation, and electricity costs, as well as the capital investment and discount rate, are sensitive to the supply cost. These parameters should be given adequate attention when determining the economic potential of the ESEIEH process. The ESEIEH process scale factor is 0.85. At this value, the supply cost of dilbit decreases with increasing plant capacity, an indication that large-scale plants are favorable to ESEIEH's profitability. The ESEIEH process is both economically viable and offers long-term environmental benefits. The results from this study show that the ESEIEH process is cost-competitive with conventional oil sands methods and can be considered as a potential replacement for new greenfield projects. The findings of this study will be useful to decision-makers in oil sands industries and to policymakers.

Chapter 4

Conclusions and recommendations

4.1. Conclusions

Solvent-based extraction process, Nsolv and Effective Solvent Extraction Incorporating Electromagnetic Heating (ESEIEH) technologies are promising means of lowering the GHG emissions of oil sands extraction that use solvents instead of steam. This study offers a novel contribution though a techno-economic assessment of Nsolv and ESEIEH technologies at a large scale. This work also developed different pathways for solvent recycling in deep and shallow reservoirs. Process models were developed for Nsolv and ESEIEH technologies to capture their equipment and energy requirements. The models were designed assuming the production of 25,000 barrels of bitumen per day. Techno-economic models were used to assess the economic feasibility of Nsolv and ESEIEH technologies. Techno-economic analysis was conducted to calculate the supply cost of diluted bitumen produced by these technologies. A deep reservoir and a high-temperature distillation column were considered in the base case for both technologies. The supply costs of dilbit produced by Nsolv and ESEIEH are C$ 62.20 and C$ 56.80 per barrel, respectively. The ranges of the supply cost for the Nsolv and ESEIEH processes are from C$48.20/bbl to C$63.70/bbl and C$55.20/bbl to C$64.40/bbl, respectively. These ranges in the supply costs are due to the variability in the input parameters. The supply cost for Nsolv is highly sensitive to solvent losses in the reservoir since it requires a high steam-to-ratio (SOR) compared to ESEIEH. Given the partial upgrading when using solvents, a reduction of around 13 vol% diluent is expected in

Nsolv and ESEIEH compared to SAGD. Thus, the cost of diluent per barrel of oil is reduced and the economic performance of these solvent-based methods is improved. The application of Nsolv and ESEIEH in deep and shallow reservoirs was also investigated. The implementation of these technologies in shallow reservoirs requires the use of butane to ensure condensation at reservoir conditions. Because butane is more expensive than propane, the supply cost increased slightly for shallow application of Nsolv and ESEIEH. Since Nsolv technology is sensitive to solvent loss and has a high capital investment, two additional pathways for solvent recovery were modeled for comparison purposes. Pathway I uses a refrigeration system to condense the solvent and remove impurities in the vapor phase. Pathway II was designed to reduce the plant size by using high-pressure separators to reduce the impurities of the solvent. The supply costs increased by 5.8% and 2.9% for pathways I and II, respectively, from the base case. Considering uncertainty, supply costs range from C$52.10/bbl to C$69.50/bbl and C$48.80/bbl to C$64.60/bbl for pathways I and II, respectively.

The effect of plant capacity on supply cost and economies of scale were also investigated. The scale factors for the Nsolv and ESEIEH technologies are 0.72 and 0.85, respectively. Thus, both technologies exhibit economic benefits at larger plant capacities because of economies of scale. Finally, the supply costs of Nsolv and ESEIEH were compared to expected future oil prices to assess their economic viability. For an average WTI price of C$ 60 per barrel, both technologies could be cost-competitive. The findings of this research can assist policymakers and industry in the decision-making regarding Nsolv and ESEIEH technologies.

4.2. Recommendations for future work

The following recommendations for future work will complement and improve the results of this research:

- A detailed study on the bitumen properties extracted by Nsolv and ESEIEH technologies such as metal and sulphur content will provide useful information on the quality of the products and their selling price.
- Integrating the models developed in this research with reservoir simulation will allow us to capture the effects of solvent recycling in the oil production rates, extraction stability, and other parameters that might influence the cost and energy requirements of the Nsolv and ESEIEH processes.
- The inclusion of the midstream and downstream stages is needed to assess more accurately the capability of Nsolv and ESEIEH technologies to reduce the carbon footprint of oil sands and thus further incentivize investments in these emerging technologies.
- An evaluation of the cost performance of implementing Nsolv and ESEIEH in an existing SAGD facility would give insight into the cost requirements and benefits of extracting remaining oil sands through less energy-intensive methods.
- Incorporating carbon capture technologies to evaluate the cost and GHG emission performance of the Nsolv and ESEIEH technology.

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Appendix A

Appendix A contains supplementary information for chapter 2.

Table A 1. Reservoir conditions

Parameter	Value
Reservoir pressure [kPa]	1,500
Reservoir initial temperature [°C]	10
Porosity [%]	30
Permeability [D]	1
Well length [m]	800

Table A 2. Operating cost assumptions

Direct supervisory/ Clerical labor	18% of Operating labor
Maintenance and repairs	6% of FCI
Operating supplies	0.9% of FCI
Laboratory charges	15% of Operating labor
Depreciation	10% of FCI
Plant overhead costs	70.8% of operating labor + 3.6% of FCI

Equation A1

$$DCF_k = \frac{NI_k}{(1+i)^k} \qquad \text{(A1)}$$

where NI_k is the net income in the year k and i is the annual discount rate.

The net present value (NPV) of the project is then calculated as the sum of the DCF_k through the lifespan of the project including the rate of return to investors and the capital investment:

Equation A2

$$NPV = \sum_{k=1}^{n} DCF_k - Capital\ Investment \qquad \text{(A2)}$$

where *n* is the lifespan of the project.

Finally, the supply cost is calculated using these equations. A DCFA was conducted assuming a real return of 10%, 2% inflation, and a 30-year project lifespan.

Appendix B

Appendix B contains supplementary information for chapter 3.

Equation B1

$$DCF_k = \frac{NI_k}{(1+i)^k} \tag{B1}$$

where NI_k is the net income in the year k, i is the annual discount rate, and DCF_k represents the present value of the net income (or loss) of the project in the year *k*.

Equation B2

$$NPV = -FCI + \sum_{k=1}^{n} \frac{NI_k}{(1+i)^k} \tag{B2}$$

where NPV is the net present value and *n* is the lifespan of the project.