Energy 173 (2019) 799–808

Contents lists available at ScienceDirect

Energy

journal homepage: www.elsevier.com/locate/energy

The development of a process simulation model for energy consumption and greenhouse gas emissions of a vapor solvent-based oil sands extraction and recovery process

Md.I.H. Soiket, A.O. Oni, A. Kumar*

Department of Mechanical Engineering, University of Alberta, 10-203 Donadeo Innovation Centre for Engineering, Edmonton, Alberta, T6G 1H9, Canada

ARTICLE INFO

Article history:
Received 13 October 2018
Received in revised form
12 February 2019
Accepted 14 February 2019
Available online 19 February 2019

Keywords:
Solvent extraction process
Energy consumption
GHG emissions
Solvent-to-oil ratio
Uncertainty analysis
SAGD

ABSTRACT

The solvent extraction process (SEP) is a promising technology for bitumen production from oil sands. This paper investigates the energy requirements and greenhouse gas (GHG) emissions in a propane-based super-heated vapor solvent extraction process as an alternative to in situ techniques, steam assisted gravity drainage, and cyclic steam simulation. A vapor solvent-based extraction model of 25,000 barrels per day of bitumen was developed to evaluate energy and GHG emissions. The energy- and emissions-intensive units were identified and a sensitivity analysis was conducted to study the impact of key operating parameters on the plant's overall emissions. Uncertainty analysis was performed on the most sensitive parameters. The results show that SEP natural gas and electricity consumption are 60.5–89.9 MJ/barrel and 19.7–19.9 kWh/barrel of bitumen, respectively. The GHG emissions range from 24.8 to 29.1 kg CO_2 eq./barrel of bitumen. The demethanizer unit, responsible for more than 90% of the electrical energy, is the most GHG emissions-intensive unit. GHG emissions are mostly influenced by the solvent-to-oil ratio. On the whole, the SEP emissions footprint is promising; however, the results must be extended to upgrading and refining emissions in order to understand the full life cycle GHG emissions of the transportation fuels produced.

1. Introduction

Bitumen is a useful resource containing a mixture of organic liquids that are highly viscous [1]. Its viscous nature makes it a challenging task to extract, process, and transport. The extraction of bitumen is crucial to the production of transportation fuel derived from oil-sands. Bitumen is extracted through surface mining or in situ technique depending on its formation, either deep beneath the earth or at the surface [2]. About 80% of the huge reserves of bitumen in Canada are in the province of Alberta, deep beneath the earth and extracted by in situ technique [2]. In 2017, the in situ technique accounted for over 50% (236.5 thousand cubic metres per day) of the bitumen extracted in Alberta. This value is expected to reach 403.3 thousand cubic metres per day by 2026. The current in situ methods is a thermal based system, which uses large amount of water and energy to generate steam for bitumen recovery. Thus, they impact on the sustainability of domestic water and increase emissions footprint. For these reasons, there has been increasing concern to develop energy efficient bitumen extraction process in the oil and gas industries in Canada.

The most commonly used in situ extraction processes are the cyclic steam simulation (CSS) and steam assisted gravity drainage (SAGD) [3]. In CSS, a single well is used for steam injection and oil production, while SAGD uses two horizontal wells; one is located slightly above than the other [4]. Steam is injected through the injection well, which heats the bitumen through condensation. The diluted bitumen, along with the condensed water, flows to the production well by gravity and is pumped to the surface for further processing [2]. Although these methods are effective for bitumen recovery, there are increasing concerns about their impact on the environment. A number of studies have been conducted to show the environmental impact of the thermal-based extraction method. In a study by Nimana et al. [4], they showed that natural gas and electrical energy consumption of SAGD is capable of releasing GHG emissions in the range of 8.0–34.0 gCO_2eq/MJ of bitumen. Although Nimana et al. suggested that SAGD emissions can be reduced significantly by considering cogeneration, no improvement was conducted to ascertain this claim. A similar study by

---

* Corresponding author.
E-mail address: Amit.Kumar@ualberta.ca (A. Kumar).

https://doi.org/10.1016/j.energy.2019.02.109
0360-5442/© 2019 Elsevier Ltd. All rights reserved.

Abbreviations

AGR	Acid Gas Removal
API	American Petroleum Institute
bbl	Barrel
bpd	Barrels Per Day
cp	Centipoise
CSS	Cyclic Steam Stimulation
DEA	Diethanolamine
FWKO	Free Water Knock-Out
GHG	Greenhouse Gas
GP	Gas Permeation
HEN	Heat Exchanger Network
IGF	Induced Gas Flotation
iSOR	Instantaneous Solvent-to-Oil Ratio
LCA	Life Cycle Assessment
LHV	Lower Heating Value
ppm	Parts Per Million
SAGD	Steam Assisted Gravity Drainage
SEP	Solvent Extraction Process
$SOR_{solvent}$	Solvent-to-Oil Ratio
SORsteam	Steam-to-Oil Ratio
TEG	Triethylene Glycol

Bergerson et al. [5] reported 9.0–16.0 gCO_2eq/MJ of bitumen for SAGD. In addition, they conducted sensitivity analysis which showed that steam production and electricity generation are the largest contributors to emissions. Other notable contributions to the study of GHG emissions from the CSS and SAGD processes have been made by Refs. [6–9]. These studies provided GHG emissions associated to the CSS and SAGD in range of 80–120 kg CO_2 eq./bbl and 51.8–220.3 kg CO_2 eq./bbl of bitumen, respectively. In most of these studies, it is clear that the resulting energy consumption and GHG emissions from bitumen recovery vary depending upon steam and electrical energy demand, operating conditions such as boiler efficiency, level of heat integration, the composition of bitumen, produced gases, water recovered, etc. Furthermore, the use of renewable energy for steam production is currently unattractive due to high production cost. Researches on the use of renewable energy for steam and power generation in bitumen recovery process are still ongoing.

Extensive research has been carried out to improve the performance of these in situ methods in order to make them less energy- and emissions intensive. For example, the work by Ashrafi et al. [10] focused on reducing the amount of natural gas used. Energy savings of approximately 8% and reduction in GHG emissions of 61,700 tonnes of CO_2 –equivalent per year was achieved. Carreon et al. [11] also evaluated energy efficiency opportunities for in situ extraction using pinch analysis. Benefits from environmental and energy savings were focused on for increased medium pressure steam. The estimated energy savings and GHG emissions reductions were approximately 6% and 5%, respectively. Extensive work done by Jacobs [12], Suncor Inc. [13], and Nadella [14] also did not lead to a significant improvement. It can be deduced from these studies that there are limitations to reduce water use, energy, and emissions of the steam-based in situ methods without addressing the possibility of using a less or non-thermal based process to increase the mobility of bitumen in the reservoir. Most of the improvement strategies addressed the surface facilities, while larger portion of the injected energy is lost or used in the reservoir. The difficult task is that the required steam necessary to raise bitumen mobility by increasing its temperature is large. Therefore, regardless of the improvement method applied to the steam-based in situ surface facilities; the overall environmental impact will not change significantly.

In the recent time, various extraction technologies have been proposed for cleaner and energy efficient operations relative to the currently used steam-based in situ methods. Some of these proposed techniques claimed that the use of hydrocarbon-based solvent requires no water and less heating for bitumen extraction. Some experimental studies have also shown that the use of hydrocarbon-based solvent such as propane, butane or pentene instead of steam is promising for cleaner operations [15,16]. Recently, N-Solv developed a pilot plant to demonstrate the hydrocarbon-based in situ extraction method [17,18]. The hydrocarbon-based in situ technique was invented based on the concept of solvent diffusion into diluted bitumen [19]. This process has a wellbore configuration similar to SAGD. In this process, a solvent such as propane, butane or pentane is injected as warm vapor through the injection well. The solvent remains in vapor phase and condenses on the extraction surface inside the reservoir [19]. Solvent selection in this process is important because bitumen must be diluted and heated using the solvent's latent heat of condensation. The most preferred solvent is propane [19] because it is less expensive than bitumen and can deliver heat at temperatures of about like 40 °C [20]. Condensed solvent diffuses with diluted bitumen and is pumped to the surface. The solvent is then separated from the bitumen and purified. The purified solvent is heated, compressed, and re-injected into the reservoir [19] to start the next cycle. Although N-Solv reported that the hydrocarbon-based solvent extraction method requires no water/steam, and will reduce energy and emissions significantly, they provided no information to ascertain their claims. To ascertain these claims, a data intensive life cycle assessment must be conducted.

While most of the published literature on bitumen extraction is focused on the thermal-based extraction method, there is currently no extensive study on the environmental impact of the hydrocarbon-based solvent extraction method. Although there a number of experimental studies which provide strong evidence for the extraction of bitumen using hydrocarbon solvent, further evaluation is required to ascertain the energy consumption and GHG emissions of the process on a commercial scale. This is a pathway to help policymakers and oil sands community provide useful information regarding the hydrocarbon-based solvent extraction method. This research is an effort to address this gap. In this study, a life cycle assessment of a hydrocarbon-based in situ technique for bitumen extraction is conducted. Using fundamental engineering principles, a data intensive model was developed to accurately represent the hydrocarbon-based solvent extraction process. Energy and life cycle greenhouse gas (GHG) emissions associated with the hydrocarbon based solvent extraction process were evaluated. The specific objectives of this study are to:

- Develop a solvent extraction process (SEP) model defining well configuration, reservoir conditions, and associated surface facilities to separate and purify the solvent from the produced emulsion.
- Evaluate the overall energy consumption and GHG emissions of SEP.
- Determine the SEP emissions uncertainty by performing a Monte Carlo simulation using a range of realistic data as input.

2. Method

2.1. Process description

Two well configuration similar to SAGD [21] was considered for the extraction process. Fig. 1 shows the schematic diagram of the solvent-based extraction and recovery of bitumen process. The production capacity of the plant is considered to be 25,000 bpd. The overall system is comprised of solvent injection, bitumen extraction, solvent and bitumen separation, and solvent compression and purification. The details of each unit are described in section 2.1.1.

2.1.1. Unit operations

2.1.1.1. Solvent injection. This unit involves the injection of vapor solvent through the injection well. The distance between the injection and production wells is considered to be 5 m [4]. Propane is purified by surface processing units (99 mol%) and injected as a vapor solvent at 70 °C at a pressure of 1800 kPa [19]. The reservoir temperature and pressure are assumed to be 8 °C [19] and 1379 kPa [20], respectively. All of the non-condensable gases are maintained at 1 mol% in the injected solvent. Solvent purification is an energy-intensive process [22] and, considering the reservoir conditions, injected solvent can bear approximately 6–7 mol% of non-condensable gases without causing solvent poisoning [19]. The reservoir wells are 300 m in length with a payzone (the zone where much of the oil is found) of 15 m [18,23]. The injection well is 148 m below the surface [18,23]. All the assumptions considered for reservoir conditions are presented in Table 1.

Table 1
Key assumptions in reservoir conditions.

Parameter	Values	Comments	Sources
Reservoir temperature	8 °C	–	[19]
Reservoir pressure	1379 kPa	Shallow reservoir	[20]
Reservoir length	300 m	-	[23]
Depth of injection well	148 m	-	[23]
Payzone	15 m	-	[23]
Porosity	35 (%)	-	[23]
Permeability	5 Darcy	-	[23]

2.1.1.2. Bitumen extraction. The heated vapor solvent is injected through the injection well at a higher pressure than the reservoir pressure. The injected solvent provides heat to dilute bitumen. The initial hot solvent accumulates as vapor at the top of the injection well but after condensed by releasing heat, the solvent falls because of the increase in density, and new injected solvent fills the empty space at the top, creating a solvent chamber at the bitumen interface. It is assumed that the chamber pressure is 1500 kPa [19]. The solvent-bitumen interface temperature is considered to be 45 °C (the bubble point of solvent). The solvent heats the bitumen to the bubble point temperature. The condensed solvent diffuses with the diluted bitumen. The diffusivity of the solvent in the diluted bitumen is significantly higher than solvent diffusing in raw bitumen [24] and thus recovers a large portion of bitumen and leaves behind a significant amount of asphaltene in the reservoir [18]. Solvent-to-oil ratio ($SOR_{solvent}$) of the extraction process is assumed to be 4 [23]. All the key extraction assumptions are presented in Table 2.

Through heating and diffusion, bitumen viscosity is lowered, and the solvent-bitumen emulsion goes into the production well by

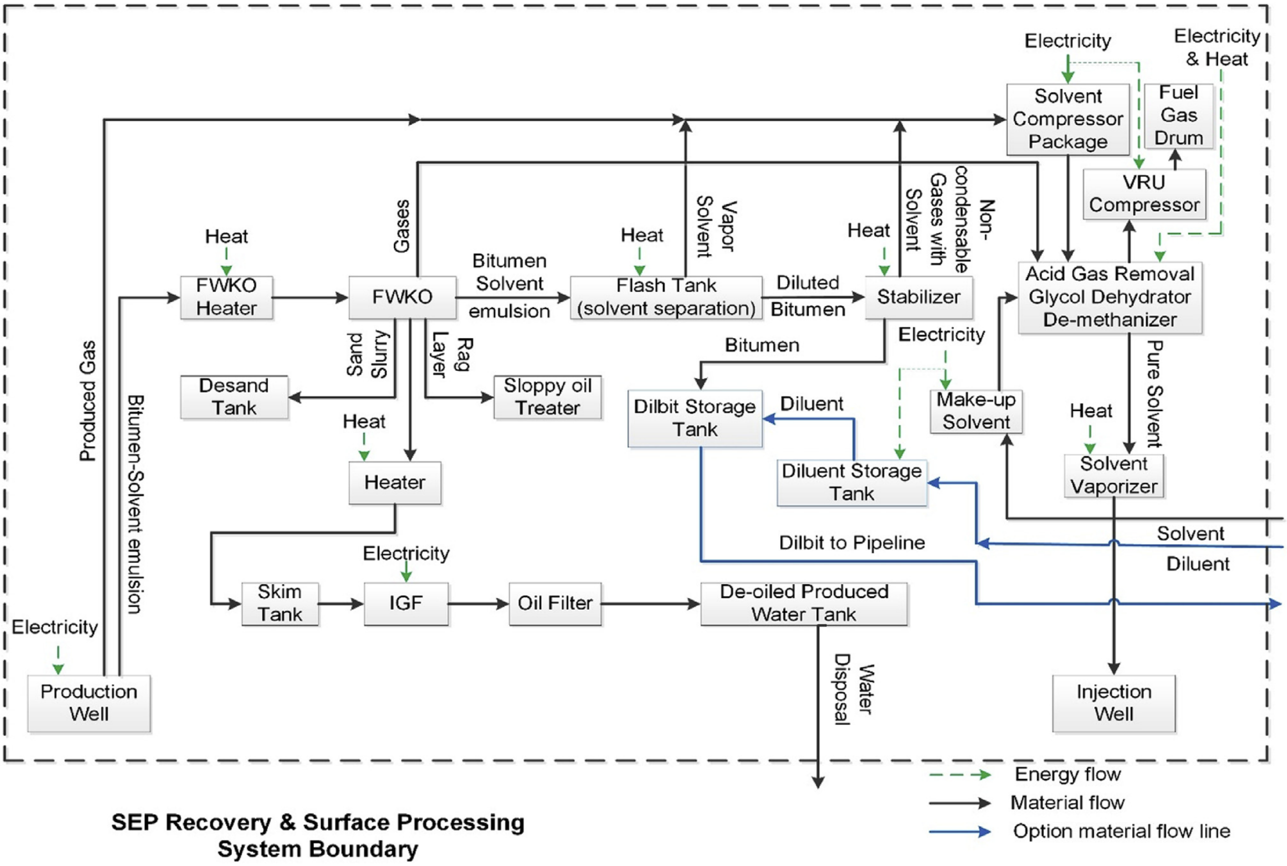

Fig. 1. Schematic diagram of the solvent-based extraction and recovery process of bitumen.

Table 2
Key assumptions for bitumen extraction.

Parameter	Values	Sources
Solvent-to-oil ratio ($SOR_{solvent}$)	4	[23]
Solvent bubble point	45 °C	
Solvent chamber pressure	1700 kPa	[19]
Solvent hold-up in the reservoir	20 (%)	[23]
Solvent composition	99 (mol%) propane and 1 (mol%) non-condensable gas	[19]
Produced gases	1–12 m^3/m^3 of bitumen	[4]
Produced water	1.75 (wt%) of bitumen	[4]
Solvent chamber pressure	1500 kPa	[19]

gravity drainage. Because of the solvent's high diffusivity and high asphaltene precipitation, the produced bitumen coming up from the production well is partially upgraded (14° API), unlike SAGD bitumen [18]. The bitumen-solvent emulsion drains at the solvent chamber pressure and is extracted to the top at 650 kPa by mechanical lift. The extraction pressure is selected based on the propane saturation curve. The emulsion needs to be extracted at a moderate pressure in order to maintain its stability. The properties of bitumen before and after extraction are provided in Table 3.

2.1.1.3. Solvent and bitumen separation. The first step after extraction is to separate the solvent from the emulsion. The solvent is separated in three stages. First, the emulsion is heated to about 40 °C and sent to a free water knock-out (FWKO) vessel. A FWKO is a pressurized closed vessel [27] used because it is a simple process and reduces electricity consumption [27]. The FWKO is operated at 1379 kPa [28], and it is assumed that more than 99 (wt%) of produced water is separated from the emulsion [29]. Moreover, due to the FWKO's operating conditions, the lighter fraction of the solvent (considered to be 10 wt% of the total solvent) is vaporized and separated as distillate from the top of the FWKO along with the produced gases. Water separated from the FWKO goes to the water de-oiling unit. It is assumed that the amount of sloppy water is negligible. Water from the FWKO bottom is heated before entering the de-oiling unit and then sent to a skim tank where suspended matter is separated by skimming. After further treatment in the induced gas flotation (IGF) unit [30] and the oil filter, the de-oiled water is trucked off site for disposal. Meanwhile, the solvent-bitumen emulsion, following water and solvent separation, is sent to the flash system. The flash system is a distillation column maintained at 64 °C and top pressure of 400 kPa, respectively [31]. Because of the temperature of the column, another 10 wt% of the solvent is separated from the tank top at high pressure. The bottom products from the flash system go to the stabilizer. Heat is provided in the stabilizer that causes the rest of the solvent, along with less condensable gases like methane, nitrogen, ethane, etc., to come out. Solvent-separated bitumen is then sent to the dilbit storage tank after being cooled to 50 °C. In the dilbit storage tank, bitumen is mixed with diluent (naphtha or natural gas condensate) in appropriate proportions and converted to dilbit, which is supplied to the upgrading or refining unit. The key operating parameters for solvent separation are provided in Table 4.

Table 3
Key assumptions in bitumen properties.

	Initial		Final	
Parameter	Values	Comments/ Sources	Values	Comments/Sources
API gravity	8°		14°	[18]
Viscosity (cp)	7 × 10ˆ6	[21]	10	[5]
Density (kg/m^3)	1020		958.1	As the produced bitumen will have very low asphaltene content, it will behave like crude oil [25]
Asphaltene (wt%)	12.7		3	[18]
Metals (ppm): Ni	137.5	[1]	30	[18]
V	68.5	[1]	77	[18]
Fe	77	[1]	5	[18]
Bitumen lower heating value (LHV) (GJ/bbl)			6.48	[26]
Emulsion pressure at production well top (kPa)			650	Assumed to maintain the stability of the emulsion

2.1.1.4. Solvent compression and purification. Gases are produced in the production well during bitumen extraction. Produced gases, along with separated solvent vapor, are compressed in a solvent compressor to 870 psi [33] because of the high pressure requirement in the demethanizer package. There is substantial solvent loss (solvent hold-up) from in situ operations and surface processing. In order to maintain a constant solvent flow rate, additional solvent is supplied from a make-up solvent unit. Solvent, along with gases, is sent to the demethanizer package. The demethanizer package consists of an acid gas removal unit, a glycol dehydrator, and a demethanizer. In this research, the amine process is considered for acid gas removal [33]. Diethanolamine (DEA) is used and regenerated to remove the H_2S and CO_2 contained in the upcoming feed. The K value for a conventional DEA load is 1.45 [34]. After acid gas removal, all gases are sent to the dehydration unit for water removal. Triethylene glycol (TEG) is used in the glycol dehydrator as a desiccant [33]. Equations from the literature were used to calculate energy consumption [33]. The detailed equations are provided in the Supporting Information. The removal of methane from the solvent is important because if the amount of methane exceeds a certain limit (>5 mol%) [19], it will cause solvent poisoning [19]. For this reason, external refrigeration is provided in the demethanizer to condense 100% of the solvent [33] and permit only 1.0 mol % methane in the solvent. The condensed solvent leaves the demethanizer bottom at about 290 psi [35] and then is converted into vapor solvent after further pressure drop. Finally, the solvent is heated in a solvent vaporizer and re-injected into the reservoir. As the amount of less-condensable gases is already low, it

Table 4
Key assumptions in the solvent separation process.

Parameter	Values	Comments	Sources
FWKO vessel pressure (kPa)	1379	-	[28]
FWKO vessel temperature (°C)	40	Temperature is selected based on the propane saturation curve	
Confined water (wt %)	3.5	-	[1]
Flash column pressure drop (kPa)	35	-	
Flash column temperature (°C)	64	-	[31]
Stabilizer pressure (kPa)	4500	-	[32]
Stabilizer temperature (°C)	110	-	[28]
Thermal efficiency (%)	90	It is assumed that the FWKO heater, flash column and stabilizer have the same thermal efficiency	
Solvent separation (wt%)	10 (FWKO) 10 (Flash Column) 80 (Stabilizer)	Solvent separation is determined base on the propane saturation curve	

is assumed that re-compression work will be supplied by the turboexpander [34]. The OPGEE model which uses an engineering-based life cycle assessment (LCA) tool [33] was used to calculate demethanizer energy consumption. The refrigeration system and compressor consume the most energy in the demethanizer [33]. The reboiler at the bottom of the fractionating column also consumes energy by providing heat in the fractionating column [33]. Reboiler efficiency values are shown in Table 5 along with other key operating parameters.

The critical parameters leading to high heat and electricity consumption were identified from the calculations. Variations in the efficiency of most energy-consuming equipment can result in significantly higher or lower energy consumption. A sensitivity analysis was performed on the energy-sensitive parameters and equipment efficiency. The parameter values were varied within ±30 in keeping with a previous study [4] to identify the most energy-sensitive parameters and equipment. An uncertainty analysis incorporating the most energy-sensitive parameters and equipment was also conducted to determine the feasibility of implementing the modeled simulation. The detailed sensitivity and uncertainty analyses, along with the results, are discussed below.

2.2. Mass and energy analysis

The following basic assumptions were made in developing the model equations:

i. The system is in a steady state flow condition
ii. Change in percentage mole fraction (mol%) is equivalent to change in percentage volume fraction (vol%) considering ideal gas flow
iii. The temperature and pressure at the reference state are $T_o = 25\,^\circ\mathrm{C}$ and $P_o = 101$ kPa, respectively.

The simulation model developed in this study is based on mass and energy balance equations.

The simplified fundamental mass and energy equations are, respectively:

$$\sum m_i = \sum m_e \tag{1}$$

$$\sum E_i + \sum Q_{CV} = \sum E_e + \sum W_{CV} \tag{2}$$

where m denotes mass flow rate, E stream energy rate, Q_{cv} heat rate into a control volume, W_{cv} work done by a control volume and the subscripts i and e denote inlet and exist, respectively.

Table 5
Key assumptions for solvent compression and purification.

Parameter	Values	Comments	Sources
Produced gas (m^3/m^3 of bitumen)	1–12	Produced gas will fit in the wide SAGD- produced gas range	[29]
K value of DEA	1.45	-	[34]
Reboiler efficiency (%)	80	Assumed	
TEG concentration (wt%)	99	-	[33]
Water removal rate (gal of TEG/lb of H_2O)	2	-	[33]
Gas pressure at compressor outlet (kPa)	6000	-	[35]
Demethanizer pressure (kPa)	3000	-	[34]
Condensed propane pressure (kPa)	2000	-	[35]
Condensed propane temperature (°C)	46.11		

Using the equations above, the energy loss was calculated in each unit operation. To calculate the amount of heat generated from the produced gas and the additional amount of natural gas required meeting the total heat (steam and heat) requirement, the following equations were used:

$$\mathit{Heat\ generated\ by\ produced\ gas} = \sum_{i=1}^{n} \frac{(LHV)_i{*}m_i}{\eta_f} \tag{3}$$

$$\mathit{Natural\ gas\ required} = \frac{\mathit{Heating\ required\ from\ natural\ gas}}{{*}\mathit{natural\ gas\ LHV}}. \tag{4}$$

where i refers to individual gas component, LHV = lower heating value (kJ/kg), m_i = mass flow rate of produced gas (kg/bbl) and η_f and η_b are the furnace and boiler efficiency, respectively.

Equation (5) was used to calculate both heat and steam production from natural gas as $\eta_f = \eta_b = 80\%$ [33]. The heating required from natural gas in equation (5) is represented in kJ/bbl unit. Steam quality was assumed to be 100% in this study. Thus, the heat generated by produced gas in equation (4) and natural gas required in equation (5) is expressed in kJ/bbl and kg/bbl unit, respectively.

Equations (6) and (7) were used to calculate the electricity consumed in the pump and compressor, respectively.

$$\mathit{Pump\ work} = \frac{m.g.h}{\eta_p} \tag{5}$$

$$\mathit{Compressor\ work} = \frac{Q.\Delta P}{\eta_c} \tag{6}$$

Here, m is the total mass of the bitumen-solvent emulsion (kg/s), h is the height of the reservoir from bottom to surface (m), Q is the total gas feed in the compressor (m^3/s), ΔP is the compressor pressure difference between input and output (Pa), and η_p (90%) and η_c (75%) [4]are the pump and compressor efficiency, respectively.

2.3. Estimation of GHG emissions

The functional unit for an GHG footprint assessment of oil sands-derived fuels is considered in terms of kg of CO_2-equivalent gas emitted per bbl of bitumen extracted (kg CO_2 eq./bbl). The lower heating value (LHV) of bitumen and fuels is considered to calculate energy consumption and GHG emissions. The data-intensive Excel-based simulation model uses default parameters with the provision to input user data. The emissions calculated for sub-unit operations include (i) combustion emissions due to burning produced fuel gas on site and (ii) upstream and combustion emissions associated with transferring and burning natural gas in order to provide required heating for extraction and surface processing. In this developed SEP model it is assumed that all the electricity will be supplied from the Alberta grid. The paper also includes fugitive emissions from leakage and the irregular release of gases as well as venting and flaring emissions but does not include emissions due to land use. Land-use emissions for in-situ production are less than 0.4 CO_2/MJ of product [36].

The emissions incurred due to the energy consumption of each process were evaluated using the emission factors from literature [7]. For natural gas, GHG emission factors include both upstream [4] and combustion emissions. For produced gas, only combustion GHG emissions are taken into consideration as those gases are produced on site. All the electricity required in the plant is assumed to be supplied from the Alberta grid, which is one of the largest hydrocarbon base in North America and where most of the oil sands in Canada are located. Electricity production in Canada is moving towards natural gas-based production due to the phasing out of coal, but there are still some operational coal-fired power plants. So upstream emissions for electricity production were calculated based on a mixture of coal and natural gas supply (70% coal and 30% natural gas), which is in good agreement with literature [7,37]. Combustion emissions related to electricity production were also taken from the NIR (national inventory report) [37]. In this research, it was assumed that GHG emissions consist of only CO_2, CH_4, and N_2O gases. Venting emissions were calculated using the global warming potential (GWP) of these gases, collected from the Intergovernmental Panel on Climate Change (IPCC) Fifth Assessment Report [38]. Venting emissions usually come from storage tanks, equipment leaks, process vents, etc. Flaring and fugitive emissions were also estimated from the GHG emissions data provided by the NIR for oil sands mining, extraction, and upgrading [37]. All the emissions factor values are provided in the Supporting Information section of this paper.

3. Results and discussion

3.1. Energy and emissions in the unit operations

Energy is consumed in the form of heat, steam, and electricity in bitumen extraction and recovery. Through pumping, bitumen-solvent emulsion is lifted from the reservoir bottom to the top. Natural gas is burned to provide heat in the FWKO heater, flash tank, and stabilizer. Water separated in the FWKO, is heated and then transferred to the water de-oiling unit. Electricity is consumed by the mechanical contractors that agitate the gas bubbles used in induced flotation. The solvents, along with the produced gases, are compressed in the compressor. The compressor energy is calculated using equation (6). The highly compressed gases are then fed into the acid gas removal (AGR) unit to remove all the CO_2 and H_2S. In the AGR unit, electricity is consumed by the booster, reflux, and circulation pumps to remove acid gases and regenerate the amine solution. The amine reboiler is a direct-fired heater that consumes natural gas to provide heating [33]. The aerial cooler needs electricity to cool the amine solution. The sweet gases from the AGR unit go to the glycol dehydrator where all the water contained in the gases is removed. This dehydration is necessary to prevent a reduction in the heating value and solid hydrate formation [33]. In the dehydrator, the reboiler provides heating to regenerate the TEG desiccant. The reboiler consumes natural gas to generate steam. Electricity is consumed by the glycol pump to circulate TEG. In the demethanizer, the propane is condensed by external refrigeration that consumes a significant amount of electricity. The gases separated from the demethanizer are sent to a fuel drum. The produced gases are burned to supplement the heat requirement.

All the equations used to calculate the total energy requirement in the demethanizer package are provided in the Supporting Information. Table 6 shows the overall energy consumption and GHG emissions from each unit operation. The range of values is included because we considered ranges for some input parameters rather than point estimates. The default energy and GHG emissions values correspond to the mean value of the input parameters. The range of output values provides a more conservative estimate of real life implementation compared to a point estimate.

The total energy required to extract and recover 1 bbl of bitumen is 132.9–198.5 MJ. The total energy required in the SEP is lower than in the thermal extraction methods (622.1–1003.1 MJ/bbl of bitumen for a SOR_{steam} of around 2.5) reported in the literature [6–9]-. In thermal extraction, a huge amount of steam is required to provide heating to dilute and extract bitumen. A significant amount of heat is also lost during extraction because of high operating pressure [19,24]. On the other hand, the SEP operates at relatively lower pressure and temperature than SAGD [19]. No steam is required in this process, and propane is more effective for separating bitumen's heavy and light components from the reservoir than steam [24]. Due to blowback [19] from previously depleted SAGD chambers, a high amount of steam might go in other directions without coming in contact with bitumen, which will increase the SOR_{steam} significantly. The blowback problem in SAGD operations can also be eliminated with solvents by controlling the amount of methane in the reservoir. So the SEP not only reduces energy consumption but also provides a better opportunity to control the extraction process.

Table 6 shows that about 39.2–43.5% of the total energy used are through electricity consumption. The reason for this high electricity consumption in SEP is the external refrigeration that is required to condense the higher amount of C_2+ compounds in the demethanizer feed. The demethanizer accounted for 91% (19 MW) of the electricity consumed, while solvent compressors and other auxiliaries are 9% (1.8 MW). In the turbo-expansion process, the higher the heavier compounds in the incoming feed, the higher the amount of external refrigeration required [69] to condense the heavier compounds. The feed to the demethanizer can be passed through a membrane-based gas separation process [39,40] to reduce the amount of C_2+ compounds in the demethanizer feed. Membranes can be used to separate gases without changing the phase by gas permeation (GP). The driving force for membrane separation can be pressure, concentration, temperature, or electrical potential. The distinguishing feature of membrane separation is the additional membrane phase, which improves separation efficiency and reduces energy consumption from separation [30]. The power required for mechanical lifting of bitumen emulsion, running different pumps and coolers in extraction, and surface processing is included in “other auxiliaries.”

Heat energy consumption in different units is shown in Fig. 2. The stabilizer consumes the highest amount of heat energy. The

Table 6
Fuel consumption and GHG emissions in the Solvent Extraction Process (SEP).

	Fuel consumption			Emissions (kg CO_2 eq./bbl of bitumen)		
	Unit	Range	Default[a]	Type	Range	Default[a]
Natural gas	(kg/bbl)	1.5–2.2	2.0	Indirect	4.4–6.5	5.2
	(MJ/bbl)	60.5[b]-89.9	82.1			
Produced gas	(MJ/bbl)	1.5–37.0	19.3	Direct	0.1–2.4	1.2
Electricity	(kWh/bbl)	19.7–19.9	19.8	Indirect	17.4–17.6	17.5
	(MJ/bbl)	70.9–71.6	71.3	Venting	0.4–4.3	2.4
				Flaring	-	0.1
				Fugitive	-	0.3

[a] The default is considered the mean value.
[b] The natural gas lower heating value (LHV) is considered to be 40,700 kJ/kg [34].

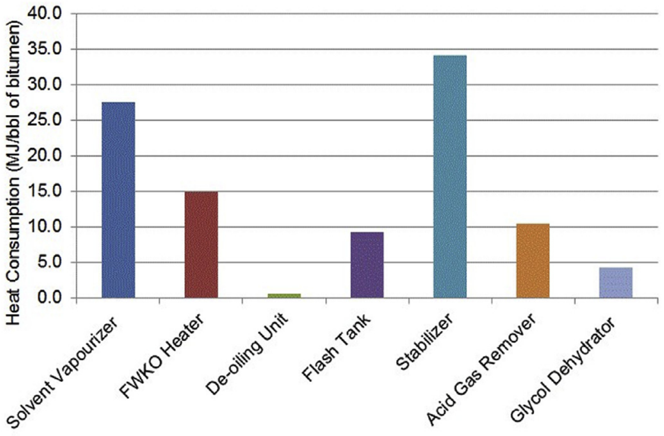

Fig. 2. Heat energy distribution in extraction and surface processing units.

majority of non-condensable gases are separated by the heat provided in the stabilizer column. Because of the relatively high operating temperature of the stabilizer, the pressure of the separated solvent increases accordingly. This reduces the work of the compressor. The solvent vaporizer consumes 27.3% of the overall heat energy. The solvent vaporizer is required to boil the condensed bottom product in the demethanizer unit. This is because a large amount of condensed solvent must be superheated before it is reinjected into the reservoir.

Another major heat consumer is the FWKO heater. The FWKO heater increases the temperature of the produced emulsion to ensure that a considerable amount of water is separated in the FWKO drum. In the FWKO drum, the lighter emulsion fractions stay on the top layer and the water-sand slurry remains at the bottom due to the density difference. The volatile fraction (considered to be 10%) of the solvent comes out from the FWKO top. The water-sand slurry is ejected from the FWKO bottom. After the sand has been filtered, almost all the water goes to the de-oiling unit. The amine reboiler in the AGR unit consumes natural gas to provide heating. Heating breaks up the chemical bonds between amine and acid gas [33].

The total GHG emissions from the SEP process vary between 24.8 and 29.1 kg CO_2 eq./bbl of bitumen. It can be inferred from Table 6 that 60–71% of the total GHG emissions are from electricity production. There are potentials to reduce the overall GHG emissions significantly. Electricity production from renewable sources can be an attractive option. The application of cogeneration is another favorable option, which has been reported to meet electricity demand in the existing extraction processes [4]. With cogeneration, it is possible to reduce upstream GHG emissions, which will result in lower overall emissions.

The GHG emissions from the SAGD extraction process range between 47.8 and 220.3 kg CO_2 eq./bbl of bitumen [6–9]. However, it is important to mention that a direct comparison with SEP is difficult due to dissimilarity in the extraction processes and the nature of bitumen that is produced. Since the solvent precipitates majority of the asphaltene, the produced bitumen properties (e.g. API gravity, viscosity, metal content) are different than the bitumen from thermal production processes like SAGD and CSS. It is difficult to compare the energy consumption and GHG emissions results with other extraction results at this stage of analysis. A well-to-wheel life cycle assessment on the vapor solvent extraction process is necessary to make meaningful deductions with other extraction methods. A common ground for comparison is at the life cycle stage where the extracted crude oil is processed into common final products such as gasoline, diesel and jet fuel.

3.2. Sensitivity analysis

A sensitivity analysis was performed on the most energy-intensive parameters and on the efficiency of the equipment. The results are given in Table 6. As discussed in section 3.1, the majority of GHG emissions are due to high electricity consumption. The solvent compressor and the demethanizer are the two most power-intensive units. As shown in Fig. 2, the solvent vaporizer consumes the highest heat energy. The heating required in the solvent vaporizer depends on the temperature of the condensed solvent from the demethanizer. Also, the higher the amount of solvent recovered from the reservoir, the higher the amount of energy required to process it. So, to see the impact of condensed solvent temperature and the amount of solvent hold-up in the reservoir on energy consumption and GHG emissions, these parameters were included in the sensitivity analysis along with the parameters mentioned above.

As illustrated in Fig. 3, the GHG emissions from the SEP are highly dependent on the external refrigeration required in the turbo-expander. External refrigeration can be reduced by transferring all the gases from the compressor to a membrane-based gas separator. As the gases from the compressor will be at a higher pressure, pressure can be used as a driving force to separate C_2+ compounds. As discussed earlier, external refrigeration in the demethanizer is required to condense C_2+ compounds so lowering the amount of C_2+ compounds in the feed will reduce power consumption in the demethanizer significantly. Since the gas pressure at the compressor outlet is slightly sensitive to GHG emissions, determining the optimum outlet gas pressure will lower energy consumption and emissions. The solvent hold-up in the reservoir has negligible impact on GHG emissions within a ±30 (vol %) variation. This is because the total energy requirement varies inversely with the solvent hold-up in the reservoir but it does not vary significantly per bbl of bitumen. As GHG emissions are related to energy consumption, the GHG emission rate also remains insensitive. Another noticeable factor is the temperature of the condensed solvent. If the temperature increases more than 10% from the base temperature, GHG emissions drop sharply. This is because if it was possible to separate all the solvents as vapor from the demethanizer, no additional heating would be required in the solvent vaporizer. The energy saved in this process would also save a significant amount of GHG emissions. But with the current technology available, it is not a feasible option.

The sensitivity of equipment efficiency is presented in Fig. 4. The

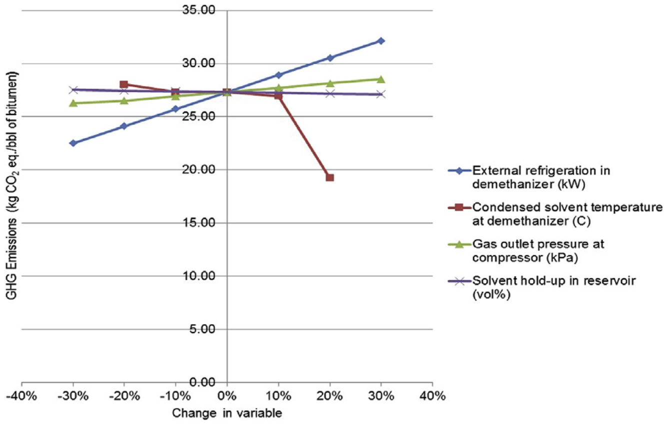

Fig. 3. The sensitivity of critical parameters on GHG emissions.

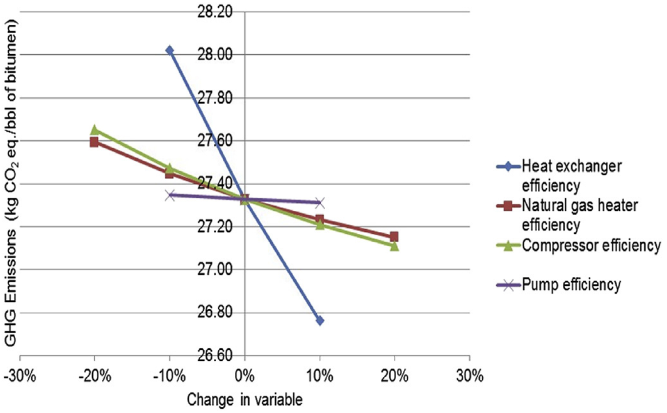

Fig. 4. Sensitivity analysis of GHG emissions to equipment efficiency.

heat consumption of the SEP significantly depends on the heat exchanger's efficiency. The compressor and the natural gas heater efficiency can be improved to optimize energy consumption. Although the technology of the heat exchanger and steam boiler has matured, their efficiency might drop after a certain period of operation. By optimizing the operating conditions and heat exchanger network (HEN), the efficiency of the heat exchangers can be improved significantly [41,42]. The fluids inside the heat exchangers deposit suspended and dissolved solids on the heat exchanger surface [43]. After a long period of operation, the deposited solids create an additional layer on the heat exchanger surface, which prevent effective heat transfer. So maintenance should be carried out regularly for consistent performance and to prevent fouling. Compressor performance can also be improved by modifying the design of the compressor blades, as shown by Lee and Kim [44]. Energy efficiency auditing of the heat exchangers, heaters, and compressors might further reduce energy consumption [45,46].

Given that the SEP is very sensitive to energy consumption, the effect of changing the instantaneous solvent-to-oil ratio (iSOR) on energy consumption was analyzed. It was found that energy consumption increases linearly with increasing iSOR, as depicted in Fig. 5. The change in heat consumption is greater than in electricity consumption. For electricity consumption, the result is reasonable compared to other solvent-based extraction processes. Heat consumption increases because of the additional heat required to heat and process the solvent. Similar analogies have been found for SAGD [4]. Electrical heating techniques like Electro-Thermal Dynamic Stripping (ET-DSP), Enhanced Solvent Extraction Incorporating Electromagnetic Heating (ESEIEH), etc., can be used to reduce the solvent requirement. These techniques use radio frequencies to polarize water molecules [47]. The water molecules then try to align with the electromagnetic field. In the negative half cycle of the microwave, the poles are flipped [48]. The agitation and flipping generate heat. The additional heat can be used to reduce the amount of solvent required to extract bitumen. Using less solvent will reduce the iSOR and total energy consumption.

Significant heat is required to separate the solvent from the solvent-bitumen emulsion. To determine how heat consumption varies with the amount of solvent separated in different units, the solvent separated in the stabilizer is decreased with an analogous increase in separation in the flash tank. Solvent separated in the FWKO is kept constant because if a higher amount of solvent is separated from the FWKO at a relatively lower pressure it will increase the work of the compressor. Increasing the compressor work will increase GHG emissions because both compressor efficiency and compressor outlet pressure are sensitive parameters. As illustrated in Fig. 6, GHG emissions increase linearly with an increase in the amount of solvent separated in the stabilizer. By looking at the ordinate values, we can infer that the amount of solvent separated in different stages insignificantly contributes to GHG emissions.

3.3. Uncertainty analysis

From the sensitivity analysis we determined that the two parameters that impact the emissions most are heat exchanger efficiency and the amount of external refrigeration in the turboexpander. The amount of external refrigeration depends on the amount of C_2+ compounds in the feed. The feed composition in the demethanizer has significantly higher amounts of C_2+ compound (77.6–87.6 mol%) than usual [49–51]. In order to make the design robust and allow for flexibility in the demethanizer feed, it was assumed that 90% of the C_2+ feed requires external refrigeration in the base model as a worst case scenario. Using the relation developed by Diaz et al. [49], we found that about 22 mol% C_2+ feed require external refrigeration. This wide range was used to develop the triangular distribution for an uncertainty analysis of external refrigeration. An uncertainty analysis was performed on these parameters along with total electricity consumption and the efficiency of the natural gas (NG) heater. The total electricity consumption was included in the uncertainty analysis to observe the impacts of compressor work and compressor efficiency. A Monte Carlo simulation was used to determine the uncertainty of the input parameters. The simulations were run in ModelRisk software. In order to perform the simulations, the statistical

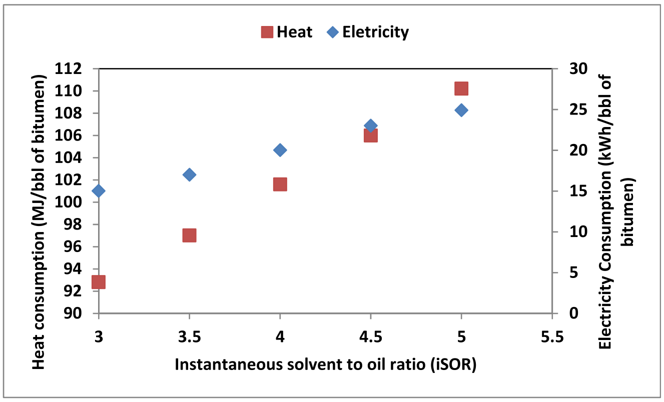

Fig. 5. Correlation between heat and electricity consumption with instantaneous SOR in the SEP.

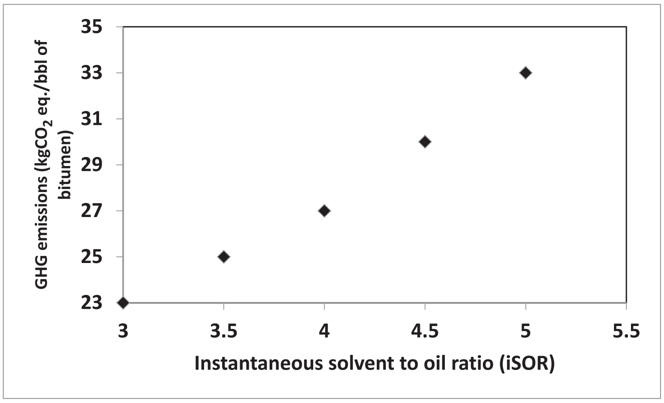

Fig. 6. Effect on GHG emissions of variations in solvent separation in the stabilizer.

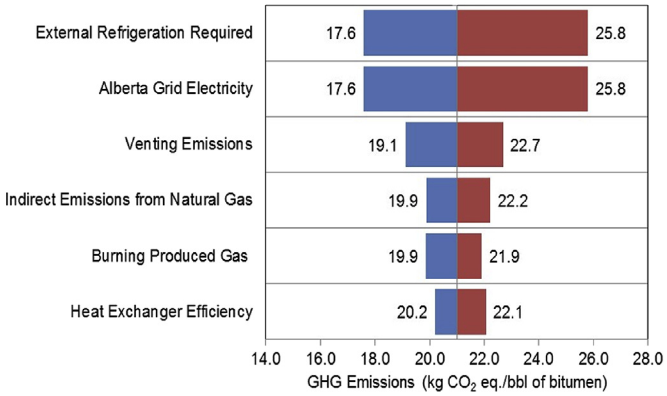

Fig. 7. Uncertainty analysis of the most sensitive parameters.

distributions of the input parameters are required. Because of limited data, triangular distributions were generated for every parameter except produced gas and venting GHG emissions. A triangular distribution gives conservative results for predictable values as well as a lower standard deviation. For produced gas and venting GHG emissions, it is hard to predict a reliable value as those parameters vary significantly from well to well. This variation is taken into account by choosing a uniform distribution as it gives the most conservative distribution and treats all the input values equally. It is estimated that to keep sampling errors below 0.1 kg CO_2 eq./bbl of bitumen, 100,000 runs are required. Formulas to calculate sampling errors are provided in the Supporting Information. The results are presented for 99% confidence intervals in Fig. 7.

For both external refrigeration and total electricity consumption, the emissions have the same values within the 0.5 and 99.5 percentile. This implies that compressor work and efficiency are relatively insensitive and the highest variations in uncertainty occur because of refrigeration in the demethanizer. The efficiency of the NG heater provides a wider range uncertainty in GHG emissions compared with the efficiency of the heat exchanger, which contradicts the results found from the sensitivity analysis shown in Fig. 4. This is because the efficiency of the heat exchanger is considered to be 90% in this research. As the efficiency of the heat exchanger cannot be higher than 100%, the efficiency was varied by ±10% and the efficiency of the NG heater was varied by ±20%. For this reason the uncertainty in emissions for the NG heater (shown in Fig. 7) has a wide range. Although a wide range for produced gas from a well is considered in this research, the uncertainties in produced gas emissions have a relatively narrow range.

4. Conclusions

This research presents a data-intensive simulation model to estimate fuel input, energy consumption, and GHG emissions for a solvent-based oil sands extraction process. Engineering first principles were used to determine mass and energy balances in each sub-unit operation. The total heat consumption is from 62.0 to 126.9 MJ per bbl of bitumen depending on the amount of gas produced from the reservoir. The total electricity consumption is relatively constant, and it is estimated that 19.7–19.9 kWh of electricity is required to extract and recover each bbl of bitumen. It was found that total GHG emissions vary from 24.8 to 29.1 kg CO_2 eq. eq./bbl of bitumen. Although the properties (e.g. API gravity, asphaltene content) of the vapor solvent extracted bitumen differs from that of CSS and SAGD extraction, the GHG emissions from the CSS and SAGD processes range from 80 to 120 kg CO_2 eq./barrel (bbl) and 51.8–220.3 kg CO_2 eq./bbl of bitumen, respectively.

More than 92% of the total electricity required is used to condense the solvent in the demethanizer. The solvent extraction process is very sensitive to the efficiency of equipment, and among the equipment, the efficiency of the heat exchanger is the most sensitive one to solvent extraction and surface processing GHG emissions. Electricity consumption can be reduced by using low energy-intensive gas separation processes. Improving the heat exchanger network and optimizing operating conditions will improve the efficiency of heat exchangers. As the total energy consumption depends on the efficiency of the equipment, proper monitoring with regular maintenance is necessary for consistent performance. Energy consumption varies linearly with the iSOR, as heat consumption is more sensitive than electricity consumption. Energy consumption can be reduced by operating at a lower iSOR. On the whole, the overall emission footprint of the solvent based extraction method is promising. However, these results must be extended to upgrading and refining emissions in order to understand the overall life cycle GHG emissions of the transportation fuels produced from solvent based extraction method.

This modeling approach will help industries find strategic pathways for sustainable oil sands derived liquid fuel production to meet global energy demands. It will also help policy makers make effective policies considering international market demand, oil price, environmental regulations, etc. that will help the oil sands industry flourish not only in Canada but also around the world.

Acknowledgements

We thank the 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. We also thank representatives from Alberta Innovates (AI), Suncor Energy Inc., Cenovus Energy Inc., Natural Resources Canada (NRCan), and Environment and Climate Change Canada (ECCC) for their valuable inputs and comments in various forms. 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). The authors are grateful to Dr. Eddy Isaacs, Strategic Advisor, Faculty of Engineering at the University of Alberta for his review comments and feedback during the study. The authors are thankful to Astrid Blodgett for editorial support.

Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.energy.2019.02.109.

References

[1] Gray MR. Upgrading oilsands bitumen and heavy oil. University of Alberta; 2015.

[2] Bolea I, Ordorica-Garcia G, Nikoo M,MC. Techno-economics of CCS in oil sands thermal bitumen extraction: comparison of CO2 capture integration options. Energy Procedia 2013;37:2754–64.

[3] Egboka CI, Yang DT. Performance of a SAGD process with addition of CO2, C3H8, and C4H10 in a heavy oil reservoir. Conference performance of a SAGD process with addition of CO2, C3H8, and C4H10 in a heavy oil reservoir. Society of Petroleum Engineers.

[4] Nimana B, Canter C, Kumar A. Energy consumption and greenhouse gas emissions in the recovery and extraction of crude bitumen from Canada's oil sands. Appl Energy 2015;143:189–99.

[5] Joule Bergerson, David K. Life cycle assessment of oil sands technologies. 2006.

[6] Keesom William, Unnasch Stefan, Moretta J. Life cycle assessment comparison of North American and imported crude. Chicago, Illinois: Jacobs Consultancy for Alberta Energy Research Institute; 2009.

[7] Wang M, Wu Y, Elgowainy A. The greenhouse gases, regulated emissions, and energy use in transportation model. Argonne, Illinois: Argonne National

Laboratory; 2014.
[8] Charpentier AD, Kofoworola O, Bergerson JA, MacLean HL. Life cycle greenhouse gas emissions of current oil sands technologies: GHOST model development and illustrative application. Environ Sci Technol 2011;45(21):9393–404.
[9] Inc. STC. GHGENIUS MODEL 4.02. Ottawa, Ontario: Natural Resources Canada; 2013.
[10] Ashrafi O, Navarri P, Hughes R, Lu D. Heat recovery optimization in a steam-assisted gravity drainage (SAGD) plant. Energy 2016;111:981–90.
[11] Carreon Carlos E, Mahmoudkhani Maryam, Alva-Argaez Alberto, Bergerson J. Evaluation of energy efficiency options in steam assisted gravity drainage oil sands surface facilities via process integration. Appl Therm Eng 2015;87:788–802.
[12] Jacobs. Consultancy SAGD energy efficiency study september. 2009.
[13] Suncor energy services Inc. Climate change emissions management corporation, Alberta Innovates - energy and environment solutions, and Jacobs consultancy Canada Inc., a greenhouse gas reduction roadmap for oil sands.
[14] Murthy Nadella N. Improving energy efficiency in thermal oil recovery surface facilities. Montreal: World Energy Congress; 2010.
[15] Butler Roger M, Mokrys IJ. A new process (V APEX) for recovering heavy oils using hot water and hydrocarbon vapour. J Can Petrol Technol 1991;30(01):97–106.
[16] Mokrys Ignor J, Butler RM. The rise of interfering solvent chambers: solvent analog model of steam-assisted gravity drainage. J Can Petrol Technol 1993;32(03):26–36.
[17] Corp. N. BEST (bitumen extraction solvent technology) field pilot plant. Sustainable Development Technology Canada; 2017.
[18] Corp N. Nsolv overview and pilot performance update. Nsolv – the solution for economic & environmentally sustainable oilsands development in a low price environment. 2015.
[19] John Nenniger, Nenniger E. Method and apparatus for stimulating heavy oil production. In: Office USPaT. United States: N-Solv Corp; 2005.
[20] Keyser T. COOL PROPANE: ideal solvent for bitumen recovery, in propane Canada. 2011. p. 16–7.
[21] Zhu Z, Zeng F, Zhao G, Laforge P. Evaluation of the hybrid process of electrical resistive heating and solvent injection through numerical simulations. Fuel 2013;105:119–27.
[22] Darvishmanesh S, Robberecht T, Luis P, Degrève J, Van Der Bruggen B. Performance of nanofiltration membranes for solvent purification in the oil industry. JAOCS, Journal of the American Oil Chemists' Society 2011;88(8):1255–61.
[23] Corporation N-S. N-solv overview, 2013 PTAC OIL SANDS FORUM. 2013.
[24] Nenniger JE, Dunn SG. How fast is solvent based gravity drainage? Canadian International Petroleum Conference. Calgary, Alberta: Petroleum Society of Canada; 2008. p. 1–14.
[25] Ancheyta J. Modeling of processes and reactors for upgrading of heavy petroleum. Taylor and Francis Group; 2013.
[26] Nimana BS. Life cycle assessment of transportation fuels from Canada's oil sands through development of theoretical engineering models. Edmonton, Alberta: University of Alberta; 2014. Edmonton, Alberta.
[27] Ergun E. Pressure changes effects of the outlet free water knock out on the oil in water: a case study for karakus field, adiyaman-Turkey. SPE Middle East oil and gas show and conference. Society of Petroleum Engineers; 2013.
[28] Arnold K, Stewart M. Surface production operations: vol 2: design of gas-handling systems and facilities. third ed. ed2014.
[29] Ender Ergun, Ali Emre Ercelebi, Sahin MO. Environmentally sustainable water management for mature oil fields: a case study for Karakus Field, Adiyaman-Turkey. SPE oilfield water management conference and exhibition, 21-22 April. Kuwait City. Kuwait: Society of Petroleum Engineers; 2014.
[30] Chen Etigz. Produced water treatment technologies. Int J Low Carbon Technol 2012;9:157–77.
[31] Stork K. Integrated low pressure depropanizer/debutanizer column. United States: kellogg brown and root. INC.; 2001.
[32] Devold H. Oil and gas production handbook an introduction to oil and gas production, transport, refining and petrochemical industry. Oslo, Norway: ABB Oil and Gas; 2013.
[33] Hassan ME-H, Masnadi Mohammad S, Vafi Kourosh, Duffy James, Brandt AR. Oil production greenhouse gas emissions estimator. Stanford school of earth, energy and environmental sciences; 2017.
[34] Manning Francis S, Thompson RE. Oilfield processing of petroleum: natural gas, volume one. Oklahoma, United States: PennWell Books; 1991.
[35] Nawaz M, Jobson M. Synthesis and optimization of demethanizer flowsheets for low temperature separation processes. In: de Haan A, Kooijman H, Gorak A, editors. Conference Synthesis and optimization of demethanizer flowsheets for low temperature separation processes; 2005. p. 79–84.
[36] Yeh S, Jordaan SM, Brandt AR, Turetsky MR, Spatari S, Keith DW. Land use greenhouse gas emissions from conventional oil production and oil sands. Environ Sci Technol 2010;44(22):8766–72.
[37] Warren Baker DB, Ana Blondel, Génier Maxime, Hickey Jason, Chang Liang, MacDonald Doug, Matin Afshin, Scott McKibbon, Frank Neitzert, Palmer Craig, Pratt Lindsay, Smith Duane, Smyth Steve. Greenhouse gas sources and Sinks in Canada 1990-2013. National inventory report. Gatineau, QC: the Canadian government's submission to the UN framework convention on climate change. 2015.
[38] The Core Writing Team RKP, Meyer Leo. Climate change 2014 synthesis report. Geneva, Switzerland: Intergovernmental Panel on climate change; 2015.
[39] Pan Y, Li T, Lestari G, Lai Z. Effective separation of propylene/propane binary mixtures by ZIF-8 membranes. J Membr Sci 2012;390–391:93–8.
[40] Zhang C, Dai Y, Johnson JR, Karvan O, Koros WJ. High performance ZIF-8/6FDA-DAM mixed matrix membrane for propylene/propane separations. J Membr Sci 2012;389:34–42.
[41] Waheed MA, Oni AO. Performance improvement of a crude oil distillation unit. Appl Therm Eng 2015;75:315–24.
[42] Waheed MA, Oni AO, Adejuyigbe SB, Adewumi BA, Fadare DA. Performance enhancement of vapor recompression heat pump. Appl Energy 2014;114:69–79.
[43] Steinhagen R, MÜller-Steinhagen H, Maani K. Problems and costs due to heat exchanger fouling in New Zealand industries. Heat Transf Eng 1993;14(1):19–30.
[44] Lee SY, Kim KY. Design optimization of axial flow compressor blades with three-dimensional Navier-Stokes solver. KSME Int J 2000;14(9):1005–12.
[45] Yang M. Air compressor efficiency in a Vietnamese enterprise. Energy Policy 2009;37(6):2327–37.
[46] Waheed MA, Oni AO, Adejuyigbe SB, Adewumi BA. Thermoeconomic and environmental assessment of a crude oil distillation unit of a Nigerian refinery. Appl Therm Eng 2014;66(1–2):191–205.
[47] Ghannadi S, Irani M, Chalaturnyk R. Induction and radio frequency heating strategies for steam-assisted gravity drainage start-up phase. Conference Induction and radio frequency heating strategies for steam-assisted gravity drainage start-up phase, vol. 1. p. 421-444.
[48] Davletbaev AY, Kovaleva LA, Nasyrov NM. An investigation of the processes of heat and mass transfer in a multilayer medium under conditions of injection of a miscible agent with simultaneous electromagnetic stimulation. High Temp 2009;47(4):574–9.
[49] Diaz MS, Serrani A, Bandoni JA, Brignole EA. Automatic design and optimization of natural gas plants. Ind Eng Chem Res 1997;36(7):2715–24.
[50] Chebbi R, Al-Amoodi NS, Abdel Jabbar NM, Husseini GA, Al Mazroui KA. Optimum ethane recovery in conventional turboexpander process. Chem Eng Res Des 2010;88(5–6):779–87.
[51] Bandoni JA, Eliceche AM, Mabe GDB, Brignole EA. Synthesis and optimization of ethane recovery process. Comput Chem Eng 1989;13(4–5):587–94.