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Life cycle assessment of greenhouse gas emissions of upgrading and refining bitumen from the solvent extraction process

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HIGHLIGHTS

- Solvent-extracted bitumen was refined into fuels with and without upgrading.
- Life cycle energy and greenhouse gas emissions were evaluated.
- Energy consumption ranges between 0.74 and 1.03 GJ/bbl of bitumen.
- Emissions range from 84.6 to 138.2 kg CO_2 eq./bbl of bitumen.
- Fuels from solvent-extracted bitumen are less energy- and emissions-intensive.

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ABSTRACT

The vapor solvent extraction process is one of the newly proposed bitumen extraction methods that use less energy and emit fewer greenhouse gases than conventional approaches. Through life cycle assessment, this paper examines the energy requirements and associated greenhouse gas emissions of three pathways of producing transportation fuels from the solvent extraction process: delayed coking upgrading, hydroconversion upgrading, and direct refining processes. A Monte Carlo simulation was performed to determine the uncertainty in the model input parameters. The hydroconversion pathway was found to be more energy- and emissions-intensive (1.03 GJ/bbl and 138.2 kg CO_2 eq./bbl) than delayed coking upgrading (0.86 GJ/bbl and 92.3 kg CO_2 eq./bbl) and direct refining (0.74 GJ/bbl and, 84.6 CO_2 eq./bbl), but it offers a higher yield than delayed coker upgrading. Direct refining of bitumen from a vapor solvent extraction process reduces the overall life cycle emissions of transportation fuels.

1. Introduction

Bitumen is a highly viscous crude and requires in situ methods such as steam-assisted gravity drainage (SAGD) and cyclic steam stimulation (CSS) for its extraction [1,2]. The high thermal energy demand to extract bitumen from a depth of 75–750 m makes those processes energy- and greenhouse gas (GHG) intensive [3]. In 2009 the GHG emissions from the oil sands industry were 39.3 metric tonne (MT) of CO_2 eq. [4]. As the demand for bitumen-based transportation fuel continues to grow, its GHG emissions are projected to increase to 112 MT of CO_2 eq. beyond 2030 [4]. The vapor solvent-based extraction process (SEP) mainly uses pure solvent injected into the reservoir at or above its dew point temperature and can offer reduced energy consumption and GHG emissions compared with the traditional SAGD process [5].

The SEP has been proven to be more effective in transferring heat than steam-based methods [6,7] but suffers from a poor oil penetration rate [8,9]. The solvent penetration rate increases dramatically with diluted bitumen [10]. Furthermore, the quality of the produced

Abbreviations: ADU, Atmospheric distillation unit; AGO, Atmospheric gas oil; API, American Petroleum Institute; bbl, Barrel; bpd, Barrels per day; cp, Centipoise; CSS, Cyclic steam stimulation; DCU, Delayed coker unit; DEA, Diethanolamine; DHT, Diesel hydrotreater; FCC, Fluid catalytic converter; GHG, Greenhouse gas; GOHT, Gas oil hydrotreater; GWP, Global warming potential; HCU, Hydroconversion unit; HEN, Heat exchanger network; HVGO, Heavy vacuum gas oil; ISO, International Organization for Standardization; KHT, Kerosene hydrotreater; LCA, Life cycle assessment; LHV, Lower heating value; LPG, Liquefied petroleum gas; LVGO, Light vacuum gas oil; MER, Maximum energy recovery; NHT, Naphtha hydrotreater; PADD, Petroleum Administration for Defense District; PFS, Plant fuel system; SAGD, Steam-assisted gravity drainage; SCO, Synthetic crude oil; SEP, Solvent extraction process; SGP, Saturated gas plant; SMR, Steam methane reforming; UGP, Unsaturated gas plant; VDU, Vacuum distillation unit; VFF, Venting, flaring, and fugitive

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bitumen is enhanced significantly due to high asphaltene precipitation in the reservoir [11]. Though at a laboratory scale, Rezaei et al.'s results show that the SEP produces more oils with relativity less energy consumption and GHG emissions than SAGD [11].

Bitumen produced from the SEP is partially upgraded and could be sent directly to a refinery or after partial upgrading. The quality of the feed determines the properties of refinery products. Energy consumption and associated GHG emissions assessment based on refinery operation alone does not reflect overall environmental performance, hence it is important to include the upgrading operation and downstream processes such as transportation and use in order to avoid burden shifting between different life cycle stages of the product system. Several studies have been conducted, mainly at laboratory scale, to evaluate the efficiency of SEP with regard to aspaltene deposition [12], to assess the oil recovery potential and environmental benefits using physical and numerical simulations [13], and to study the effects of asphaltene content and solvent concentration on crude viscosity. To the best of the authors' knowledge, there is no study that extensively evaluates and characterizes the environmental performance of the SEP. Quantifying the energy and environmental benefits of this technology at its early stage of development provides useful information to be considered in its design at a relatively lower cost than the technology is widely deployed in the future.

This paper, therefore, aims at characterizing the life cycle environmental profile of transportation fuel from bitumen extracted using the SEP through life cycle assessment (LCA). The life cycle stages are the extraction, upgrading, refinery, transportation, and end use combustion. LCA has been widely used to assess the environmental impacts of a product or a product system by calculating energy and material requirements and related emissions outputs at each stage of its life cycle. An LCA of fuel products can help industry identify the critical issues in the product supply chain and suggest ways to reduce potential environmental impacts in order to comply with low-carbon fuel regulations around the world [14,15].

Several LCA case studies have been carried out to evaluate the environmental performances of bitumen upgrading and refinery processes using different open-source LCA models [16,17]. These models can provide estimates of energy demand and GHG emissions of oil sands operations, but these estimates cannot be attributed to a specific project. Lazzaroni et al. calculated the energy demand and GHG emissions from a project-specific integrated extraction and upgrading plant [18]. Giacchetta et al. investigated the economic and environmental factors of an industrial-scale SAGD extraction plant [19]. However, these models could not accurately model bitumen from the SEP, which has specific crude properties that are widely different from SAGD bitumen and other conventional crudes. Furthermore, the models do not show the energy and GHG emissions contributions of each upgrading and refinery unit operation with changes in the configuration (e.g., the replacement of delayed coking upgrader by a hydroconversion upgrader) [20]. The energy consumption and associated GHG emissions could be affected significantly for several reasons. Brandt identified the following factors to have significant impact on the variability and uncertainty of the results: the choice of specific projects or industry average, the scope of modeling, the variability in the assumption of energy intensities of extraction and upgrading processes, the fuel mix assumptions, the treatment of fugitive and venting emissions, and the consideration of emissions due to land-use change [21].

Therefore, this paper develops a bottom-up, Excel-based LCA model to trace the energy consumption and GHG emissions from each sub-unit operation of bitumen extracted using SEP technology. A similar approach has been used by others [22,23], but only for certain crude types. Some studies analyzed the impact of crude quality and refinery configuration but did not consider upgrading and refinery operations together [24,25]. Charpentier et al. [26] and Bergerson et al. [27], however, reported energy consumption and emissions ranges based on confidential industry data for the thermal extraction process, but the results are specific to the models and are not applicable for other bitumen assays. In order to fill these research gaps and accurately estimate energy consumption and GHG emissions from upgraders and refinery operations of partially upgraded bitumen from the solvent-based extraction process, this paper presents a detailed data-intensive simulation and an LCA model. Three project-specific production pathways were considered in the model to compare their life cycle environmental sustainability performance. The specific objectives are to:

- Estimate path-specific life cycle energy consumption and GHG emissions from bitumen processing in three different pathways and determine the best production pathway.
- Conduct uncertainty analyses to generate conservative ranges of GHG emissions through variations in the most energy- and emissions-sensitive parameters

2. Method

2.1. Goal and scope definition

The main purpose of the study is, through process modeling, to estimate path-specific energy consumption and GHG emissions of transportation fuel production from bitumen. Three pathways – delayed coker upgrading followed by refining as pathway I, direct refining as pathway II, and hydroconversion upgrading followed by refining as pathway III – are compared based on the after extraction to the refinery fuel life cycle. Uncertainty and sensitivity analyses are done to measure how realistic the results are and to identify parameters that are sensitive to energy and GHG emissions. Such a comparative LCA helps to understand the environmental trade-offs among alternative upgrading and refinery options in order to make informed decisions.

2.1.1. System boundary

The system boundary defines the main processes included in the assessment. Energy and material requirements at each life cycle stage are calculated using the functional unit of refining, 1 bbl of bitumen. Energy is mainly in the form of natural gas, purge gas, and electricity from Alberta's grid mix. Natural gas is burned in the furnace to heat the crude to its vaporization temperature, and steam is used to strip distillation products from fractionating columns [24]. Fig. 1 illustrates the main systems included in each alternative pathway. The dotted green lines show the boundary of the production pathways considered in this study. The SCO produced from upgrading has properties similar to conventional crudes [1] and can be converted into transportation fuels through refining operations. However, since the bitumen produced from the SEP is partially upgraded, it is important to assess whether the deep conversion refineries can convert the bitumen into transportation fuels without upgrading. Detailed flow diagrams for delayed coker and hydroconversion upgraders are given in Fig. 2(a) and (b), respectively.

2.1.2. Upgrading

The upgraders were modeled in Aspen HYSYS 8.8 software. Diluent is separated in the atmospheric distillation unit (ADU). For both upgraders, naphtha and diesel are separated from the ADU and sent to naphtha and diesel hydrotreaters, respectively. Atmospheric residue from the ADU bottom goes to the vacuum distillation unit (VDU) and is further fractioned into light and heavy vacuum gas oils (VGOs). The delayed coker and hydroconversion upgraders are similar in configuration except that vacuum residue is treated in a different conversion unit (Fig. 2). Vacuum residue is fed into a coker (delayed coker upgrader) or hydrocracker (hydroconversion upgrader). In delayed coker upgrader operations, the coking process requires a long reaction time in the liquid phase (12–14 h per cycle) to convert the residue fraction of the feed into gases, distillates, and coke [1]. The coke is a highly aromatic byproduct rich in sulfur, nitrogen, and metals [1]. The hydrocracker process, on the other hand, uses high-pressure hydrogen and a

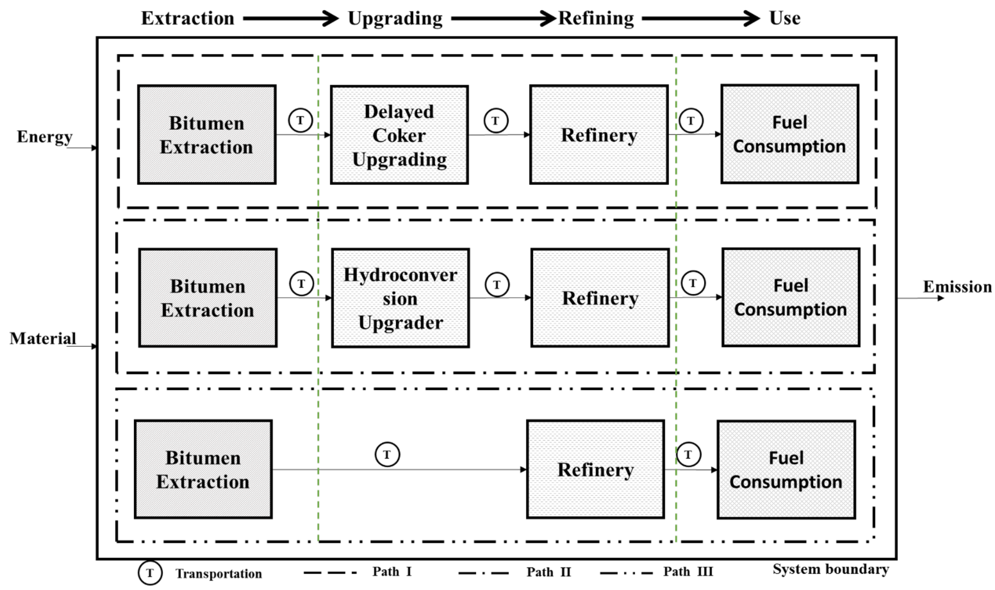

Fig. 1. System boundary for the three bitumen processing pathways.

bifunctional acid-cracking catalyst to convert the vacuum residue [1]. The hydrocracking process first hydrotreats the feed to reduce sulfur and nitrogen levels and make the acid-cracking catalyst effective before the actual conversion takes place.

Naphtha, diesel, and gas oils produced in primary upgrading are fed to the hydrotreaters in secondary upgrading to remove selectively the heteroatoms with little attendant cracking [1]. The operating temperatures considered were found in work by Robinson and Dolbear [28]. Hydrogen is supplied from the on-site steam methane reforming (SMR) plant. Though it depends on the type of feed and quality of hydrocracker feed [20], a huge amount of hydrogen is normally required in the hydrocracker. Natural gas is supplied in the SMR plant both as feedstock and as fuel to produce hydrogen [29]. The amount of natural gas, steam, and electricity required to produce a per unit volume of hydrogen is taken from a report by Spath and Mann [30].

Both delayed coking and hydroconversion operations produce purge

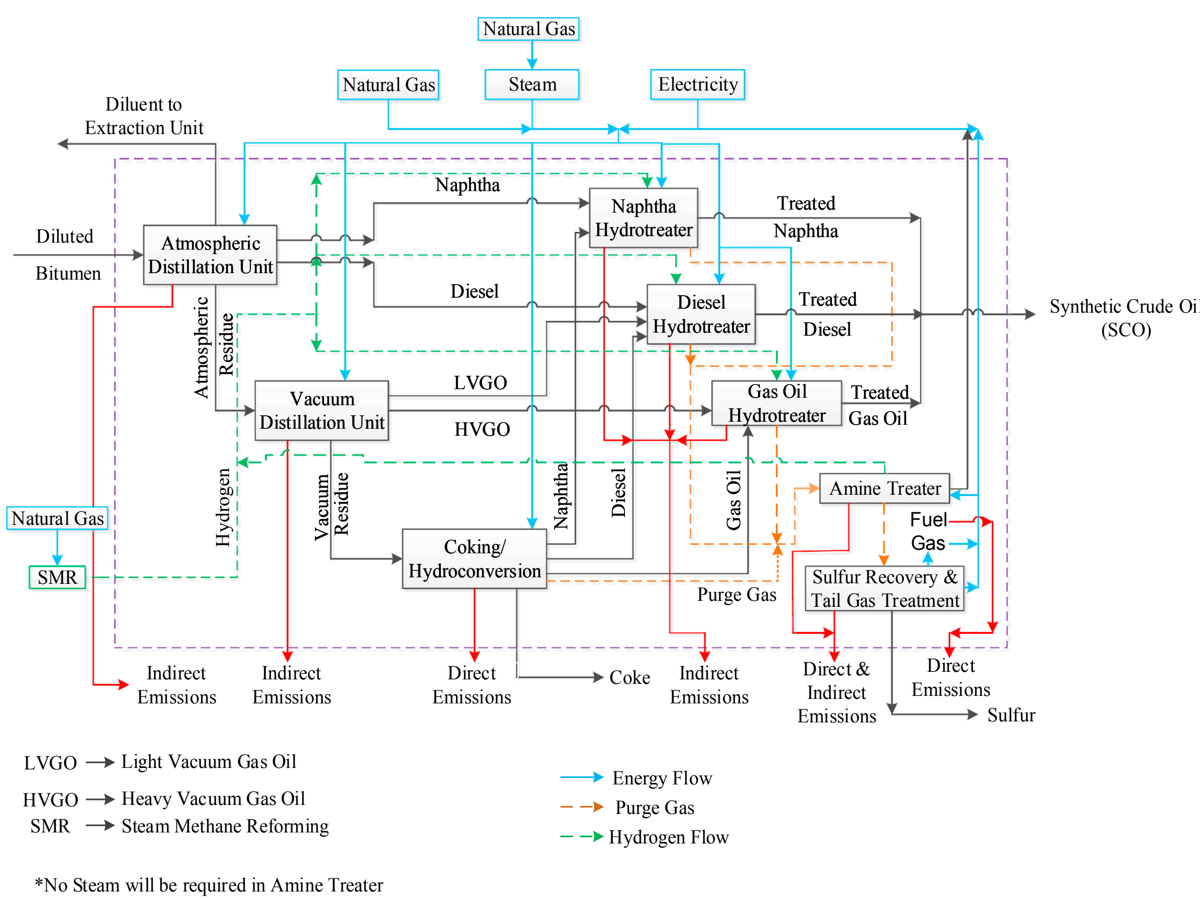

Fig. 2. Sub-unit operations of the delayed coker and hydroconversion upgrader.

gases such as effluent [1]. Since purge gases have low hydrogen and olefin content, they cannot be recovered as separate streams [1]. Ammonia and hydrogen sulfide in the purge gases are removed and used as fuel gas in the fuel gas burner with the assumption that they have the same efficiency as a natural gas (NG) furnace. Diethanolamine (DEA) is assumed to be used to remove the acid gases from the feed. The hydrogen produced in the amine treater is used to supplement the hydrogen required for the process. All the acid gases are transferred to the sulfur recovery unit to convert H_2S into raw sulfur. A modified Claus process with 92–98% efficiency was considered because of its low cost. A tail gas treatment unit is used to treat additional sulfur from the effluent gas. Steam produced as a byproduct from the Claus unit is used to supplement the steam requirement.

2.1.3. Refining

SCO from upgraders and dilbit from the extraction unit are fed into a deep conversion refinery to obtain gasoline, diesel, jet fuel, and other end products [31]. Diluent is separated from the ADU of the upgrader or refinery. A refinery is a complex process plant that makes products based on feedstocks, correlations, and processing units.

A typical deep conversion refinery has a coker and a catalytic cracking unit [32]. Processing units like hydrotreaters (HTs), fluid catalytic crackers (FCC), reformers, hydrocrackers, and alkylation units are simulated in the HYSYS petroleum shift reactors, which operate on the delta base-shift concept. All the input and output fractions are calculated in HYSYS using equations and correlations from the literature. The gas streams containing saturates are directed to the saturated gas plant (SGP). After separation, the methane gas goes to the SMR plant to supplement the hydrogen requirement. The total C_2 and fractions of C_3 and C_4 gases go to the plant fuel system (PFS). The produced fuel gas is burned to generate heat. Unsaturated gases containing olefins from the FCC and coker unit go to the unsaturated gas plant (UGP). From the UGP, ethylene is directed to the PFS, and propylene and butylene are transferred to the alkylation unit. Alkylate produced in the alkylation unit is mixed with other components to form liquefied petroleum gas (LPG). Similar to upgraders, the utilities in the refinery are fuel, steam, and electrical power. The model uses base correlations to predict the utility energy consumption values in each sub-unit operation of the refinery.

Energy is consumed in the refinery in the form of heat, electricity, and steam. Natural gas and refinery fuel gas are used to meet heating and steam demands. Steam produced in the SMR and sulfur recovery unit is used to meet additional steam demand. For the refinery operation, all of the electricity is supplied from the grid. Similar to upgraders, natural gas is supplied as fuel and feedstock in the on-site SMR plant to produce hydrogen. All the hydrogen required as feedstock in the SMR plant is supplemented by the methane produced in the SGP. The hydrogen produced in the reformer supplements the entire refinery hydrogen requirement.

Given the lack of data, all the refining operations are simulated in the default deep conversion refining model developed in HYSYS. In HYSYS, the dependent variables such as product flow rates, qualities, utilities, etc., are determined by the base correlations specified for different independent variables such as feed type and flow rate [33]. In order to adjust the deviations in the independent variables from a base point, HYSYS uses a set of linear equations that are developed by differentiating the independent variables with respect to the dependent variables at their base points [20]. The derivatives of the independent variables represent the coefficients of the equations. The step-by-step simulation procedures for the upgrading and refinery are as follows:

I. A fluid package is selected as input feed. In this study, to analyze the thermodynamics of bitumen and SCO refinery operations, the Peng-Robinson equation of state [34] is considered appropriate.
II. Bitumen properties are characterized.
III. Upgrading and refining process flow models are developed.
IV. Process flows are simulated in models to generate data for energy and emissions analysis.

2.1.4. Feedstock transportation

Although bitumen produced from the SEP has better properties than thermal extraction, it does not meet pipeline requirements. Hence bitumen is mixed with lighter hydrocarbons like naphtha or natural gas condensate to meet the pipeline density (API gravity > 19) and viscosity (< =350 centistokes [cSt]) [1]. This mixture is called dilbit. The distance between the extraction site and upgrading facility is considered to be 500 km, which is the distance between the extraction site located in Fort McMurray and the upgrading unit in Fort Saskatchewan [32] (see Supporting Information). It is assumed that the refineries are 1884.5 km from both the upgrading and the extraction units. This is the distance of the Keystone Pipeline that transports bitumen from Hardisty, Alberta to the American Midwest markets in Illinois and Oklahoma [35]. A pipeline transportation model was developed to estimate the energy consumption and GHG emissions from transporting dilbit and SCO including the diluent return to the extraction site. The Keystone Pipeline can transport 590,000 bpd of feed [35].

2.2. Life cycle inventory analysis

The life cycle inventory analysis includes the data collection process for both the foreground and the background product systems. The data are used both in process modeling and in energy and GHG emissions calculations.

Table 1 lists the bitumen properties for the feed in the HYSYS upgrader and refinery models. These numbers are from Nsolv [36].

Tables 2 and 3 list the input data used to develop the upgrader models and purge gas recovery processes, respectively. The amount of natural gas, steam, and electricity required to produce a unit volume of hydrogen is taken from a study by Spath and Mann [30].

2.2.1. Mass and energy analysis

The basic modeling equations are derived based on the following considerations:

- The system is assumed to be in a steady-state flow condition.
- Changes in kinetic and potential energy are negligible.
- The reference states are considered to be: $T_o = 25\,^{\circ}\mathrm{C} = 298.15\,\mathrm{K}$ and $P_o = 101\,\mathrm{kPa}$.
- Change in mole percent is assumed to equal change in volume percent.

The upgrading and refinery simulation models were developed based on engineering first principles, using mass balance (Eq. (1)) and energy balance (Eq. (2)) equations:

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

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

where i refers to input, e refers to output, and m and E are mass and

Table 1
Properties of bitumen assay [36].

API gravity	14.0
Density (kg/m^3)	972.5
Sulfur (wt%)	4.6
Asphaltene (wt%)	3.0
Vacuum residue yield (538 °C+) (vol%)	45.4
Metals:	
Nickel (ppm)	30.0
Vanadium (ppm)	77.0
Ferrous (ppm)	5.0

Table 2
Input data used to develop upgrading models.

Hydrogen consumption	Delayed coker	Hydroconversion	Source
Naphtha hydrotreater (scf/bbl)	142	100	[53]
Diesel hydrotreater (scf/bbl)	162	259	[53]
Gas oil hydrotreater (scf/bbl)	736	690	[53]
Hydroconverter (scf/bbl)	–	1218	[53]
Hydrogen production in SMR plant:			
NG required (m^3/N m^3 of H_2)	0.04		[30]
NG feedstock required (m^3/N m^3 of H_2)	0.36		[30]
Electricity consumption (kWh/N m^3 of H_2)	0.03		[24]
Steam produced (lb/N m^3 of H_2)	0.86		[24]
Efficiency of equipment:			
NG furnace (%)	87		[62]
NG fired boiler (%)	85		[26]
Heat exchanger (%)	90		[63]

Table 3
Input data used for the purge gas recovery process.

Purge gas composition	Delayed coker	Hydroconversion	Source
Hydrocarbons (mol%)	42.20[a]	24.17 [a]	
Hydrogen (mol%)	53.83[a]	70.74[a]	
Acid gas (mol%)	3.97[a]	5.09[a]	
Amine treater:			
Acid gas pickup (m^3/liter of solution)		4.63	[1]
Reboiler heating (MJ/m^3 of DEA)		280	[31]
Electricity consumption (kWh/m^3 of DEA)		2.60	[31]
Claus sulfur recovery plant:			
Electricity consumption (kWh/ton of S)		98	[31]
Steam production (lb/ton of S)		1215	[31]
Tail gas treater:			
Electricity consumption (kWh/ton of S)		463	[31]

[a] Values are obtained from the HYSYS delayed coker and hydroconversion upgrader models.

energy flow rates, respectively. Q_{cv} and W_{cv} are input heat flow rate and external work rate for a control volume, respectively. Using input energy E_i and output energy E_e, the efficiency of a piece of equipment is calculated as:

$$\eta = \frac{E_e}{E_i} \tag{3}$$

Eq. (4) from Waheed et al. [37] is used to calculate the energy balance across the heat exchangers:

$$(E_{hot,in} - E_{hot,out}) - (E_{cold,in} - E_{cold,out}) = 0 \tag{4}$$

The amount of heat and steam required in upgrader and refinery models was obtained from HYSYS simulation models. Produced gas and natural gas meet the heat and steam requirements. The heat generated from the gas in upgraders and refineries was estimated using Eq. (5):

$$Heat_{produced\ gas}\,(\text{kJ/bbl}) = \sum_{i=1}^{n} \frac{(LHV)_i * m_i}{\eta_f} \tag{5}$$

where LHV refers to the lower heating value (kJ/kg) of an individual gas component i, m is the mass flow rate (kg/bbl), and η_f is the furnace efficiency. The amount of C_4 and C_{4+} compounds in the produced fuel gas was less than 5% (vol) [38], therefore, a LHV of C_4H_{10} was used to calculate the total amount of heat produced from upgrader fuel gas. However, to simplify the calculation, the steam was assumed to be from natural gas and was calculated using Eq. (6):

$$NG\ \ required\left(\frac{\text{kg}}{\text{bbl}}\right) = \frac{Heating\ \ required\ \ from\ \ NG\left(\frac{\text{kJ}}{\text{bbl}}\right)}{\eta_b * NG\ \ LHV\left(\frac{\text{kJ}}{\text{kg}}\right)} \tag{6}$$

where η_b is boiler efficiency, assuming 100% steam quality [39].

The electricity consumed in different sub-units of the upgrader and refinery was determined using correlations from the Abella and Bergerson's PRELIM model [24].

2.2.2. Pipeline pressure requirement and energy consumption equations

To evaluate the total pressure head loss in pipeline transportation, the diluent ratio (DR) is calculated based on the specific gravity of dilbit, diluent, and bitumen (the equations are provided in the Supporting Information). The target velocity and shipped volume of crude are used to determine the actual velocity and approximate inner pipe diameter [40]. The approximate pipe diameter has been compared to a list of API 5L standard pipe diameters. The designed capacity of the pipeline system is calculated based on the diluent ratio and shipped volume of feed. Fluid velocities are calculated considering a standard pipe diameter larger and smaller than the approximate inner pipe diameter. Finally, the standard pipe diameter for which the calculated fluid velocity matches best with the target velocity was selected.

The friction factor was calculated to determine the pressure drop due to friction between the feed and pipe material. For a Reynolds number of the feed below 2100, the flow is assumed to be laminar and the friction factor is determined by $f = 64/Re$, where Re is the Reynolds number. When the Reynolds number is greater than 2100, however, the Haaland friction factor equation is used to calculate the initial friction factor value [41]. The exact friction factor is obtained using the Colebrook friction factor equation with several iterations until the input and output difference is less than 10^{-5} [42]. Finally, pressure loss due to pipe friction is calculated using Eq. (7):

$$P_{fricton} = \frac{fL\rho}{2D}v^2 \tag{7}$$

where L is the pipeline length (m), ρ is the fluid density (kg/m^3), D is the pipeline inner diameter (m), and v is the fluid velocity (m/s).

Eq. (8) calculates the pressure drop due to elevation change:

$$P_{elevation} = h\rho g \tag{8}$$

where h is change in elevation (m) and g is gravity (9.81 m^2/s).

The change in elevation from Alberta to refineries in the American Midwest is considered to be 632 m [40]. The elevation change from Fort McMurray to Fort Saskatchewan is 350 m [43,44]. Once the total pressure loss has been determined, it is used to calculate pipeline pumping energy intensity (see Supporting Information). In pipeline transportation, only electricity is consumed to drive the pumps [26]. The total GHG emissions from pipeline transportation are calculated considering the energy intensity and GHG emission factors. The GHG emission factors are taken from GREET [16]. Table 4 summarizes the dilbit and pipeline specifications.

2.3. Life cycle impact assessment

This research determines the life cycle energy consumption and GHG emissions of bitumen processing. As in an earlier study [22], the three main GHG emissions, carbon dioxide (CO_2), methane (CH_4), and nitrous oxide (N_2O) and their respective characterization factors according to Intergovernmental Panel on Climate Change (IPCC), were considered to calculate global warming potential (GWP) as CO_2 Eq. [45].

The total GHG emissions reported in the model developed for this work include both direct emissions from on-site combustion of natural gas and produced fuel gas and indirect emissions from the recovery,

Table 4
Feed properties and pipeline specifications.

Parameters	Dilbit	Diluent	SCO	Sources
Shipped volume (bpd)	590,000.0	132336.2[α]	590,000.0	[35]
API	21.4	55.0	40.0 (coker), 37.3 (hydroconversion)	[32,40] (SCO API is obtained from the HYSYS model
Fluid velocity (m/s)	1.4	1.4	1.4	[40]
Pipe inner diameter (inch)	39.1	18.5	39.1	Calculated
Pump efficiency (%)	85.0	85.0	85.0	[40]
Outlet pressure (bar)	1.0	1.0	1.0	[40]
Relative roughness (10^{-5})	4.6	4.6	4.6	[40]
Friction factor (10^{-3})	8.0	9.7	8.0	Calculated

[α] Diluent amount is calculated based on a diluent ratio (DR) of 22.43:77.57 of diluent and bitumen

processing, and transportation of natural gas, electricity generation, and the life cycle of materials used as inputs in the defined product system. The model considers venting, flaring, and fugitive (VFF) emissions but not land-use emissions.

The GHG emission factors to determine the total GWP from the natural gas used in the upgraders are taken from GREET [16]. The GHG emission factors for the fuel gas are taken from Netzer [46] (2419.40 g-CO_2 eq./kg). Only upstream emissions were considered for the natural gas supplied as feedstock in the SMR plant. Emissions from steam reforming reactions are consistent with those of Nimana et al. [20]. For the electricity emissions factor in upgraders, Alberta's grid mix numbers provided in the National Inventory Report for 2014 are used [47]. The LHV was used to define the energy content. The LHVs of the produced fuels are provided in the Supporting Information.

The GHG emission factors for the refinery are consistent with upgrading units except for the electricity grid. The US electricity grid mix emissions factor of 581 g-CO_2 eq./kWh [20] is considered. The VFF GHG emissions for refinery operations are also extracted from the National Inventory Report [48].

2.4. Uncertainty analysis

Sensitivity analysis allows us to observe the impacts of input parameters on output results. A sensitivity analysis was conducted to determine the impacts of energy-intensive parameters on net GHG emissions from upgrading and refining. The parameter values were varied by ± 30% from their base values as consistent with work by Nimana et al. [20]. For some parameters, the variation range is reduced to ± 10% to generate meaningful results.

To improve the model reliability and determine the impact of the most sensitive parameters on energy consumption and GHG emissions, uncertainty analysis using a Monte Carlo simulation was performed. The sampling error is less than 0.1 kg CO_2 eq./bbl of bitumen. To maintain that sampling error, sample size is calculated with Eq. (9):

$$\textit{Sampling error}, \bar{X} = \frac{z * \sigma}{\sqrt{n}} \tag{9}$$

where σ is the standard deviation of the mean and n is the number of samples. z values are determined based on the confidence interval of the standard normal distribution. The z values are provided in the Supporting Information.

3. Results and discussion

This section discusses the main results in terms of energy consumption and GHG emissions for each life cycle stage of bitumen processing.

3.1. Upgrading

The SCO produced from hydroconversion and delayed coker upgrading is 127.2 and 117.9 vol% bitumen, respectively. The SCO quality from the two upgraders differs. The volume yield from SAGD bitumen ranges from 78% to 94% in delayed coking and 95–106% in hydroconversion [20,26,27,49]. A higher SCO yield is obtained from both upgrading units because of the quality of bitumen feed processed.

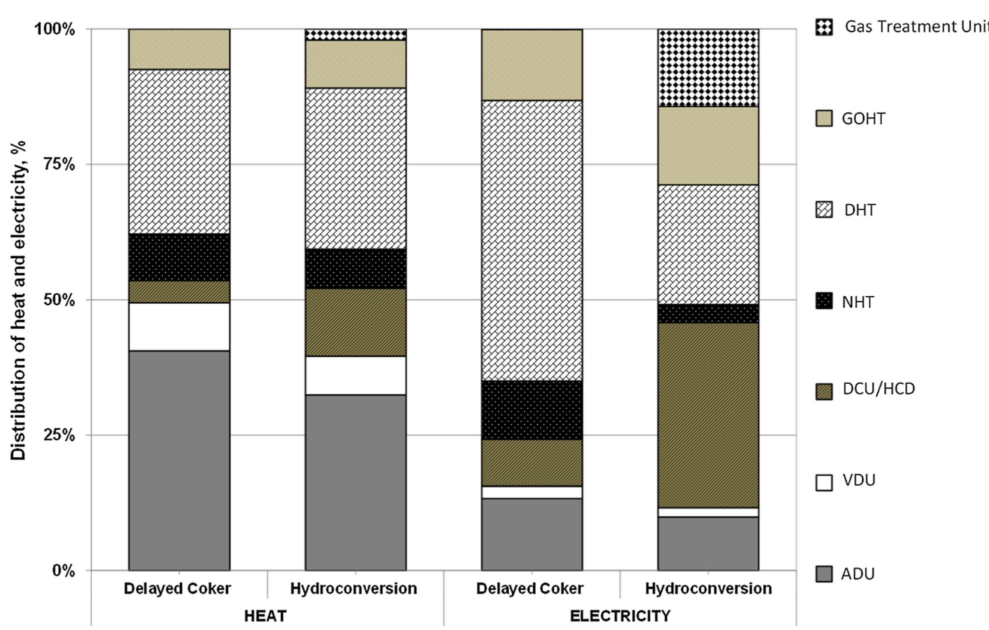

Fig. 3. Distribution of heat and electricity consumed in delayed coker (0.40 GJ/bbl) and hydroconversion upgrading (0.50 GJ/bbl).

Table 5
Upgrading energy consumption and GHG emissions.*

	Energy consumption (kg/bbl of bitumen)		GHG emissions (kg CO_2 eq./bbl of bitumen)	
	Delayed coker	Hydroconversion	Delayed coker	Hydroconversion
Fuel gas	0.08	6.21	11.18	15.14
Natural gas:				
Heating	8.74	11.86	25.63	22.41
Steam[α]	2.22	4.23	6.50	12.39
Electricity	6.77	9.07	5.98	8.01
VFF emissions[β]			4.77	4.77

* The results do not include SMR energy consumption and GHG emissions.
[α] Natural gas required for steam production is determined considering a steam LHV of 2.79 MJ/kg [55].
[β] VFF emissions are calculated using a bitumen LHV of 6.48 GJ/bbl [20] and emission factors from the National Inventory Report [49].

The total energy consumed to upgrade bitumen depends on the feed quality and the process used. Excluding energy consumption in the SMR plant, the total energy from delayed coker and hydroconversion upgraders is 0.40 GJ and 0.50 GJ per bbl of bitumen, respectively. Fig. 3 shows the distribution of heat and electricity consumed in the delayed coker and hydroconversion upgrading units. The total GHG emissions, excluding the SMR plant, are estimated to be 54.28 and 80.30 kg CO_2 eq./bbl of bitumen for delayed coker and hydroconversion, respectively. Table 5 summarizes the GHG emissions and energy consumption contribution from each source.

A pinch analysis was performed to design the heat exchanger network. The maximum energy recovery (MER) design as described by Waheed et al. [50] was obtained for a minimum temperature difference of 21 °C between hot and cold streams in the delayed coking. For hydroconversion, the MER design was obtained at a minimum temperature difference of 20 °C. In delayed coker upgrading about 46.13% of the total heat energy is recovered through heat exchange between the feed and product streams. In the case of hydroconversion upgrading, the minimum temperature difference is 20 °C and the maximum heat energy recovered is 36.81% of the total heat required. Since the design has been optimized and the bitumen fed to the upgraders is upgraded, the energy required in both upgrader types is lower than in SAGD. Of the net energy required, electricity demand ranges from 6.77 kWh/bbl of bitumen in the delayed coker unit (DCU) to 9.07 kWh/bbl of bitumen in the hydroconversion unit (HCU). Total heat and electricity consumption distributions in DCU and HCU sub-unit operations are shown in Figs. 5 and 6, respectively.

The distillation columns, both atmospheric and vacuum, are the main energy-intensive components in both upgrading units. Our results show differences in heat and electricity consumption compared with recent results published by Pacheco et al. [51] for both upgrading operations. The variations reflect the impact of crude quality on upgrading sub-unit operations. The diesel hydrotreater (DHT) consumes considerably more heat energy than other units such as the NHT and gas oil hydrotreaters (GOHT). This is because the amount of middle distillate fractions, like diesel and light vacuum gas oil (LVGO) from the ADU and VDU units, are significantly higher than other fractions in both processes.

As shown in Table 5, due to the relatively higher amount of light ends produced in the hydrocracker, the fuel gas produced in the hydroconversion upgrader is significantly higher than in the delayed coker. As the feed has relatively heavier fractions, high temperature and high pressure hydrocracking reactions break the vacuum residue and produce a large amount of light ends [1]. The light ends, after treatment, are used as fuel gas in the hydroconversion upgrading process. VFF emissions depend on the type of reservoir and vary significantly from site to site [52]. Due to the lack of site-specific VFF emissions data, Canada's national inventory data were used and are considered to be the same for both upgrading units [48].

In addition to energy, hydrocrackers and hydrotreaters consume a considerable amount of hydrogen [20,51]. The hydrogen requirements in N m^3/bbl of bitumen are 7.70 and 17.70 for delayed coker and hydroconversion, respectively. The on-site SMR plant supplies hydrogen to all the hydrotreating units in coking upgrading and also to the hydrocracking unit in hydroconversion upgrading. More hydrogen is required in the hydrocracking unit than in the hydrotreaters [53], and the amount of hydrogen used increases the total amount of hydrogen consumed in hydroconversion upgrading. To produce hydrogen, natural gas is supplied both as fuel and feedstock. Steam is produced as a byproduct and used in the process to reduce the total steam requirement in the upgraders. The amount of natural gas and electricity required and the amount of steam produced from the SMR plant are calculated with the data in Table 2. In the SMR plant, emissions occur in the transportation, combustion, and reforming of natural gas into hydrogen. The GHG emissions from reforming reactions in kg CO_2 eq./ bb bitumen are 1.08 and 2.49 for delayed coker and hydroconversion upgrading units, respectively.

Since the bitumen from the SEP is partially upgraded, the SCOproduced in both upgraders has a high API gravity and low sulfur content compared with the thermal extraction methods. The API gravity and sulfur content of the produced SCOs are given in the next section.

3.2. Refining

The deep conversion refinery process model described in the method section was used to determine the final products, total energy consumption, and associated emissions from processing SCO and bitumen. The API gravity, composition, and sulfur content of the SCO obtained from the upgraders are presented in Table 6.

SCO produced from SEP bitumen has relatively higher API gravity than the average API gravity of the SCO from SAGD bitumen as seen in Table 6. The sulfur content of DC and HC SCOs is significantly lower than that from SAGD bitumen as the bitumen obtained from the SEP has inherently low sulfur content.

The feed is assumed to be supplied in the refinery at a rate of 25,000 bpd. The yield obtained per bbl of feed is presented in Table 7. The distillation fractions from refinery ADU and VDU are provided in the Supporting Information. SCO contains more lighter fractions (e.g., naphtha, diesel, and kerosene) than bitumen, which contains more heavier fractions such as gas oil and residue [1,20].

As shown in Table 7, the fuel gas produced from refining per bbl of SCO is significantly higher than the fuel gas produced from refining SCO from SAGD bitumen (0–0.15 bbl/bbl of feed [20,24]). Bitumen from the SEP has heavier fractions than SCO and a negligible amount of fuel gas is produced when bitumen is directly fed into the refinery. As the bitumen is partially upgraded, more liquid petroleum gas (LPG) is produced per bbl of bitumen [20,24]. Gasoline produced from the upgraded bitumen is 73 (vol%) and from SAGD-upgraded SCO is 54 (vol %) [20,24], which indicates that the bitumen produced from the SEP has a higher naphtha fraction than conventional SAGD bitumen [24]. Refining HC SCO produces more gasoline and diesel than refining DC SCO. This is because in hydroconversion upgrading, the feed is

Table 6
Properties of SCO from upgraders [1]

Properties	DC SCO	HC SCO	SCO produced from SAGD bitumen
API gravity	40.04	37.30	19.60–38.50
Naphtha (vol%)	20.29	17.16	8.00–25.00
Diesel (vol%)	57.97	59.39	40.00–59.00
Gas oil (vol%)	21.74	23.45	20.10–40.00
Sulfur content (wt%)	0.02	0.01	0.04–0.40

Table 7
Yield (bbl/bbl) of feed obtained from refining per bbl of different feeds.

	DC SCO	HC SCO	Bitumen
Fuel gas	0.51	0.42	0.00
Liquefied petroleum gas	0.05	0.04	0.05
Diesel	0.06	0.12	0.25
Kerosene/Jet fuel	0.05	0.05	0.01
Gasoline	0.37	0.40	0.73
Fuel oil	0.04	0.08	0.18
Coke (kg/bbl of feed)	0.00*	0.01	0.98

* Coke formation is negligible in DC SCO.

hydrogenated and cracked more intensely. As the fuel gas is not a desired product, so bitumen produced from the SEP should directly be refined without being upgraded to SCO. The produced yield fractions from both SCO and bitumen fall within the broad range of yield fractions reported for the wide range of crude assays [54]. It was found in this study that, unlike the bitumen extracted using SAGD technology, bitumen from solvent extraction may require partial or no upgrading to maximize the amount of gasoline and diesel fuels in the refinery. This technology will provide oil sands-derived transportation fuels producers better control and flexibility to distillate the bitumen during upgrading and refinery operation.

Refining bitumen is the most energy-intensive pathway, and energy consumption is an estimated 68.5% higher than for refining coker SCO. This is because distilling bitumen in the ADU and VDU is far more energy intensive than distilling SCO. The SCO produced from both upgraders also has significantly higher vapor fractions, which results in low energy requirement to distillate fuel gas. The detailed breakdown of energy consumption for refining coker and hydroconversion SCOs and bitumen is shown in Fig. 4.

For SCO distillation, 12.3–12.8% of the total energy is required in the distillation columns, but for bitumen, more than 20% is required. As mentioned earlier, a major fraction of the SCO (both from coker and hydroconversion) is light components (C_4 or less) that distillate at low temperatures, thereby reducing the energy required in SCO distillation. In the case of SCO, the catalytic reformer is responsible for a large portion of energy consumption, mainly to preheat the reactor beds and to regenerate the catalysts [55]. The dehydrogenation and dehydrocyclization reactions, which are highly endothermic, also consume significant amounts of energy. As SCO has a high naphtha content, the catalytic reforming is the most energy-intensive operation. This implies that the SCO produced from the upgraders has a relatively high amount of lighter fractions. For bitumen refining, energy consumption in the reformer is within the range found in other studies [20,24]. A significant amount of energy is also consumed by the SGP unit in refining SCO. The high energy consumption is from treating the high amount of fuel gas produced from SCO, as shown in Table 7. As bitumen is the heaviest feed, energy consumption in fluid catalytic cracking (FCC) is higher for bitumen (19.8%) than for SCO (8.5–13.7%). SCO requires less cracking to convert heavier fractions into final products as it is lighter than bitumen [1]. Energy consumption in the UGP ranges from 5.9 to 14.5%, depending on the feed. The higher production of unsaturated gases from bitumen leads to higher energy consumption than from SCO [20].

As with the upgraders, hydrogen is supplied to the hydrotreaters and hydrocrackers from an on-site SMR plant. Hydrogen produced in the reformer supplements the total H_2 requirement in the refinery. The total energy consumption calculated in the SMR plant is 7.4%–24% of the total refinery energy consumption. Similar energy consumption results have been reported by Abella and Bergerson [24]. The feedstock required to produce hydrogen comes from the SGP and external natural gas supplies. Steam produced as a byproduct from the reforming reaction is supplied to the refinery to reduce the refinery steam requirement.

All of the energy required in the refinery comes from steam, electricity, fuel gas, and natural gas. Steam production accounts for 25.1%–49.6% of the total energy required. Bitumen demands more steam than SCO due to the high energy requirement to distillate heavier fractions [56]. Electricity makes up 8.3%–11.5% to the total energy consumption. In the case of SCO, the fuel gas produced is the main source of heat for the refinery. Although the fuel gas reduces the energy consumption by supplementing the heat requirement, it is not beneficial as the fuel gas is generated from the unstable lighter fractions that ultimately reduce the production yield.

GHG emissions from processing bitumen and SCO are 26.6 kg CO_2 eq./bbl of SCO to 84.3 kg CO_2 eq./bbl of bitumen. As more energy is required to refine heavier feeds, bitumen refining results in higher emissions than SCO.

Since refineries produce different products, proper allocation is

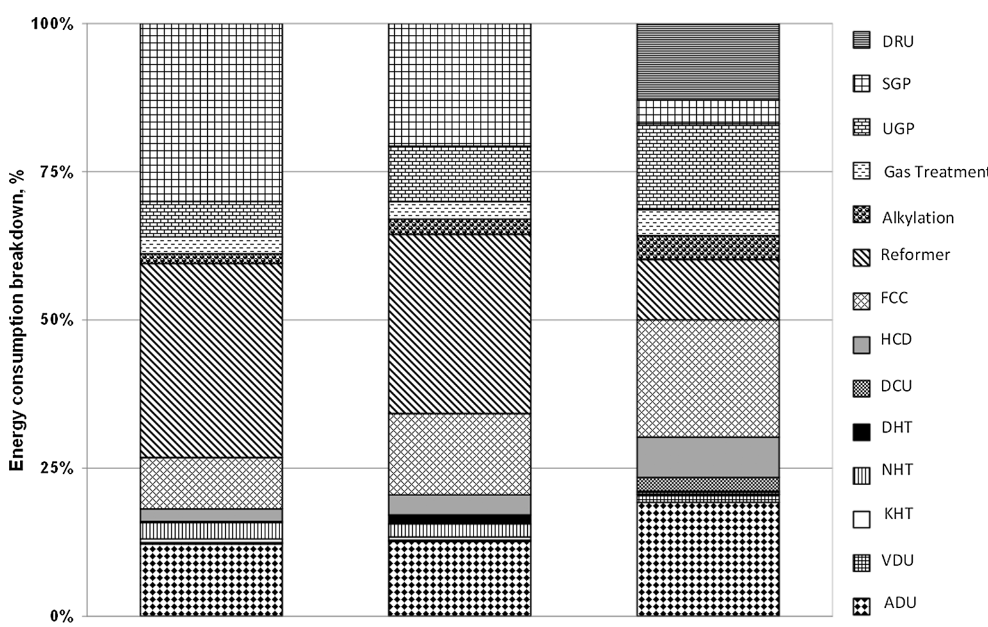

Fig. 4. Energy consumption breakdown in each sub-unit operation of refining coker SCO, hydroconversion SCO, and bitumen.

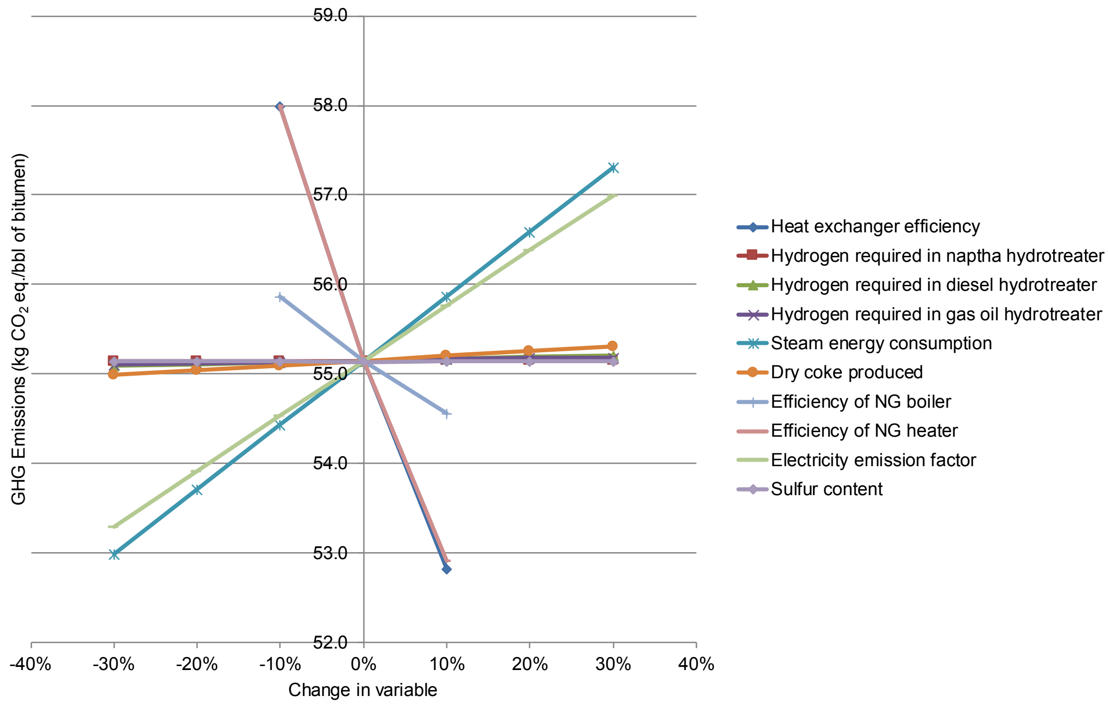

Fig. 5. Sensitivity of key parameters on GHG emissions in delayed coker upgrading.

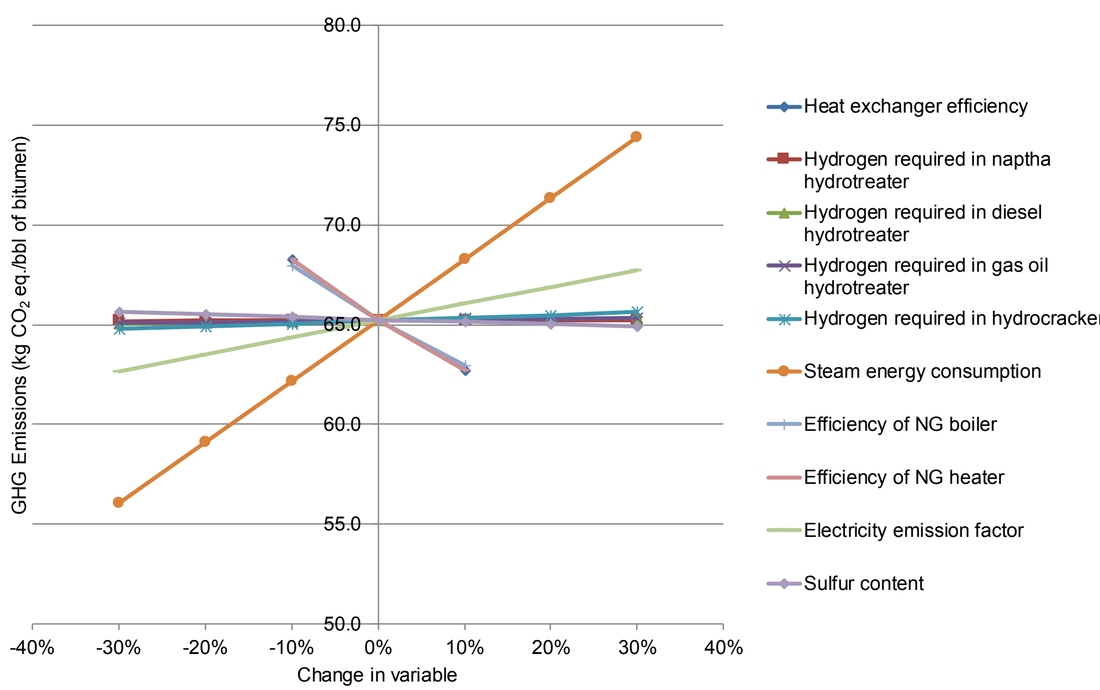

Fig. 6. Sensitivity of key parameters on GHG emissions in hydroconversion upgrading.

required to evaluate the impact of the energy and emissions from the production of individual transportation fuels [57]. For a specific amount of feed, variation in the production volume of a single refined product changes the production volume of other products, ultimately affecting the overall energy consumption and emissions. To allocate energy consumption and emissions more accurately to individual products, the International Organization for Standardization recommends allocation at the refinery sub-process level instead of the refinery level [57]. According to Wang et al., the most energy-intensive transportation fuels are gasoline and diesel [57]. Therefore, refinery sub-process level allocation is used in this study to determine the emissions from gasoline and diesel production and compare them with the results from other models.

There is a proportionate relation between energy consumption and emissions. Because refining bitumen requires more energy, it generates more GHG emissions. GHG emissions from gasoline produced from coker SCO and bitumen (11.2 g CO_2 eq./MJ of gasoline and 11.6 g CO_2 eq./MJ of gasoline, respectively) are almost equivalent but lower than the emissions from gasoline produced from hydroconversion SCO (14.5 g CO_2 eq./MJ of gasoline). For diesel, however, GHG emissions from bitumen are significantly higher (15.5 g CO_2 eq./MJ of diesel) than from coker (9.5 g CO_2 eq./MJ of diesel) or hydroconversion SCO

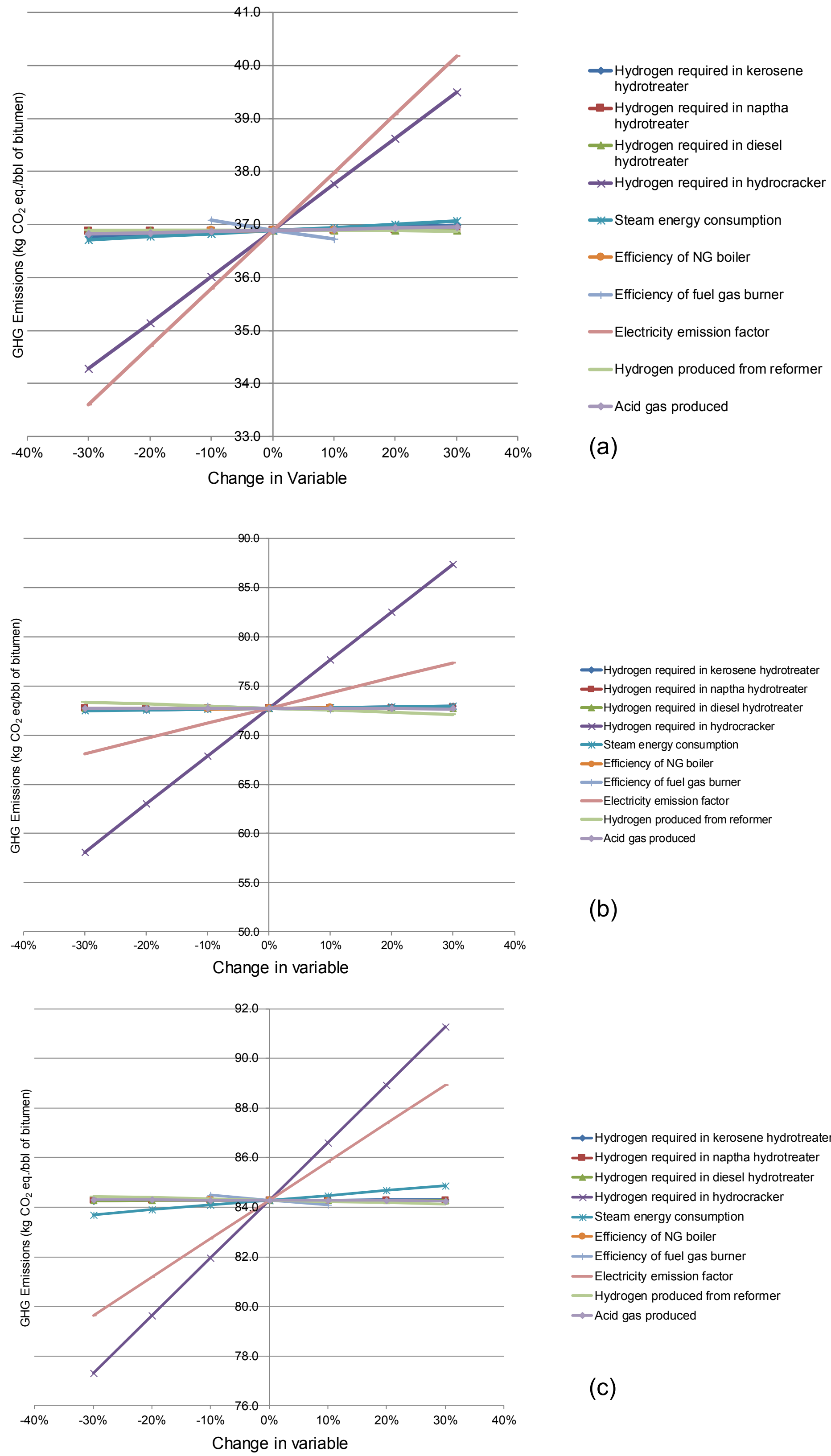

Fig. 7. Impact of key parameters on GHG emissions from refining (a) coker SCO (b) hydroconversion SCO, and (c) bitumen.

(14.3 g CO_2 eq./MJ of diesel). Although the overall energy consumption in refining coker SCO and hydroconversion SCO is less than bitumen's, the lower gasoline and diesel yield from SCO makes the emissions comparable when allocated at the refinery sub-process level. This again justifies the need to allocate emissions at the sub-process level to determine what percentage of energy we are getting back from investing energy and causing emissions.

The energy consumption and GHG emissions in the upgrading and

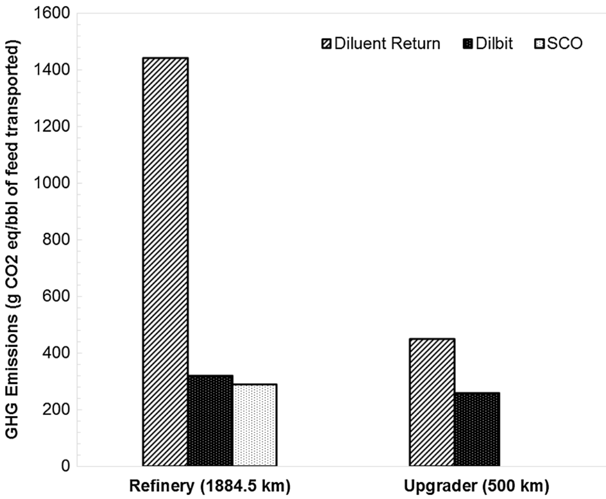

Fig. 8. GHG emissions from the pipeline transportation of feed to the upgrader and refinery.

refining operations calculated in this study depend solely on the bitumen assay provided in a Sustainable Development Technology Canada project [58]. The results are also susceptible to other factors such as plant capacity, operating conditions, etc., of the upgrader and refinery.

3.3. Sensitivity analysis

For the upgrader, energy-intensive parameters were selected from the results shown in Table 5. The parameters investigated are heat exchanger efficiency, NG heater and boiler efficiency, the sulfur content in the feed, electricity emission factors, and hydrogen required in hydrotreating units. The amount of dry coke produced in delayed coking and the hydrogen required in the hydrocracking unit of the hydroconversion upgrader is also considered as sensitive parameters. The effects of steam conditions and steam energy requirement are reflected in the steam energy parameter.

The results in Fig. 5 illustrate that changing parameters by ± 30% changes the net GHG emissions by ± 5.5% in delayed coking. The GHG emissions from coking depend mostly on equipment efficiency, steam energy, and electricity emission factors. The hydrogen required in hydrotreaters has less significance on coking emissions. This might be due to the inherently higher carbon/hydrogen ratio in the bitumen that reduces the hydrogen requirement in the hydrotreaters.

In hydroconversion upgrading, GHG emissions vary with steam consumption, as shown in Fig. 6.

For a ± 30% change in steam energy requirement, GHG emissions vary by ± 11.5%. As with delayed coking, the electricity GHG emissions factor and NG heater efficiency influence the GHG emissions significantly. The hydrogen required in the hydrocracker is less important because the hydrocracker uses bitumen feed of better quality, which requires less hydrogen for hydrocracking.

For the refinery, a sensitivity analysis was performed on the key parameters that impact the energy efficiency and performance of the operations. The following parameters are considered: the hydrogen required in the hydrotreaters and hydrocracker, the total steam energy consumption, the hydrogen produced from the reformer, the total acid gas produced (CO_2 + H_2S), and the efficiency of the equipment. SGP and UGP unit operating conditions were not included in the sensitivity analysis because the energy consumption calculated in these units was due to higher fuel gas production, which is unusual in a refinery. To determine sensitivity, all the parameters except equipment efficiency were varied by ± 30% from base conditions.

The efficiency of the equipment was varied by ± 10% from base conditions. As shown in Fig. 7, the two most sensitive parameters are the hydrogen required in hydrocracking and the electricity GHG emissions factor for electricity production. As mentioned earlier, the hydrogen required in hydrocracking is higher than in hydrotreating, and varying the H_2 requirement in the hydrocracker significantly affects the energy consumption and ultimately the GHG emissions. The electricity emission factor depends on the technology and the type of fuel used to generate electricity. In upgrading, the GHG emissions factor of on-site combustion of natural gas and the distribution of electricity is considered to be 820 g CO_2 eq./kWh [47]. It is higher than the refinery electricity emissions. But depending on the technology and type of fuel used to generate electricity, emissions can be as low as 428.31 g CO_2 eq./kWh [16].

Apart from the two main parameters, the efficiency of the fuel gas burner in the refining coker SCO case and the total steam consumption in the refining bitumen case also influence the GHG emission results. The total steam consumption depends on steam pressure, quality, and water inlet temperature[59].

3.4. Pipeline transportation emissions

The GHG emissions from the transportation of bitumen and SCO and to return diluent to the extraction site are shown in Fig. 8. Returning diluent is more emissions-intensive than transporting bitumen or SCO. This is because a smaller pipeline is used to transport diluent due to the lower shipping volume. A smaller pipe causes more friction due to the high friction factor and thus more energy is lost. Transporting dilbit to the refinery is more energy- and emissions-intensive than transporting SCO because SCO is lighter and less viscous than dilbit. The results are consistent with the results presented by Nimana et al. [40]. It is also noticeable that with the increase in distance, energy consumption and emissions also increase.

GHG emissions from transporting diluent, dilbit, and SCO are significantly lower than emissions from upgrading and refinery operations. Because the emissions are low, the transportation parameters are not included in the uncertainty analysis.

3.5. Pathway comparison

The production of gasoline and diesel from SCO is less energy- and GHG emissions-intensive than from bitumen. But a significant amount of energy is required in upgrading operations, which results in considerable GHG emissions. Since the GHG emissions from gasoline production are almost the same for coker SCO and bitumen, pathway II would be less energy- and emissions-intensive than pathways I and III, given that more gasoline is produced than diesel (as shown in Table 7). A comparative analysis was made by allocating upgrading emissions on a mass basis following the approach taken by Wang et al. [57] and adding them to the refinery emissions for pathways I and III. The equation for allocation is provided in the Supporting Information.

As shown in Fig. 9(a), the GHG emissions from gasoline production are 34.2, 11.9, and 42.8 g-CO_2 eq./MJ of gasoline in pathways I, II, and III, respectively. Depending on the pathway, the GHG emissions for diesel production range from 15.7 to 117.5 g-CO_2 eq./MJ of diesel. The GHG emissions shares are shown in Fig. 9(b).

As shown in Fig. 9 for both gasoline and diesel, pathway II (refining bitumen) is the least emissions-intensive path. If upgrading is not required and bitumen can be refined directly, energy consumption and emissions will be reduced, and the production path will be similar to the production of transportation fuels from conventional crudes. This production pathway will also reduce transportation fuel production costs significantly because it will eliminate the need to build upgrading units. Refining extracted bitumen is only possible when it is partially upgraded during the extraction process.

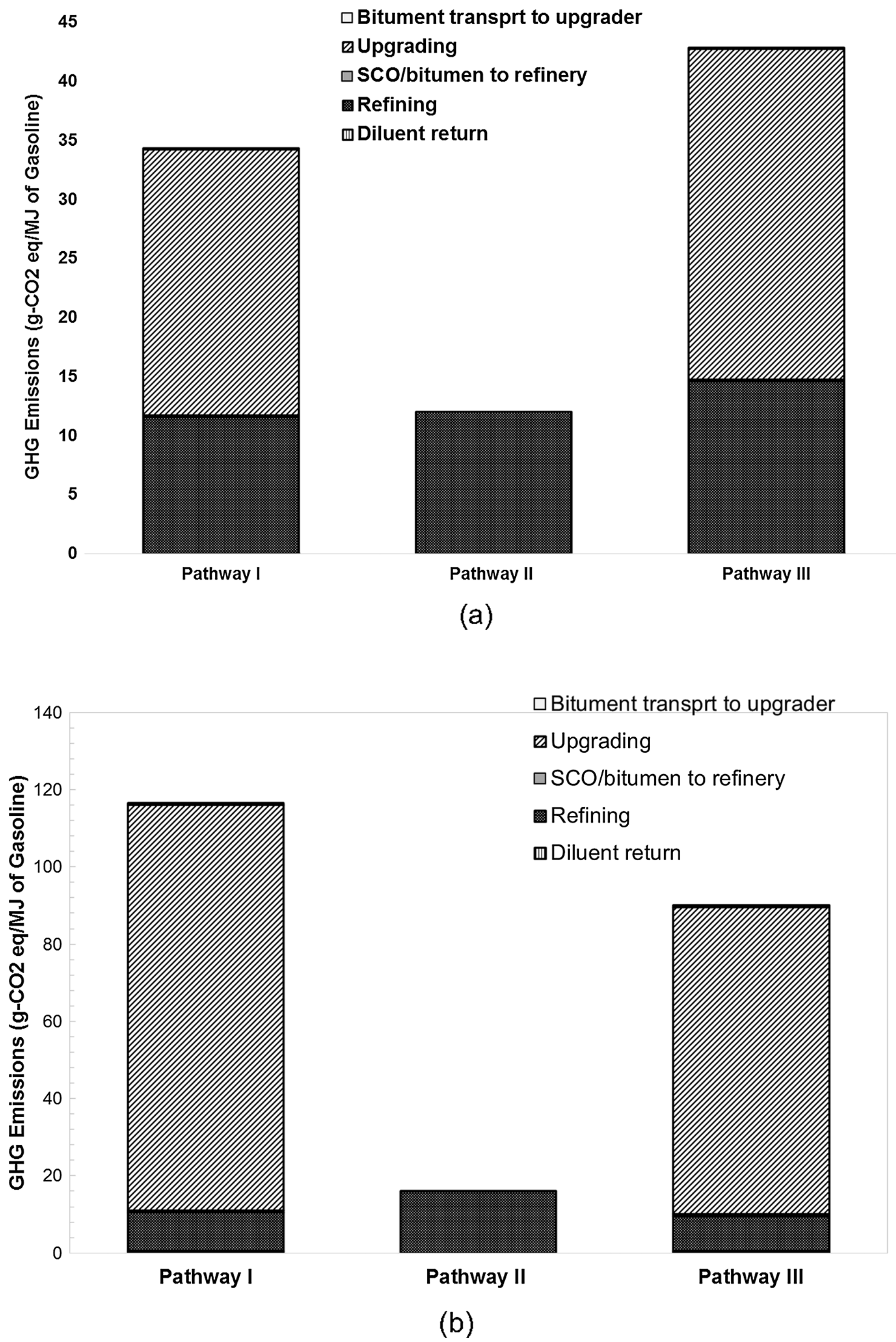

Fig. 9. (a) Gasoline and (b) Diesel GHG emissions' comparison among the pathways.

3.6. Uncertainty analysis

Emissions also vary significantly because of changes in key parameters, as described in Sections 3.3. In order to accommodate the uncertainties/changes in emissions, we conducted an uncertainty analysis through a Monte Carlo simulation. To conduct a Monte Carlo simulation, the distributions of the input parameters first need to be defined. The range of data considered to develop the input distributions is provided in the Supporting Information. For equipment efficiency and steam heating value, a PERT distribution was used in the Monte Carlo simulation. A PERT distribution emphasizes the most likely value rather than maximum and minimum estimates [60]. A PERT distribution was thus considered for NG heaters, boilers, and fuel gas burners, as their efficiency does not vary extensively. PERT generates an output distribution similar to normal distributions without the input parameter value. As a result, a more conservative range of output values will be obtained, which means that output estimates will not vary significantly from the exact value. For the rest of the sensitive input parameters (electricity emission factors, heat exchanger efficiency, and hydrogen required in the refinery hydrocracker), a triangular distribution was used. Although electricity can be produced from a wide range of technologies, electricity production is shifting from coal to natural gas and renewable energy sources, which will reduce electricity GHG emissions [47]. To obtain more conservative outputs from the most

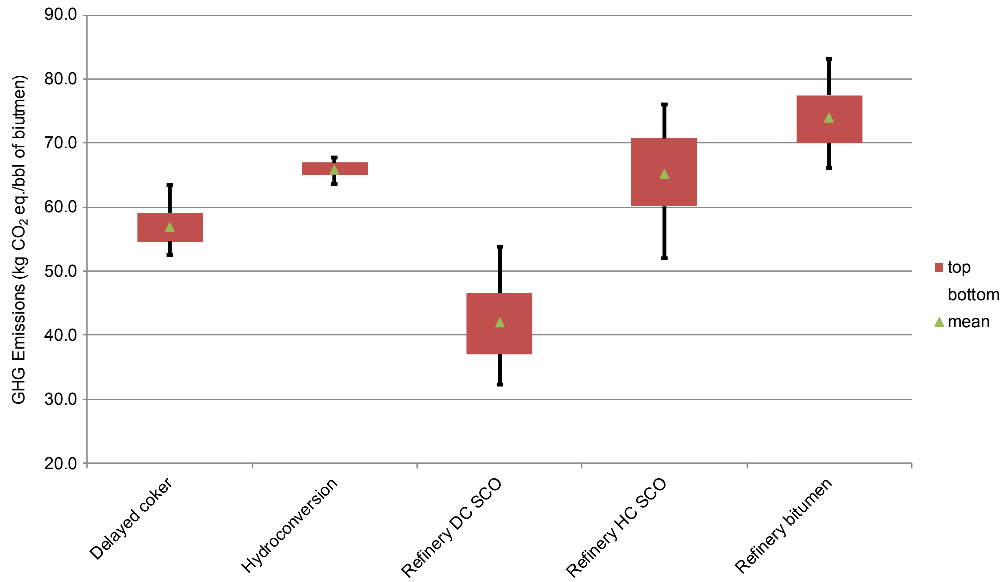

Fig. 10. Uncertainty in emissions from upgrading and refining operations.

predictable input values and to get a lower standard deviation, a triangular distribution was considered for the heat exchanger efficiency. Di Lullo et al. have done similar work [20]. The hydrogen required in the refinery hydrocracker depends on the feed quality and the feed rate to the hydrocracker. As the feed quality causes higher uncertainties in emissions results, a triangular distribution with a wider range was generated using the value calculated here as the mode value. The maximum and minimum values were collected from existing studies [16,20,22,24]. The wide range with the triangular distribution generates an output emission range that emphasizes the feed quality.

To ensure the sampling error is less than 0.1 kg CO_2 eq./bbl of bitumen, approximately 100,000 samples are required for each simulation. The sample size has been determined considering the highest standard deviation from the input distributions to minimize variations in reproducing the experiments. In addition to the highest standard deviation, results were generated within a 99% confidence interval. The ranges taken for developing all the input distributions except the hydrogen required for refining hydrocracking operation are established. To develop conservative distributions for the hydrogen required for refining hydrocracking operations in order to maintain accuracy and reproducibility of results, a wide range of hydrogen required considering light to extra heavy feed is considered. So, the output results are reproducible with almost negligible variation unless the experimental conditions are changed significantly. The samples were simulated in ModelRisk [61]. In Fig. 10, emissions are presented within 5%–95% confidence intervals.

Detailed uncertainty emissions including the impact of each critical parameter for upgrading and refinery operations considered for all the pathways are provided in the Supporting Information. Only the uncertainties in emissions from upgrading and refining operations are discussed here. The mean is the 50th percentile while the top and bottom of the rectangular boxes are the 75th and 25th percentiles, respectively. Accordingly, the top and bottom error bars show emissions at the 95 and 5 percentiles, respectively. Refinery operations show a wider range of emissions than upgrading operations. Also, the emissions range is higher for refining SCOs than for bitumen. The wider range signifies the range of hydrogen required to hydrocrack different qualities of refinery feed (see Supporting Information). For upgrading operations, the uncertainty in emissions is mainly because of the electricity emission factor, which depends on the fuel and technology used to generate electricity. Depending upon the heat exchanger, emissions can vary considerably for coker upgrading, but as mentioned earlier, the efficiency of the heat exchangers does not vary significantly.

4. Conclusions

A detailed comparative analysis of transportation fuels produced from solvent-extracted bitumen was conducted by developing three production pathways. Refining bitumen directly consumes 68.5% more energy than refining SCO. But bitumen needs to be upgraded to produce SCO, and upgrading is an energy- and emissions-intensive process that increases overall GHG emissions, as found in pathways I and III but not in pathway II as it does not require upgrading. The transportation of bitumen, dilbit, and SCO contributes negligible shares of the total GHG emissions for gasoline ($\geq 1\%$) and diesel ($> 1.5\%$) production. When we allocate emissions at the refinery process level, we find the emissions in pathway II are comparable to those from gasoline and diesel produced from conventional crudes.

On the whole, when the pathways are compared, pathway II is the preferable option to convert bitumen from solvent extraction into transportation fuels. Bitumen extracted from thermal extraction methods, like SAGD or CSS, requires more energy in the upgrader before refining into products such as gasoline, diesel, etc. The bitumen extracted from the vapor solvent extraction process has better properties. It can be easily refined without upgrading, as shown in pathway II. In addition, pathway II reduces production costs with the elimination of upgrading operations. The overall range of uncertainty in refinery operations is wider than in upgrading operations, which is attributed to the higher number of energy- and emissions-intensive parameters in refining than in upgrading operations. The variations due to changes in energy and emissions-insensitive parameters have minor impacts on the results.

The findings of this study will help the oil sands industry to implement a better extraction technology to produce transportation fuels from oil sands with life cycle GHG emissions comparable to those of conventional crudes. In addition to minimizing emissions, there might be significant economic benefits from solvent extraction, as our study shows that solvent-extracted bitumen can be refined directly without being upgraded. Furthermore, the production pathways described here are transparent, which will help policy makers and economists determine how to allocate Canada's oil sands reserves nationally and internationally for successful expansion.

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Appendix A. Supplementary material

Supplementary data to this article can be found online at https://doi.org/10.1016/j.apenergy.2019.02.039.

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