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Integrated Energy Optimization Model for Oil Sands Operations

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ABSTRACT: This article presents a new energy model that predicts the energy infrastructure required to maintain oil production in the Canadian Oil Sands operation at minimum cost. Previous studies in this area have focused on the energy infrastructure for fixed energy demands (i.e., the production schemes that produce synthetic crude oil (SCO) and commercial diluted bitumen remained fixed in the calculation of the optimal infrastructure). The key novelty of this work is that the model searches simultaneously for the most suitable set of oil production schemes and the corresponding energy infrastructures that satisfy the total production demands under environmental constraints, namely, CO_2 emissions targets. The proposed modeling tool was validated using historical data and previous simulations of the Canadian Oil Sands operation in 2003. Likewise, the proposed model was used to study the 2020 Canadian Oil Sands operations under three different production scenarios. Also, the 2020 case study was used to show the effect of CO_2 capture constraints on the oil production schemes and the energy producers. The results show that the proposed model is a practical tool for determining the production costs of the Canadian Oil Sands operations, evaluating future production schemes and energy demand scenarios, and identifying the key parameters that affect Canadian Oil Sands operations.

1. INTRODUCTION

The Athabasca region in Canada represents one of the largest oil reserves in the world. In 2008, Alberta's total oil reserves were estimated to be 171.8 billion barrels, which makes Canada second behind Saudi Arabia.[1] Several political, economic, and technological aspects have motivated the sudden expansion of oil-related activities in the Canadian Oil Sands. Bitumen is a highly viscous mixture of condensed polycyclic aromatic hydrocarbons that is found in large proportions in the Canadian Oil Sands. This term is widely used in the oil industry to identify a tar form of petroleum that needs to be diluted before it can be transported in a pipeline. In 2008, the production of bitumen in the province of Alberta was 1.3 million barrels per day.[1] It is expected that the daily production will reach 3 million barrels by the year 2020.[2] The crude bitumen can be extracted by surface or in situ methods. Surface methods require mining the oil sand using shovels and trucks, whereas in situ methods require the injection of an external agent into the underground oil reservoir to extract the bitumen from the reservoir. According to geologists, 80% of the present oil deposits in Alberta should be recovered using in situ methods such as steam-assisted gravity drainage (SAGD),[2] whereas the remaining 20% can be extracted using mining. The increase in Canadian Oil Sands production also generates more CO_2 emissions. The combined CO_2 emissions from synthetic crude oil (SCO) and diluted crude bitumen (DCB) productions make the Canadian Oil Sands the largest contributor to greenhouse-gas (GHG) emissions in Canada.[3]

Based on the above considerations, it is important to have a reliable modeling tool that can be used to determine the production and environmental costs of Canadian Oil Sands operations for the upcoming years. Also, this modeling tool can be used to synthesize the potential energy infrastructure that will be needed for the Canadian Oil Sands in the future; that is, the model can be used for scheduling and planning of the energy production demands for the Canadian Oil Sands. Likewise, uncertainty in key production factors, such as natural gas prices and CO_2 emissions, can also be incorporated by considering worst-case, expected, and optimistic scenarios for the Canadian Oil Sands operations in the future. These analyses can provide a broader scope of what can be expected in the future and the potential (environmental) consequences.

Models for Canadian Oil Sands operations have recently been developed.[4−6] In those studies, the most suitable infrastructure and expected production capacities of the different types of energy producers are estimated for a given oil production infrastructure. That is, the capacities and infrastructure available to produce oil from the Canadian Oil Sands need to be specified a priori and are assumed to provide the input into their proposed models. Thus, the results obtained with those models are limited because the future project planning and scheduling of the Canadian Oil Sands operation can be done only for the specified energy infrastructure (i.e., determination of the oil production schemes was not considered in the analysis). Also, the energy producers obtained by those models might not be optimal because they were calculated to satisfy fixed oil production schemes. That is, there is no guarantee that combinations among the different oil production schemes can return a more economically attractive energy infrastructure.

The aim of this study is to present an integrated optimization model in which the energy producers and production schemes of the Canadian Oil Sands are simultaneously considered within the analysis. The energy producers are the plants used for the generation of the energy commodities, including steam produced in natural gas boilers, electricity produced in power plants, and hydrogen. The production schemes from bitumen and SCO producers are modeled using different combinations of bitumen

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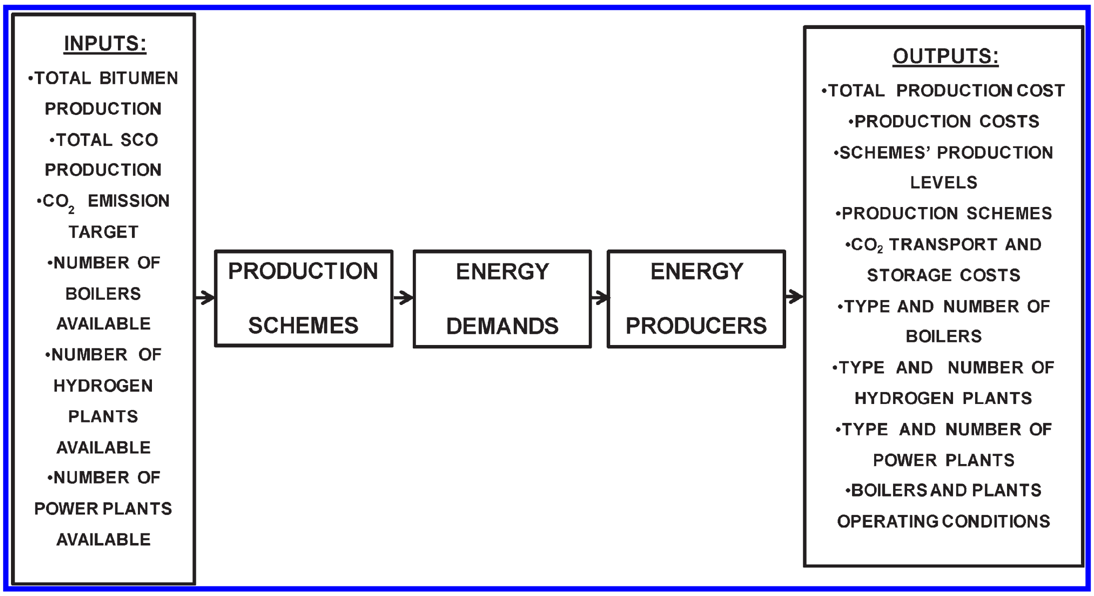

Figure 1. General layout of the integrated model.

extractions methods and upgrading technologies, such as mining and SAGD as extraction methods and delayed coking (DC), LC-fining (LCF), and LC-fining plus fluid coking (FC) as upgrading technologies. The proposed optimization model takes into account CO_2 emissions and the production forecasts for bitumen and SCO to select the most suitable configuration among production schemes with their corresponding levels and energy producers that minimize the energy production costs of the Canadian Oil Sands. To the authors' knowledge, the model proposed in this work is the first that simultaneously solves for bitumen and SCO production schemes and levels and the corresponding energy producers' infrastructure at minimum cost with a constraint on CO_2 emissions for Canadian Oil Sands operations. The integrated modeling tool proposed in this work was validated using data from the Alberta Energy and Utilities Board[7] and simulation results reported from a previous study.[4] A case study that aims at reducing the production costs for Canadian Oil Sands in 2020 is presented. The integrated model determines the most suitable combination of production schemes and energy producers with and without CO_2 emissions constraints. Because of uncertainties in the bitumen and SCO productions for 2020, the integrated model was used for different scenarios corresponding to the highest, lowest, and reference oil production forecasts for that year.[2]

This article is organized as follows: Section 2 presents the main features of the integrated optimization modeling tool proposed in this work. Section 3 presents a case study for the year 2003 that was used for model validation. Comparisons between the results obtained by the present model and those reported by other studies and sources in the open literature are presented in this section. Section 4 reports the results obtained using the proposed integrated model to predict the Canadian Oil Sands operations for the year 2020. Concluding remarks and future work are presented in section 5.

2. INTEGRATED OPTIMIZATION MODEL

This section presents the main features of the integrated optimization model that is proposed in this work to determine the energy production costs of Canadian Oil Sands operations.

2.1. Problem Statement. The energy optimization model presented in this work aims to minimize the energy costs associated with oil production from unconventional resources in the province of Alberta, Canada. Figure 1 shows the key model inputs and outputs considered for the present energy model. The model consists of oil producers, namely, SCO and commercial crude bitumen, and energy commodity producers such as boilers and hydrogen and power plants. The oil production schemes and energy plants represent the energy consumers and producers, respectively. The oil producers require energy commodities to extract and convert the crude bitumen into marketable products, whereas the energy producers supply the commodities to meet the oil producers' demands. Therefore, the integrated relationship between energy consumers and producers was considered in the present energy model. The model's key inputs are (see Figure 1) the total expected commercial bitumen and SCO productions, the carbon dioxide (CO_2) emissions target, and the numbers of energy producers available (i.e., boilers and power and hydrogen plants). To meet the total expected SCO and bitumen production levels, the present energy model minimizes the energy costs by simultaneously selecting the most suitable type of oil production schemes and the energy producer plants subject to an environmental (CO_2 emissions) constraint. Therefore, the model's key outputs are (see Figure 1) the total energy cost for production, the individual costs (energy, capital, operation, and maintenance), the oil schemes' producers with their corresponding capacities, the carbon dioxide (CO_2) transport and storage costs, and the types and numbers of energy producers with their operating conditions (plant's capacities). The present energy model includes the selection of continuous, binary, and integer variables. Thus, the resulting model is formulated as a mixed-integer nonlinear optimization problem. Each of the sections shown in Figure 1 is described next.

2.2. Inputs. In the proposed model, the inputs are represented by the total diluted bitumen (TDB) and total SCO (TSCO) productions expected for a given year, the carbon dioxide emissions target (CO_2E), and the maximum number of energy producers available. The oil production values and the CO_2 emissions target are obtained from forecasts[2] or are specified by

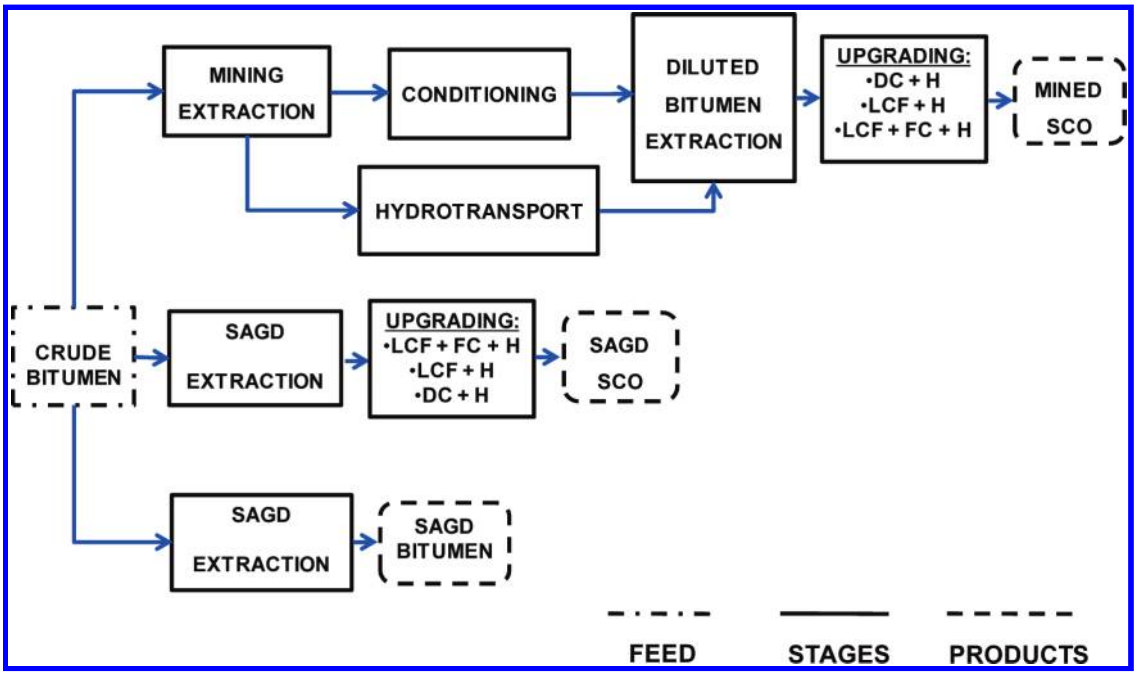

Figure 2. Diluted bitumen and SCO production schemes.

the user. The maximum number of energy producers available is also defined by the user. The total diluted bitumen production (TDB) is formulated as

$$\mathrm{TDB} = \mathrm{CDB} \tag{1}$$

where TDB is a function of the bitumen obtained by SAGD extraction (CDB) for commercialization (bbl of bitumen/day). That is, the present work assumes that the total diluted bitumen is produced only by SAGD extraction. The mathematical formulations for carbon dioxide emissions and the total SCO production are presented next.

2.3. CO_2 Emissions Target. The carbon dioxide emissions target (CO_2E) is calculated as

$$\mathrm{CO_2E} = \mathrm{CO_2B}(1 - \mathrm{CCO_2}) \tag{2}$$

where CO_2E (specified by the user) is a function of the operations' baseline carbon dioxide emissions without CO_2 capture [CO_2B (t of CO_2/h)] and the percentage of CO_2 captured (CCO_2).

2.4. Production Schemes. The integrated model considers two different crude bitumen extraction methods: surface and in situ. The surface method requires mining the oil sand, whereas the in situ method involves injecting an external agent into the underground reservoir. SAGD is the only in situ method considered in this study for diluted bitumen production. Three different products are considered in the model: mined bitumen upgraded to SCO (integrated mining/upgrading production schemes), SAGD bitumen upgraded to SCO (integrated SAGD/upgrading production schemes), and SAGD diluted bitumen. These products and their corresponding production schemes are shown in Figure 2. Mining and SAGD bitumen extraction upgraded to SCO are the production schemes modeled to produce SCO. The bitumen upgrading technologies considered in this study are the currently leading technologies in the Athabasca region.[7] These leading technologies follow the upgrading routes

$$\text{delayed coking (DC)} + \text{hydrotreatment (H)} \tag{3}$$

$$\text{LC-fining (LCF)} + \text{hydrotreatment (H)} \tag{4}$$

$$\text{LC-fining (LCF)} + \text{fluid coking (FC)} + \text{hydrotreatment (H)} \tag{5}$$

The total SCO production (TSCO) is estimated in the model as

$$\mathrm{TSCO} = \sum_{i=1}^{N} \mathrm{SO}_i \tag{6}$$

where subscript i represents a specific SCO production scheme and N is the total number of production schemes considered in the model. SO_i represents the mined and SAGD bitumen upgraded to SCO produced by each scheme (bbl of SCO/day). Equation 6 is used in the model to select the most suitable SCO production schemes and their corresponding production levels. Each of the stages involved in the schemes is described below.

2.4.1. Mining Extraction. This is a surface method used only for SCO production in the model. The amount of mined oil sand in the model depends on the characteristics of the integrated mining/upgrading production schemes and their corresponding production levels. The energy resource for this process is diesel, which is consumed by the fleets of shovels and trucks used for mining the oil sand. The models and numbers of vehicles included in the fleets correspond to those used in a typical Canadian Oil Sands operation.[5] The total amount of diesel (D) consumed by the fleets depends on the specifications of each individual vehicle (i.e., fuel consumption parameters) and the numbers of trucks and shovels used in the fleets. The diesel consumed by the fleet of shovels is formulated as

$$\mathrm{DSH} = \sum_{k=1}^{K} \mathrm{SH}_k D_k \tag{7}$$

where DSH is the amount of diesel (L/h) consumed by the shovel fleet, K is the total number of models of shovels available in the fleet, SH_k is the number of vehicles of model k used in the fleet, and D_k is the fuel consumption of the kth model (L/h).[5] The diesel consumption by the truck fleet (DT) is calculated as

$$\mathrm{DT} = \sum_{l=1}^{L} T_l D_l \tag{8}$$

where L represents the total number of truck models in the fleet, T_l is the number of vehicles of model l used in the fleet, and D_l is the diesel consumption of model l (L/h).[5] The total diesel demand for the Canadian Oil Sands operation is thus estimated as

$$D = \mathrm{DSH} + \mathrm{DT} \tag{9}$$

2.4.2. SAGD Extraction. SAGD extraction is an in situ method[8] that is used in the model for the production of commercial diluted bitumen and SCO. The amount of SAGD bitumen is calculated based on the characteristics of the integrated SAGD/upgrading production schemes (i.e., SCO conversion) and the DB scheme and their corresponding production levels. The demands of this method are (i) SAGD steam (at 8000 kPa with a quality of 80%), which is injected into the oil reservoir at a typical steam-to-oil ratio (SOR), and ii) power. The SAGD steam demand for this process is estimated as

$$\mathrm{SSE} = \mathrm{SOR}\left[\mathrm{DBR} + \sum_{i=1}^{N} \mathrm{PS_S}(i)\,\mathrm{BIT}_i\right]$$
$$\mathrm{PS_S}(i) = \begin{cases} 1 & \text{if } i \text{ follows an integrated SAGD/} \\ & \quad \text{upgrading production scheme} \\ 0 & \text{otherwise} \end{cases} \tag{10}$$

where SSE represents the total steam consumption in SAGD extraction, including that used for diluted bitumen and SCO production by SAGD schemes. Accordingly, SSE (tonne/h) is a function of the bitumen production rate by SAGD for SCO BIT_i (tonne/h) and diluted bitumen DBR (tonne/h) productions. BIT_i and DBR are the bitumen inputs required by the production schemes to meet the demands for SCO and diluted bitumen productions, respectively. Also, SSE is a function of the steam-to-oil ratio (SOR) parameter. This parameter is typically set to 2.4 for Canadian Oil Sands operations.[8] Similarly, the power demand (PSE) is calculated as

$$\mathrm{PSE} = P_{\mathrm{SE}}\left[\mathrm{DBR} + \sum_{i=1}^{N} \mathrm{PS_S}(i)\,\mathrm{BIT}_i\right] \tag{11}$$

where P_{SE} is a parameter used to indicate the power requirements for SAGD extraction (P_{SE} = 3.1 kW per tonne of bitumen[8]).

2.4.3. Conditioning. Conditioning constitutes the first process that separates the crude bitumen from the sand. In the model, this stage is used to condition only 25% of the oil sand processed, whereas the remaining 75% is processed using hydrotransport. Conditioning consists of mixing the mined oil sand with hot water at 35−50 °C and caustic soda. The resulting slurry is agitated in rotary drums known as tumblers where the temperature is maintained constant by steam injection. Then, the slurry rich in oil obtained from this process is treated in the bitumen extraction plant. The main energy consumption for conditioning is for hot water ($\mathrm{HW_C}$) and steam (S_C). The hot water demands for this process are calculated as

$$\mathrm{HW_C} = \mathrm{WOSR_C} \sum_{i=1}^{N} \mathrm{UR_F}(i)\,\mathrm{LF_C}(i)\,\mathrm{OSR}_i$$
$$\mathrm{UR_F}(i) = \begin{cases} 1 & \text{if } i \text{ follows the upgrading route in eq 5} \\ 0 & \text{otherwise} \end{cases}$$
$$\mathrm{LF_C}(i) = \begin{cases} 1 & \text{if } i \text{ follows a production scheme with a conditioning stage} \\ 0 & \text{otherwise} \end{cases} \tag{12}$$

As shown in eq 12, $\mathrm{HW_C}$ (tonne/h) depends on the rate at which oil sand is mined [OSR_i (tonne/h)] and the water-to-oil sand ratio for conditioning ($\mathrm{WOSR_C}$ = 0.333 t of water per tonne of oil sand[9]). The steam requirement in this stage (S_C) is formulated as

$$S_\mathrm{C} = \mathrm{SOSR_C} \sum_{i=1}^{N} \mathrm{UR_F}(i)\,\mathrm{LF_C}(i)\,\mathrm{OSR}_i \tag{13}$$

where $\mathrm{SOSR_C}$ is the steam-to-oil sand ratio in conditioning ($\mathrm{SOSR_C}$ = 0.036 t of steam per tonne of oil sand[9]).

2.4.4. Hydrotransport. In this process, the mined oil sand is mixed with hot water at 35 °C, creating a slurry that is pumped through pipelines to the bitumen extraction stage. This process can be considered as two processes occurring simultaneously, because the slurry is being conditioned and transported at the same time. The energy demands in this stage include those for hot water and power. The hot water demands for this stage are calculated as

$$\mathrm{HW_H} = \mathrm{WOSR_H} \sum_{i=1}^{N} \mathrm{PS_M}(i)\,\mathrm{OSR}_i$$
$$\mathrm{PS_M}(i) = \begin{cases} 1 & \text{if } i \text{ follows an integrated mining/} \\ & \quad \text{upgrading production scheme} \\ 0 & \text{otherwise} \end{cases} \tag{14}$$

where $\mathrm{HW_H}$, the hot water consumed in hydrotransport, is a function of the rate at which oil sand is mined (OSR_i) and the water-to-oil sand ratio for hydrotransport ($\mathrm{WOSR_H}$ = 0.30 t of water per tonne of oil sand). The power demand (P_H) is calculated as

$$P_\mathrm{H} = \sum_{i=1}^{N} \mathrm{PS_M}(i)\,\mathrm{ST}_i d_i \mathrm{SPF}_i \tag{15}$$

where ST_i (tonne/h) is the amount of slurry (70% solids content), d_i (m) is the distance from the mining site to the extraction plant, and SPF_i [kWh (t of slurry)$^{-1}$ m^{-1}] is the slurry pumping factor. ST_i depends on the rate at which mined oil sand is processed, and d_i is a model parameter.

2.4.5. Diluted Bitumen Extraction. The mined SCO production schemes considered in the model follow a two-step hot-water process for bitumen extraction.[9] In the primary extraction, the bitumen froth from hydrotransport and conditioning is separated from the slurry using steam and hot water. In this step, a set of chemical and physical processes are used to break the bonds that hold together the slurry components, namely, sand, crude bitumen, clay, and water. In secondary extraction, the bitumen froth is diluted in naphtha and then centrifuged to separate the remaining sand and water from the bitumen. This last process is a mechanical separation method. Thus, the energy requirements associated with this stage are those for hot water, steam, and electricity. The hot water demand in bitumen extraction is

$$HW_{BE} = WOSR_{BE} \sum_{i=1}^{N} PS_M(i)\ OSR_i \tag{16}$$

where HW_{BE} (tonne/h), the total hot water demand in primary extraction,[9] is a function of OSR_i and the water-to-oil sand ratio for diluted bitumen extraction ($WOSR_{BE}$ = 0.41 t of water per tonne of oil sand[9]), which is a parameter that indicates the wash water requirement in this stage. The steam demands for this process in secondary extraction are calculated as[9]

$$S_{BE} = SFR \sum_{i=1}^{N} PS_M(i)\ BF_i \tag{17}$$

where BF_i (t of froth/h) is the amount of crude bitumen froth coming from primary extraction and SFR is a parameter that defines the steam requirement for secondary extraction in the stage (SFR = 0.040 t of steam per tonne of froth[9]). The power demand is calculated from the equation

$$P_{BE} = \sum_{i=1}^{N} PS_M(i)(PT_i + PC_i) \tag{18}$$

where P_{BE} (kW) is the total power demand for this stage, which comprises the power requirements to pump the tailings to disposal (PT_i) and power for centrifugation (PC_i).

2.4.6. Upgrading. In the upgrading stage, the bitumen feed proceeding from diluted bitumen extraction is upgraded to SCO. The present model considers three upgrading routes shown in Figure 2. Bitumen upgrading requires large amounts of energy, namely, steam, hydrogen, power, and process fuel (NG) for heating. The first step in the stage consists of recovering the naphtha (NT) used as a solvent for bitumen dilution in the diluent recovery unit (DRU). Then, the naphtha is recycled and reused in diluted bitumen extraction. Once the naphtha is recovered, the remaining products, namely, atmospheric topped bitumen (ATB) and light gas oil (LGO), require further treatment to be upgraded into SCO. The ATB is sent to the vacuum distillation unit (VDU)[10] in the following production schemes: integrated mining/upgrading that follows the upgrading routes in eqs 3 and 4 and integrated SAGD/upgrading. For the remaining production scheme, integrated mining/upgrading that follows the upgrading route in eq 5, the ATB flow splits between the VDU and LC-finer unit (LCFU). The products obtained from the VDU, namely, LGO and heavy gas oil (HGO), are sent to hydrotreatment together with the LGO coming from the DRU. In the hydrotreatment process, the sulfur and nitrogen contents are removed to produce a sweet SCO product. The VDU bottom product, known as vacuum topped bitumen (VTB), is mixed with ATB from the DRU and sent to the LC-finer in production schemes following the upgrading route described in eq 5 or to the delayed coker units (DCU) in schemes following the upgrading route in eq 3.

The LC-finer is a hydrocracking technology based on a catalytic chemical process in which high-boiling and high-molecular-weight hydrocarbons, namely, VTB and ATB, are cracked into lower-boiling and lower-molecular-weight hydrocarbons, namely, LGO, HGO, and NT products. Then, these products are sent to hydrotreatment. The low-conversion LC-finer (upgrading route in eq 5) bottoms are transported to the fluid coker. The coker units, namely, fluid coker (FCU) and delayed coker, use thermal cracking to yield lighter hydrocarbons. The coker units process the bottoms from upstream units to produce additional NT, LGO, and HGO products. Also, this process yields petroleum coke as a byproduct. The low-conversion LC-finer and the fluid coker were modeled according to the approaches in Sunderland,[11] Schumacher,[12] and Van Driesen et al.,[13] whereas the model for the high-conversion LC-finer (upgrading route in eq 4) was taken from Meyers.[9] The last process in this stage consists of the hydrodesulfurization of the products (NT, LGO, and HGO) that have not been passed through hydrotreaters, where the sulfur is removed. In the last stage of this process, the oil products are blended together to form SCO. The total steam demand for upgrading (S_U) is formulated as

$$S_U = \sum_{i=1}^{N} [DBIT_i SDRU + (ATB - ATBF)_i SVDU + FB_i SFCU] \tag{19}$$

where $DBIT_i$ (tonne/h) is the amount of diluted bitumen entering the upgrading stage; SDRU is a parameter defining the steam requirements in the DRU (SDRU = 0.30 t of steam per tonne of diluted bitumen); and SVDU and SFCU are parameters in the model that correspond to the steam requirements for the VDU and FCU, respectively (SVDU = 0.07 t of steam per tonne of diluted bitumen[10]). The term ATB_i (tonne/h) is the amount of atmospheric topped bitumen, $ATBF_i$ (tonne/h) is the amount of atmospheric topped bitumen feeding the LC-finers, and FB_i represents the LC-finer bottom oil fractions.

The hydrogen demand for upgrading considers the hydrogen needed for hydrocracking and hydrodesulfurization. The hydrogen for hydrocracking is calculated as

$$H_{HC} = \frac{1}{\rho_{H_2}} \sum_{i=1}^{N} HC(i)[(VTB_i + ATBF_i)HLF + VTB_i HHF]$$
$$HC(i) = \begin{cases} 1 & \text{if } i \text{ follows hydrocracking upgrading schemes} \\ & \text{(i.e., eqs 4 and 5)} \\ 0 & \text{otherwise} \end{cases} \tag{20}$$

where H_{HC} is the total hydrogen used for hydrocracking, VTB_i is the amount of vacuum topped bitumen (tonne/h), and HLF is a parameter that indicates the hydrogen requirements for low-conversion LC-finers (HLF = 6.046 scf of H_2 per tonne of bitumen). This first term in the equation represents the hydrogen consumption of the low-conversion LC-finer,[11−13] whereas the second term represents that of the high-conversion LC-finer.[9] The parameter HHF is the hydrogen required for the high-conversion LC-finers (HHF = 8.464 scf of H_2 per tonne of bitumen), and ρ_{H_2} is the hydrogen density (ρ_{H_2} = 423 000 scf/tonne). The total hydrogen (H_{HT}) demand for hydrotreatment was modeled as[11]

$$H_{HT} = \frac{1}{UCF\rho_{H_2}} \sum_{i=1}^{N} \left[\left(\frac{LGO_i}{DLGO}\right) HLGO + \left(\frac{HGO_i}{DHGO}\right) HHGO + \left(\frac{NT_i}{DNT}\right) HNT \right] \tag{21}$$

where DLGO, DHGO, and DNT are the average densities of the oil fractions entering the hydrotreaters for the LGO, HGO, and NT streams, respectively. The numerical values for these parameters were taken from Yui et al. as follows:[10] DLGO = 0.9125 t/m^3, DHGO = 0.9713 t/m^3, and DNT = 0.744 t/m^3). HLGO, HHGO, and HNT are parameters that specify the hydrogen requirements in hydrotreaters for LGO, HGO, and NT, respectively. The numerical values for these parameters were taken from Sunderland:[11] HLGO = 1150 scf/bbl, HHGO = 1150 scf/bbl, and HNT = 930 scf/bbl. Finally, UCF is a unit conversion factor (UCF = 0.1589873 m^3/bbl). The total hydrogen demand in upgrading (H_U) is thus defined as

$$H_U = H_{HC} + H_{HT} \tag{22}$$

The power demands in upgrading depend on the upgrading route. For schemes following the upgrading route in eq 3, the total power requirements is given by the equation

$$P_{UD} = \frac{\mathrm{PDDC}}{\mathrm{DVTB}} \sum_{i=1}^{N} \mathrm{UR_D}(i)\ \mathrm{VTB}_i$$
$$\mathrm{UR_D}(i) = \begin{cases} 1 & \text{if } i \text{ follows the upgrading route in eq 3} \\ 0 & \text{otherwise} \end{cases} \tag{23}$$

where P_{UD} (kW) is the power demand in delayed-coking-based schemes, PDDC is a parameter that defines the electricity requirement for delayed coking (PDDC = 3.9 kWh/bbl), and DVTB the density of vacuum topped bitumen (DVTB = 0.16805 t/bbl). The power requirement for the production schemes following the upgrading route in eq 4 is estimated from the expression

$$P_{UL} = \frac{\mathrm{PDLF}}{\mathrm{DLF}} \sum_{i=1}^{N} \mathrm{UR_L}(i)\ \mathrm{VTB}_i$$
$$\mathrm{UR_L}(i) = \begin{cases} 1 & \text{if } i \text{ follows the upgrading route in eq 4} \\ 0 & \text{otherwise} \end{cases} \tag{24}$$

where P_{UL} is the power demand in LC-fining-based schemes, PDLF is a parameter that indicates the power demands per bitumen feed in high-conversion LC-finers[14] (PDLF = 16.5 kWh/bbl), and DLF is the average density of the LC-finer feed (DLF = 0.1654 t/bbl[10]). The total power demand from schemes that include the upgrading route in eq 5 is defined as

$$P_{UF} = \frac{1}{\mathrm{DLF}} \sum_{i=1}^{N} \mathrm{UR_F}(i)[\mathrm{PDHF}(\mathrm{VTB} + \mathrm{ATBF})_i + \mathrm{FB}_i\mathrm{PDFC}] \tag{25}$$

where P_{UF} is the power demand for LC-fining plus fluid coking schemes (eq 5) and the model parameters PDHF and PDFC correspond to the power requirements for low-conversion LC-finer[14] and fluid coking[9] processes, respectively (PDHF = 16.5 kWh/bbl, PDFC = 6 kWh/bbl). The total electricity demands for the upgrading stage (P_U) can thus be calculated as

$$P_U = P_{UD} + P_{UL} + P_{UF} \tag{26}$$

As for the power demands (eq 26), the process fuel requirements for upgrading depend on the upgrading route. Process fuel (NG) is consumed in different steps of the upgrading stages. The corresponding energy demands for NG are calculated as

$$\mathrm{PF_{UD}} = \frac{\mathrm{FDDC}}{\mathrm{DVTB \cdot HVNG}} \sum_{i=1}^{N} \mathrm{UR_D}(i)\ \mathrm{VTB}_i \tag{27}$$

$$\mathrm{PF_{UL}} = \frac{\mathrm{FDLCF}}{\mathrm{DLF \cdot HVNG}} \sum_{i=1}^{N} \mathrm{UR_L}(i)\ \mathrm{VTB}_i \tag{28}$$

$$\mathrm{PF_{UF}} = \frac{\mathrm{FDLCF}}{\mathrm{DLF \cdot HVNG}} \sum_{i=1}^{N} \mathrm{UR_F}(i)(\mathrm{VTB} + \mathrm{ATBF})_i \tag{29}$$

where FDLCF and FDDC are parameters that represent the process fuel requirements for LC-fining[14] and delayed coking,[14] respectively (FDLCF = 93.47 MJ/bbl, FDDC = 153 MJ/bbl), and HVNG is the Western Canadian gas heating value (HVNG = 38.05 MJ/m^3). The total process fuel demand for upgrading ($\mathrm{PF_U}$) is calculated as

$$\mathrm{PF_U} = \mathrm{PF_{UD}} + \mathrm{PF_{UL}} + \mathrm{PF_{UF}} \tag{30}$$

2.4.7. Additional Power Requirements. The proposed integrated model also considers additional power demands, including the power requirements of the hydrogen plants and the power requirements to transport the CO_2 captured in hydrogen and power plants from Fort McMurray to depleted oil fields near Edmonton, such as Red Water Field. The model considers different hydrogen plants (see section 2.6). From these plants, only steam methane reforming (SMR) requires energy to operate. The remaining plants (gasification) coproduce power to maintain themselves and to add electricity to the Canadian Oil Sands supply. The power demand for the SMR hydrogen plants (PHP) is calculated as

$$\mathrm{PHP} = \sum_{j=1}^{J} \mathrm{HPS}(j)\ \mathrm{HP}_j\mathrm{PC}_j$$
$$\mathrm{HPS}(j) = \begin{cases} 1 & \text{if hydrogen plant } j \text{ is of type SMR} \\ 0 & \text{otherwise} \end{cases} \tag{31}$$

where J represents the total number of types of hydrogen plants considered in the model. HP_j is the amount of hydrogen produced in a plant of type j (t of H_2/h), and PC_j is the power consumption in a plant of type j [kWh/(t of H_2)]. The power demand to transport the CO_2 captured in hydrogen plants (PCTH) is formulated as

$$\mathrm{PCTH} = \mathrm{PL} \sum_{j=1}^{J} \mathrm{HPC}(j)\ \mathrm{CCH}_j\mathrm{CPCT}$$
$$\mathrm{HPC}(j) = \begin{cases} 1 & \text{if hydrogen plant } j \text{ captures } CO_2 \\ 0 & \text{otherwise} \end{cases} \tag{32}$$

where CCH_j is the amount of CO_2 captured in a hydrogen plant of type j (t of CO_2/h), CPCT is the compression power for CO_2 transport [kWh (t of CO_2)$^{-1}$ km^{-1}], and PL is the pipeline length (km). The power demand to transport the CO_2 captured in power plants (PCTP) is calculated as

$$\mathrm{PCTP} = \mathrm{PL}\sum_{m=1}^{M} \mathrm{PPC}(m)\,\mathrm{CCP}_m\mathrm{CPCT}$$
$$\mathrm{PPC}(m) = \begin{cases} 1 & \text{if power plant } m \text{ captures } CO_2 \\ 0 & \text{otherwise} \end{cases} \quad (33)$$

where M represents the number of types of power plants considered in the model. CCP_m is the amount of CO_2 captured in power plants (t of CO_2/h), and CPCT is the compression power for CO_2 transport.

2.5. Energy Demands. The total energy demands are represented by the sum of the energy requirements of the production schemes considered in the model. Thus, the energy demands considered in the model are those for power, steam, hot water, hydrogen, diesel, and process fuel. As described in the previous section, the energy demands are estimated based on the energy requirements needed by each of the production schemes. Equations 34−36 show the expressions used to determine the power, process steam, and hot water demands for the Canadian Oil Sands production schemes

$$\mathrm{PD} = \mathrm{PSE} + P_{\mathrm{H}} + P_{\mathrm{BE}} + P_{\mathrm{U}} + \mathrm{PHP} + \mathrm{PCTH} + \mathrm{PCTP} \quad (34)$$

$$\mathrm{SD} = S_{\mathrm{C}} + S_{\mathrm{BE}} + S_{\mathrm{U}} \quad (35)$$

$$\mathrm{HWD} = \mathrm{HW}_{\mathrm{C}} + \mathrm{HW}_{\mathrm{H}} + \mathrm{HW}_{\mathrm{BE}} \quad (36)$$

The total power demand, PD (kW), represents the electricity demands from the different production schemes. Likewise, the total process steam demand, SD (tonne/h), is a function of the steam requirements in conditioning (S_{C}), diluted bitumen extraction (S_{BE}), and upgrading (S_{U}). Similarly, HWD, the total demand for hot water, is calculated based on the hot water consumption in conditioning (HW_{C}), hydrotransport (HW_{H}), and diluted bitumen extraction ($\mathrm{HW}_{\mathrm{BE}}$). The expressions to estimate the energy demands for diesel, SAGD steam, and hydrogen for upgrading are defined in eqs 9, 10, and 22, respectively.

2.6. Energy Producers. The energy producers considered in the present model to satisfy the energy demands in the model are (i) boilers, which are used to satisfy the energy demands for process steam (SD), hot water (HWD), and SAGD extraction steam (SSE); (ii) power plants for electricity generation (PD); and (iii) hydrogen plants to cover the hydrogen demands for upgrading (H_{U}). A detailed description on the energy producers is presented next.

2.6.1. Boilers. The present model considers conventional natural-gas-fired boilers.[15] Process steam is generated at 6300 kPa and 500 °C. This type of steam is used for conditioning, diluted bitumen extraction, and upgrading. The total cost associated with the production of process steam (S_{TC}) in this type of boiler (SB) is calculated as

$$S_{\mathrm{TC}} = t\left[\frac{1}{\mathrm{EC}}(\mathrm{NSB}\cdot\mathrm{NGSB}\cdot\mathrm{CS}\cdot\mathrm{HVNG}\cdot\mathrm{PNG}) + \mathrm{SD}\cdot\mathrm{CFW}\right] \quad (37)$$

where NGSB is the consumption of NG per boiler (N m^3/h), NSB is the number of boilers selected by the model to produce process steam, CS is the percentage of the boiler capacity used to generate steam (82%), HVNG is the heating value of NG (38.05 MJ/Nm^3), PNG is the price of NG, SD (eq 35) is the total amount of steam (tonne/h) produced by the boilers, CFW is the cost of the boiler feedwater, t is the annual operating hours (8760 h/year), and EC is an energy conversion factor (1000 MJ/GJ).

Hot water at 35 °C is used for conditioning, hydrotransport, and diluted bitumen extraction. The proposed model assumes that the capacity of the boilers is used to produce process steam and hot water. The total cost of hot water ($\mathrm{HW}_{\mathrm{TC}}$) is calculated as

$$\mathrm{HW}_{\mathrm{TC}} = t\left\{\frac{1}{\mathrm{EC}}[\mathrm{NSB}\cdot\mathrm{NGSB}(1-\mathrm{CS})\mathrm{HVNG}\cdot\mathrm{PNG}] + \mathrm{HWD}\cdot\mathrm{CFW}\right\} \quad (38)$$

where HWD (eq 36) is the amount of hot water (tonne/h) produced in the boilers (SB).

The present model also includes boilers that produce SAGD steam at 8000 kPa and 80% quality (SSEB). The cost for SAGD steam ($\mathrm{SSE}_{\mathrm{TC}}$), used only for in situ bitumen extraction, is calculated as

$$\mathrm{SSE}_{\mathrm{TC}} = t\left[\frac{1}{\mathrm{EC}}(\mathrm{NSEB}\cdot\mathrm{NGSEB}\cdot\mathrm{HVNG}\cdot\mathrm{PNG}) + \mathrm{SSE}\cdot\mathrm{CFW}\right] \quad (39)$$

where NGSEB is the consumption of NG per boiler (Nm^3/h), NSEB is the number of boilers producing SAGD steam, and SSE is the amount of SAGD steam produced in the boilers (see eq 10). The installed capacity of the boilers considered in the model is 340 t of steam/h.[15] The capital cost of the boilers is not considered in this model given that it can be neglected when compared to the annual fuel consumption cost.

2.6.2. Hydrogen Plants. The present model considers steam methane reforming (SMR) and gasification as the technologies for hydrogen production. The SMR plants considered in this model are based on those used in previous studies.[16−18] The model assumes SMR hydrogen plants both without and with CO_2 capture. The gasification plants were modeled using data from different sources.[19−21] The hydrogen producers in this model also include gasification plants without and with CO_2 capture. The total cost to produce hydrogen with the two technologies can be estimated as

$$H_{\mathrm{TC}} = \sum_{j=1}^{J} \mathrm{NHP}_j(\mathrm{ACC}_j + \mathrm{OMC}_j) + \frac{tF_j\mathrm{FHV}_j\mathrm{FC}_j}{\mathrm{EC}} \quad (40)$$

where NHP_j represents the number of plants of type j considered in the model. ACC_j is the annual capital cost of a hydrogen plant of type j (\$/year), OMC_j is the annual operation and maintenance cost for a plant of type j (\$/year), F_j is the fuel consumed by a plant of type j (NG in N m^3/h or coal in kg/h), FHV_j is the fuel heating value (NG = 38.05 MJ/Nm^3 or coal = 24.05 MJ/kg), and FC_j the fuel cost (\$/GJ) for a plant of type j. Equation 40 is related to the total hydrogen demand as

$$H_{\mathrm{U}} = \sum_{j=1}^{J} \frac{F_j\mathrm{FHV}_j}{\mathrm{HR}_j} \quad (41)$$

where H_{U} is the total hydrogen demand (see eq 22) and HR_j is the heating rate required to produce 1 t of H_2 [MJ/(t of H_2)] for a hydrogen plant of type j. The annual capital cost (ACC_j) of each type of plant is calculated as

$$ACC_j = HPIC_j PCC_j ACF_j, \quad j = 1, ..., J \tag{42}$$

where the annual capital cost is a function of the plant installed capacity, $HPIC_j$ (t of H_2/h); the plant capital cost, PCC_j [(\$ h)/(t of H_2)]; and ACF_j, which is an amortized capital factor given as a percentage (%). The annual operation and maintenance cost (OMC_j) of each type of plant is calculated as

$$OMC_j = HPIC_j PCC_j OMF_j, \quad j = 1, ..., J \tag{43}$$

where OMF_j is an operation and maintenance economic factor given as a percentage (%).

2.6.3. Power Plants. The power plants included in the model are integrated gasification combined cycle (IGCC),[22] oxyfuel,[23] natural gas combined cycle (NGCC),[24] and supercritical pulverized coal (SCPC).[24] Three CO_2 capture methods are considered in the model: precombustion, postcombustion, and oxy-combustion. Precombustion is modeled in IGCC plants, postcombustion in NGCC and SCPC plants, and oxy-combustion in oxyfuel power plants. The IGCC power plants in the model used coal as the feedstock. The model considers IGCC plants both without and with CO_2 capture. Likewise, the oxyfuel plants included in the model are natural gas and coal with CO_2 capture. Moreover, the model considers NGCC plants both without and with CO_2 capture. Furthermore, two SCPC plants are included in the model: SCPC without CO_2 capture and SCPC with CO_2 capture. The total cost for power generation by the fleet (P_{TC}) is formulated as

$$P_{TC} = \sum_{m=1}^{M} NPP_m(ACC_m + OMC_m) + \frac{tPG_m HR_m FC_m}{EC} \tag{44}$$

where NPP_m represent the number of power plants of type m, ACC_m is the capital cost, OMC_m is the annual operation and maintenance cost, PG_m is the power generated (kW), HR_m is the heating rate (MJ/kWh), and FC_m is the fuel cost (NG and coal) for plants of type m. Equation 44 is related to the total power demands as follows

$$\begin{aligned} PD &= \sum_{m=1}^{M} PG_m + \sum_{j=1}^{J} HPG(j)\, PG_j \\ HPG(j) &= \begin{cases} 1 & \text{if hydrogen plant } j \text{ is of type gasification} \\ 0 & \text{otherwise} \end{cases} \end{aligned} \tag{45}$$

where PD (see eq 34) is the total power required in the model (kW) and PG_j is the power cogenerated in a gasification hydrogen plant of type j (kW). The annual capital cost of the power plants (ACC_m) is calculated as

$$ACC_m = HPIC_m\, PCC_m\, ACF_m, \quad m = 1, ..., M \tag{46}$$

where the annual capital cost is a function of the plant installed capacity, $HPIC_m$ (kW); the plant capital cost, PCC_m (\$/kW); and ACF_m, an amortized capital factor given as a percentage (%). Every energy producer considered in the model includes an installed capacity. However, the operating condition (OC) is a decision variable within the optimization formulation (see eq 52). The annual operation and maintenance cost (OMC_m) of each type of plant is calculated as

$$OMC_m = HPIC_m\, PCC_m\, OMF_m, \quad m = 1, ..., M \tag{47}$$

where OMF_m is an operation and maintenance economic factor given as a percentage (%).

2.7. Additional Costs and Outputs. The diesel and process fuel feedstocks are assumed to be supplied by external providers. As shown in eq 48, the total cost of diesel (D_{TC}) is calculated from the total fuel diesel demand, D (see eq 9), and the cost of the diesel (CD). Similarly, the total cost of process fuel demand (PF_{TC}) in eq 49 is a function of the total process fuel consumption, PF_U (see eq 30)

$$D_{TC} = D \cdot CD \cdot t \tag{48}$$

$$PF_{TC} = \frac{tPF_U HVNG \cdot PNG}{EC} \tag{49}$$

The costs associated with the transport of the CO_2 captured in power and hydrogen plants is calculated as

$$CTC = \left[\sum_{j=1}^{J} HPC(j)\, CCH_j + \sum_{m=1}^{M} PPC(m)\, CCP_m\right](t \cdot UCTC \cdot PL) \tag{50}$$

where the subscripts j and m represent the types of hydrogen and power plants, respectively. CTC is the total annual CO_2 transport cost (\$/year); CCH_j and CCP_m are the total amounts of CO_2 captured in hydrogen and power plants (t of CO_2/h), respectively; UCTC is the unitary CO_2 transport cost (\$0.014 per tonne of CO_2 per kilometer); and PL is the length of the pipe used to transport the CO_2 from Fort McMurray to depleted oil fields nearby Edmonton (600 km). The annual carbon dioxide storage cost (CSC) is calculated as

$$CSC = t \cdot UCSC\left[\sum_{j=1}^{J} HPC(j)\, CCH_j + \sum_{m=1}^{M} PPC(m)\, CCP_m\right] \tag{51}$$

where UCSC is a parameter representing the carbon dioxide underground injection cost.

2.8. Optimization Model. Based on the inputs, the bitumen and SCO production schemes, the energy demands, and the energy producers discussed in the above sections, the optimization model considered in this work is formulated as

$$\begin{aligned} \min_{\eta}\ CF = {} & P_{TC} + H_{TC} + S_{TC} + SSE_{TC} + HW_{TC} \\ & + PF_{TC} + D_{TC} + CTC + CSC \end{aligned} \tag{52}$$

subject to total energy demands (eqs 9, 10, 22, 30, and 34–36), production schemes (OPS_i), production levels (eqs 1 and 6), energy producers (eqs 37–47), energy producers' installed capacities, and environmental constraints (eq 2), where η = [OPS_i, SO_i, NSB, NSEB, NHP_j, NPP_m, OC].

η represents the set of decision variables specified by the production schemes (OPS_i), the scheme production levels (SO_i), the number of process steam boilers (NSB), the number of SAGD steam boilers (NSEB), the number of hydrogen plants of type j, (NHP_j), the number of power plants of type m (NPP_m), and the energy producers' operating conditions (OC). The optimization model will find the most suitable combination of production schemes with corresponding levels and energy producers that minimize the energy production costs of the Canadian Oil Sands operation. The resulting integrated energy optimization model is a comprehensive mathematical model that considers the energy demands, the diluted bitumen and SCO

Table 1. Synthetic Crude Oil and Diluted Bitumen Production Schemes

production scheme	stage(s)[a]
	Integrated Mining/Upgrading
OPS_1	mining → hydro → DBE → DC → H
OPS_2	mining → hydro → DBE → LCF → H
OPS_3	mining → hydro → DBE → LCF → FC → H
OPS_4[b]	Mining → Cond → DBE → LCF → FC → H Mining ↘ Hydro ↗ DBE
	Integrated SAGD/Upgrading
OPS_5	SAGD → DC → H
OPS_6	SAGD → LCF → H
OPS_7	SAGD → LCF → FC → H
	Diluted Bitumen
DB	SAGD

[a] cond = Conditioning, DBE = diluted bitumen extraction, DC = delayed coking, FC = fluid coking, H = hydrotreatment, hydro = hydrotransport, LCF = LC-fining, SAGD = steam-assisted gravity drainage. [b] OPS_4 assumed that 25% of the oil sand processed with this scheme was treated using conditioning, whereas the remaining 75% was processed using hydrotransport.

production schemes, the production levels of each scheme, and environmental constraints such as targets on carbon dioxide emissions. The proposed mixed integer nonlinear program (MINLP) model was developed in the General Algebraic Modeling System (GAMS)[25] and was executed using the Discrete and Continuous Optimizer (Dicopt) as a solver, which is based on the outer-approximation algorithm.[26]

The integrated model features a new spectrum of possibilities to determine, plan, and schedule future Canadian Oil Sands energy demands. The main advantage of the proposed model over previous models[4−6] is that it considers the production schemes (OPS_i) and the SCO production levels (SO_i) simultaneously as decision variables within the optimization formulation. This expands the energy producers' feasible region to search for a combination in the energy producers' infrastructure that can satisfy the total energy demands for the Canadian Oil Sands operation at a lower cost. Therefore, the model can be used as a practical tool to determine the production costs for the Canadian Oil Sands operations, generate future production schemes and energy demand scenarios, and also identify the key parameters that directly affect the Canadian Oil Sands operation.

3. CASE STUDY 2003

The first step considered in the present study was to validate the proposed energy model for the Canadian Oil Sands. Thus, the optimization model described in the previous section was initially used to simulate the Canadian Oil Sands operations in 2003. The year 2003 was selected in this study because information regarding the 2003 production levels for the Canadian Oil Sands operations is available in the literature.[27] Also, a study that shows the energy demands for the specific production schemes and their corresponding production levels for 2003 is available.[4] In addition, the unit costs per barrel of SCO and commercial bitumen produced in 2003 have been reported in the literature.[28]

Table 2. Key Inputs for Case Study 2003[a]

parameter[b]	units	value
boiler feedwater cost	\$/t	1.5
natural gas cost	\$/GJ	5.8
diesel cost	\$/L	0.7
natural gas heating value	MJ/Nm^3	38.05
heat for process steam (SB)	MJ/(t of steam)	3415
heat for SAGD steam (SSEB)	MJ/(t of steam)	2469
boiler capacity	t of steam/h	340
annual operating hours	h/year	8760
plant capacity factors	%	0.90

[a] Costs expressed in 2003 U.S. dollars for this case study. [b] SB = natural gas boilers for process steam at 6300 kPa and 500 °C, SSEB = natural gas boilers for SAGD steam at 80% quality and 8000 kPa.

The optimization modeling tool proposed in this work was validated for a specific production scenario, namely, fixed OPS_i (see Table 1, OPS_1-OPS_4) and SO_i (see eq 6, SO_1-SO_4). Integrated mining/upgrading production schemes were the only schemes considered in this case study. Thus, the number of production schemes (N) was set to 4. Similarly, the potential benefits of using an integrated model were explored assuming that only the total SCO and bitumen productions are given as inputs. In this case, the proposed integrated modeling tool was employed to select the most suitable OPS_i and SO_i ($N = 4$) that minimize the fleet's energy costs.

The key inputs for the 2003 case study are listed in Table 2. For the present case study, SMR hydrogen plants and NGCC power plants without CO_2 capture were considered as the only hydrogen and power plants in this case study (see Table 3), that is, HP_1, $J = 1$, and PP_1, $M = 1$. This was done to mimic the conditions for hydrogen and power production in 2003[4] (see Table 4 for plant details). Because the energy producers considered for 2003 do not account for CO_2 capture, the CO_2 capture constraint shown in the integrated optimization model (see eq 2 and 52) was neglected in the optimization formulation for this case study. Hence, the costs associated with CO_2 capture that appear in the model's objective function shown in eq 52, namely, CO_2 transport costs (CTC) and storage costs (CSC), were set to zero for the present analysis. Furthermore, the present case study assumed that the only process fuel considered for heating during upgrading was natural gas (NG) and that the shovels and trucks fleets used for mining the oil sand consisted of four and five different models, respectively (i.e., $K = 4$ and $L = 5$).

3.1. Model Validation. To validate the model proposed in this work, the production schemes and their corresponding production levels (i.e., OPS_i and SO_i) were specified a priori and represent inputs into the model. This approach, henceforth referred to as the sequential mode, selects only the energy infrastructure (energy plants) and the corresponding operating conditions that minimize the annual production costs of the Canadian Oil Sands for specific settings in the production schemes. As shown in Figure 3, the sequential mode considers that OPS_i and SO_i remain fixed during the optimization calculations. Thus, the energy demands in the sequential mode, which are functions of the production levels in each scheme, also remain fixed during the optimization. Consequently, the energy producers' infrastructure obtained by the optimization model is restricted by the inputs specified for the production schemes. Therefore, the sequential mode is focused on the search for more

Table 3. Energy Producers

energy producer[a]	ref(s)
Boilers	
NG at 6300 kPa and 500 °C steam without CO_2 capture (SB)	15
NG with 80% steam at 8000 kPa without CO_2 capture (SSEB)	15
Power Plants	
NGCC without CO_2 capture (PP_1)	24
supercritical coal without CO_2 capture (PP_2)	24
IGCC without CO_2 capture (PP_3)	22
IGCC with 88% CO_2 capture with Selexol (PP_4)	22
IGCC with 88% CO_2 + H_2S cocapture with Selexol (PP_5)	22
NGCC with 90% CO_2 capture with MEA (PP_6)	24
supercritical coal with 90% CO_2 capture with MEA (PP_7)	24
NG oxyfuel with CO_2 capture (PP_8)	23
coal oxyfuel with CO_2 capture (PP_9)	23
Hydrogen Plants	
SMR without CO_2 capture (HP_1)	16 and 17
coal gasification without CO_2 capture (HP_2)	19 and 20
SMR with 90% CO_2 capture with MEA (HP_3)	16 and 17
coal gasification with 90% CO_2 capture with Selexol (HP_4)	19 and 20
coal gasification with 90% CO_2 + H_2S cocapture with Selexol (HP_5)	19 and 20
gasification of coal without CO_2 capture (HP_6)	19 and 20

[a] NG = natural gas, NGCC = natural gas combined cycle power plants, IGCC = integrated gasification combined cycle power plants, SMR = steam methane reforming hydrogen plants, MEA = monoethanolamine.

Table 4. Energy Producer Modeling Factors

Power Plants				
energy producer	installed capacity (kW)	heating rate (MJ/kWh)	capital cost (\$/kW)	operation and maintenance economic factor (% of capital cost)
PP_1	507000	7.17	570	0.018
PP_2	524000	9.16	1,230	0.038
PP_3	539000	8.76	1,760	0.026
PP_4	448000	11.06	2,400	0.025
PP_5	513000	10.17	1,890	0.026
PP_6	432000	8.41	930	0.037
PP_7	492000	12.04	1,980	0.049
PP_8	440000	7.70	1,250	0.086
PP_9	532000	9.72	1,950	0.076
Hydrogen Plants				
energy producer	installed capacity (t/h)	heating rate [MJ/(t of H_2)]	capital cost [(MM\$ h)/(t of H_2)]	operation and maintenance economic factor (% of capital cost)
HP_1	6.25	174900	11.130	0.060
HP_2[a]	32.09	209000	23.780	0.036
HP_3	6.25	204200	17.760	0.060
HP_4[a]	32.09	209000	25.070	0.036
HP_5	32.09	209000	23.400	0.036
HP_6[a]	32.09	209000	25.070	0.036

[a] HP_2, HP_4, and HP_6 cogenerate 2240, 1210, and 1210 kWh/(t of H_2), respectively.

economically attractive scenarios that can account for different combinations of the energy producers and their corresponding operating conditions. This sequential mode was employed to mimic the 2003 Canadian Oil Sands operations for validation purposes. The values for OPS_i and SO_i for 2003 were obtained from the literature.[27] The results obtained by the model are

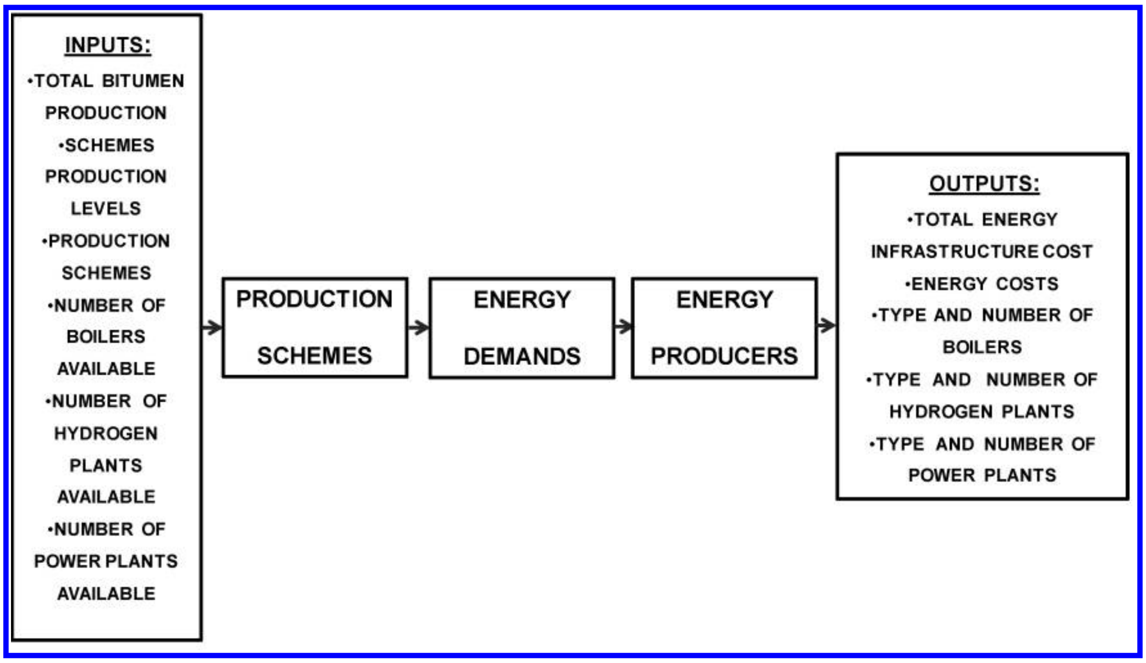

Figure 3. General layout for sequential mode.

Table 5. Simulation Results for Case Study 2003

variable[a]	units	sequential mode	integrated model
	Production Schemes		
OPS_1	t of oil sand mined/year	152 469 006	1238.79
OPS_2	t of oil sand mined/year	45 291 746	841.04
OPS_3	t of oil sand mined/year	43 900 129	308 200 000
OPS_4	t of oil sand mined/year	108 347 364	1216.19
OPS	t of oil sand mined/year	350 008 245	308 203 296
	bbl of SCO/day	538 200	538 200
DB	bbl/day	350 000	350 000
	Energy Demands		
power	kWh	638 640	323 570
steam	t/h	3088	3271.02
hot water	t/h	28 462	24 987.82
diesel	L/h	43 486	38 313.23
hydrogen	t/h	71.77	68.51
process fuel (NG) for DC	N m^3/h	25 103	0.20
process fuel (NG) for LCF	N m^3/h	8325	7286.37
	Annual Costs		
capital	MM$/year	130.2	105.08
operation and maintenance	MM$/year	49.09	39.62
fuel	MM$/year	2,809.83	2,625.93
water	MM$/year	496.52	521.58
total cost	MM$/year	3,485.64	3,292.2

[a] OPS = total oil sand mined, DB = total diluted bitumen production.

reported in Table 5 (sequential mode). The 2003 energy demands and the energy producers' infrastructure obtained with the modeling tool match with those reported by a previous study[4] (see Tables 5 and 6).

One of the key parameters in the current optimization model is the natural gas price. The present case study assumed that the energy producers used only natural gas as a fuel. Also, natural gas was assumed to be the only process fuel in the upgrading stage for heating purposes. Thus, the costs associated with natural gas consumption are expected to have a significant effect on the model's cost function. As shown in Figure 4, the natural gas prices for Alberta in 2003 fluctuated between a minimum of $4.60 and a maximum of \$8.94, with an average cost of \$5.80.[29] To evaluate the significance of this parameter on the unit production costs of SCO and bitumen, the proposed (sequential mode) model was simulated using different natural gas prices for 2003. The selected natural gas price range goes from the lowest to the highest recorded price in that year. Figure 5 shows the sensitivity analysis results obtained from the optimization model and the historical data for the SCO production costs for 2003. As shown in the figure, the predictions on the unit cost of SCO for integrated mining/upgrading production schemes (see Table 1 OPS_1-OPS_4) agree reasonably well with the historical data reported in the literature.[28] Figure 5 also shows the unit cost of SCO corresponding to the average natural gas price and its standard deviation ($\pm$\$1.7). As shown in this figure, the model predicts that these costs are within the range of values reported for the price per barrel of SCO produced in 2003 (\$9−13.5). Although the unit cost for the SCO production that corresponds to the maximum value in the natural gas price is outside the range reported in the literature, that value was considered as rare in the NG prices for 2003 and is not representative of the natural gas prices for 2003 (see Figure 4). Moreover, a similar analysis was made for the commercial diluted bitumen production. The results shown in Figure 6 suggest that the unit production costs per barrel of bitumen produced by SAGD obtained by the proposed model agrees with the range of unit costs reported for 2003.[28] Figure 6 also shows the unit costs for the bitumen when the average value and the corresponding standard deviation were used in the model for this case study.

As can be seen, the results obtained with the optimization model presented in this work agree with those reported in a previous study[4] and with historical data reported for the Canadian Oil Sands in the year 2003.[27,28] Therefore, the optimization

Table 6. Energy Producers' Infrastructure for Case Study 2003

	sequential mode		integrated model	
energy producer	number of units	capacity	number of units	capacity
PP_1	2	319 320 kWh	1	323 570 kWh
HP_1	13	5.52 t/h	13	5.27 t/h

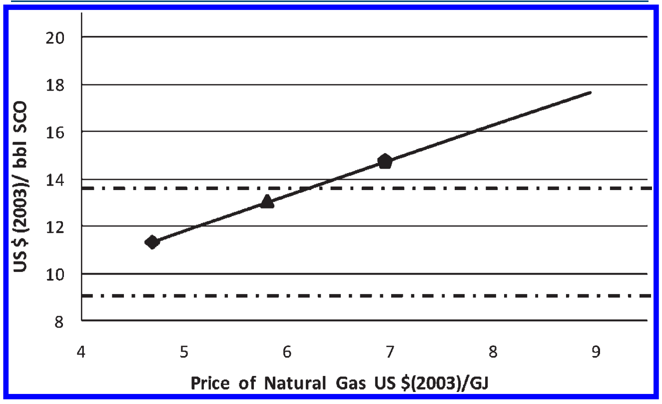

Figure 5. Influence of Alberta's natural gas price on SCO unit production costs for the year 2003.

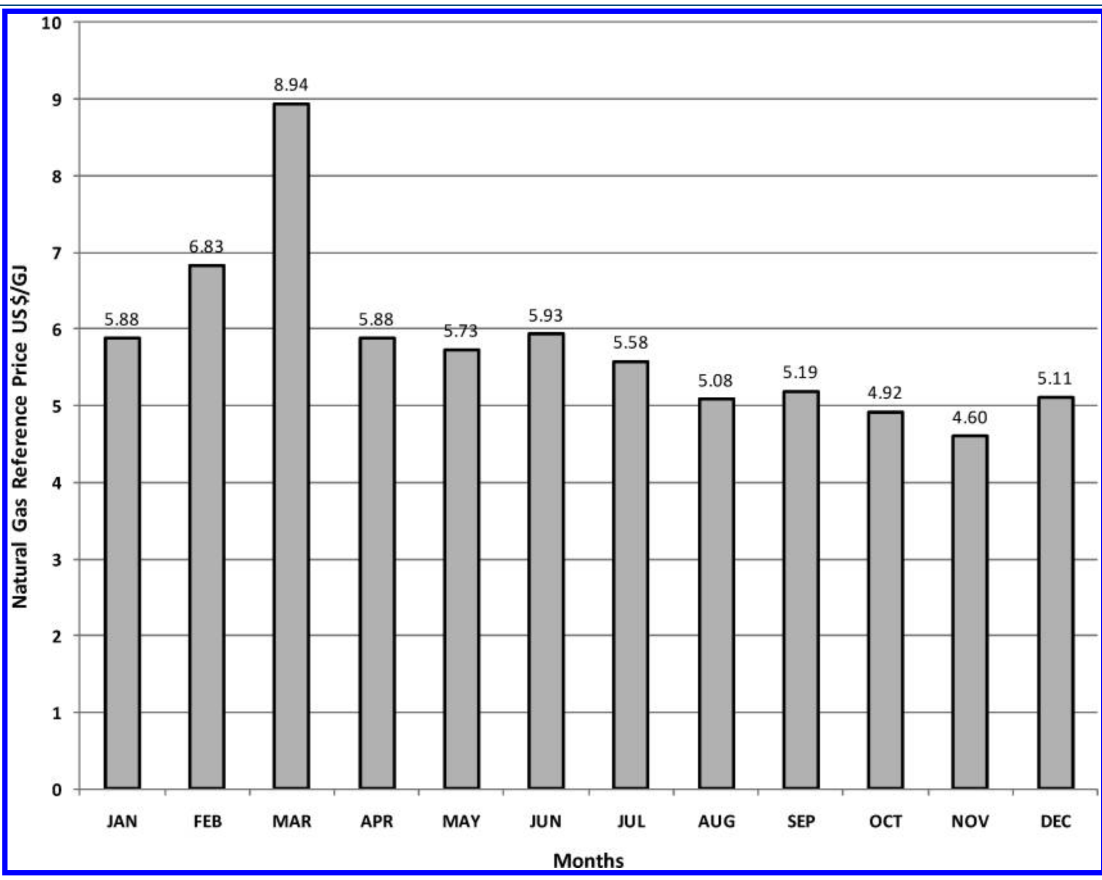

Figure 4. Alberta natural gas reference price history for the year 2003.[29]

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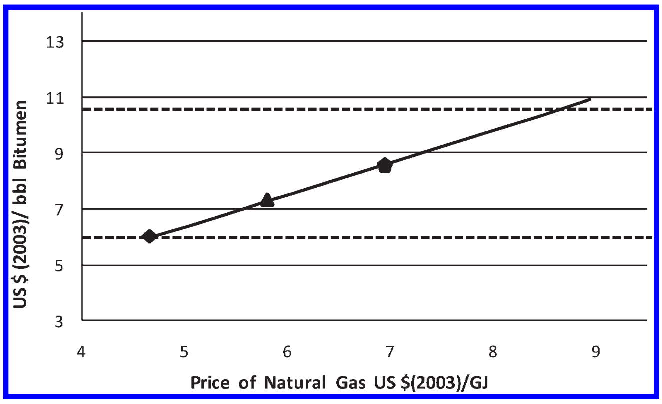

Figure 6. Influence of Alberta's natural gas price on SAGD bitumen unit production costs for the year 2003.

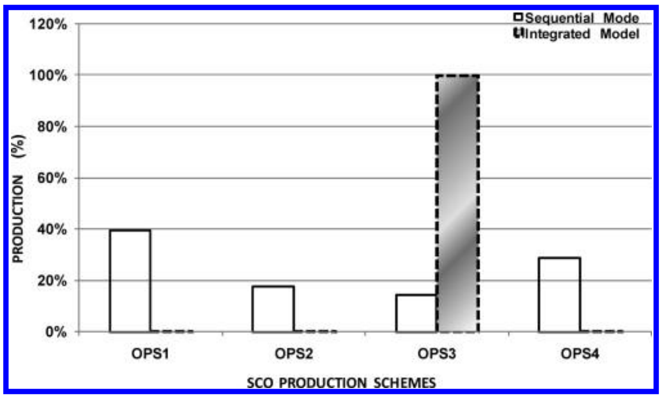

Figure 7. Comparison of SCO production schemes between the sequential mode and the integrated model for the year 2003.

model presented in this study can be used as a complementary tool to predict future production energy costs and potential scenarios for the energy demands and the energy infrastructure for the Canadian Oil Sands operations. However, the current optimization tool can be further validated in the future with updated reports from the Energy Resources Conservation Board[30] and the Canadian National Energy Board.[31] This energy model can be used by Canadian Oil Sands operators, such as Syncrude Canada Ltd., Suncor Energy Inc., and Shell Canada Limited, because the oil production technologies considered in the model are based on the processes used by these traditional oil companies. Also, the model can be used by governmental planning energy entities such as the National Energy Board of Canada to help forecast future energy scenarios that can be used in Energy Roadmaps.

3.2. Simulation of the Integrated Model for 2003. To illustrate the potential benefits of using the proposed integrated model, the 2003 case study was redone with the total diluted bitumen (TDB) and SCO production (TSCO) as the only inputs defined in the model. That is, the production schemes (OPS_1-OPS_4) and their corresponding production levels (SO_1-SO_4) are selected by the optimization algorithm. This represents a main advantage with respect to the sequential mode because the model also selects the most suitable production schemes that need to be used to minimize the total energy costs for the Canadian Oil Sands operations. Therefore, OPS_i and SO_i are treated as decision variables within the optimization model. The optimization results obtained with the sequential mode were used as the initial guesses for this simulation. In this particular scenario, the optimization algorithm searches for combinations in the production schemes, their corresponding levels of operation, the energy infrastructure, and the corresponding operating conditions that minimize the energy costs for the 2003 SCO and diluted bitumen productions. This scenario for the 2003 case study was solved using the MINLP solver DICOPT through the GAMS modeling system. The MINLP algorithm inside DICOPT solves a series of NLP (nonlinear programming) and MIP (mixed integer programming) subproblems. These subproblems were solved using MINOS and CPLEX as NLP and MIP solvers, respectively. MINOS is based on an augmented Lagrangian objective function, and the CPLEX algorithm is based on an implementation of a branch-and-bound search. The proposed optimization problem considered for this scenario consists of 1318 continuous variables and 203 discrete variables. The optimization problem converged after 60 CPU s in a 2.00 GHz machine with 2.038 GB of RAM memory. The solution was found after four major iterations of the outer approximation algorithm. The time required for solving the NLPs represented 64% of the total solution time, whereas the MIP problem required the remaining 36% of the time.

Table 5 (integrated model) shows a summary of the results obtained by the integrated model for 2003. As shown in the table, the integrated model returned a solution that is more economically attractive than that proposed by the sequential mode. The integrated model returned energy savings that are 5.6% (193.4 MM$) higher than those obtained by the sequential mode. Particularly, the average cost per barrel of SCO produced was reduced from 13 to 12 $/bbl (7.7% cost reduction). On the other hand, the cost of the bitumen produced remained constant ($7.28/bbl). This is because only one production scheme was considered in this case study for commercial diluted bitumen production. Figure 7 shows a comparison between the production schemes selected by the integrated model and the production schemes reported for 2003 that were used for the model validation using the sequential mode approach. As shown in this figure, OPS_3 (see Table 1) is the only and preferred SCO production scheme selected by the integrated model. These results suggest that the production schemes that include a combination of thermal cracking and hydrocraking (fluid coking and LC-fining) are the most suitable to be selected compared to those that only use thermal cracking (OPS_1, delayed coking) or hydrocracking (OPS_2). Also, OPS_4, which includes conditioning and hydrotransport as parallel oil sand treating stages, was not selected by the model. Although OPS_4 is based on a combination of thermal and hydrocracking technologies (fluid coking and LC-fining) as is OPS_3, the OPS_4 combination of conditioning and hydrotransport stages is less energy-efficient than treating the total mined oil sand by hydrotransport (scheme OPS_3) because conditioning requires larger amounts of hot water per tonne of oil sand processed than does hydrotransport. Thus, higher costs can be expected from this production scheme as higher energy requirements are needed for the conditioning stage. In addition, OPS_4 consumes steam, which is not used in hydrotransport. Moreover, the considered distance from mining to the extraction plants is 6 times larger for OPS_4 than for OPS_3 (d_4 = 3000 m versus d_3 = 500 m). Thus, the electricity requirements to pump the slurry to the extraction plants are expected to be higher for OPS_4 than for OPS_3. Furthermore, OPS_4 also consumes more process fuel per barrel of SCO produced than does OPS_3. These characteristics favored the selection of OPS_3 over OPS_4 for the present scenario.

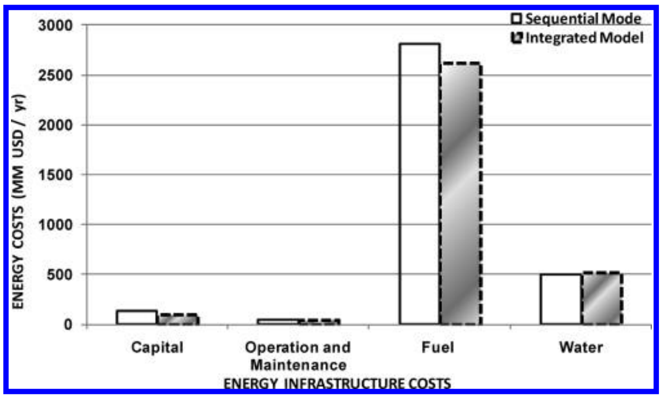

Figure 8. Energy cost comparison between the sequential mode and the integrated model for the year 2003.

The production scheme OPS_2 was not selected by the integrated model because it consumes 2.25 times more electricity than OPS_3. This is mainly because the distance between mining and the extraction plant is 6 times larger for OPS_2 than for OPS_3 (d_2 = 3000 m versus d_3 = 500 m); that is, larger energy requirements are needed for OPS_2. Also, the hydrogen demands are 1.85 times larger for OPS_2 that for OPS_3. This is because OPS_2 uses hydrocracking as the only cracking technology. This technology is highly intensive in consuming hydrogen, which is produced by SMR hydrogen plants that use natural gas as a feedstock. Also, OPS_2 requires 4 times more process fuel in upgrading than OPS_3. Note that the only process fuel considered in the present case study is natural gas. As mentioned above, natural gas is one of the most influential factors in determining the total energy infrastructure cost in the model.

The production scheme OPS_1 was not selected because it consumes 1.25 times more hot water per barrel of SCO produced than OPS_3. Although the hot water requirements per tonne of oil sand processed are the same for both schemes, the output (bbl of SCO) from OPS_3 per tonne of oil sand processed is greater than that from OPS_1, which makes OPS_3 a more efficient scheme. Also, the electricity demands are 2.3 times higher in OPS_1 when compared to OPS_3 mainly because the distance from the mining site to the extraction plant is 5.8 times larger in OPS_1 (d_1 = 2900 m versus d_3 = 500 m); that is, pumps with larger energy consumptions are needed to transport the slurry to the extraction stage. In addition, the process fuel consumption in OPS_1 is 8.7 times larger than OPS_3, because OPS_1 uses thermal cracking as the only cracking technology. Thus, more heating is required during upgrading for this production scheme.

As shown in Table 5, the proposed integrated model reduced the process fuel and electricity demands by 78% and 50%, respectively, with respect to the sequential mode approach. Similarly, the hot water and diesel demands were reduced by 12%, whereas the hydrogen requirement was reduced by 4.5%. Moreover, only one power plant was needed to satisfy the electricity demands. This power plant is an NGCC plant which requires natural gas for the electricity supply. On the other hand, the information reported in a previous study[4] suggests that two NGCC power plants were required to meet the electricity demands (see Table 6). This difference can be attributed to the power demand reduction of 50% obtained with the integrated optimization model proposed in this work.

The annual cost distributions for both the integrated and sequential approaches are shown in Figure 8. This figure shows that the fuel consumed by the production schemes and the energy producers dominate the costs for this year. Hence, the optimization algorithm focuses on these variables to minimize the cost function represented by the annual energy supply costs of the Canadian Oil Sands industry (see cost function in eq 52). The fuel cost is reduced by 6.5% when the integrated model is used. Although the capital and the operation costs are significantly reduced (19.3%), these last two costs represent no more than 5% of the total energy costs. On the other hand, the fuel costs are roughly 80% of the total energy costs. As discussed previously, the process fuel for heating in the upgrading stage and the power demands are the two key process variables that were reduced the most in the integrated approach because they are very sensitive to fuel consumption. The capital costs do not constitute a large contribution to the objective cost function, because it is amortized over the energy producers' book life (30 years). Likewise, the capital cost is distributed along this period of time and does not represent a major financial burden in the model. Water is the other significant cost because of its high consumption for steam and hot water production. Steam is commonly used for SAGD extraction, bitumen upgrading, and process operations, whereas hot water is mostly used for conditioning and hydrotransport.

A key resource for the operation of the Canadian Oil Sands is the water used to separate crude bitumen from the sand in both mining and SAGD operations. Approximately 10 and 3 barrels of water are needed in the mining and SAGD operations, respectively, per barrel of bitumen produced. Although most of the water used is recycled into the process, about 20% of freshwater is required to make up the water losses in the process, which creates serious concerns over the need for water conservation and future sustainability of the operations.[32] The water consumption rate from the Athabasca River for oil operations needs to be improved; otherwise, there may be risks regarding the availability of sufficient water to support the expected expansion of mining-based operations in the near future, especially during winter seasons when the river flow is typically low.[33] Freshwater withdrawal from the river is already limited to protect fish and birds habitat. Canadian Oil Sands operators are required to obtain water licenses that specify both the annual volume of water that can be extracted and the maximum rate of extraction from the Athabasca River.[34] The water license also indicates the annual extraction limits from other water sources, such as groundwater, surface runoff, and tributaries to the Athabasca River. The regulations regarding the water extraction rates of industrial facilities in Alberta were introduced in February 2007 as part of the Lower Athabasca Water Management Framework.[35] The Framework consists of two phases. Whereas the first phase monitors and classifies weekly flow conditions of the Lower Athabasca River, the second phase takes into account environmental, social, and economic factors into the project.[34] The Framework aims to reduce the environmental impacts of industrial water consumption and promote the improvement in water use efficiency to protect aquatic ecosystem during relative sensitive periods (e.g., winter season).

Climate-based models have been developed to predict future changes in flow from the Athabasca region rivers.[34,36−38] According to the simulation studies conducted by Mannix et al.,[34] an estimated average of 6 weeks of water restriction can be expected by the year 2025 based on the current growth of the Canadian Oil Sands operations. These water restrictions are more likely to occur during the winter season. Nevertheless, the variability in

Table 7. Production Scenarios for Case Study 2020

production scenario	units	SCO production (TSCO)	bitumen production (TDB)
high	bbl/day	1 647 000	1 426 000
reference	bbl/day	1 491 000	1 291 000
low	bbl/day	1 130 000	851 000

Table 8. Key Inputs for Case Study 2020[a]

parameter	units	value
boiler feedwater cost	\$/t	1.50
natural gas cost	\$/GJ	6.82
coal cost	\$/GJ	0.74
diesel cost	\$/L	1.25
CO_2 transport cost	(\$)(100 km)/(t of CO_2)	1.30
CO_2 injection cost	\$/(t of CO_2)	7.0
natural gas heating value	MJ/(N m^3)	38.05
coal heating value	MJ/kg	24.05
heat for process steam (SB)	MJ/(t of steam)	3415
heat for SAGD steam (SSEB)	MJ/(t of steam)	2469
boiler capacity	t of steam/h	340
annual operating hours	h/year	8760
plant capacity factors	%	0.90
boiler capacity used for steam	%	0.82

[a] Costs expressed in 2007 U.S. dollars for this case study.

the key factors associated with climate change (e.g., the Pacific decadal oscillation) makes it difficult to detect and predict patterns in the streamflow of the Athabasca region rivers for the upcoming years.[39] The disposal of process water also represents a possible constraint in the Canadian Oil Sands growth given that oil producers operate under a zero-discharge policy. Thus, oil producers companies are required to store the process water and tailings on-site, which has led to the construction of over 70 km^2 of tailings ponds, deposits of residues from oil operations, and a considerably large inventory of waste.[33] In 2004, the volume of impounded process water at Syncrude's Lease 17/22 was approaching 1 billion m^3, whereas the volume of impounded tailings sludge generated from the Canadian Oil Sands operators has exceeded 700 million m^3.[33] The present energy model estimates the fresh water consumption costs associated with the production of steam and hot water. Current work on this research considers the addition of water management within the energy model to include a freshwater consumption constraint, water recycling in the mining production schemes, and water treatment technologies for the Canadian Oil Sands operations.

4. CASE STUDY 2020

The integrated model presented in this work was also used to determine the energy infrastructure and the potential energy costs for the operation of the Canadian Oil Sands in year 2020. This year was selected as a case study because current estimates of energy prices and economic projections with governmental programs are available in a recent report issued by the National Energy Board of Canada (NEB).[2] Although the NEB released a report in 2007 with projections for the year 2030,[40] this report did not consider the financial crisis in the energy sector that occurred in 2008. Hence, the updated report used in this case study takes

Table 9. Simulation Results for Case Study 2020 without CO_2 Capture

variables	units	low production	reference	high production
		Production Schemes		
OPS_1	bbl/day	0	0	0
OPS_2	bbl/day	115 750	161 570	0
OPS_3	bbl/day	550 500	716 130	1 021 570
OPS_4	bbl/day	0	0	0
OPS_5	bbl/day	0	0	121 280
OPS_6	bbl/day	463 750	613 300	504 150
OPS_7	bbl/day	0	0	0
OPS	bbl/day	1 130 000	1 491 000	1 647 000
DB	bbl/day	851 000	1 291 000	1 426 000
		Energy Demands		
power	kWh	783 910	1 063 800	1 031 600
steam	t/h	5081	6682	8017
hot water	t/h	30 053	39 521	47 421
SAGD steam	t/h	20 557	29 836	32 767
hydrogen	t/h	180.1	238.26	240
process fuel (NG)	N m^3/h	39 522	52 576	56 046
diesel	L/h	46 079	60 597	72 709
		Annual Costs		
power	MM\$/year	224.91	381.28	379.31
hydrogen	MM\$/year	1,907.80	2,460.20	2,462.20
hot water	MM\$/year	624.31	821.01	985.11
process steam	MM\$/year	1,105.50	1,453.90	1,744.50
SAGD steam	MM\$/year	2,719.50	3,947.10	4,334.90
process fuel	MM\$/year	89.84	119.52	127.41
diesel	MM\$/year	504.56	663.54	796.16
total cost	MM\$/year	7,176.42	9,846.55	10,829.59
		Unitary Costs		
mined SCO	\$/bbl	12.69	12.71	12.58
SAGD SCO	\$/bbl	13.32	13.30	13.34
diluted bitumen	\$/bbl	5.92	5.94	5.95

into account this unforeseeable event that changed the economic perspective and forecasts for the Canadian Oil Sands operations. The key factors that affected the upcoming scenarios for the Athabasca region was the unexpected increase in the oil prices (\$147/barrel[2]) followed by a sudden reduction in the value of the oil (\$60/barrel[2]) during the early stages of the financial crisis in 2008. These factors, together with new environmental policies, have changed the global oil business perspective for the future.

Table 7 lists the highest, lowest, and reference SCO and bitumen productions, that is, total SCO and bitumen production (TSCO and TDB), expected for the year 2020 in the Canadian Oil Sands. These production scenarios were used as inputs in the integrated optimization model to predict the most suitable combination of production schemes and energy infrastructures that minimize the production costs for 2020. To propose a more realistic scenario, all of the production schemes shown in Table 1 are considered for this case study, namely, OPS_1-OPS_7 ($N = 7$) Also, Table 3 shows all of the energy producers considered to supply the energy demands for 2020, namely, HP_1-HP_6 ($J = 6$) and PP_1-PP_9 ($M = 9$). In addition, the shovel and truck fleets used for mining the oil sand were assumed to be composed of

Table 10. Energy Producers' Infrastructure for Case Study 2020 without CO_2 Capture

	low		reference		high	
energy producer	no. of units	capacity	no. of units	capacity	no. of units	capacity
SB	20	254.05 t/h	29	230.41 t/h	35	229.06 t/h
SSEB	61	337 t/h	90	331.51 t/h	98	334.36 t/h
HP_2	3	25.466 t/h	4	26.27 t/h	4	26.635 t/h
		186 640 kW		256 730 kW		260 270 kW
HP_6	4	25.922 t/h	5	26.635 t/h	5	26.635 t/h
		125 670 kW		161 410 kW		161 410 kW
PP_1	–	–	1	174 090 kW	1	174 510 kW
PP_2	1	471 600 kW	1	471 600 kW		435 440 kW

four ($K = 4$) and five ($L = 5$) different models, respectively. This first scenario for 2020 does not consider CO_2 capture, that is, the terms associated with the CO_2 capture in the cost function [i.e., CO_2 transport costs (CTC) and storage costs (CSC) in eq 52] were initially neglected. However, an additional scenario that considers CO_2 capture for this case study is presented at the end of this section. The main economic parameters included in the optimization model, namely, natural gas, coal, and CO_2 storage and transport costs, are listed in Table 8. As in the 2003 case study, the resulting MINLP optimization model was coded in GAMS and solved using the MINLP solver DICOPT. The problem consisted of 4126 continuous variables and 840 discrete variables. The solutions for the high, reference, and low production scenarios required CPU times (2.00 GHz, 2.038 GB RAM) of 125 006, 138 468, and 124 642 s, respectively.

The results obtained for these scenarios are reported in Tables 9 and 10. These tables show that over 62% of the total energy costs are represented by the hydrogen and SAGD steam generation costs. The average unitary costs are \$12.66, \$13.32, and \$5.94 for mined SCO, SAGD SCO, and diluted bitumen, respectively. The hydrogen producers are coal gasification plants, and the power producers are NGCC and supercritical pulverized coal power plants. Figure 9 shows the distribution between the production schemes selected by the integrated model for each scenario. As shown in this figure, the most suitable synthetic crude oil production schemes are OPS_3 and OPS_6. Although the 2003 case study did not include integrated SAGD/upgrading schemes (OPS_5−OPS_7), the predictions obtained for 2020 shows that OPS_3 remains as the main oil producer. Thus, the results obtained for the present case study are consistent with those obtained for the 2003 case study. As mentioned above, OPS_3 is the preferred scheme because it is the most energy-efficient per barrel of SCO produced among the integrated mining/upgrading schemes. Historically, SAGD bitumen extraction has been more expensive than mined bitumen extraction, according to reports from the National Energy Board of Canada.[28,41] Therefore, the mining/upgrading scheme (OPS_3) is less expensive than the integrated SAGD/upgrading scheme (OPS_6).

Among the integrated mining/upgrading schemes, OPS_2 is the second largest production scheme selected by the optimization model. This is because OPS_2 is based on hydrocracking, which uses hydrogen to upgrade the bitumen. For this case study, IGCC hydrogen plants are available that use coal as the feedstock fuel and cogenerate power. According to information reported by the National Energy Board of Canada,[2] the reference prices of natural gas and coal in the year 2020 are expected to be US\$ (2008) 7.50/MMBtu and C\$ (2008) 0.82/GJ, respectively. Thus, coal is considered to be 9.2 times less expensive than natural gas, and the hydrogen produced with IGCC for OPS_2 is less expensive than producing oil from OPS_1 because it requires large amounts of NG as the process fuel. The results for the three scenarios show that the model selected only IGCC hydrogen plants to cover the hydrogen requirements. This is indeed the cheapest technology considered in the model to produce hydrogen and power simultaneously. Likewise, OPS_2 is more economically attractive than OPS_4 because it requires less hot water and process steam per barrel of SCO produced. These two energy commodities are produced in natural-gas-fired boilers.

OPS_6 is the most suitable integrated SAGD/upgrading scheme because it is based on hydrocracking, which is a cheap process because the hydrogen is produced in IGCC plants. Moreover, OPS_6 was modeled with the highest SCO conversion among these schemes (95%). OPS_5 is the other selected scheme from this group (SAGD/upgrading scheme). Although this last scheme depends on thermocracking, which consumes large amounts of process fuel, OPS_7 also includes thermocracking as part of its upgrading process. Additionally, OPS_7 consumes 5.77 times more power than OPS_5 during upgrading.

Consequently, the results show that the optimization model focuses on reducing the natural gas consumption, which is the dominating cost in the model's cost function. Thus, the production schemes that required considerable amounts of process fuel, hot water, and process steam are lessened by the optimization model because they rely on natural gas as the feedstock fuel.

The energy cost breakdowns for the scenarios considered for this case study are shown in Figure 10. As shown in this figure, steam production for SAGD extraction is the dominant energy cost. This is because SAGD steam is used in SAGD SCO and diluted bitumen production. Moreover, SAGD steam production is highly energy-intensive; that is, it requires 2469 MJ per tonne of SAGD steam produced. During SAGD extraction, 2.4 t of steam is required per tonne of bitumen recovered. The average unit costs per integrated scheme for SCO and bitumen production are shown in Table 9. (Costs are in 2007 U.S. dollars.) The average production cost per barrel of mined SCO is lower than SAGD SCO because mining extraction is cheaper than SAGD extraction. This is because a significant amount of SAGD steam is used to recover the bitumen contained in the sand. Although SAGD extraction is more expensive, it is estimated that 80% or more of the bitumen reserves in the Canadian Oil Sands required thermal extraction methods for its recovery.[1] Therefore, in this study, thermal extraction takes an important place given its promising future as the leading extraction method in the future

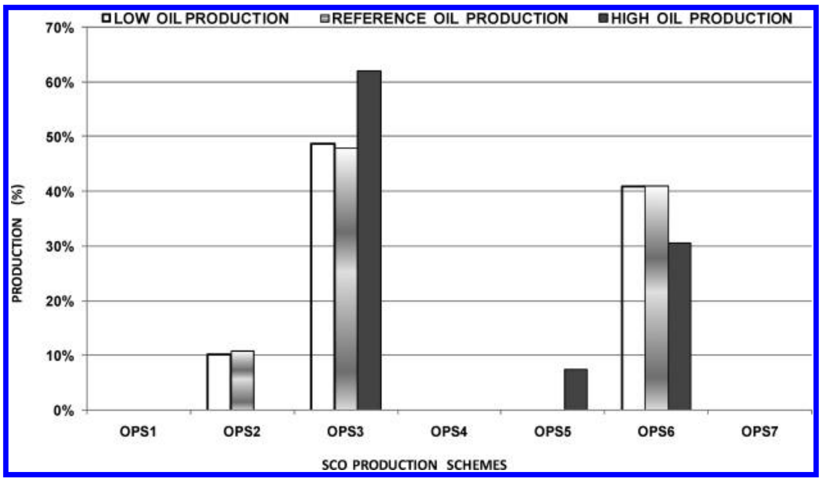

Figure 9. Comparison of SCO production schemes for different price scenarios for the year 2020.

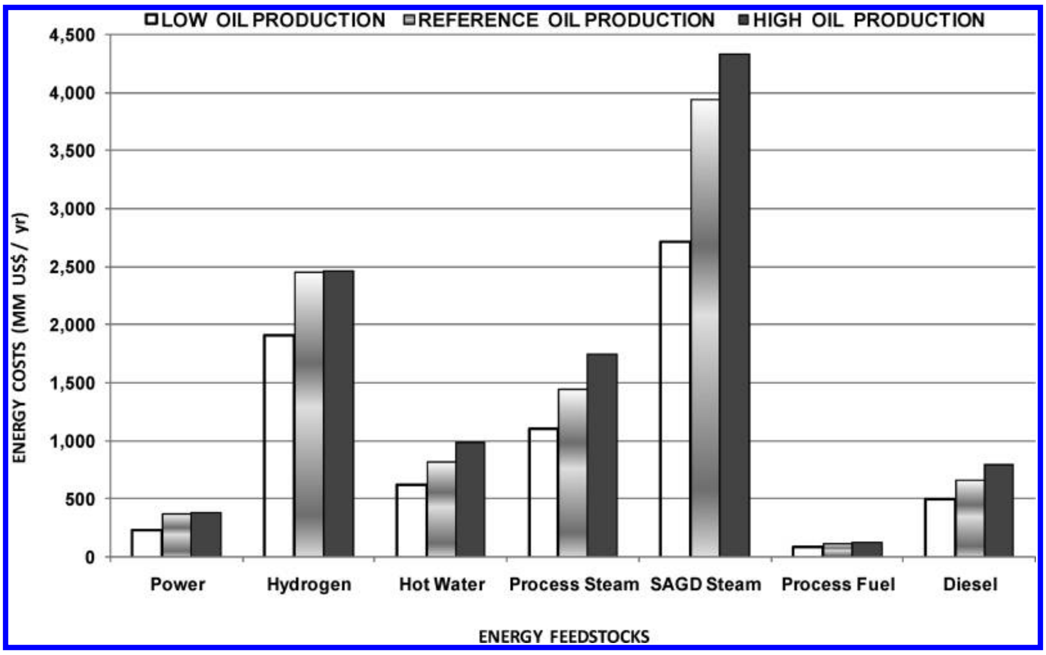

Figure 10. Energy costs for the different price scenarios for the year 2020.

of Canadian Oil Sands. Furthermore, by 2020, thermal bitumen extraction is projected to overtake mining extraction combining SCO and diluted bitumen production. Figure 10 also shows that the second highest energy cost is the hydrogen cost. According to the results, hydrogen will be required in three out the four production schemes selected by the optimization model (OPS_2-OPS_3 and OPS_6). This is because the upgrading processes for these schemes are based only on hydrocracking (OPS_2 and OPS_6) or a combination of hydro- and thermocracking (OPS_3). Moreover, hydrocracking is part of the two schemes that produced almost 90% of the total SCO. On the other hand, the process fuel (natural gas) represents the lowest energy cost. This is because the integrated model selects hydrocracking-based schemes over thermal-cracking-based schemes. As mentioned above, schemes that include hydrocracking are most suitable to be selected because hydrogen production is cheaper because IGCC plants are available in the model. The IGCC plants consume coal as the fuel feedstock, whereas the thermal cracking depends mainly on process fuel (NG) for heating purposes, to crack the bitumen in upgrading.

The Canadian Oil Sands industry is sensitive to global and regional economics, technological developments, and changes in governmental programs and policies. However, one of the key factors that affects and drives the operation of this industry is related to the natural gas price. The most significant change to North American natural gas markets in the future is closely related to the development of new technologies that can be implemented to recover natural gas from shale gas reservoirs and other unconventional gas resources. These technological developments are advancing rapidly in the United States and are starting to emerge as an attractive option for natural gas production in Canada.[2] Commercial-scale production from unconventional gas reservoirs can be achieved through technological advances in horizontal drilling and hydraulic fracturing. These two technologies for unconventional gas reservoirs enable the recovery of vast natural gas resources, thus boosting the natural gas supply in North America and making up for the reduction in the production of natural gas from conventional resources. Canadian natural gas is connected to the North American market through a network of pipelines that allows gas buyers to purchase and transport gas from a number of supply sources across the continent.[42] The recent increase in unconventional natural gas resources can ease the tight supply/demand balance and contribute to a decrease in natural gas prices in the future. Thus, Canadian Oil Sands producers can continue relying on natural gas as the main feedstock fuel for energy commodity productions and plant processes. Accordingly, new technological developments for alternative energy production, especially those related to coal gasification processes, might not be considered as key developments for the Canadian Oil Sands because the natural gas price will likely become closer to the coal price. From the environmental point of view, namely, GHG emissions, the use of natural gas is preferred over the use of coal technologies to comply with expected CO_2 emissions regulations in the Canadian Oil Sands industry. Furthermore, natural gas prices can also be sensitive to crude oil prices, as some consumers in the United States can switch between natural gas and fuel oil to cover their energy needs. Thus, this competition produces a relationship between oil and natural gas prices; specifically, an increase in oil prices generate an increase in the gas price.[42]

One of the challenges faced by the Canadian Oil Sands industry is the development of new transportation networks to deliver the hydrocarbon products to the market. In this regard, the Canadian and the U.S. government are negotiating the terms to construct and operate a crude oil pipeline and related infrastructure to transport crude oil from an oil supply hub near Hardisty, Alberta, Canada, to the south central region of the United States (e.g., Oklahoma and Texas). The proposed project, Keystone XL, would consist approximately of 327 and 1384 miles of pipeline in Canada and the United States, respectively. The project would have an initial nominal transport capacity of 700 000 bbl of crude oil per day. Because the pipeline project crosses the international border between Canada and the United States, a presidential permit issued by the U.S. Department of State is required for the project to proceed.[43] This subjects the Keystone XL Project to the National Environmental Policy Act (NEPA), which requires the disclosure of potential environmental impacts and the consideration of possible alternatives.[43] Currently, the U.S. Environmental Protection Agency (EPA) has rejected the Environmental Impact Statements for this project submitted by the U.S. Department of State. This is because significant environmental impacts have not been evaluated, and additional information and analyses are required to thoroughly weigh the environmental costs and benefits of transporting Canadian Oil Sands crude from Canada to the south central region of the United States.[44] The major environmental impacts include potential oil spills in communities and environmental justice concerns, environmental and health impacts to communities along the pipeline and adjacent to the refineries, lifecycle GHG emissions associated with oil sands crude, and impacts to wetlands and migratory bird populations. Also, TransCanada, the company that would likely build the pipeline, has recognized the challenges related to restore the productive capability of the lands disturbed by the construction of the pipeline network, especially the native rangeland from North and South Dakota and central Nebraska.[45] The approval of the project will guarantee the further development and expansion of the Canadian Oil Sands industry because the south central region of the United States is the largest refining region in the United States, with the capacity to process a wide range of both light and heavy crude oil types. A key aspect that will promote the development of this project is the implementation of environmental remediation,[45] which includes CO_2 capture and storage, return of the subsoil to the trench, topsoil respread to original reclamation lands, revegetation of the affected areas with native species, the use of straw or native prairie hay to prevent wind erosion in the soil, the use of hodder gaugers or imprinters to create impression in the soil thus reducing erosion, improvement of moisture retention, creation of microsites for seed germination, use of sediment logs (straw wattles) or slope breakers to manage soil erosion, and use of biodegradable materials in place of metal when possible. In the present energy model, only reduction of GHG emissions was considered as an environmental remediation, which is one of the key environmental concerns for this industry. Future work on this research considers the addition of other environmental remediation options in the energy model to predict the operation of the Canadian Oil Sands under these additional environmental restrictions.

Table 11. Simulation Results for Case Study 2020 with CO_2 Capture

variable	units	value
	Production Schemes	
OPS_1	bbl/day	0
OPS_2	bbl/day	0
OPS_3	bbl/day	993 520
OPS_4	bbl/day	0
OPS_5	bbl/day	460 800
OPS_6	bbl/day	36 680
OPS_7	bbl/day	0
OPS	bbl/day	1 491 000
DB	bbl/day	1 291 000
CO_2 emissions	t/h	5588
total cost	MM$/year	10,390

4.1. Simulation under Governmental Plan to Reduce Greenhouse Gases for 2020. The present case study was also used for determining the energy infrastructure and the corresponding costs following a report published by the Canadian Federal Government: *Turning the Corner*.[46] This report is a notice of intent to develop and implement regulations for reducing greenhouse gas (GHG) and air pollution emissions from the industry. According to this plan, the Canadian Oil Sands GHG emissions for 2020 should be under 50 Mt of CO_2 equivalent. Therefore, the model proposed in this work was used to simulate the Canadian Oil Sands operations subject to this environmental constraint (see eq 2 and the CO_2 constraint in problem 52). This scenario was run for the 2020 reference oil production scenario (see Table 7). Tables 11 and 12 summarize the key results obtained with the CO_2 emissions constraint for the year 2020. As shown in these tables, the most suitable synthetic crude oil production schemes are OPS_3 and OPS_5. Both schemes represent over 97.5% of the total SCO production expected for 2020 with this environmental restriction. Moreover, OPS_3 constitutes two-thirds (2/3) of the total SCO production, i.e., OPS_3 remains as the main oil producer. This is because OPS_3 is an integrated mining/upgrading scheme (cheaper scheme). In situ production schemes, such as OPS_5−OPS_7, create higher

Table 12. Energy Producers' Infrastructure for Case Study 2020 with CO_2 Capture

energy producer	no. of units	capacity
SB	29	268.86 t/h
SSEB	91	333.33 t/h
HP_2	1	26.635 t/h
		65 069 kW
HP_4	6	26.635 t/h
		193 690 kW
PP_1	1	395 920 kW
PP_6	1	215 520 kW

GHG emissions; that is, in situ production (without upgrading) generate on average 2.5 times more CO_2 emissions than mining (without upgrading) per barrel of bitumen produced.[47] Therefore, the model selects integrated mining/upgrading schemes over integrated SAGD/upgrading schemes. The schemes with mining present very similar GHG emissions per barrel of SCO produced. However, OPS_3 is the most energy-efficient, as described in previous sections.

The principal integrated SAGD/upgrading scheme selected is OPS_5 because it uses thermocracking instead of hydrocracking in the upgrading stage. The cheapest process to produce hydrogen in the model is through IGCC plants, which have the highest rate of CO_2 emissions per tonne of hydrogen produced (17.26 t of CO_2 per tonne of H_2). Therefore, the optimization model selects thermocracking over hydrocracking-based schemes to meet the user-specified emissions target. Also, around 2.5% of the SCO is produced by scheme OPS_6 because it assumes the highest SCO conversion among the integrated SAGD/upgrading schemes. The results reported in Table 11 also suggest that the Canadian Oil Sands energy costs are expected to be 5.3% higher when compared to the reference production case without CO_2 capture. This is because the model selects a new distribution of production schemes and energy producers that generates less CO_2 at a higher cost. Also, part of the increase in the energy costs for this scenario was due to the costs associated with CO_2 storage and transport that were considered in the energy cost function.

Current technological developments have the potential to impact the future of the Canadian Oil Sands industry. One of these developments is the introduction of new processes for moderate primary upgrading, namely, Snamprogetti and Hydrocarbon Research Institute (HRI) licensors.[48] Snamprogetti is a technology based on a slurry catalyst reactor whose operation is similar to that of a multiphase flow reactor, that is, the reactant (gas) is bubbled through a solution that contains catalyst particles (solid). A key feature of this new technology compared to current technologies is that it enables heat recovery and a more adequate temperature control.[49] Similarly, HRI is a technology based on an ebullated bed reactor in which the catalyst particles are held in suspension by the upward flowing stream of the liquid reactant and gas flow. The velocity of the fluid is adequate to hold the particles in suspension, but not sufficiently large to carry the particles out of the vessel. The solid particles revolve around the bed, creating excellent mixing among them.[49] Both the HRI and Snamprogetti technologies can consider the use of byproduct, such as heavier asphaltene-rich residues, to produce hydrogen and other energy commodities, such as steam. Therefore, the use of these byproducts can reduce the dependence of the Canadian Oil Sands operators on natural gas.[48] Consequently, the use of more moderate primary upgrading processes that consider byproducts to generate hydrogen could significantly decrease the upgrading costs. The synthesis gas produced from the gasification of residues also offers the potential to generate hydrocarbons that are suitable for producing high-quality distillates with controlled carbon chain size.[48]

Moreover, the introduction of nanoengineered catalysts with novel materials, such as Azko's Nebula,[50] has the potential to significantly increase SCO conversion by an order of magnitude when compared to their predecessors.[48] The key for these improvements is the conversion of low-value hydrocarbons, such as aromatic compounds, to ultra-low-sulfur diesel, which will substantially increase the SCO quality and value. The energy model proposed in this work may be expanded by incorporating the technological developments mentioned above. Therefore, the energy model can include new SCO production schemes that consider moderate primary upgrading processes, namely, Snamprogetti and HRI, and gasification plants that use the process residues to produce hydrogen. Also, SCO producers with upgrading stages that take into account nanoengineered catalysts can also be considered in the future development of the current energy model. The addition of these new technological developments within the energy model will enable a sensitivity analysis that predicts the effects of these developments on the energy costs and the Canadian Oil Sands operations. Also, the oil producers and energy commodity producers' infrastructure obtained from simulations that consider these developments can be compared to the infrastructure obtained when those developments are not considered in the analysis. Thus, niches for new process improvements in the Canadian Oil Sands operations may be identified from these analyses.

5. CONCLUSIONS

A comprehensive integrated model is proposed in the current work for reducing the production costs and forecasting the Canadian Oil Sands energy demands. The proposed model minimizes the total energy cost of the Canadian Oil Sands operations by selecting the most suitable production schemes with the corresponding production levels, the energy producer infrastructure (power and hydrogen plants, and boilers) along with its operating conditions for a given CO_2 emissions target. The proposed energy model was validated using the Canadian Oil Sands operation reported for 2003. The simulation results obtained with the sequential mode, that is, production schemes and corresponding operating conditions fixed at constant values, showed that the energy demands and the energy infrastructure correspond to those reported in a previous study[4] and the production costs match with the 2003 historical data reported in the literature.[28] To demonstrate the potential benefits of using an integrated modeling approach, the 2003 case study was solved assuming that the total SCO and bitumen production for 2003 were given as inputs to the model. The results showed that the integrated approach returned savings that are 5.6% higher than those obtained by the sequential mode. The proposed model was also used to forecast the Canadian Oil Sands operations for 2020. The proposed 2020 case study was simulated for three different production scenarios where the corresponding SCO and bitumen production forecast values fluctuates among low, high, and reference values. The results obtained with the integrated model suggest that hydrocracking-based schemes are more attractive than thermocracking-based production technologies. This is because hydrogen can be produced in IGCC plants that use coal as feedstock fuel and also cogenerate power. On the other hand, thermocracking depends mainly on process fuel (natural gas), which costs much more than coal. Moreover, OPS_3 (see Table 1) is the most suitable production scheme in the model because is an integrated mining/upgrading scheme that includes hydrocracking and is an energy efficient scheme. When CO_2 capture was included as an environmental constraint for 2020, OPS_3 remained as the main SCO producer, whereas OPS_5, the cleanest integrated SAGD/upgrading scheme, became the second largest producer. Also, the total energy costs for the 2020 case study increased by 5.3% when CO_2 capture was considered as an environmental constraint in the study. This result indicates the level of compromise between capturing or not CO_2 in the Canadian Oil Sands.

The results presented in this work show that the integrated model can be used as a practical tool to analyze the production costs of the Canadian Oil Sands. Also, this tool can be used for planning and scheduling the current and future energy producers' infrastructure. Current work on this area is focused on the addition of alternative energy production schemes, such as nuclear energy, and an in-depth analysis of the CO_2 emissions and water consumption under different scenarios for the Oil Sand operations. Furthermore, future work on this research includes the development of a multiperiod energy optimization model that takes into account fluctuations of the model's key process variables, such as fluctuations in time of the gas prices. The current modeling tool predicts the potential oil producers and energy commodity producers' infrastructures for projected total SCO and commercial crude bitumen levels at steady state. However, a model that describes the operation of the Canadian Oil Sands in the presence of fluctuations in the key process variables over a finite time horizon or variations in the economic and environmental constraints, e.g., CO_2 capture and storage costs, is also of interest. Thus, the energy model presented in this work can be used as a basis to develop a multiperiod energy optimization model for the Canadian Oil Sands. The multiperiod model would enable the users to determine medium- to long-term economic and environmental impacts due to variations in the key process variables that are expected to affect the operation of this industry.

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NOMENCLATURE

Acronyms

ATB = atmospheric topped bitumen
CDB = commercial diluted bitumen
DC = delayed coking
DCB = diluted crude bitumen
DCU = delayed coker unit
DRU = diluent recovery unit
EPA = U.S. Environmental Protection Agency
FC = fluid coking
FCU = fluid coker unit
GHG = greenhouse gas
H = hydrotreatment
HGO = heavy gas oil
IGCC = integrated gasification combined cycle
LCF = LC-fining
LCFU = LC-finer unit
LGO = light gas oil
MEA = monoethanolamine
NEPA = National Environmental Policy Act
NG = natural gas
NGCC = natural gas combine cycle
NT = naphtha
SAGD = steam-assisted gravity drainage
SB = natural-gas-fired boilers for process steam and hot water production
SCO = synthetic crude oil
scf = standard cubic foot
SCPC = supercritical pulverized coal
SMR = steam methane reforming
SSEB = SAGD steam boilers
VDU = vacuum distillation unit
VTB = vacuum topped bitumen

Model Variables

ACC_j = annual capital cost of hydrogen plant j (\$/year)
ACC_m = annual capital cost of power plant m (\$/year)
ACF_j = amortization capital factor for hydrogen plant j (%)
ACF_m = amortization capital factor for power plant m (%)
$ATBF_i$ = atmospheric topped bitumen feeding the LC-finers in SCO scheme i (t/h)
ATB_i = atmospheric topped bitumen in SCO scheme i (t/h)
BF_i = bitumen froth in extraction plant for scheme i (t of froth/h)
BIT_i = bitumen rate from SAGD for SCO production in scheme i (t/h)
CCH_j = amount of CO_2 captured in hydrogen plant j (t/h)
CCP_m = amount of CO_2 captured in power plant m (t/h)
CDB = commercial diluted crude bitumen (bbl/day)
CF = objective cost function (\$/year)
CO_2B = baseline carbon dioxide emissions of the Canadian Oil Sands operations (t/h)
CCO_2 = percentage CO_2 capture (%)
CSC = annual carbon dioxide storage cost (\$/year)
CTC = annual transportation cost of the CO_2 captured (\$/year)
D = total diesel demand (L/h)
D_{TC} = total annual cost of diesel (\$/year)
$DBIT_i$ = diluted bitumen entering into upgrading in SCO scheme i (t/h)
DBR = bitumen rate from SAGD for commercialization (bbl/day)
DSH = diesel consumed by the shovels' fleet (L/h)
DT = diesel consumption by the trucks' fleet (L/h)
FB_i = LC-finer bottom oil fractions in SCO scheme i (t/h)
F_j = fuel consumed by hydrogen plant j (m^3/h)
H_{HC} = hydrogen demand for hydrocracking (t/h)
H_{HT} = hydrogen demand for hydrotreatment (t/h)
H_{TC} = total annual cost of hydrogen production (\$/year)
H_U = total hydrogen demand (t/h)
HP_j = hydrogen produced in plant j (t/h)
HW_{BE} = hot water demand in bitumen extraction plant (t/h)
HW_C = hot water demand in conditioning (t/h)
HW_H = hot water demand in hydrotransport (t/h)
HW_{TC} = annual cost of hot water (\$/year)
HWD = total hot water demand (t/h)
NGSB = consumption of natural gas in process steam boilers (Nm^3/h)
NGSEB = consumption of natural gas in SAGD steam boilers (Nm^3/h)

OC = operating conditions for energy and oil producers
OMC_j = annual operation and maintenance cost for hydrogen plant j (\$/year)
OMC_m = annual operation and maintenance cost for power plant m (\$/year)
OMF_j = operation and maintenance economic factor for hydrogen plant j (%)
OMF_m = operation and maintenance economic factor for power plant m (%)
OSR_i = mined oil sand rate for SCO production scheme i (t/h)
P_{BE} = power demand in bitumen extraction plant (kW)
P_H = power demand in hydrotransport stage (kW)
P_{TC} = total annual cost of power (\$/year)
P_U = total electricity demand in upgrading (kW)
P_{UD} = power demand in delayed-coking-based schemes (kW)
P_{UF} = power demand for scheme including LC-fining and fluid coking (kW)
P_{UL} = power demand in LC-fining-based scheme (kW)
PC_i = power demand in centrifugation for scheme i (kW)
PCTH = power demand to transport CO_2 captured in hydrogen plants (kW)
PCTP = power demand to transport CO_2 captured in power plants (kW)
PD = total power demand (kW)
PF_{TC} = annual cost of process fuel (\$/year)
PF_U = total process fuel demand in upgrading (m^3/h)
PF_{UD} = process fuel demand on delayed-coking-based scheme (m^3/h)
PF_{UF} = process fuel demand for scheme including LC-fining and fluid coking (m^3/h)
PF_{UL} = process fuel demand in LC-fining-based scheme (m^3/h)
PG_j = power cogenerated in gasification hydrogen plant j (kW)
PG_m = power generated by plant m (kW)
PHP = power demand of SMR hydrogen plants (kW)
PSE = power demand in SAGD extraction (kW)
PT_i = power demand to pump tailings to disposal from scheme i (kW)
S_{BE} = steam demand in bitumen extraction plant (t/h)
S_C = steam demand in conditioning stage (t/h)
S_{TC} = annual cost of process steam (\$/year)
S_U = steam demand for upgrading (t/h)
SD = total process steam demand (t/h)
SH_k = number of vehicles of model k used in shovels' fleet
SO_i = mined and SAGD bitumen upgraded to SCO produced by scheme i (bbl of SCO/day)
SSE = total steam consumption in SAGD extraction (t/h)
SSE_{TC} = annual SAGD steam cost (\$/year)
ST_i = slurry in hydrotransport for scheme i (t/h)
T_l = number of vehicles of model l used in the truck's fleet
VTB_i = vacuum topped bitumen in SCO scheme i (t/h)

Binary Variables

HC(i) = 1 if i follows hydrocracking upgrading scheme (i.e., eqs 4 and 5), 0 otherwise
HPC(j) = 1 if hydrogen plant j captures CO_2, 0 otherwise
HPG(j) = 1 if hydrogen plant j is of type gasification, 0 otherwise
HPS(j) = 1 if hydrogen plant j is of type SMR, 0 otherwise
$LF_C(i)$ = 1 if i follows production scheme with a conditioning stage, 0 otherwise
OPS_i = 1 if SCO production scheme i exists in the operation, 0 otherwise
PPC(m) = 1 if power plant m captures CO_2, 0 otherwise
$PS_M(i)$ = 1 if i follows an integrated mining/upgrading SCO production scheme, 0 otherwise
$PS_S(i)$ = 1 if i follows an integrated SAGD/upgrading SCO production scheme, 0 otherwise
$UR_D(i)$ = 1 if i follows the upgrading route in eq 3, 0 otherwise
$UR_F(i)$ = 1 if i follows the upgrading route in eq 5, 0 otherwise
$UR_L(i)$ = 1 if i follows the upgrading route in eq 4, 0 otherwise

Integer Variables

NSB = number of boilers that produce process steam
NSEB = number of boilers that produce SAGD steam
NHP_j = number of hydrogen plants of type j
NPP_m = number of power plants of type m

Indices

i = SCO production scheme
j = type of hydrogen plant
J = number of hydrogen plant types
k = model of shovel in the fleet
K = number of shovel models available in the fleet
l = model of truck in the fleet
L = number of truck models available in the fleet
m = type of power plant
M = number of power plant types
N = number of SCO production schemes

Model Parameters

CD = cost of the diesel (\$/L)
CFW = cost of the boiler feedwater (\$/tonne)
CO_2E = carbon dioxide emissions target (t/h)
CPCT = compression power for CO_2 transport [kWh (t of CO^2)$^{-1}$ km^{-1}]
CS = percentage of the boiler's capacity used to generate process steam (%)
DB = SAGD bitumen production scheme
DHGO = average density of HGO fraction in hydrotreatment (t/m^3)
d_i = distance from the mining site to extraction plant i (m)
D_k = fuel consumption in shovel model k (L/h)
D_l = diesel consumption in truck model l (L/h)
DLF = average LC-finer feed density (t/bbl)
DLGO = average density of LGO fraction in hydrotreatment (t/m^3)
DNT = average density of NT fraction in hydrotreatment (t/m^3)
DVTB = vacuum topped bitumen density (t/bbl)
EC = energy conversion factor (MJ/GJ)
FC_j = fuel cost for hydrogen plant j (\$/GJ)
FC_m = fuel cost in power plants m (\$/GJ)
FDDC = process fuel requirements for delayed cokers (MJ/bbl)
FDLCF = process fuel requirements in LC-finers (MJ/bbl)
FHV_j = fuel heating value in hydrogen plant j (MJ/m^3)
HHF = hydrogen requirements in high-conversion LC-finers [scf/(t of bitumen)]
HHGO = hydrogen requirements for HGO in hydrotreatment (scf/bbl)
HLF = hydrogen requirements in low-conversion LC-finers [scf/(t of bitumen)]
HLGO = hydrogen requirements for LGO in hydrotreatment (scf/bbl)
HNT = hydrogen requirements for naphtha in hydrotreatment (scf/bbl)

$HPIC_j$ = installed capacity of hydrogen plant j (t/h)
$HPIC_m$ = installed capacity of power plant m (kW)
HR_j = heating rate to produce hydrogen in plant j [MJ/(t of H_2)]
HR_m = heating rate to produce power in plant m (MJ/kWh)
HVNG = Western Canadian Gas heating value (MJ/m^3)
P_{SE} = power requirement in SAGD extraction [kW/(t of bitumen)]
PC_j = power demand in hydrogen plant j [kWh/(t of H_2)]
PCC_m = power capacity cost factor in plant m (\$/kW)
PCC_j = capital cost of hydrogen plant j [(\$ h)/(t of H_2)]
PDDC = electricity requirement in delayed coking (kWh/bbl)
PDFC = power requirement in fluid cokers (kWh/bbl)
PDHF = power requirement in low-conversion LC-finers (kWh/bbl)
PDLF = power requirement in high-conversion LC-finers (kWh/bbl)
PL = pipeline length (km)
PNG = price of natural gas (\$/GJ)
SDRU = steam requirement in the DRU [t of steam/(t of diluted bitumen)]
SFCU = steam requirements in the FCU [t of steam/(t of diluted bitumen)]
SFR = steam requirement in bitumen extraction plant [t of steam/(t of froth)]
SOR = steam-to-oil ratio [t of steam/(t of bitumen)]
$SOSR_C$ = steam-to-oil sand ratio in conditioning [t of steam/(t of oil sand)]
SPF_i = slurry pumping factor in SCO scheme i [kWh (t of slurry)$^{-1}$ m^{-1}]
SVDU = steam requirement in the VDU [t of steam/(t of diluted bitumen)]
t = annual operating hours (h/year)
TDB = total diluted bitumen production (bbl/day)
TSCO = total synthetic crude oil production (bbl/day)
UCF = volumetric conversion factor (m^3/bbl)
UCSC = carbon dioxide underground injection cost [\$/(t of CO_2)]
UCTC = unitary CO_2 transport cost [\$ (t of CO_2)$^{-1}$ km^{-1}]
$WOSR_{BE}$ = water-to-oil sand ratio in bitumen extraction plant [t of water/(t of oil sand)]
$WOSR_C$ = water-to-oil sand ratio for conditioning [t of water/(t of oil sand)]
$WOSR_H$ = water-to-oil sand ratio for hydrotransport [t of water/(t of oil sand)]
ρ_{H_2} = hydrogen density (scf/t)

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