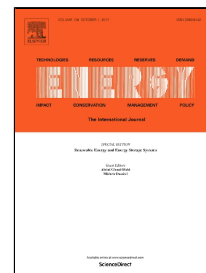# Accepted Manuscript

Design and simulation of a petcoke gasification polygeneration plant integrated with a bitumen extraction and upgrading facility and net energy analysis

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Please cite this article as: E. Lazzaroni, M. Elsholkami, E. Martelli, A. Elkamel, Design and simulation of a petcoke gasification polygeneration plant integrated with a bitumen extraction and upgrading facility and net energy analysis, *Energy* (2017), doi: 10.1016/j.energy.2017.09.072

- A simulation-based modelling of energy demands of oil sands operations is proposed.
- Aspen simulations used to simulate a petcoke polygeneration facility.
- A self-sufficient power, heat and hydrogen facility with a capacity of 190 tonne/h.
- The net efficiency of polygeneration is in the range of 48% - 58%.

Design and simulation of a petcoke gasification polygeneration plant integrated with a bitumen extraction and upgrading facility and net energy analysis

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Abstract – The in-situ extraction of bitumen from oil sands, particularly steam assisted gravity drainage, has been the fastest growing production technology in the industry. Integrated with upgrading operations to enhance the fuel quality, the process consumes significant amounts of energy, which are currently mostly derived from burning natural gas. On the other hand, considerable amounts of petroleum coke residues are generated in the refineries. This petcoke ends up stockpiled as a waste byproduct with associated environmental concerns. The aim of this study is to evaluate the feasibility of integrating a petroleum coke residue gasification plant to the energy infrastructure of an integrated SAGD/upgrading facility. The petcoke gasification process is specifically designed to fulfill the demands of of a facility processing 112,500 barrels per day of Athabasca bitumen. Two plant configurations are compared, one without and one with CO_2 capture and storage. The gasification-based polygeneration plant is modeled with the Aspen Plus flowsheeting software. Two levels of energy demands (i.e. high and low energy scenarios), reflecting the range of variability in the energy requirements of extraction and upgrading operations (e.g. steam to oil ratio), are considered. The net efficiency for polygeneration plant was determined to be in the range of 48 – 58%. The gasification of approximately 190 t/h of petroleum coke is required to achieve the power, thermal and hydrogen demands. The incorporation of carbon capture imposes significant energy penalties, which requires the addition of natural gas fueled gas turbines to meet the power requirements.

Keywords – Oil sands, Hydrogen production, Gasification, Polygeneration, Petroleum coke

1. Introduction

The demand for oil is continuing to increase worldwide and is expected to reach 111 million barrels per day by 2040 [1]. The production of unconventional oil has increased more than twofold over the past decade, and will almost double by 2020 accounting for 11% of total oil production [2]. Heavy oil deposits exist worldwide, one of the largest being the Athabasca Oil Sands in Alberta, which contribute to Canada's total proven oil reserves to be the third largest in the world [3]. Currently heavy oil production accounts for one third of Canada's total oil production, and this share is expected to reach approximately 75% by 2030 [4].

The two prominent production technologies used to extract bitumen from oil sands are mining and in-situ operations. Approximately 80% of bitumen is only recoverable using in-situ production methods, which are currently experiencing the highest growth in the industry [5]. In-situ methods rely on the use of steam, solvents or thermal energy for bitumen extraction, which enhance its flow allowing it to be pumped to the surface. The in-situ production technology mostly applied in major in-situ projects is the steam assisted gravity drainage (SAGD), in which large quantities of steam are injected into the reservoir to reduce the viscosity of bitumen. The produced bitumen is then typically diluted and sold, used as a fuel for energy production on-site, or upgraded to a form suitable to direct processing in refineries. Primary upgrading routes typically consist of thermal cracking, hydrocracking processes or a combination of both, which are used in order to decompose large hydrocarbon molecules in heavy bitumen into smaller ones.

Studies have been done in the past in order to quantify the energy demands of oil sands operations. Ordorica-Garcia et al. [6] modeled the energy demands of major oil sands operations based on data provided by commercial producers. The mathematical model was used to quantify power, thermal, hydrogen, fuel requirements, and greenhouse gas (GHG) emissions associated with mining and in-situ bitumen extraction operations, and thermal- and hydro-cracking upgrading routes. Charpentier et al. [7] developed a life cycle based-model in order to quantify the GHG intensities of various process configurations based on confidential operating data of oil sands producers and their energy requirements. In these studies it was assumed that a natural gas based energy infrastructure will supply the energy requirements of the industry. Based on the results they obtained it was concluded that the utilization of natural gas is currently an economical and clean approach to provide energy for the industry; however, the planned increase in production of oil sands will offset achieved reductions in emissions [8].

Significant amounts of natural gas are utilized in the oil sands industry to produce steam, hydrogen and electricity to support the energy requirements of its operations. In 2014 oil sands operators consumed approximately 60,000,000 m^3/day of natural gas (approximately 26 Nm^3/bbl of bitumen), which is expected to considerably increase considering the major future expansions in the industry [9]. Recently there have been rising concerns regarding the availability of natural gas to support the continuously increasing energy requirements of the oil sands industry [10]. Natural gas price volatility has been another factor that is encouraging energy producers, including oil sands operators, to replace the highly consumed commodity with an alternative fuel [11]. The utilization of natural gas for energy production entails various advantages compared to the use of other fossil fuels (e.g. efficient production, lower greenhouse gas emissions, etc.) resulting in its contribution to a significant share of primary energy supply in the industrial sector. Therefore, changes in a natural gas-based energy infrastructure will involve significant techno-economic and environmental consequences, which should be thoroughly assessed prior to a course of action addressing its replacement [12].

Various fossil fuel replacement options can be potentially used in the industry, such as coal, asphaltenes, petroleum coke, which vary in their energy contents and environmental impacts requiring different production technologies. Ordorica-Garcia et al. [13] and Betancourt et al. [14] have used mathematical optimization techniques to determine the infrastructure required to

satisfy the energy demands of oil sands operations. They considered various fossil-fuel based alternatives for the energy commodity producers. Other energy production alternatives have lower environmental impacts associated with their utilization, including renewable sources, such as biomass and geothermal energy [15, 16]. However, the dependency on fossil fuels is expected to prevail until cleaner energy alternatives can potentially satisfy the increasing energy demands while competing with the relatively cheaper, accessible and reliable conventional resources [12].

Canadian petcoke produced at upgraders in Alberta and Saskatchewan has currently reached annual levels of at least 10 Mt, and most of it ends up stockpiled as a waste byproduct in limited inventory due to the significant costs associated with transporting it out of the region. Petcoke stockpiles in Alberta has currently reached levels of 80 Mt and is growing at a rate of approximately 4 Mt/yr [17, 18]. Even though the byproduct is a burden for many oil sands produces, it has significant potential in supporting the increasing energy requirements of bitumen production and upgrading operations. The stockpiles of petcoke in Alberta's upgraders represent a unique energy reserve if untapped using gasification processes, owing to its favorable properties, such as high heating value, low ash content, adequate particle size, etc [19]. Despite its high heating value, oil sands petcoke is characterized by having high sulphur content and low reactivity (i.e. low composition of volatile matter) making it a less suitable fuel for other combustion based processes (e.g. boilers, furnaces, etc.) [20].

Various energy forms that are highly demanded by oil sands operations can be produced by the utilization of petcoke as a feed for co-generation gasification plants, which include power, steam and hydrogen. Several studies that focus on assessing the feasibility of the commercialization of oil sands petcoke gasification processes have been conducted addressing the utilization options of the potentially valuable energy resource. These include studies utilizing oil sands petcoke on a laboratory and pilot-plant to commercial scale in which its properties are investigated under various operating conditions [17]. The limitation of the use of the petcoke on a commercial range is associated with the cost of co-generation energy production facilities, as well as the maturing of carbon capture and sequestration technologies [20].

Studies have been done in the past to investigate the suitability of oil sands coke gasification as an alternative fuel to natural gas, which incorporated both economic and environmental aspects. Vartivarian et al. [21] determined that the gasification of petcoke for the production hydrogen is an economically attractive alternative at high natural gas prices. They also concluded that it has several environmental advantages as it facilitates the capture of sulphur dioxide and carbon dioxide emissions [22, 23, 24]. Recent studies addressed the feasibility of integrating oil sands petcoke gasification technologies with oil sands operations. Based on a life cycle framework, McKellar et al. [12] indicated that the gasification of coke byproducts will considerably offset recently achieved greenhouse gas emission reductions in the industry. Using Aspen Hysys simulations, El Gemayel et al. [24] investigated the feasibility of incorporating an integrated gasification combined cycle utilizing oil sands petcoke as a feed stock in bitumen upgrading facilities. The aim of the study was to generate a process that is self-sufficient for power, hydrogen and steam requirements while meeting carbon dioxide emission constraints using carbon capture technologies.

A class of methods that is commonly used in comparing alternative fuels is the net energy analysis (NEA), which provide a quantification of the effectiveness of an energy system [25]. Net energy is defined as the energy yield of a certain fuel minus the energy inputs required to produce the particular fuel [26]. The net energy of a certain resource is the adequate measure representing the true value of a fully developed resource to society [27]. NEA methods are based on life cycle concepts and are generally acceptable in assessing the potential of energy systems. Several studies in the literature have utilized these methods to conduct comparative analyses of various types of energy systems (e.g. fossil fuel based) [28, 29, 30, 31, 32, 33]. In this scenario NEA can provide a tool in assessing the viability of an energy system based on the gasification of petcoke and comparing its performance to that of the NG based infrastructure currently utilized for oil sands operations. A commonly used parameter for representing the energy return of a process is the net energy ratio (NER), which is defined as the ratio of the net output of primary energy to the sum of external primary energy and self-consumption of primary energy [34]. The net external energy ratio (NEER) considers only the inputs that are consumed from the existing industrial energy system, excluding any self-use (e.g. produced oil burned on site to power oil producing operations), and it is defined as the ratio of the net output of primary energy to the input external primary energy. The NEER can measure the potential growth in energy supply to society because it only counts those inputs that must be produced and delivered externally from and to the process through the existing energy supply system. This comparison indicates to what extent a certain process is energy independent.

The gasification of petcoke to produce various energy forms (i.e. power, heat and hydrogen) represents an attractive alternative to the natural gas-based energy infrastructure that provides the energy commodities required for oil sands operations. However, this replacement will entail techno-economic and environmental consequences that must be thoroughly investigated. Therefore, this study aims to contribute to evaluating the feasibility of integrating a petcoke gasification plant into the energy infrastructure of a SAGD/upgrading process. The technical feasibility of the petcoke based polygeneration plant will be investigated using Aspen Plus simulations [35]. To the authors' knowledge, the petcoke gasification simulation model proposed in this work is the first that accounts for satisfying the power, thermal and hydrogen demands of an integrated SAGD/upgrading process. Combining the gasification plant with the SAGD/upgrading process facilitates achieving a power, hydrogen and thermal self-sufficient facility with GHG mitigation measures (i.e. CO_2 capture), which provides an incentive to reduce reliance on imports of external fossil fuels (e.g. natural gas). A net energy analysis of the proposed system is conducted and compared to that of an energy system utilizing NG based technologies that represents the current energy infrastructure supplying oil sands operators. In this study the trade-offs between GHG emissions and energy use associated with substituting oil sands petcoke for natural gas will also be examined, and the results will provide an insight of these trade-offs for oil sands producers and government decision makers concerned with addressing GHG emissions regulations.

2. Current and proposed energy infrastructure – Case study

The case study investigated in this work is based on an integrated SAGD/upgrading plant processing 112.5 thousand barrels per day (TBD) [36]. The diluted bitumen produced from SAGD extraction is sent to an upgrading facility to produce synthetic crude oil (SCO) at a rate of 150 TBD. The energy forms consumed in the bitumen extraction and upgrading operations are power, hydrogen, and thermal (steam and heat). The bitumen upgrading configuration considered in this case study consists of atmospheric and vacuum distillation units (ADU and VDU) for diluent recovery and diesel production, delayed coking (DC) of vacuum residue and hydrotreating (HTR) of liquid fractions. The diluent and portion of the liquid fractions of bitumen are separated in the ADU, and the VDU maximizes their recovery. The thermal cracking of the remaining heavy ends takes place in the DC unit from which vapor, liquid and solid fractions are produced. The produced liquid fractions are hydrotreated to produce sweet SCO.

The majority of energy consumed in SAGD extraction operations is in the form of high pressure steam, which is injected to the well in order to mobilize the bitumen. In SAGD extraction operations electricity is consumed to operate bottom hole pumps, surface circulating pumps and the compressors of the vapor recovery units. The amounts of steam and electricity required for SAGD extraction are proportional to the rate of bitumen production through the steam- and electricity-to-oil ratios, respectively, which are defined based on operational data of existing facilities [6, 37, 38, 39, 40]. During bitumen recovery there is a significant amount of gas produced from the dissolution of non-condensable components, which can be utilized as a fuel for energy production.

Steam (20 bar / 400 – 510°C) used for bitumen upgrading is consumed in distillation columns to improve distillation efficiency, and the DC to avoid coke deposition into the furnace. The electricity requirements of the ADU, VDU and DC are relatively lower compared to that of HTRs. The lowest electricity consumption takes place in the ADU as it is only limited to fluid pumping, while higher electricity requirements are needed for the VDU in order to support the vacuum conditions. The DC requires electricity to operate the hydraulic decoking pump. The high electricity consumption associated with the operation of the HTRs is attributed to the high compression requirements for the hydrogen make up and recycle streams. The high sulphur, nitrogen and aromatic content of bitumen imposes the significant hydrogen requirements for hydrotreating operations.

The energy requirements for the SAGD extraction and bitumen upgrading operations have been modeled by Lazzaroni et al. [36], which are considered as the basis for energy infrastructure developed in this study (Table 1). The authors quantified the energy requirements of upgrading operations using Aspen HYSYS simulations, which provide an adequate estimation of the energy consumed by process units and their associated CO_2 emissions. Two levels of demand have been considered for the energy commodities considered in this case study, which are referred to as the high energy scenario (HES) and the low energy scenario (LES). The upper and lower limits of the energy consumption factors associated with process units were used to define the conditions

of the HES and LES, respectively. The breakdown of energy consumption in upgrading operations is shown in Figure 1.

The current infrastructure (Table 2) that is used to supply the energy requirement for oil sands operations is mainly utilizing natural gas as a fuel. Considered as a clean and easily accessible option, natural gas was selected in order to reduce the GHG emissions associated with the industry. However, due to the challenges (e.g. increasing prices, reduced availability, etc.) associated with the continued reliance on its utilization, oil sands operators are becoming increasingly interested in exploring other available alternatives. The ease of accessibility, high energy content, and reduction of inventory in limited storage associated with the utilization of stockpiled petcoke present an attractive alternative for oil sands operators.

Therefore, in this study an energy infrastructure based on the utilization of petcoke and supplementary gas as a feedstock is proposed. The objective is to assess the feasibility of the system in satisfying the energy requirements of the integrated SAGD/upgrading facility presented in this case study. The entire energy system proposed in Table 2 is modeled using the Aspen Plus simulation software. The performance of the simulated energy system will be compared to that of currently used technologies utilizing NG (Base case). The performance of the base case energy infrastructure (Figure 2) satisfying the demands of the SAGD/upgrading facility considered in this work has been modeled in a previous study by Lazzaroni et al. [36]. The results they obtained are summarized in the following lines, which will be compared to the performance of the proposed energy system in order to illustrate the trade-offs associated with using petcoke as a fuel alternative to NG.

From the data in Table 3 it can be observed that significant amounts of NG are consumed in the LES and HES in order to satisfy the power, steam and hydrogen demand of the SAGD extraction and bitumen upgrading operations. The NG consumption in the HES is 47.6 % higher than that of the LES, which is mainly associated with the higher steam production from the OTSGs. The NEER parameter is defined as the ratio of the total energy output (SCO and produced gas) to the external energy input (natural gas). In other words, the ratio indicates the total worth of energy produced for every unit of energy input to the system. The NEER values associated with the current practice of providing energy (NG based technologies) for oil sands production are considerably low in comparison to those associated with the production of conventional oil resources (15 – 30 GJ/GJ) [41], which provides an indication of the current energetic disadvantage of SCO production in comparison to conventional oil resources.

Figure 3 provides an indication of the CO_2 emissions specific to each energy production unit. The CO_2 emissions associated with the NG cogeneration facility are mainly attributed to the cogeneration gas turbines with postfiring. The OTSGs are the largest contributor to CO_2 emissions accounting to approximately 50% of total emissions, and are the major producers of emissions in SAGD extraction operations. The majority of CO_2 emissions associated with upgrading are associated with the operation of furnaces, followed by the reformers and steam generators.

3. Process description and simulation

The overall process shown in Figure 4 presents the gasification process investigated in this study. The proposed polygeneration process can be integrated with oil sands SAGD production and/or bitumen upgrading facilities in order to fulfill their power, steam and hydrogen requirements. The petcoke residue from the upgrading section and supplemental natural gas are used as a feed for the gasifier in which a synthetic gas (syngas) is produced that is mainly composed of carbon monoxide and hydrogen. The syngas is cooled and treated to eliminate sulfur species (i.e. H_2S and COS). The resulting sweet stream then passes through water-gas shift reactors in which carbon monoxide reacts with water to produce carbon dioxide and more hydrogen. The syngas can be used as a fuel for gas turbines, and the heat from exhaust gases can be recovered using once through heat recovery steam generators (OT-HRSGs), from which the cogeneration of electricity and thermal energy can be achieved. Steam can also be produced through heat recovered by syngas coolers, and OTSGs fueled by sweet produced gas, natural gas and fuel gas. The steam and power produced supply the electrical and thermal requirements of the gasification process, upgrading facility and SAGD operations. Heat requirements of the upgrading facility (e.g. coking and distillation units) are also supplied by furnaces utilizing produced gas. Carbon capture is utilized in order to produce a hydrogen rich stream that can be recycled to hydrotreaters in the upgrading facility. For the scenario in which carbon capture is considered an additional natural gas cogeneration plant is incorporated in the model in order to satisfy the additional power requirements.

The process simulation (Figure 5) was conducted using Aspen Plus [35], which is equipped with a variety of property packages and contains thousands of pure components allowing the user to define hypothetical compounds for which various properties can be estimated using embedded thermodynamic correlations. The simulation of the gasification process was conducted using the Peng-Robinson thermodynamic model.

3.1. Gasification model

The selection of the type of gasifier depends on the petcoke feed characteristics. Petcoke has a graphite-like structure that results from the prolonged exposure to high temperatures. In comparison to coal, which is a common feedstock to various commercial gasification units, petcoke has a higher level of fixed carbon (80 – 90%), lower levels of ash (3 – 8%), higher sulfur content (5 – 7%), and negligible moisture content. The properties of petcoke used in this study are summarized in Table 4 [17], which are based on the ultimate and proximate analyses of petcoke produced by the delayed coking unit in the upgrading facility of Suncor. The high content of fixed carbon and the low content of ash raise the lower heating value of the petcoke to values similar to the highest rank anthracites. The selection of the gasification technology also depends on the particle size distribution of the fuel injected into the gasifier. Particle size distribution will affect the flow and pressure drop in the gasifier, and the reactivity of the fuel.

The particle size distribution of petcoke produced from Suncor's delayed coker is summarized in Table 4 [17].

The gasification technology that is suitable and most commonly used for bitumen derived petcoke is the entrained bed gasifier [17, 42]. Considering the characteristics of petcoke considered in this study, there are several limitations associated with the use of fluidized and moving bed gasifiers, which include: 1) The considerable amount of fine particles is unsuitable for the type of flow occurring in these gasifiers, 2) The low reactivity of petcoke requires considerably high operating temperatures, 3) due to the long residence time which limits the capacity (kW_{th}/m^3) of these reactors, a huge reactor volume would be necessary, 4) if air is used as oxidizing agent, the high sulfur content of petcoke will require significant amount of limestone used for insitu sulfur capture in fluidized bed gasifiers. A penalty of the benefits associated with entrained flow gasifiers, which include fuel flexibility, tar and oils-free syngas production, high carbon conversion, low methane production, and high throughput, is the relatively high oxygen consumption required. This relatively increases the requirement of pure oxygen for the gasification reactions, which requires the additional installation of cryogenic air separation units that has considerably high capital and operating costs associated with their operation.

The gasification technology selected for this study is the Shell/Prenflo gasifier, for which Suncor's petcoke is considered to be an ideal feedstock [43]. The technology consists of a dry feed upflow reactor with a membrane wall vessel. The produced syngas is cooled to a temperature range of 250 – 280°C, from which heat is recovered to produce steam. A portion of the cooler syngas (up to 50%) is recycled as a syngas quench stream. A variation of this process involves the utilization of a water quench stream.

The petcoke stream was simulated as a non-conventional stream and its properties were defined using HCOALGEN property method. The input requirements for this model include the proximate, ultimate and sulfur analyses, as well as the fuel heating value, which are all summarized in Table 4. It is necessary to convert the non-conventional stream into a conventional stream, which is achieved using the RYield model by specifying reaction yields of the individual components. The low moisture content of the fuel does not impose the requirement of a drying unit. The low ash content of petcoke requires the addition of ash to the petcoke feed stream before it enters the gasification reactor. This is essential to adjust for the required working conditions of the membrane walls and refractory liners of slagging gasifiers [42]. It has been assumed that setting the amount of ash added to the feed stream by recycling the flyslag separated by ceramic filters is sufficient to correct the operating conditions to those required for the membrane wall. Another option is to mix the petcoke with a stream of coal with high ash content.

The gasifier has been modeled with an RGibbs reactor with a feed of petcoke, steam, oxygen and nitrogen, which are the components required for the gasification reactions to take place (Table 5). The membrane wall of the gasifier is simulated as a steam generator producing medium

pressure steam. A separator is used to isolate the unconverted carbon from the coarse slag, which simulates what occurs at the bottom of the reactor. After the separator a mixer is added to represent the syngas quench. Part of the cooled syngas is recycled by a blower, and the amount of syngas used in the recycle stream is regulated by fixing the quench temperature to 900°C. The parameters that have the highest effect on the quality of syngas produced are the flowrates of oxygen and steam, and the outlet temperature of the gasifier. The quality of syngas produced can be measured using the cold gas efficiency (CGE), which is defined as the ratio of the energy content of syngas to that of the fuel used as a feed for the gasifier. The optimization model analysis tool in Aspen Plus was used to maximize the CGE by varying these parameters and constraining heat losses to the membrane wall, as well as the CO_2 and CH_4 content in the produced syngas. The value of the CGE obtained is slightly lower than the values obtained with commercial coal (approximately 80%), which is primarily attributed to the quality of the fuel. The syngas obtained from the gasification of petcoke has a considerably higher carbon monoxide to hydrogen ratio compared to that obtained from conventional coal (typically 2:1).

3.2. Syngas treatment model

For the gasification based polygeneration plant two scenarios were simulated, the first involves the venting (VS) of the produced carbon dioxide to the atmosphere, and the other incorporates a carbon capture and sequestration (CS) technology in which the carbon dioxide contained in the syngas is separated, compressed and stored in extinguished reservoirs. The main processes included in the syngas treatment section for the VS are wet scrubbing, COS hydrolysis and acid gas removal (Table 6). The syngas generated is then split between the gas turbines for cogeneration, and further treatment for pure hydrogen production. The production of a hydrogen rich stream is achieved through the use of a water gas shift model and a pressure swing adsorber (PSA).

3.2.1. Wet scrubbing unit

The wet scrubbing unit removes solid particulates remaining after the candle filter. The scrubbing takes place below the dew point of the gas. The finest particles act as nuclei for condensation, thus ensuring the complete removal of solids. The solid removal aspect of wet scrubbing is difficult to simulate in Aspen Plus. However, it is important to take account of this unit due to the large amount of water used in it, which has a considerable effect on the syngas temperature and composition. The scrubber is simulated as a RadFrac column, which incorporates a rigorous model for simulating all types of multistage vapor-liquid fractionation operations. The L/G (liquid over gas mass flow rate) for scrubbing was set equal to the flowrate of syngas at a temperature of 166 °C. The model used in the RadFrac column is based on Henry's law, which is not supported by the Peng-Robinson property method. Therefore, the syngas stream was switched to the ELECNRTL property method. The water outlet of the scrubber is purified from the acid components and recycled to the scrubber. Since some of the water evaporates into the syngas, a water make-up stream is required. The recycled and make-up streams are reheated to the inlet temperature before entering the column. The heating requirement is recovered from the syngas cooling process occurring after the water gas shift process used for hydrogen production.

3.2.2. COS hydrolysis, water gas shift reactor and PSA

Sulphur species in the syngas produced include COS and H2S. In the CO_2 venting case, to complete the desulphurization process, the COS should be converted into H2S before the acid gas removal stage. The COS can be selectively converted into H_2S through a catalytic hydrolysis reaction [42, 43]. The hydrolysis reactor has been modeled as an RGibbs reactor with restricted equilibrium where only the COS hydrolysis reaction is allowed.

An adiabatic RGibbs block is used to simulate the water gas shift (WGS) reactor. Typically two shift reactors are used: The first one is operating with high temperature catalysts, while the second operates with low temperature catalysts. The first reactor is operating with iron oxide–chromium oxide catalysts with a usual inlet temperature of 300-450 °C. The second reactor uses Cu-ZnO-Al$_2$O$_3$ catalysts that usually operates in a range of 190-250 °C [36]. The ©inlet temperatures for HTS and LTS reactors are respectively 400 and 200 °C. All the components were considered inert except the ones involved in the reaction. An adiabatic pre-reformer unit is also included in the process, in which hydrocarbon molecules react with water to produce carbon monoxide and hydrogen. After the WGS reactor the syngas is cooled and sent to the purification units.

The hydrogen in the syngas should be separated from the other components in order to obtain the desired purity, which is achieved in the PSA unit. Since modelling the PSA process is extremely challenging and out of the scope of this work, the PSA has been modelled as a simple separator block with performance measures taken from literature data for industrially available PSA systems [44]. More in detail, the H_2 recovery rate is assumed to be equal to 90%. The hydrogen produced is sent to the hydrotreaters while the off-gas of the PSA, which is mainly composed of CO_2, H_2, CH_4, CO and H_2O, is sent to the burners of the steam reformer reactor. The steam reformer burners use the off-gas from the PSA unit as a fuel, as well as some of the produced gas from the upgrading process. The burners were modeled with an adiabatic RStoic unit in which the combustion reactions occur. The exhaust gases are then cooled, and the heat is recovered for steam production. The heat for steam production is also provided from the syngas coolers after the reforming reactors and the WGS reactors. The heat at high temperature is used to produce steam while the heat at low temperature is used for water preheating. The steam is produced at 38.5 bar and 500°C.

The main processes involved in the syngas treatment for the carbon capture scenario are the same as considered for the venting scenario. However, the acid gas removal step involves the elimination of carbon dioxide. In the carbon capture scenario all the syngas produced is sent to the WGS, which requires a high amount of medium pressure superheated steam that is produced in the polygeneration plant. The amount of steam required is proportional to the syngas processed, therefore, the amount of steam required in the CS is considerably higher than that needed for the VS. The WGS process takes place before the acid gas removal stage, which imposes the requirement of using a sulfur-tolerant cobalt-molybdenum alumina catalysts. The carbon capture scenario does not require a COS hydrolysis unit. This is because the COS

hydrolysis takes place in the WGS reactors, where the abundance of water and the thermodynamic conditions meet the requirements of the reaction. The WGS converts most of the CO in the syngas to CO_2, resulting in the production a of H_2 rich stream (up to 95%) after the CO_2 removal stage. The characteristics of the syngas produced in the two scenarios are completely different. The syngas produced in the VS is mainly composed of CO (approximately 75%) and has a considerably lower heating value compared to that produced from the CS.

3.2.3. ASU, Acid Gas Removal and CO_2 compressor

Given the complexity of the involved physical/chemical phenomena and the availability of literature data, the air separation unit (ASU), the acid gas removal (AGR) process and the CO_2 compressor were not modelled in detail but their performance (energy consumption and separation efficiency) were taken from the work by Martelli et al. [43] who considered a similar Shell-based IGCC with/without CO_2 capture and storage, with a net electric production capacity within the range of 270 MW – 315 MW. Without loss of accuracy, the power consumption of the ASU (providing 95% purity O_2) is assumed to be proportional to the mass flow rate of produced oxygen (2.482 MJ/kg O_2). Similarly, the petcoke handling/milling consumption is proportional to the amount of petcoke used as a feedstock for the gasifier.

In the venting case, the SULFINOL-M process is considered for H_2S removal, and the same separation efficiencies and specific energy consumption (heat for the reboiler and electricity per mole of H_2S) used in [43] are assumed. Also for the capture case, the same capture process (i.e., SELEXOL), CO_2 compressor configuration and performance assumptions (heat for the reboiler and electricity consumption per mole of CO_2) used in [43] are assumed.

3.3.   Cogeneration model

Steam is generated by heat recovery from the membrane wall of the gasifier, syngas coolers, and shift reactor coolers (High and low temperature). Steam requirements for the gasification plant include medium pressure (42 bar) steam used for the gasification and water gas shift reactions, as well as high pressure (100 bar) steam used for the stripping column in the acid gas removal section. The low temperature shift reactor coolers preheat the water for medium pressure steam generation. They also produce hot water for the scrubber and low pressure (20 bar) steam required for the acid gas removal section. The generation of medium pressure steam is mainly from the high temperature shift reactor coolers, the syngas coolers, and the gasifier's membrane wall. The production of low pressure steam required for the upgrading section is from heat recovered by the syngas coolers. The superheating of the medium pressure steam takes place at the entrance of the OT-HRSG. The final stream of syngas produced is sent to the cogeneration model. The composition of the syngas produced from the VS and CS is summarized in Table 7.

The gas turbine used for the cogeneration model is the 47.5 MW Siemens SGT-800 [44]. The most important concern is the mitigation of NOx emissions, which becomes critical due to the very high H_2 flame temperature. Two different methods can be used to control NO_X emissions from gas turbines, which are premixed lean combustor and diffusive flame dilution with inert species (steam, water or nitrogen). The second technology is mature and fully developed [45].

By using a saturator and/or the addition of the nitrogen available from the ASU dilution of the syngas can be achieved, which leads to a lower flame temperature and lower NO_x emissions. However, the addition of inert gases for diffusive combustion imposes a penalty associated with efficiency reduction, as shown [45]. Although premixed lean combustors are not yet available in the market due to the challenges posed by the high reactivity of the hydrogen and the difficulty in designing premixed combustors with acceptable pressure drops [45, 46], successful pilot scale tests are reported in the literature and full scale combustors will soon be available [45]. Therefore, it was assumed that the syngas-fueled SIEMENS SGT-800 gas turbine is equipped with a premixed Dry Low NOx combustor. The performance of the SGT-800 gas turbine was simulated with the software GT-PRO [47], capable of accounting for the off-design behavior of the machine. The composition of the syngas (Table 7) has been used as a fuel input for the GT-PRO simulation. The postfiring system has been modeled with another RStoic reactor and the excess of air used in the gas turbine supports the occurrence of another combustion process. The limit to the amount of fuel injected for the postfiring is the maximum allowed temperature of the heat recovery steam generator case, estimated to be about 750 °C. Since the amount of syngas allocated for the postfiring is not sufficient to reach the desired temperature, a small amount of supplemental natural gas is used. Due to the impurities in the water, which is coming from the CPF, is not possible to use drum evaporators. In order to keep the liquid in contact to the pipes, it is mandatory to maintain low steam quality. If a complete evaporation occurs, the harnesses in the water could deposit over the surface of the pipes, creating fouling and several problems related. The water and the steam produced by the OT-HRSG are then separated, and the steam is sent to the wells while the water is used for preheating the boiling feed water.

4. Results and discussion

The performances measurements of the gasification based polygeneration plant are summarized in Table 8. The auxiliary power (i.e. difference between gross and net power) consumed by the gasification plant is considerably high, which is mainly due to the air separation unit power requirements (VS: >85%, CS: >65% of auxiliary power). The power consumption of AGR, sulphur recovery and petcoke handling account for less than 20% of total auxiliary power. A significant amount of power (approximately 20% of auxiliary power) is required in the carbon CS for the compression of the produced CO_2 stream, as well as for the AGR and sulphur recovery facility. Therefore, the net electric power generated is substantially lower in the CS compared to the VS. In the CS the net electricity output turns out to be too low to satisfy the total electricity demand, even for the low energy scenario, which as a result requires the installation of additional natural gas-fueled gas turbines.

The gasification based polygeneration plant results have a value of efficiency (defined as the energy content of the produced commodities, which are hydrogen, steam and power divided by the energy content of the petcoke fuel) of 58% for the VS and 48.26% for CS, which are considered to be in the higher range of typical efficiencies of gasification plants [48]. The net efficiency is calculated based on the following equation:

$$\eta_{net} = \frac{\dot{m}_{H2} * LHV_{H2} + \dot{m}_{LPS} * \Delta H_{MPS} + \dot{m}_{MPS} * \Delta H_{MPS} + \dot{m}_{HPS} * \Delta H_{HPS} + PWR_{net}}{\dot{m}_{NG} * LHV_{NG} + \dot{m}_{PCK} * LHV_{PCK}} \qquad (1)$$

where $\dot{m}_{H2}$, $\dot{m}_{LPS}$, $\dot{m}_{MPS}$, $\dot{m}_{HPS}$, $\dot{m}_{PCK}$ and $\dot{m}_{NG}$ are the mass flow rates of hydrogen, low-pressure steam, medium-pressure steam, high-pressure steam, gasified petcoke, and natural gas, respectively. LHV_{H2} is the lower heating value of hydrogen. ΔH_{LPS} is the specific variation of enthalpy of the LP steam produced at 20 bar. ΔH_{MPS} is the specific variation of enthalpy of the LP steam produced (at 42 bar and 500 °C). ΔH_{HPS} is the specific variation of enthalpy of the HP steam produced at 100 bar. LHV_{NG} is the lower heating value of natural gas, and PWR_{net} is the net electricity produced excluding electricity utilized for auxiliaries consumption. The detailed components of the net efficiency are summarized in Table 9.

The polygeneration plant also offers very good performances in the oil sands energy infrastructure. In both the VS and CS, the gasification based polygeneration plant produce all the hydrogen and steam required for the upgrading operations. In the venting scenario, all the electricity for the processes can be produced with syngas-fueled gas turbines. However, in the CS additional NG-fueled gas turbines are required in order to produce all the electricity demand for the SAGD extraction and upgrading processes. The polygeneration plant is capable of producing up to approximately 40% of the total steam requirements of the SAGD extraction facility. The remaining steam is produced by OTSGs that mostly utilize produced gas as a fuel. As a result, the dependence on natural gas is reduced from 74% in the base case to 19-23% in the gasification case.

The gasification plant can achieve a steam/power ratio of 3.09 and 6.67 for the venting and the capture scenario, respectively. These values of the steam/power ratio compared to the value of 1.65 of NG cogenerations systems (with postfiring) are closer to the demand requirements of the integrated SAGD/upgrading facility (13.6-13.99 MW_steam/MW_power). The MP steam demands of the WGS are higher in the CS scenario than in the VS, which reduces the capacity available for the production of HP steam required for SAGD extraction. The venting scenario had satisfactory output of both net power (158.42 MW) and steam (616.2 MW). The net efficiency is adequate for a plant with a considerable energy penalty factor associated with carbon capture. Despite the high efficiency of the carbon dioxide venting scenario there is a significant amount of CO_2 emissions associated with its operation.

The gasification based polygeneration plant is simulated in order to fully satisfy the energy demand of the integrated SAGD/upgrading facility defined in Table 1. The developed model was simulated for the LES and HES with and without the consideration of carbon capture and sequestration. After obtaining the energy stream for the various investigated scenarios, a net energy analysis was then conducted and the CO_2 emissions were calculated. The results obtained were then compared to those obtained for the base case energy infrastructure.

In the gasification model developed the fuels utilized include petcoke, natural gas, produced gas (PG) from wells, and fuel gas (FG) from the DC unit. Specific fuel consumption was related to each unit in the energy system. The attribution of fuel consumption to the gasification reactor was based on dividing the energy streams into two groups, which are electricity and steam and hydrogen production. The natural gas consumption for electricity generation is the syngas stream used in the simple cycle of the gas turbine. The steam generation and hydrogen production is proportional to the syngas processed in the hydrogen production process. The offgas produced in this process is used as a fuel for steam generation.

The results obtained for the LES and HES with and without the consideration of carbon capture are shown in Figure 6. It can be observed that the cogeneration system accounts for the highest consumption of energy, which mostly utilize petcoke as a fuel. A small amount of supplementary NG is used as a fuel in the gasification system; however, the noticeable increase in the CS is due to the energy requirements of the additional natural gas-fueled turbines required in the cogeneration system. The highest amount of natural gas consumption is due to the OTSGs required to fulfill the HP steam requirements of the SAGD extraction operations, which results in a total increase of energy consumption of approximately 35% between the LES and the HES. Even though a significant amount of natural gas was still consumed in the energy system, it is still considerably lower than the total consumption of natural gas in the base case energy infrastructure. In order to have an adequate representation of the comparison between the base case and the proposed energy infrastructure proposed in this work, a net energy analysis was conducted. The breakdown of fuel consumption in the cogeneration system is shown in Figure 7.

The net energy analysis was based on evaluating the NER and NEER parameters, which are used to represent the total energy return from the proposed energy infrastructure. The NER considers all input and output energy streams in its evaluation, as shown in Eq. (2):

$$NER = \frac{Net\ Output\ of\ Primary\ Energy}{External\ Primary\ Energy + Self\ Consumption\ Primary\ Energy} \tag{2}$$

The input streams involved in the evaluation of the NEER consider only the external input energy (i.e. natural gas), as indicated in Eq. (3):

$$NEER = \frac{Net\ Output\ of\ Primary\ Energy}{External\ Primary\ Energy} \tag{3}$$

This provides an indication of the contribution of each individual fuel to the performance of the energy system. The NER and NEER values obtained for the proposed energy infrastructure are summarized in Table 10. The NER value for the HES is lower than that obtained for the LES, similarly the values obtained for the CS are lower than those obtained for the VS. This is a consequence of the higher energy requirements of the HES and CS. The values of NEER are very high in both cases. In the VS, the values of NEER are close to those obtained for the production conventional crude oil (NEER = 15-30) [41]. For every MJ of external fuel input for the production of SCO, 9.19-26 MJ of energy is produced. The values of NEER of the CS are

lower than those obtained for the VS, which is a result of the additional natural gas-fueled turbines used in the cogeneration system to satisfy the additional power requirements.

The energy consumption associated with the incorporation of the petcoke gasification plant is higher than the scenario in which natural gas was the only fuel considered for energy production, which was determined to be higher by 22.39% and 28.49% for the VS and CS, respectively. This is mainly attributed to the considerably lower energy production efficiencies associated with the utilization of petcoke as a fuel for energy production in comparison with natural gas based technologies. However, it is important to note that this resource is typically stockpiled as it is considered to be a waste byproduct by oil sands operators. Therefore, its utilization on a large scale for energy production with an adequate efficiency represents an attractive alternative to the reliance on the increasingly expensive cleaner fossil fuels (i.e. natural gas). In the proposed energy system most of the bitumen components are utilized for energy production. This efficient conversion of the bitumen is more sustainable in the long term than the large input of external natural gas.

This was evident from the results obtained as the share of natural gas consumed, which was less than 25% in all the scenarios involving the utilization of petcoke as the main fuel for energy production. This can be better illustrated by the comparison of the NEER value associated with the utilization of petcoke to those of the base case energy infrastructure. The NEER obtained for the gasification of petcoke are in the range of 12.6 – 17.75, which are considerably higher than those associated with the base case scenario. The base case resulted in low values of energy return ratios, which were determined to be 3.02-3.9 and 3.85-5.42 for the NER and the NEER, respectively. This is because the petcoke, PG and FG, which are the major sources of energy in the new proposed energy system are extracted from bitumen. Therefore, they are not considered as external energy sources; however, the base case energy infrastructure relies mostly on purchased natural gas.

Despite the considerable benefits of the reduced reliance on natural gas, the CO_2 emissions associated with the gasification of petcoke (Figure 8) are considerably higher due to its content of larger carbon to hydrogen ratio. Petcoke is essentially fixed carbon, and due to its very low hydrogen content, it is one of the fossil fuels that generates the highest amount of CO_2 emissions per unit of fuel consumed. On the other hand, NG has a high hydrogen to carbon ratio, resulting in considerably lower CO_2 emissions. The allocation of emissions to the production of power, hydrogen and steam in both the VS and CS is shown in Figure 8. It can be observed that the majority of CO_2 emissions are associated with the production of power from petcoke gasification. In the CS the utilization of additional NG-fueled gas turbines for the production of power contributed to a substantial share of the total emissions.

For the scenario in which carbon capture is not implemented, the production of the same amount of energy from petcoke gasification incorporates CO_2 emissions that are 50% higher than base case energy system. In an operational year the VS of the gasification system would vent around 2.5 Mt of CO_2 per year more than the base case. These number are unacceptable in the current global scenario, especially in a CO_2 constrained industry as Alberta's oil sands. However, these

are considerably mitigated with the integration of carbon capture into the process. Carbon dioxide can be separated and captured more efficiently and at a lower cost if integrated with a gasification plant compared to a conventional combustion based energy production technology [50]. The emissions associated with the CS are approximately 30% lower than those associated with the base case energy system, which equates to approximately 1.5 Mt of CO_2 emissions. Additionally, gasification of petcoke produces very low emissions of SO_2 as most of it converted to H_2S and eliminated in the sulphur recovery unit. This reduction in the emissions intensity is much higher than the one currently required by the Alberta government (12% of current emission levels), and therefore, the proposed energy system can provide an adequate alternative in in the long term planning of a more environmentally sustainable energy infrastructure.

5. Conclusions

In this work a gasification process with carbon capture was simulated and integrated into the energy infrastructure of an integrated SAGD/upgrading facility in order to satisfy its power, thermal and hydrogen requirements. The entrained bed gasifier was selected for the gasification of the delayed coker petroleum coke residue at oxygen and steam to carbon ratios of approximately 0.9 and 0.2, respectively. Part of the syngas generated from the gasifier is used to produce the pure hydrogen required for upgrading use, while the remaining syngas was used as a fuel for the gas turbines to produce power. The syngas coolers and the OT-HRSG downstream of the syngas-fired gas turbines were used to recover the heat required for steam production. The steam to power ratio of the gasification plant obtained to be 3.09 and 6.67 for the venting and capture scenarios, respectively, which is closer to the ratio of steam to power demanded for the extraction and upgrading operation than those obtained from NG based cogeneration systems. The efficiency range of the gasification based polygeneration plant is 48.3 – 58%, which correspond to the CS and VS, respectively. The gasification plant provides all the hydrogen and steam requirements for the upgrading operations in all scenarios investigated. The syngas fired gas turbines can satisfy all the power requirements in the VS. However, the energy penalties associated with carbon capture requires the addition of NG fired turbines to satisfy the electricity demand. Approximately 22-37% of the steam required for bitumen extraction is produced from gasification plants, which considerably reduces the dependence on NG for SAGD steam production. The remaining amount is produced using OTSGs utilizing produced gas and natural gas. The overall efficiency of the gasification polygeneration plant is lower than NG based technologies. However, the NEER was determined to be considerably higher due to the low dependence on external energy sources (i.e. NG), which were determined to be 5.42 and 3.85 for the low and high energy scenarios of the base case, respectively, 26.30/9.19 for the low/high energy scenarios of the venting scenario of the petcoke polygeneration plant, and 17.4/7.79 for the low/high energy scenarios of the carbon capture scenario. The CO_2 emissions associated with the petcoke gasification polygeneration case are considerably higher than that obtained from firing NG. The VS of the gasification case results in producing around 50 % more CO_2 than the base case energy infrastructure (using natural gas in state-of-the-art cogeneration gas turbines and upgrading processes), which are not adhering to the emission regulations of the government

of Alberta. The integration of carbon capture reduced the CO_2 emissions by 30% of that of the base case.

In summary, the deployment of petcoke gasification with CO_2 capture represents an energetically and environmentally sustainable alternative to NG based energy production that would allow for the efficient disposal of growing petcoke waste stockpiles. Besides, the co-gasification with biomass, which is an abundant resource in Alberta that has been receiving a growing interest from oil sands operators, would help to decrease the CO_2 emissions of the polygeneration plant making it more attractive from an environmental point of view. However, gasification based processes are capital intensive, which is even more pronounced with the necessity of incorporating expensive carbon capture and storage technologies. There are also numerous challenges being faced in the development of carbon capture and storage in Canada (e.g. lack of large-scale demonstration, public acceptance, time pressure for retrofitting existing technologies, viability of integrated systems etc.) [50, 51]. Therefore, given also the current low prices, the continued reliance on NG is expected to persist in the near future. Future works shall incorporate a comparison among the viability of various fuels, including biomass, and a detailed techno-economic analysis of the possible plant configurations.

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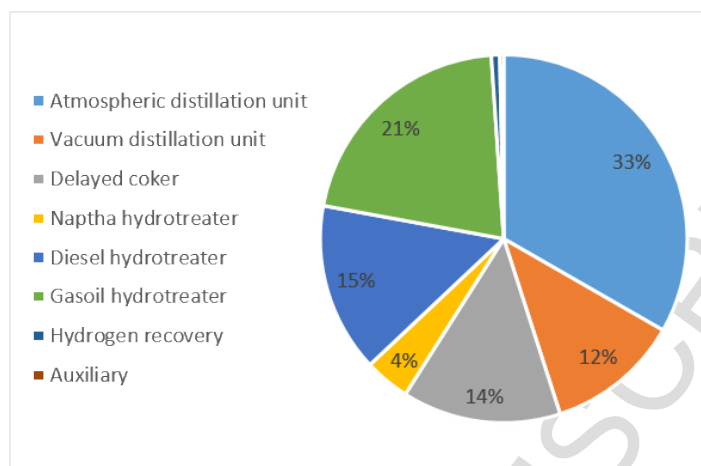
Figure 1 Breakdown of energy consumption of upgrading operations
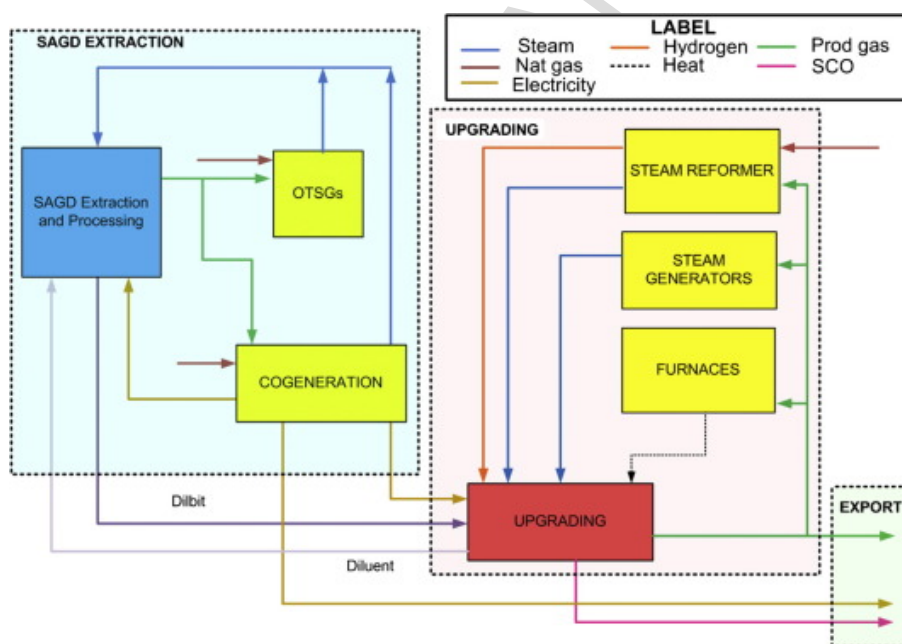

Figure 2 Flowsheet of the base case energy infrastructure model
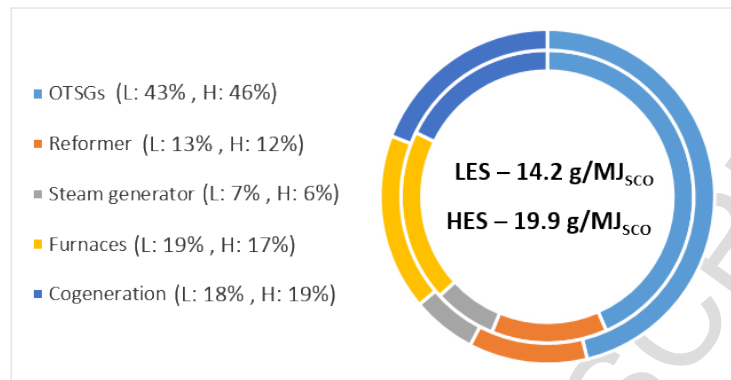
Figure 3 CO$_2$ emissions and breakdown of energy producers
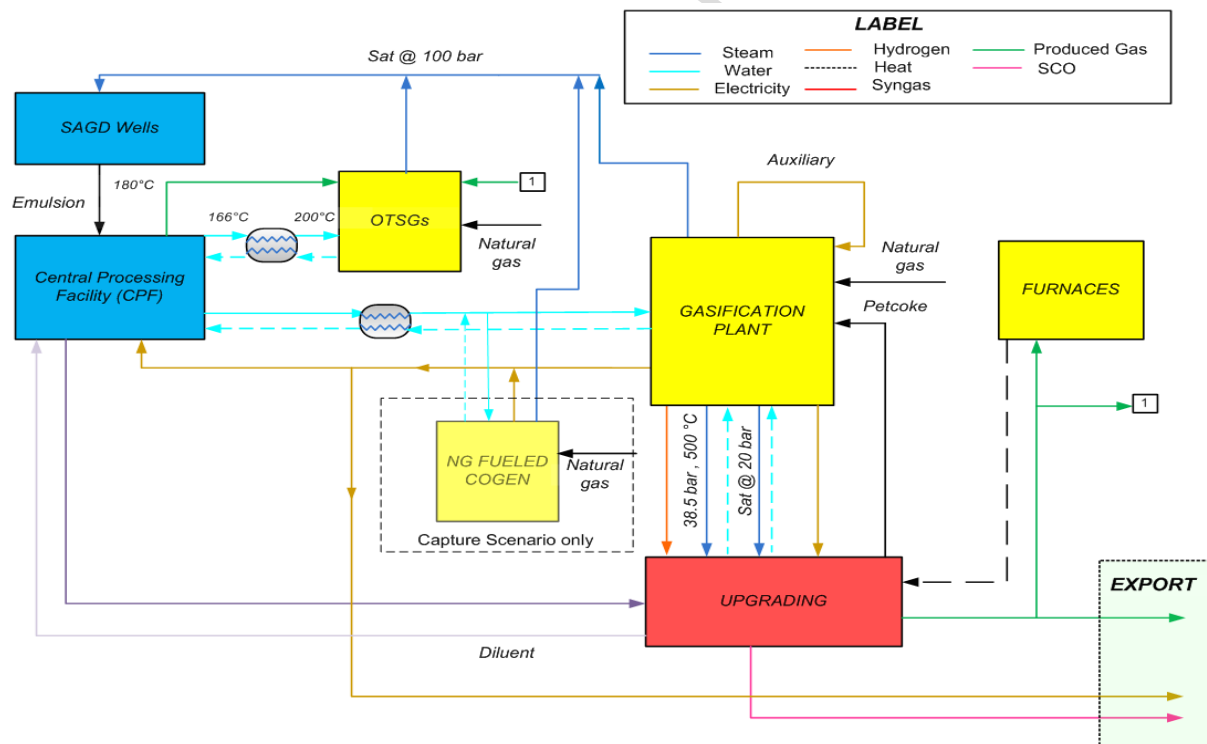
Figure 4 Integrated bitumen extraction and upgrading facility and new proposed energy system

Figure 5 Simplified Aspen Plus flowsheet of the petcoke gasification polygeneration plant

Figure 6 Fuel consumption of energy producers
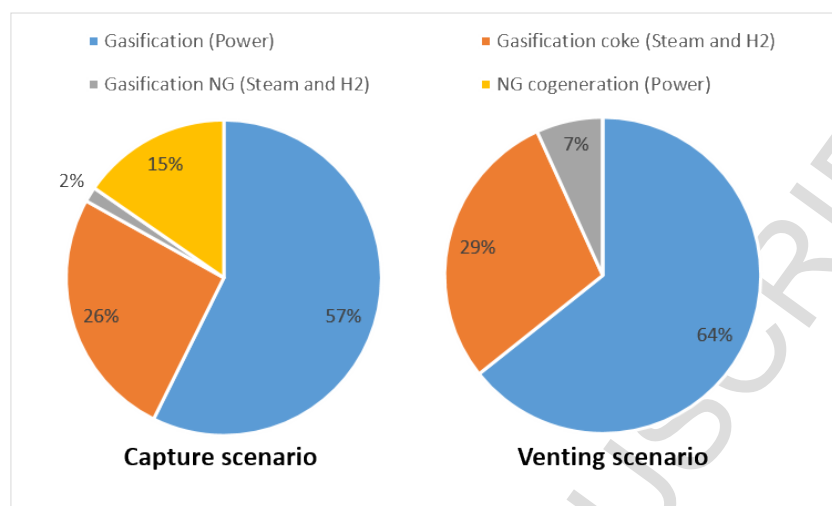
Figure 7 Breakdown of fuel consumption in the cogeneration system

Figure 8 Allocation of emissions to energy producers

Table 1 Energy requirements of SAGD extraction and bitumen upgrading in MW

	SAGD	Upgrading
Heat [LES / HES]	-	327 / 409
Steam [LES / HES]	928 / 1392	141 / 176
Electricity [LES / HES]	36 / 67	26 / 33
Hydrogen [LES / HES]	-	192 / 240

Table 2 Current and proposed energy infrastructure of the integrated SAGD/upgrading facility

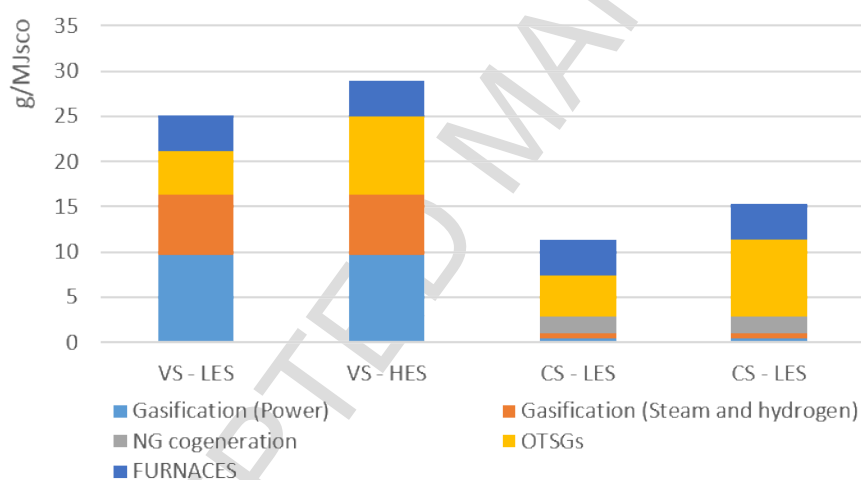| Base case energy infrastructure	Proposed energy infrastructure

• Natual gas cogeneration system (Power / SAGD steam)
   - Gas turbines with postfiring
   - Once-through heat recovery steam generators
• OTSGs (SAGD & Process steam)
• Steam methane reformers (Hydrogen / Steam)
• Furnaces (Heat)
	• Petcoke gasification polygeneration system (Power / Steam / Hydrogen)
	   - Gasification reactor
	   - Syngas treatment (Shift reactors, carbon capture, acid gas removal, etc.)
	   - Gas turbines and OT-HRSGs
	• OTSGs (SAGD & Process steam)
	• Furnaces (Heat)

Table 3 Fuel consumption and performance of the base case energy infrastructure

Parameter	LES	HES
Produced gas consumed (MJ/GJ_{SCO})	76.2	93.3
Natural gas consumed (MJ/GJ_{SCO})	186.4	274.5
NER (GJ produced /GJ consumed)	3.9	3.02
NEER (GJ produced /GJ consumed)	5.52	3.69
CO_2 emissions (g/MJ_{SCO})		
SAGD	7.9	12.0
Upgrading	6.3	7.9

Table 4 Petroleum coke properties

Proximate Analysis	(%)	Properties	
Moisture	0.4	Gross heating value (MJ/kg)	35.2
Volatile	12.45	Initial AFT (ºC)	1074
Fixed carbon	83.37	Softening AFT (ºC)	1285
Ash	3.78	Hemispherical AFT (ºC)	1338
		Fluid AFT (ºC)	1446
Ultimate Analysis	(%)	Particle size distribution	(wt%)
Carbon	83.7	d > 4.76 mm	2.5
Hydrogen	3.7	1.4 mm < d < 4.7 mm	1
Nitrogen	1.8	841 μm< d < 1.4 mm	4
Chlorine	0	250 μm < d < 841 μm	27.6
Sulphur	5.7	149 μm < d < 250 μm	51
Oxygen	1.3	d < 149 μm	12.9

Table 5 Operating parameters of the gasifier

Gasifier reactor

Gasification temperature (°C)	1480
Gasification pressure (bar)	38.5
Carbon conversion (%)	99.5
Oxygen inlet pressure (bar)	46
Oxygen inlet temperature (°C)	100
Steam inlet temperature (°C)	320
Steam inlet pressure (bar)	42
Membrane wall inlet temperature (°C)	250
Membrane wall outlet temperature (°C)	320
Coarse slag percentage (wt%)	35
Fly slag percentage (wt%)	65
Syngas quench temperature (°C)	900
Syngas recycle (%)	49.25

Table 6 Operating parameters of the syngas treatment model

Wet Scrubber	
Scrubber water inlet temperature (°C)	166
L/G ratio in Scrubber	0.25
COS hydrolysis reactor	
COS hydrolysis reactor temperature (°C)	200
COS hydrolysis conversion (%)	>99
WGS operating parameters	
Steam to CO ratio	2.5
Steam temperature (°C)	270
High Temperature Shift inlet Temperature (°C)	250
Low Temperature Shift inlet Temperature (°C)	250
HTS/LTS reactor approach temperatures (°C)	10
Condenser Temperature (°C)	38
PSA Hydrogen Split fraction (%)	0.9
PSA hydrogen Quality (%)	1
Sulfinol-M for removal of H_2S	
LT heat for stripping (MJ/kg)	13.4
CO co-absorbed (%)	0.265
H_2 co-absorbed (%)	0.268
CO_2 co-adsorbed (%)	16
Selexol for selective removal of CO_2 & H_2S	
LT heat for stripping (MJ/kg)	44.7
CO_2 adsorbed (%)	96.54
CO co-absorbed (%)	0.44
H_2 co-absorbed (%)	0.553

Table 7 Composition of syngas produced from gasification (%vol)

Component	Venting scenario	Carbon capture scenario
H_2	27.98	94.15
CO	69.83	1.35
CO_2	0.02	2.34
AR	1.07	1.05
H_2O	0.01	0.01
CH_4	0.08	0.08
N_2	1.01	1.02

Table 8 Performance parameters of the petcoke gasification polygeneration plant for LES/HES

Parameter	VS	CS
Petcoke feed (kg s^{-1} / MW)	51 / 1781	51 / 1781
Natural gas feed (kg s^{-1} / MW)	2.1 / 101.7	0.7 / 32.9
H_2 produced (kg s^{-1} / MW)	2.7 / 317.9	2.7 / 317.9
LPS (kg s^{-1} / MW)	16.7 / 32.7	16.7 / 32.7
MPS (kg s^{-1} / MW)	16.1 / 55.4	16.1 / 55.4
HPS (kg s^{-1} / MW)	240.9 / 528.1	182.6 / 400.9
Gross power (MW)	301.02	261.25
Net power (MW)	158.42	68.48
Steam to power	3.89	6.67
Net efficiency (%)	58	48.26
CO_2 produced (kg/s)	158.5	155.3
CO_2 captured (kg/s)	0	144.84
CO_2 emissions (kg/s)	158.5	10.45

Table 9 Components of net efficiency and performance parameters of the petcoke polygeneration plant

Parameter	Unit	VS	CS
Petcoke feed	kg/s	50.66	50.66
	MW	1781.05	1781.05
Natural gas	kg/s	2.07	0.67
	MW	101.74	32.93
H2	kg/s	2.65	2.65
	MW	317.89	317.89
LPS	kg/s	16.7	16.7
	MW	32.73	32.73
MPS	kg/s	16.11	16.11
	MW	55.42	55.42
HPS	kg/s	240.86	182.62
	MW	528.06	400.88

Overall steam	MW	616.2	456.3
Gross power	MW	301.02	261.25
Net power	MW	158.42	68.48
Steam/Netpower Ratio	MW_{steam}/MW_{power}	3.89	6.67
Net efficiency	%	58	48.26

Table 10 Net energy analysis of the petcoke gasification polygeneration plant

Parameter	Base case LES/HES	VS - LES	VS - HES	CS - LES	CS - HES
SCO production (MW)	9505.47	9505.47	9505.47	9505.47	9505.47
Electricity Export (MW)	2.39/10.7	62.61	21.99	70.88	30.25
Electricity Export in primary energy[1] (MW)	5.97/26.75	156.525	54.97	177.2	75.62
Fuel consumption (MW)	637.73	3317	3990	3504	4177
"External" fuel consumption (MW)	-	364	1037	550.5	1223.4
NER	3.90/3.02	2.88	2.39	2.76	2.3
NEER	5.42/3.85	26.30	9.19	17.4	7.79

[1] To calculate the equivalent primary energy associated to the electricity export, an average net electric efficiency of 40% has been considered for Alberta's electric grid [49]

Acronyms

ADU	Atmospheric distillation unit
AGR	Acid gas removal
ASU	Air separation unit
CGE	Cold gas efficiency
CS	Carbon capture scenario
DC	Delayed coking
FG	Fuel gas
GHG	Greenhouse gas
HES	High energy scenario
HPS	High pressure steam
HTR	Hydrotreater
LES	Low energy scenario
LHV	Lower heating value
LPS	Low pressure steam
MPS	Medium pressure steam
NEA	Net energy analysis
NG	Natural gas
NEER	Net external energy ratio
NER	Net energy ratio
OT-HRSG	Once through heat recovery steam generator
OTSG	Once through steam generator
PG	Produced gas
PSA	Pressure swing adsorption
SAGD	Steam assisted gravity drainage

SCO	Synthetic crude oil
TBD	Thousand barrels per day
VDU	Vacuum distillation unit
VS	Venting scenario
WGS	Water gas shift reactor