Efficient Electrochemical Refrigeration Power Plant using Natural Gas with ~ 100 % CO₂ capture

Easa I. Al-musleh^a, Dharik S. Mallapragada^a, and Rakesh Agrawal^{a,b}

^a School of Chemical Engineering,

Purdue University, West Lafayette, IN-47907

^bTo whom correspondence should be addressed. Email: agrawalr@purdue.edu

Abstract

We propose an efficient Natural Gas (NG) based Solid Oxide Fuel Cell (SOFC) power plant equipped with ~ 100 % CO₂ capture. The power plant uses a unique refrigeration based process to capture and liquefy CO₂ from the SOFC exhaust. The capture of CO₂ is carried out via condensation and purification using two rectifying columns operating at different pressures. The uncondensed gas mixture, comprising of relatively high purity unconverted fuel, is recycled to the SOFC and found to boost the power generation of the SOFC by 22%, when compared to a stand alone SOFC. If Liquefied Natural Gas (LNG) is available at the plant gate, then the refrigeration available from its evaporation is used for CO₂ Capture and Liquefaction (CO₂CL). If NG is utilized, then a Mixed Refrigerant (MR) vapor compression cycle is utilized for CO₂CL. Alternatively, the necessary refrigeration can be supplied by evaporating the captured liquid CO₂ at a lower pressure, which is then compressed to supercritical pressures for pipeline transportation. From rigorous simulations, the power generation efficiency of the proposed processes is found to be 70-76% based on lower heating value (LHV). The benefit of the proposed processes is evident when the efficiency of 73% for a conventional SOFC-Gas turbine power plant without CO₂ capture is compared with an equivalent efficiency of 71.2% for the proposed process with CO₂CL.

Keywords: Liquefied natural gas, Carbon capture and sequestration, solid oxide fuel cell, refrigeration

1. Introduction

Concerns over the role of anthropogenic Greenhouse Gas (GHG) emissions in affecting climate change have prompted widespread research and development efforts into reducing future emissions from stationary and distributed sources [1, 2]. As of 2010, plants producing heat and electricity from the combustion of fossil fuels like coal and Natural Gas (NG) accounted for ~41% of global CO₂ emissions [3, 4]. One option towards reducing future CO₂ emissions from electrical power generation is implementing carbon-free renewable energy sources like solar, wind etc. The global installed generation capacity of solar and wind power grew by ~73% (29 GW) and ~20% (41 GW) in 2011 [5]. Overall, the global installed generation capacity of non-hydroelectric renewable power grew by 77 GW in 2011 [6], due to a combination of several factors including favorable government policies and substantial cost reductions. However, in the absence of cost-effective energy storage options, the intermittent nature of the energy source currently limits the use of renewable power generation to peak shaving rather than baseload applications.

Alternative approaches for reducing CO₂ emissions from baseload power generation rely on using fossil fuels more efficiently with or without CO₂ Capture and Sequestration (CCS) [2, 7]. Implementing CCS for electric power generation involves capturing 80-90% of the CO₂ in the exhaust gas from the power plant, compressing it to supercritical conditions (for pipeline transportation), and subsequently storing it in a suitable geological location [8]. The energy input for these steps comes at the expense of about 9-15% reductions in the efficiency of converting primary energy to electricity [7, 9, 10]. If ships, road or rail tankers are utilized for CO₂ transportation, then CO₂ needs to be

maintained as a low pressure subambient liquid [11], which may require additional electricity input due to the need for a refrigeration cycle. Currently, the focus of CCS deployment efforts has been towards power plants using coal rather than NG [7]. This is because a coal power plant has roughly double the CO₂ emissions when compared to the same size state-of-art Natural Gas Combined Cycle (NGCC) power plant [7]. The recent increased interest in developing shale gas reserves, notably in the United States (U.S.), points to the possibility of increased use of NGCC power plants for electricity generation [12]. For example, the share of total U.S. electricity generation from NG has increased from ~21% in 2008 to ~25% in 2011 [13]. However, despite using the lower GHG emitting NGCC power plants, it is estimated that CCS technologies would be necessary to achieve 50-80% GHG emissions reduction compared to the present levels [4, 7].

In this paper, we present process designs for high efficiency electrical power generation along with CO₂ Capture and Liquefaction (CO₂CL) from NG, available either as Liquefied Natural Gas (LNG) or compressed gas from a pipeline. The process, referred as the *Electrochemical Refrigeration Power Plant* (ERPP), is based on an integrated steam reforming-Solid Oxide Fuel Cell (SOFC) system followed by a unique process for CO₂CL. The ERPP is capable of achieving ~100% CO₂ capture and liquefaction while attaining a power generation efficiency of 70.4-76.0%. Here, power generation efficiency (Eq. 1) is defined as the ratio of the net electrical energy output of the plant to the primary energy input of the feed fuel based on its Lower Heating Value (LHV). In Eq. 1, the first, second, third, and fourth term of the numerator refer to the work output from turbines (> 0), SOFC (> 0), compressors (< 0) and pumps (< 0), respectively. When compared to a SOFC plant that does not capture CO₂, the CO₂CL section in our process

enables the plant to have nearly zero CO₂ emissions with an energy penalty of less than 2%. While the CO₂CL process is demonstrated in this work for SOFC exhaust, the process may be applied for other applications and technologies such as oxygen firing NGCC [10]. In the following sections, we describe the different ERPP designs and their simulation via Aspen PlusTM [14]. This is followed by a summary of the key results, sensitivity analysis, and comparison with other NG power plant designs. Additional details and sensitivity analysis are provided in the supporting information (SI).

$$\text{Power generation efficiency} = \frac{\dot{a}_{\text{turbines}} W_{T-i} + W_{\text{SOFC}} + \dot{a}_{\text{compressors}} W_{K-i} + \dot{a}_{\text{pumps}} W_{P-i}}{\text{Feed}_{\text{LHV}}} \times 100 \quad (1)$$

The concept of the ERPP is shown in Fig.1. If NG is available as LNG, it is first evaporated and then fed to an integrated steam reforming-SOFC system for electrical power generation. The exhaust from the SOFC anode, composed primarily of the oxidation products CO₂ and H₂O (~80-90 mol%), is processed for CO₂CL using the cold refrigeration released from the evaporating LNG and a small amount of additional electricity. If NG is available from a pipeline, then a greater portion of the generated electrical power (i.e. greater than the case of LNG) is utilized to supply the refrigeration needs of the CO₂CL process. In both cases, the CO₂CL process is the same, except for the source of refrigeration.

1.1. SOFC exhaust gas utilization

The exhaust of a SOFC, which operates using gas leaving the steam reformer, contains unconverted H₂ and CO in amounts that depends on the extent of fuel utilization (i.e. operating current) [15-17]. A suggested process to capture CO₂ from the SOFC exhaust stream is to first convert CO to H₂ via a high-temperature and then a low-

temperature Water-gas Shift (WGS) reaction, followed by the CO₂ capture through a SelexolTM process [18]. The purified H₂ rich stream leaving the SelexolTM process is then combusted for additional power generation [18]. Although such a design leads to high power generation efficiency from NG (~69% according to [18]), the CO₂ recovery is generally limited to ~90%. It has also been proposed to use two SOFC devices in series, the first one produces electrical power and the second acts as an afterburner that achieves near 95 % combustion (with negligible power generation, see Fig. 5 and ref. [18, 19]). Here, the system benefits from fuel combustion using pure oxygen and thereby avoids diluting the CO₂ exhaust. The reported power generation efficiency at a CO₂ recovery near 90% is around 68%, which is 5% lower than a process that does not capture CO₂ (see section 6).

To achieve close to 100% CO₂ recovery, it has also been suggested to combust the SOFC exhaust with a pure O₂ stream supplied from an Air Separation Unit (ASU). This leads to a power generation efficiency of ~63% LHV basis [20], a 10% efficiency loss when compared to a process that does not capture CO₂ (see section 6). While use of an enriched O₂ stream, instead of pure O₂ could reduce the power demand of the ASU, it may not necessarily improve the overall process efficiency. This is because the flue gas leaving the combustion unit will need to be purified from its associated inerts (i.e. N₂, Ar, from the ASU) to obtain the required CO₂ sequestration purity [21]. A trade off between CO₂ purification and ASU power is presented by Wilkinson et al. [22] for CO₂ capture from a furnace exhaust combusting fuel with enriched O₂. Also it is worth mentioning that, the relatively low composition of H₂ and CO (i.e. low LHV) in the SOFC anode

exhaust (at an operating current corresponding to the maximum SOFC efficiency) could necessitate the use of expensive catalytic combustion [23].

The proposed ERPP design, through the use of a novel refrigeration based process for CO₂CL, can achieve nearly 100% CO₂ capture with a purity of 99.6 mole %. This means that nearly all the unconverted fuel (i.e. H₂ and CO) from the SOFC exhaust is recovered at a much higher purity (containing ~ 12 mole % CO₂) and is available for recycle to the SOFC anode. The recycle stream is responsible for the nearly 22% increase in the SOFC power output compared to a standalone SOFC operation, while achieving near zero CO₂ emissions. Just as the above studies need capital equipment for CO₂ capture (using solvent absorption, WGS, or ASU etc), our proposed process also requires capital equipment for CO₂ capture (i.e. multistream cryogenic heat exchangers, compressors etc.). However, our unique proposal of recycling relatively rich unconverted fuel stream to the SOFC anode avoids the need for additional capital equipment, such as catalytic combustor and or a combustion based turbine, for its utilization. A process economic analysis, out of the scope of this work, can quantify the relative capital needs of our process compared to the other processes.

1.2. Other features of ERPP

Depending on the storage site (or end user) location, the transportation of CO₂ may either require supercritical compression for pipeline transportation, or liquefaction for sea, road or rail transportation. The ERPP design produces low pressure liquid CO₂, which can be readily shipped [8, 11]. The use of LNG is among the preferred methods for shipping large quantities of NG over long distances. In 2011, the world wide LNG imports was estimated to be ~ 331 billion standard m³ (at 15 °C and 1 bar) [5], which is

equivalent to 6.61×10^5 TWh on a LHV basis (see SI). However, liquefaction of NG is energy intensive, typically resulting in ~6% of the feed NG being consumed for the process [24]. Around 0.2 kWh of compression energy is needed to produce one standard m^3 of LNG using a propane-mixed refrigerant vapor compression cycle [25]. By considering LNG rather than NG as feed to the ERPP, the design enables the recovery of a portion of the energy invested in the upstream NG liquefaction process. Previous works have addressed the recovery of LNG refrigeration for N_2 liquefaction, power generation, and CO_2 liquefaction [8, 24, 26-30]. In one example configuration for CO_2 liquefaction, the LNG is first evaporated and then split for pipeline NG supply and power generation using O_2 combustion (O_2 supplied from ASU) [26, 27]. Subsequently, CO_2 produced from the power generation is liquefied using all of the available LNG refrigeration. In contrast, the ERPP design relies on a unique CO_2CL configuration that captures and liquefies ~ 100% of the generated CO_2 using the refrigeration from the LNG feed to the power plant along with a small amount of the generated electric power.

The ERPP design also benefits from the use of SOFC systems versus combustion-based turbine power generation NGCC systems [16, 31, 32]. The SOFC, producing electricity in a single electrochemical conversion step, reduces the fuel chemical exergy losses associated with multistep systems, such as combustion-based methods [31, 32]. SOFC also avoids direct mixing between the fuel and air, instead allowing for oxygen from the cathode (air feed) to diffuse as an oxygen ion to the anode (fuel feed) for oxidation [31, 32]. This means that the anode exhaust is not diluted with nitrogen (or other diluents), which makes the subsequent CO_2 capture less energy intensive relative to the NGCC plant using air.

2. Design and simulation of the ERPP

A detailed configuration of the ERPP design, with LNG as feed, is shown in Fig. 2 and Fig.3. Aspen PlusTM is used to rigorously simulate the process using the PSRK thermodynamic property method and steam tables (see SI for stream data). For the range of temperature and pressures associated with the ERPP, the vapor-liquid equilibrium predictions of the PSRK method for system compare reasonably well with the available experimental data. This comparison is provided in section 7 of the SI. Aside from the SOFC unit (discussed later), the remaining part of the process consists of heat exchange and compression processes that are modeled using built-in blocks within Aspen PlusTM and operated within their known temperature and pressure limits. The process is designed for a net power output of ~158 MW, with details of the simulation basis shown in Table 1.

The modeling parameters for the SOFC are adopted from the relevant references shown in Table 1 [15, 16, 36-38] and implemented in Aspen Plus using Aspen Calculator tool [14]. Here, the oxidation reactions in the SOFC are modeled to be at chemical equilibrium, which is deemed reasonable due to the high operating temperature of the reformer-SOFC unit. We also include experimentally derived relations (from [15, 16]) for electrochemical losses due to anode/cathode activation/diffusion polarization, electrolyte resistance, and interconnectors resistance. Since the adopted model from Van herle et. al. [15, 16] is based on atmospheric SOFC operation, the model is corrected to include the effect of pressure by adding to the calculated voltage a pressure adjustment factor obtained from Lundberg et al. [36]. The ERPP design consists of the three steps of power

generation and heat recovery, CO₂ compression and dehydration, and CO₂CL. Detailed descriptions of each of these steps are provided below.

2.1. Power generation and heat recovery

Since the direct use of NG in the SOFC causes carbon deposition, the NG feed needs to be converted to H₂-rich synthesis gas (syngas) [45]. In the process of Fig. 2, NG is steam reformed internally (but separated from the anode) by soaking up the SOFC waste heat. This heat integration is a unique feature of SOFC systems [31, 39]. The steam reforming reactions are modeled using chemical equilibrium at the reformer operating conditions. This modeling approach of the reformer-SOFC has been adapted from the publications that have presented SOFC models and their validation with experimental data. The assumption of chemical equilibrium for the reformer and the oxidation reactions at the anode is deemed reasonable based on the high operating temperature of the unit (~ 950 °C) [15, 16, 39]. Referring to Fig. 2, the stored LNG is first evaporated and its refrigeration is utilized for CO₂CL (discussed in section 2.3). The vaporized NG (i.e. stream 6) is then heated up to 680°C in heat exchanger E-2, mixed with 10 bar superheated steam (at 680°C) sourced from reboiled column C-1, and then fed to the SOFC reforming section. Before reaching 680°C, a portion of the NG leaves heat exchanger E-2 via stream R1 at 310°C to regenerate the molecular sieve adsorber V-6 (see section 3.2). NG plus desorbed water leave the regenerating adsorber V-6 via stream R2 to be heated up to 680°C in E-2 and finally mixed with stream 7 for NG steam reforming. In heat exchanger E-2, heat is supplied by cooling the exhaust stream from turbine T-4 (i.e. stream 10), where the turbine discharge pressure is adjusted to achieve a minimum temperature approach of 28°C (design basis) in the heat exchanger. Increasing

the temperature of the SOFC improves the efficiency of SOFC power generation and increases the conversion of NG to syngas [37, 39]. Therefore, we have chosen to operate the SOFC at 950°C, which is close to the maximum limit of 1,000°C. The reformer is operated at 922°C to maintain a minimum temperature approach for heat exchange of 28°C (Table 1). Further, increasing the SOFC operating pressure improves power generation efficiency by reducing the heat dissipation due to the aforementioned electrochemical losses. Here, we set the SOFC pressure to be 10 bar, close to the maximum allowable pressure for existing commercial systems [36]. The generated syngas is fed to the SOFC anode to be electrochemically oxidized at 950°C and 10 bar according to the following reactions.

The fuel utilization in the SOFC is defined as the percent syngas conversion via the oxidation reactions 2-3. In agreement with the current corresponding to the maximum power generation efficiency of the ERPP (see section 2.3), the fuel utilization is set to ~80%. This means that the amount of O₂ consumed by reactions 1-2 is ~80% of the stoichiometric requirement. The O₂ needed for the fuel oxidation is provided by compressed and heated air stream 24 fed to the cathode side of the SOFC. Stream 24, at 680°C and 10 bar, is produced from feed air stream 22 via multistage intercooled air compressor K-1 followed by heat exchanger E-1. At the SOFC cathode, oxygen ions are generated via reaction 4 and diffuse through the yttria-stabilised zirconia electrolyte to the anode side for fuel oxidation [15, 16]. For the chosen fuel utilization of 80%, excess air (equal to 1.07 times the oxidation stoichiometry requirement) is fed to the SOFC

cathode to absorb the SOFC waste heat (of ~ 23 MW) that is in excess of the steam reforming heat requirements (of ~ 70 MW). Depleted air stream 25 leaves the cathode side of the SOFC to be expanded in turbine T-2, partially cooled to 180 °C against inlet air stream 23 in heat exchanger E-1, and finally expanded to 1.1 bar in turbine T-3. Similar to turbine T-4, the discharge pressure of turbine T-2 is set to achieve a minimum temperature approach of 28°C in heat exchanger E-1.

2.2. CO₂ compression and dehydration

Dehydration of CO₂ prior to storage and transportation is essential for the following reasons. 1) Storage of CO₂ at low pressure requires subambient cooling that may freeze water, if not removed. Freezing of a component in a stream could block pipelines and heat exchanger passages that would disrupt normal process operation. 2) When contacted with water, CO₂ tends to form carbonic acid (see reaction 5), which tends to corrode typical storage and pipeline materials of construction (e.g. carbon steel). The stream containing CO₂ needs to be dehydrated to the extent such that the water dew point (i.e. the temperature at which water starts to condenses) is below the storage temperature of CO₂, which is -52°C in this study. This extent of dehydration can be achieved using molecular sieve adsorbers (V-4, V-5, and V-6 in Fig. 2) [46].

Referring to Fig 2, the SOFC anode exhaust, after its passage through heat exchanger E-2, stream 11, is fed to the reboiler of stripping column C-1 to provide the column with the heat duty (at 157°C) necessary for boilup. Stream 12 leaving the reboiler, is then cooled to 43°C in heat exchanger E-3 (using cooling water) to condense a portion of the

water present in the stream. The uncondensed gases are separated from the condensed water in a vapor-liquid separator V-1 and compressed to 15 bar in the multistage intercooled compressor K-2. The compression of the SOFC exhaust via K-2 to 15 bar improves the power generation efficiency and per pass liquid CO₂ recovery in the downstream CO₂CL section (see section 3.2 and SI for details). Compression also favors water condensation to produce a gaseous CO₂-rich stream. This reduces the dehydration load (e.g. amount of water removal) on the downstream molecular sieve adsorbers (i.e. V-4, V-5, and V-6). As the gaseous stream is compressed in K-2, condensed water plus small amounts of dissolved gases (i.e. H₂, N₂, CO, CO₂) leaving the interstage coolers are separated via vapor-liquid separators V-2 and V-3. The water-rich liquid separated in V-3 is pressure reduced from 15 bar to 12 bar and then fed to the upstream separator V-2. This enables the partial recovery of the dissolved gases in the liquid leaving separator V-3 at intermediate pressure of 12 bar, thereby avoiding low-pressure recompression work. Similarly, the water-rich liquid leaving V-2 is pressure reduced from 12 bar to 9 bar to further separate the dissolved gases in separator V-1. Lastly, the collected water-rich liquid leaving separator V-1 (stream 17 ~99.8 mol% H₂O) is pressure reduced from 9 bar to 2 bar and fed to stripping column C-1. Column C-1 strips stream 18 of the remaining dissolved gases. The stripped gases are sent to separator V-1 via compressor K-3.

The CO₂-rich gas, stream 14, leaving separator V-3 is sent to the molecular sieve adsorbers, V-4 and V-5, to reduce the stream water dew point (by water removal) to -60°C. While two adsorbers dehydrate gas stream 14, a fully water saturated adsorber (in this case V-6) goes through regeneration by means of hot NG stream (stream R1). When

the regeneration is completed, the adsorber is brought online for dehydration of CO₂ stream 14 and another water saturated adsorber goes through regeneration.

2.3. CO₂CL

The capture of CO₂ using a refrigeration based processes needs to be designed such that stream temperatures remain above -55°C to avoid the possibility of freezing CO₂. While cooling at higher pressures could eliminate the possibility of CO₂ freezing, condensation of the other components (such as CO and H₂) may take place due to the increased dew points of these components. This is undesirable as some of the unconverted fuel (from the SOFC) will be lost with the captured CO₂. In our proposed process (details discussed below) for CO₂CL, maximum recovery of CO₂ is achieved by cooling the exhaust to a temperature not less than -55°C using three stages at which each stage is operating at different pressure. Each stage produces recovered liquid rich CO₂ and unrecovered gaseous lean CO₂. The unrecovered gaseous CO₂ from the first stage is compressed to higher pressure for recovery in the second stage and so on. To minimize the condensation of the unconverted fuel (i.e. lost of unconverted fuel with the captured CO₂), the recovered liquid rich CO₂ from the second and third stages are reduced in pressure and used as reflux streams in a first and second rectifying columns, respectively. Inside each column the vapor will strip the CO and H₂ from the liquid reflux. At the same time, the liquid reflux will absorb the CO₂ associated with the vapor (see below for detailed explanation). In other words, each column operating at a different pressure, achieves some degree of separation, to ultimately recovery high purity (>99%) of liquid CO₂.

Referring to Fig.3, the dry CO₂-rich gas stream 29 leaving the molecular sieve adsorbers is fed to the CO₂CL section, described in Fig. 3. Almost 85 mole % per pass of the feed gaseous CO₂ is recovered and stored as liquid CO₂ at a purity of ~99.6 mole %, at -52°C and 10 bar. The remaining CO₂ is recycled via stream 48 to the SOFC and contains ~ 12 mole % CO₂. Such high recovery and purity are achieved using three separation stages, represented by C-2, C-3, and V-7, and the vaporizing LNG provides a good portion of the refrigeration required by the process.

In the first separation stage, stream 29 is cooled to a temperature of 0°C and then its pressure is reduced to 10 bar (stream 31). This pressure reduction further cools the stream to -3°C using the Joule-Thomson (J-T) effect. In an alternative configuration, the J-T valve may be replaced with a turbine device to achieve greater cooling by power generation (see section 4). The vapor portion of the cooled stream leaving the J-T valve is then refluxed in rectifying column C-2 with liquid CO₂-rich stream (~98 mole % CO₂) sourced from rectifying column C-3. This produces a 99.6 mol% CO₂ stream 32 from the bottom of C-2. The uncondensed gases, stream 33, leave column C-2 for refrigeration recovery in heat exchangers E-5 and E-4. Stream 34 leaving exchanger E-4, completes the first separation stage and is then compressed to a pressure of 34 bar via compressor K-4.

The compressed gas, stream 35, is fed to the second separation stage where it is cooled via heat exchangers E-4 and E-5 to 1°C. The cooled stream 36 is then fed at the bottom of the rectifying column C-3 and the rising vapor is refluxed with stream 42 (containing ~ 92 mole % CO₂) sourced from vapor-liquid separator V-7. The uncondensed gas, stream 38, leaving column C-3, is fed to E-5 and E-4 to recover

refrigeration, producing stream 39 which completes the second separation stage. Stream 39 is recompressed in compressor K-5 to pressure of 100 bar for the last separation stage. Stream 40 is partially condensed by cooling via E-4, E-5, E-6, and E-7 to -53°C . The uncondensed gases are separated in vapor-liquid separator V-7 to produce stream 43, which is heated via E-6, E-5 and E-4 and then expanded in turbine T-5. The discharge pressure of turbine T-5 is limited to ensure that the discharge temperature of stream 45 should be greater than -55°C . This temperature limit is necessary to avoid the undesirable CO_2 freezing conditions. Stream 45 is fed to E-6, E-5 and E-4 for refrigeration recovery. Leaving heat exchanger E-4, stream 46 is further expanded in turbine T-6 to around 10 bar. The turbine discharge stream 47 is fed to heat exchanger E-4 for refrigeration recovery before being recycled to the SOFC as stream 48.

The LNG sourced from the storage tank (at -160°C and 1.3 bar) is pumped to a pressure of 100 bar via pump P-1 and fed to E-7, E-6, E-5, and E-4 to provide the primary process refrigeration needs. Stream 3 leaving heat exchanger E-4 is expanded in turbine T-1 to a pressure of ~ 40 bar, which cools down the NG from 35°C to -24°C . Any further pressure reduction will cause the heavier components in the NG stream (i.e. propane and nbutane) to condense in the turbine and is therefore avoided. Instead, a J-T valve is utilized to generate stream 5 at 11 bar and -46°C . This stream is fed to E-6, E-5, and E-4 to provide the remaining process refrigeration, and leaves as stream 6 at 35°C for the high temperature heat exchanger E-2.

The above CO_2CL operating conditions are identified using the optimization problem formulated in Table 2. Here, the compression power (K-2, K-4, K-5) is minimized subject to heat exchangers minimum temperature approach $\geq 1.1^{\circ}\text{C}$, per pass liquid CO_2

recovery of ≥ 85 mole %, and liquid CO₂ purity ≥ 99.5 mole % using Aspen PlusTM Successive Quadratic Programming (SQP) method. The optimization is carried out using multiple starting points for the various independent variables to ensure a good quality optimum solution. This is further supported by sensitivity analysis on the key process variables, presented in the SI.

3. Discussion

3.1 Thermodynamic analysis

The presented ERPP design operates with an electrical power generation efficiency of 71.2 % while achieving ~100 % CO₂ capture. This efficiency is based on the process heat integration between the hot and cold streams, depicted by the composite curves of Fig. 4a. Here, cold streams below the temperature of the cooling water utility (at 32°C) are not included. In addition, Fig. 4a does not consider integration of the SOFC waste heat. Instead, the integrated steam reformer-SOFC system is modeled to operate adiabatically, with the SOFC waste heat soaked up for heating the air stream 24 (from 680°C to 922°C) and the NG reforming reactions (at 922°C). According to Fig. 4a, the process is deemed to be heat sufficient, which implies that there is no need to burn a portion of the feed NG for heat supply. Fig.4a also suggests that a portion of the heat available from the hot streams (~155°C and below), rather being rejected to the cold utility, can be used for additional electrical power generation. The heat rejected to the cold utility translates into ~18.4 MW of exergy or 11.6% of the plant net power output. A portion of this exergy could be recovered by use of Rankine cycle with maximum temperature near 155°C. At such relatively low temperatures, using an organic fluid such as propane instead of steam may be more suitable for the Rankine cycle [47]. Although

we have not included such a cycle in our analysis, considering about 20% of the exergy of the rejected heat may be recovered as electricity, the power generation efficiency could increase from 71.2% to 72.8%. Alternatively, the rejected heat could be utilized to co-generate low pressure steam to be exported.

$$W^{CO_2CL}_{Min.}=EX_{32}+EX_{48}-EX_{29} \quad (6)$$

$$W^{LNG}_{Max.}=EX_1-EX_6 \quad (7)$$

According to Eq. 6, the minimum power for capturing and liquefying CO₂ is 1.7 MW (or 5.9 MJ kmol⁻¹ liquid CO₂ stored). Here, EX_j refers to the rate of exergy flow associated with a stream j (see SI). On the other hand, the rate at which exergy (i.e. maximum power) is available from vaporizing LNG (Eq.7) is 2.8 MW (or 9.9 MJ kmol⁻¹ liquid CO₂ stored) as per Eq.6. The difference between Eq. 6 and Eq.7 represents the maximum rate at which exergy may be available from the vaporizing LNG, and is equal to ~1.1 MW. In other words, for a thermodynamically reversible CO₂CL section, ~ 1.1 MW of power could be produced. However, the simulation results in Table 3 suggest that, in addition to the available LNG refrigeration, around 5.1 MW of compression power ($W_{K-4}+W_{K-5}+W_{P-1}-W_{T-1}-W_{T-5}-W_{T-6}$ from Table 3) is required. This additional power requirement is a consequence of the process exergy losses (irreversibilities) originating from temperature differences, pressure drops, compressors/turbines inefficiencies, etc. For example, the process composite curves for the CO₂CL heat exchangers (E-4, E-5, E-6 and E-7 in Fig. 3), shown in Fig. 4b, reveal large temperature differences (i.e. large irreversibilities) in the coldest section of the process (-52°C to -153°C). Such large temperature differences are contributed by the hot streams physical properties (i.e. heat capacities, latent heat, etc.) as well as CO₂ minimum temperature constraint. As mentioned before, cooling of

streams containing CO₂ is set to be not below -55°C to avoid the undesirable CO₂ freezing conditions. It should be noted that when natural gas from a pipeline rather than LNG is used, the losses at cold temperatures are much smaller than the ones shown in Figure 4(b) (see SI).

3.2 Sensitivity analysis

Fig. 5 shows the effect of the SOFC fuel utilization on the process power generation efficiency for different values of Steam to Carbon ratios (STC) used for NG reforming. For the base case, NG reforming is carried with STC of 2.5 that is typically used without carbon deposition issues in hydrogen production plants [48]. For fixed STC, increasing the fuel utilization beyond 77% corresponds to decreasing power generation efficiency. The higher currents (i) associated with increased fuel utilization translate into a greater fraction of SOFC energy input lost as heat (losses scale as i^2). Even though a portion of the SOFC waste heat is recovered to produce power via air expansion, the cascading process energy losses (due to air compression and expansion) result in reduced work output and lower power generation efficiency. If the fuel utilization value is less than 77 %, then the SOFC waste heat is observed to be insufficient to meet the heat requirements of the NG steam reforming. In such a case, a portion of the NG needs to be combusted to supply the deficit heat required, which is not considered in this article. Also shown in Fig. 5 is the improvement in power generation efficiency if the steam reformer can be operated at lower STC ratios. As the steam requirement is reduced, the turbine T-4 discharge pressure can be lowered to produce additional power without violating the minimum temperature approach in heat exchanger E-2. For instance, the maximum power generation efficiency of 76.0% is achievable at STC ratio of 1.5, which is close to

the critical STC ratio for steam methane reforming [49]. Currently, this is made impractical due to the need for excess steam to suppress carbon formation during reforming [49].

The process power generation efficiency can also be improved by reducing the minimum temperature approach of the heat exchangers and SOFC. For example, for a STC ratio of 2.5, reducing the minimum temperature approach from 28°C to 5°C raises the power generation efficiency from 71.2% to 72.2% (see SI for details). Designing heat exchanger E-1 and E-2 to operate with lower minimum temperature approaches (i.e. lower than 28°C) will give the opportunity to further expand streams 9 and 25 in turbines T-4 and T-2, respectively. This will increase the net power output of the plant. However, it should be noted that reducing the minimum temperature approach would increase the heat transfer area and result in increased capital costs. Similarly, a lower minimum temperature approach than 28°C in the SOFC will reduce the air required for waste heat absorption in the SOFC. Consequently, the multistage compressor K-1 will process less air and require less compression power.

The sensitivity of the power generation efficiency to the power needed for CO₂ compression and dehydration is represented in Fig. 6a. For all other process variables remaining constant, the maximum power generation efficiency is achieved for a compressor K-2 discharge pressure of 15 bar. Increasing the discharge pressure beyond 15 bar will allow for condensing additional water from stream 14 in Fig. 2. This will reduce the regeneration energy penalty for the molecular sieve adsorbers (see Fig. 6b) at the expense of increasing the K-2 compressor power input. On the other hand, pressures below 15 bar will reduce K-2 compression power but will increase the regeneration

energy requirements of the molecular sieve adsorbers. As a result greater fraction of the NG feed needs to be diverted via stream R1 to the adsorber being regenerated (see Fig. 6b). The overall effect of decreasing K-2 pressure below 15 bar is a reduction in the power generation efficiency. As shown in Fig. 6b, for a K-2 discharge pressure below 12 bar, the regeneration gas flow is identified to be more than the NG feed flow to the plant. Therefore, it is preferable to keep the K-2 discharge pressure to be higher than 12 bar.

3.3 Simplified configuration of the ERPP

The proposed process of Fig. 2 is meant to represent an optimized version of the ERPP where all possible power generation opportunities are exploited to achieve the highest possible power generation efficiency. Therefore, while this optimized version could appear complex, one can always eliminate a few units at the expense of a reduction in power generation efficiency. For example, turbine T-4 (which is associated with low pressure ratio) could be eliminated in an attempt to simplify the process. The power generation efficiency, however, is estimated to drop from 71.2 to 70.8%. Another possible process simplification is the air compression (i.e. K-1) currently using three stages with intercoolers. While single stage compression will reduce the complexity of the process, the compression power will increase from near 27.4 to 30.4 MW. Another concern with single stage compression is the compressor discharge temperature. For example, single stage air compression will lead to a discharge temperature near 374°C which may require expensive alloys for manufacturing. The need for a simplified version of the process of Fig. 2 is also likely to depend the power generation capacity of the power plant. Above a certain power capacity, the economy of scale could make it cost-effective to consider the optimized version of the process shown in Fig. 2. Eventually,

detailed efficiency and cost analysis, which is outside the scope of this work, would be essential to determine the precise process layout to be selected.

4. Alternative designs for CO₂ capture

While the CO₂CL section of Fig. 3 is carried out using three separation stages, the process objective (i.e. CO₂ recovery and purity) can also be achieved using a configuration with two separation stages. However, as discussed in the SI, a major drawback of such configuration is its higher compression power requirements relative to the three separation stages configuration of Fig. 3. Further net power reduction is expected if one chooses to use four or more separation stages.

As mentioned earlier, an alternative CO₂CL configuration to the one of Fig. 3 would use a turbine device instead of a J-T valve for expanding stream 30 to 10 bar, resulting in stream 31 temperature of -32 °C. Such lower temperature (lower than the case of Fig. 3) is a consequence of the power generation (i.e. enthalpy around the expander is not constant) associated with the turbine and leads to an improvement in the process efficiency from 71.2% to 71.7%. In this case, the optimal discharge pressure of the compressors K-2, K-3, and K-5 are calculated to be 18, 25, and 83 bar, respectively. Additionally, stream 29, 35, and 40 would need to be cooled to temperatures of -5°C, -6°C, and -51°C, respectively.

Rather than LNG, if NG is available as a compressed gas from a pipeline (desulfurized and at 31 bar, 38°C) at the plant gate, then the CO₂CL section can be modified to either of the processes shown in Fig. 7 or 8. In both of these configurations, NG is expanded in turbine T-1 to 11 bar (SOFC pressure plus pressure drops), and then used in the CO₂CL section to recover the refrigeration available from expansion. In other

words, both configurations make use of the available exergy in the NG feed to produce near 0.46 MW of electrical power from T-1, and provide nearly 0.61MW of refrigeration (or heat) duty (for temperature range of -20.8 to 41.8°C) in heat exchangers E-4 and E-5 (see Fig. 7 and 8). Overall, the combined exergy of the feed NG is lower than the combined exergy recovered from LNG in the process of Fig. 3 (0.35 MW of electrical power from T-1 and 4.2MW of refrigeration duty, for temperature range of -152.8 to 34.7°C). LNG is associated with higher exergy than feed NG because of its availability as liquid at low temperature. If the exergy of the feed NG or LNG is not available for utilization, then a vapor compression cycle will be needed to meet the refrigeration duty of the CO₂CL. With a thermodynamic efficiency of 50% (ratio of minimum work to actual work needed by the cycle), the vapor compression cycle is estimated to require 3.4MW of additional compression power. In the process of Fig. 7, the pressure of liquid CO₂ (99.6 mol%) in stream 32 is reduced to 6 bar (set to give a minimum temperature approach of 1.1 °C) and then evaporated in the heat exchangers to provide the remaining process refrigeration. The CO₂ gas, stream A, is compressed in intercooled multistage compressor K-6 to supercritical pressure of 100 bar, condensed, and finally pumped to 150 bar for pipeline transportation.

Instead of supercritical CO₂, the alternative configuration of Fig. 8 is proposed to produce low pressure liquid CO₂. Here, a mixed refrigerant (MR) vapor compression cycle [25] is chosen to provide the process refrigeration requirements (along with the expanded NG of stream 2). The power required by the vapor compression cycle is equal to the work input for compressor K-6 that is near 1.76MW. In this case, compressor K-6 is used to compress the MR (optimal molar composition of 14 % CH₄, 58 % C₂H₆, 1.2 %

C_3H_8 , and 27 % nC_4H_{10}) to 19 bar which is subsequently cooled and condensed as it flows through heat exchangers E-4, E-5, and E-6. Leaving heat exchanger E-6, the high pressure liquid MR is reduced to 13 bar pressure and then rerouted via stream D to the heat exchangers to supply the process refrigeration. As seen in Table 3, use of the configurations of Fig. 7 and Fig. 8 in the developed ERPP results in power generation efficiencies of 70.5 and 70.4 %, respectively. In both cases, the additional work required by the compressor K-6 is responsible for lowering the power generation efficiency below the efficiency of the ERPP using LNG. Detailed material and energy balance, heat exchangers composite curves, and optimization approach are available in the SI.

5. ERPP vs conventional SOFC power plant

The conventional SOFC power generation plant [31], shown in Fig. 9, is rigorously simulated to compare it with the developed ERPP designs. Here, the process does not capture CO_2 and the SOFC anode exhaust is combusted with the depleted air stream leaving the cathode side of the device. The combustion effluent gas is then expanded in gas turbine T-1 to a pressure of 3 bar, cooled by heat recovery in heat exchanger E-1 and E-2 to a temperature of $179^\circ C$ and $94^\circ C$, respectively, and finally expanded to 1.1 bar in turbine T-2. Water is condensed for recycle to the SOFC reformer via pump P-2 and the CO_2 rich gas (stream 13) is discharged to the atmosphere. The process power generation efficiency of the plant is calculated to be 73.0 %, which is slightly more efficient than the ERPP fed with LNG or NG as seen in Table 3. Also, notice the lower SOFC power output compared to the ERPP design. The difference is indicative of the additional power generated in the SOFC due to the recycle stream 48 in the ERPP design. Sensitivity analysis for the power plant of Fig. 9 is presented in the SI.

It is interesting to note that the power penalty associated with the recovery and liquefaction of CO₂ is only 1.8% when compared with ERPP-LNG. This reflects the synergistic nature of the overall ERPP flowsheet.

6. Conclusion

The presented ERPP designs allow for efficient electrical power generation using LNG or NG, coupled with a novel process that achieves ~ 100% CO₂ capture. The following are the unique system integration features of the ERPP designs. 1) If LNG is to be utilized, then its evaporation step is integrated into the CO₂CL section. Here, captured CO₂ is stored as a low pressure liquid, thereby offering flexibility in storage and shipping. If NG is used, then an MR vapor compression cycle is deployed to capture and liquefy the CO₂. 2) Alternatively, if supercritical CO₂ is the preferred means of transportation, then the captured liquid CO₂ may be evaporated to supply the refrigeration need for its capture. This eliminates the need for an MR refrigeration cycle. 3) The use of three stages for capturing CO₂ along with the utilization of liquid phases as reflux streams enables the recovery of near pure CO₂. 4) In all of the above cases, the separated unconverted H₂ and CO are recycled to the SOFC. The calculations show that the recycle stream increases the SOFC power output by 22%. 5) The power and heat integration schemes enable to operate the system without the need to combust portion of the NG to provide process heat. The design power generation efficiency, as identified by rigorous modeling and simulations, is in the range of 70.4 to 76.0% on LHV basis with ~ 100% CO₂ capture. When compared to a SOFC-gas turbine plant that does not capture CO₂, the ERPP achieves the benefit of nearly zero CO₂ emissions with reduced energy penalty.

Glossary

EX_j : Rate of exergy flow associated with a stream j

$W_{Min}^{CO_2CL}$: Minimum power (exergy) for capturing and liquefying CO₂

W_{Max}^{LNG} : Maximum power (exergy) available from vaporizing LNG

Acknowledgments

Research supported by Qatar University, by the Centre for Direct Catalytic Conversion of Biomass to Biofuels, an Energy Frontier Research Centre funded by the U.S. Department of Energy (DOE), Office of Science, Basic Energy Sciences (BES), under Award # DE-SC0000997, by the National Science Foundation Solar Economy IGERT (0903670-DGE) and EFRI (0938033-DGE).

References

- [1] S. Chu, A. Majumdar, *Nature.*, 488 (2012) 294-303.
- [2] S. Pacala, R. Socolow, *Science.*, 305 (2004) 968-972.
- [3] IEA, International Energy Agency, Paris, 2012.
- [4] M.L. Szulczewski, C.W. MacMinn, H.J. Herzog, R. Juanes, *P. Natl. Acad. Sci. USA.*, 109 (2012) 5185-5189.
- [5] BP, British Petroleum, London, 2013.
- [6] REN21, Renewable Energy Policy network for The 21st century (REN21), Paris, 2011.
- [7] E.S. Rubin, H. Zhai, *Environ. Sci. Technol.*, 46 (2012) 3076-3084.
- [8] A. Aspelund, T. Gundersen, *Appl. Energ.*, 86 (2009) 781-792.
- [9] H. Zhai, E.S. Rubin, *Energ. Fuel.*, 27 (2013) 4290-4301.
- [10] IPCC, Intergovernmental Panel on Climate Change, New York, 2005.
- [11] A. Aspelund, M.J. Molnvik, G.D. Koeijer, *Chem. Eng. Res. Design* 84 (2006) 847-855.
- [12] A. Lee, O. Zinaman, J. Logan, National Renewable Energy Laboratory, Golden, CO, 2012.
- [13] EIA, <http://www.eia.gov/>, Washington, DC., (Accessed on June 2011).
- [14] Aspen Technology, ASPEN Plus users guide Aspen Technology Inc., Burlington, Massachusetts, 2006.
- [15] J. Van herle, F. Maréchal, S. Leuenberger, Y. Membrez, O. Bucheli, D. Favrat, *J. Power. Sources.*, 131 (2004) 127-141.

- [16] J. Van herle, F. Maréchal, S. Leuenberger, D. Favrat, *J. Power. Sources.*, 118 (2003) 375-383.
- [17] T.A. Adams, P.I. Barton, *J. Power. Sources.*, 195 (2010) 1971-1983.
- [18] S. Campanari, *J. Power. Sources.*, 112 (2002) 273-289.
- [19] M. Ola, B. Rune, B. Olav, K. H.M., S. Morgen, in: *Greenhouse Gas Control Technologies 7*, Canada, 2005, pp. 273-281.
- [20] L. Duan, K. Huang, X. Zhang, Y. Yang, *Energy.*, 58 (2013) 66-77.
- [21] R. Allam, V. White, N. Ivens, M. Simmonds, in: D.C. Thomas (Ed.), *Carbon Dioxide Capture for Storage in Deep Geologic Formations-Results from the CO₂ Capture Project*, Elsevier, New York, 2005.
- [22] M.B. Wilkinson, J.C. Boden, R.S. Panesar, M. Babcock, R.J. Allam, in: *First National Conference on Carbon Sequestration*, Washington DC, 2001.
- [23] R.S. Gemmen, in: *International Joint Power Generation Conference*, Baltimore, MD, 1998, pp. 23-26.
- [24] A. Aspelund, T. Gundersen, *Appl. Energ.*, 86 (2009) 793-804.
- [25] E.I. Al-musleh, in: *Chemical Engineering*, Purdue Univeristy, IN, USA, 2010.
- [26] N. Zhang, N. Lior, *Energy.*, 31 (2006) 1666-1679.
- [27] N. Zhang, N. Lior, M. Liu, W. Han, *Energy.*, 35 (2010) 1200-1210.
- [28] R. Agrawal, in: *US Patent, patent number US5141543 A*, Air Products and Chemicals, Inc., United States, 1992.
- [29] W. Wong, *P. I. Mech. Eng. A-J. Pow.*, 208 (1994) 3-12.
- [30] J. Szargut, I. Szczygiel, *Energy.*, 34 (2009) 827-837.
- [31] K. Hassmann, *Fuel. Cells.*, 1 (2001) 78-84.
- [32] A.B. Stambouli, E. Traversa, *Renew. Sust. Energ. Rev.*, 6 (2002) 433-455.
- [33] W.D. Seider, J.D. Seader, D.R. Lewin, S. Widagdo, *Product and Process Design Principles*, John Wiley & Sons Ltd., New York, 2009.
- [34] Heatric, http://www.heatric.com/temp_pressure_capabilities.html, (Accessed on August 2013).
- [35] C.F. McDonald, *Appl. Therm. Eng.*, 16 (1996) 3-32.
- [36] W.L. Lundberg, R.A. Israelson, R.R. Moritz, S.E. Veyo, R.A. Holmes, P.R. Zafred, J.E. King, R.E. Kothmann, *Siemens Westinghouse Power Corporation*, Pittsburgh, PA 2000.
- [37] EG>echnicalServices, *US Department of Energy*, Morgantown, West Virginia, 2004.
- [38] T.A. Adams, P.I. Barton, *AIChE. J.*, 56 (2010) 3120-3136.
- [39] J.R. Rostrup-Nielsen, K. Aasberg-Petersen, in: W. Vielstich, H.A. Gasteiger, A. Lamm (Eds.), *Handbook of Fuel Cells-Fundamentals, Technology and Applications*, Wiley, New York, 2003, pp. 160-176.
- [40] R. Allam, O. Bolland, J. Davison, P. Feron, F. Goede, A. Herrera, M. Iijima, D.I. Jansen, I. Leites, P. Mathieu, E. Rubin, D. Simbeck, K. Warmuzinski, M. Wilkinson, R. Williams, in: B. Metz, O. Davidson, H.d. Coninck, M. Loos, L. Meyer (Eds.), *Intergovernmental Panel on Climate Change Special Report on Carbon Dioxide Capture and Storage*, Cambridge University Press, New York, 2005.
- [41] J.M. Campbell, *Gas Conditioning and Processing: The Basic Principles*, 8 ed., John M Campbell and Company, Norman, Oklahoma, 2001.
- [42] J.H. Foglietta, *Hydrocarb. Process.*, 83 (2004) 39-44.

- [43] Hydrocarbon Processing, Gas Processes Handbook 2004, Gulf publishing Company, Houston, Texas, 2004.
- [44] R.N. Maddox, D.J. Morgan, Gas conditioning and processing: Gas treating and sulfur recovery, 4 ed., Campbell Petroleum Series, Norman, Oklahoma, 1998.
- [45] S.A. Barnett, Direct hydrocarbon SOFCs, in: Handbook of Fuel Cells, John Wiley & Sons, Ltd, 2010.
- [46] J.M. Campbell, R.A. Hubbard, Gas Conditioning and Processing: The Equipment Modules, John M Campbell and Company, Norman, Oklahoma, 2001.
- [47] A. Bourji, A. Winstead, CEP, 109 (2013) 35-39.
- [48] N.R. Udengaard, Prepr. Pap.-Am. Chem. Soc., Div. Fuel Chem., 49 (2004) 906-907.
- [49] J.J. Hartvigsen, S. Elangovan, A.C. Khandkar, System design, in: Handbook of Fuel Cells, John Wiley & Sons, Ltd, 2003

Figure Legends

Fig. 1 A conceptual sketch of the ERPP. SOFC = Solid Oxide Fuel Cell. LNG = Liquefied Natural Gas.

Fig. 2 A detailed flowsheet for the ERPP concept of Fig.1 capturing ~ 100% CO₂. Stream data available in SI.

Fig. 3 Detailed configuration of the CO₂CL section of the ERPP design of Fig. 2. Stream data from modeling is available in SI.

Fig. 4 (a) Process composite hot and cold curves for the ERPP design of Fig. 2, including all process streams with a temperature greater than that of the available cold utility (32 °C). (b) Composite hot and cold curves of the heat exchangers E-4, E-5, E-6, and E-7 in the CO₂CL section of Fig. 3.

Fig. 5 Impact of fuel utilization and NG reforming Steam To Carbon (STC) ratio on the power generation efficiency of the ERPP design of Fig. 2.

Fig. 6 (a) Impact of K-2 discharge pressure on the power generation efficiency of the ERPP design of Fig. 2. (b) K-2 discharge pressure effect on the regeneration requirements of molecular sieve adsorber and stream 14 water content in Fig. 2.

Fig. 7 Alternative CO₂CL process configuration in which NG is available from a pipeline at the plant gate and the captured CO₂ is compressed to supercritical conditions for pipeline transportation. Stream data available in SI.

Fig. 8 Alternative CO₂CL process configuration in which NG is available from a pipeline at the plant gate and the captured CO₂ is stored as a low pressure liquid. Stream data available in SI.

Fig. 9 Conventional SOFC power plant without CO₂ capture. Unconverted H₂ and CO in SOFC exhaust gas is combusted in air for additional power generation. Stream data available in SI.

Tables

Table 1 Design basis for ERPP simulation.

Description	Reference	Value/description
Ambient conditions	Basis	1.013 bar (1 atm), 25 °C, 50% relative humidity
Cooling water supply temperature	[33]	32 °C
Expansion minimum pressure	Assumption	1.1 bar
Heat exchangers maximum pressure	[34]	100 bar
Heat exchangers maximum temperature	[34, 35]	800 °C
<u>Pressure drops in heat exchangers</u>		
For boiling/condensation	[33]	0.10 bar
For gases		0.21 bar
For low viscosity liquid		0.34 bar
<u>Heat exchangers min temperature approach</u>		
T < 25 °C	[33]	1 °C
T = 32.2 to 148 °C		11 °C
T > 148 °C		28 °C
Heat exchangers heat leakage	Assumption	0
Storage tanks heat leakage	Assumption	0
<u>SOFC</u>		
Modeling and simulation approach	[15, 16]	
Maximum temperature approach	[36]	270 °C
Maximum temperature	[37]	1,000 °C
DC to AC efficiency	[38]	96%
Maximum pressure	[36]	10 bar
Fuel reforming	[31, 39]	Internal steam reforming
Phase separators pressure drop	Assumption	0.21 bar
Compressor discharge maximum temperature	Assumption	200 °C
Gas turbines maximum inlet temperature	[40]	1,300 °C
Compressor isentropic efficiency	[41]	80 %
Turbines isentropic efficiency	[42]	85 %
Turbines motor efficiency	Assumption	97 %
Adsorber maximum pressure	[43]	120 bar
Carbon dioxide dehydration molecular sieve	[41, 44]	3A

Table 2 Optimization problem formulation for CO₂CL section of Fig. 3.

Minimize	Compression power (K-2,4 and 5)
Subject to	Stream 32 CO ₂ mole % ≥ 99.5 Per pass liquid CO ₂ recovery ≥ 85 mole % E-4,5,6, and 7 minimum temperature approach ≥ 1.1 °C Stream 31 temperature ≥ -55 °C (if J-T is replaced by turbine)
Independent variables	(1) K-2 discharge pressure (10 to 100 bar) (2) K-4 discharge pressure (10 to 100 bar) (3) K-5 discharge pressure (10 to 100 bar) (4) Stream 30 temperature (43 to -55 °C) (5) Stream 36 temperature (43 to -55 °C) (6) Stream 41 temperature (43 to -55 °C)
Method	Aspen Plus SQP [14]

Table 3 Simulation results for the different ERPP designs as well as a comparable design for a conventional SOFC power plant using NG or LNG. The basis for all the power plant designs is a constant NG feed rate of 895 kmol/hr. Negative values indicate power consumption. Positive values indicate power production.

	ERPP-LNG	ERPP-NG Supercritical CO ₂	ERPP-NG Liquid CO ₂	Conventional SOFC without CO ₂ capture
Compressors, MW				
K-1	-27.42	-27.42	-27.42	-26.12
K-2	-1.41	-1.41	-1.41	-
K-3	-0.02	-0.02	-0.02	-
K-4	-3.89	-2.68	-3.98	-
K-5	-3.05	-3.08	-3.09	-
K-6	-	-2.89	-1.76	-
Turbines, MW				
T-1	0.35	0.46	0.46	40.40
T-2	6.81	6.81	6.81	4.80
T-3	9.99	9.99	9.99	-
T-4	0.71	0.71	0.71	-
T-5	1.06	1.12	1.15	-
T-6	0.56	0.51	0.49	-
SOFC AC power, MW	174.33	174.33	174.33	142.81
Pumps, kW				
P-1	-165.7	-	-	-16.20
P-2	-18.91	-18.91	-18.91	-19.84
Net power output, MW	157.83	156.41	156.28	161.85
Feed LHV, MW			221.93	
Power generation efficiency, %	71.2	70.5	70.4	73.0

To sequestration