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Early-Phase Evaluation of CO_2 Recovery at Geothermal Facilities

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ABSTRACT

A study was performed to evaluate process schemes and early-phase economics for the recovery of four different CO_2 products from a non-condensable gas (NCG) stream at a geothermal power plant. The four products include: 1) low-pressure, lower-purity CO_2 for greenhouse gas use, 2) high-pressure, dense-phase CO_2 for enhanced oil recovery applications, 3) a refrigerated-liquid CO_2 product of beverage grade quality, and 4) subsurface reinjection for sequestration of the CO_2-rich NCG for carbon capture credits. The estimate for Product 3 is intended to also apply to other refrigerated liquid CO_2 grades, such as dry ice, food grade, and industrial grade. Although other grades of liquefied CO_2 could possibly be slightly less expensive to produce compared to the beverage grade, the differences are likely within the error margins of this early phase work. Even though the CO_2 is not actually recovered in Product 4, it is considered a "product" due to the value of the carbon-capture credit from avoidance of venting the CO_2 to the atmosphere. The NCG feed to the recovery process in the study was 50 tonne per hour for a hypothetical binary geothermal power plant with high-gas content in the geothermal fluid. The CO_2 content of the NCG feed to the CO_2 recovery process is 98.4 mol% (dry basis). The capital cost, major process operating cost (power and chemicals), and the estimated total treat cost per tonne of CO_2 product (includes power and chemical cost, operating labor, maintenance and G&A, and capital amortization) were estimated for the four products. The removal of other impurities (Hg, COS, and C_2H_6) in the NCG that could impact the CO_2 product recovery process was also considered.

1. Introduction

Geothermal steam usually contains a small amount of non-condensable gases (NCG) often in the range of a half to a few percent by volume. The NCG is a mixture that generally contains carbon dioxide (CO_2) as a major constituent and often significant concentrations of hydrogen sulfide (H_2S), nitrogen (N_2), hydrogen (H_2), ammonia (NH_3), argon (Ar), methane (CH_4), other light hydrocarbons, sulfur species, and possibly many other species. The composition of the NCG varies depending upon the geographical location and the physical conditions in and the rock composition of the geothermal reservoir. Oxygen (O_2) is not a component in the geothermal NCG at reservoir conditions but may be introduced to the NCG in surface installations due to leakage of air into the process under the vacuum conditions common in the final power generation stage of conventional geothermal power plants. However, the ingress of oxygen is not an issue for binary geothermal power plants because the gas is always maintained at positive pressures.

The NCG is vented to the atmosphere in most geothermal power plants, although some pollution abatement steps may be required first; for some cases, it may be possible to compress the NCG and reinject it with condensate/brine (Dippo, 2012). It is common for H_2S, when it is present in NCG at significant concentrations, to be removed from the

NCG before venting to the atmosphere due to its toxicity, odor, and eventual fate in the atmosphere as sulfuric acid (causing acid rain).

In 2015, Trimeric Corporation conducted a study funded by the Energy Sector Management Assistance Program (ESMAP) of The World Bank to evaluate the technical and economic viability of CO_2 capture from NCG, with the intent of converting the captured CO_2 to a commercially useful form (Mamrosh et al., 2015). Trimeric's evaluation was a part of a larger study carried out by the ESMAP team and resulted in a technical report addressing greenhouse gas (GHG) emissions from geothermal power production (ESMAP, 2016). This paper summarizes the results of Trimeric's work for the World Bank. CO_2 capture from binary geothermal power plants with high-NCG-content geothermal fluids was the focus, although the information in this paper may be useful in other situations also. The conversion of the CO_2 to a commercially useful form included the removal of impurities, compression, and/or liquefaction of the CO_2. The final report that Trimeric produced for ESMAP can be found at: http://pubdocs.worldbank.org/pubdocs/publicdoc/2016/2/477131454921995311/Geothermal-CO2-Recovery-Analysis-Final and also at Trimeric's website (www.trimeric.com). The technical report prepared by ESMAP can be found at: http://documents.worldbank.org/curated/en/2016/06/26519830/greenhouse-gases-geothermal-power-production.

In this study, four specific grades of CO_2 were considered as potential products:

- Greenhouse: Low-pressure CO_2 gas suited for addition to greenhouses, which would be transferred to the user via a low-pressure pipeline.
- Enhanced Oil Recovery (EOR): High-pressure CO_2 fluid suited for use in EOR, which would be transferred to the user via a high-pressure pipeline.
- Beverage Grade: High-purity liquid CO_2 suited for food, beverage, dry ice, or general industrial uses. Although the CO_2 specifications for these products are often slightly different, the technologies used to obtain the different grades are usually very similar. The differences in capital and operating costs to produce the different grades are probably not large enough to consider them separately in this early-phase study.
- Reinjection of NCG for Carbon Capture Credits: CO_2 is not actually recovered but instead the NCG is subjected to the minimal treatment necessary for recompression and sequestration by injection into disposal wells. It is assumed that the sequestration of this stream, which would previously have been vented to the atmosphere, would provide economic benefit from carbon capture credits.

The grade of CO_2 product is dependent on the end-user's needs. The overall treatment steps required to reach the desired product will vary depending on the conditions (i.e., composition and flow rate) of the NCG and, thus, need to be evaluated on a case-by-case basis. For example, with some products, other impurities that are present in minor amounts – carbonyl sulfide (COS), ethane (C_2H_6), and mercury (Hg) – must also be taken into account since additional specialized unit operations would be required to remove these components from the CO_2. The impact on the economics for removal of these species is presented in this paper.

This evaluation is early-phase in nature, and only considers the use of commercialized technologies that could reasonably be applied to produce the desired products from the NCG. The economic evaluation includes preliminary estimates for the capital cost, energy usage, and other major operating costs of the technologies based on initial sizing data. Complete process-design-type information was not generated, thus the use of the term "early-phase".

The remainder of this paper presents the design basis for the NCG stream. Conceptual process descriptions and schematics are summarized for the different CO_2 grade products. The approximate major capital and operating costs are given, as well as the total treating costs for the overall processes. The economic impact of the removal of Hg, COS, and C_2H_6 is also discussed.

2. Design Basis

The characteristics of the NCG stream considered in the study are shown in Table 1. The NCG feed to the recovery process is assumed to be at a rate of 50,000 kilograms per hour. This flow rate would correspond to all of the NCG from a 50 MW geothermal power plant emitting 1,000 g/kW-h NCG, or a 100 MW plant emitting 500 g/kW-h. The characteristics of the NCG were based on a hypothetical binary geothermal power plant case characterized by high-gas content in the geothermal fluid, with a very high CO_2 content in the NCG (98.4 mol. %).

The economic and design data that are presented in this paper are applicable to the NCG characteristics shown in Table 1 and the product specifications that were considered, as discussed in the next section. Consideration of the composition and conditions of the NCG stream are key factors in applying the results of this analysis to other geothermal power plants; the results of this study may not necessarily be meaningful for NCG streams with substantially different composition. Later in this paper, the concentrations of Hg, COS, and C_2H_6 are presented and evaluated separately for removal from the CO_2 stream.

Table 1. NCG Stream Conditions and Composition.

Parameter	Value
Flow rate, kg/hr	50,000
Pressure, bar g	4
Temperature, °C	70
Composition (dry basis), vol. %	
CO_2	98.445
H_2S	0.200
N_2	0.500
Ar	0.005
H_2	0.050
CH_4	0.700
NH_3	0.100
H_2O	Saturated at stream temperature and pressure
Total sulfur content, tonne/d	1.76

This economic analysis also assumes that a natural gas pipeline is not located near the CO_2 recovery facility, thus heating cannot be supplied by burning natural gas, and electrical power would be the main source of energy for the processes considered; some facilities may have excess steam and/or other waste heat that could be considered for use in CO_2 recovery processes. The cost of electrical power assumed is $0.105/kW-h. For minor fuel needs, a liquid fuel (e.g., propane) was assumed to be provided at the site via truck delivery.

Note also that the NCG contains flammable species, such as H_2, CH_4, and H_2S. Theoretically, once the species have been separated from the bulk NCG, energy recovery could be considered by burning these components and recovering heat to generate additional steam in a waste heat boiler (Avery et al., 2015). Waste heat recovery was not included in the analyses for this study.

3. Recovery Processes for CO_2 Products

The CO_2 product specifications and technologies selected for manufacturing each product are described in the following subsections. A number of specific technologies are required for each product. Each technology performs a specific task such as the removal of ammonia or H_2S, compression, dehydration, etc. Some of the same technologies are used for more than one CO_2 product. The technologies were selected as being reasonable for the requirements based on the authors' extensive experience in gas treating and CO_2 purification in the geothermal, industrial gas, and enhanced oil recovery industries. Other technologies can be applied to manufacture these products, and optimizing the technology selection was not attempted. More background on technology selection can be found in the full report for the study (Mamrosh et al., 2015).

3.1 CO_2 Gas for Use in Greenhouse

Since a specific, widely accepted purity requirement for the use of CO_2 in greenhouses was not available for this project, it was assumed that any component in the gas should not be present at levels assumed to be unacceptable based on published data from the American Conference of Governmental Industrial Hygienists (ACGIH) 8-hr threshold limit value (TLV) once it is diluted into the greenhouse air. Table 2 shows the TLV values that were used for the study.

Table 2. TLV Values for the Purity Requirement of CO_2 Greenhouse Gas.

Compound	ACGIH 8-hr TLV, ppmv
CO_2	5,000
H_2	Not Applicable
NH_3	25
CH_4	Not Applicable
H_2S	1

The dilution of CO_2 was estimated by assuming that CO_2 will be added to increase the CO_2 concentration in the greenhouse air from 400 ppmv (parts per million by volume) to 1,000 ppmv. Based on this assumption, the only components that must be removed from the CO_2 to meet the purity requirements in Table 2 are H_2S and ammonia. Therefore, H_2S removal is included in the overall process flow scheme for this product, although "deep" removal to less than ~10 ppmv is not required. Likewise, only a limited amount of ammonia removal is needed for the gas to be acceptable for greenhouse use. After the ammonia and hydrogen sulfide have been removed, water removal would also be required to prevent condensation and corrosion in the pipeline between the geothermal facility and the greenhouse.

The conceptual process to produce CO_2 greenhouse gas is shown in Figure 1. Since the NCG has to be cooled for dehydration, the simplest, cost-effective method to remove a significant portion of the ammonia would probably be to cool the NCG; as water condenses from the cooled, saturated NCG, a significant fraction of the ammonia is absorbed into that aqueous liquid phase. Cooling the NCG to 40°C should result in enough water condensing such that the ammonia remaining in the NCG is expected to be reduced to about 40 ppmv. (Note - when diluted into the greenhouse air, the NH_3 concentration will be less than the TLV limit of 25 ppmv.) The condensed liquids are separated and would be stored for reinjection with other process water streams.

The NCG then flows to a liquid redox sulfur recovery (LRSR) unit for H_2S removal. LRSR processes are considered to be economically viable for the removal of H_2S from gases for cases where the amount of H_2S to be removed is in a range of roughly 0.25 to 20 tonne per day on an elemental sulfur basis; the amount of sulfur to be removed from the

NCG for this case is 1.76 tonne/day. The LRSR process shown in Figure 1 uses redox chemistry that converts the H_2S in the NCG directly to elemental (solid) sulfur. Chelated iron is used in the redox chemistry and the iron is continuously regenerated with air and recirculated in the process. The solid sulfur formed in the process is a slurry that is removed as a wet cake using a filter. The LRSR process can reduce H_2S to low levels (e.g., 1 ppmv). The most common applications of LRSR technologies to NCG treatment are for low-pressure gases, such as the case for NCG coming from a vacuum pump or ejector from the last turbine stage of a conventional geothermal power plant (Mamrosh, et. al., 2012). For the application considered in this paper, the LRSR unit will operate at pressure (~5 bara), which is within the capability of the technology, but may be prone to more operating issues due to the high CO_2 concentration and pressure.

The NCG will be saturated with water after treatment by the LRSR unit. The NCG will be transferred to the greenhouses via pipeline, and it is possible that the NCG would cool during the pipeline transfer which would result in the condensation of corrosive, CO_2-saturated water. To prevent the condensation of corrosive water in the pipeline, the NCG flows to a chiller dehydration unit to remove water. A chiller arrangement (to 7.2°C) is used because the dehydration requirement to prevent condensation in the pipeline is assumed to be minimal and the costs are expected to be lower than other alternatives, such as glycol dehydration system. The gas will be transported in an underground pipeline at a ground temperature assumed to be greater than 7.2°C.

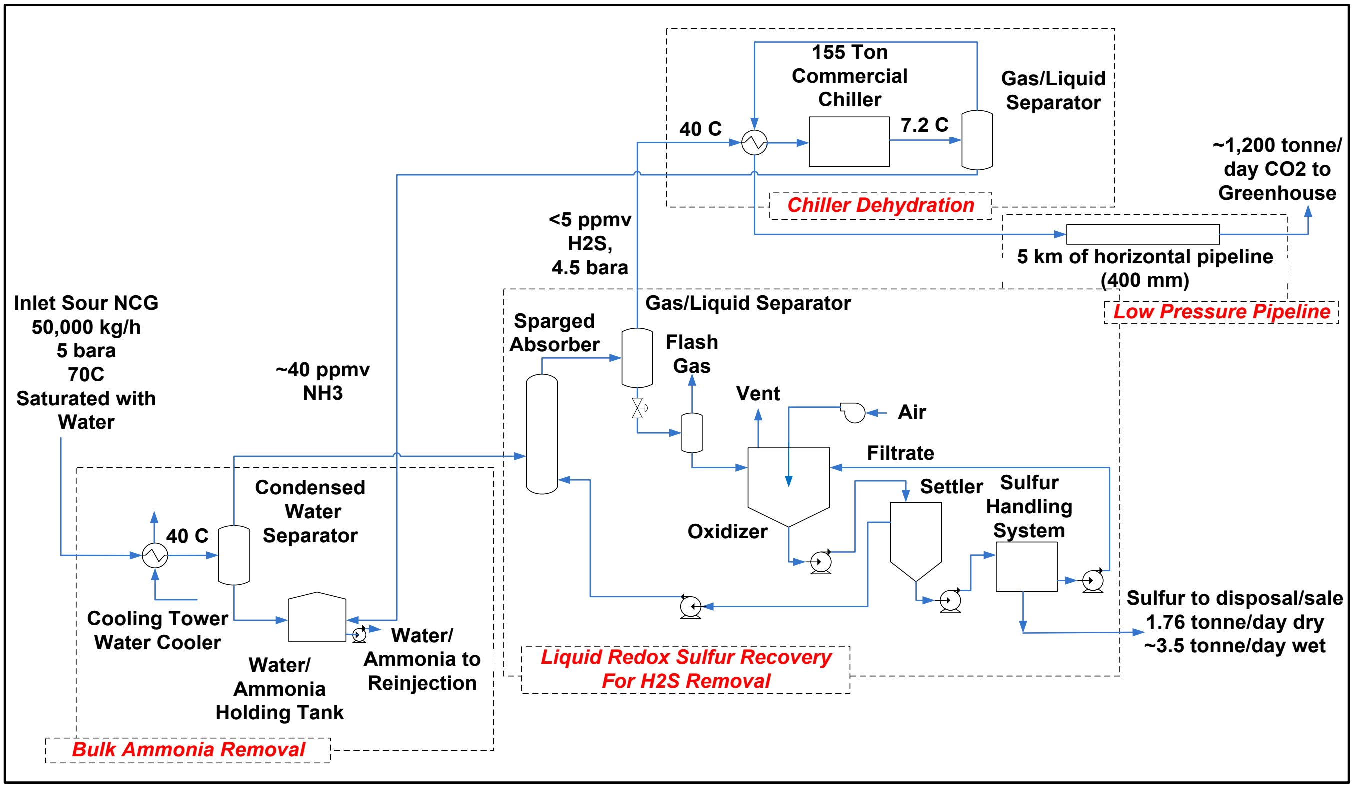

Figure 1. Conceptual Process for Production of CO_2 Gas Product for Use in Greenhouses.

The NCG is then transferred about 5 km using only the pressure available in the gas after dehydration (estimated to be ~ 4.3 bara). For this distance, it was determined to be more economic to build a larger diameter pipeline than to compress the gas, which would allow use of a smaller diameter pipeline. A 400 millimeter diameter pipeline was estimated to be sufficient to allow for the transfer of the ~1,200 tonne/day of CO_2 product to the greenhouse(s). The pressure drop through the line was 1.8 bara, and the inlet and outlet velocities through the pipe are 14.5 and 25 m/s. Elevation changes were not included in the pipeline pressure drop calculations.

3.2 CO_2 for Use in EOR

The composition and conditions listed in Table 3 are thought to be typical for EOR CO_2 since universally-accepted composition

Table 3. Typical CO2 Conditions for EOR.

Component	Value	Level
CO_2, vol. %	95	Minimum
N_2, vol. %	4	Maximum
Hydrocarbon, vol. %	5	Maximum
Water, lb/MMscf (ppmv)	30 (633)	Maximum
O_2, ppmv	10	Maximum
H_2S, ppmv	10-200	Maximum
Glycol, L/Sm^3 gal/MMscf	4.02e-5 0.3	Maximum Maximum
Temperature, °C	50	Maximum

MMscf = millions of standard cubic feet (standard conditions are 60 °F & 1 atmosphere)

Sm^3 = standard cubic meters (standard conditions are 15 °C & 1 atmosphere)

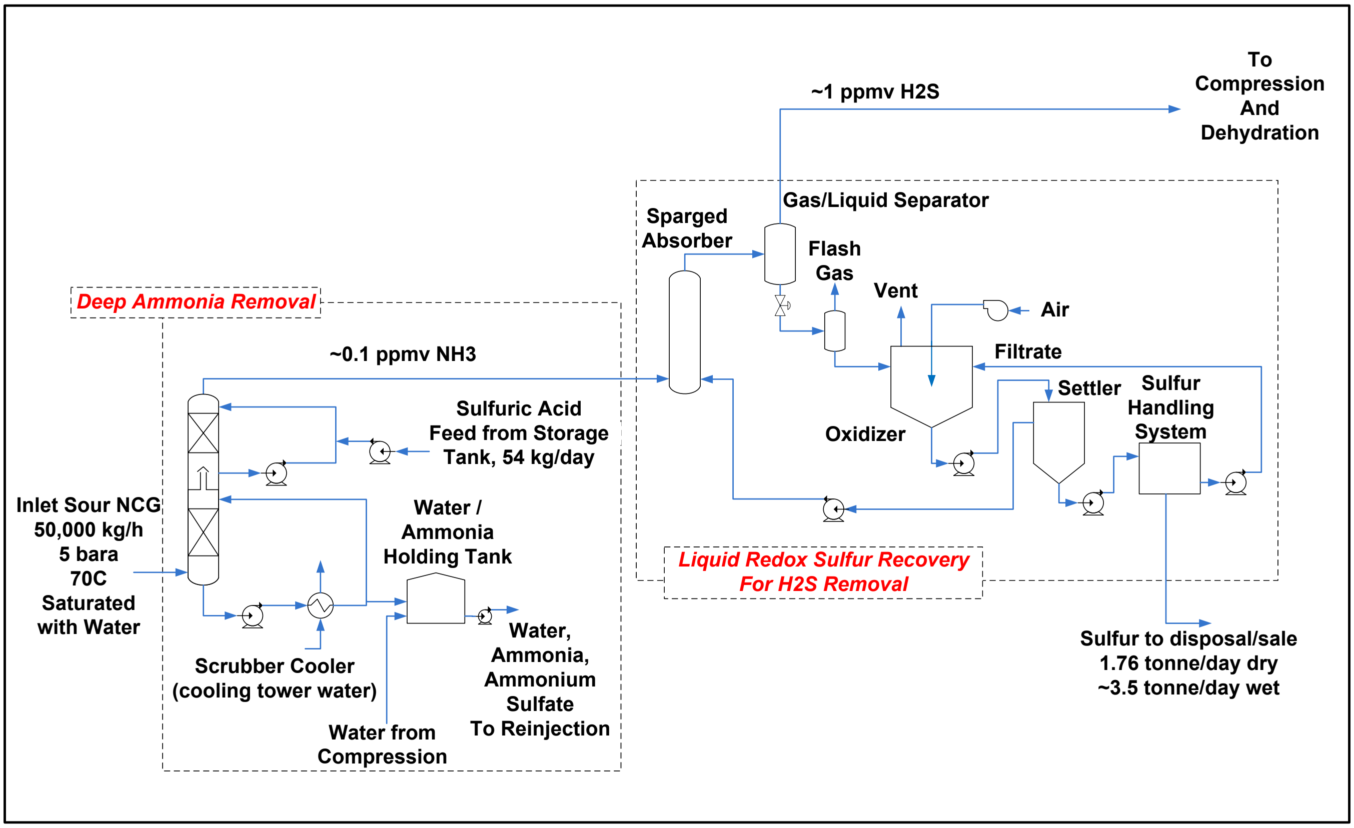

Figure 2. Conceptual Process for Production of CO_2 High-Pressure Fluid for Use in EOR (1 of 2).

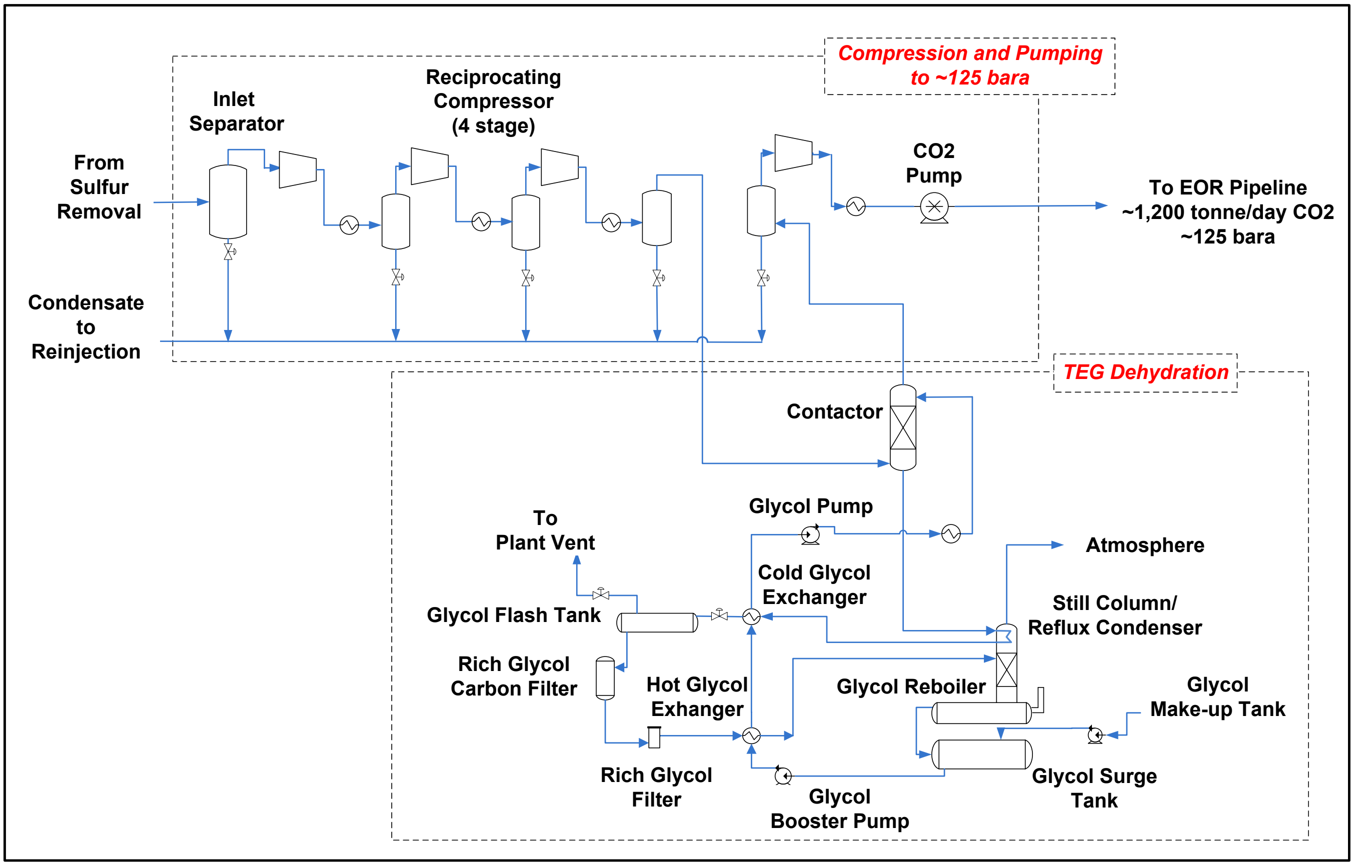

Figure 2. (continued). Conceptual Process for Production of CO_2 High-Pressure Fluid for Use in EOR (2 of 2).

specifications were not available for this project. The product would also be compressed to a pressure typical of EOR CO_2 applications (125 bara); at these conditions, the CO_2 would be a supercritical fluid of relatively high density.

To meet the purity requirements and to avoid problems during compression and/or pipeline transfer, it is assumed that deep ammonia removal, H_2S removal to 1 ppmv, compression, and dehydration are all necessary. Pipeline costs are not included for the EOR case because potential users of EOR CO_2 may be located much farther away than the 5 km assumed with the greenhouse case. Long distance pipelines can be an important factor in the overall economics and should be considered if the use of CO_2 for EOR is further developed.

Figure 2 shows a schematic of a process to produce EOR CO_2. Although the CO_2 product specification for EOR use does not require that ammonia be removed down to low levels, it is known that ammonia present in the bulk CO_2 gas can cause problems during compression (solids formation). When it is present in CO_2, ammonia is typically removed to low levels prior to compression. Thus, it was assumed that deep ammonia removal (from 0.1 vol% down to ~0.1 ppmv) was required. A scrubbing process using sulfuric acid would be used to achieve the deep ammonia removal. Sulfuric acid is a strong acid that is relatively inexpensive and readily available, and should function adequately in this application. Also, the product salt solution (mostly aqueous solution of ammonium sulfate) could probably be mixed with the geothermal power plant condensate/brine stream and disposed of by injection. A two-stage scrubbing system is used to minimize acid usage. The bulk of the ammonia is removed by the condensed and recirculating water in the bottom of the contactor tower. The top section serves as a polishing step for deep ammonia removal with the recirculating sulfuric acid solution. Spent sulfuric acid solution (ammonium sulfate solution) overflows through a chimney tray between the two sections, and is mixed with the water (with ammonium carbonate) in the bottom of the column.

After ammonia scrubbing, the NCG then flows to an LRSR unit for H_2S removal similar to the greenhouse case. From the LRSR unit, the NCG is compressed using a four-stage reciprocating compressor to raise the pressure of the gas to 83.8 bara. After the fourth stage, the CO_2 stream is a dense-phase supercritical fluid, and the density is high enough (~0.57 kg/L) that a multistage centrifugal pump can be used to raise the pressure to 125 bara.

The bulk of the water in the CO_2 is removed from the CO_2 as it is compressed, via interstage coolers, but further dehydration of the CO_2 is required to meet the common pipeline water content specification. Glycol dehydration was selected because it can reach the required water content and is usually less expensive than competing technologies, such as molecular sieves. Triethylene glycol (TEG) absorbs water from the gas in the contactor and then the TEG is regenerated in a still column/reboiler unit by evaporating the water from the TEG. The TEG reboiler is heated using electric resistance heating due to the assumption that natural gas would not be available.

For this case, it is economically and technically advantageous to dehydrate the gas at an intermediate pressure rather than at the compression train feed or discharge pressure. Dehydration at the supercritical pressure after compression would result in excessive glycol losses into the dense-phase CO_2 stream, and dehydration at the low pressure of the compression train feed would be expensive due to the large quantity of water that would need to be removed. This would lead to high glycol circulation rates and high energy requirements for glycol regeneration as well as larger equipment due to the higher flow rate of glycol and the higher volumetric flow rate of gas at low pressure. To minimize dehydration system capital and operating costs, the glycol dehydration unit is located at an intermediate pressure in the process that is closest to the pressure at which the solubility of water in CO_2 is near the minimum (between the third and fourth compression stage for this case).

3.3 Liquefied, Beverage-Grade CO_2

To represent food, beverage, and dry-ice grades of CO_2, a beverage-grade specification (Grade I) found in the literature (Compressed Gas Association, Inc., 2011) was selected. In an early-stage process screening evaluation, there is little difference in the processes that are commonly used to create food, beverage, and dry-ice grades of CO_2, so the more stringent specification was used. Some of the key items from the beverage-grade specification (e.g., limits for components in the NCG composition) are listed in Table 4.

The process is more complex for the production of liquefied, beverage-grade CO_2 compared to the production of greenhouse or EOR CO_2, because nearly all of the impurities must be removed to lower levels. A process to produce beverage-grade CO_2 is shown in Figure 3. Deep ammonia removal is required, so the same ammonia removal process that was specified for the EOR CO_2 process – scrubbing with sulfuric acid – was chosen. An LRSR process alone cannot be relied upon to meet the H_2S removal requirement, because beverage-grade CO_2 requires very deep removal of H_2S (to meet the 0.1 ppmv total sulfur specification). Other technologies, commonly referred to as H_2S scavengers, are often used to meet the very low sulfur specifications for beverage grade

Table 4. Beverage-Grade CO_2 Product Specifications.

Parameter	Value
CO_2 minimum (vol. %)	99.9
Ammonia (ppmv)	2.5
Total sulfur (ppmv)	0.1
Total hydrocarbon content as methane (ppmv)	50
Total non-methane hydrocarbons as methane (ppmv)	20
Water (ppmv)	20

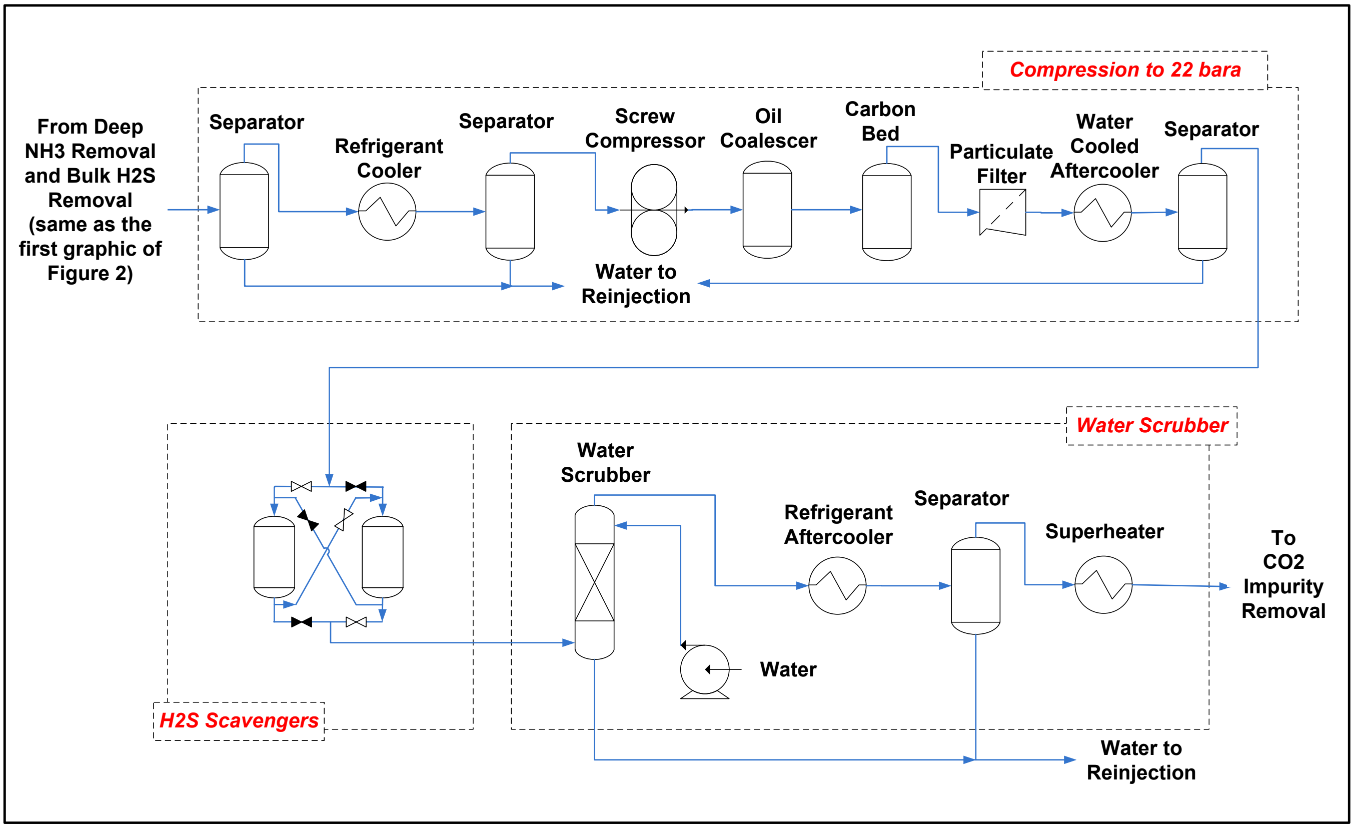

Figure 3. Conceptual Process for Production of Liquefied Beverage-Grade CO_2 (1 of 2). (Deep NH_3 removal and H_2S removal with LRSR will be required initially as shown in Figure 2).

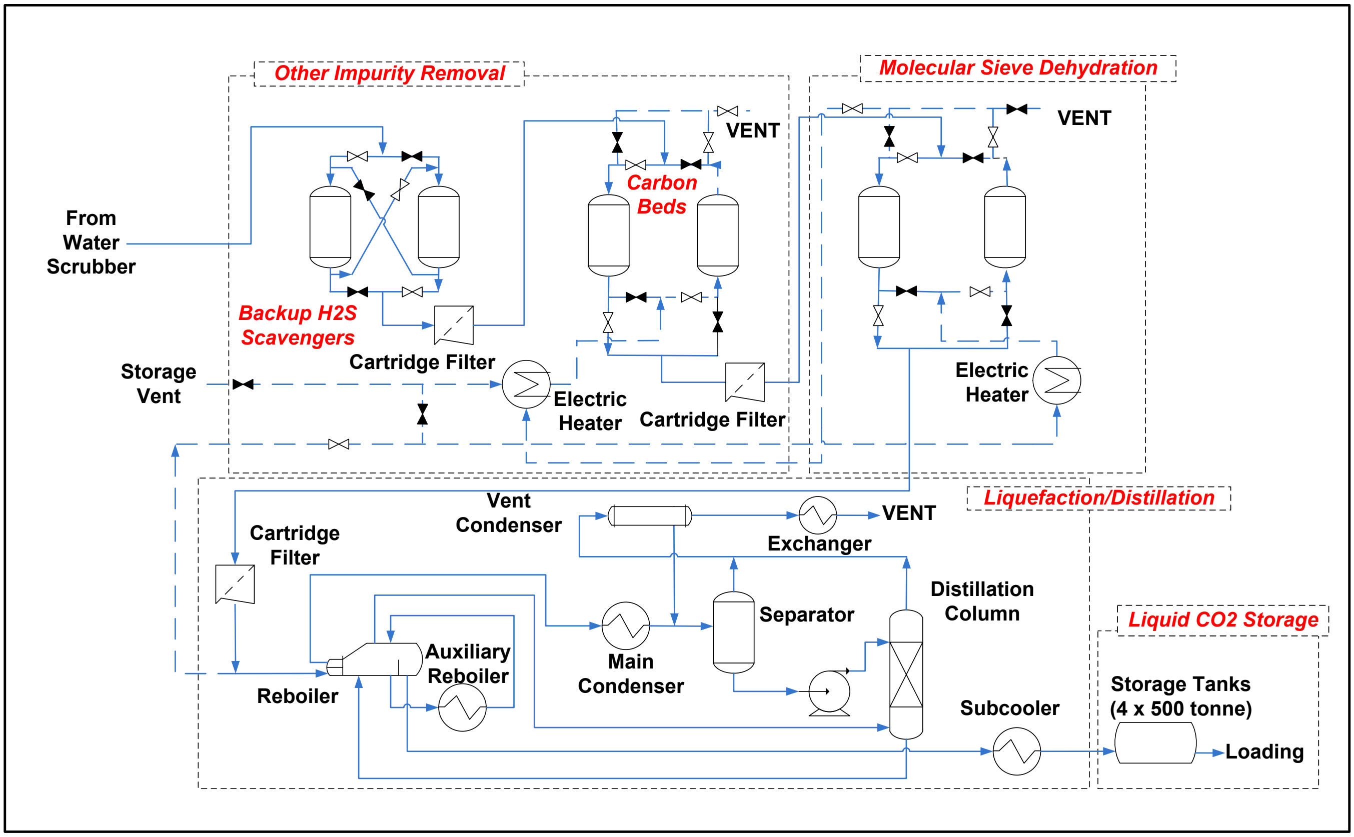

Figure 3. (continued). Conceptual Process for Production of Liquefied Beverage-Grade CO_2 (2 of 2).

CO_2; however, H_2S scavengers are not economical for the removal of the entire amount of H_2S (1.76 tonne/day) that must be removed from this NCG stream. Therefore, a two-step process for H_2S removal is used: an LRSR unit for bulk H_2S removal (to ~1 ppmv) and a solid scavenger unit to remove essentially all of the remaining H_2S (specification is <0.1 ppmv). It is common in the design of beverage-grade CO_2 production systems to include a backup H_2S scavenger system (Fisher & McKaskle, 2006), and this study assumes the presence of a backup scavenger. Figure 3 does not show the deep NH_3 and bulk H_2S removal steps since they are the same as in Figure 2.

As previously discussed regarding the schematic shown in Figure 2, NH_3 is removed first (to ~0.1 ppmv), followed by bulk H_2S removal (to ~1 ppmv). Figure 3 shows the remainder of the beverage-grade CO_2 process. The gas is then compressed to about 22 bara. A conventional (oil-flooded) screw compressor is often most economical for this application, and the power requirement is estimated at 2,240 kW.

The CO_2 gas at about 22 bara then undergoes a series of further purification processes before entering the liquefaction process. The first step is the use of scavenger chemicals to get to less than 0.1 ppmv H_2S. It is assumed that a conventional solid scavenger composed of metal oxides on a solid adsorbent substrate would be used for this application. Solid scavenger systems are typically installed with two or three vessels, sometimes in a lead-lag configuration, with the beds changed out when the absorbent is spent (Fisher, Lundeen, & Leppin, 1999).

After the H_2S scavengers, the gas undergoes once-through water scrubbing to remove small amounts of miscellaneous water-soluble impurities (in the ppm and ppb concentration range) that might be in the gas. The required water feed to the scrubber and the resulting wastewater flowrate from the scrubber are highly dependent on which species are present and on their concentration.

The gas is then contacted with the Backup H_2S scavenger bed (two in lead-lag configuration) to ensure that there is no breakthrough of H_2S above 0.1 ppmv. After the Backup H_2S scavenger beds, carbon beds are used to remove any low-level organic impurities such as benzene and other light hydrocarbon compounds. The carbon beds are regenerable, and electric heaters (266 kW) are used to regenerate the off-line bed.

Since the liquefaction of CO_2 involves low temperatures (typically -26°C), the CO_2 must be dehydrated to very low water concentrations (<1 ppmv), which is well below the treatment specification (20 ppmv), in order to prevent solid CO_2-H_2O hydrate formation and freezing of residual water in the CO_2. Molecular sieves can achieve the lower outlet water concentrations more efficiently than conventional glycol dehydration systems. Two molecular sieve beds are used

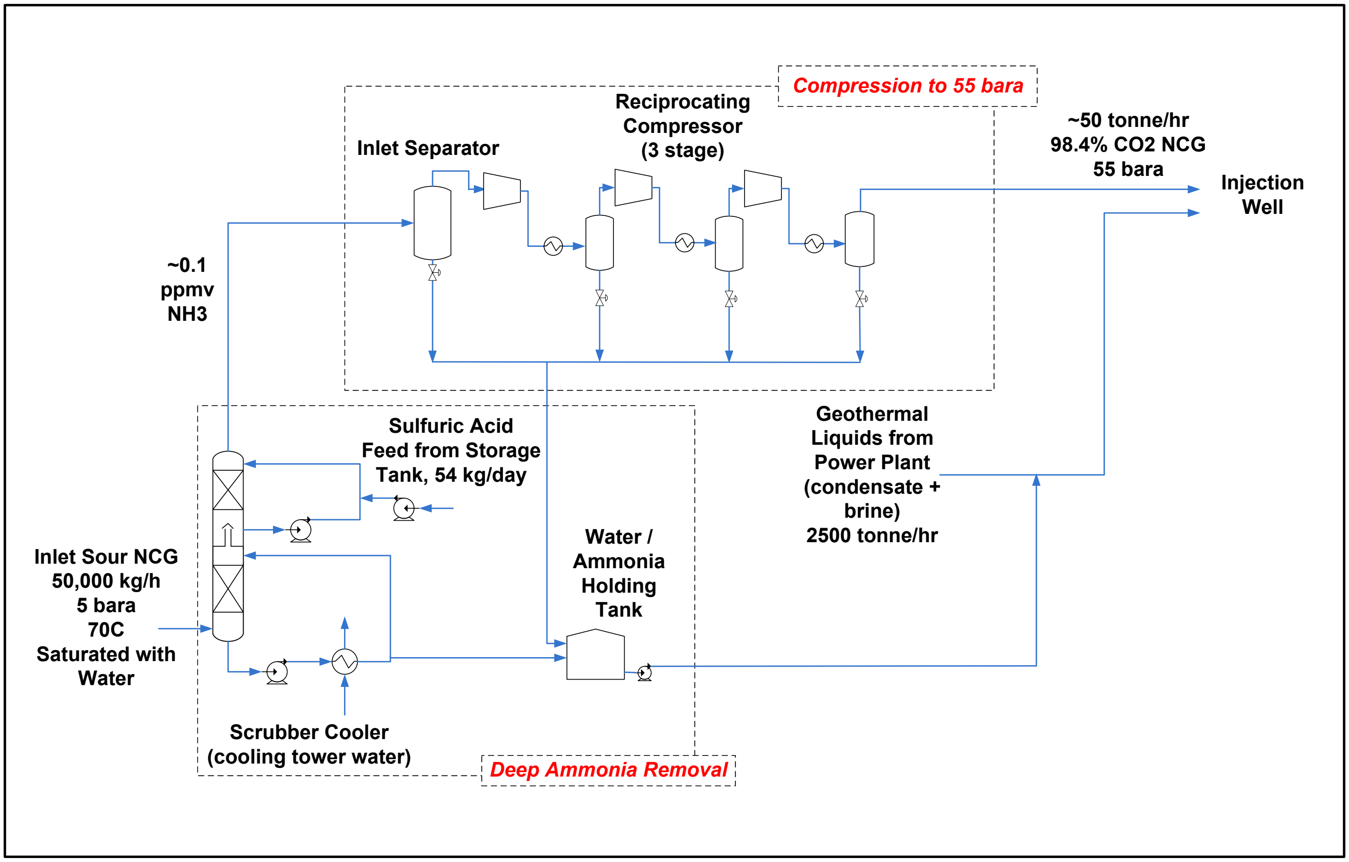

Figure 4. Conceptual Process for Production of Reinjection Grade CO_2.

so that one is on-line while the other is being regenerated. Regeneration occurs by passing a heated dry gas slip-stream (3,400 kg/hr) through the bed. Electric heating (266 kW) is used as the heat source.

The liquefaction/distillation process first liquefies the CO_2 by refrigeration and then purifies the product by distillation. The main impurities removed in the distillation are the gaseous compounds with significantly higher volatility than CO_2 (nitrogen, argon, hydrogen and methane). These impurities are removed from the CO_2 product as part of the overhead gas stream (vented to the atmosphere) from the distillation system.

Finally, the liquefied CO_2 product is often stored on-site until it is transferred by tank car, ship, or tank truck as a refrigerated liquid. Since geothermal power plants are often in remote locations, it was assumed that the product would be shipped by tank truck. Four storage tanks (500 tonne CO_2 capacity each) are used for 1.7 days of storage. Longer storage capacity was not deemed necessary, because the geothermal power plant could stop CO_2 production temporarily and vent the NCG to the atmosphere without impacting operation of the power plant.

3.4 Reinjection of NCG for Carbon Capture Credits.

Figure 4 shows the processing steps to reinject the NCG for carbon capture credits. Because the NCG is mostly CO_2, it is assumed in this case that reinjection of the CO_2 instead of venting it to the atmosphere would have an economic benefit due to carbon capture credits. The stream only requires the minimum extent of processing needed for reinjection since the CO_2 is not recovered as a product. The compressed NCG would be dissolved in the geothermal fluids (condensate plus brine) from the geothermal power plant and disposed of with the fluids in the injection well. Dissolution of the NCG into the fluids requires high pressure, so the main costs for this product are associated with compression of the NCG. The NCG would be compressed to the pressure needed to dissolve it in the liquid, but the dissolution into the liquid would occur at the reservoir depth where the liquid is at the required pressure for dissolution of the NCG due to hydrostatic pressure. Therefore, pumping of the geothermal liquids to the required pressure was not included; rather only the unit operations needed to compress the NCG to the required pressure were included. Given the liquid conditions (flow rate, temperature, and allowable dissolved gas content), a pressure of about 55 bara is required to achieve the dissolution of the NCG into the geothermal liquid stream for reinjection. Prior to compression, deep ammonia removal is required to avoid the formation of solids during compression.

4. Economic Evaluation

Estimates of capital and operating costs were generated for the processes considered based on various sources such as literature information, past experience from work in geothermal, industrial-gas, and enhanced-oil-recovery industries, and capital cost estimation software. The time basis for the costs is the first quarter of 2014, and US dollars are used. The cost estimate data cover only the actual processes used to recover the CO_2, including the cost of major equipment and the cost of installing the major equipment (e.g., piping, electrical, foundations, instrumentation). Items such as land purchase, site preparation, access roads, office and control buildings, and utilities that are not specifically provided in the process design are not included. This type of capital expense is often referred to as inside battery limits (ISBL) costs. The major operating costs include utilities, chemical and other disposable material costs, operations labor, and maintenance and G&A (general and administrative overhead). Labor is based on a $100,000 annual cost per full-time equivalent. Maintenance is estimated at 2% of the total installed cost per year while G&A is 2.5% per year.

Capital amortization was included to estimate the total treating cost. Capital amortization was based on a simple 20-year, straight-line, no-interest method. A 20% contingency was included in the capital costs as well for this early-phase study.

Figure 5 shows the total capital costs estimated on the ISBL basis for each CO_2 product. The capital costs are also shown separately by major processing step (i.e., NH_3 removal, H_2S removal, compression, dehydration, etc.) to demonstrate which steps contribute most to the overall cost. The "other" category for the beverage-grade product includes the costs of compression, impurity removal, dehydration, liquefaction, distillation, and storage. The products have very different costs due to the purity and physical states that are required for the products. The total estimated ISBL project costs for the less expensive products – greenhouse gas (including a 5 km pipeline) and NCG reinjection – are $13.1 and $14.7 million, respectively. The more expensive cases – EOR (for which pipeline costs are not included) and beverage-grade liquid CO_2 product – have estimated total ISBL project costs of $25.3 and $37.8 million, respectively. For the greenhouse product, H_2S removal and the delivery pipeline represent the largest capital costs; if the pipeline cost was not included, the ISBL cost for the greenhouse product would be $8.5 million. The EOR costs were heavily dependent on H_2S removal and the compression/pumping required to achieve a CO_2 pipeline inlet pressure of 125 bara. The major cost for reinjection is compression to 55 bara. Costs for the beverage-grade product were not segregated in the same manner as with the other products; H_2S removal does, however, include treating costs for bulk and deep removal. The cost of ammonia removal is higher for the EOR, beverage-grade, and reinjection processes, compared to the greenhouse process, because deep removal is needed to prevent solids formation during compression.

Figure 5. Capital Cost (Installed ISBL) for CO_2 Product Recovery, Million $US

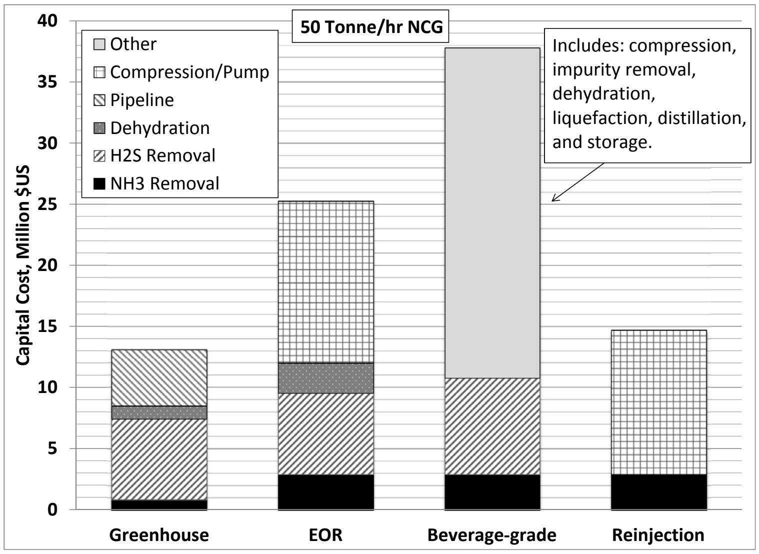

Figure 6 shows the major operating costs for CO_2 product recovery. Major operating costs for the greenhouse case are much lower than the other cases: only $1.4 million per year versus $5.0 million, $6.6 million, and $3.4 million per year for the EOR, beverage-grade, and reinjection products, respectively. As shown in Figure 6, one major reason that the operating costs for the greenhouse case are so much lower is that this process does not require any compression. The operating costs do not include capital amortization. Much of the operating costs are due to electricity usage for compression.

The total treating costs as shown in Figure 7 were estimated by adding the annual amount for capital amortization and the annual operating costs for the CO_2 products and then dividing this sum by the tonne per year of CO_2 production for each option. The total treating cost refers to the cost to recover the CO_2 per tonne of the product. This number is used to evaluate the feasibility of manufacturing the product because the total treating cost can be compared directly to the potential sales value of the product. In lowest to highest order, the total treating costs ($/tonne) are estimated to be $5.00, $10.30, $15.40, and $21.10 for the greenhouse, reinjection, EOR, and beverage-grade products, respectively. Most of the treating costs are related to H_2S removal and/or compression.

It should be noted that the case considered in this study has a high-gas content in the geothermal fluid. For a plant with half the gas but the same power output, the unit cost for most of the CO_2 treatment options would increase slightly. However, the cost of reinjection would actually decrease because the pressure needed to dissolve the gas into the liquid would be lower (half the gas in the same volume of liquid).

5. Removal of Some Other Impurities That Can Be Present in NCG

Geothermal NCG can contain many other components, such as Hg, COS, and C_2H_6, that can have an impact on the CO_2 product recovery processes. For example, Hg removal is potentially required for the greenhouse and liquefied beverage-grade CO_2 products due to the potential safety/health risks; activated carbon beds can be used to remove the mercury to low levels. COS removal is

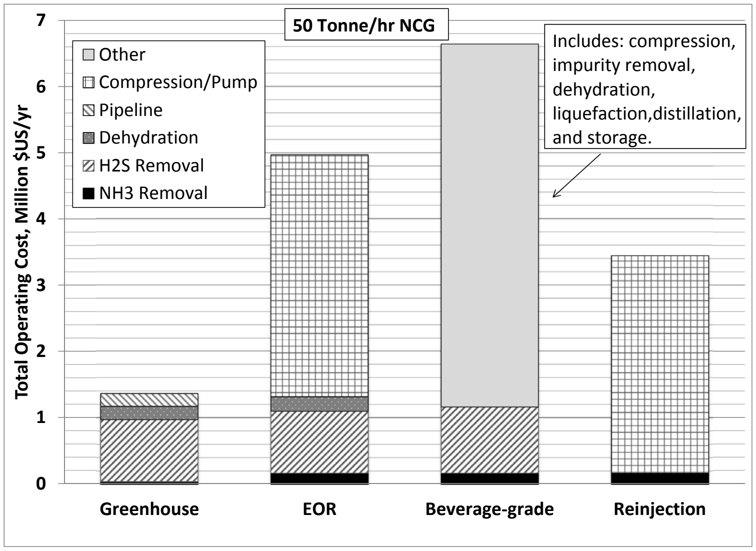

Figure 6. Total Operating Costs for CO_2 Product Recovery, Million $US/yr.

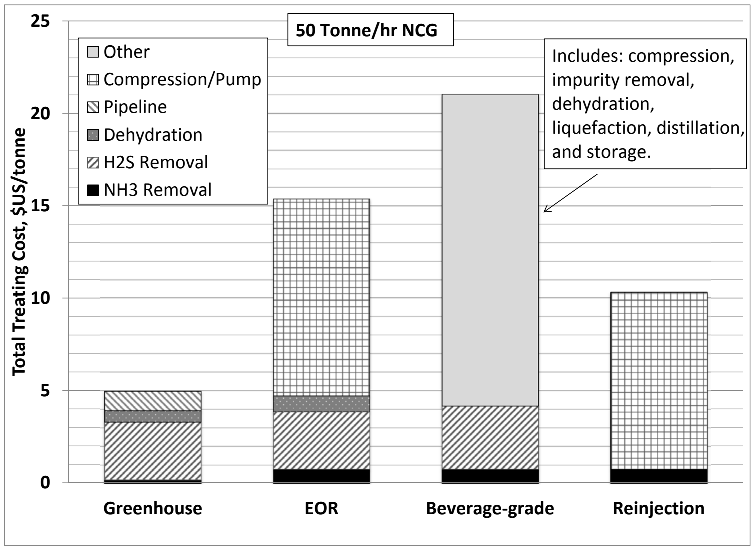

Figure 7. Total Treating Cost of CO_2 Product Recovery, $US/tonne.

likely required for only the beverage-grade product and a catalytic hydrolysis process can be used (Udengaard & Berzins, 1985). Finally, C_2H_6 removal would also be warranted for the beverage-grade product using a catalytic oxidation method. The addition of Hg removal for the greenhouse gas product has only a small impact (~5% increase) on the overall capital and recovery cost of the CO_2. If Hg, COS, and C_2H_6 are all in the NCG, then there is a significant increased cost for the beverage-grade CO_2 product; the capital cost increases by 14% and the total treating cost by 9.5%. Details of these processes can be found in the full report for this work (Mamrosh, et al., 2015).

6. Summary

The processing steps and estimated economics for recovery of four different CO_2 products from the NCG stream of a geothermal power plant were evaluated. The four products include: 1) a low-pressure, lower-purity CO_2 gas that could be transported a short distance and used to supplement CO_2 concentrations in greenhouses, 2) a high-pressure, dense-phase CO_2 fluid of intermediate purity suited for use in enhanced oil recovery (EOR) applications, 3) a refrigerated-liquid CO_2 product of beverage-grade quality, and 4) subsurface reinjection for sequestration of the CO_2-rich NCG stream for presumed carbon capture credits. The economics varied widely based on the purity requirements and physical state of the CO_2. The capital, operating, and total treating costs per tonne of product varied by a factor of about 2.9, 4.9, and 4.3, respectively, when comparing the greenhouse product (least expensive) with the beverage-grade CO_2 product (most expensive). The economics for the EOR and reinjection products differed from each other by a factor of 1.5 to 1.7 with EOR being higher because of the need for H_2S removal and dehydration, and the different injection pressures (125 and 55 bara, respectively). While the estimated costs for recovering these products vary widely, the value of each of the products will also be very different. This screening study serves as a starting point for more detailed, case-specific comparisons of product recovery costs and product values. Treatment for removal of Hg, COS, and C_2H_6 can increase the base case economics by 5 to 14% depending on which of these impurities are present and the grade of CO_2 product produced.

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