Illinois State Geological Survey

Evaluation of CO_2 Capture Options from Ethanol Plants

Topical Report

October 1, 2005–September 30, 2006

Principal Investigator: Robert Finley
Illinois State Geological Survey
(217) 244-8389
finley@isgs.uiuc.edu

Report Issued: October 31, 2006

U.S. DOE Contract: DE-FC26-05NT42588

The Board of Trustees of the University of Illinois
K. L. Williams, Director
c/o Grants & Contracts, Post Award
1901 S. First St., Suite A
Champaign, IL 61820

Illinois State Geological Survey
Kentucky Geological Survey
Indiana Geological Survey
And the MGSC Project Team

Disclaimer

Illinois State Geological Survey

Evaluation of CO_2 Capture Options from Ethanol Plants

Prepared for

Dr. Robert Finley and Dr. Scott Frailey
Illinois State Geological Survey
615 E. Peabody Dr.
Champaign, IL 61820-6964

Prepared by

TRIMERIC CORPORATION
P.O. Box 826
Buda, TX 78610
512-295-8118
www.trimeric.com

Technical Contacts
Kevin Fisher and Ray McKaskle

Contents

Abstract

The Illinois State Geological Survey and the Midwest Geological Sequestration Consortium are conducting CO_2 sequestration and enhanced oil recovery testing at six different sites in the Illinois Basin. The capital and operating costs for equipment to capture and liquefy CO_2 from ethanol plants in the Illinois area were evaluated so that ethanol plants could be considered as an alternate source for CO_2 in the event that successful enhanced oil recovery tests create the need for additional sources of CO_2 in the area.

Estimated equipment and operating costs needed to capture and liquefy 68 metric tonnes/day (75 tons/day) and 272 tonnes/day (300 tons/day) of CO_2 for truck delivery from an ethanol plant are provided. Estimated costs are provided for food/beverage grade CO_2 and also for less purified CO_2 suitable for enhanced oil recovery or sequestration. The report includes preliminary plant and equipment designs and estimates major capital and operating costs for each of the recovery options. Availability of used equipment was assessed.

Executive Summary

The Illinois State Geological Survey and the Midwest Geological Sequestration Consortium are conducting CO_2 sequestration and enhanced oil recovery testing at six different sites in the Illinois Basin. The capital and operating costs for equipment to capture and liquefy CO_2 from ethanol plants in the Illinois area were evaluated so that ethanol plants could be considered as an alternate source for CO_2 in the event that successful enhanced oil recovery tests create the need for additional sources of CO_2 in the area.

Preliminary plant and equipment designs and major operating cost estimates are provided for capture and liquefaction of 68 tonnes/day (75 tons/day) of CO_2 for a local demand and for 272 tonnes/day (300 tons/day), which represents the maximum capture rate for a typical 151,000,000-L/yr (40,000,000-gal/yr) ethanol plant. Estimated costs for food/beverage grade purification are provided for situations where a commercially saleable product is needed and cost estimates for less purified CO2 that could be used for sequestration or enhanced oil recovery are also provided.

Table 1 summarizes the results for each of the recovery plant scenarios. The estimated total installed capital costs to install food/beverage grade CO_2 liquefaction facilities are $2.5 million for a 68-tonnes/day (75-tons/day) facility and $5.8 million for a 272-tonnes/day (300-tons/day) facility. The estimated total installed capital costs to install lower purity CO_2 suitable for enhanced oil recovery or sequestration are $2.1 million for a 68-tonnes/day (75-tons/day) facility and $4.7 million for a 272-tonnes/day (300-tons/day) facility.

Labor costs were not estimated as part of the operating costs. Electricity is the largest operating cost aside from labor. Assuming $0.10/kWh, estimated electrical costs for the food/beverage grade facilities are $19.46/tonne ($17.66/ton) CO_2 produced for the 68-tonnes/day (75-tons/day) facility and $18.18/tonne ($16.50/ton) CO_2 produced for the 272-tonnes/day (300-tons/day) facility. The estimated electrical costs for the lower purity CO_2 facilities are $21.15/tonne ($19.19/ton) CO_2 produced for the 68-tonnes/day (75-tons/day) facility and $18.93/tonne ($17.18/ton) CO_2 produced for the 272-tonnes/day (300-tons/day) facility.

Used equipment searches showed that the used equipment market is limited due to current business conditions in the oil and gas industry. Merchants in the food/beverage grade CO_2 industry may also avoid putting used equipment on the market to prevent increased supply of food/beverage grade CO_2. In future projects, it may be possible to save on the order of 30% for refurbished CO_2 and refrigerant compressors relative to new equipment costs.

Introduction

The Illinois State Geological Survey (ISGS) is conducting pilot tests involving the injection of CO_2 into geologic formations at six different pilots to determine the viability of CO_2 sequestration and enhanced oil and gas recovery in those formations. If the oil reservoir project demonstrates an increase in oil and gas recovery with CO_2 injection in the Illinois Basin, that increase may create the need for additional sources of CO_2 in this area, such as ethanol plants. Trimeric Corporation was asked to evaluate the feasibility of recovering CO_2 from ethanol plants that are planned for the Illinois Basin.

The primary objectives of this study were to determine what process equipment would be required to recover CO_2 from ethanol plants and to estimate the major capital and operating costs associated with CO_2 capture and liquefaction operation. The basis for the study was to produce CO_2 suitable for transport and delivery using tank trucks. This mode of delivery allows flexibility for CO_2 to be sold to industrial consumers or transported to nearby enhanced oil recovery operations where a suitable pipeline network is unavailable. Longer term, large-scale enhanced oil recovery and sequestration operations would likely be supported with a CO_2 pipeline infrastructure. The larger investment in processing equipment required to produce a salable liquid product would not be necessary when CO_2 could be delivered directly to a pipeline for sequestration or enhanced oil recovery.

Equipment Selection and Cost Estimating Approach

Selection for Capacities of 68 tonnes/day (75 tons/day) and 272 tonnes/day (300 tons/day)

This study determines the cost to meet a local demand (e.g., a pilot test) versus the cost to install a full-scale commercially viable facility for 68 metric tonnes/day (75 tons/day) and for 272 tonnes/day (300 tons/day), respectively. The 272 tonnes/day (300 tons/day) is representative of the CO_2 available for recovery at an ethanol plant producing 151,000,000 L/yr (40,000,000 gal/yr). Cost differences in these two facility sizes can be used to compare the cost of capturing the amount of CO_2 required to meet a local demand versus the cost to install a full-scale commercially viable facility for sale of CO_2. Discussions with industry experts and with CO_2 liquefaction system suppliers indicate that the economies of scale and costs of operation result in very few CO_2 liquefaction facilities with a capacity of less than 90.7 tonnes/day (100 tons/day) being built.

CO_2 Recovery and Purification Options

Food or beverage grade CO_2 is not required for enhanced oil and gas recovery or sequestration. However, grassroots CO_2 plants would likely be designed for production of food/beverage grade CO_2 in order to have a broad client base. Trimeric estimated the cost of producing both food/beverage grade and non-food/beverage grade CO_2 in order to understand the incremental cost difference between food/beverage and non-food/beverage grade CO_2. Food grade requires <0.4 ppmv of sulfur, and beverage grade requires <0.1 ppmv of sulfur. The equipment required to recover food and beverage grade CO_2 is the same. Incoming feed gases are carefully managed and blended to produce as much beverage grade CO_2 as possible.

Transportation of non-food/beverage grade CO_2 is likely to include the extra cost of a dedicated delivery trailer, because trailers in food/beverage grade CO_2 service cannot be used for non-food/beverage grade deliveries.

Efforts to Find Used CO_2 Liquefaction Plants and Used Components

Trimeric surveyed several used equipment dealers to determine the availability of used equipment for CO_2 capture. The used equipment market is limited at this time due to the current business conditions in the oil and gas industry. Specific major equipment pieces, such as a CO_2 compressor or a skid-mounted refrigeration system, could possibly be purchased used at the beginning of a project for CO_2 capture at an ethanol plant. Recently similar projects have achieved savings of about 30% from refurbishing and re-engineering compressors compared with the cost of new compressors.

The costs presented in the remainder of this document, however, unless specifically noted, are for new equipment. Additional details regarding searches for used CO_2 capture equipment are provided in Appendix A.

Equipment Required for CO_2 Capture

To a large degree, the type of equipment required for CO_2 recovery does not depend on the recovery rate. Size and cost of the equipment increase with increased recovery rate. There are differences in the equipment and costs associated with purification of CO_2 for food and beverage uses and for non-food and beverage uses. This section discusses the equipment required for CO_2 capture and purification for food/beverage applications and for non-food/beverage grade applications.

Food/Beverage Grade Case

Figure 1 shows the equipment required for CO_2 capture for the food/beverage grade case. Figure 2 shows the corresponding refrigeration equipment required for CO_2 capture for the food/beverage grade case. A description of this equipment follows.

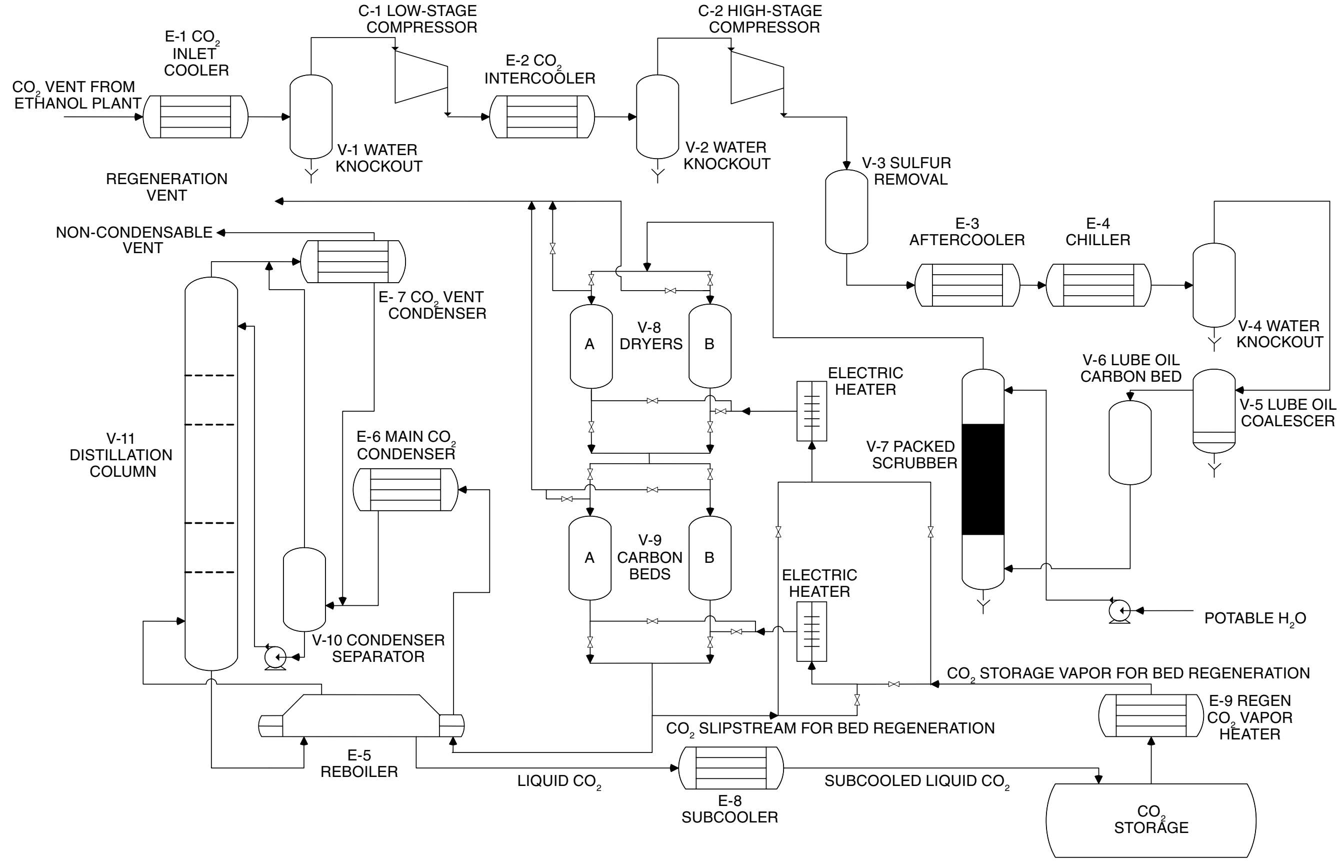

Figure 1 *Food/beverage grade CO_2 process flow diagram.*

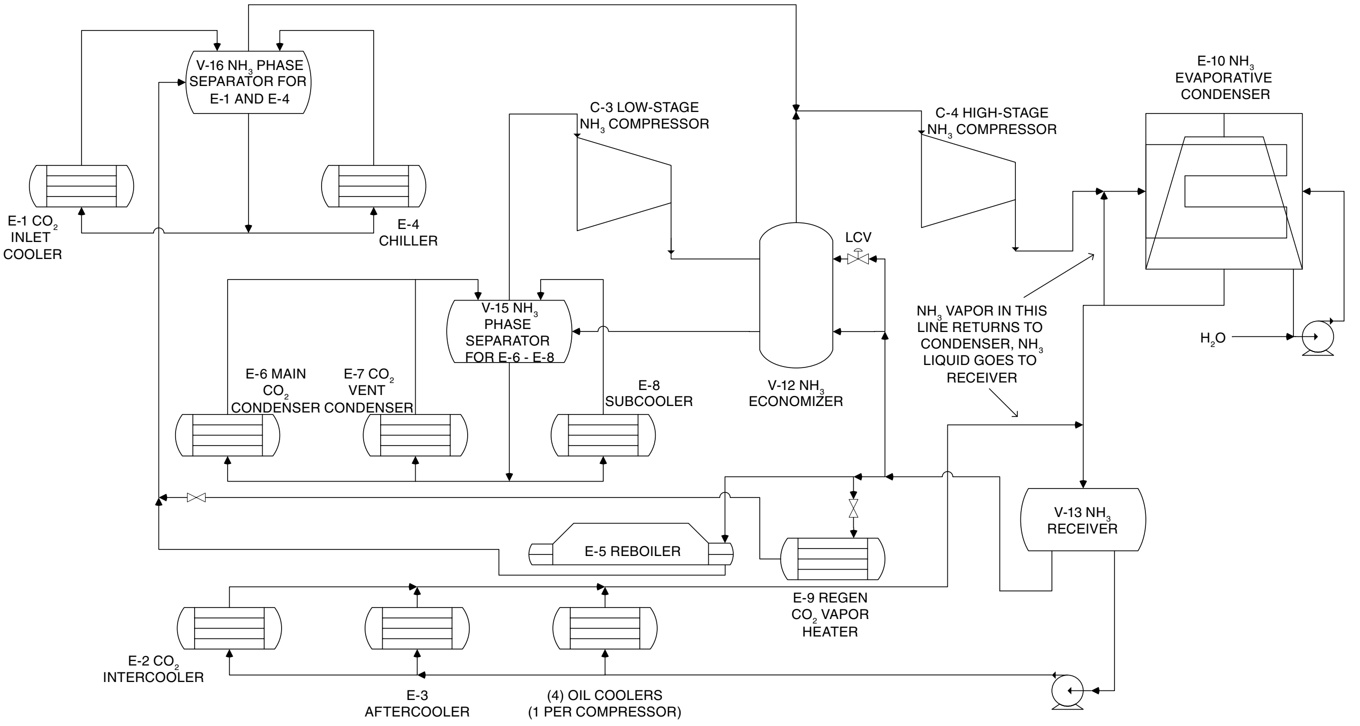

Figure 2 *Food/beverage grade NH_3 refrigeration process flow diagram.*

The sizing and types of units are preliminary and are subject to confirmation after further process engineering. The basic design assumptions, based on previous ethanol CO_2 recovery experience, may require modification after gas analysis on an actual source is performed. Temperatures, pressures, and other parameters in the following description are approximate. Either plant would typically be designed for unattended operation using a programmable logic controller (PLC) unit, or in some cases, a small distributed control system (DCS). Plant operator preference governs this decision.

Lubricant-injected, rotary screw compressors for the main compression services have been selected. In general, screw compressors suit this size of facility and provide lower maintenance costs than reciprocating compressors due to their rotary movement and fewer moving parts. They also offer superior power characteristics at part load and excellent control characteristics. A lubricant management system would be incorporated to ensure an oil-free product.

Inlet design flow rates for the food/beverage grade case are 2,834 kg/hr (6,250 lb/hr) for the facility producing 68 tonnes/day (75 tons/day) and 11,340 kg/hr (25,000 lb/hr) for the facility producing 272 tonnes/day (300 tons/day). For design purposes, the vapor produced by the warming of the stored CO_2 is assumed at an average flow rate of 90.7 kg/hr (200 lb/hr) for the 68-tonne/day (75-ton/day) facility and 227 kg/hr (500 lb/hr) for the 272-tonne/day (300-ton/day) facility. This assumption is to account for heat gain in the storage tanks and the effects of truck loading. Recompression and recycle of these vapors add to the horsepower and electricity requirements for the facilities. The high-stage CO_2 compressor would be used to maintain the pressure in the CO_2 storage tanks.

The plant is designed to accept the CO_2 from the source at 0.987 atm (14.5 psia) and 27°C (80°F). An assumption is that an inlet separator for incoming gas is not required. This assumption would be reviewed after further design discussions. The incoming CO_2 enters a refrigerated shell and tube exchanger (E-1) and is cooled to lower the water content and volumetric flow rate and to prevent condensation in the low-stage CO_2 compressor (C-1). The condensed water is separated in a phase separator (V-1), and the gas is then compressed to 5.1 atm (75 psia) in the low-stage CO_2 compressor (C-1). The CO_2 discharge gas is then cooled in the intercooler (E-2) and enters a second phase separator (V-2).

The gas then enters the high-stage CO_2 compressor (C-2) for compression to approximately 20.4 atm (300 psia). The gas stream then enters the sulfur removal bed (V-3) and then flows through an aftercooler (E-3), a chiller (E-4), and a separator (V-4). The aftercooler and chiller exchangers are arranged in series to achieve the required cooling after compression as economically as possible by utilizing cooling water or high-temperature refrigerant in the aftercooler, followed by lower (intermediate) temperature refrigerant in the chiller. The stream then enters the secondary coalescer lubricant separator (V-5), removing any entrained lubricant before entering a single bed of carbon adsorbent (V-6). The adsorbent bed is sized for 1 yr, operating at design conditions.

The CO_2 then enters a packed bed scrubber (V-7) at approximately 37.8°C (100°F). The scrubbing water is required to be fresh, clean, potable, and odor-free for removal of the water-soluble components.

Next, the CO_2 enters the dryer units (V-8 A and B), where the dew point is lowered to specification. The dryers are arranged for a nominal 12-hr on-line adsorption cycle, with a regeneration cycle of approximately 8 to 10 hr. A slipstream of the primary CO_2 compressed vapor stream is used for dryer unit regeneration. The dryer regeneration gas system is set up so that a gas source is always available. The backup gas source, CO_2 vent vapors from the liquid CO_2 storage tanks, can be manually selected.

Returning to the main gas stream after drying, the CO_2 flows through carbon bed units (V-9 A and B) for final cleanup. The beds are designed for a minimum 24-hr adsorption cycle with a 16-hr regeneration cycle. The regeneration cycle is the same as used for the dryers. The main CO_2 stream then flows to the reboiler (E-5), providing reboiling heat by cooling the main gas stream. The main CO_2 then enters the main CO_2 condenser (E-6) where most of the vapor is condensed. The resulting effluent from the main condenser is mixed with condensate from the distillation column vent condenser (E-7) and flows to the condenser separator (V-10). From this vessel, the liquid CO_2 is pumped to the distillation column (V-11). Non-condensable vapor from the distillation column is vented to atmosphere. The liquid flows down the distillation column, countercurrent to the stripping vapor generated in the reboiler. Note that the reboiler contains two heating bundles, using the heat in the CO_2 main stream for part of the reboiling duty. The balance is provided by the liquid ammonia flow to the main CO_2 condenser.

The main CO_2 liquid stream, having been purified in the column and reboiler, then exits to the subcooler (E-8). The subcooler cools the liquid stream to storage conditions, and the liquid CO_2 flows to the storage unit. Heat exchanger E-9 is used to heat vapors from the CO_2 storage tank when vapors from the CO_2 storage tank are used to regenerate the dryers (V-8) and the carbon beds (V-9) instead of using a slipstream of the main CO_2 gas stream leaving the carbon beds. Ammonia from the NH_3 receiver is used to heat the vapors from the CO_2 storage tank in exchanger E-9 as shown in Figure 2.

The 272-tonne/day (300-ton/day) refrigeration system shown in Figure 2 utilizes a conventional two-stage type system; a low-stage compressor (C-3) provides compression for the required refrigeration on the main CO_2 condenser and subcooler. The second-stage machine (C-4) provides the high-stage compression for the low stage, along with the high-temperature loads. The ammonia refrigeration system for the 68-tonne/day (75-ton/day) plant could be an economized type using a single compressor with a side port for the refrigeration cycle.

The ammonia condenser (E-10) is an evaporative type. The ammonia receiver (V-13) is sized to hold the entire charge of ammonia for pump down. Two small pumps (each 100% duty) are installed on the receiver for ammonia coolant recirculation.

The refrigerant ammonia is at temperatures of approximately 7.2°C (45°F) and –31.1°C (–24°F) with a condensing temperature of 35°C (95°F). A circulation system using ammonia at 35°C (95°F) is also used for the high-temperature cooling. Using the ammonia as the higher-temperature coolant instead of cooling tower water eliminates fouling in the exchangers that are normally water-cooled and allows the use of fixed-tube bundle type units instead of removable bundle types.

The use of pumped ammonia for high-level cooling in the plant eliminates the use of water in all exchangers except the main ammonia condenser. This non-fouling system is used in many different plants and has been highly satisfactory to the users. Exchanger cleaning is essentially eliminated in all plant exchangers except the main ammonia condenser. Although this system does require slightly more surface area in the main ammonia condenser, the lowered maintenance on the other units quickly pays for the additional surface area cost.

As an alternate feature, the ammonia high-stage compressor may be set up so that, when the plant is down, it operates as the CO_2 storage tank condenser compressor. This is accomplished by using the main CO_2 condenser as the storage tank condenser and the column pump to return the liquid to storage. Minimal valving changes are required for changeover and could be automated if desired. The compressor would cycle on storage system pressure.

Additional details regarding equipment required for CO_2 capture for the food/beverage grade case are provided in Appendix B. Information provided in Appendix B includes preliminary equipment sizes and details, consumable requirements, electrical requirements, other utility requirements, feed and product stream composition, and applicable equipment design standards.

Non-Food/Beverage Grade Case

Food/beverage grade CO_2 is not required for enhanced oil and gas recovery and sequestration. Lower-purity CO_2 from natural or industrial sources is used in other regions for enhanced oil recovery. Eliminating some of the equipment required for food/beverage grade CO_2 purification reduces capital costs. These savings are somewhat offset by the requirement to process approximately 15% more CO_2, due to higher vent gas flashing losses in the non-food/beverage grade purification process, which results in higher horsepower and electricity requirements. Figure 3 shows the equipment required for CO_2

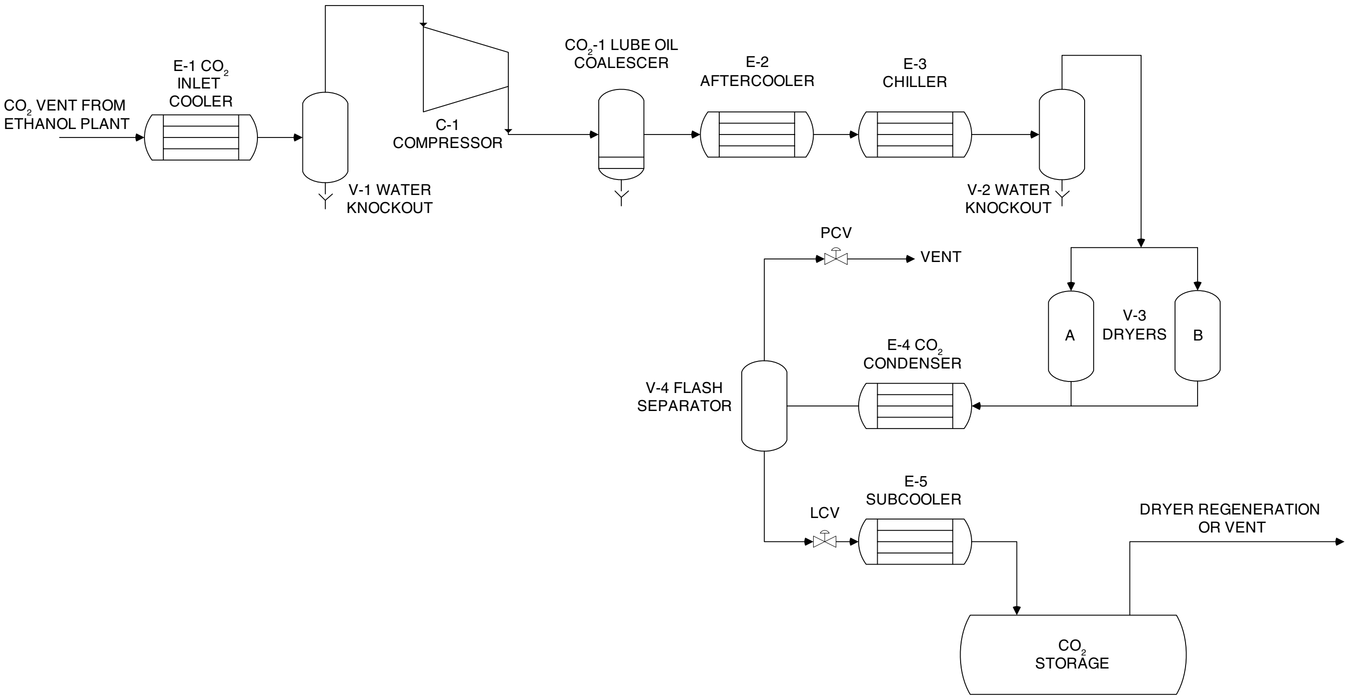

Figure 3 *Non-food/beverage grade* CO_2 *process flow diagram.*

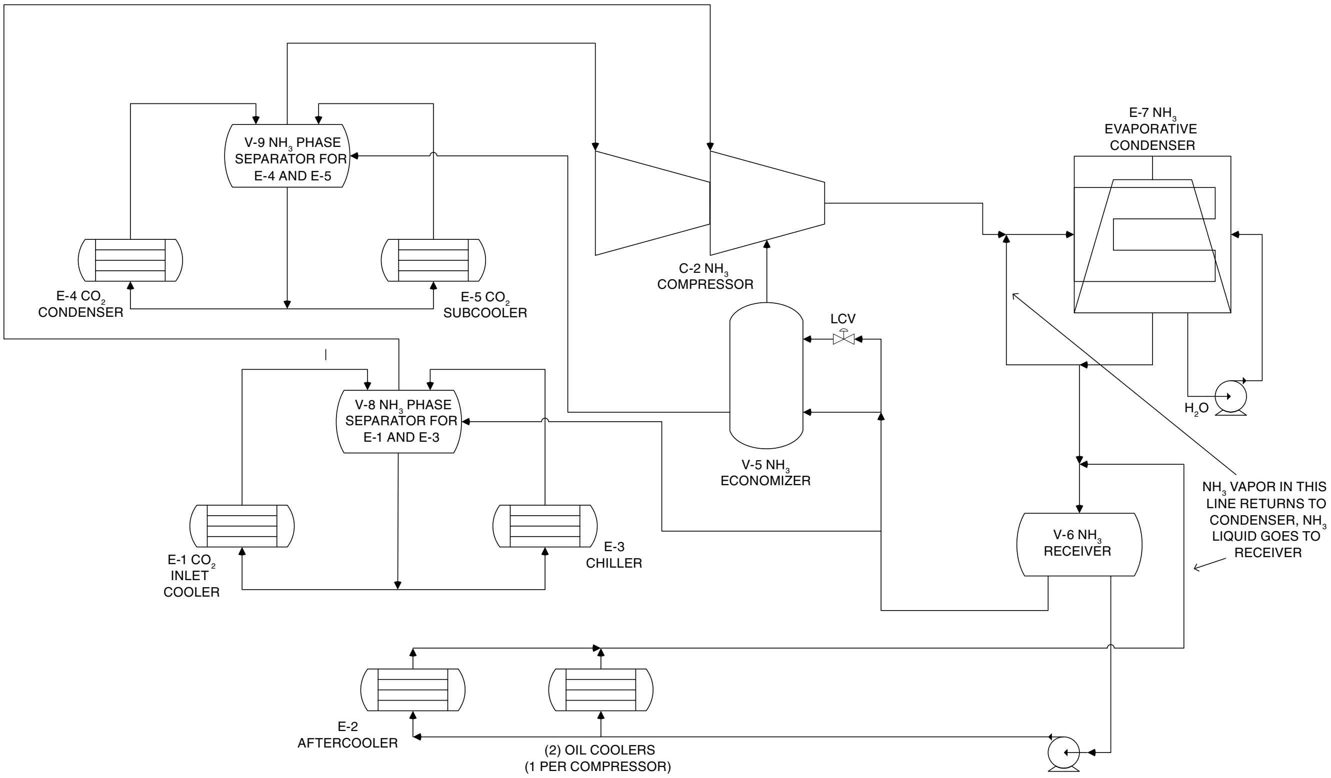

Figure 4 *Non-food/beverage grade* NH_3 *refrigeration process flow diagram.*

capture for the non-food/beverage grade case. Figure 4 shows the corresponding refrigeration equipment required for the non-food/beverage grade case.

The sizing and types of units are preliminary and would be subject to confirmation after further process engineering. The basic design assumptions are based on previous ethanol CO_2 recovery experience and may require modification after gas analysis on the actual source. Temperatures, pressures, and other parameters in the following description are approximate. Either plant would typically be designed for unattended operation using a PLC unit, or in some cases, a small DCS. Plant operator preference would govern this decision.

Lubricant-injected, compound (two-stage) rotary screw compressors have been selected for the main compression services. These models consist of two casings assembled as a single unit, thus requiring only one lube system.

Fifteen percent has been added to the flows of 68 and 272 tonnes/day (75 and 300 tons/day), making the design flows 3,266 kg/hr (7,200 lbs/hr) and 13,041 kg/hr (28,750 lb/hr), respectively for the two units. This increase in flow is a preliminary estimate of the CO_2 lost by the flash of the liquid after liquefaction. In this system, the ammonia refrigerant is at temperatures of approximately 7.2°C (45°F) and –31.1°C (–24°F). The condensing temperature is 35°C (95°F). Tower water or an evaporative condenser may be used for high-level cooling and ammonia condensing service.

The plant is designed to accept the CO_2 from the source at 0.987 atm (14.5 psia) and 27°C (80°F). It is assumed that an inlet separator for the incoming gas is not required, but this assumption will be reviewed after further discussion. The incoming CO_2 enters a refrigerated shell and tube exchanger (E-1) and is cooled to lower the water content and the volumetric flow rate and to prevent condensation in the CO_2 compressor. The condensed water is separated in a phase separator (V-1), and the gas is then compressed to 20.4 atm (300 psia) in the two-stage CO_2 compressor (C-1).

The stream then enters the secondary coalescer lubricant separator (CO_2-1), lowering the lube oil content in the gas to approximately 1 mg/L (1 ppmw). The gas stream then flows through an aftercooler (E-2), a chiller (E-3), and another separator (V-2). Aftercooler and chiller exchangers are arranged in series to achieve the required cooling after compression as economically as possible by utilizing cooling water or high-temperature refrigerant in the aftercooler followed by lower (intermediate) temperature refrigerant in the chiller.

The CO_2 then enters the dryer units (V-3 A and B), where the dew point is lowered to specification. The dryers are arranged for a nominal 12-hr on-line adsorption cycle with an approximate 8- to 10-hr regeneration cycle. A slipstream of the primary CO_2 compressed vapor stream is used for dryer unit regeneration. The dryer regeneration gas system is set up so that a gas source is always available and includes manual selection of the backup gas source. The backup source is CO_2 vent vapors from the liquid CO_2 storage tanks. For simplicity, the regeneration streams have been omitted from Figure 3. Refer to Figure 1, the food/beverage grade CO_2 process flow diagram, for additional details on the routing of the primary and backup dryer regeneration streams.

Returning to the main gas stream after drying, the CO_2 then enters the main CO_2 condenser (E-4), where most of the vapor is condensed. The condensed liquid flows to the flash separator (V-4), where flash vapor is vented to the atmosphere.

The main CO_2 liquid stream then exits to the subcooler (E-5), which cools the liquid stream to storage conditions, and the liquid flows to the storage unit.

The ammonia refrigeration system for the 68-tonne/day (75-ton/day) plant is an economized type, using a single compressor with a side port for the refrigeration cycle. The 272-tonne/day (300-ton/day) plant utilizes a conventional two-stage type system with a compound refrigeration compressor similar to the CO_2 compressor providing refrigeration on the main CO_2 condenser and subcooler. The ammonia condenser (E-7) may be a shell and tube or an evaporative type if cooling tower water is in short supply. A shell and coil subcooler/economizer (V-5) would be utilized. The ammonia receiver (V-6) is sized to hold the entire charge of ammonia for pump down. Two small pumps (each 100% duty) are installed on the receiver for ammonia coolant recirculation.

The use of pumped ammonia for high-level cooling in the plant eliminates the use of water in all exchangers except the main ammonia condenser. This non-fouling system is used in many different plants and has been highly satisfactory to the users. Exchanger cleaning is essentially eliminated in all of the plant exchangers except the main ammonia condenser. Although this system requires slightly more surface area in the main ammonia condenser, the lowering of maintenance on the other units quickly pays for the additional capital cost for the larger ammonia condenser.

A CO_2 storage tank vapor condenser that uses a small reciprocating compressor could be installed if required but is not included at this time. Typically, the vapor from this type of system is not recovered, as the higher inert content makes vapor more difficult to condense.

Additional details regarding equipment required for CO_2 capture for the non-food/beverage grade case are provided in Appendix C. Information provided in Appendix C includes preliminary equipment sizes and details, consumable requirements, electrical requirements, other utility requirements, feed and product stream composition, and applicable equipment design standards.

Cost Estimate Summary for CO_2 Recovery Equipment

Budgetary cost estimates for new equipment for 68-tonne/day (75-ton/day) and 272-tonne/day (300-ton/day) CO_2 capture from an ethanol plant for food/beverage grade and non-food/beverage grade cases are provided in this section. Electrical cost is a critical factor in the economic viability of CO_2 capture. Thus, electrical costs are given the same level of importance as capital equipment costs in this section. An assumed cost of $0.10/kWh was used in this economic analysis. Details regarding electrical requirements and other consumable materials are provided in Appendix B for the food/beverage grade case and Appendix C for the non-food/beverage grade case.

Food/Beverage Grade Case

The estimated cost is $1,200,000 for the 68-tonne/day (75-ton/day) CO_2 capture and food/beverage grade purification equipment (Appendix B) plus a $40,000 freight allowance. This total includes the estimates of $80,000 for the CO_2 compressors and $72,000 for the NH_3 refrigeration compressors. Storage, as described in the following equipment list, would be an additional $340,000. The other equipment included in this cost estimate that would be necessary for 68-tonne/day (75-ton/day) CO_2 capture and food/beverage grade purification is as follows:

- Truck scale
- Metal building approximately 12 m × 21 m (40 ft × 70 ft) with 5.5-m (18-ft) eave height, control room, manager's office, and driver area
- Electrical gear, including motor starters and associated switch gear
- Labor and materials to install and place the plant in service
- Three days of storage in two units: use two 90.7-tonne (100-ton) or 109-tonne (120-ton) capacity units

The cost of this installation is estimated at $950,000, which, with the $1,240,000 in equipment and $340,000 in storage, provides a total installed equipment cost of $2,530,000, or $37,174/tonne ($33,733/ton) of daily capacity. The estimated cost of installation is based on discussions with companies that have built CO_2 liquefaction facilities during the past 2 yr.

The estimated cost of the 272-tonne/day (300-ton/day) CO_2 capture and food/beverage grade purification equipment that is listed in Appendix B is $2,850,000 plus a $100,000 freight allowance. This total includes the estimates of $225,000 for the CO_2 compressors and $185,000 for the NH_3 refrigeration compressors. Storage, as described in the following equipment list, would be an additional $1,070,000. The other equipment included in this cost estimate that would be necessary for 272-tonne/day (300-ton/day) CO_2 capture and food/beverage grade purification is as follows:

- Truck scale
- Metal building approximately 18 m × 37 m (60 ft × 120 ft) with 6.7-m (22-ft) eave height, control room, manager's office, and driver area
- Electrical gear, including transformers, motor starters, and associated switch gear
- CO_2 pipeline (inside plant limits)
- Labor and materials to install and place the plant in service
- Three days of storage in at least two units: use two 454-tonne (500-ton) capacity units

The cost of this installation is estimated at approximately $1,750,000, which, with the $2,950,000 in equipment and $1,070,000 in storage cost, provides a total installed equipment cost of $5,770,000, or $21,195/tonne ($19,233/ton) of daily capacity. The estimated cost of installation is based on discussions with companies that have built CO_2 liquefaction facilities during the past 2 yr.

Please note that these costs may vary considerably depending on union versus non-union labor, site condition and suitability, contractor availability, distance from source, and other site-specific items.

Non-Food/Beverage Grade Case

The estimated cost of the 68-tonne/day (75-ton/day) CO_2 capture and non-food/beverage grade purification equipment as listed in Appendix C is $800,000 plus a $30,000 freight allowance. This total includes the cost of compressors, estimated at $80,000 for the CO_2 compressors and $72,000 for the NH_3 refrigeration compressors. Storage, as described in the following equipment list, would be an additional $340,000. A dedicated trailer would have to be purchased for non-food/beverage grade CO_2 transportation. Trailers used for food/beverage grade CO_2 transportation could not also be used for non-food/beverage grade CO_2 transportation. The cost for one delivery trailer, estimated at $100,000, is added to each of the non-food/beverage grade equipment cost estimates. The other equipment included in this cost estimate that would be necessary for 68-tonne/day (75-ton/day) capture and non-food/beverage grade purification is as follows:

- Truck scale
- Transportation trailer
- Metal building approximately 12 m × 18 m (40 ft × 60 ft) with 5.5-m (18-ft) eave height, control room, manager's office, and driver area

- Electrical gear, including motor starters and associated switch gear
- Labor and materials to install and place the plant in service
- Three days of storage in two units: use two 90.7-tonne (100-ton) or 109-tonne (120-ton) capacity units

Cost of this installation is estimated at $825,000, which, with the $930,000 in equipment and $340,000 in storage, provides a total estimated cost of $2,095,000, or a cost of $30,782/tonne ($27,933/ton) of daily capacity. The estimated cost of installation is based on discussions with companies that have built CO_2 liquefaction facilities during the past 2 yr.

The estimated cost of the 272-tonne/day (300-ton/day) CO_2 capture and non-food/beverage grade purification equipment as listed in Appendix C is $2,100,000 plus an $80,000 freight allowance. This total includes the estimates of $225,000 for the CO_2 compressors and $185,000 for the NH_3 refrigeration compressors. A dedicated trailer would have to be purchased for non-food/beverage grade CO_2 transportation. Trailers used for food/beverage grade CO_2 transportation could not also be used for non-food/beverage grade CO_2 transportation. The cost for one delivery trailer, estimated at $100,000, is added to each of the non-food/beverage grade equipment cost estimates. Storage, as described in the following equipment list, would be an additional $1,070,000. The other equipment included in this cost estimate that would be necessary for 272-tonne/day (300-ton/day) capture and non-food/beverage grade purification is as follows:

- Truck scale
- Transportation trailer
- Metal building approximately 18 m × 30.5 m (60 ft × 100 ft) with 6.7-m (22-ft) eave height, control room, manager's office, and driver area
- Electrical gear, including transformers, motor starters, and associated switch gear
- CO_2 pipeline (inside plant limits)
- Labor and materials to install and place the plant in service
- Three days storage in at least two units: use two 454-tonne (500-ton) capacity units

Cost of this installation is estimated at approximately $1,350,000, which, with the $2,280,000 in equipment and $1,070,000 in storage cost, provides a total estimated cost of $4,700,000, or a cost of $17,265/tonne ($15,667/ton) of daily capacity. The estimated cost of installation is based on discussions with companies that have built CO_2 liquefaction facilities during the past 2 yr.

Please note that costs may vary considerably depending on union versus non-union labor, site condition and suitability, contractor availability, distance from source, and other site-specific items.

There is more uncertainty in the non-food/beverage grade cost estimates and their electrical requirement estimates due to a lack of historical data and current fabrication of these types of plants. Additional process simulation would be required to improve the accuracy of the electrical requirements for the non-food/beverage grade facilities.

Table 1 provides the cost summary for the four approaches considered in this report.

Table 1. Summary of estimated costs for CO_2 recovery options.

Cost	75 tons/day food/ beverage grade	75 tons/day non-food/beverage grade	300 tons/day food/ beverage grade	300 tons/day non-food/beverage grade
Process equipment cost, $	1,200,000	800,000	2,850,000	2,100,000
Freight, $	40,000	30,000	100,000	80,000
Storage equipment, $	340,000	340,000	1,070,000	1,070,000
Installation, $	950,000	825,000	1,750,000	1,350,000
Delivery trailer, $	N/A	100,000	N/A	100,000
Total installed equipment, $	2,530,000	2,095,000	5,770,000	4,700,000
Total installed equipment, $/ton daily capacity	33,733	27,933	19,233	15,667
Design, kWh	553	601	2,062	2,148
Electricity, $/ton CO_2 produced[1]	17.66	19.19	16.50	17.18
Connected hp, hp	952	947	3,393	2,895

[1] Electricity cost of $0.10/kWh assumed.

Acknowledgment

Paul Selz, an independent consultant, was a key contributor in the preparation of this report. Selz has decades of engineering experience in the CO_2 recovery, refrigeration, and related industries. His experience and his extensive knowledge of the equipment providers in these industries were very helpful in the preparation of this report. His contributions are much appreciated by Trimeric Corporation.

Appendix A: Used Equipment Searches

Discussion of the Used Equipment Market

Trimeric interviewed a major refrigeration system manufacturer on January 6, 2006. The representative from this company summarized the current difficulties in finding a complete, used CO_2 liquefaction plant that could be purchased and relocated to an ethanol plant. In recent years, some used CO_2 liquefaction plants came on the market as natural gas prices spiked, making uneconomical the operation of ammonia plants that rely on natural gas as a feedstock and produce CO_2 as a recoverable by-product. The plants that were shut down have already sold their CO_2 liquefaction systems, and the plants that remain in operation are likely to survive. The refrigeration system manufacturer said that any remaining used CO_2 liquefaction plant on the market is likely to be in unacceptable condition.

A CO_2 industry expert consultant explained one further difficulty in acquiring used CO_2 liquefaction plants is that merchant CO_2 suppliers sometimes decommission their own equipment or buy other companies' equipment and decommission it rather than letting it go to other competitors.

A change in the industry may cause more used CO_2 liquefaction plants to come on the market in the future. Requirements for lower-sulfur fuels are causing refineries to move toward the use of pressure swing adsorption (PSA) units to produce a higher-purity H_2 needed to meet more stringent sulfur limitation in fuel products. The PSA units do not generate CO_2 that is economical to capture, whereas previous process configurations employing alkanolamine-based purification did produce a recoverable CO_2 stream. Discussions with CO_2 industry experts indicate that at least two CO_2 liquefaction facilities in the United States are shutting down during spring 2006 for this reason.

Used Equipment Internet Searches and Inquiries

Trimeric contacted several sources of used equipment early in 2006 in search of compressors, refurbished CO_2 storage tanks, liquefaction plants, and component equipment that could be used to assemble a 68-tonne/day (75-ton/day) CO_2 capture facility for non-food/beverage grade CO_2 purification on a piece-by-piece basis.

Trimeric made numerous additional telephone and Internet contacts in search of used equipment in order to determine the availability and cost of used equipment for this study. However, no significant additional opportunities for used equipment were identified. Refurbished CO_2 storage tank costs are on the order of \$1,433/tonne (\$1,300/ton) of storage capacity in the tank. These costs include a refrigeration system to reduce evaporative CO_2 losses during storage and an electrically heated vaporizer to maintain storage tank pressure during liquid drawdown from the storage tank. According to refurbished tank vendors and other CO_2 industry experts, refurbished tanks typically cost about 75% of the cost of a new CO_2 storage tank.

Trimeric made numerous additional telephone and Internet contacts in search of used equipment but found no significant additional opportunities for used equipment. It is important to note that the used equipment market is fluid, and additional efforts to make these kinds of used equipment searches would be worthwhile in the future as part of a CO_2 capture project at an ethanol plant.

Appendix B: Equipment Details for CO_2 Capture: Food/Beverage Grade Case

The following tables summarize the preliminary design of the equipment, consumable materials and energy required for 68-tonne/day (75-ton/day) and 272-tonne/day (300-ton/day) CO_2 capture and food/beverage grade purification cases.

Table B-1. 75-ton/day CO_2 capture: food/beverage grade case plant vessels.

Vessel no.	Vessel function	Vessel diameter (inches)	Vessel length seam/seam (inches)	Material of construction[1]	Design working pressure (psi)
V-1	Inlet separator	24	60	304 SS	25
V-2	Intercooler separator	18	48	304 SS	150
V-3	Sulfur removal bed	14	120	304 SS	350
V-4	Aftercooler separator	12	48	304 SS	350
V-5	Coalescer	8	36	CS	350
V-6	Lube oil carbon bed	14	96	CS	350
V-7	Wash column	18	192	304 SS	350
V-8 A, B	CO_2 dryer	20	96	CS	350
V-9 A, B	Carbon bed	20	96	CS	350
V-10	Pump receiver	18	60	CS	350
V-11	CO_2 distillation column	8	384	CS	350
V-12	Ammonia economizer	24	60	CS	250
V-13	Ammonia receiver	30	240	CS	250
V-14	Oil still	8	36	CS	350
V-15	Ammonia phase separator for E-6 and E-8	18	144	CS	250
V-16	Ammonia phase separator for E-1 and E-4	Integral with E-1			

[1] CS, carbon steel; SS, stainless steel.

Table B-2. 75-ton/day CO_2 capture: food/beverage grade case heat exchangers.

Exchanger no.	Function, type, and size	Surface area (ft^2)	Tube side material[1]	Design working pressure (psi)	Shell side material[1]
E-1	CO_2 inlet cooler type BEM 12-120	242	304 SS	50	CS
E-2	CO_2 intercooler type BEM 10-120	163	304 SS	150	CS
E-3	CO_2 aftercooler type BEM 10-120	163	304 SS	350	CS
E-4	CO_2 chiller type BJM 8-120	77	304 SS	350	CS
E-5C (CO_2)	CO_2 reboiler type BKU 8/6-204	70 (120 length)	CS	350	CS
E-5A (NH_3)	CO_2 reboiler type BKU 8/6-204	21 (72 length)	CS	250	CS
E-6	CO_2 condenser type BJM 15-192	637	CS	350	CS
E-7	CO_2 vent condenser type BKM 10-96	130	SS	350	SS
E-8	CO_2 subcooler type BJM 6-120	35	CS	350	CS
E-9	Heat recovery pipe-in-pipe	TBD[2]	CS	350	CS
E-10	Ammonia evaporative condenser	Galvanized coils, panels, and basin			

[1] CS, carbon steel; SS, stainless steel. [2] To be determined.

Table B-3. 300-ton/day CO_2 capture: food/beverage grade case plant vessels.

Vessel no.	Vessel function	Vessel diameter (inches)	Vessel length seam/seam (inches)	Material of construction[1]	Design working pressure (psi)
V-1	Inlet separator	48	72	304 SS	25
V-2	Intercooler separator	36	60	304 SS	150
V-3	Sulfur removal bed	48	168	304 SS	350
V-4	Aftercooler separator	30	60	304 SS	350
V-5	Coalescer	14	36	CS	350
V-6	Lube oil carbon bed	30	120	CS	350
V-7	Wash column	24	192	304 SS	350
V-8 A, B	CO_2 dryer	30	120	CS	350
V-9 A, B	Carbon bed	30	120	CS	350
V-10	Pump receiver	30	72	CS	350
V-11	CO_2 distillation column	20	384	CS	350
V-12	Ammonia economizer	42	168	CS	250
V-13	Ammonia receiver	48	240	CS	250
V-14	Oil still	8	36	CS	350
V-15	Ammonia phase separator for E-6 and E-8	30	168	CS	250
V-16	Ammonia phase separator for E-1 and E-4	Integral with E-1			

[1] CS, carbon steel; SS, stainless steel.

Table B-4. 300-ton/day CO_2 capture: food/beverage grade case heat exchangers.

Exchanger no.	Function, type, and size	Surface area (ft^2)	Tube side material[1]	Design working pressure (psi)	Shell side material[1]
E-1	CO_2 inlet cooler type BEM 18-192	774	304 SS	50	CS
E-2	CO_2 intercooler type BEM 12-16	353	304 SS	150	CS
E-3	CO_2 aftercooler type BEM 12-144	254	304 SS	350	CS
E-4	CO_2 chiller type BJM 12-192	341	304 SS	350	CS
E-5C (CO_2)	CO_2 reboiler type BKU 25/13-240	787 (96 length)	CS	350	CS
E-5A (NH_3)	CO_2 reboiler type BKU 25/13-240	177 (72 length)	CS	250	CS
E-6	CO_2 condenser type BJM 27-216	2,350	CS	350	CS
E-7	CO_2 vent condenser type BKM 17-120	512	SS	350	SS
E-8	CO_2 subcooler type BJM 12-120	211	CS	350	CS
E-9	Heat recovery pipe-in-pipe	TBD[2]	CS	350	CS
E-10	Ammonia evaporative condenser	Galvanized coils, panels, and basin			

[1] CS, carbon steel; SS, stainless steel. [2] To be determined.

Table B-5. Preliminary estimate of consumable materials required for CO_2 recovery and food/beverage grade purification.

Tag no. of associated item	Component	Quantity required for 75-ton/day unit	Quantity required for 300-ton/day unit
V-12	Ammonia refrigerant	2,600 lbs	6,000 lbs
V-3	"Sulfa Treat" (bed sized for 180-day operation)	30 ft^3 (approx. 2,100 lbs)	200 ft^3 (approx. 14,000 lbs)
V-6	Oil removal carbon (bed sized for 365-day operation)	10 ft^3 (approx. 400 lbs)	47 ft^3 (approx. 1,880 lbs)
V-8 A, B; V-9 A, B	Dryer/carbon bed support material	6 ft^3 (approx. 240 lbs)	20 ft^3 (approx. 800 lbs)
V-8	Alumina ("Selexsorb CD" or equal)	26 ft^3 (approx. 1,560 lbs)	80 ft^3 (approx. 4,800 lbs)
V-9	Carbon ("Calgon PCB" or equal)	26 ft^3 (approx. 1,000 lbs)	80 ft^3 (approx. 3,200 lbs)

Table B-6. Preliminary estimate of electricity required for 75-ton/day CO_2 capture: food/beverage grade case.

Item[1]	Description	Swept volume (cfm)	Motor size (hp)	Required Bhp	kWh	Efficiency	Item kWh
C-1	Low-stage CO_2 compressor	1,110	250	190	141.7	95.0	149.2
CP-1	Oil pump (C-1)		5	3.5	2.6	87.0	3.0
C-2	High-stage CO_2 compressor	292	250	200	149.2	92.0	162.2
CP-2	Oil pump		3	2	1.5	87.0	1.7
C-3	NH_3 high-stage compressor	853	350	270	201.4	95.0	212.0
CP-3	Oil pump		3	2	1.5	87.0	1.7
H-1	Dryer regen. heater	24-hr average			30.0	56.0	2.2
H-2	Carbon regen. heater	24-hr average			30.0	56.0	2.2
P-1	Column pump		2	1.5	1.1	60.0	1.9
P-2	Wash column pump		5	2.5	1.9	50.0	3.7
P-3	Cooling pump for oil coolers		2	1.3	1.0	60.0	1.6
EC-1	Evaporative condenser pump		1.5	1.0	0.7	85.0	0.9
EC-2	Evaporative condenser fan		10	8.5	6.3	90.0	7.0
P-4A	CO_2 loading pump (avg. for 15 loads/day)		15	10	7.5	60.0	3.3
P-4B	CO_2 loading pump (spare)		15				
Design power draw, kWh							552.7
Liquid CO_2 produced, tons/hr							3.13
Estimated kWh/ton of liquid CO_2 produced							176.6
Approximate connected hp for equipment in 75-ton/day plant							952

[1] For simplicity, Figures 1, 2, 3, and 4 do not show all essential equipment in the plant. However, all equipment consumes electricity and so is listed in this table.

Table B-7. Preliminary estimate of electricity required for 300-ton/day CO_2 capture: food/beverage grade case.

Item[1]	Description	Swept volume (cfm)	Motor size (hp)	Required Bhp	kWh	Efficiency	Item kW/h
C-1	Low-stage CO_2 compressor	5,760	1,250	980	731.1	95.0	769.6
CP-1	Oil pump (C-1)		15	11	8.2	90.0	9.1
C-2	High-stage CO_2 compressor	1,110	700	590	440.1	95.0	463.3
CP-2	Oil pump		15	12	9.0	87.0	10.3
C-3	Low-stage NH_3 compressor	2,236	600	490	365.5	95.0	384.8
CP-3	Oil pump		10	7	5.2	87.0	6.0
C-4	NH_3 high-stage compressor	1,110	500	405	302.1	95.0	318.0
CP-4	Oil pump		10	7	5.2	87.0	6.0
H-1	Dryer regen. heater	24-hr average			125.0	56.0	9.3
H-2	Carbon regen. heater	24-hr average			125.0	56.0	9.3
P-1	Column pump		5	4	3.0	60.0	5.0
P-2	Wash column pump		20	17	12.7	50.0	25.4
P-3	Cooling pump for oil coolers		7.5	6	4.5	60.0	7.5
EC-1	Evaporative condenser pump		7.5	5	3.7	85.0	4.4
EC-2	Evaporative condenser fan		50	40	29.8	90.0	33.2
P-4A	CO_2 loading pump (avg. for 15 loads/day)		20	16	11.9	60	1.3
P-4B	CO_2 loading pump (spare)		15				
Design power draw, kWh							2,062.4
Liquid CO_2 produced, tons/hr							12.5
Estimated kWh/ton of liquid CO_2 produced							165.0
Approximate connected hp for equipment in 300-ton/day plant							3,393

[1] For simplicity, Figures 1, 2, 3, and 4 do not show all essential equipment in the plant. However, all equipment consumes electricity and so is listed in this table.

Applicable Codes and Standards

The system would typically be built to the following codes and standards:

- American Society of Mechanical Engineers Code for Unfired Pressure Vessels, Section VIII, Division I for all pressure vessels
- Tubular Equipment Manufacturer’s Association, Class “C” for all shell and tube exchangers
- National Electrical Code USA for wiring and electrical components

- American National Standards Institute, Section B31.5 for ammonia piping and Section B31.3 for CO_2 piping
- American National Standards Institute, ANSI/ASHRAE 15-84 Safety Code for Mechanical Refrigeration for the ammonia system
- National Electrical Manufacturer's Association for electric motors and enclosures

Plant Production

Plant production level will need to be a minimum of 68 tonnes/day (75 tons/day) or 272 tonnes/day (300 tons/day), net to storage, averaged over 24 hr. Plant storage pressure design is 16.3 atm (240 psia) at –25.6°C (–14°F).

Inlet and Outlet Gas Composition

The inlet conditions assumed are 0.987 atm (14.5 psia) at 27°C (80°F) at the inlet separator, saturated with water vapor. The design atmospheric pressure is 1.0 atm (14.7 psia). Table B-8 summarizes the typical inlet and outlet stream conditions for CO_2 recovery and food/beverage grade purification from an ethanol plant.

Table B-8. Typical CO_2 inlet and outlet stream conditions for CO_2 recovery from an ethanol plant and food/beverage grade purification.

Component	Inlet design case maximum amount	Product maximum amount
Water	Saturated	–70°F dew point (7.8 ppmv)[1]
Acetaldehyde	3 ppmv	0.2 ppmv
Methanol	50 ppmv	10 ppmv
Ethanol	950 ppmv	10 ppmv
Acetone	1.5 ppmv	Total allowable of 1.0 ppmv for these six components
Ethyl acetate	30 ppmv	
n-Propanol	0.7 ppmv	
i-Butanol	3 ppmv	
n-Butanol	0.5 ppmv	
Isoamyl acetate	0.6 ppmv	
Hydrogen sulfide	1 ppmv	0.2 ppmv
Dimethyl sulfide	0.5 ppmv	0.2 ppmv
Nitrogen	600 ppmv	120 ppmv
Oxygen	60 ppmv	30 ppmv
Methane	3 ppmv	20 ppmv
Carbon dioxide	Balance	99.985 minimum %

[1] ppmv, parts per million, by volume.

Utility Water Requirements

The ammonia evaporative condenser will require approximately 0.41 L/s (6.5 U.S. gpm) for the 68-tonne/day (75-ton/day) unit and 1.44 L/s (22 gpm) for the 272-tonne/day (300-ton/day) unit. Typically, recommended blowdown is the same; thus, total water consumption for the condenser would be in the area of 0.82 L/s (13 gpm) for the 68-tonne/day (75-ton/day) unit and 2.8 L/s (44 gpm) for the 272-tonne/day (300-ton/day) unit. Blowdown rates may vary depending on the type of water treatment utilized.

Potable water for the scrubber is estimated at approximately 0.4 to 0.5 L/s (6 to 8 gpm) for the 68-tonne/day (75-ton/day) unit and 1.1 to 1.6 L/s (18 to 25 gpm) for the 272-tonne/day (300-ton/day) unit.

Instrument Air Requirements

The instrument air requirement is approximately 0.71 m^3/hr (25 scf/hr) for either plant size. The system would be designed for CO_2 vent gas to be used as the primary source of instrument air.

Appendix C: Equipment Details for CO_2 Capture: Non-Food/Beverage Grade Case

The following tables summarize the preliminary design of the equipment, consumable materials and energy required for 68 tonne/day (75-ton/day) and 272-tonne/day (300-ton/day) CO_2 capture and non-food/beverage grade purification.

Table C-1. 75-ton/day CO_2 capture: non-food/beverage grade case plant vessels.

Vessel no.	Vessel function	Vessel diameter (inches)	Vessel length seam/seam (inches)	Material of construction[1]	Design working pressure (psi)
V-1	Inlet separator	24	60	304 SS	25
CO_2-1	Lube oil coalescer	8	36	CS	350
V-2	Aftercooler separator	12	48	304 SS	350
V-3 A, B	CO_2 dryer	20	96	CS	350
V-4	Flash separator	12	48	CS	350
V-5	Ammonia economizer	24	60	CS	250
V-6	Ammonia receiver	30	240	CS	250
V-7	Oil still	8	36	CS	350
V-8	Ammonia phase separator for E-1 and E-3	18	144	CS	250
V-9	Ammonia phase separator for E-4 and E-5	Integral with E-4			

[1] CS, carbon steel; SS, stainless steel.

Table C-2. 75-ton/day CO_2 capture: non-food/beverage grade case heat exchangers.

Exchanger no.	Function, type, and size	Surface area (ft^2)	Tube side material[1]	Design working pressure (psi)	Shell side material[1]
E-1	CO_2 inlet cooler type BEM 12-120	242	304 SS	50	CS
E-2	CO_2 aftercooler type BEM 10-120	163	304 SS	350	CS
E-3	CO_2 chiller type BJM 8-96	80	304 SS	350	CS
E-4	CO_2 condenser type BJM 12-180	365	CS	350	CS
E-5	CO_2 subcooler type BJM 8-120	192	CS	350	CS
E-6	Heat recovery pipe-in-pipe	TBD[2]	CS	350	CS
E-7	Ammonia evaporative condenser	Galvanized coils, panels, and basin			

[1] CS, carbon steel; SS, stainless steel. [2] To be determined.

Table C-3. 300-ton/day CO_2 capture: non-food/beverage grade case plant vessels.

Vessel no.	Vessel function	Vessel diameter (inches)	Vessel length seam/seam (inches)	Material of construction[1]	Design working pressure (psi)
V-1	Inlet separator	48	72	304 SS	25
CO_2-1	Lube oil coalescer	14	36	CS	350
V-2	Aftercooler separator	30	60	304 SS	350
V-3 A, B	CO_2 dryer	30	120	CS	350
V-4	Flash separator	36	96	CS	350
V-5	Ammonia economizer	42	168	CS	250
V-6	Ammonia receiver	48	240	CS	250
V-7	Oil still	8	36	CS	350
V-8	Ammonia phase separator for E-1 and E-3	30	168	CS	250
V-9	Ammonia phase separator for E-4 and E-5	Integral with E-4			

[1] CS, carbon steel; SS, stainless steel.

Table C-4. 300-ton/day CO_2 capture: non-food/beverage grade heat exchangers.

Exchanger no.	Function, type, and size	Surface area (ft^2)	Tube side material[1]	Design working pressure (psi)	Shell side material[1]
E-1	CO_2 inlet cooler type BEM 18-192	774	304 SS	50	CS
E-2	CO_2 aftercooler type BEM 10-192	353	304 SS	350	CS
E-3	CO_2 chiller type BEM 12-144	254	304 SS	350	CS
E-4	CO_2 condenser type BJM 24-192	1,472	CS	350	CS
E-5	CO_2 subcooler type BJM 12-120	211	CS	350	CS
E-6	Heat recovery pipe-in-pipe	TBD[2]	CS	350	CS
E-7	Ammonia evaporative condenser	Galvanized coils, panels, and basin			

[1] CS, carbon steel; SS, stainless steel. [2] To be determined.

Table C-5. Preliminary estimate of consumable materials required for CO_2 recovery and non-food/beverage grade purification.

Tag no. of associated item	Component	Quantity required for 75-ton/day unit	Quantity required for 300-ton/day unit
V-5	Ammonia refrigerant	2,600 lbs	6,000 lbs
V-3 A, B	Dryer bed support material	3 ft^3 (approx. 120 lbs)	10 ft^3 (approx. 400 lbs)
V-3 A, B	Alumina ("Selexsorb CD" or equal)	26 ft^3 (approx. 1,560 lbs)	80 ft^3 (approx. 4,800 lbs)

Table C-6. Preliminary estimate of electricity required for 75-ton/day CO_2 capture: non-food/beverage grade case.

Item[1]	Description	Swept volume (cfm)	Motor size (hp)	Required Bhp	kWh	Efficiency	Item kW/h
C-1	Two-stage CO_2 compressor	1,390	500	435	324.5	95.0	341.6
CP-1	Oil pump (C-1)		10	6	4.5	87.0	5.1
C-2	NH_3 compressor (economized type)	853	350	302	225.3	95.0	237.1
CP-2	Oil pump		3	2	1.5	87.0	1.7
H-1	Dryer regen. heater				30.0	56.0	2.2
P-1	Cooling pump for oil coolers		2	1.3	1.0	60.0	1.6
EC-1	Evaporative condenser pump		1.5	1	0.7	85.0	0.9
EC-2	Evaporative condenser fan		10	8.5	6.3	90.0	7.0
P-2A	CO_2 loading pump (avg. for 3.75 loads/day)		15	10	7.46	60	3.3
P-2B	CO_2 loading pump (spare)		15				
Design power draw, kWh							600.7
Liquid CO_2 produced, tons/hr							3.13
Estimated kWh/ton of liquid CO_2 produced							191.9
Approximate connected hp for equipment in 75-ton/day plant							947

[1] For simplicity, Figures 1, 2, 3, and 4 do not show all essential equipment in the plant. However, all equipment consumes electricity and so is listed in this table.

Table C-7. Preliminary estimate of electricity required for 300-ton/day CO_2 capture: non-food/beverage grade case.

Item[1]	Description	Motor size (hp)	Required Bhp	kWh	Efficiency	Item kWh
C-1	Two-stage CO_2 compressor	1,750	1,680	1,253.3	95.0	1,319.2
CP-1	Oil pump (C-1)	25	19	14.2	90.0	15.7
C-2	NH_3 compressor (economized type)	1,000	950	708.7	95.0	746.0
CP-2	Oil pump	15	13	9.7	87.0	11.1
H-1	Dryer regen. heater			125.0	56.0	9.3
P-1	Cooling pump for oil coolers	7.5	6.0	4.5	60.0	7.5
EC-1	Evaporative condenser pump	7.5	5.0	3.7	85.0	4.4
EC-2	Evaporative condenser fan	50	40	29.8	90.0	33.2
P-2A	CO_2 loading pump (avg. for 15 loads/day)	20	16	11.9	60	1.3
P-2B	CO_2 loading pump (spare)	20				
Design power draw, kWh						2,147.8
Liquid CO_2 produced, tons/hr						12.5
Estimated kWh/ton of liquid CO_2 produced						171.8
Approximate connected hp for equipment in 300-ton/day plant						2,895

[1] For simplicity, Figures 1, 2, 3, and 4 do not show all essential equipment in the plant. However, all equipment consumes electricity and so is listed in this table.

Applicable Codes and Standards

The system would typically be built to the following codes and standards:

- American Society of Mechanical Engineers Code for Unfired Pressure Vessels, Section VIII, Division I for all pressure vessels
- Tubular Equipment Manufacturer's Association, Class "C" for all shell and tube exchangers
- National Electrical Code USA for wiring and electrical components
- American National Standards Institute, Section B31.5 for ammonia piping and Section B31.3 for CO_2 piping
- American National Standards Institute, ANSI/ASHRAE 15-84 Safety Code for Mechanical Refrigeration for the ammonia system
- National Electrical Manufacturer's Association for electric motors and enclosures

Plant Production

Plant production level will need to be a minimum of 68 tonnes/day (75 tons/day) or 272 tonnes/day (300 tons/day), net to storage, averaged over 24 hr, based on a 15% allowance for flash losses. Plant storage pressure design is 16.3 atm (240 psia) at –25.6°C (–14°F).

Inlet and Outlet Gas Composition

The inlet conditions assumed are 0.987 atm (14.5 psia) at 27°C (80°F) at the inlet separator, saturated with water vapor. The design atmospheric pressure is 1.0 atm (14.7 psia). Table C-8 summarizes the typical inlet and outlet stream conditions for non-food/beverage grade CO_2 recovery from an ethanol plant.

Table C-8. Typical CO_2 inlet and outlet stream conditions for CO_2 recovery from an ethanol plant and non-food/beverage grade purification.

Component	Inlet design case maximum amount	Product maximum amount
Water	Saturated	– 70°F dew point (7.8 ppmv)[1]
Acetaldehyde	3 ppmv	NS[2]
Methanol	50 ppmv	NS
Ethanol	950 ppmv	NS
Acetone	1.5 ppmv	NS
Ethyl acetate	30 ppmv	NS
n-Propanol	0.7 ppmv	NS
i-Butanol	3 ppmv	NS
n-Butanol	0.5 ppmv	NS
Isoamyl acetate	0.6 ppmv	NS
Hydrogen sulfide	1 ppmv	NS
Dimethyl sulfide	0.5 ppmv	NS
Nitrogen	600 ppmv	NS
Oxygen	60 ppmv	NS
Methane	3 ppmv	NS
Carbon dioxide	Balance	NS

[1] ppmv, parts per million, by volume. [2] NS, no specification.

Utility Water Requirements

Cooling tower water, if available, can be used for the ammonia condenser and for the aftercooler. Approximately 22.1 L/s (350 U.S. gpm) of 24°C (75°F) tower water will be required for the 68-tonne/day (75-ton/day) non-food/beverage grade plant and approximately 90.47 L/s (1,450 gpm) of 24°C (75°F) tower water for the 272-tonne/day (300-ton/day) plant.

If cooling tower water is not available, then the ammonia evaporative condenser will require approximately 0.41 L/s (6.5 gpm) for the 68-tonne/day (75-ton/day) unit and 22 gpm for the 272-tonne/day (300-ton/day) unit. Typically, recommended blowdown is the same; thus, total water consumption for the condenser would be in the area of 0.824 L/s (13 gpm) for the 68-tonne/day (75-ton/day) unit and 2.8 L/s (44 gpm) for the 272-tonne/day (300-ton/day) unit. Blowdown rates may vary depending on the type of water treatment utilized.

Instrument Air Requirements

The instrument air requirement is approximately 0.71 m^3/hr (25 scf/hr) for either plant size. The system would be designed for CO_2 vent gas to be used as the primary source of instrument air.