Evaluation of Carbon Dioxide Capture Options from Ethanol Plants

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Front cover: *Left: Distillation column in a* CO_2 *EOR facility. Right: Molecular sieve dryer beds in a* CO_2 *EOR facility. Photographs courtesy of Chaparral Energy.*

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EXECUTIVE SUMMARY

The Illinois State Geological Survey and the Midwest Geological Sequestration Consortium (MGSC) have been conducting carbon dioxide (CO_2) storage and enhanced oil recovery (EOR) testing in the Illinois Basin since 2003. Capital and operating costs for the equipment required to capture and liquefy CO_2 from ethanol plants in the Illinois area were evaluated in 2005 to 2006 so that ethanol plants could be considered as a source of CO_2 for the U.S. Department of Energy-sponsored MGSC CO_2 pilot projects planned at that time. Continued and sustained public and private interest in the 2006 report provided the impetus to update and expand the report.

The estimated capital and operating costs to capture, purify, and liquefy 75 U.S. ton/day (68 tonne/day) and 300 ton/day (272 tonne/day) of CO_2 have been updated, and a larger 1,000 ton/day (907 tonne/day) case has been added. Carbon dioxide used for food and beverage applications is typically transported by truck or rail as a refrigerated liquid at approximately 290 psig (pounds per square inch gauge; 20 bar g [gauge pressure]) and 0 °F (–18 °C). Larger amounts of CO_2 used for EOR are typically transported via a pipeline (vs. truck and tanker trailers), so this report includes a summary of capital and operating costs for equipment that could be added to raise the CO_2 pressure to feed it to a pipeline for the 1,000 ton/day (907 tonne/day) case. For each facility size, estimated costs are provided for producing food and beverage grade CO_2 as well as for producing less purified CO_2 that would be suitable for EOR or storage. The report includes preliminary plant and equipment designs and estimates for major capital and operating costs for each of the recovery options. The availability of used equipment was also assessed.

Table ES1 summarizes the capital and operating cost estimates for each of the recovery plant scenarios. The estimated total installed capital costs for food and beverage grade CO_2 liquefaction facilities are \$4.7 million for a 75 ton/day (68 tonne/day) facility, \$10.5 million for a 300 ton/day (272 tonne/day) facility, and \$21.7 million for a 1,000 ton/day (907 tonne/day) facility. The estimated total installed capital costs for nonfood and nonbeverage grade CO_2 liquefaction facilities generating lower purity CO_2 suitable for EOR or sequestration are \$4.3 million for a 75 ton/day (68 tonne/day) facility, \$9.8 million for a 300 ton/day (272 tonne/day) facility, and \$20.2 million for a 1,000 ton/day (907 tonne/day) facility. The total installed capital cost estimates are based on the average of total facility costs estimated by two to three firms that build these facilities. Therefore, itemized costs for each piece of equipment that is added up to the total facility costs are not available for this report.

Electricity is the largest single operating cost. In the 2006 report, electrical costs were estimated at \$0.10/kWh because that was the price many operators were using at that time to evaluate projects. However, the actual rates in 2006 were more in the range of \$0.04/kWh to \$0.065/kWh; thus, electrical costs were often not as strong of a consideration as they were in 2014. In the 2014 market, costs of \$0.10/kWh were becoming more realistic and operators were assigning more importance to electrical operating costs. Thus, plant designs have evolved that are more focused on reducing electricity consumption. For example, the use of a distillation column to increase the recovery of raw gas and thereby avoid energy costs for the compression, dehydration, and liquefaction of CO_2 that could not be recovered in a plant designed without a distillation column was another factor considered in the decision to use a distillation column, even if the column is not needed to meet CO_2 purity specification requirements. Estimated labor costs have also been included as part of the operating cost estimates.

Electrical costs estimated based on an electricity price of \$0.10/kWh for the current food and beverage grade design facilities are \$16.23/ton (\$17.89/tonne) of CO_2 produced for the 75 ton/day (68 tonne/day) facility, \$14.70/ton (\$16.20/tonne) of CO_2 produced for the 300 ton/day (272 tonne/day) facility, and \$12.96/ton (\$14.28/tonne) of CO_2 produced for the 1,000 ton/day (907 tonne/day) facility. The estimated electrical costs for the lower purity, nonfood and nonbeverage grade CO_2 facilities are \$16.20/ton (\$17.82/tonne) of CO_2 produced for the 75 ton/day (68 tonne/day) facility, \$14.66/ton (\$16.16/tonne) of CO_2 produced for the 300 ton/day (272 tonne/day) facility, and \$12.94/ton (\$14.26/tonne) for the 1,000 ton/day (907 tonne/day) facility.

Used equipment searches in 2006 showed that the used equipment market was limited because of business conditions in the oil and gas industry. More recent inquiries of used equipment dealers and people interested in building these kinds of plants suggest that this remains the case. Merchants in the food and beverage grade CO_2 industry may also naturally avoid putting used equipment on the market when their competitors could acquire it. More recent CO_2 plants have been built as the "packaged" type (in which major equipment is installed on skids so that field piping is minimal) using smaller screw-type compressors, so the plant is simpler and less costly to relocate as compared with the old-style "built-in-place" plants. This trend has the tendency to reduce the availability of used equipment on the market because it is easier for operators to relocate packaged equipment for use elsewhere within their company. With the current outlook in the oil and gas industry and the trend toward more packaged-type plants, used equipment will probably continue to be difficult to locate.

Table ES1 Summary of capital and operating cost estimates

Item	75 ton/day		300 ton/day		1,000 ton/day	
	Food and beverage grade	Nonfood and nonbeverage grade	Food and beverage grade	Nonfood and nonbeverage grade	Food and beverage grade	Nonfood and nonbeverage grade
Purchased equipment cost, $	2,049,000	1,684,000	4,680,000	4,069,000	11,821,000	10,538,000
Freight, $ (included above)	36,500	36,500	103,000	103,000	194,000	194,000
Storage tanks, $	361,000	361,000	1,375,000	1,375,000	2,700,000	2,700,000
Installation, $	2,267,000	2,251,000	4,431,000	4,338,000	7,143,000	6,979,000
Total installed capital cost, $	4,676,000	4,296,000	10,485,000	9,783,000	21,664,000	20,217,000
Total installed capital cost, $/capacity in ton/day	62,000	57,000	35,000	33,000	22,000	20,000
Design, kWh/ton to storage	162.3	162.0	147.0	146.6	130.0	129.4
Power cost/ton, $ (for $0.10/kWh)	16.23	16.20	14.70	14.66	12.96	12.94
Power cost/ton, $ (for $0.055/kWh)	8.92	8.91	8.09	8.06	7.13	7.11
Connected motor horsepower, hp (kW) (not including regeneration heaters)	806 (601)	806 (601)	2,655 (1,981)	2,655 (1,981)	8,210 (6,125)	8,210 (6,125)
Regeneration heater, hp (kW)	67 (50)	34 (25)	214 (160)	107 (80)	603 (450)	302 (225)
Total connected horsepower, hp (kW)	873 (651)	839 (626)	2,869 (2,140)	2,762 (2,060)	8,813 (6,575)	8,512 (6,350)

INTRODUCTION

The Illinois State Geological Survey (ISGS) and the Midwest Geological Sequestration Consortium (MGSC) have been conducting carbon dioxide (CO_2) storage and enhanced oil recovery (EOR) testing in the Illinois Basin since 2003. If testing shows that it is possible to increase oil and gas recovery with CO_2 injection in the Illinois Basin, this may create the need for additional commercial sources of CO_2 in this region, such as ethanol plants. As part of the MGSC project, the Trimeric Corporation evaluated the costs of recovering CO_2 from ethanol plants in the Illinois Basin. This report was originally issued in 2006. Continued and sustained public and private interest in the 2006 report provided the impetus to update and expand the report.

The primary objectives of this and the previous 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 operations. The basis for these studies was to produce CO_2 suitable for transport and delivery by tank trucks. This mode of delivery allows flexibility for CO_2 to be sold to industrial consumers or transported to nearby EOR operations if a suitable pipeline network is unavailable. Longer term, large-scale EOR and sequestration operations would likely be supported with a CO_2 pipeline infrastructure. This updated report also provides estimates of the capital and operating costs that would be associated with adding a multistage centrifugal pump, recycle valve, and other equipment to the 1,000 ton/day (907 tonne/day) facility to deliver the CO_2 produced in these facilities to a pipeline if a pipeline is available.

PLANT CAPACITY SELECTION, EQUIPMENT SELECTION, AND COST-ESTIMATING APPROACH

This study compares the cost to produce 75 ton/day (68 tonne/day) of CO_2 (e.g., a pilot test) with the cost to install a full-scale commercial facility with a capacity of either 300 ton/day (272 tonne/day) or 1,000 ton/day (907 tonne/day) of CO_2. The 300 ton/day (272 tonne/day) capacity is representative of the CO_2 available for recovery at an ethanol plant producing 40 million gal/year (151 million liter/year) of ethanol. The 1,000 ton/day (907 tonne/day) capacity is representative of the CO_2 available for recovery at an ethanol plant producing 130 million gal/year (492 million liter/year) of ethanol, which represents one of the larger ethanol plants in the Illinois Basin. Cost differences in these three facility sizes can be used to compare the cost of capturing the amount of CO_2 required to meet a small local demand with the cost to install a full-scale commercial facility for the sale of CO_2. Economies of scale and costs of operation result in very few CO_2 liquefaction facilities with a capacity of less than 200 ton/day (181 tonne/day) being built.

CARBON DIOXIDE RECOVERY AND PURIFICATION OPTIONS

Food and beverage grade CO_2 is not required for enhanced oil and gas recovery or storage. However, "grassroots" CO_2 plants would likely be designed for the production of food and beverage grade CO_2 to have a broad client base, particularly in areas like Illinois, where a mature CO_2 EOR market has not yet developed. However, some nonfood and nonbeverage grade liquefaction plants have been built and are in operation today. In at least some of these instances, liquefaction for EOR use required using distillation to meet CO_2 purity specifications for oxygen content. Trimeric estimated the costs of producing both food and beverage grade and nonfood and nonbeverage grade CO_2 to show the estimated incremental cost difference between food and beverage grade and nonfood and nonbeverage grade CO_2. Food grade and beverage grade typically allow a maximum of 0.1 ppmv (parts per million by volume) of total sulfur content (excluding sulfur dioxide). For the purposes of this report, and for most practical purposes when recovering CO_2 from ethanol plants, specifications for food grade or beverage grade are nearly equivalent.

Much of the equipment required to produce food and beverage grade CO_2 is the same as that required to produce nonfood and nonbeverage grade CO_2, with the primary difference between food or beverage grade and nonfood or nonbeverage grade plants being the additional purification equipment required to remove sulfur, hydrocarbons, and other organic contaminants. In either food and beverage grade or nonfood and nonbeverage grade plants, minimized power costs and maximized recovery of the available raw product are often the governing requirements of plant design. Transportation of nonfood and nonbeverage grade CO_2 may need to include the extra cost of a dedicated delivery trailer fleet because trailers in food and beverage grade CO_2 service are not used for nonfood and nonbeverage grade CO_2 transportation.

Distillation is required to meet product purity specifications for food and beverage grade CO_2. Therefore, distillation was included in the 2006 design for the food and beverage grade cases. However, the 2006 report did not include distillation for the nonfood and nonbeverage grade CO_2 cases. That design was a lower capital cost, higher operating cost approach that used a simple two-phase flash separation instead of distillation. Trimeric decided to include distillation in the updated report, even for the nonfood and nonbeverage grade cases. In addition to removing oxygen from the CO_2, adding the distillation column reduces venting losses of CO_2 as compared with the 2006 design. This also reduces operating costs because a higher percentage of the feed would have been undergoing the energy-intensive processes of compression, dehydration, and liquefaction before being lost to the vent in the two-phase flash of the 2006 design.

Although not accounted for in the 2006 study or the current report, purchasing the raw CO_2 feed stream from the CO_2 source facility usually involves a cost to the operator of the CO_2 recovery facility. Companies in the CO_2 business consider the cost of the raw feed gas confidential. Therefore, Trimeric has no documentable basis for estimating the costs associated with the higher venting losses without distillation, but we do think these costs would drive operators toward a design with distillation. We understand from industry contacts that the cost of raw feed gas has generally been on an upward trend for several years, often more than double the prices in 2006. The location of the raw gas source may also play an important role in the feed gas cost, particularly in the food and beverage industry. Transportation costs are high for food and beverage grade CO_2, and a source closer to large markets for this product would therefore be of greater value to the CO_2 producer. Similarly, the proximity to EOR fields would be a consideration with respect to transportation costs in EOR applications.

In recent years, Trimeric has observed increasing concern about oxygen in CO_2 that is transported in pipelines and used for EOR, which further substantiates the need to distill CO_2 for EOR applications if it contains significant amounts of oxygen. Higher levels of oxygen in CO_2 lead to concerns with biological growth in oil and gas reservoirs and with an increased potential for cathodic reactions and thus corrosion from the higher oxygen content in CO_2 recycle streams that have been in contact with formation water. Adding distillation makes it possible to meet stringent CO_2 pipeline specifications for oxygen, which are currently approximately 10 to 20 ppmv. Meeting this specification would usually not be possible with the simple two-phase flash design that was used in the 2006 report for the nonfood and nonbeverage grade CO_2 cases. Figure 1 shows a distillation column in a plant used to produce CO_2 for EOR.

EFFORTS TO FIND USED CARBON DIOXIDE LIQUEFACTION PLANTS AND USED COMPONENTS

In 2006, Trimeric surveyed several used equipment dealers to determine the availability of used equipment for CO_2 recovery. The used equipment market was limited at that time because of the business conditions in the oil and gas industry. For the present study, inquiries with used equipment dealers and individuals interested in building these kinds of plants suggest that the availability of used equipment is still relatively limited. Merchants in the food and beverage grade CO_2 industry may also avoid putting used equipment on the market when their competitors could acquire it. More recent CO_2 plants have been built as the "packaged" type (in which major equipment is installed on skids so that field piping is minimal) by using smaller screw-type compressors so that the plant is simpler and less costly to relocate as compared with the old "built-in-place" style plants. This trend has the tendency to reduce the availability of used equipment on the market because it is easier for operators to relocate packaged equipment for use elsewhere within the company. With the current outlook in the oil and gas industry and the trend toward more packaged-type plants, the used equipment market will continue to be limited. Figure 2 shows a typical screw compressor package.

Specific major equipment pieces, such as a CO_2 compressor or a skid-mounted refrigeration system, could possibly be purchased for use at the beginning of a CO_2 recovery project at an ethanol plant. Past projects have achieved savings of approximately 30% by refurbishing and reengineering compressors compared with purchasing new compressors. Unless specifically noted otherwise, the costs presented in the remainder of this document are for new equipment.

Figure 1 Distillation column in a CO_2 EOR facility. Photograph courtesy of Chaparral Energy.

EQUIPMENT REQUIRED FOR CARBON DIOXIDE CAPTURE

To a large degree, the type of equipment required for CO_2 recovery does not depend on the recovery rate. The size and cost of the equipment increase with increasing recovery rate. Differences in the equipment and operating costs are associated with purification of CO_2 for food and beverage uses compared with nonfood and nonbeverage uses. These costs are largely due to the equipment required to remove sulfur, hydrocarbons, and other organic contaminants to meet food and beverage grade CO_2 specifications that are not required for nonfood and nonbeverage uses. A water scrubber,

Figure 2 Screw compressor package. Photograph courtesy of GEA FES.

sulfur removal beds, and carbon beds are removed from the food and beverage grade CO_2 plant design for the nonfood and nonbeverage grade cases. Otherwise, the plant designs are the same.

Since the 2006 version of the report was issued, the capacity range of two-stage compound screw compressors has increased as manufacturers have made more models available. The rating software has also improved, which allows designers to better match the first-stage (low-stage) and second-stage (high-stage) compressor bodies to the required flow rate. The use of two-stage compound screw compressors usually decreases the operating cost for the facility. The use of the two-stage compound machines also lowers the facility capital cost because they require only one lubrication system instead of two and one main drive motor and starter instead of two, eliminate one interstage cooler and separator, and reduce piping and insulation requirements. Recent developments in two-stage compound screw compressors allow the use of this type machine in almost all instances, instead of the use of separate first- and second-stage compressor systems. Thus, this reduces the overall number of CO_2 and ammonia (refrigerant) compressors in the updated cases in this report as compared with the 2006 report.

The capabilities of plant control systems, such as a distributed control system (DCS) or programmable logic controller (PLC), have greatly increased in recent years. This reduces the amount of operator and supervisor labor required to operate and maintain the facility. Labor costs were not included in the 2006 report. However, labor is often the second highest operating cost (after electricity). Trimeric included estimated operator and supervisor labor costs in this updated report.

The remainder of this section contains a detailed description of the equipment required for CO_2 recovery and purification for food and beverage grade CO_2 applications and for nonfood and nonbeverage grade applications. As mentioned, increases in the costs of electricity and feed gas and the increasing demand for low-oxygen-content CO_2 for EOR have made the two types of plants more similar in makeup; thus, distillation is included in all plant designs in the updated version of this report.

Food and Beverage Grade Cases

Figures 3 and 4 show the process flow diagram for the equipment required for CO_2 capture for the food and beverage grade cases. Figure 5 shows the corresponding process flow diagram for the refrigeration equipment required for CO_2 capture for the food and beverage grade cases. A detailed description of this equipment is provided later in this report. The sizing and types of units are preliminary and are 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 an actual source is performed. Temperatures, pressures, and other parameters in the following description are approximate. These plants would typically be designed for unattended operation when using a PLC unit or, in some cases, a small 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 do reciprocating compressors because of their rotary movement and smaller number of moving parts. They also offer superior power characteristics at part load and excellent load/capacity control characteristics. A lubricant management system would be incorporated to ensure an oil-free product. Two-stage screw compressors (often termed “compound compressors”) now have a much greater size range than when the 2006 report was issued, allowing a much broader application range than previously. As discussed, this usually results in reduced capital and operating costs.

Flow rates for each case are based on actual anticipated capacities for specific equipment models; thus, there are some differences relative to the nominal design rates. Feed rates, nominal rates, and actual product rates are summarized in Table 1.

The 75 ton/day (68 tonne/day) and 300 ton/day (272 tonne/day) facilities are designed with one train consisting of one CO_2 compressor and one ammonia (refrigerant) compressor. The 1,000 ton/day (907 tonne/day) facility is based on two compressor trains or two 50% compressors for both services. Screw compressors are available that are large enough to allow for a single-train facility with this capacity, but their size and horsepower make them difficult to install on a package system.

For design purposes, the vapor produced by warming the stored CO_2 is assumed to have an average flow rate of 200 lb/h (90.7 kg/h) for the 75 ton/day (68 tonne/day) facility, 500 lb/h (227 kg/h) for the 300 ton/day (272 tonne/day) facility, and 900 lb/h (408 kg/h) for the 1,000 ton/day (907 tonne/day) facility. This assumption was made to account for heat gain in the storage tanks and the effects of truck loading. Recompression and recycling of these vapors add slightly to the horsepower and electricity requirements for the facilities. The high stage of the CO_2 compressor would be used to remove and recompress vapors from the CO_2 storage tanks to maintain the pressure in the tanks. Losses for storage vapors generally make up a higher percentage of the product rate for smaller facilities, which can lead to a greater difference between the feed and product rates in these plants. The amount of loss is dependent on a number of factors, including the number and size of the product storage tanks and whether vapors are recompressed or lost to the atmosphere.

The plants are designed to accept CO_2 from the source at 13.5 psia (pounds per square inch absolute; 0.9 atm [atmosphere]) and 100 °F (38 °C). The incoming gas from the host plant enters a two-phase separator (VB-1, Figure 3) in which any mechanically entrained water is separated from the gas stream, with the water discharged for treatment or disposal. The gas is then compressed to approximately 28.5 psia (1.9 atm) by a multistage centrifugal-type blower (B-1, Figure 3) and is cooled in a heat exchanger (BE-1, Figure 3) by using recirculated cooling tower water from the host plant. After passing through a second two-phase separator (VB-2, Figure 3) that removes any additional condensed water, the raw gas then enters a pipeline to the CO_2 plant inlet.

In most of today’s plants, the blower unit is installed in the host ethanol plant near

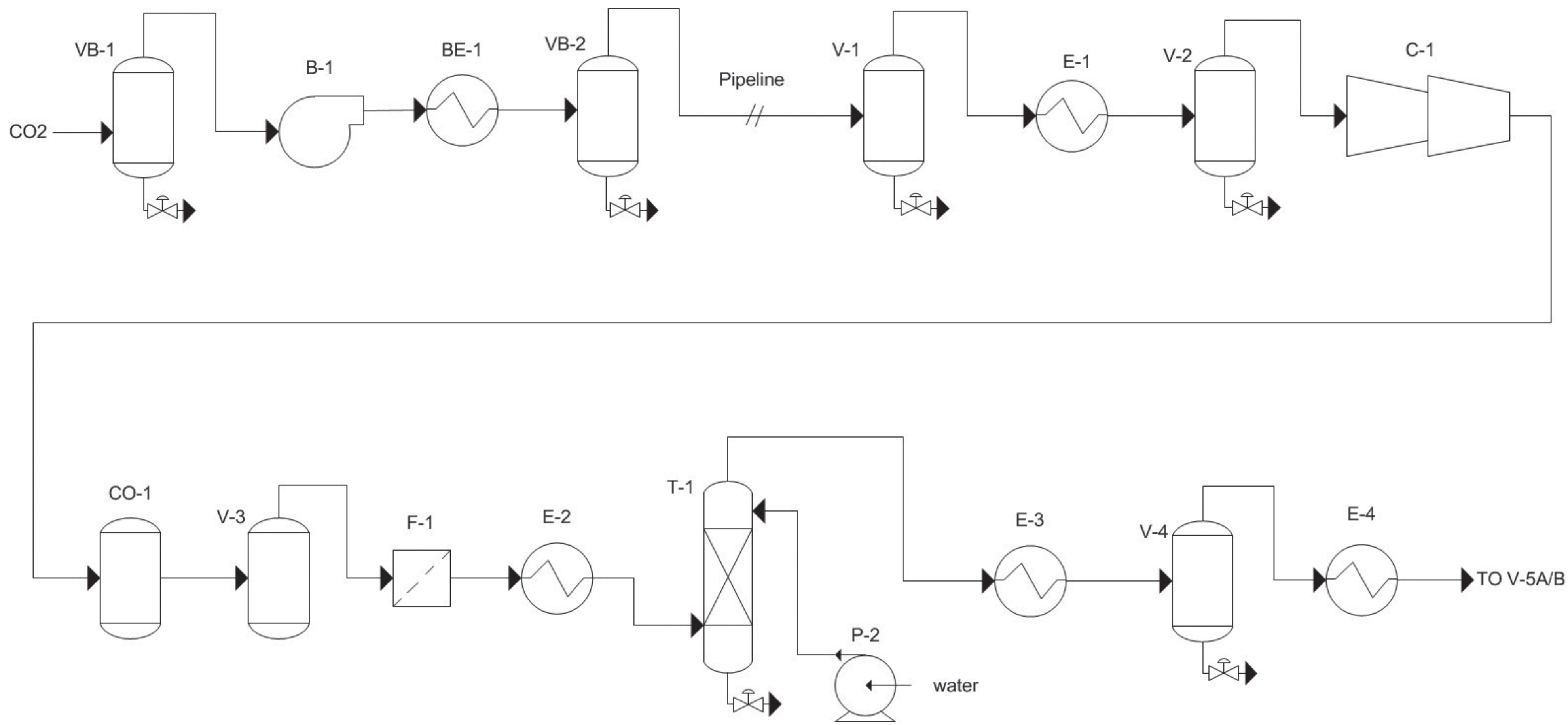

Figure 3 Carbon dioxide process flow diagram for food and beverage grade cases (1 of 2).

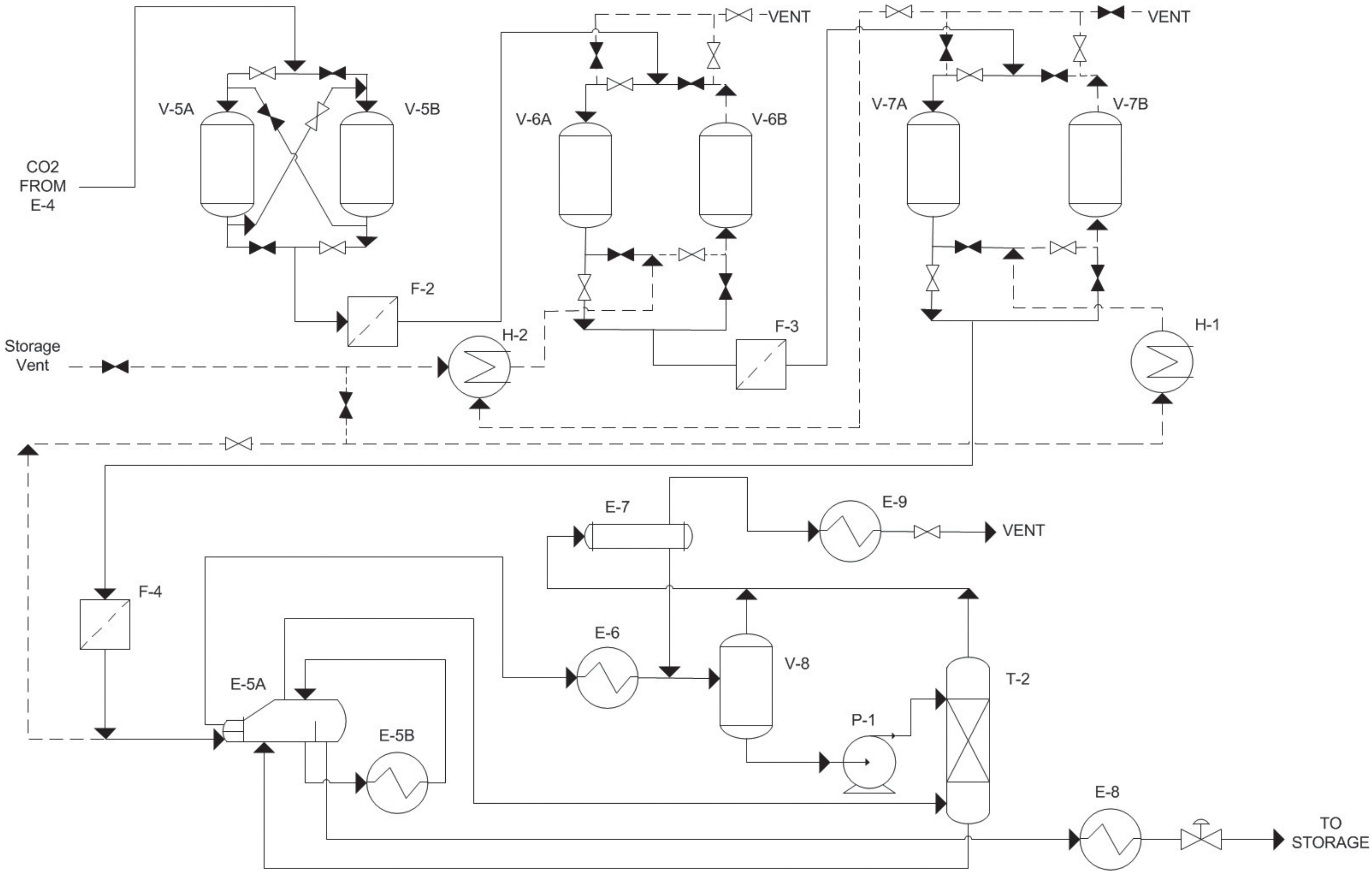

Figure 4 Carbon dioxide process flow diagram for food and beverage grade cases (2 of 2).

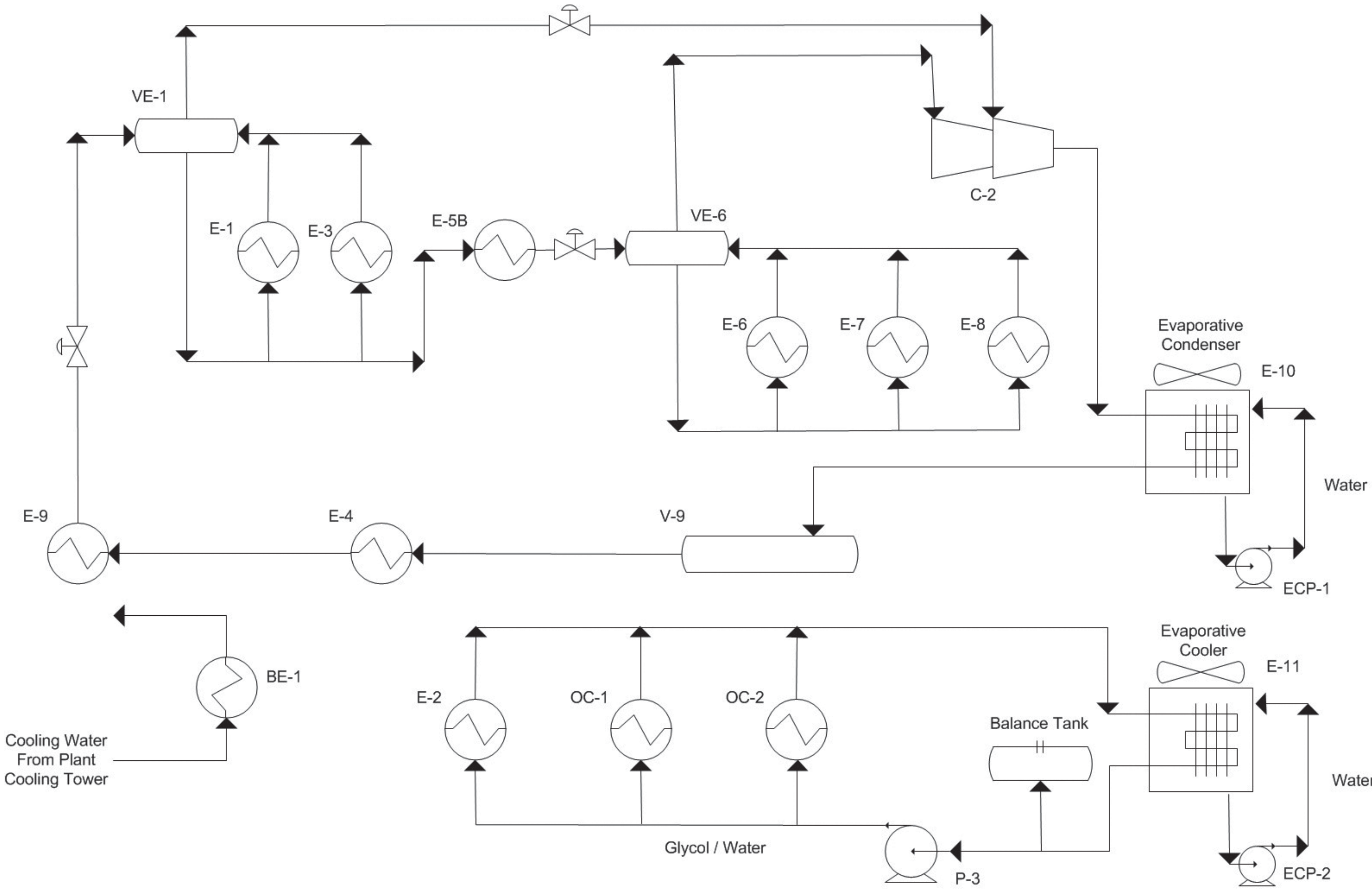

Figure 5 Ammonia refrigeration process flow diagram.

Table 1 Product rates for the food and beverage grade CO_2 cases

	Nominal product rate, ton/day (tonne/day)		
Rate	75 (68)	300 (272)	1,000 (907)
Feed rate, lb/h (kg/h)	7,006 (3,178)	28,025 (12,715)	93,417 (42,385)
Actual product rate, lb/h (kg/h)	6,376 (2,892)	25,126 (11,407)	89,232 (40,486)
Actual product rate, ton/day (tonne/day)	77 (69)	302 (274)	1,071 (972)

the final scrubber of the host plant. This blower unit serves several purposes: (1) it allows a smaller pipeline from the host plant to the CO_2 facility, (2) it requires a smaller main compression unit because of the lower actual volumetric suction flow rate at the inlet to the main compression unit, thus lowering the compressor size and cost, (3) the blower and compressor power requirements combined are usually lower than the compressor power requirements without a blower, and (4) the cost of the blower is more than offset by the savings in the main compression unit.

At the CO_2 plant inlet, the CO_2 enters a phase separator (V-1, Figure 3) to remove any moisture that condensed in the pipeline. The CO_2 then enters a refrigerant-cooled shell and tube exchanger (E-1, Figure 3), where the CO_2 is cooled to lower the water content and volumetric flow rate and to prevent moisture condensation in the two-stage CO_2 compressor (C-1, Figure 3). The condensed water is separated in a phase separator (V-2, Figure 3), and the gas is then compressed to approximately 315 psia (21 atm) in the two-stage CO_2 compressor (C-1, Figure 3). The CO_2 discharge gas enters a high-efficiency oil coalescer (CO-1, Figure 3), and then flows into a carbon bed (V-3, Figure 3) arranged to remove residual oil from the CO_2 gas stream to very low levels (ppbv [parts per billion by volume]). A cartridge-type filter (F-1, Figure 3) then removes particulate matter from the CO_2 gas stream. Next, the CO_2 is cooled in the water-cooled aftercooler (E-2, Figure 3) and flows to a packed-bed water scrubber (T-1, Figure 3) at approximately 100 °F (38 °C) for removal of any water-soluble contaminants (e.g., ethanol and acetaldehyde) in the CO_2 stream. The scrubbing water supplied by pump P-2 is required to be fresh, clean, potable, and odor free.

The CO_2 gas stream then flows through a refrigerant-cooled aftercooler (E-3, Figure 3) and a separator (V-4, Figure 3). The gas then flows through a superheater (E-4, Figure 3), which uses liquid ammonia to slightly warm the CO_2 to minimize the chance of moisture condensation in the adsorbent beds. The slightly superheated CO_2 then enters the primary beds (V-5A and V-5B, Figure 4) for removal of sulfur compounds. A mixed-metal oxide formed on a carbon or alumina substrate, such as HydroCAT GTS 2007 or equivalent, is a typical choice for the primary sulfur removal agent. Two beds are used, installed in a manually changed “lead–lag” type system, in which the CO_2 flows through the beds in series. The bed life at design quantities of sulfur is estimated at approximately 250 to 300 days, after which the “lead” bed adsorbent is replaced and becomes the “lag” bed. Leaving the sulfur removal beds, the gas passes through a cartridge-type filter (F-2, Figure 4) that removes particulate matter from the gas stream.

Next, the CO_2 enters the carbon bed units (V-6A and V-6B, Figure 4), where any remaining trace sulfur and hydrocarbon contaminants are removed. The CO_2 then enters the dryer units (V-7A and V-7B, Figure 4), where the dew point (water content) is lowered to specification. Figure 6 shows an illustration of a molecular sieve-type dryer system in a CO_2 facility.

Both sets of beds are designed for a minimum 24-hour adsorption cycle with a nominal 16- to 18-hour regeneration cycle. The regeneration cycle is arranged so that the same regeneration gas used in the dryers is used for the carbon beds. One carbon bed is regenerated simultaneously with one dryer bed. A slipstream of the primary CO_2 compressed, dehydrated vapor stream is used for dryer and carbon bed regeneration. The regeneration gas system is set up so that a regeneration gas source is always available. The backup gas source, CO_2 vent vapors from the liquid CO_2 storage tanks, can be manually selected. If required by any nonstandard operating conditions, the dryers and carbon beds may be regenerated separately. The regeneration stream is heated to temperatures of approximately 450 °F (232 °C) via a heater (H-1, Figure 4), which is typically an electric or gas-fired heater. After passing through the dryer bed, the gas is reheated in an electric heater (if required; H-2, Figure 4) and then used to regenerate the carbon bed. This joint cycle saves on both heater power and the amount of regeneration gas required. Additional cartridge-type filters (F-3 and F-4, Figure 4) are installed after the carbon beds and after the dryers, respectively, to remove particulate matter from the gas stream.

The main CO_2 stream then flows to the reboiler (E-5A, Figure 4), providing heat to the reboiler. An auxiliary reboiler (E-5B, Figure 4) is also installed for use in conjunction with the main reboiler if the heat available in the main reboiler is insufficient. The auxiliary reboiler uses liquid ammonia as its heat source. The main CO_2 flow then enters the main CO_2 condenser (E-6, Figure 4), where most of the CO_2 vapor stream is condensed. The resulting two-phase effluent CO_2 stream from the main condenser is mixed with condensate from the distillation column vent condenser (E-7, Figure 4) and flows to the condenser separator (V-8). From this vessel, the liquid CO_2 is pumped by a column pump (P-1, Figure 4) to the distillation column (T-2, Figure 4). Vapor from the distillation column is mixed with vapor from the condenser separator and then flows to the vent condenser (E-7, Figure 4), where additional CO_2 is condensed by evaporating refrigerant on the shell side of the condenser. This liquid rejoins the main liquid CO_2 stream and flows back to the condenser separator. Vapor from the vent condenser flows to

Figure 6 Molecular sieve dryer beds in a CO_2 EOR facility. Photograph courtesy of Chaparral Energy.

the heat exchanger (E-9, Figure 4), where the cold vent stream is used to subcool the ammonia refrigerant before the CO_2 vent stream is discharged into the atmosphere.

Oxygen and nitrogen are stripped from the liquid CO_2 as the liquid CO_2 flows down the distillation column (T-2, Figure 4), countercurrent to the stripping vapor generated in the reboilers. The main CO_2 liquid stream then flows from the bottom of the distillation column to the reboilers (E-5A and E-5B, Figure 4). After purification in the column and reboilers, the liquid CO_2 then flows to the subcooler (E-8, Figure 4). The subcooler cools the liquid stream to storage conditions, and then the liquid CO_2 flows to the storage tanks. In many plants, an additional small heat exchanger (not shown) is used to heat vapors from the CO_2 storage tank when they are used to regenerate the carbon beds (V-6A and V-6B, Figure 4) and the dryer beds (V-7A and V-7B, Figure 4). Ammonia from the ammonia receiver is used to heat the vapors from the CO_2 storage tank in this additional small exchanger, and the additional subcooling of the ammonia refrigerant resulting from heating the CO_2 storage tank vapor improves the efficiency of the plant.

The refrigeration cycle also uses a two-stage compound screw compressor (C-2, Figure 5) for ammonia compression. This equipment greatly simplifies the system and makes it more compact as compared with the design in the 2006 report. The ammonia condenser (E-10, Figure 5) is an evaporative type with cooling water circulated by pump ECP-1. Ammonia flows inside the tubes of the exchanger while recirculated water is sprayed down on the tube bundle and forced air flows up through the tube bundle. The ammonia condensing temperature is approximately 95 °F (35 °C). The ammonia receiver (V-9, Figure 5) is sized to hold the entire charge of ammonia for pump-down (storage when the refrigeration unit is not operating). Subcooling of the refrigerant, which improves the efficiency of the refrigeration cycle, is achieved in exchangers E-4 and E-9 (Figure 5) before it is used for the low-temperature cooling applications. For more detail on exchangers E-4 and E-9, refer to the description of the CO_2 processing equipment in the previous section.

After subcooling, the ammonia refrigerant is flashed to the intermediate temperature, which is typically in the 40 to 50 °F (4 to 10 °C) range, and enters the high-pressure ammonia separator (VE-1, Figure 5). Liquid ammonia from the separator is evaporated in a CO_2 cooler (E-1, Figure 5) upstream of the main CO_2 compressor (C-1, Figure 3) and in a CO_2 cooler (E-3, Figure 5) downstream of the water scrubber (T-1, Figure 3). Vapor leaves these exchangers and returns to the ammonia separator (VE-1, Figure 5) before returning to the second stage of the ammonia compressor. Some refrigerant from the ammonia separator is further subcooled in an auxiliary reboiler (E-5B, Figure 5), and then passes through a valve to lower its pressure before entering the low-pressure ammonia separator (VE-6, Figure 5). Liquid ammonia from the ammonia separator at approximately –24 °F (–31 °C) is used for the lower ammonia pressure cooling services, which include the main CO_2 condenser (E-6, Figure 5), the distillation column vent condenser (E-7, Figure 5), and the product CO_2 subcooler (E-8, Figure 5). Ammonia vapor from these heat exchangers returns to the low-pressure ammonia separator (VE-6, Figure 5) and then to the first stage of the ammonia compressor.

These types of plants tend to lower the amount of ammonia stored or in use because of Process Safety Management requirements and general concerns regarding the hazards of ammonia, which are associated with its toxicity. Currently, a common method of cooling at these kinds of plants includes the use of a recirculated propylene glycol–water solution instead of ammonia for some cooling services to lower the amount of ammonia used and stored on-site. In this study, the compressor oil coolers (OC-1 and OC-2, Figure 5) and the CO_2 cooler (E-2, Figure 5) located upstream of the water scrubber (T-1, Figure 3) use glycol–water instead of ammonia for cooling. The glycol–water solution is cooled by an evaporative-type cooler for heat rejection before it returns to these heat exchangers.

The use of the recirculated glycol–water solution at 95 °F (35 °C) for these high-level cooling services in the plant eliminates the use of cooling tower water in all exchangers except for the blower aftercooler (BE-1, Figure 5, which is often in the ethanol plant), the main ammonia condenser (E-10, Figure 5), and the glycol–water evaporative cooler (E-11, Figure 5). Pump ECP-2 circulates the glycol–water solution in E-11, and pump P-3 is used to supply this solution to the heat exchangers. A balance tank is provided to accommodate changes in the volume of the glycol–water solution attributable to ambient temperature variations. Using the recirculated glycol–water solution instead of cooling tower water eliminates fouling in the exchangers that would otherwise be water-cooled and allows the use of smaller and lower cost fixed-tube bundle-type heat exchangers instead of the removable bundle types that would otherwise be required to facilitate heat exchanger cleaning. For wet CO_2 service, which requires stainless steel contact surfaces for the CO_2, this process also allows the use of stainless steel tubes (tube side only) and a carbon steel shell instead of a totally stainless steel heat exchanger, thus greatly lowering the cost of the heat exchanger. This nonfouling 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 two evaporative units. Although this system does require a second evaporative unit, the lowered maintenance on the other units quickly pays for the additional cost of adding the glycol–water evaporative cooler. Using the glycol–water cooler also reduces the chances for ammonium carbonate salt formation if there are tube leaks in the heat exchangers, and it reduces the amount of ammonia on-site, which can reduce environmental and safety compliance costs for the plant.

Additional details regarding the equipment required for CO_2 capture for the food and beverage grade cases are provided in Appendix A. The information provided in Appendix A includes preliminary equipment sizes and details, consumable requirements, electrical and labor requirements, other utility requirements, feed and product stream compositions, and applicable equipment design standards.

Nonfood and Nonbeverage Grade Cases

Food and beverage grade CO_2 is not required for enhanced oil and gas recovery or for sequestration. Lower purity CO_2 from natural or industrial sources is used for EOR in several regions in the United States. In years past, these plants were often relatively simple “flash”-type plants in which some of the equipment needed to produce food and beverage grade CO_2 was not required. Although equipment costs were lower, this style of plant typically required more power per ton (tonne) of CO_2 product because approximately 15% more feed CO_2 was required per ton (tonne) of product CO_2 owing to the higher vent gas flashing losses in the plant. However, in today’s market, the increasing costs of both power and raw feed gas, as well as the low oxygen content requirement of 10 to 20 ppmv for CO_2 entering many CO_2 pipelines, have reduced the use of the flash-type plant. As explained previously, these more recent developments led to a change for the nonfood and nonbeverage cases from a two-phase flash plant design in the 2006 report to a distillation-based plant design in this updated report.

Even though distillation has been incorporated into the plant design for the nonfood and nonbeverage grade CO_2 cases, some differences still exist in the equipment and operating costs because purification of CO_2 to meet specifications for food and beverage uses is not required. These differences are largely due to exclusion of the equipment required to remove sulfur compounds, hydrocarbons, and other organic contaminants for the food and beverage grade CO_2 cases. A water scrubber, sulfur removal beds, and carbon beds are required in the food and beverage grade CO_2 plant design, but they are not required for the nonfood and nonbeverage grade cases. Otherwise, the plant designs are the same.

Figures 7 and 8 show the process flow diagram for the equipment required for CO_2 capture for the nonfood and nonbeverage grade cases. The process flow diagram for the refrigeration equipment required for CO_2 capture for the nonfood and nonbeverage grade cases is identical to that for the food and beverage grade cases (see Figure 5). A detailed description of this equipment follows. The sizing and types of units are preliminary and are subject to confirmation after further process engineering. The basic design assumptions are based on previous experience with ethanol CO_2 recovery and may require modification after gas analysis is performed on an actual source. Temperatures, pressures, and other parameters in the following description are approximate. These plants would typically be designed for unattended operation using a PLC unit or, in some cases, a small DCS. Plant operator preference governs this decision.

Lubricant-injected rotary screw compressors have been selected for the main compression services. In general, screw compressors suit this size of facility and provide lower maintenance costs than do reciprocating compressors because of their rotary movement and smaller number of 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. The two-stage screw compressors (often termed “compound compressors”) now have a much greater size range than when the 2006 report was issued, allowing a much broader application range than previously. As discussed, this usually results in a reduction in capital and operating costs, and all cases in this report are based on the use of two-stage (compound) screw compressors.

Flow rates for each case are based on actual anticipated capacities for specific equipment models; thus, rates differ somewhat relative to the nominal design rates. Feed rates, nominal rates, and actual product rates are summarized in

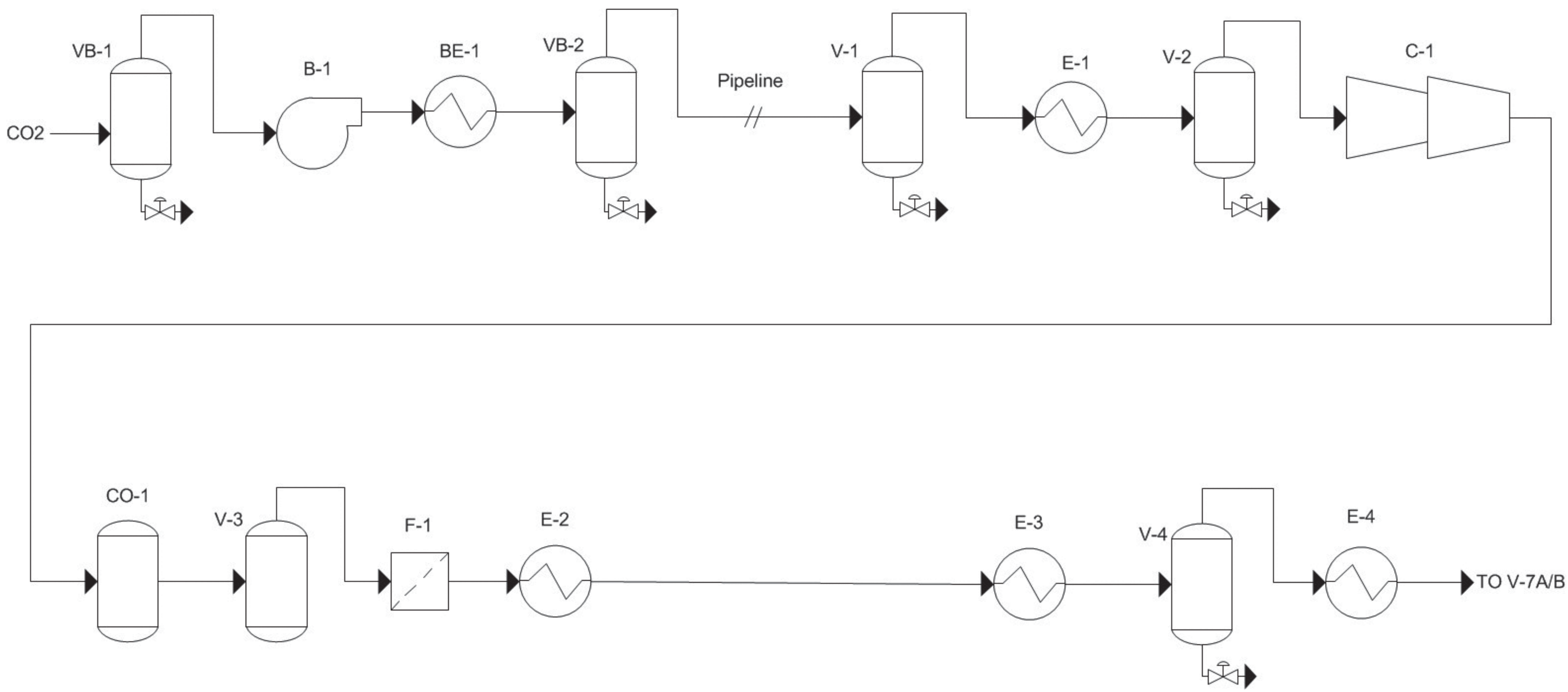

Figure 7 Carbon dioxide process flow diagram for nonfood and nonbeverage grade cases (1 of 2).

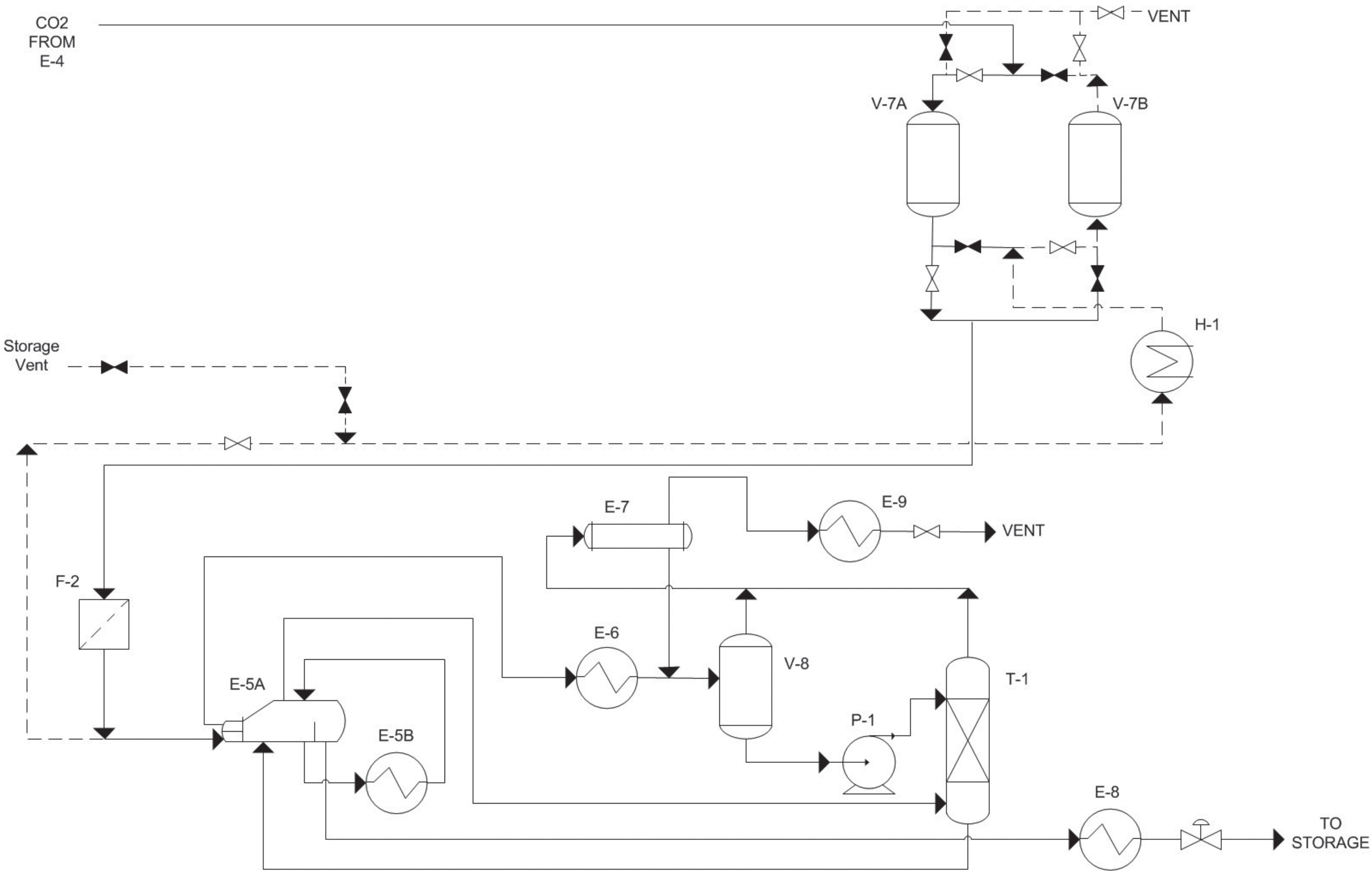

Figure 8 Carbon dioxide process flow diagram for nonfood and nonbeverage grade cases (2 of 2).

Table 2 Product rates for the nonfood and nonbeverage grade CO_2 cases

	Nominal product rate, ton/day (tonne/day)		
Rate	75 (68)	300 (272)	1,000 (907)
Feed rate, lb/h (kg/h)	7,006 (3,178)	28,025 (12,715)	93,417 (42,385)
Actual product rate, lb/h (kg/h)	6,376 (2,892)	25,126 (11,407)	89,232 (40,486)
Actual product rate, ton/day (tonne/day)	77 (69)	302 (274)	1,071 (972)

Table 2. The feed and product rates are the same as for the food and beverage grade cases because the main differences are sulfur and hydrocarbon removal vessels, which are not required for the nonfood and nonbeverage cases. These steps do not materially affect CO_2 product recovery rates.

The 75 ton/day (68 tonne/day) and 300 ton/day (272 tonne/day) facilities are designed with one train consisting of one CO_2 compressor and one ammonia (refrigerant) compressor. The 1,000 ton/day (907 tonne/day) facility is based on two compressor trains or two 50% compressors for both services. Screw compressors large enough to allow for a single-train facility are available for this capacity, but their size and horsepower make them difficult to install on a package system.

For design purposes, the vapor produced by warming of the stored CO_2 is assumed to have an average flow rate of 200 lb/h (90.7 kg/h) for the 75 ton/day (68 tonne/day) facility, 500 lb/h (227 kg/h) for the 300 ton/day (272 tonne/day) facility, and 900 lb/h (408 kg/h) for the 1,000 ton/day (907 tonne/day) facility. This assumption is made to account for heat gain in the storage tanks and the effects of truck loading. Recompression and recycling of these vapors add slightly to the horsepower and electricity requirements for the facilities. The high stage of the CO_2 compressor would be used to remove and recompress vapors from the CO_2 storage tanks to maintain the pressure in the storage tanks.

The plants are designed to accept the CO_2 from the source at 13.5 psia (0.9 atm) and 100 °F (38 °C). The incoming gas from the host plant enters a two-phase separator (VB-1) in which any mechanically entrained water is separated from the gas stream and the water is discharged for treatment or disposal. The gas is then compressed to approximately 28.5 psia (1.9 atm) by a multistage centrifugal-type blower (B-1) and cooled in a heat exchanger (BE-1) by using recirculated cooling tower water from the host plant. After passing through a second two-phase separator (VB-2) that removes any additional condensed water, the raw gas then enters a pipeline to the CO_2 plant inlet.

In most of today's plants, the blower unit is installed in the host ethanol plant near the final scrubber of the host ethanol plant. This blower unit serves several purposes: (1) it allows use of a smaller pipeline from the host plant to the CO_2 facility; (2) it requires smaller main compression units because of the lower actual volumetric suction flow rate at the inlet to the main compression unit, thus lowering the compressor size and cost; and (3) the blower and compressor power requirements combined are usually lower than the compressor power requirements without a blower.

At the CO_2 plant inlet, the CO_2 enters a phase separator (V-1, Figure 7) to remove any moisture that condensed in the pipeline. The CO_2 then enters a refrigerant-cooled shell and tube exchanger (E-1, Figure 7), where the CO_2 is cooled to lower the water content and volumetric flow rate and to prevent moisture condensation in the two-stage CO_2 compressor (C-1, Figure 7). The condensed water is separated in a phase separator (V-2, Figure 7), and the gas is then compressed to approximately 315 psia (21 atm) in the two-stage CO_2 compressor (C-1, Figure 7). The CO_2 discharge gas enters CO-1 (Figure 7), a high-efficiency oil coalescer, and then flows into V-3 (Figure 7), a carbon bed arranged to remove residual oil from the CO_2 gas stream to very low (ppbv) levels. A cartridge-type filter (F-1, Figure 7) then removes particulate matter from the CO_2 gas stream. Next, the CO_2 is cooled in a water-cooled aftercooler (E-2, Figure 7) and a refrigerant-cooled aftercooler (E-3, Figure 7) in series before it goes to a separator (V-4, Figure 7) to remove the condensed water. The gas then flows through a superheater (E-4, Figure 7), which uses liquid ammonia to slightly warm the CO_2 to minimize the chance of moisture condensation in the adsorbent beds.

The slightly superheated CO_2 then enters the dryer units (V-7A and V-7B, Figure 8), where the dew point (water content) is lowered to specification. The dryer beds are designed for a minimum 24-hour adsorption cycle with a nominal 16- to 18-hour regeneration cycle. A slipstream of the primary compressed, dehydrated CO_2 vapor stream is used for dryer bed regeneration. The regeneration gas system is set up so that a regeneration gas source is always available. The backup gas source, CO_2 vent vapors from the liquid CO_2 storage tanks, can be manually selected. The regeneration stream is heated to temperatures of approximately 450 °F (232 °C) via a heater (H-1, Figure 8), which is typically an electric or gas-fired heater. A cartridge-type filter (F-2, Figure 8) is installed after the dryer beds to remove particulates from the CO_2 gas stream.

The main CO_2 stream then flows to the reboiler (E-5A, Figure 8), providing heat to the reboiler by cooling the main gas stream. An auxiliary reboiler (E-5B, Figure 8) is also installed for use in conjunction with the main reboiler if the heat available in the main reboiler is insufficient. The auxiliary reboiler uses liquid ammonia as its heat source. The main CO_2 flow then enters the main CO_2 condenser (E-6, Figure 8), where most of the CO_2 vapor stream is condensed. The resulting two-phase effluent CO_2 stream from the main condenser is mixed with condensate from the distillation column vent condenser (E-7, Figure 8) and flows to the condenser separator (V-8, Figure 8). From this vessel, the liquid CO_2 is pumped to the distillation column (T-1, Figure 8). Vapor from the distillation column is mixed with vapor from the condenser separator and then flows to the vent condenser (E-7, Figure 8), where additional liquid is condensed. This liquid flows back to the condenser separator and rejoins the main liquid CO_2 stream. Vapor from the vent condenser flows to the heat exchanger (E-9, Figure 8), where the cold vent stream is used to subcool the ammonia refrigerant before the CO_2 vent stream is discharged to the atmosphere.

Oxygen and nitrogen are stripped from the liquid CO_2 as the liquid CO_2 flows down the distillation column, countercurrent to the stripping vapor generated in the reboilers. The main CO_2 liquid stream then flows from the bottom of the distillation column to the reboilers. After purification in the column and reboilers, the liquid CO_2 then flows to the subcooler (E-8, Figure 8). The subcooler cools the liquid stream to storage conditions, and the liquid CO_2 flows to the storage tanks. In many plants, an additional small heat exchanger (not shown) is used to heat vapors from the CO_2 storage tanks when they are used to regenerate the dryer beds (V-7A and V-7B, Figure 8). Ammonia from the ammonia receiver is used to heat the vapors from the CO_2 storage tank in this additional small exchanger, and the additional subcooling of the ammonia refrigerant resulting from heating the CO_2 storage tank vapor improves the efficiency of the plant.

The refrigeration cycle for the nonfood and nonbeverage grade CO_2 cases is identical to that of the food and beverage grade cases, so the description is not repeated in this section. This includes the use of both a glycol–water and an ammonia-based cooling system.

Appendix B provides additional details regarding the equipment required for CO_2 capture for the nonfood and nonbeverage grade cases. The information provided in Appendix B includes preliminary equipment sizes and details, consumable requirements, electrical and labor requirements, other utility requirements, feed and product stream compositions, and applicable equipment design standards.

COST ESTIMATE SUMMARY FOR CARBON DIOXIDE RECOVERY EQUIPMENT

Budgetary cost estimates for new equipment for 75 ton/day (68 tonne/day), 300 ton/day (272 tonne/day), and 1,000 ton/day (907 tonne/day) are provided in this section for food and beverage grade CO_2 or nonfood and nonbeverage grade CO_2 cases. The cost of electricity is a critical factor in the economic viability of CO_2 recovery. 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, but comparisons on a basis of $0.055/kWh are also provided in the Executive Summary of this report. Additional details regarding electrical requirements and other consumable materials are provided in Appendix A for the food and beverage grade cases and in Appendix B for the nonfood and nonbeverage grade cases.

Food and Beverage Grade Cases

The estimated purchased equipment cost for the 75 ton/day (68 tonne/day) food and beverage grade CO_2 case is $2,048,530. This cost estimate includes a $36,500 freight allowance. Storage, as described in the following equipment list, would be an additional $360,800 based on using two 120 ton (109 tonne) capacity, factory-insulated tanks. The other equipment and related items included in this cost estimate that would be necessary for 75 ton/day (68 tonne/day) CO_2 capture and food and beverage grade purification are as follows:

- Engineering for typical installation
- Site work
- Truck scale
- Metal building with approximate dimensions of 40 × 70 ft (12 × 21 m) with an 18 ft (5.5 m) eave height
- Control room, manager's office, and driver area
- Electrical gear, including motor starters and associated switch gear
- Three days of total storage capacity based on two units, each with a 120 ton (109 tonne) capacity

The cost of installation is estimated at $2,266,715, which, when combined with the $2,048,530 in equipment and $360,800 for storage tanks, gives a total installed equipment cost of $4,676,045 or $62,347/ton ($68,765/tonne) of nominal daily capacity. Purchased equipment costs and installation costs were developed internally by using a bottom-up method based on estimating costs for each equipment component and aspect of facility construction. The estimated facility costs were then validated based on discussions with companies that have built CO_2 liquefaction facilities in the past few years. These costs may vary considerably depending on construction labor costs, site conditions and suitability, contractor availability, distance from the source, and other site-specific items.

The estimated purchased equipment cost for the 300 ton/day (272 tonne/day) food and beverage CO_2 grade case is $4,679,750. This cost estimate includes a $103,000 freight allowance. Storage, as described in the following equipment list, would be an additional $1,375,000. The other equipment included in this cost estimate that would be necessary for 300 ton/day (272 tonne/day) of CO_2 capture and food and beverage grade purification is as follows:

- Engineering for typical installation
- Site work
- Truck scale
- Metal building with approximate dimensions of 60 × 120 ft (18 × 37 m) with a 22 ft (6.7 m) 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)
- Three days of total storage capacity based on two units, each with a 500 ton (454 tonne) capacity

The cost of installation is estimated at $4,430,615, which, with the $4,679,750 in equipment and $1,375,000 in storage cost, gives a total installed equipment cost of $10,485,365 or $34,951/ton ($38,549/tonne) of nominal daily capacity. Purchased equipment costs and installation costs were developed internally by using a bottom-up method based on estimating costs for each equipment component and aspect of facility construction. The estimated facility costs

Table 3 Cost estimate summary for the food and beverage grade cases

	Nominal capacity, ton/day (tonne/day)		
Cost	75 (68)	300 (272)	1,000 (907)
Purchased equipment cost, $	2,048,530	4,679,750	11,820,750
Storage cost, $	360,800	1,375,000	2,700,000
Installation cost, $	2,266,715	4,430,615	7,142,760
Total installed equipment cost, $	4,676,045	10,485,365	21,663,510
Total installed equipment cost, $/ton of nominal daily capacity ($/tonne)	62,347 (68,765)	34,951 (38,549)	21,664 (23,885)

were then validated based on discussions with companies that have built CO_2 liquefaction facilities in the past few years. These costs may vary considerably depending on construction labor costs, site conditions and suitability, contractor availability, distance from the source, and other site-specific items.

The estimated purchased equipment cost for the 1,000 ton/day (907 tonne/day) food and beverage grade CO_2 case is $11,820,750. This cost estimate includes a $194,000 freight allowance. Storage, as described in the following equipment list, would be an additional $2,700,000. The other equipment included in this cost estimate that would be necessary for 1,000 ton/day (907 tonne/day) of CO_2 capture and food and beverage grade purification is as follows:

- Engineering for typical installation
- Site work
- Truck scale
- Metal building with approximate dimensions of 60 × 140 ft (18 × 43 m) with a 22 ft (6.7 m) 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)
- Two or more days of total storage capacity based on at least four units, each with a 500 ton (454 tonne) capacity

The cost of installation is estimated at $7,142,760, which, with the $11,820,750 in equipment and $2,700,000 in storage, gives a total installed equipment cost of $21,663,510 or $21,664/ton ($23,885/tonne) of nominal daily capacity. Purchased equipment costs and installation costs were developed internally using a bottom-up method based on estimating costs for each equipment component and aspect of facility construction. The estimated facility costs were then validated based on discussions with companies that have built CO_2 liquefaction facilities in the past few years. These costs may vary considerably depending on construction labor costs, site conditions and suitability, contractor availability, distance from the source, and other site-specific items. Table 3 provides a summary of the cost estimates for the food and beverage cases for each of the three nominal capacities.

Nonfood and Nonbeverage Grade Cases

The estimated purchased equipment cost for the 75 ton/day (68 tonne/day) nonfood and nonbeverage grade CO_2 case is $1,684,210. This cost estimate includes a $36,500 freight allowance. Storage, as described in the following equipment list, would be an additional $360,800. Trailers used for food and beverage grade CO_2 transportation cannot be used for nonfood and nonbeverage grade CO_2 transportation, which could lead to additional costs. The other equipment and related items included in this cost estimate that would be necessary for 75 ton/day (68 tonne/day) of CO_2 capture and nonfood and nonbeverage grade purification are as follows:

- Engineering for typical installation
- Site work
- Truck scale
- Metal building with approximate dimensions of 40 × 70 ft (12 × 21 m) with a 18 ft (5.5 m) eave height
- Control room, manager's office, and driver area
- Electrical gear, including motor starters and associated switch gear
- Three days of storage in two units, each with a 120 ton (109 tonne) capacity

The cost of installation is estimated at $2,251,480, which, with the $1,684,210 in equipment and $360,800 in storage, gives a total installed cost of $4,296,490, or a cost of $57,287/ton ($63,184/tonne) of nominal daily capacity. Purchased equipment costs and installation costs were developed internally by using a bottom-up method based on estimating costs for each equipment component and aspect of facility construction. The estimated facility costs were then validated based on discussions with companies that have built CO_2 liquefaction facilities in the past few years. These costs may vary considerably depending on construction labor costs, site conditions and suitability, contractor availability, distance from the source, and other site-specific items.

The estimated purchased equipment cost for the 300 ton/day (272 tonne/day) nonfood and nonbeverage grade CO_2 case is $4,069,250. This cost estimate includes a freight allowance of $103,000. Trailers used for food and beverage grade CO_2 transportation cannot be used for nonfood and nonbeverage grade CO_2 transportation, which could lead to additional costs. Storage, as described in the following equipment list, would be an additional $1,375,000. The other equipment included in this cost estimate that would be necessary for 300 ton/day (272 tonne/day) of CO_2 capture and nonfood and nonbeverage grade purification is as follows:

- Engineering for typical installation
- Site work
- Truck scale
- Metal building with approximate dimensions of 60 × 120 ft (18 × 37 m) with a 22 ft (6.7 m) 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)
- Three days of storage in two units, each with a 500 ton (454 tonne) capacity

The cost of installation is estimated at $4,338,325, which, with the $4,069,250 in equipment and $1,375,000 in storage, gives a total installed cost of $9,782,575, or a cost of $32,609/ton ($35,965/tonne) of nominal daily capacity. Purchased equipment costs and installation costs were developed internally by using a bottom-up method based on estimating costs for each equipment component and aspect of facility construction. The estimated facility costs were then validated based on discussions with companies that have built CO_2 liquefaction facilities in the past few years. These costs may vary considerably depending on construction labor costs, site conditions and suitability, contractor availability, distance from the source, and other site-specific items.

The estimated purchased equipment cost for the 1,000 ton/day (907 tonne/day) nonfood and nonbeverage grade CO_2 case is $10,538,250. This cost estimate includes a freight allowance of $194,000. Trailers used for food and beverage grade CO_2 transportation cannot be used for nonfood and nonbeverage grade CO_2 transportation, which could lead to additional costs. Storage, as described in the following equipment list, would be an additional $2,700,000. The other equipment included in this cost estimate that would be necessary for 1,000 ton/day (907 tonne/day) of CO_2 capture and nonfood and nonbeverage grade purification is as follows:

- Engineering for typical installation
- Site work
- Truck scale
- Metal building with approximate dimensions of 60 × 140 ft (18 × 43 m) with a 22 ft (6.7 m) 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)
- Two or more days of total storage capacity based on at least four units, each with a 500 ton (454 tonne) capacity

The installation cost is estimated at $6,978,760, which, with the $10,538,250 in equipment and $2,700,000 in storage, gives a total installed equipment cost of $20,217,010 or $20,217/ton ($22,290/tonne) of nominal daily capacity. Purchased equipment costs and installation costs were developed internally by using a bottom-up method based on estimating costs for each equipment component and aspect of facility construction. The estimated facility costs were then validated based on discussions with companies that have built CO_2 liquefaction facilities in the past few years. These costs may vary considerably depending on construction labor costs, site conditions and suitability, contractor availability, distance from the source, and other site-specific items. Table 4 provides a summary of the cost estimates for the nonfood and nonbeverage cases for each of the three nominal capacities.

COST ESTIMATE FOR ADDING EQUIPMENT TO GET CARBON DIOXIDE INTO A PIPELINE

The pressure of the liquid CO_2 generated by these kinds of facilities is generally approximately 315 to 415 psia (21 to 28 atm). If a pipeline became available near an existing facility of this type, equipment could be added to raise the pressure to feed the CO_2 into the pipeline. Figure 9 shows a multistage centrifugal pump used to feed CO_2 into a pipeline. Pipeline inlet pressures vary depending on the diameter, length, flow rate, and surface injection pressure at the EOR field but can be expected to be in the range of 1,015 to 3,015 psia (69 to 205 atm). A multistage centrifugal pump is often used to boost CO_2 to the pressures needed to enter a pipeline. These pumps contain approximately 30 stages (impellers) on a single shaft. They are equipped with a motor and often come with a variable-frequency drive that is used to change the speed of the pump to control suction pressure, other process parameters, or both. These pumps are inexpensive compared with other compression options and are energy efficient. However, it is critical to maintain the process parameters, such as suction density and discharge pressure, for these pumps within acceptable limits. Otherwise, mechanical failure of the pumps can occur. In some cases, a simple, low-cost centrifugal booster pump is added upstream of the multistage centrifugal pump to ensure that the liquid CO_2 entering the multistage centrifugal pump is vapor free.

Table 4 Cost estimate summary for the nonfood and nonbeverage grade cases

	Nominal capacity, ton/day (tonne/day)		
Cost	75 (68)	300 (272)	1,000 (907)
Purchased equipment cost, $	1,684,210	4,069,250	10,538,250
Storage cost, $	360,800	1,375,000	2,700,000
Installation cost, $	2,251,480	4,338,325	6,978,760
Total installed equipment cost, $	4,296,490	9,782,575	20,217,010
Total installed equipment cost, $/ton of nominal daily capacity ($/tonne)	57,287 (63,184)	32,609 (35,965)	20,217 (22,290)

Figure 9 Multistage centrifugal pump used to feed CO_2 into a pipeline. Photograph courtesy of Chaparral Energy.

The flow rate for the 1,000 ton/day (907 tonne/day) facility could be handled by one multistage centrifugal pump. This system would be installed downstream of the reboilers and upstream of refrigerant subcooler E-8 (see Figure 8). The estimated purchased equipment costs for one of these systems is $237,583. This cost estimate is based on a vendor quote for a recent, similar facility. The estimated total installed capital cost to add this pumping system to the 1,000 ton/day (907 tonne/day) facility is $515,165. This estimate includes the following items:

- Booster pump with motor
- One multistage centrifugal pump with motor and variable-frequency drive
- Recycle valve installed downstream of the multistage centrifugal pump for facility capacity control
- Engineering
- Installation

For a discharge pressure of 2,015 psia (137 atm), the power requirements for the pumping system are estimated at 195 hp (145 kW). At a purchased electricity cost of $0.10/kWh, annual power costs for the pumping system would be $127,020. At a purchased electricity cost of $0.055/kWh, annual power costs for the pumping system would be $69,861. The booster pump and multistage centrifugal pump together would raise the temperature of the CO_2 by approximately 15 °F (8 °C). This results in some reduction in the efficiency improvement that subcooler E-8 provides for the refrigeration system.

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APPENDIX A: EQUIPMENT DETAILS FOR CARBON DIOXIDE CAPTURE: FOOD AND BEVERAGE GRADE CASES

The following tables summarize the preliminary design of the equipment, consumable materials, and energy required for the 75 ton/day (68 tonne/day; Tables A1–A5), 300 ton/day (272 tonne/day; Tables A6–A10), and 1,000 ton/day (907 tonne/day; Tables A11–A15) CO_2 capture and food and beverage grade purification cases.

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 1 for all pressure vessels
- Tubular Exchanger Manufacturers Association, Inc. 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 15-2010 Safety Code for Mechanical Refrigeration for the ammonia system
- National Electrical Manufacturers Association for electric motors and enclosures

Inlet and Outlet Gas Composition

The inlet conditions assumed are 13.5 psia (0.9 atm) at 100 °F (38 °C) at the inlet separator, saturated with water vapor. The design atmospheric pressure is 14.7 psia (1.0 atm). Table A16 summarizes the typical inlet stream conditions for CO_2 recovery and food and beverage grade purification from an ethanol plant. This represents a "typical" analysis with concentration ranges of components usually found in the raw CO_2 gas from an ethanol plant. The gas analysis will typically vary over a period of days because of differences in corn batches, types of enzyme, and fermentation cycles. The oxygen–nitrogen quantity analysis may vary daily depending on factors such as the number of fermenters in the alcohol plant and alcohol plant operations.

Table A17 shows typical product specifications based on guidelines from the International Society of Beverage Technologists, which is widely used as the acceptable product standard by many companies purchasing liquid CO_2. In some cases, however, companies purchasing CO_2 may have their own maximum limits on components that are more stringent than those shown in the table.

Utility Water Requirements

The ammonia evaporative condenser will require approximately 15 U.S. gallons/min (gpm; 57 L/min [lpm]) for the 75 ton/day (68 tonne/day) facility, 36 gpm (136 lpm) for the 300 ton/day (272 tonne/day) facility, and 86 gpm (326 lpm) for the 1,000 ton/day (907 tonne/day) facility. Typically, recommended blowdown is the same; thus, total water consumption for the ammonia evaporative condenser would be approximately 30 gpm (114 lpm) for the 75 ton/day (68 tonne/day) facility, 72 gpm (273 lpm) for the 300 ton/day (272 tonne/day) facility, and 172 gpm (651 lpm) for the 1,000 ton/day (907 tonne/day) facility. Blowdown rates may vary depending on the type of water treatment utilized.

The glycol–water evaporative condenser will require approximately 12 U.S. gpm (45 lpm) for the 75 ton/day (68 tonne/day) facility, 30 gpm (114 lpm) for the 300 ton/day (272 tonne/day) facility, and 108 gpm (409 lpm) for the 1,000 ton/day (907 tonne/day) facility. Typically, recommended blowdown is the same; thus, total water consumption for the glycol–water condenser would be approximately 24 gpm (91 lpm) for the 75 ton/day (68 tonne/day) facility, 60 gpm (227 lpm) for the 300 ton/day (272 tonne/day) facility, and 216 gpm (818 lpm) for the 1,000 ton/day (907 tonne/day) facility. Blowdown rates may vary depending on the type of water treatment utilized.

Potable water for the scrubber is estimated at approximately 8 gpm (30 lpm) for the 75 ton/day (68 tonne/day) facility, 20 gpm (76 lpm) for the 300 ton/day (272 tonne/day) facility, and 40 gpm (151 lpm) for the 1,000 ton/day (907 tonne/day) facility.

Instrument Air Requirements

The instrument air requirement is approximately 25 scf/h (standard cubic feet per hour [0.7 m^3/h]) for the 75 ton/day (68 tonne/day) facility as well as for the 300 ton/day (272 tonne/day) facility and approximately 40 scf/h (1.1 m^3/h) for the 1,000 ton/day (907 tonne/day) facility. The system would be designed to use CO_2 vapors from the storage tanks instead of instrument air under normal, steady-state operations.

Table A1 Vessel details for a 75 ton/day (68 tonne/day) food and beverage grade facility[1]

Tag no.	Description	Internal components	Diameter, inches	Seam/seam, inches	Material of construction	Design pressure, psig
VB-1	Blower inlet separator	Demister	36	84	304 SS	50
VB-2	Blower aftercooler separator	Demister	24	72	304 SS	50
V-1	Plant inlet separator	Demister	30	72	304 SS	50
V-2	Precooler separator	Demister	30	72	304 SS	50
CO-1	Coalescer	Coalescing elements	8	36	CS	350
V-3	Carbon oil absorber	Johnson screens	18	72	CS	350
V-4	Aftercooler separator	Demister	14	48	CS	350
V-5A/B	Sulfur removal beds	Johnson screens	30	156	CS	350
V-6A/B	Carbon beds	Johnson screens	24	120	CS	350
V-7A/B	Dryer beds	Johnson screens	24	120	CS	350
V-8	Condenser separator	Vortex breaker	20	96	CS	350
V-9	Ammonia receiver	Dip tube	24	216	CS	250
VE-1	Ammonia separator for E-1, E-3	Demister	12	144	CS	250
VE-6	Ammonia separator for E-6, E-7, E-8	Demister	14	192	CS	250
T-1	Water scrubber	Packing and supports, distributor	16	216	304 SS	350
T-2	Distillation column	Packing and supports, distributor	16	360	CS	350

[1]psig, pounds per square inch gauge; SS, stainless steel; CS, carbon steel.

Table A2 Heat exchanger details for a 75 ton/day (68 tonne/day) food and beverage grade facility[1]

Tag no.	Service	Exchanger details	Duty, Btu/h	Surface area, sq. ft	Material and design pressure–tube side, psi	Material and design pressure–shell side, psi
OC-1	Oil cooler for C-1	By compressor vendor	1,011,700	TBD	CS, 150	CS, 400
OC-2	Oil cooler for C-2	By compressor vendor	349,700	TBD	CS, 150	CS, 300
BE-1	Blower aftercooler	BEM 17-144	218,436	552	304 SS, 150	304 SS, 150
E-1	Precooler	BEM 10-144	166,543	197	304 SS, 150	CS, 250
E-2	CO_2 compressor aftercooler	BEM 10-144	161,768	197	304 SS, 350	CS,150
E-3	Refrigerant-cooled CO_2 aftercooler	BEM 8-120	95,899	106	304 SS, 350	CS, 250
E-4	CO_2 superheater	BEM 6-96	17,813	41	CS, 350	CS, 250
E-5A	Main reboiler	BKU 8/16-120	77,769	77	CS, 350	CS, 350
E-5B	Auxiliary reboiler	BEM 6-96	91,432	41	CS, 250	CS, 350
E-6	Main condenser	BEM 15-192	831,097	637	CS, 350	CS, 250
E-7	Vent condenser	BEM 10-144	104,459	197	CS, 350	CS, 250
E-8	CO_2 subcooler	BEM 6-120	50,629	52	CS, 350	CS, 250
E-9	Vent gas refrigerant subcooler	Jacketed pipe	12,285	9	CS, 350	CS, 250
E-10	Ammonia refrigerant condenser	BAC model no. PC-2111-0412-10	1,413,000			
E-11	Glycol–water plant cooling	BAC model no. FXV-0806B-28D-K	1,523,000			

[1]psi, pounds per square inch; TBD, to be determined; CS, carbon steel; SS, stainless steel.

Table A3 Utility requirements for a 75 ton/day (68 tonne/day) food and beverage grade facility[1]

Tag no.	Item, units	Full load capacity	Swept volume, cfm	Connected motor hp	Required bhp at design	Voltage	Motor efficiency, %	Power at design, kW
B-1	Centrifugal blower, HSI or equivalent, lb/h	7,144	TBD	125	110	480/3/60	95	86
C-1	CO_2 compressor, Mycom model 2016 SSC, lb/h	7,072	570 LS, 292 HS	350	293	480/3/60	95	230
OP-1	Oil pump, gpm	80		5	4	480/3/60	94	3
C-2	NH_3 compressor, Mycom model 2016 MSC, tons refrigeration	84	710 LS, 292 HS	250	200	480/3/60	95	157
OP-2	Oil pump, gpm	44		3	3	480/3/60	94	2
P-1	Column pump, gpm	17		5	4	480/3/60	92	3
P-2	Scrubber water pump, gpm	10		5	3	480/3/60	90	2
	CO_2 product loading pumps, gpm	200		20	1	480/3/60	90	1
	Evaporative condenser fans, BAC PC2-2111-0412-10			10	9	480/3/60	95	7
	Evaporative condenser pump			1.5	1	480/3/60	95	1
	Closed-circuit cooler fan, BAC FXV-0806B 28D-K			9	9	480/3/60	95	7
	Cooler water pump			2	1	480/3/60	95	1
	Coolant circulating pump			20	15	480/3/60	95	12
H-1	Dryer regenerator heater, kW	25		34	5	480/3/60		4
H-2	Carbon regenerator heater, kW	25		34	1	480/3/60		1
	Dryer regenerator flow, lb/h	634						
	Dryer regenerator flow, MMscfd	0.1328						
	Total connected hp			806				
	bhp[2]				659			
	Total kW at design							517
	CO_2 to storage							
	lb/h	6,376						
	ton/day	76.51						
	tonne/day	69.43						
	MMscfd	1.3357						
	kWh/ton	162.26						
	kWh/tonne	178.82						

[1]cfm, cubic feet per minute; hp, horsepower; bhp, brake horsepower; LS, low stage; HS, high stage; gpm, gallons per minute; MMscfd, million standard cubic feet per day; TBD, to be determined.

[2]Includes dryer kilowatts converted to brake horsepower for power calculation purposes.

Table A4 Operating costs for a 75 ton/day (68 tonne/day) food and beverage grade facility[1,2]

Item	Amount	Units	Unit cost, $	Units	Cost/ton, $
Variable manufacturing cost items					
Power	162.26	kWh/ton	0.100	kWh	16.23
Desiccant (estimated 4-year life), including disposal	0.020	lb/ton	6.00	lb	0.12
Water makeup for evaporative condenser and cooler	13	gpm	0.10	Mgal	0.02
Water makeup for scrubber	8	gpm	0.10	Mgal	0.003
Sulfur removal (based on GTS), including disposal	0.17	lb/ton	0.72	lb	0.12
Carbon (estimated 2.5-year life), including disposal	0.015	lb/ton	6.50	lb	0.10
Water blowdown disposal, returned to host plant	14	gpm	0.05	Mgal	0.01
Variable manufacturing cost					16.61
Labor and overhead cost items					
Labor and benefits					
Plant manager	0.25		80,000	year	0.81
Shift foremen	0.25		65,000	year	0.66
Shift operator	0.50		50,000	year	1.02
Base labor cost			61,250	year	
Benefits factor of 40%			24,500	year	1.00
Total labor and benefits			85,750	year	3.48
Overhead					
Maintenance	2.00	% of capital	93,521	year	3.72
Taxes	1.00	% of capital	46,760	year	1.86
Insurance	1.50	% of capital	70,141	year	2.79
Labor and overhead costs					11.86
Total manufacturing cost					28.47
Principal and interest					
Principal and interest costs			418,360	year	16.99
Estimated total cost/ton					45.46

[1]Background specifications: Total installed cost: $4,676,045; interest rate/year: 6.50%; years: 20; type of plant: food and beverage grade; production capacity (ton/day): 75; annual operating days: 335; annual tons produced: 24,623; percent nameplate capacity: 98.00%; average ton/day: 74.
[2]gpm, gallons per minute; Mgal, thousand gallons; GTS, HydroCAT type GTS 2007.

Table A5 Consumables for a 75 ton/day (68 tonne/day) food and beverage grade facility[1]

Tag no.	Function	Component required, size or units	Quantity
CO-1	Oil removal	Coalescing elements manufactured by Balston, Parker, and Zander. Material of construction: borosilicate glass fiber with carbon steel retainers and fluorocarbon O-rings; size: 3.23 inches OD/1.72 inches ID × 25 inches long	~4
V-3	Oil removal	Typical coconut shell or coal-based activated carbon, e.g., carbon type 208 C or equivalent, lb	300
V-5A/B	Sulfur removal	Mixed-metal oxide formed on a carbon or alumina substrate, e.g., HydroCAT type GTS 2007 or equivalent, lb	3,900 × 2 units
V-6A/B	Final polish and cleanup	Typical coconut shell or coal-based activated carbon, e.g., 2/3 carbon type CJ and 1/3 carbon type 208 C or equivalents, lb	700 × 2 units + 300 × 2 units
V-7A/B	Aldehyde and moisture removal	Activated alumina adsorbent, e.g., Selexsorb type CD or equivalent, lb	1,100 × 2 units
V-9	Refrigeration	Ammonia, lb	1,650

[1]OD, outside diameter; ID, inside diameter.

Table A6 Vessel details for a 300 ton/day (272 tonne/day) food and beverage grade facility[1]

Tag no.	Description	Internal components	Diameter, inches	Seam/seam, inches	Material of construction	Design pressure, psig
VB-1	Blower inlet separator	Demister	72	180	304 SS	50
VB-2	Blower aftercooler separator	Demister	48	120	304 SS	50
V-1	Plant inlet separator	Demister	48	120	304 SS	50
V-2	Precooler separator	Demister	48	120	304 SS	50
CO-1	Coalescer	Coalescing elements	16	48	CS	350
V-3	Carbon oil absorber	Johnson screens	36	120	CS	350
V-4	Aftercooler separator	Demister	30	60	CS	350
V-5A/B	Sulfur removal beds	Johnson screens	60	180	CS	350
V-6A/B	Carbon beds	Johnson screens	42	120	CS	350
V-7A/B	Dryer beds	Johnson screens	42	120	CS	350
V-8	Condenser separator	Vortex breaker	30	72	CS	350
V-9	Ammonia receiver	Dip tube	42	288	CS	250
VE-1	Ammonia separator for E-1, E-3	Demister	20	144	CS	250
VE-6	Ammonia separator for E-6, E-7, E-8	Demister	30	192	CS	250
T-1	Water scrubber	Packing and supports, distributor	30	192	304 SS	350
T-2	Distillation column	Packing and supports, distributor	30	360	CS	350

[1]psig, pounds per square inch gauge; SS, stainless steel; CS, carbon steel.

Table A7 Heat exchanger details for a 300 ton/day (272 tonne/day) food and beverage grade facility[1]

Tag no.	Service	Exchanger details	Duty, Btu/h	Surface area, sq. ft	Material and design pressure–tube side, psi	Material and design pressure–shell side, psi
OC-1	Oil cooler for C-1	By compressor vendor	2,068,400	TBD	CS, 150	CS, 400
OC-2	Oil cooler for C-2	By compressor vendor	1,288,400	TBD	CS, 150	CS, 300
BE-1	Blower aftercooler	BEM 27-192	905,077	2,116	304 SS, 150	304 SS, 150
E-1	Precooler	BEM 17-192	690,064	827	304 SS, 150	CS, 250
E-2	CO_2 compressor aftercooler	BEM 17-192	670,278	827	304 SS, 350	CS,150
E-3	Refrigerant-cooled CO_2 aftercooler	BEM 15-144	397,353	475	304 SS, 350	CS, 250
E-4	CO_2 superheater	BEM 12-60	71,250	120	CS, 350	CS, 250
E-5A	Main reboiler	BKU 10/20-216	322,230	223	CS, 350	CS, 350
E-5B	Auxiliary reboiler	BEM 10-144	378,842	197	CS, 250	CS, 350
E-6	Main condenser	BEM 27-240	3,443,604	2,656	CS, 350	CS, 250
E-7	Vent condenser	BEM 17-192	432,822	827	CS, 350	CS, 250
E-8	CO_2 subcooler	BEM 13-192	209,777	467	CS, 350	CS, 250
E-9	Vent gas refrigerant subcooler	BEM 6-120	49,140	52	CS, 350	CS, 250
E-10	Ammonia refrigerant condenser	BAC model no. PC2-374-1212-30	6,042,279			
E-11	Glycol–water plant cooling	BAC model no. FXV-0812B-16D-M	4,932,155			

[1]TBD, to be determined; CS, carbon steel; SS, stainless steel.

Table A8 Utility requirements for a 300 ton/day (272 tonne/day) food and beverage grade facility[1,2]

Tag no.	Item, units	Full load capacity	Swept volume, cfm	Connected motor hp	Required bhp at design	Voltage	Motor efficiency, %	Power at design, kW
B-1	Centrifugal blower, HSI or equivalent, lb/h	28,000	TBD	350	318	480/3/60	96	247
C-1	CO_2 compressor, Mycom model 3225 SSC, lb/h	27,181	2,240 LS, 1,119 HS	1,250	1,059	4,160/3/60	96	823
OP-1	Oil pump, gpm	260		10	8	480/3/60	95	6
C-2	NH_3 compressor, Mycom model 3225 MSC, tons refrigeration	349	2,790 LS, 1,110 HS	900	834	4,160/3/60	96	648
OP-2	Oil pump, gpm	100		7.5	6	480/3/60	95	5
P-1	Column pump, gpm	65		7.5	4	480/3/60	95	3
P-2	Scrubber water pump, gpm	35		15	11	480/3/60	95	9
	CO_2 product loading pumps, gpm	175		20	5	480/3/60	95	4
	Evaporative condenser fans, BAC PC2-374-1212-30			30	26	480/3/60	95	20
	Evaporative condenser pump			5	4	480/3/60	95	3
	Closed-circuit cooler fan, BAC FXV-0812B-16D-M			20	18	480/3/60	95	14
	Cooler water pump			5	4	480/3/60	95	3
	Coolant circulating pump			35	35	480/3/60	95	27
H-1	Dryer regenerator heater, kW	80		107	40	480/3/60		30
H-2	Carbon regenerator heater, kW	80		107	8	480/3/60		6
	Dryer regenerator flow, lb/h	2,153						
	Dryer regenerator flow, MMscfd	0.4510						
	Total connected hp			2,655				
	Average bhp[2]				2,380			
	kW at design							1,848
	CO_2 to storage							
	lb/h	25,142						
	ton/d	301.70						
	tonne/d	273.78						
	MMscfd	5.2671						
	kWh/ton	147.04						
	kWh/tonne	162.03						

[1]cfm, cubic feet per minute; hp, horsepower; bhp, brake horsepower; LS, low stage; HS, high stage; gpm, gallons per minute; MMscfd, million standard cubic feet per day; TBD, to be determined.

[2]Includes dryer kilowatts converted to brake horsepower for power calculation purposes.

Table A9 Operating costs for a 300 ton/day (272 tonne/day) food and beverage grade facility[1,2]

Item	Amount	Units	Unit cost, $	Units	Cost/ton, $
Variable manufacturing cost items					
Power	147.04	kWh/ton	0.100	kWh	14.70
Desiccant (estimated 4-year life), including disposal	0.020	lb/ton	6.00	lb	0.12
Water makeup for evaporative condenser and cooler	52	gpm	0.10	Mgal	0.02
Water makeup for scrubber	20	gpm	0.10	Mgal	0.01
Sulfur removal (based on GTS), including disposal	0.17	lb/ton	0.72	lb	0.12
Carbon (estimated 2.5-year life), including disposal	0.015	lb/ton	6.50	lb	0.10
Water blowdown disposal, returned to host plant	46	gpm	0.05	Mgal	0.01
Variable manufacturing cost					15.09
Labor and overhead cost items					
Labor and benefits					
Plant manager	0.25		80,000	year	0.20
Shift foremen	0.5		65,000	year	0.33
Shift operator	1		50,000	year	0.51
Base labor cost			102,500	year	
Benefits factor of 40%			41,000	year	0.42
Total labor and benefits			143,500		1.46
Overhead					
Maintenance	2.00	% of capital	209,707	year	2.09
Taxes	1.00	% of capital	104,854	year	1.04
Insurance	1.50	% of capital	157,280	year	1.56
Labor and overhead costs					6.15
Total manufacturing cost					21.24
Principal and interest					
Principal and interest costs			938,113	year	9.52
Estimated total cost/ton					30.77

[1]Background specifications: Total installed cost: $10,485,365; interest rate/year: 6.50%; years: 20; type of plant: food and beverage grade; production capacity (ton/day): 300; annual operating days: 335; annual tons produced: 98,490; percent nameplate capacity: 98.00%; average ton/day: 294.
[2]gpm, gallons per minute; Mgal, thousand gallons; GTS, HydroCAT type GTS 2007.

Table A10 Consumables for a 300 ton/day (272 tonne/day) food and beverage grade facility[1]

Tag no.	Function	Component required, size or units	Quantity
CO-1	Oil removal	Coalescing elements manufactured by Balston, Parker, and Zander. Material of construction: borosilicate glass fiber with carbon steel retainers and fluorocarbon O-rings; size: 3.23 inches OD/1.72 inches ID × 25 inches long	~12
V-3	Oil removal	Typical coconut shell or coal-based activated carbon, e.g., carbon type 208 C or equivalent, lb	2,500
V-5A/B	Sulfur removal	Mixed metal oxide formed on a carbon or alumina substrate, e.g., HydroCAT type GTS 2007 or equivalent, lb	14,000 × 2 units
V-6A/B	Final polish and cleanup	Typical coconut shell or coal-based activated carbon, e.g., 2/3 carbon type CJ and 1/3 carbon type 208 C or equivalents, lb	3,600 × 2 units
V-7A/B	Aldehyde and moisture removal	Activated alumina adsorbent, e.g., Selexsorb type CD or equivalent, lb	2,000 × 2 units + 1,200 × 2 units
V-9	Refrigeration	Ammonia, lb	4,800

[1]OD, outside diameter; ID, inside diameter.

Table A11 Vessel details for a 1,000 ton/day (907 tonne/day) food and beverage grade facility[1]

Tag no.	Description	Internal components	Diameter, inches	Seam/seam, inches	Material of construction	Design pressure, psig
VB-1	Blower inlet separator	Demister	120	180	304 SS	50
VB-2	Blower aftercooler separator	Demister	96	120	304 SS	50
V-1	Plant inlet separator	Demister	96	120	304 SS	50
V-2	Precooler separator	Demister	96	120	304 SS	50
CO-1A	Coalescer	Coalescing elements	20	48	CS	350
CO-1B	Coalescer	Coalescing elements	20	48	CS	350
V-3A	Carbon oil absorber	Johnson screens	36	96	CS	350
V-3B	Carbon oil absorber	Johnson screens	36	96	CS	350
V-4	Aftercooler separator	Demister	42	60	CS	350
V-5A/B	Sulfur removal beds	Johnson screens	108	180	CS	350
V-6A/B	Carbon beds	Johnson screens	66	148	CS	350
V-7A/B	Dryer beds	Johnson screens	66	148	CS	350
V-8	Condenser separator	Vortex breaker	54	144	CS	350
V-9	Ammonia receiver	Dip tube	48	288	CS	250
VE-1	Ammonia separator for E-1, E-3	Demister	30	144	CS	250
VE-6	Ammonia separator for E-6, E-7, E-8	Demister	36	192	CS	250
T-1	Water scrubber	Packing and supports, distributor	48	192	304 SS	350
T-2	Distillation column	Packing and supports, distributor	48	360	CS	350

[1]psig, pounds per square inch gauge; SS, stainless steel; CS, carbon steel.

Table A12 Heat exchanger details for a 1,000 ton/day (907 tonne/day) food and beverage grade facility[1]

Tag no.	Service	Exchanger details	Duty, Btu/h	Surface area, sq. ft	Material and design pressure–tube side, psi	Material and design pressure–shell side, psi
OC-1A	Oil cooler for C-1A	By compressor vendor	3,445,400	TBD	CS, 150	CS, 400
OC-1B	Oil cooler for C-1B	By compressor vendor	3,445,400	TBD	CS, 150	CS, 400
OC-2A	Oil cooler for C-2A	By compressor vendor	1,979,000	TBD	CS, 150	CS, 300
OC-2B	Oil cooler for C-2B	By compressor vendor	1,979,000	TBD	CS, 150	CS, 300
BE-1	Blower aftercooler	BEM 31-240	2,963,536	3,563	304 SS, 150	304 SS, 150
E-1	Precooler	BEM 27-240	2,259,507	2,656	304 SS, 150	CS, 250
E-2	CO_2 compressor aftercooler	BEM 27-240	2,194,723	2,656	304 SS, 350	CS,150
E-3	Refrigerant-cooled CO_2 aftercooler	BEM 21-216	1,301,073	1,436	304 SS, 350	CS, 250
E-4	CO_2 superheater	BEM 15-144	237,500	475	CS, 350	CS, 250
E-5A	Main reboiler	BKU 17/31-192	1,055,092	782	CS, 350	CS, 350
E-5B	Auxiliary reboiler	BEM15-168	1,240,461	556	CS, 250	CS, 350
E-6	Main condenser	BEM 39-288	11,275,550	6,260	CS, 350	CS, 250
E-7	Vent condenser	BEM 19-192	1,417,208	2,785	CS, 350	CS, 250
E-8	CO_2 subcooler	BEM 21-216	686,883	1,426	CS, 350	CS, 250
E-9	Vent gas refrigerant subcooler	BEM 12-48	30,713	96	CS, 350	CS, 250
E-10	Ammonia refrigerant condenser	BAC model no. PC2-948-1268-60	20,464,521			
E-11	Glycol–water plant cooling	BAC model no. FXV-0812C-16Q-O	13,043,523			

[1]TBD, to be determined; CS, carbon steel; SS, stainless steel.

Table A13 Utility requirements for a 1,000 ton/day (907 tonne/day) food and beverage grade facility[1]

Tag no.	Item, units	Full load capacity	Swept volume, cfm	Connected motor hp	Required bhp at design	Voltage	Motor efficiency, %	Power at design, kW
B-1	Centrifugal blower, HSI or equivalent, lb/h	97,804	TBD	1,250	1,190	4,160/3/60	97	920
C-1A	CO_2 compressor, Mycom model 3225 LLSC, lb/h	46,315	3,970 LS, 1,660 HS	2,000	1,833	4,160/3/60	97	1,417
OP-1A	Oil pump, gpm	365		30	19	480/3/60	95	15
C-1B	CO_2 compressor, Mycom model 3225 LLSC, lb/h	46,315	3,970 LS, 1,660 HS	2,000	1,833	4,160/3/60	97	1,417
OP-1B	Oil pump, gpm	365		30	19	480/3/60	95	15
C-2A	NH_3 compressor, Mycom model 3225 LSC, tons refrigeration	548	3,359 LS, 1,119 HS	1,250	1,084	4,160/3/60	97	838
OP-2A	Oil pump, gpm	215		15	9	480/3/60	94	7
C-2B	NH_3 compressor, Mycom model 3225 LSC, tons refrigeration	548	3,359 LS, 1,119 HS	1,250	1,084	4,160/3/60	97	838
OP-2B	Oil pump, gpm	215		15	9	480/3/60	94	7
P-1	Column pump, gpm	225		20	14	480/3/60	94	11
P-2	Scrubber water pump, gpm	90		30	26	480/3/60	94	21
	CO_2 product loading pumps, gpm	175		40	36	480/3/60	95	28
	Evaporative condenser fans			60	50	480/3/60	95	39
	Evaporative condenser pump			30	28	480/3/60	95	22
	Closed-circuit cooler fan			75	70	480/3/60	95	55
	Cooling water pump			15	12	480/3/60	95	9
	Coolant circulating pump			100	76	480/3/60	95	60
H-1	Dryer regenerator heater, kW	225		302	70	480/3/60		53
H-2	Carbon regenerator heater, kW	225		302	15	480/3/60		11
	Dryer regenerator flow, lb/h	6,323						
	Dryer regenerator flow, MMscfd	1.3246						
	Total connected hp			8,210				
	bhp[2]				7,479			
	Total kW at design							5,782
	CO_2 to storage							
	lb/h	89,232						
	ton/day	1,070.78						
	tonne/day	971.67						
	MMscfd	18.69						
	kWh/ton	129.60						
	kWh/tonne	142.82						

[1]cfm, cubic feet per minute; hp, horsepower; bhp, brake horsepower; LS, low stage; HS, high stage; gpm, gallons per minute; MMscfd, million standard cubic feet per day; TBD, to be determined.
[2]Includes dryer kilowatts converted to brake horsepower for power calculation purposes.

Table A14 Operating costs for a 1,000 ton/day (907 tonne/day) food and beverage grade facility[1]

Item	Amount	Units	Unit cost, $	Units	Cost/ton, $
Variable manufacturing cost items					
Power	129.6	kWh/ton	0.100	kWh	12.96
Desiccant (estimated 4-year life), including disposal	0.020	lb/ton	6.00	lb	0.12
Water makeup for evaporative condenser and cooler	172	gpm	0.10	Mgal	0.02
Water makeup for scrubber	40	gpm	0.10	Mgal	0.02
Sulfur removal (based on GTS), including disposal	0.17	lb/ton	0.72	lb	0.12
Carbon (estimated 2.5-year life), including disposal	0.015	lb/ton	6.50	lb	0.10
Water blowdown disposal, returned to host plant	126	gpm	0.05	Mgal	0.01
Variable manufacturing cost					13.36
Labor and overhead cost items					
Labor and benefits					
Plant manager	0.5		80,000	year	0.12
Shift foremen	1		65,000	year	0.20
Shift operator	1		50,000	year	0.15
Base labor cost			155,000	year	
Benefits factor of 40%			62,000	year	0.19
Total labor and benefits			217,000	year	0.66
Overhead					
Maintenance	2.00	% of capital	433,270	year	1.29
Taxes	1.00	% of capital	216,635	year	0.65
Insurance	1.50	% of capital	324,953	year	0.97
Labor and overhead costs					3.57
Total manufacturing cost					16.93
Principal and interest					
Principal and interest costs			1,938,208	year	5.90
Estimated total cost/ton					22.83

[1]Background specifications: Total installed cost: $21,663,510; interest rate/year: 6.50%; years: 20; type of plant: food and beverage grade; production capacity (ton/day): 1,000; annual operating days: 335; annual tons produced: 328,300; percent nameplate capacity: 98.00%; average ton/day: 980.

[2]gpm, gallons per minute; Mgal, thousand gallons; GTS, HydroCAT type GTS 2007.

Table A15 Consumables for a 1,000 ton/day (907 tonne/day) food and beverage grade facility[1]

Tag no.	Function	Component required	Quantity
CO-1	Oil removal	Coalescing elements manufactured by Balston, Parker, and Zander. Material of construction: borosilicate glass fiber with carbon steel retainers and fluorocarbon O-rings; size: 3.23 inches OD/1.72 inches ID × 25 inches long	~14 × 2 units
V-3	Oil removal	Typical coconut shell or coal-based activated carbon, e.g., carbon type 208 C or equivalent, lb	2,100 × 2 units
V-5A/B	Sulfur removal	Mixed-metal oxide formed on a carbon or alumina substrate, e.g., HydroCAT type GTS 2007 or equivalent, lb	50,000 × 2 units
V-6A/B	Final polish and cleanup	Typical coconut shell or coal-based activated carbon, e.g., 2/3 carbon type CJ and 1/3 carbon type 208 C or equivalents, lb	8,700 × 2 units
V-7A/B	Aldehyde and moisture removal	Activated alumina adsorbent, e.g., Selexsorb type CD or equivalent, lb	6,000 × 2 units + 3,000 × 2 units
V-9	Refrigeration	Ammonia, lb	7,700

[1]OD, outside diameter; ID, inside diameter.

Table A16 Typical inlet gas composition[1]

Component	Concentration range, ppmv
Moisture	Saturated
Acetaldehyde	3–75
Methanol	1–50
Ethanol	25–950
Acetone	0–2.5
Ethyl acetate	2–30
n-Propanol	0–1.0
i-Butanol	0–3
n-Butanol	0.5–1.0
Isoamyl acetate	0.6–3.0
Hydrogen sulfide	1–5
Dimethyl sulfide	0.5–1.5
Nitrogen	50–600
Oxygen	10–100
Methane	0–3
Carbon dioxide	Balance

[1]ppmv, parts per million by volume.

Table A17 Typical product specification limits[1]

Item	Limit
Purity of CO_2	99.9% vol. min.
Moisture (H_2O)	20 ppmv max.
Oxygen (O_2)	30 ppmv max.
Carbon monoxide (CO)	10 ppmv max.
Ammonia (NH_3)	2.5 ppmv max.
Nitrogen monoxide (NO)	2.5 ppmv max.
Nitrogen dioxide (NO_2)	2.5 ppmv max.
Nonvolatile residue (NVR)	10 ppmw max.
Nonvolatile organic residue (NVOR)	5 ppmw max.
Methanol (MeOH)	10 ppmv max.
Total volatile hydrocarbons (THC, as methane)	50 ppmv max. (including 20 ppmv max. as total nonmethane hydrocarbons)
Acetaldehyde (AA)	0.2 ppmv max.
Aromatic hydrocarbon (AHC)	20 ppbv max.
Total sulfur content (TSC as S) (total sulfur-containing impurities, excluding sulfur dioxide)	0.1 ppmv max.
Sulfur dioxide (SO_2)	1 ppmv
Odor of solid CO_2 (snow)	No foreign odor
Appearance of solid CO_2 (snow)	No foreign appearance
Odor and taste in water	No foreign odor or taste
Appearance in water	No color or turbidity

[1]ppmv, parts per million by volume; ppmw, parts per million by weight; ppbv, parts per billion by volume.

APPENDIX B: EQUIPMENT DETAILS FOR CARBON DIOXIDE CAPTURE: NONFOOD AND NONBEVERAGE GRADE CASES

The following tables summarize the preliminary design of the equipment, consumable materials, and energy required for the 75 ton/day (68 tonne/day, Tables B1–B5), 300 ton/day (272 tonne/day, Tables B6–B10), and 1,000 ton/day (907 tonne/day, Tables B11–B15) CO_2 capture and nonfood and nonbeverage grade purification cases.

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 1 for all pressure vessels
- Tubular Exchanger Manufacturers Association, Inc. 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 15-2010 Safety Code for Mechanical Refrigeration for the ammonia system
- National Electrical Manufacturers Association for electric motors and enclosures

Inlet and Outlet Gas Composition

The inlet conditions assumed are 13.5 psia (0.9 atm) at 100 °F (38 °C) at the inlet separator, saturated with water vapor. The design atmospheric pressure is 14.7 psia (1.0 atm). Table B16 summarizes the typical inlet stream conditions for CO_2 recovery and nonfood and nonbeverage grade purification from an ethanol plant. This represents a "typical" analysis with concentration ranges of components usually found in the raw CO_2 gas from an ethanol plant. The gas analysis will typically vary over a period of days because differences in corn batches, types of enzyme, and fermentation cycles. The oxygen–nitrogen quantity analysis may vary daily depending on factors such as the number of fermenters in the alcohol plant and alcohol plant operations.

Product specifications for nonfood and nonbeverage grade plants will vary based on the use and the user. Typical pipeline specifications for minimum water content of approximately 30 lb of H_2O/MMscf of CO_2 (633 ppmv) and oxygen of 10 to 20 ppmv may be encountered. A plant that liquefies CO_2 will have a much lower water content than the pipeline specification. In some cases, however, companies purchasing CO_2 may have their own maximum limits on components that are more stringent than the typical pipeline specifications.

Utility Water Requirements

The ammonia evaporative condenser will require approximately 15 U.S. gpm (57 lpm) for the 75 ton/day (68 tonne/day) facility, 36 gpm (136 lpm) for the 300 ton/day (272 tonne/day) facility, and 86 gpm (326 lpm) for the 1,000 ton/day (907 tonne/day) facility. Typically, recommended blowdown is the same; thus, total water consumption for the ammonia evaporative condenser would be approximately 30 gpm (114 lpm) for the 75 ton/day (68 tonne/day) facility, 72 gpm (273 lpm) for the 300 ton/day (272 tonne/day) facility, and 172 gpm (651 lpm) for the 907 tonne/day (1,000 ton/day) facility. Blowdown rates may vary depending on the type of water treatment utilized.

The glycol–water evaporative condenser will require approximately 12 U.S. gpm (45 lpm) for the 75 ton/day (68 tonne/day) facility, 30 gpm (114 lpm) for the 300 ton/day (272 tonne/day) facility, and 108 gpm (409 lpm) for the 1,000 ton/day (907 tonne/day) facility. Typically, recommended blowdown is the same; thus, total water consumption for the glycol–water condenser would be approximately 24 gpm (91 lpm) for the 75 ton/day (68 tonne/day) facility, 60 gpm (227 lpm) for the 300 ton/day (272 tonne/day) facility, and 216 gpm (818 lpm) for the 1,000 ton/day (907 tonne/day) facility. Blowdown rates may vary depending on the type of water treatment utilized.

Potable water for the scrubber is estimated at approximately 8 gpm (30 lpm) for the 75 ton/day (68 tonne/day) facility, 20 gpm (76 lpm) for the 300 ton/day (272 tonne/day) facility, and 40 gpm (151 lpm) for the 1,000 ton/day (907 tonne/day) facility.

Instrument Air Requirements

The instrument air requirement is approximately 25 scf/h (0.7 m^3/h) for the 75 ton/day (68 tonne/day) facility as well as for the 300 ton/day (272 tonne/day) facility and approximately 40 scf/h (1.1 m^3/h) for the 1,000 ton/day (907 tonne/day) facility. The system would be designed to use CO_2 vapors from the storage tanks instead of instrument air under normal, steady-state operations.

Table B1 Vessel details for a 75 ton/day (68 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Description	Internal components	Diameter, inches	Seam/seam, inches	Material of construction	Design pressure, psig
VB-1	Blower inlet separator	Demister	36	84	304 SS	50
VB-2	Blower aftercooler separator	Demister	24	72	304 SS	50
V-1	Plant inlet separator	Demister	30	72	304 SS	50
V-2	Precooler separator	Demister	30	72	304 SS	50
CO-1	Coalescer	Coalescing elements	8	36	CS	350
V-3	Carbon oil absorber	Johnson screens	18	72	CS	350
V-4	Aftercooler separator	Demister	14	48	CS	350
V-7A/B	Dryer beds	Johnson screens	24	120	CS	350
V-8	Condenser separator	Vortex breaker	20	96	CS	350
V-9	Ammonia receiver	Dip tube	24	216	CS	250
VE-1	Ammonia separator for E-1, E-3	Demister	12	144	CS	250
VE-6	Ammonia separator for E-6, E-7, E-8	Demister	14	192	CS	250
T-1	Distillation column	Packing and supports, distributor	16	360	CS	350

[1]psig, pounds per square inch gauge; SS, stainless steel; CS, carbon steel.

Table B2 Heat exchanger details for a 75 ton/day (68 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Service	Exchanger details	Duty, Btu/h	Surface area, sq. ft	Material and design pressure–tube side, psi	Material and design pressure—shell side, psi
OC-1	Oil cooler for C-1	By compressor vendor	1,011,700	TBD	CS, 150	CS, 400
OC-2	Oil cooler for C-2	By compressor vendor	349,700	TBD	CS, 150	CS, 300
BE-1	Blower aftercooler	BEM 17-144	218,436	552	304 SS, 150	304 SS, 150
E-1	Precooler	BEM 10-144	166,543	197	304 SS, 150	CS, 250
E-2	CO_2 compressor aftercooler	BEM 10-144	161,768	197	304 SS, 350	CS,150
E-3	Refrigerant-cooled CO_2 aftercooler	BEM 8-120	95,899	106	304 SS, 350	CS, 250
E-4	CO_2 superheater	BEM 6-96	17,813	41	CS, 350	CS, 250
E-5A	Main reboiler	BKU 8/16-120	77,769	77	CS, 350	CS, 350
E-5B	Auxiliary reboiler	BEM 6-96	91,432	41	CS, 250	CS, 350
E-6	Main condenser	BEM 15-192	831,097	637	CS, 350	CS, 250
E-7	Vent condenser	BEM 10-144	104,459	197	CS, 350	CS, 250
E-8	CO_2 subcooler	BEM 6-120	50,629	52	CS, 350	CS, 250
E-9	Vent gas refrigerant subcooler	Jacketed pipe	12,285	9	CS, 350	CS, 250
E-10	Ammonia refrigerant condenser	BAC model no. PC-2111-0412-10	1,413,000			
E-11	Glycol–water plant cooling	BAC model no. FXV-0806B-28D-K	1,523,000			

[1] psi, pounds per square inch; TBD, to be determined; CS, carbon steel; SS, stainless steel.

Table B3 Utility requirements for a 75 ton/day (68 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Item, units	Full load capacity	Swept volume, cfm	Connected motor hp	Required bhp at design	Voltage	Motor efficiency, %	Power at design, kW
B-1	Centrifugal blower, HSI or equivalent, lb/h	7,144	TBD	125	110	480/3/60	95	86
C-1	CO_2 compressor, Mycom model 2016 SSC, lb/h	7,072	570 LS, 292 HS	350	293	480/3/60	95	230
OP-1	Oil pump, gpm	80		5	4	480/3/60	94	3
C-2	NH_3 compressor, Mycom model 2016 MSC, tons refrigeration	84	710 LS, 292 HS	250	200	480/3/60	95	157
OP-2	Oil pump, gpm	44		3	3	480/3/60	94	2
P-1	Column pump, gpm	17		5	4	480/3/60	92	3
P-2	Scrubber water pump, gpm	10		5	3	480/3/60	90	2
	CO_2 product loading pumps, gpm	200		20	1	480/3/60	90	1
	Evaporative condenser fans, BAC PC2-2111-0412-10			10	9	480/3/60	95	7
	Evaporative condenser pump			1.5	1	480/3/60	95	1
	Closed-circuit cooler fan, BAC FXV-0806B 28D-K			9	9	480/3/60	95	7
	Cooler water pump			2	1	480/3/60	95	1
	Coolant circulating pump			20	15	480/3/60	95	12
H-1	Dryer regenerator heater, kW	25		34	5	480/3/60		4
H-2	Carbon regenerator heater, kW	25		34	1	480/3/60		1
	Dryer regenerator flow, lb/h	634						
	Dryer regenerator flow, MMscfd	0.1328						
	Total connected hp			806				
	bhp[2]				658			
	Total kW at design							516
	CO_2 to storage							
	lb/h	6,376						
	ton/day	76.51						
	tonne/day	69.43						
	MMscfd	1.3357						
	kWh/ton	162.03						
	kWh/tonne	178.55						

[1]cfm, cubic feet per minute; hp, horsepower; bhp, brake horsepower; LS, low stage; HS, high stage; gpm, gallons per minute; MMscfd, million standard cubic feet per day; TBD, to be determined.

[2]Includes dryer kilowatts converted to brake horsepower for power calculation purposes.

Table B4 Operating costs for a 75 ton/day (68 tonne/day) nonfood and nonbeverage grade facility[1,2]

Item	Amount	Units	Unit cost, $	Units	Cost/ton, $
Variable manufacturing cost items					
Power	162.03	kWh/ton	0.100	kWh	16.20
Desiccant (estimated 4-year life), including disposal	0.020	lb/ton	6.00	lb	0.12
Water makeup for evaporative condenser and cooler	13	gpm	0.10	Mgal	0.02
Water makeup for scrubber	0	gpm	0.10	Mgal	0.00
Sulfur removal (based on GTS), including disposal	0.00	lb/ton	0.72	lb	0.00
Carbon (estimated 2.5-year life), including disposal	0.000	lb/ton	6.50	lb	0.00
Water blowdown disposal, returned to host plant	6	gpm	0.05	Mgal	0.01
Variable manufacturing cost					16.35
Labor and overhead cost items					
Labor and benefits					
Plant manager	0.25		80,000	year	0.81
Shift foremen	0.25		65,000	year	0.66
Shift operator	0.50		50,000	year	1.02
Base labor cost			61,250	year	
Benefits factor of 40%			24,500	year	1.00
Total labor and benefits			85,750	year	3.48
Overhead					
Maintenance	2.00	% of capital	85,930	year	3.42
Taxes	1.00	% of capital	42,965	year	1.71
Insurance	1.50	% of capital	64,447	year	2.57
Labor and overhead costs					11.18
Total manufacturing cost					27.53
Principal and interest					
Principal and interest costs			384,402	year	15.61
Estimated total cost/ton					43.14

[1]Background specifications: Total installed cost: $4,296,490; interest rate/year: 6.50%; years: 20; type of plant: nonfood and nonbeverage grade; production capacity (ton/day): 75; annual operating days: 335; annual tons produced: 24,623; percent nameplate capacity: 98.00%; average ton/day: 74.

[2]gpm, gallons per minute; Mgal, thousand gallons; GTS, HydroCAT type GTS 2007.

Table B5 Consumables for a 75 ton/day (68 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Function	Component required, size or units	Quantity
CO-1	Oil removal	Coalescing elements manufactured by Balston, Parker, and Zander. Material of construction: borosilicate glass fiber with carbon steel retainers and fluorocarbon O-rings; size: 3.23 inches OD/1.72 inches ID × 25 inches long	~4
V-3	Oil removal	Typical coconut shell or coal-based activated carbon, e.g., carbon type 208 C or equivalent, lb	300
V-7A/B	Aldehyde and moisture removal	Activated alumina adsorbent, e.g., Selexsorb, type CD or equivalent, lb	1,100 × 2 units
V-9	Refrigeration	Ammonia, lb	1,650

[1]OD, outside diameter; ID, inside diameter.

Table B6 Vessel details for a 300 ton/day (68 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Description	Internal components	Diameter, inches	Seam/seam, inches	Material of construction	Design pressure, psig
VB-1	Blower inlet separator	Demister	72	180	304 SS	50
VB-2	Blower aftercooler separator	Demister	48	120	304 SS	50
V-1	Plant inlet separator	Demister	48	120	304 SS	50
V-2	Precooler separator	Demister	48	120	304 SS	50
CO-1	Coalescer	Coalescing elements	16	48	CS	350
V-3	Carbon oil absorber	Johnson screens	36	120	CS	350
V-4	Aftercooler separator	Demister	30	60	CS	350
V-7A/B	Dryer beds	Johnson screens	42	120	CS	350
V-8	Condenser separator	Vortex breaker	30	72	CS	350
V-9	Ammonia receiver	Dip tube	42	288	CS	250
VE-1	Ammonia separator for E-1, E-3	Demister	20	144	CS	250
VE-6	Ammonia separator for E-6, E-7, E-8	Demister	30	192	CS	250
T-1	Distillation column	Packing and supports, distributor	30	360	CS	350

[1]psig, pounds per square inch gauge; SS, stainless steel; CS, carbon steel.

Table B7 Heat exchanger details for a 300 ton/day (272 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Service	Exchanger details	Duty, Btu/h	Surface area, sq. ft	Material and design pressure–tube side, psi	Material and design pressure–shell side, psi
OC-1	Oil cooler for C-1	By compressor vendor	2,068,400	TBD	CS, 150	CS, 400
OC-2	Oil cooler for C-2	By compressor vendor	1,288,400	TBD	CS, 150	CS, 300
BE-1	Blower aftercooler	BEM 27-192	905,077	2,116	304 SS, 150	304 SS, 150
E-1	Precooler	BEM 17-192	690,064	827	304 SS, 150	CS, 250
E-2	CO_2 compressor aftercooler	BEM 17-192	670,278	827	304 SS, 350	CS,150
E-3	Refrigerant-cooled CO_2 aftercooler	BEM 15-144	397,353	475	304 SS, 350	CS, 250
E-4	CO_2 superheater	BEM 12-60	71,250	120	CS, 350	CS, 250
E-5A	Main reboiler	BKU 10/20-216	322,230	223	CS, 350	CS, 350
E-5B	Auxiliary reboiler	BEM 10-144	378,842	197	CS, 250	CS, 350
E-6	Main condenser	BEM 27-240	3,443,604	2,656	CS, 350	CS, 250
E-7	Vent condenser	BEM 17-192	432,822	827	CS, 350	CS, 250
E-8	CO_2 subcooler	BEM 13-192	209,777	467	CS, 350	CS, 250
E-9	Vent gas refrigerant subcooler	BEM 6-120	49,140	52	CS, 350	CS, 250
E-10	Ammonia refrigerant condenser	BAC model no. PC2-374-1212-30	6,042,279			
E-11	Gycol–water plant cooling	BAC model no. FXV-0812B-16D-M	4,932,155			

[1]TBD, to be determined; CS, carbon steel; SS, stainless steel.

Table B8 Utility requirements for a 300 ton/day (272 tonne/day) nonfood and nonbeverage grade facility[1,2]

Tag no.	Item, units	Full load capacity	Swept volume, cfm	Connected motor hp	Required bhp at design	Voltage	Motor efficiency, %	Power at design, kW
B-1	Centrifugal blower, HSI or equivalent, lb/h	28,000	TBD	350	318	480/3/60	96	247
C-1	CO_2 compressor, Mycom model 3225 SSC, lb/h	27,181	2,240 LS, 1,119 HS	1,250	1,059	4,160/3/60	96	823
OP-1	Oil pump, gpm	260		10	8	480/3/60	95	6
C-2	NH_3 compressor, Mycom model 3225 MSC, tons refrigeration	349	2,790 LS, 1,110 HS	900	834	4,160/3/60	96	648
OP-2	Oil pump, gpm	100		7.5	6	480/3/60	95	5
P-1	Column pump, gpm	65		7.5	4	480/3/60	95	3
P-2	Scrubber water pump, gpm	35		15	11	480/3/60	95	9
	CO_2 product loading pumps, gpm	175		20	5	480/3/60	95	4
	Evaporative condenser fans, BAC PC2-374-1212-30			30	26	480/3/60	95	20
	Evaporative condenser pump			5	4	480/3/60	95	3
	Closed-circuit cooler fan, BAC FXV-0812B-16D-M			20	18	480/3/60	95	14
	Cooler water pump			5	4	480/3/60	95	3
	Coolant circulating pump			35	35	480/3/60	95	27
H-1	Dryer regenerator heater, kW	80		107	40	480/3/60		30
H-2	Carbon regenerator heater, kW	80		107	8	480/3/60		6
	Dryer regenerator flow, lb/h	2,153						
	Dryer regenerator flow, MMscfd	0.4510						
	Total connected hp			2,655				
	Average bhp[2]				2,372			
	kW at design							1,842
	CO_2 to storage							
	lb/h	25,142						
	ton/day	301.70						
	tonne/day	273.78						
	MMscfd	5.2671						
	kWh/ton	146.56						
	kWh/tonne	161.50						

[1]cfm, cubic feet per minute; hp, horsepower; bhp, brake horsepower; LS, low stage; HS, high stage; gpm, gallons per minute; MMscfd, million standard cubic feet per day; TBD, to be determined.

[2]Includes dryer kilowatts converted to brake horsepower for power calculation purposes.

Table B9 Operating costs for a 300 ton/day (272 tonne/day) nonfood and nonbeverage grade facility[1,2]

Item	Amount	Units	Unit cost, $	Units	Cost/ton, $
Variable manufacturing cost items					
Power	146.56	kWh/ton	0.100	kWh	14.66
Desiccant (estimated 4-year life), including disposal	0.020	lb/ton	6.00	lb	0.12
Water makeup for evaporative condenser and cooler	52	gpm	0.10	Mgal	0.02
Water makeup for scrubber	0	gpm	0.10	Mgal	0.00
Sulfur removal (based on GTS), including disposal	0.00	lb/ton	0.72	lb	0.00
Carbon (estimated 2.5-year life), including disposal	0.000	lb/ton	6.50	lb	0.00
Water blowdown disposal, returned to host plant	26	gpm	0.05	Mgal	0.01
Variable manufacturing cost					14.81
Labor and overhead cost items					
Labor and benefits					
Plant manager	0.25		80,000	year	0.20
Shift foremen	0.5		65,000	year	0.33
Shift operator	1		50,000	year	0.51
Base labor cost			102,500	year	
Benefits factor of 40%			41,000	year	0.42
Total labor and benefits			143,500	year	1.46
Overhead					
Maintenance	2.00	% of capital	195,652	year	1.95
Taxes	1.00	% of capital	97,826	year	0.97
Insurance	1.50	% of capital	146,739	year	1.46
Labor and overhead costs					5.84
Total manufacturing cost					20.64
Principal and interest					
Principal and interest costs			875,235	year	8.89
Estimated total cost/ton					29.53

[1]Background specifications: Total installed cost: $9,782,575; interest rate/year: 6.50%; years: 20; type of plant: nonfood and nonbeverage grade; production capacity (ton/day): 300; annual operating days: 335; annual tons produced: 98,490; percent nameplate capacity: 98.00%; average ton/day: 294.

[2]gpm, gallons per minute; Mgal, thousand gallons; GTS, HydroCAT type GTS 2007.

Table B10 Consumables for a 300 ton/day (272 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Function	Component required, size or units	Quantity
CO-1	Oil removal	Coalescing elements manufactured by Balston, Parker, and Zander. Material of construction: borosilicate glass fiber with carbon steel retainers and fluorocarbon O-rings; size: 3.23 inches OD/1.72 inches ID × 25 inches long	~12
V-3	Oil removal	Typical coconut shell or coal-based activated carbon, e.g., carbon type 208 C or equivalent, lb	2,500
V-7A/B	Aldehyde and moisture removal	Activated alumina adsorbent, e.g., Selexsorb type CD or equivalent, lb	2,000 × 2 units + 1,200 × 2 units
V-9	Refrigeration	Ammonia, lb	4,800

[1]OD, outside diameter; ID, inside diameter.

Table B11 Vessel details for a 1,000 ton/day (907 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Description	Internal components	Diameter, inches	Seam/seam, inches	Material of construction	Design pressure, psig
VB-1	Blower inlet separator	Demister	120	180	304 SS	50
VB-2	Blower aftercooler separator	Demister	96	120	304 SS	50
V-1	Plant inlet separator	Demister	96	120	304 SS	50
V-2	Precooler separator	Demister	96	120	304 SS	50
CO-1A	Coalescer	Coalescing elements	20	48	CS	350
CO-1B	Coalescer	Coalescing elements	20	48	CS	350
V-3A	Carbon oil absorber	Johnson screens	36	96	CS	350
V-3B	Carbon oil absorber	Johnson screens	36	96	CS	350
V-4	Aftercooler separator	Demister	42	60	CS	350
V-7A/B	Dryer beds	Johnson screens	66	148	CS	350
V-8	Condenser separator	Vortex breaker	54	144	CS	350
V-9	Ammonia receiver	Dip tube	48	288	CS	250
VE-1	Ammonia separator for E-1, E-3	Demister	30	144	CS	250
VE-6	Ammonia separator for E-6, E-7, E-8	Demister	36	192	CS	250
T-1	Distillation column	Packing and supports, distributor	48	360	CS	350

[1]psig, pounds per square inch gauge; SS, stainless steel; CS, carbon steel.

Table B12 Heat exchanger details for a 1,000 ton/day (907 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Service	Exchanger details	Duty, Btu/h	Surface area, sq. ft	Material and design pressure–tube side, psi	Material and design pressure–shell side, psi
OC-1A	Oil cooler for C-1A	By compressor vendor	3,445,400	TBD	CS, 150	CS, 400
OC-1B	Oil cooler for C-1B	By compressor vendor	3,445,400	TBD	CS, 150	CS, 400
OC-2A	Oil cooler for C-2A	By compressor vendor	1,979,000	TBD	CS, 150	CS, 300
OC-2B	Oil cooler for C-2B	By compressor vendor	1,979,000	TBD	CS, 150	CS, 300
BE-1	Blower aftercooler	BEM 31-240	2,963,536	3,563	304 SS, 150	304 SS, 150
E-1	Precooler	BEM 27-240	2,259,507	2,656	304 SS, 150	CS, 250
E-2	CO_2 compressor aftercooler	BEM 27-240	2,194,723	2,656	304 SS, 350	CS,150
E-3	Refrigerant-cooled CO_2 aftercooler	BEM 21-216	1,301,073	1,436	304 SS, 350	CS, 250
E-4	CO_2 superheater	BEM 15-144	237,500	475	CS, 350	CS, 250
E-5A	Main reboiler	BKU 17/31-192	1,055,092	782	CS, 350	CS, 350
E-5B	Auxiliary reboiler	BEM15-168	1,240,461	556	CS, 250	CS, 350
E-6	Main condenser	BEM 39-288	11,275,550	6,260	CS, 350	CS, 250
E-7	Vent condenser	BEM 19-192	1,417,208	2,785	CS, 350	CS, 250
E-8	CO_2 subcooler	BEM 21-216	686,883	1,426	CS, 350	CS, 250
E-9	Vent gas refrigerant subcooler	BEM 12-48	30,713	96	CS, 350	CS, 250
E-10	Ammonia refrigerant condenser	BAC model no. PC2-948-1268-60	20,464,521			
E-11	Glycol–water plant cooling	BAC model no. FXV-0812C-16Q-O	13,043,523			

[1]TBD, to be determined; CS, carbon steel; SS, stainless steel.

Table B13 Utility requirements for a 1,000 ton/day (907 tonne/day) nonfood and nonbeverage grade facility[1,2]

Tag no.	Item, units	Full load capacity	Swept volume, cfm	Connected motor hp	Required bhp at design	Voltage	Motor efficiency, %	Power at design, kW
B-1	Centrifugal blower, HSI or equivalent, lb/h	97,804	TBD	1,250	1,190	4,160/3/60	97	920
C-1A	CO_2 compressor, Mycom model 3225 LLSC, lb/h	46,315	3,970 LS, 1,660 HS	2,000	1,833	4,160/3/60	97	1,417
OP-1A	Oil pump, gpm	365		30	19	480/3/60	95	15
C-1B	CO_2 compressor, Mycom model 3225 LLSC, lb/h	46,315	3,970 LS, 1,660 HS	2,000	1,833	4,160/3/60	97	1,417
OP-1B	Oil pump, gpm	365		30	19	480/3/60	95	15
C-2A	NH_3 compressor, Mycom model 3225 LSC, tons refrigeration	548	3,359 LS, 1,119 HS	1,250	1,084	4,160/3/60	97	838
OP-2A	Oil pump, gpm	215		15	9	480/3/60	94	7
C-2B	NH_3 compressor, Mycom model 3225 LSC, tons refrigeration	548	3,359 LS, 1,119 HS	1,250	1,084	4,160/3/60	97	838
OP-2B	Oil pump, gpm	215		15	9	480/3/60	94	7
P-1	Column pump, gpm	225		20	14	480/3/60	94	11
P-2	Scrubber water pump, gpm	90		30	26	480/3/60	94	21
	CO_2 product loading pumps, gpm	175		40	36	480/3/60	95	28
	Evaporative condenser fans			60	50	480/3/60	95	39
	Evaporative condenser pump			30	28	480/3/60	95	22
	Closed-circuit cooler fan			75	70	480/3/60	95	55
	Cooling water pump			15	12	480/3/60	95	9
	Coolant circulating pump			100	76	480/3/60	95	60
H-1	Dryer regenerator heater, kW	225		302	70	480/3/60		53
H-2	Carbon regenerator heater, kW	225		302	15	480/3/60		11
	Dryer regenerator flow, lb/h	6,323						
	Dryer regenerator flow, MMscfd	1.3246						
	Total connected hp			8,210				
	bhp[2]				7,464			
	Total kW at design							5,771
	CO_2 to storage							
	lb/h	89,232						
	ton/day	1,070.78						
	tonne/day	971.67						
	MMscfd	18.69						
	kWh/ton	129.35						
	kWh/tonne	142.54						

[1]cfm, cubic feet per minute; hp, horsepower; bhp, brake horsepower; LS, low stage; HS, high stage; gpm, gallons per minute; MMscfd, million standard cubic feet per day; TBD, to be determined.
[2]Includes dryer kilowatts converted to brake horsepower for power calculation purposes.

Table B14 Operating costs for a 1,000 ton/day (907 tonne/day) nonfood and nonbeverage grade facility[1]

Item	Amount	Units	Unit cost, $	Units	Cost/ton, $
Variable manufacturing cost items					
Power	129.35	kWh/ton	0.100	kWh	12.94
Desiccant (estimated 4-year life), including disposal	0.020	lb/ton	6.00	lb	0.12
Water makeup for evaporative condenser and cooler	172	gpm	0.10	Mgal	0.02
Water makeup for scrubber	0	gpm	0.10	Mgal	0.00
Sulfur removal (based on GTS), including disposal	0	lb/ton	1.66	lb	0.00
Carbon (estimated 2.5-year life), including disposal	0	lb/ton	4.50	lb	0.00
Water blowdown disposal, returned to host plant	86	gpm	0.05	Mgal	0.01
Variable manufacturing cost					13.09
Labor and overhead cost items					
Labor and benefits					
Plant manager	0.5		80,000	year	0.12
Shift foremen	1		65,000	year	0.20
Shift operator	1		50,000	year	0.15
Base labor cost			155,000	year	
Benefits factor of 40%			62,000	year	0.19
Total labor and benefits			217,000	year	0.66
Overhead					
Maintenance	2.00	% of capital	404,340	year	1.21
Taxes	1.00	% of capital	202,170	year	0.60
Insurance	1.50	% of capital	303,255	year	0.91
Labor and overhead costs					3.38
Total manufacturing cost					16.46
Principal and interest					
Principal and interest costs			1,808,791	year	5.51
Estimated total cost/ton					21.97

[1]Background specifications: Total installed cost: $20,217,010; interest rate/year: 6.50%; years: 20; type of plant: nonfood and nonbeverage grade; production capacity (ton/day): 1,000; annual operating days: 335; annual tons produced: 328,300; percent nameplate capacity: 98.00%; average ton/day: 980.
[2]gpm, gallons per minute; Mgal, thousand gallons; GTS, HydroCAT type GTS 2007.

Table B15 Consumables for a 1,000 ton/day (907 tonne/day) nonfood and nonbeverage grade facility[1]

Tag no.	Function	Component required	Quantity
CO-1	Oil removal	Coalescing elements manufactured by Balston, Parker, and Zander. Material of construction: borosilicate glass fiber with carbon steel retainers and fluorocarbon O-rings; size: 3.23 inches OD/1.72 inches ID × 25 inches long	~14 × 2 units
V-3	Oil removal	Typical coconut shell or coal-based activated carbon, e.g., carbon type 208 C or equivalent, lb	2,100 × 2 units
V-7A/B	Aldehyde and moisture removal	Activated alumina adsorbent, e.g., Selexsorb type CD or equivalent, lb	6,000 × 2 units + 3,000 × 2 units
V-9	Refrigeration	Ammonia, lb	7,700

[1]OD, outside diameter; ID, inside diameter.

Table B16 Typical inlet gas composition[1]

Component	Concentration range, ppmv
Moisture	Saturated
Acetaldehyde	3–75
Methanol	1–50
Ethanol	25–950
Acetone	0–2.5
Ethyl acetate	2–30
n-Propanol	0–1.0
i-Butanol	0–3
n-Butanol	0.5–1.0
Isoamyl acetate	0.6–3.0
Hydrogen sulfide	1–5
Dimethyl sulfide	0.5–1.5
Nitrogen	50–600
Oxygen	10–100
Methane	0–3
Carbon dioxide	Balance

[1]ppmv, parts per million by volume.