Task 2
Low-Quality Natural Gas Sulfur
Removal / Recovery

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SUMMARY

The accompanying Task 2 Report presents conceptual designs of the CFZ and CNG-Claus Processes for treatment of low quality natural gas. The conceptual designs are based on a low quality gas flow rate of 200 MMSCFD and two low quality gas compositions: 1) a composition representative of LaBarge reservoir with 66% carbon dioxide, 20% methane, 5% hydrogen sulfide, 7% nitrogen, and 1% other (helium, ethane, COS, water) and 2) a generic low quality gas composition with 20% carbon dioxide, 80% methane and 0.5% hydrogen sulfide. The low quality gas flowrate chosen for the conceptual design is consistent in magnitude with quantities of low quality gas now being produced from the LaBarge reservoir.

The integrated CFZ and CNG Claus process and the stand alone CNG Claus Process have a number of attractive features for treatment of low quality natural gas. Acid gases are separated from methane at pressure, as opposed to conventional chemical and physical solvent-based operations which dissipate the acid gas partial pressure during solvent regeneration. Maintenance of acid gas pressure is especially important if CO₂ is to be produced for EOR applications. The processes are well suited to handle a variety of trace contaminants normally associated with low quality natural gas processing and Claus sulfur recovery. The processes can accommodate large changes in feed composition over the life of a producing reservoir. The processes achieve virtually undetectable levels of sulfur compounds and significantly less than 2% CO₂ in the product methane. Sulfur recovery with the proposed high pressure sulfur recovery unit is nearly complete; waste carbon dioxide is the only intentional vent stream containing hydrogen sulfide (16 ppm). The processes can produce EOR or food grade carbon dioxide if markets exist. Finally, the processes have synergy with cryogenic nitrogen rejection and with liquefied natural gas production.

Process flow diagrams, material and energy balances, and equipment specifications are developed for three conceptual design cases: 1) a LaBarge gas processed with combined CFZ and CNG Claus technology; 2) a LaBarge gas processed with standalone CNG Claus technology independent of CFZ; and 3) a generic gas processed with combined CFZ and CNG technology. Process economics, evaluated for high and low product price scenarios, are presented on the basis of a breakeven allowance per MSCF of raw gas, i.e., the raw gas cost per MSCF at which plant net revenues become zero.

Breakeven allowances range from 20c to \$2.14/MSCF of raw gas for the low and high product price scenarios, assuming a simple five year capital payout. Capital and operating costs estimated for the conceptual process compare favorably with costs derived for the Selexol process treating low quality LaBarge gas [32]. Comparable breakeven allowances derived for the Selexol process are 8c and 73c.

The high pressure sulfur recovery process examined in this work is based on the Claus thermal reaction operating with excess hydrogen sulfide and deficient oxygen, with recycle of unconverted hydrogen sulfide. Western Research studied the thermodynamic and mechanical aspects of this technology and concluded a process scheme is feasible using conventional unit operations. The process achieves virtually 100% sulfur recovery at costs competitive with conventional Claus plus tail gas processes. Western Research has identified technical uncertainties and developed a research plan (submitted under separate cover) to permit further evaluation of the process.

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I. INTRODUCTION

The primary purpose of this Task 2 Report is to present conceptual designs developed to treat a large portion of proven domestic natural gas reserves which are low quality. The conceptual designs separate hydrogen sulfide and large amounts of carbon dioxide (>20%) from methane, convert hydrogen sulfide to elemental sulfur, produce a substantial portion of the carbon dioxide as EOR or food grade CO₂, and vent residual CO₂ virtually free of contaminating sulfur containing compounds. A secondary purpose of this Task 2 Report is to review existing gas treatment technology and identify existing commercial technologies currently used to treat large volumes of low quality natural gas with high acid gas content.

The subject technology of this Task 2 Report comprises four process technologies integrated to produce pipeline specification methane from low quality gas by efficient separation of relatively large amounts of hydrogen sulfide and carbon dioxide. These process technologies are:

- III Exxon's Controlled Freeze Zone (CFZ) Process
- III CNG's Liquid CO₂ Absorption of Sulfur Containing Contaminants
- III CNG's Triple-Point Crystallization (TPC) Process, and
- III CNG's High Pressure Sulfur Recovery Process (HPSRU).

Exxon developed the CFZ process to separate methane from acid gases in a single cryogenic distillation operation. CNG Research Company has developed the TPC process to purify carbon dioxide and concentrate hydrogen sulfide by alternately freezing and melting CO₂ at or near its triple point conditions (-69.9°F, 75.1 psia). The CNG HPSRU combines the high initial conversion obtained in a Claus thermal reactor with recycle of unconverted H₂S to the thermal reactor to effectively remove all sulfur components in the HPSRU feed gas.

The Task 2 objective is to prepare conceptual process designs to establish the technical feasibility and cost of upgrading large volumes of low quality natural gas to pipeline quality by integration of the three new gas processing technologies identified above. Two representative low quality gas compositions are examined at a scale of 200 MMSCFD: low quality gas from the LaBarge Anticline of southwest Wyoming with 66% carbon dioxide, 5% hydrogen sulfide, 20% methane and 9% other (predominately nitrogen), and a generic low quality gas with 20% carbon dioxide, 0.5% hydrogen sulfide and 80% methane. Exxon currently upgrades about 500 MMSCFD of low quality natural gas produced from the LaBarge formation.

The Task 2 Report is organized as follows. Section II defines low quality gas and speaks briefly to the motivation for seeking technology to develop low quality gas reserves. The target low quality gas to be treated with the proposed technology is identified, and barriers to the production of this gas are reviewed. Section III provides a description of the CFZ-CNG technologies, their features, and perceived advantages. The three conceptual process designs prepared under Task 2 are presented in Section IV along with the design basis and process economics. Process flow diagrams, material and energy balances, and equipment specifications are discussed (greater detail is included as Appendices B, C, D, and E). Section V presents an overview of existing gas treatment technologies, organized into acid gas removal technology, and sulfur recovery technology.

II. LOW QUALITY NATURAL GAS

A. Low Quality Natural Gas Definition

Natural gas is generally categorized as low quality if its acid gas content or inert content is above the minimum specifications of a natural gas transmission company. Pipeline specifications for acid gas content are usually a maximum of ¼ grain H₂S per 100 SCF (4 ppm H₂S) and less than 2% CO₂. Hugman's [1] definition of low quality gas appears as suitable as any other: " . . . any volume of gas containing levels of carbon dioxide above 2 percent or of nitrogen above 4 percent or gas with carbon dioxide plus nitrogen above 4 percent, or significant (more than trace) quantities of H₂S."

B. Motivation to Develop Low Quality Gas Reserves

Increased use of natural gas (methane) across the board in the domestic energy market will force the development of large domestic non-producing gas reserves now considered low quality.

New Federal regulations embodied in the Clean Air Act of 1990, Title III and Title V, are creating strong new market opportunities for natural gas. These include natural gas co-firing and reburn in coal-fired power plants to help reduce SO₂ and NO_x emissions; natural gas fired turbines for peak-shaving and on-site generation of electrical power and steam; and natural gas as an alternative clean transportation fuel, either compressed, liquefied, or converted to methanol. These new markets alone will increase substantially the demand for natural gas. The American Gas Association forecasts these new markets will increase the annual demand for natural gas by 2 TCF, more than 13% of current production. To meet these anticipated gas supply demands, domestic production of natural gas must begin to exploit the large reserves of low quality gas available but not now produced for technical, economic, or environmental reasons.

C. Target Low Quality Gas for Proposed Process Technology

Vast reserves of low quality natural gas contaminated with hydrogen sulfide, carbon dioxide, and nitrogen are not suitable for conventional gas treating. Domestic reserves of low quality natural gas are estimated to be about 200 trillion cubic feet (TCF) [1,2]. Upgrading this vast reserve of low quality gas to pipeline specifications with current technology, primarily amine absorption of acid gases and sulfur recovery by conventional Claus technology, is generally neither economic nor environmentally acceptable.

The major proven non-producing reserve of low quality gas is located in the LaBarge reservoir of southwestern Wyoming. The amounts are large, and dwarf all other proven reserves combined. Estimates place the LaBarge reservoir recoverable gas at about 167 TCF, of which at least 33 TCF is methane. The LaBarge reservoir, prone to all three contaminants H₂S, CO₂ and N₂, is one of the largest geologic formations in North America.

The CFZ/CNG-Claus process is capable of treating and has targeted the relatively large quantity of low quality natural gas containing >10% CO₂ which may also contain either or both of H₂S and N₂ and which cannot be economically treated on a large scale by other known technologies.

D. Low Quality Natural Gas Production Barriers

There are many technical barriers which now prevent the economic up-grade of low quality natural gas to pipeline standards [3] These include but are not limited to:

- ▮ removal of H₂S to very low levels -- pipeline specs are 4 ppm H₂S;
- ▮ removal of N₂ to increase heating value, decrease transportation cost;
- ▮ regeneration of fluid or solid separating agents [Rouseau, 4]
- ▮ degradation of polymeric materials used in membrane-based separations;
- ▮ poor selectivity for CH₄ in presence of acid gases, or for H₂S in presence of CO₂;
- ▮ separation of isolated acid gases into pure CO₂ and concentrated H₂S;
- ▮ recovery of sulfur from separated H₂S;
- ▮ dissipation of the substantial CO₂ partial pressure available in high-CO₂ LNG;
- ▮ recompression of CO₂ for commodity use or sales.

The above list mentions explicitly only the three major contaminants (aside from water) of LNG: nitrogen, hydrogen sulfide, and carbon dioxide. The presence of additional contaminants, especially sulfur species such as carbonyl sulfide (COS) and mercaptans (RSH), increases the difficulty of treating LNG.

III. CFZ-CNG LOW QUALITY GAS PROCESS

A. Description of Component Process Technologies

A.1. Controlled Freeze Zone (CFZ) Process

The Controlled Freeze Zone (CFZ) Process is a cryogenic technology for the separation of carbon dioxide from natural gas by distillation. CFZ is a proprietary process being developed by Exxon Production Research Company [5].

Natural gas processing involves the separation and recovery of valuable hydrocarbon components, and the removal of undesirable components such as H_2S , CO_2 and water. Low temperature physical separations, based on turbo-expander and Joule-Thomson (J-T) operations, and direct fractionation are the preferred methods for processing natural gas whenever possible. However, cryogenic fractionation of a gas containing more than about 5% CO_2 can lead to the solidification of CO_2 . This freezing phenomena would occur at an intermediate point in a cryogenic demethanizer rendering such tower inoperative. Thus alternative technologies, using solvents or freeze-prevention additives, have been utilized in the past. The CFZ process, in contrast, achieves a direct cryogenic separation of methane and CO_2 . In an otherwise conventional distillation tower, solid CO_2 is confined to a special section of the tower, *the CFZ section*, specifically designed to control the formation and melting of solid CO_2 .

Figure III-1 is a simplified illustration of a CFZ tower. Viewed from outside, its operating conditions are essentially those of a conventional demethanizer. Typical operating pressures are in the 500 to 650 psia range. Overheads, rich in methane, leave the column at temperatures in the -120 to -150°F range. Corresponding temperatures for the bottoms stream, rich in CO_2 and in heavier hydrocarbons (and containing any sulfur compounds) are in the 30 to 60°F range. The CFZ tower has three sections. In the lower section, the bulk separation of methane and CO_2 is accomplished and methane is stripped off the bottoms product. The upper section performs the final removal of sour gas components and of heavier hydrocarbons down to sales or additional processing specifications. The solidification and melting of CO_2 occurs in the middle, or CFZ, section.

The upper and lower sections are of standard design and incorporate conventional internals in the form of trays or packed sections. The CFZ section does not have conventional internals. Instead it consists of spray banks at its upper end and a specially designed melt tray at its lower end. Rising vapors enter this section still containing about 15 to 20 mol% CO_2 (approaching solidification conditions and thus the limit for conventional cryogenic distillation). They are contacted with a cold CO_2 lean liquid sprayed from the upper section. Solid CO_2 forms and drops to the melting tray; In doing so it allows a CO_2 lean gas to progress upwards to the upper section of the tower. The CO_2 crystals fall onto a warm liquid layer held atop the melting tray. The temperature of this liquid is maintained above the freezing temperature of CO_2 . Thus no solid CO_2 leaves the CFZ section. The formation of solid CO_2 is effectively *controlled* in this section. The CO_2 lean vapors enter the upper rectification section containing about 3 to 5% CO_2 and are further purified via conventional distillation to the desired purity. The CFZ concept has

been successfully demonstrated in a 600 MSCFD pilot plant at Clear Lake, near Houston, Texas [6,7].

A.1.1. Thermodynamic Description of CH₄-CO₂ System.

A pressure-temperature diagram for the methane-CO₂ system [8] illustrates the previous limitations of direct fractionation of methane and CO₂. The diagram in Figure III-2 shows the pure methane and the pure CO₂ saturation lines, as well as the locus of critical points corresponding to the various mixtures of methane and CO₂. It also shows the CO₂ solidification dome defining conditions at which CO₂ will solidify. Operation at pressures above the solidification dome is limited by the approach to the critical locus, where vapor and liquid become indistinguishable from each other and prevent any further separation. This limitation is encountered much before commercial methane purities of 2 to 4% CO₂ can be achieved. Operation at lower pressures is limited by approach to the solidification dome where the methane product still contains 15 to 20% CO₂. The CFZ process enables operation at these lower pressures where CO₂ freezes by carefully controlling the formation and subsequent remelting of solid CO₂.

The operation of the CFZ tower can be related to a temperature-composition diagram for the methane-CO₂ system. Figure III-3 shows the regions where vapor, liquid, and solid phases exist. In a normal system only vapor, liquid, and vapor-liquid regions would be present, and a classical distillation operation would be confined to the vapor-liquid region. The CFZ tower, however, involves operations in a vapor-liquid-solid region in addition to the classical vapor-liquid region. The feed to the CFZ tower is typically introduced at the bottom section. The rising vapors in this section become colder and leaner in CO₂ (Point A_v). Eventually these vapors reach the region where, upon further cooling, a solid CO₂ phase forms in equilibrium with a methane-CO₂ vapor (Point B_v, already inside the CFZ section). Further cooling of these vapors depletes them of CO₂ as pure CO₂ crystals form and drop away. Vapors very lean in CO₂ enter the vapor-liquid-only region (Point C_v) where conventional distillation can be resumed in the upper section of the tower.

A very cold, CO₂-lean liquid is withdrawn from the upper distillation section (Point D_L) and sprayed into the CFZ section to provide the cooling required for the formation of solid CO₂. At the bottom of the CFZ section, the melt tray holds a layer of liquid sufficiently warm to melt the falling CO₂ crystals. The surface of this liquid layer is at the conditions represented by point B_L. The liquid in the melting tray is kept warm by the warm vapors rising from the bottom section of the tower. Excess liquid from this melt tray is directed to the lower section of the tower where conventional distillation is resumed.

A.1.2. Potential Applications and Advantages of the CFZ Process

Technically the CFZ Process can remove CO₂ from natural gas streams containing concentrations of CO₂ above about 5%. Economically, however, the CFZ process becomes competitive for low quality natural gas streams with 10% or more CO₂. Commercially available large-scale CO₂ removal processes use chemical and physical solvents (amines, Selexol, Rectisol, and many others more obscure), freeze preventive additives (Ryan Holmes type processes), or membranes. Chemical solvents absorb CO₂ and subsequently

Figure III-1 Simplified CFZ Column

Figure III-2 CFZ Thermodynamic Basis, P vs T

Figure III-3 CFZ Thermodynamic Basis, T vs x,y

release it when heated in the regeneration step; the cost of gas treatment increases in direct proportion to its CO₂ content. Thus chemical solvents are preferred when low concentrations of CO₂ are present. Physical and hybrid solvents extend the range; but they frequently absorb hydrocarbons as well, and since they are still based on absorbing the impurities, their costs go up with CO₂ content. Cryogenic distillation, while far less sensitive to the CO₂ content, involves expensive refrigeration systems. Thus its economic competitiveness begins at concentrations of 10 to 15% CO₂. Relative to distillations based on freeze-preventive additives, the CFZ process has the advantage of accomplishing the separation in a single step, saving costs associated with the recovery and regeneration of the freeze-prevention additive. Membranes are good CO₂ removal candidates but high recoveries and purities involve several stages of processing and expensive recompressions between them. Economies of scale are not realized for treating large quantities of gas.

There are some specific instances, within the above generalized guidelines, where the CFZ process appears to be well suited for certain applications, e.g., EOR applications using CO₂ floods. The associated gas would contain high and varying amounts of CO₂ after breakthrough. The CFZ unit could easily produce a methane gas stream and a liquid CO₂/NGL miscible reinjection stream. Since the reinjection stream is liquid, it can easily be pumped to any required reinjection pressure. Another example involves applications that require further cryogenic processing of the separated streams, as in LNG production or when nitrogen rejection units (NRU) are needed. A cryogenic integration of the entire facility can provide significant overall savings.

A.2. Triple Point Crystallization (TPC) Process

Acid gas removal from gases with a high CO₂ to H₂S ratio requires the separation of CO₂ and H₂S to produce a CO₂ byproduct or vent stream free of sulfur compounds and a concentrated H₂S Claus feed. Distillation of CO₂ and H₂S to produce a pure CO₂ product is not practical due to the low relative volatility of CO₂ to H₂S and high CO₂ purity requirements at the pure CO₂ end (top) of the distillation column.

The continuous triple point crystallizer separates pure carbon dioxide from a variety of contaminants such as H₂S, COS, CH₃SH and hydrocarbons. The process has been developed and patented by Consolidated Natural Gas Company [9,10,11,12,13,14]. The crystallizer operates at or near the triple point of CO₂. Solid CO₂ crystals are formed by adiabatic flashing at pressures slightly *below* the CO₂ triple point, and melted by adiabatic contact with CO₂ vapor at pressures slightly *above* the CO₂ triple point. No solid CO₂ is formed on heat exchange surfaces by indirect heat exchange; all solid CO₂ is formed and melted by direct contact heat exchange.

Figure III-4 shows a single stage of the triple point crystallizer. The liquid feed enters the flash zone where the pressure is lowered sufficiently to form vapor and solid. The ratio of the heat of vaporization to the heat of fusion for triple point CO₂ is about 1.75, thus 1.75 lb of solid is formed for every pound of vapor flashed. Solid crystals approximately 1 mm in diameter fall through the liquid (solid density 1.5 g/cm³, liquid density 1.2 g/cm³) and form a packed solid bed. The crystal bed is washed by a small amount of upward flowing liquid from the melt zone at the bottom of the packed solid bed. The wash liquid, about 10 to 15% of the downward solid flow, washes H₂S from the crystal

Figure III-4**Single Stage Carbon Dioxide
Triple-Point Crystallizer**

surfaces and prevents diffusion of H_2S into the melt zone. In the melt zone at the bottom of the packed solid bed, solid is contacted with clean CO_2 vapor which condenses in the melter providing the heat to melt the solid. Part of the clean liquid formed in the melter is returned upwardly through the packed solid bed as wash liquid, another portion is the pure CO_2 product, and a third portion is sent to a heat exchanger where it is vaporized to produce melt gas. Vapor from the flash zone is compressed and condensed on the other side of this heat exchanger. The condensed liquid is then returned to the flash zone. The concentrated H_2S -rich product is withdrawn as liquid from the flash zone.

An alternate arrangement of the triple point crystallizer separates the flash zone and the washer-melter zone into two vessels. Solid CO_2 is pumped as a slurry from the flasher to the washer-melter. This arrangement allows the difference in flashing and melting pressure to be controlled by the slurry pump rather than the height of the liquid leg in the crystallizer. This crystallizer configuration is convenient when large contaminant concentrations in the mother liquor lower the flasher temperature and pressure, causing large pressure differences (several 10's of psi as opposed to 10 or less) to exist between the flash zone and the melt zone. (The melt zone, where pure CO_2 is produced, always operates at conditions near the triple point of CO_2 .)

Experimental operation of a laboratory crystallizer has demonstrated that a very high degree of separation can be achieved in a single stage of crystallization. Experimentally measured separation factors, the ratios of contaminant concentrations in the flash zone (solid forming) to the melt zone, are 1000 to 1500 for H_2S and over 3000 for COS [14]. Pure CO_2 containing less than 1 ppm by volume H_2S has been produced by triple point crystallization from contaminated CO_2 mixtures. In this particular low quality natural gas processing application, two stages of triple point crystallization produce pure CO_2 and a concentrated H_2S product.

A.3. CNG High Pressure Sulfur Recovery (HPSRU) Process

A.3.1. Introduction

A new sulfur recovery process [15] based on the Claus thermal reaction, with no catalytic stages or conventional tailgas cleanup, is made possible by the inherent capability of triple-point crystallization technology to separate hydrogen sulfide and other contaminants from carbon dioxide. A schematic diagram of the CNG-Claus high pressure sulfur recovery process is shown in Figure IV-4, Section IV; there are four steps: 1) a Claus thermal reactor to convert hydrogen sulfide to sulfur, 2) waste heat exchange and sulfur condensation, 3) a hydrogenator to convert all sulfur species in the Claus reactor effluent to hydrogen sulfide, and 4) a quench and dehydration. All hydrogen sulfide is recycled to the TPC section which concentrates acid gases by rejecting carbon dioxide and other inert species such as nitrogen

The conventional Claus plant, in order to recover 99+% sulfur, normally comprises a thermal reactor, several catalytic reactors in series, and a tail gas cleanup unit. As noted by Hyne [16], "more than 50% of the conversion of hydrogen sulfide to elemental sulphur takes place in the front end reaction furnace; (while the) downstream components do no more than convert that part of the sour gas feed stream that is not converted to product sulphur in the front end reaction furnace." The high conversion of hydrogen sulfide to

sulfur achieved in the front end reaction furnace is achieved at relatively modest cost -- about 20% of the total Claus plant capital cost. The down-stream components, which accomplish roughly 40% of hydrogen sulfide conversion to sulfur, account for about 80% of the capital cost (Figure III-5).

The new sulfur recovery process retains the high recovery/low cost of the Claus thermal reactor, but eliminates the low recovery/high cost catalytic stages and tailgas cleanup unit. Unreacted hydrogen sulfide is recycled to the thermal reactor inlet via the TPC section; this tail gas recycle approach enables essentially 100% sulfur recovery, free of thermodynamic and kinetic limitations imposed by the Claus reaction.

A.3.2. Process Description

The flow diagram of the HPSRU is shown Figure IV-4, Section IV. The process steps encompass the Claus thermal reaction, waste heat exchange and sulfur condensation, hydrogenation of sulfur species to hydrogen sulfide, quench, and dehydration. Each of these unit operations is discussed briefly below.

A.3.2.1. Claus thermal reaction furnace

Three design parameters of CNG's proposed Claus thermal reaction furnace are fundamentally different than those of a conventional Claus thermal reaction furnace:

- 1) excess hydrogen sulfide, oxygen deficient operation,
- 2) high oxygen concentration (90% oxygen), and
- 3) high operating pressure (typically 100 psia).

Acid gas from the TPC section enters the Claus thermal reaction furnace where it reacts with oxygen to ultimately produce (primarily) sulfur and water according to overall reaction (R-I). This overall reaction results from an exothermic and irreversible combustion reaction (R-II) and the endothermic, reversible Claus reaction (R-III).

Hydrogen sulfide is an excess reactant and comprises 70% of the acid gas feed. The excess hydrogen sulfide serves as a thermal moderator to maintain an acceptable reaction temperature (about 2,000°F), and to reduce the presence of unreacted SO₂ in the reactor effluent to less than 2 mol%. The conversion of hydrogen sulfide to sulfur is about 60%. Since sulfur is produced only in the thermal reaction furnace, the hydrogen sulfide converted to sulfur coincides with the hydrogen sulfide contained in the raw low quality natural gas feed. Thus, for 60% conversion, the hydrogen sulfide flow to the thermal reaction furnace is about 1.67 times the hydrogen sulfide flow in the raw low quality natural gas feed.

Figure III-5

The Claus thermal reaction furnace operates in an oxygen deficient mode compared to the conventional Claus thermal reaction furnace which operates close to the stoichiometric ratio of $2 \text{ H}_2\text{S}:\text{O}_2$. Enriched air containing 90% O_2 is used as the oxygen source. This concentration was selected after examination of oxygen cost versus purity, and the burden which inerts (nitrogen and argon) place on downstream and TPC section process costs.

The HPSRU process train operates at about 100 psia to maintain operating pressure close to that of TPC, and avoid large gas compression requirements to recycle acid gas to the TPC section. The modest pressure level should not complicate the mechanical design or materials selection for the reaction furnace.

A.3.2.2. waste heat exchanger/sulfur condenser

These two units serve to recover thermal energy as high pressure steam, and to cool the reaction furnace effluent to permit condensation of product sulfur. The 100 psia operating pressure of the exchanger/condenser should not influence its design or operation significantly. The elevated pressure provides a slight advantage for sulfur condensation by lowering the gas phase sulfur partial pressure for any given condensation temperature, and permits thermal energy recovery at higher temperature levels than in conventional atmospheric Claus plants. The sulfur condenser effluent is cooled to about 300°F.

A.3.2.3. reducing gas generator

The reducing gas generator performs two functions: 1) heating the condenser effluent gas to the desired hydrogenation reactor temperature, and 2) producing reductant gases (H_2 and CO) for the hydrogenation reactions. These functions are achieved in parallel by burning a hydrocarbon fuel (methane) with deficient oxygen in a conventional burner. The hot combustion products are mixed with the HPSRU tail gas before the hydrogenation reactor. About 0.3% (LaBarge case) and 0.02% (generic case) of the raw low quality natural gas methane, obtained from the product gas, supplies the necessary hydrocarbon feed to the reducing gas generator.

A.3.2.4. hydrogenation reactor

The hydrogenation reactor selectively converts SO_2 , COS , CS_2 and residual elemental sulfur vapor to H_2S . The conversion is performed to eliminate SO_2 and the possibility of the Claus reaction (R-III) occurring and forming elemental sulfur either in the recycle stream to the TPC section, or the TPC section itself. The reduction reaction of SO_2 to H_2S is the major consumer of reducing gases, and thus the motivation to operate the thermal reactor at conditions which minimize SO_2 in the thermal reactor effluent is evident.

Sulfur and SO_2 are hydrogenated to H_2S ; COS and CS_2 are hydrolyzed to H_2S . The reactor operates in the temperature range 450 to 750°F and employs a cobalt-molybdenum catalyst. The hydrogenation reactor design is conventional, and conventional Claus plant experience indicates that the cobalt-moly hydrogenation catalyst outlasts Claus reaction catalysts by factors of two to three.

A.3.2.5. waste heat exchange/quench/dehydration

The function of these units is to cool and dry the HPSRU tail gas leaving the hydrogenation reactor. The gas is cooled to about 275°F, well above the water dew point, and then enters a single stage water wash designed to further cool the gas to about 70°F and condense water. Gas leaving the quench contains less than 1 mol% water. The gas is then further dehydrated to a water dew point of about -80°F in a mol sieve desiccant system, and compressed to about 125 psia for return to the TPC section.

B. Integrated Technologies for Low Quality Gas Treatment

B.1. The CFZ - CNG Claus Process

The integrated process is shown in Figure III-6. After dehydration and cooling the feed gas is sent to the CFZ tower which separates the CH_4 and other light components such as He and N_2 from CO_2 , H_2S and other trace heavy components such as COS and C_2H_6 . If the CH_4 overhead product contains significant amounts of N_2 it is sent to a cryogenic nitrogen rejection unit (NRU). The CO_2 containing bottoms product is sent to the TPC section of the process.

The TPC section contains an absorber-stripper which strips out the small amount of CH_4 (approximately 1%) and C_2H_6 carried over in the CFZ bottoms. The H_2S in this vent stripping gas is re-absorbed with clean liquid CO_2 from the crystallizer. This vent gas also contains the inert components N_2 , argon, and hydrogen brought to the TPC section with the HPSRU recycle stream. The stripped CO_2 is sent to the crystallizer which produces

Figure III-6
Conceptual CFZ/CNG Claus
Low Quality Natural Gas Treatment Process

pure CO₂ and a CO₂ stream concentrated in H₂S. This pure CO₂ product can be used for EOR or commodity CO₂ applications. The H₂S rich TPC product is further enriched in H₂S by stripping out CO₂ and is sent to the Claus plant. Tail gas from the HPSRU is returned to the TPC section for reconcentration of the unreacted H₂S.

B.2. CNG AGR Technology for Low Quality Gas Treatment

The process is shown in Figure III-7; it does not employ CFZ process technology to separate methane from acid gases. Instead, the gas is dehydrated and cooled to condense a portion of the CO₂ in the raw gas stream. It enters the sulfur compound absorber-stripper column. This column has three sections. The top portion absorbs H₂S and other sulfur compounds from the gas; the center portion strips methane and other light components from the liquid; the bottom section concentrates H₂S by stripping out CO₂. In the top section of the column, vapor from the feed and lower stripping section is contacted with pure liquid CO₂ from the crystallizer and reflux from the condenser. The vapor leaves the column free of sulfur compounds and with a reduced CO₂ content. In the center section the liquid is stripped of methane and light ends by vapor from the bottom section. All of the methane free liquid is withdrawn from the column and fed to the crystallizer. The H₂S enriched product from the crystallizer is returned to the bottom section of the column where it is stripped of CO₂ by vapor from the reboiler. Concentrated H₂S Claus feed is withdrawn from the bottom of the column.

The product gas, free of H₂S and other sulfur containing components, is contacted with cold methanol to absorb residual CO₂. The gas is sent to a cryogenic nitrogen rejection unit (NRU) for separation into methane, crude helium and N₂. Methanol is regenerated by stripping near atmospheric pressure with nitrogen from the NRU.

Tail gas from the HPSRU is cooled and partially condensed. The liquid, rich in H₂S, is recycled to the HPSRU. The vapor is contacted with liquid CO₂ absorbent from the crystallizer to absorb H₂S. The H₂S free vapor is vented; the absorbent liquid is returned to the crystallizer.

C. CURRENT PROCESS UNCERTAINTIES

C.1. Heavy (C₂+) Hydrocarbons

The feed gases specified for this study contained no significant amounts of heavier hydrocarbons. The process designs of this study would concentrate ethane (if it were present in the raw gas) in the tail gas vent and would direct propane and heavier hydrocarbons into the Claus feed for combustion along with hydrogen sulfide. The design of the absorber-stripper can be modified to accommodate significant amounts of ethane in the feed gas by adding a condenser to the top of the column to produce a fuel gas product containing methane, ethane and carbon dioxide instead of the tail gas vent. Heavier hydrocarbons can be separated from the acid gases in the Claus feed by absorption or distillation before they enter the Claus plant. All light hydrocarbons which might enter the crystallizer with the acid gas feed remain with the non-crystallizable material constituting the mother liquor and exit the crystallizer in the concentrated hydrogen sulfide stream. High separation factors for ethane and ethylene have been measured experimentally in a laboratory crystallizer [14]. Butane has been added during flash rate (crystal formation rate)

Figure III-7
Conceptual CNG-Claus
Low Quality Natural Gas Treatment Process

tests [17] without detrimental effect on flasher performance. Since light hydrocarbons anticipated in raw natural gas do not form solid solutions with carbon dioxide [18], the crystallizer is expected to effectively separate carbon dioxide from feeds containing these C₂+ hydrocarbons.

C.2. Triple Point Crystallizer

The triple point crystallizer has been tested extensively in the laboratory at bench-scale (up to 6-inch vessel diameters) [14]. With the laboratory bench-scale equipment continuous runs of up to 72 hours duration were achieved and terminated routinely. A large scale flash vessel (18-inch diameter) built to test rates of solid carbon dioxide production and slurry pumping characteristics has been operated in continuous runs of up to 40 hours duration at production rates of 25 tons of solid CO₂ per day. No unusual wear or abrasion on the slurry pump was observed after many tests totaling hundreds of hours. Slurries of 25 wt% solid CO₂ were routinely pumped.

C.3. HPSRU

The high pressure Claus design is based on an equilibrium model of the thermal reactor. The model has been shown valid for many systems with a similar array of reacting crystallizer is expected to effectively separate carbon dioxide from feeds containing these C₂+ hydrocarbons.components. Reaction kinetics may have an effect on conversion efficiency, reaction temperature and the extent of side reactions such as hydrogen or carbon monoxide formation. These effects will be quantified by operation of a small scale reactor. Other parts of the high pressure sulfur recovery unit are much more conventional. Concern for corrosion should be limited to the quench tower where liquid water and hydrogen sulfide are present. However, quench towers performing comparable service are presently used in the SCOT and Beavon tail gas cleanup processes. Also, sulfur dioxide, which is much more acidic in aqueous solution than hydrogen sulfide, is not present in the quench tower because of upstream hydrogenation.

D. Summary of Process features/Advantages

D.1. Controlled Freeze Zone

- Bulk acid gas removal,
- Single distillation step,
- No solvents or additives necessary,
- Contaminant insensitivity,
- High pressure acid gas,
- Synergy with cryogenic NRU,
- Synergy with LNG product,
- Non-corrosive process streams.

D.2. Liquid Carbon Dioxide Absorption of Contaminants

- Attractive physical properties (density/viscosity/mol weight),
- Favorable contaminant equilibrium,
- High stage efficiency,
- No solvents or additives necessary, available from raw gas,
- Not combustible

D.3. Triple-Point Crystallization

- Concentrated hydrogen sulfide,
- Pure liquid carbon dioxide for EOR or absorption of H₂S
- Sharp separation of trace contaminants
- No solvents or additives necessary,
- Direct contact heat exchange,
- Small pressure driving forces cause phase changes.

D.4. CNG High Pressure Claus Sulfur Recovery

- Claus thermal stage only
- Eliminates catalytic stages, tail gas unit, and incinerator,
- Elevated pressure,
- Oxygen or enriched air,
- No solvents or additives necessary,
- High sulfur recovery,
- Low sulfur emissions.

IV. CONCEPTUAL LOW QUALITY GAS PROCESS DESIGN

A. Design Basis

Two feed gas mixtures, each 200 MMSCF/day, were studied having CO₂ contents covering a wide range of process applicability. One case is a high CO₂ gas now processed at the Exxon's Shute Creek facility near LaBarge, Wyoming. The feed gas for the second case is a lower CO₂ content generic subquality gas. These crude gas streams are more fully described below.

A.1. LaBarge Low Quality Case

This case examines treatment of subquality gas produced from the LaBarge field in southwestern Wyoming. The formation is estimated to contain 167 TCF of low Btu raw gas [1]. This crude gas is characterized below in Table IV-1.

The product slate includes methane (pipeline gas), elemental sulfur, helium, and EOR grade carbon dioxide. The CFZ methane product has 50 ppm CO₂ and less than 4 ppm H₂S (¼ grain/SCF). The low CO₂ content of the methane prevents formation of solid CO₂ in the cryogenic NRU. The final methane product specifications after nitrogen and helium rejection and recompression are 97% CH₄ at 1000 psia. The EOR-grade liquid CO₂ product has less than 16 ppm H₂S and is supplied at 2000 psia.

A.2. Generic Low Quality Case

For the purpose of this study, a generic subquality gas composition shown in Table IV-1. This composition is typical of gas produced from the San Juan Basin area of northwest New Mexico. Product specifications for the generic case include a methane product with 4 ppm H₂S (¼ grain/SCF) and less than 2% CO₂ (0.25% CO₂ is achieved as dictated by the controlling spec on H₂S). In the design of this case, no market was assumed to exist for CO₂. Pipeline methane and a small stream of elemental sulfur are the only saleable products produced from this generic subquality gas.

A.3. Common Design Parameters and Assumptions

The CFZ and TPC sections of the process were simulated with the HYSIM (HYPROTECH Ltd, 14th Street N.W., Calgary, Alberta T2N 4Z6) simulator. In the TPC section simulation, the interaction constants for the CO₂-H₂S and CO₂-COS binaries in the Peng-Robinson Stryjek-Vera equation of state were modified to fit CNG proprietary vapor liquid equilibrium data. HYSIM resident data cannot be used to accurately simulate the equilibrium behavior of these two binary systems.

The equipment sparing philosophy used for both LaBarge and generic cases was that pumps would be spared and compressors would not be spared. In determining total capital costs, a major equipment installation factor of 3.5 was used for the generic case and 4.52 was used for the LaBarge case. The higher installation factor for the LaBarge case reflects costs associated with the remote Wyoming location. The generic factor represents a typical, domestic U.S. installation. A simple five year payback on capital was used in the analysis of annual costs.

Table IV-1
Gas Compositions and Process Conditions

	Case 1: LaBarge		Case 2: Generic	
Specie:	mol%	lb mol/hr	mol%	lb mol/hr
CH ₄	20.50	4494.	80.50	17679.
CO ₂	66.50	14578.	19.00	4172.
H ₂ S	5.00	1096.	0.50	110.
N ₂	7.35	1644.		
COS	0.05	10.		
C ₂ H ₆	0.01	2.		
He	0.60	132.		
TOTAL	100.00	21956.	100.00	21961.
Pressure	1,060 psia		715 psia	
Temperature	60°F		90°F	
Gas Flow	200 MMSCF/day		200 MMSCF/day	
Water	10 lb/MMSCF		75 lb/MMSCF	
Claus Plant	High Pressure Oxygen		High Pressure Oxygen	
Oxygen	90% O ₂ 210 Ton/day		90% O ₂ 21 Ton/day	
EOR CO ₂	16 ppm H ₂ S 2,000 psia		None	
Methane	50 ppm CO ₂ undetectable H ₂ S 40 MMSCFD		0.25% CO ₂ 4 ppm H ₂ S 160 MMSCFD	
Sulfur	375 LT/day 99.9% recovery		38 LT/day 99.9% recovery	
CO ₂ Vent	16 ppm H ₂ S 13 Ton H ₂ S/yr		16 ppm H ₂ S 8.5 Ton H ₂ S/yr	

B. Process Design

B.1. LaBarge Composition Case- CFZ-CNG Claus Process

B.1.1. CFZ Section

The process flow diagram for the CFZ section is shown in Figure IV-1. The feed gas, entering the plant at 60°F and 1060 psia, is dehydrated to a -80°F water dew point in a DRIZO unit. The dry gas is then cooled to -50°F using three levels of propane refrigeration and an intermediate level of ethylene refrigeration. Additional cooling is also provided by recovering refrigeration from the pure CO₂ product stream discharged from the TPC unit. A portion of the feed is liquefied during this cooling. The vapors are sent to an expander, and the liquids to a Joule-Thomson (J-T) valve to reduce their pressure to the CFZ tower operating pressure of 600 psia.

The overhead stream from the CFZ tower contains less than 50 ppm CO₂ and no detectable amounts of H₂S. The overhead is condensed at -150°F with an ethylene refrigerant. A 5°F temperature approach is used in the plate fin condenser. The low condensing temperature is due to the presence of nitrogen and helium. Nitrogen makes up 7.5% of the feed, and helium 0.6%. Being the lightest components in the feed, they end up in the CFZ overheads in concentrations of about 27% for nitrogen and 2% for helium. The CFZ overheads are sent to a nitrogen rejection unit (NRU) and a helium plant for removal of nitrogen from the methane sales stream, and for recovery of helium as a saleable product. The NRU and helium plants are based on standard, proven technologies and beyond the scope of this study.

Exxon's Shute Creek (LaBarge) sub-quality gas processing plant in Wyoming includes a NRU and helium plant. CFZ overheads would be sent to the existing NRU and helium plant after recovering available refrigeration potential. CFZ overheads are used to subcool and condense ethylene in the ethylene refrigeration loop; in doing so it is allowed to warm up from -150°F to -10°F. After going through a J-T valve, the CFZ overhead stream would have conditions similar to the current NRU feed at the Shute Creek plant.

The CFZ bottoms exit the tower at +41°F and 600 psia, containing at most 1% methane, and are sent to the TPC section. In a stand-alone application of CFZ, the bottoms could be expanded and their refrigeration potential used to cool the CFZ feed, thus reducing the external refrigeration requirements.

The refrigeration system incorporates a two-level ethylene loop, and a three-level propane loop. Propane is provided at +40°F, -10°F, and -40°F. High pressure propane is air cooled to 100°F and condensed in the CFZ reboiler. Propane refrigerant cools CFZ feed and condenses ethylene refrigerant. Ethylene, in turn, is provided at -60°F and -155°F. Ethylene is condensed at -30°F using propane and CFZ overheads.

B.1.2. TPC Section

The bottoms liquid from the CFZ tower are sent to the TPC section. The process flow diagram for the TPC section is shown in Figure IV-2. The liquid is cooled by heat exchange with the product streams and expanded to the H₂S absorber-light ends stripper pressure of 100 psia. The bottom portion of this tower strips N₂ and CH₄ from the liquid and the top portion of the column absorbs the H₂S from the stripping gas. Heat is provided

Figure IV-1
CFZ SECTION
Process Flow Diagram
LaBarge Composition Case

STREAM	RAW FEED 1	COOLED FEED 2	CFZ FEED 3	CFZ FEED 4	CFZ OVERHEAD 5	TO NRU 6	CFZ BOTTOMS 7
P (psia)	1060	1000	600	600	600	570	600
T (°F)	60	-50	-75	-58	-150	-10	41
FLOW MSCFD	200						
lbmoles/hr	21961	21956	2423	19533	6111	6111	15845
COMPOSITION							
Helium	0.0060	0.0060	0.0424	0.0015	0.0215	0.021	--
Nitrogen	0.0749	0.0749	0.2771	0.0498	0.2689	0.268	0.0001
C ₁	0.2046	0.2047	0.4659	0.1723	0.7095	0.709	0.0100
C ₂	0.0001	0.0001	0.0001	0.0001	--	--	0.0001
CO ₂	0.6638	0.6640	0.2019	0.7213	50 ppm	50ppm	0.9200
H ₂ S	0.0499	0.0499	0.0126	0.0546	--	--	0.0692
COS	0.0005	0.0005	0.0000	0.0005	--	--	0.0006
H ₂ O	0.0002	--	--	--	--	--	--

at the bottom of the tower by cooling and condensing the Claus offgas stream. The bottom liquid containing only ppm levels of CH₄ and N₂ is sent to the triple point crystallizer.

Two stages of crystallization produce pure CO₂ liquid and vapor with only 3 to 4 ppm of H₂S. Part of this pure CO₂ liquid is taken as EOR grade CO₂ product. It is pumped to 2000 psia and warmed against the CFZ bottoms liquid and the feed gas to the CFZ section. The remaining pure CO₂ liquid is pumped to the top of the H₂S absorber-light ends stripper tower to be used as absorbent. Some of the pure CO₂ produced by the crystallizer is vaporized to provide cooling within the crystallizer.

A concentrated contaminant liquid, containing 20% H₂S is the other product of the crystallizer. This liquid is pumped to the distillation column and warmed by the CFZ bottoms. This column operates at 100 psia and produces a 10% H₂S overhead product and a 70% H₂S bottoms product. The vapor at the top of the column is compressed and condensed with reboiled vapor at the bottom of the column. Part of this condensed liquid is used as reflux to the column and the remainder is returned to the crystallizer. All uncondensed vapor is sent to the H₂S absorber-light ends stripper. The 70% H₂S bottoms product is vaporized by heat exchange with the CFZ bottoms and sent to the high pressure Claus plant. The offgas from the Claus plant is cooled and partially condensed by heat exchange with the stripper bottoms. The liquid is sent to the distillation column for reconcentration of the H₂S and the vapor containing the N₂ and H₂ from the enriched oxygen and hydrogenator is sent to the H₂S absorber-light ends stripper. Clean vapor from the top of the absorber and clean vapor from the crystallizer are combined, warmed by heat exchange with the CFZ bottoms, expanded to near atmospheric pressure and warmed again by heat exchange with the CFZ bottoms and vented.

The two stage triple point crystallizer is shown in Figure IV-3. The liquid feed enters the top stage flasher. This flasher operates at 51 psia and -81°F. Approximately 410 tons/hr of solid CO₂ are produced in the flasher by adiabatic flash vaporization. The vapor from the flasher is compressed to 100 psia and condensed and cooled to -58°F by heat exchange with bottom stage liquid CO₂ product, generating melt vapor in the process. The liquid is returned to the flasher. The solid CO₂ is pumped as a slurry to the top stage washer melter vessel. The liquid from the slurry is returned to the flasher. A portion of this liquid is taken as the concentrated H₂S product (20% H₂S). The solid is melted with compressed vapor from the bottom stage flasher. The melt zone operates at 78 psia and -70°F. The melted liquid is sent to the bottom stage flasher which operates at 73 psia, -70°F and 0.4% H₂S (liquid). The vapor from this flasher returns to the top stage melter and the solid is pumped as slurry to the bottom stage washer melter. The liquid from the slurry is returned to the flasher and the solid is melted with vapor from the heat exchanger. The liquid from the melter contains 4 ppm H₂S and a few ppb of COS. Part of the liquid is taken as pure CO₂ product and the remainder is vaporized while condensing the top flasher stage compressed vapors. A portion of this vapor is used in the melter and the rest is vented.

B.1.3. HPSRU Section

The HPSRU process flow diagram is shown in Figure IV-4. The high pressure Claus feed is mixed with 90% oxygen from a PSA (pressure swing adsorption) plant and

Figure IV-3
Triple Point Crystallizer

Figure IV-4

HP CLAUS SECTION Process Flow Diagram
CFZ/CNG LaBarge Composition Case

Stream Name	1	2	3	4	5	6	7	8
Vapour_Frac	1.00	1.00	1.00	0.00	1.00	1.00	0.00	1.00
Temperature	50.00	50.00	2006.00	275.00	50.00	775.00	100.00	100.00
Pressure	95.00	95.00	94.00	93.00	52.00	91.00	89.00	110.00
Flow	2633.30	594.30	3300.60	1104.00	26.40	2756.10	1085.80	1670.30
Hydrogen	0.0000	0.0000	0.0266	0.0000	0.0000	0.0196	0.0000	0.0000
Oxygen	0.0000	0.9011	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Nitrogen	0.0000	0.0989	0.0178	0.0000	0.0530	0.0218	0.0000	0.0000
Methane	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0350	0.0000	0.0000	0.0014	0.0000	0.0000
CO2	0.0000	0.0000	0.2076	0.0000	0.0000	0.2949	0.0000	0.0000
H2S	0.6940	0.0000	0.2050	0.0000	0.0000	0.2677	0.0000	0.4117
COS	0.0060	0.0000	0.0015	0.0000	0.0000	0.0007	0.0000	0.0011
S02	0.0000	0.0000	0.0171	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0000	0.0000	0.3221	0.0000	0.0000	0.3940	1.0000	0.0000
S_Vapour	0.0000	0.0000	0.1674	0.0000	0.0000	0.0000	0.0000	0.0000
S_Liq	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000	0.0000

combusted in the Claus thermal reaction furnace. The oxygen flow is approximately 60% of the stoichiometric requirement. The reactor operates at 2000°F and 95 psia. The reactor effluent is cooled in the waste heat exchanger by making 400 psia steam, and in the sulfur condenser with boiler feed water, to 300°F. The liquid sulfur is drawn off and the vapor is heated to 430°F in a heat exchanger and sent to the reducing gas generator. Methane and a sub-stoichiometric amount of O₂ are combusted in the reducing gas generator to heat the gas to 500°F and provide additional reducing gas for hydrogenation. All of the sulfur vapor and SO₂ are converted to H₂S over the Co-Mo catalyst in the hydrogenator. The only sulfur containing compounds in the hydrogenator effluent are COS and H₂S. The hydrogenator effluent gas is cooled in another waste heat exchanger by making low pressure 55 psia steam. The gas passes through the quench tower where the water vapor in the gas is condensed by contacting it with cold water. Most of the quench water is cooled in an air cooler and pumped to near the top of the tower. A portion of the cooled quench water is cooled to 65°F by heat exchange with the EOR liquid CO₂ product and sent to the top of the tower. The rest of the quench water is stripped of H₂S at the bottom of the tower, cooled and released. The gas leaves the quench tower at 70°F with 0.4% water and is sent to the mol sieve dehydration beds. After final water removal in the dehydration bed the gas is compressed to 115 psia and returned to the TPC section. About 15% of the dried compressed gas is used as regeneration gas. It is heated with high pressure steam, passed through the bed to be regenerated and returned to the quench tower. The sulfur production rate is 375 metric tons/day.

B.2. LaBarge Composition - CNG-Claus Process

B.2.1. Bulk CO₂ and H₂S Removal Section

The process flow diagram of the bulk CO₂ and H₂S removal section is shown in Figure IV-5. The H₂S feed gas is dehydrated to a -80°F water dew point in a DRIZO unit. The dry gas is cooled to -30°F and partially condensed with two levels of propane refrigeration (0°F, -40°F). Additional cooling is provided by recovering refrigeration from the product gas, Claus feed, fuel gas, and vent gas from the methanol regenerator. Approximately 85% is condensed.

The feed enters an absorption and stripping tower which operates at 700 psia. The overhead product stream contains 2 ppm of H₂S, an undetectable amount of COS, 18% CO₂, 21% N₂, 2% He, and 59% methane. The overhead product is sent on to the final CO₂ removal section. A portion of the vapor from the tower is condensed at -67°F with an ethane refrigerant. This reflux is combined with pure liquid CO₂ from the TPC to provide the absorbent for the H₂S in the tower.

B.2.2. TPC Section

A liquid side draw below the feed tray provides the feed to the TPC. All the liquid is withdrawn from the tower at this point. The section of the tower between the feed and the liquid side draw reduces the methane content of the liquid to 0.02% methane.

The H₂S enriched product liquid from the TPC is returned to the stage below the liquid side draw. The bottom section of the tower concentrates the H₂S. The bottoms product exits the tower at 94°F and 600 psi and contains 70% H₂S.

Figure IV-5

CNG Claus Process Flow Diagram
H₂S and Bulk CO₂ Removal Section / LaBarge Composition Case

Stream Name		FEED	TAIL GAS	S-FREE GAS	EORCO2	VENT GAS	CLAUSTRIZER FEED	TPC FEED	CO2 ARS1	CO2 ARS2
Vapour Frac		1.00	1.00	1.00	0.00	1.00	1.00	0.00	0.00	0.00
Temperature	F	60.00	100.00	-67.00	72.61	47.63	24.99	-38.00	-64.53	-64.53
Pressure	psia	1060.00	160.55	700.00	2000.00	15.00	95.00	695.00	700.00	700.00
Flow	lbmole/hr	21960.00	1670.91	7615.01	6058.64	7337.65	2623.06	33270.84	2410.00	1407.00
Helium	Mole frac	0.005999	0.000000	0.017301	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000
Methane	Mole frac	0.204979	0.000000	0.590186	0.000000	0.000961	0.000000	0.000219	0.000000	0.000000
Nitrogen	Mole frac	0.073493	0.018691	0.211939	0.000000	0.008763	0.000124	0.000001	0.000000	0.000000
CO2	Mole frac	0.664934	0.406400	0.180494	0.999996	0.982691	0.295636	0.882097	0.991196	0.999996
H2S	Mole frac	0.049995	0.441501	0.000002	0.000004	0.000010	0.699428	0.116983	0.000004	0.000004
COS	Mole frac	0.000500	0.001107	0.000000	0.000000	0.000000	0.004891	0.000596	0.000000	0.000000
Ethane	Mole frac	0.000100	0.000000	0.000000	0.000000	0.000216	0.000006	0.000104	0.000000	0.000000
Hydrogen	Mole frac	0.000000	0.032300	0.000000	0.000000	0.007351	0.000015	0.000000	0.000000	0.000000

The feed to the TPC containing 12% H₂S is cooled from the tower temperature of 58°F to -38°F by heat exchange with the TPC H₂S enriched product, the EOR CO₂ product and the vent gas product. Two stages of crystallization produce pure CO₂ liquid and vapor with only 3 to 4 ppm of H₂S and an enriched product containing 23% H₂S. Part of the pure liquid CO₂ is taken as EOR grade CO₂ product. It is warmed by heat exchange with the TPC feed and pumped to 2000 psia. The remaining liquid is used as absorbent. Some is pumped up to 700 psi and returned to the top of the absorption and stripping column and another portion is used as absorbent in the tail gas absorption column. Some of the pure CO₂ produced by the crystallizer is vaporized to provide cooling within the crystallizer. The H₂S enriched product is pumped to 700 psi, warmed by heat exchange with the TPC feed and returned to the bottom section of the absorber-stripper.

The two stage TPC is essentially the same as that used in the CFZ-CNG Claus design. Noncondensables in the flash gas from the higher concentration H₂S stage of the TPC are sent to the tail gas absorber. The solid production rate for this case is 400 tons/hr.

The 70% H₂S bottoms product from the bottom of the absorption stripping tower is reduced in pressure, mixed with tail gas condensate and vaporized by heat exchange with the tail gas and the feed and sent to the HPSRU. The tail gas from the HPSRU is cooled and partially condensed by heat exchange with the HPSRU feed. The tail gas condensate is returned to the HPSRU and the vapor is sent to the tail gas absorption tower where it is contacted with pure liquid CO₂ from the TPC. The liquid from the tail gas absorber is returned to the TPC. The vapor from the tail gas absorber is mixed with CO₂ vapor from the TPC, warmed by heat exchange with the TPC feed, expanded to near atmospheric pressure, warmed again with the TPC feed and vented. The vented gas contains only ppm levels of H₂S.

B.2.3. HPSRU Section

The HPSRU is the same as that used in the CFZ-CNG Claus design.

B.2.4. Final CO₂ Removal

The H₂S free gas is sent to the final CO₂ removal section. A process flow diagram is shown in Figure IV-6. The gas is contacted with cold methanol to remove CO₂, and leaves the tower at -79°F and 680 psi containing 50 ppm of CO₂. The gas is warmed to -10°F by heat exchange with the feed and by sub cooling the ethane and propane refrigerants and is sent to the NRU. The NRU is the same as that used in the CFZ-CNG Claus design.

The absorbent is flashed to recover methane and is sent to the regeneration column where it is contacted with N₂ from the NRU at 18 to 20 psia. The regenerated methanol is pumped back to the top of the absorption tower. The vent gas and fuel gas are warmed by heat exchange with the feed.

B.3. Generic Composition Case - CFZ-CNG Claus Process

The Process Flow Diagram for the CFZ section is shown in Figure IV-7. As in the LaBarge case, the feed gas, entering the plant at 90°F and 715 psia, is first dehydrated to a -80°F dew point in a DRIZO unit. The dry gas is then cooled to about -50°F using propane refrigeration and the CFZ overhead stream. The availability of sufficiently cold CFZ overheads eliminates the need for the intermediate ethylene level of refrigeration used in

Figure IV-6

CNG Claus Process Flow Diagram Final CO2 Removal Section / LaBarge Composition Case

Stream Name		S_Free_Gas	NRU_N2	PROD	UENT	FUEL_GAS	MeOH
Vapour_Frac	F	1.00	1.00	1.00	1.00	1.00	0.00
Temperature		-66.10	-20.00	-10.00	45.00	43.00	-80.01
Pressure	psia	700.00	20.00	665.00	15.00	215.00	20.00
Flow	lbmole/hr	7615.01	1350.00	6032.01	2753.35	179.16	6149.69
Methane	Mole frac.	0.590233	0.199998	0.712283	0.125079	0.687084	0.000213
Nitrogen	Mole frac.	0.211956	0.799992	0.265883	0.392376	0.053934	0.000065
Methanol	Mole frac.	0.000000	0.000000	0.000011	0.000105	0.000093	0.999702
CO2	Mole frac.	0.180508	0.000010	0.000032	0.482438	0.257264	0.000020
Helium	Mole frac.	0.017302	0.000000	0.021791	0.000002	0.001624	0.000000

the LaBarge case. The cold vapors are then simply passed through a J-T valve prior to being introduced to the CFZ tower which operates at 600 psia.

To meet the 4 ppm H₂S (¼ grain per 100 SCF) specification, the CO₂ content of the overhead stream from the CFZ tower is reduced to less than 0.25% CO₂. CFZ overheads are condensed at -123°F with an ethylene refrigerant. A 7 F° temperature approach is allowed in the plate fin condenser. The condensing temperature is much warmer than in the LaBarge composition case due to the absence of any components lighter than methane. The CFZ overheads are used to subcool the ethylene refrigerant and to cool the plant feed prior to being sent to the sales pipeline.

The bottoms of the CFZ tower exit at 42°F and 600 psia, containing at most 1% methane, and are sent to the TPC unit. The refrigeration system incorporates a single level ethylene loop, and a three-level propane loop. Propane is provided at 40°F, -10°F, and -40°F. High pressure propane is air cooled to 100°F and condensed in the CFZ reboiler. Propane is used to cool the CFZ feed and to condense ethylene refrigerant. Ethylene, in turn, is provided at -130°F. Ethylene is condensed at -30°F using propane and a side stream of a reboiler in the TPC section.

The bottoms liquid from the CFZ tower are sent to the TPC section. The process flow diagram for the TPC section is shown in Figure IV-8. The liquid is cooled by heat exchange with the product streams and expanded to the H₂S absorber-stripper pressure of 100 psia. The bottom portion of this tower strips CO₂ from H₂S, the center portion of the column strips N₂ and CH₄ from the liquid to be fed to the crystallizer and the top portion of the column absorbs the H₂S from the stripping gas. Heat is provided at the bottom of the tower by cooling and condensing ethylene from the CFZ refrigeration system. All of the liquid from the bottom of the upper section of the tower, which contains only ppm levels of CH₄ and N₂, is sent to the triple point crystallizer.

Two stages of crystallization produce pure CO₂ liquid and vapor with only 3 to 4 ppm of H₂S. This pure CO₂ liquid is pumped to the top of the H₂S absorber-stripper tower to be used as absorbent. Some of the pure CO₂ produced by the crystallizer is vaporized to provide cooling within the crystallizer. A concentrated contaminant liquid, containing 20% H₂S is the other product of the crystallizer. This liquid is pumped back to the top of the lower section of the absorber-stripper tower and warmed by the Claus offgas to near its bubble point.

The 70% H₂S bottoms product from the tower is vaporized by heat exchange with the CFZ bottoms and sent to the high pressure Claus plant. The offgas from the Claus plant is cooled and partially condensed by heat exchange with the H₂S rich condensate and the H₂S product from the crystallizer. The liquid is revaporized by heat exchange with the Claus offgas and is mixed with the tower bottoms and sent to the Claus plant. The vapor containing the N₂ and H₂ from the enriched oxygen and hydrogenator is sent to the upper part of the H₂S absorber-stripper. Clean vapor from the top of the absorber and clean vapor from the crystallizer are combined, warmed by heat exchange with the CFZ bottoms, expanded to near atmospheric pressure and warmed again by heat exchange with the CFZ bottoms and vented.

Figure IV-8
TPC SECTION Process Flow Diagram
Generic Case

Stream Name		1	2	3	4	5	6	7	8	9
Vapour frac		0.11	0.00	1.00	0.00	0.00	1.00	1.00	0.25	1.00
Temperature	F	59.74	-56.11	-68.75	-68.54	-63.21	-60.00	74.60	-60.00	11.97
Pressure	psia	100.00	100.00	75.00	100.00	100.00	105.00	95.00	105.00	10.00
Flow	lbmole/hr	4291.90	7300.90	730.67	3399.33	3255.27	41.11	261.57	162.70	4190.59
Methane	Mole frac	0.010000	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	0.010242
CO2	Mole frac	0.964392	0.954741	0.999998	0.999998	0.897436	0.558313	0.299340	0.482999	0.987000
H2S	Mole frac	0.025600	0.045257	0.000002	0.000002	0.102561	0.164774	0.700022	0.415309	0.000015
Nitrogen	Mole frac	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	0.000000	0.030099	0.001479
CO2	Mole frac	0.000000	0.000001	0.000000	0.000000	0.000002	0.000144	0.000638	0.001024	0.000000
Hydrogen	Mole frac	0.000000	0.000000	0.000000	0.000000	0.000000	0.128661	0.000000	0.032569	0.001264

The two stage triple point crystallizer is a smaller scale version of the LaBarge case crystallizer described previously. The crystallizer operates at essentially the same composition, temperature and pressure as the LaBarge Case. The solid CO₂ production rate for this case is 116 tons/hr.

The high pressure Claus plant is also a smaller scale version of the LaBarge case. The operating temperatures, pressures and compositions are the same as the LaBarge case. The sulfur production rate for this case is 38 metric tons/day.

C. Equipment Sizing

Sizing of equipment is an intermediate step in establishing the cost of such equipment. For costing purposes, the sizing units are those for which cost correlations have been established in the past. Typically, they are horsepower (hp) for pumps and compressors, heat transfer area or duty for heat exchangers, and weight for towers and vessels. The process simulator HYSIM provided most of this information with the notable exception of weights. With the exception of the TPC flasher and washer melter, Acrion sized the towers and vessels of the CNG sections of the process with HYSIM. HYSIM uses liquid residence times and flooding vapor velocities to size towers and vessels. The CFZ section equipment was scaled up from a prior 20 MMSCFD design by Exxon Production Research Company [45]. The results of equipment sizing are presented in detail in Appendix D. Some specifics are discussed below.

In the triple point crystallizer a separation factor, ratio of contaminant concentration in the mother liquor to melted solid, of 1000 was used for H₂S and COS. This value is about 30% below the separation factor of 1500 for H₂S routinely observed in TPC operation. An experimentally measured solid production rate of 3000 lb/hr/ft² (10% less than measured experimentally) was used to size the vessels [17]. A 15% backwash in the washer-melter was used, although backwash rates as low as 10% have been verified experimentally.

A stage efficiency of 80% was used to size the H₂S absorber. This efficiency is based on the scale-up of experimental measurements of the absorption of H₂S from gas mixtures with liquid CO₂ [14]. Experimental gas mixtures used in contaminant absorption experiments included N₂ and CO₂, CH₄ and CO₂, and CH₄-CO-H₂ and CO₂.

Residence times for the Claus reactor and hydrogenator, as well as heat transfer coefficients for the sulfur condenser and waste heat boiler, were provided by Western Research. A residence time of 2 seconds was used in the thermal reactor. Typical residence times of Claus thermal reactors are from 0.6 to 0.8 sec [48] so the 2 second value is conservative. The waste heat exchanger and sulfur condenser were designed using a conventional sulfur plant design heat transfer coefficient of 12 Btu/hr/ft²/F. The higher pressure and pressure drop allowed in the HPSRU should provide improved heat transfer coefficients. The hydrogenation reactor is designed using the conventional cobalt molybdenum catalyst that is used commercially in other tail gas treatment processes.

All vessels operating below -50°F were fabricated of stainless steel. This assumption provides conservative cost estimates as process equipment operating at or near the CO₂ triple point (-70°F) could be fabricated from less expensive nickel alloy steels. Carbon steel was used for temperatures above -50°F.

D. Conceptual Process Economic Estimates

Capital costs, operating costs, and revenue generation potential have been estimated to evaluate the economic viability of the combined technologies.

D.1. Capital Costs

Overall capital investment costs were estimated by conventional factoring of major equipment costs. The equipment costs associated with each of the three sections, CFZ, TPC, Claus, were determined separately. The costs for the CFZ unit were estimated from vendor quotes and contractor estimates for a 20 MMSCFD CFZ demonstration facility that Exxon planned to install in Wyoming. The TPC equipment costs were estimated by Acrlon using cost estimating charts [46] and, for selected items, vendor quotes.

In factoring equipment costs to an overall capital investment figure, a typical gas industry factor of 3.5 was used for the generic composition case. A higher installation factor of 4.52 was used in the LaBarge composition case. The higher factor reflected the additional costs associated with the remote Wyoming location, and the need to winterize the processing facilities. An overall project contingency of 7.7% was applied to the generic case and is included in the overall factor of 3.5. Due to its remote location, an overall project contingency of 20% was applied to the LaBarge cases and is included in the overall factor of 4.52.

Tables IV-2 through IV-4 provide the capital investment figures for each case. The overall installed costs are \$148 million for the LaBarge composition CFZ-CNG Claus case, \$120 million for the LaBarge composition CNG Claus case, and \$75 million for the generic case. The higher cost for the LaBarge cases reflects the higher installation factor as well as additional refrigeration requirements. In the generic case, CO₂ is vented as a gas slightly above atmospheric pressure after recovery of its refrigeration potential. In the LaBarge cases, however, CO₂ is a high pressure liquid product, and thus a substantial amount of refrigeration potential remains with it.

The CFZ unit costs (including refrigeration systems) ranged from 57 to 75% of the overall costs and the TPC unit represented 16 to 34% of the overall cost for the two cases with a CFZ unit. The Claus unit represented 2 to 7%, and the feed dehydration 5 to 7% of the total cost for all the cases.

The major equipment cost for the dehydration unit was the same in all cases. The higher installed cost in the LaBarge case reflects the higher LaBarge installation factor.

The cost of major equipment for the CFZ unit is 16% higher for the CFZ LaBarge case than the Generic case; mainly reflecting additional refrigeration requirements. The installed cost of the CFZ unit, however, is 50% higher in the CFZ LaBarge case than in the Generic. Such is the impact of the higher installation factor for a remote location.

The TPC unit equipment cost for the CFZ LaBarge case is about 4 times that of the Generic case. This substantial difference reflects the larger and slightly more complex TPC section required for the LaBarge composition case. The raw feed gas in the LaBarge case contains 70% sour gas components, whereas the Generic case has only about 20%. As these components become the main constituents of the TPC section feed, it is easy to see that for the same volumetric raw gas feed, the LaBarge case requires a larger TPC section.

Table IV-2
Capital Costs
CFZ/CNG Claus Process
LaBarge Composition Case, 200 Million SCFD Unit

	<u>Feed Dehydration</u>	<u>CFZ</u>	<u>TPC</u>	<u>Claus</u>	<u>Overall Process</u>
DIRECT COSTS					
<i>Major Equipment</i>	\$1,476	\$18,559	\$11,092	\$1,648	\$32,775
<i>Other Direct Costs</i>	\$2,195	\$27,590	\$16,490	\$2,450	\$48,724
<i>Total Direct Costs</i>	\$3,671	\$46,149	\$27,582	\$4,097	\$81,499
INDIRECT COSTS	\$1,886	\$23,706	\$14,168	\$2,105	\$41,865
TOTAL BEFORE CONT.	\$5,557	\$69,855	\$41,750	\$6,202	\$123,364
CONTINGENCY (20 %)	\$1,111	\$13,971	\$8,350	\$1,240	\$24,673
TOTAL INSTALLED COST	\$6,668	\$83,826	\$50,100	\$7,443	\$148,037

Table IV-3
Capital Costs
CFZ/CNG Claus Process
Generic Composition Case, 200 Million SCFD Unit

	<u>Feed Dehydration</u>	<u>CFZ</u>	<u>TPC</u>	<u>Claus</u>	<u>Overall Process</u>
DIRECT COSTS					
<i>Major Equipment</i>	\$1,476	\$15,983	\$3,460	\$402	\$21,322
<i>Other Direct Costs</i>	\$1,849	\$20,018	\$4,333	\$504	\$26,703
<i>Total Direct Costs</i>	\$3,325	\$36,001	\$7,793	\$906	\$48,025
INDIRECT COSTS	\$1,478	\$15,998	\$3,463	\$403	\$21,341
TOTAL BEFORE CONT.	\$4,803	\$51,999	\$11,256	\$1,309	\$69,366
CONTINGENCY	\$369	\$3,996	\$865	\$101	\$5,330
TOTAL INSTALLED COST	\$5,172	\$55,995	\$12,120	\$1,410	\$74,697

Table IV-4
Capital Costs
CNG Claus Process
LaBarge Composition Case, 200 Million SCFD Unit

	<u>Feed Dehydration</u>	<u>CFZ</u>	<u>Claus</u>	<u>Overall Process</u>
DIRECT COSTS				
<i>Major Equipment</i>	\$1,476	\$23,174	\$1,927	\$26,578
<i>Other Direct Costs</i>	\$2,195	\$34,451	\$2,865	\$39,511
<i>Total Direct Costs</i>	\$3,671	\$57,625	\$4,793	\$66,089
INDIRECT COSTS	\$1,886	\$29,611	\$2,463	\$33,960
TOTAL BEFORE CONT.	\$5,557	\$87,237	\$7,255	\$100,049
CONTINGENCY (20 %)	\$1,111	\$17,447	\$1,451	\$20,010
TOTAL INSTALLED COST	\$6,669	\$104,684	\$8,707	\$120,059

Likewise, the equipment cost for the sulfur recovery unit (HPSRU) is five to six times as high for the LaBarge cases as it is for the Generic case. This difference is again due to the larger feed stream to the HPSRU.

Returning to overall plant costs, it is worth observing that processing a gas with a LaBarge type composition but in a less remote location would have resulted in a total installed cost of \$115 and \$93 million for the CFZ and CNG cases respectively (using an installation factor of 3.5). This is about 25% lower than the estimated \$148 and \$120 million for the CFZ and CNG cases at remote location. Therefore the high LaBarge case costs should be kept in the proper perspective - *they apply to a specific remote location application.*

The CNG process has a lower capital cost than the CFZ-CNG process for the same LaBarge gas feed. The higher CFZ costs are due to the lower refrigeration temperature and high reflux ratio in the CFZ tower.

D.2. Operating Costs

The operating costs, shown in Tables IV-5 through IV-7, ranged from 6¢/MSCF of plant feed for the generic case to 24¢ to 26¢ for the LaBarge cases. As in the capital costs, the higher operating costs reflect the increased refrigeration requirements, as well as the higher operating and maintenance costs of a more expensive and remote facility.

Utilities and chemicals costs ranged from less than 1¢ to 12¢/MSCF of feed gas. Electric motors were used throughout for the LaBarge case, since relatively low cost hydroelectric power, 3¢/kWh, is available at the site. Electricity represented the bulk of this cost category, from 9¢ to 10¢/MSCF. Enriched-oxygen air represented over 95% of the chemical portion of this cost category for the LaBarge cases. Fuel gas was used to power the compressors in the Generic case; as would be the case in a more typical scenario. As the fuel gas is taken from the processed gas stream itself, no explicit operating costs are recorded; instead there is a corresponding reduction in the amount of product gas available for sale. About 2¢/MSCF utilities cost is shown for the Generic case, primarily associated with enriched-oxygen air.

Labor and overhead costs are 1.5¢/MSCF for all cases. This cost includes two 24 hour a day, seven day a week operators, a 5 day a week full time supervisor and overhead for both at 175% of labor.

Other operating costs included maintenance materials, local taxes, insurance, and in the LaBarge cases the costs associated with the nitrogen rejection unit (NRU) and the helium processing plant. As indicated in previous technical sections and in the capital cost section, the light gases present in the raw feed gas, nitrogen and helium, exit with the overhead stream. Before this stream becomes pipeline gas, nitrogen is removed, and helium, being very valuable, is also recovered as a separate product. Cryogenic NRU's and helium plants are considered established technologies and they were not part of the scope of this evaluation. To provide an overall economic picture, however, the costs of further processing the overhead stream by an NRU, a helium plant and a methane gas recompression facility were estimated. Such estimates ranged from 5 to 10¢/MSCFD of raw gas feed. A mid-point value of 7.5¢ was used in this study.

Table IV-5

**Annual Operating Costs
CFZ/CNG Claus Process
LaBarge Composition Case, 200 Million SCFD Unit**

	<u>Annual Cost (\$M)</u>	<u>Cost per MSCF of Plant Feed</u>
Utilities	7182	\$0.103
Chemicals	1070	\$0.015
Total Utilities & Chemicals	8252	\$0.118
Direct Labor	370	\$0.005
Overhead	648	\$0.009
Total Labor & Overhead	1018	\$0.015
Maintenance Materials	1480	\$0.021
Local Taxes & Insurance	2287	\$0.033
NRU/He Processing	5250	\$0.075
Total Other Costs	9018	\$0.129
ANNUAL OPERATING COSTS	18,288	\$0.261

* Capital Investment (\$M)	\$148,037
* On stream days per year	350
	95.9%
* Total MSCF gas process	70,000,000

Table IV-6
Annual Operating Costs
CFZ/CNG Claus Process
Generic Composition Case, 200 Million SCFD Unit

	Annual Cost (\$M)	Cost per MSCF of Plant Feed
Utilities	440	\$0.006
Chemicals	255	\$0.004
<i>Total Utilities & Chemicals</i>	694	\$0.010
Direct Labor	370	\$0.005
Overhead	648	\$0.009
<i>Total Labor & Overhead</i>	1018	\$0.015
Maintenance Materials	747	\$0.011
Local Taxes & Insurance	2018	\$0.029
<i>Total Other Costs</i>	2765	\$0.039
ANNUAL OPERATING COSTS	4,477	\$0.064

* Capital Investment (\$M)	\$74,697
* On stream days per year :	350
	95.9%
* Total MSCF gas processed	70,000,000

Table IV-7
Annual Operating Costs
CNG Claus Process
LaBarge Composition Case, 200 Million SCFD Unit

	<u>Annual Cost (\$M)</u>	<u>Cost per MSCF of Plant Feed</u>
Utilities	5967	\$0.085
Chemicals	1085	\$0.016
Total Utilities & Chemicals	7052	\$0.101
Direct Labor	370	\$0.005
Overhead	648	\$0.009
Total Labor & Overhead	1018	\$0.015
Maintenance Materials	1201	\$0.017
Local Taxes & Insurance	1944	\$0.028
NRU/He Processing	5250	\$0.075
Total Other Costs	8394	\$0.120
ANNUAL OPERATING COSTS	16,464	\$0.235

* Capital Investment (\$M)	\$120,059
* On stream days per year :	350
	95.9%
* Total MSCF gas processed	70,000,000

Maintenance materials, taxes, and insurance were estimated as percentages of capital investment. A certain portion of taxes was based on revenues.

The CNG process has a lower operating cost than the CFZ-CNG process for the same LaBarge gas feed. The higher CFZ costs are due to greater electric power requirements for refrigeration.

Appendix E provides additional details on the operating requirements and operating costs.

D.3. Revenues

Natural gas prices have been depressed for some time now, but have shown recent strength with a forecasted average 1993 price of \$2/MCF, and supplies finally beginning to tighten [47]. For process revenue estimates, we have simply bracketed the anticipated gas and by-product prices with reasonable high and low values. Table VI-8 shows the prices used in this evaluation. The sulfur produced in the LaBarge cases must be transported to a sulfur market and as such its net revenue is lower than that used for the Generic case.

The anticipated yearly revenues for each case are shown in Tables VI-9 through VI-11. Gross revenues range from \$0.81 (CNG-LaBarge) to \$1.21 /MSCF (Generic) of feed gas in a low product price scenario. The range is \$1.32 (CNG LaBarge) to \$2.42 (Generic) in a high product price scenario. The CFZ-CNG process recovers a greater amount of carbon dioxide for EOR use than the CNG process and therefore has a greater gross revenue than the CNG process.

It is interesting to note the relative contributions to the annual revenue. In the LaBarge cases, methane is present in the feed at 20.5% and contributes 31 to 41% of the revenues. Carbon dioxide makes up 66.5% of the feed and contributes 21 to 30 % of the revenue. Sulfur, as H₂S, makes up 5% of the feed but contributes 13 to 16% of the revenue. Helium, by far the most valuable component, makes up only 0.6% of the feed but contributes 19 to 27% of the annual revenue. In the Generic case, methane makes up 80.5% of the feed and contributes about 99% of the revenue, the small remainder being sulfur revenue. The H₂S content of the Generic case feed gas is 0.5%. CO₂ present in the feed at 19% is vented.

On an equivalent per MSCF of raw feed gas basis, the total annual costs range from 28¢ to 68¢. These annual costs include operating costs and a capital charge based on a simple five year payout period, but do not include the cost of the raw feed gas. The net revenue (exclusive of raw gas costs) ranges from 20¢ to \$2.14/MSCF of feed gas. The BTU content of the feed gas varies substantially, from 240 BTU/SCF for the LaBarge cases to 822 BTU/SCF in the Generic case. On this basis, the breakeven allowance for raw gas cost ranged from 82¢ to \$1.14/MMBTU content of the raw gas in the low price scenario to \$2.61 to \$3.09 per MMBTU in the high price scenario.

For the LaBarge composition, the CNG process has greater net annual revenues (16% higher) than the CFZ-CNG process for the low price scenario and essentially equal net annual revenues for the high price scenario. This is due to the lower cost and lower carbon dioxide recovery of the CNG process.

Table IV-8
Assumed Product Prices

Product		High Price	Low Price
<i>LaBarge Composition Case</i>			
Methane	(\$MMBTU)	3.00	1.50
CO ₂	(\$MSCF)	1.00	0.70
Sulfur	(\$/LT)	110	60
Helium (liq)	(\$MSCF)	50	40
<i>Generic Composition Case</i>			
Methane	(\$MMBTU)	3.00	1.50
Sulfur	(\$/LT)	130	70

Table IV-9
Annual Revenues
CFZ/CNG Claus Process, LaBarge Composition Case

	Rate		Unit Price	High Prices			Low Prices		
				Annual Revenue (\$M)	Per MSCF of Feed Gas		Annual Revenue (\$M)	Per MSCF of Feed Gas	
Methane Residue Gas	37.50	MMSCFD	2.94 \$/MSCF	\$38,548			1.47 \$/MSCF	\$19,274	
			3 \$/MBTU				1.5 \$/MBTU		
CO ₂	75.80	MMSCFD	1 \$/MSCF	\$26,530			0.7 \$/MSCF	\$18,571	
Sulfur	378.00	LT/d	110 \$/LT	\$14,553			60 \$/LT	\$7,938	
Helium	1.10	MMSCFD	50 \$/MSCF (l)	\$19,250			40 \$/MSCF (l)	\$15,400	
Steam									
400 psia	42500	lb/hr	2 \$/M-lb	\$714			1 \$/M-lb	\$357	
55 psia	11500	lb/hr	1.5 \$/M-lb	\$145			0.75 \$/M-lb	\$72	
Total Annual Gross Revenue				\$99,740	\$1.42		\$61,613	\$0.88	
Total Operating Costs				\$18,288	\$0.26		\$18,288	\$0.26	
Capital Investment				\$148,037			\$148,037		
Payout Period (yrs)				5.0			5.0		
Annual Capital Charge				\$29,607	\$0.42		\$29,607	\$0.42	
Total Annual Costs				\$47,895	\$0.68		\$47,895	\$0.68	
Net Annual Revenue (exclusive of raw gas costs)				\$51,845	\$0.74		\$13,717	\$0.20	
<hr/>									
Breakeven Allowance for Raw Gas Cost									
(\$/MSCF)					\$0.74			\$0.20	
(\$/MBTU)					\$3.09			\$0.82	
On-stream days per year :	350								
	95.9%								
Total MMSCF gas processed	70,000								
Raw Gas HHV (BTU/SCF)	240								
Residue Gas HHV (BTU/SCF)	979								

Table IV-10
Annual Revenues
CFZ/CNG Claus Process, Generic Composition Case

	Rate		High Prices			Low Prices		
			Unit Price	Annual Revenue (\$M)	Per MSCF of Feed Gas	Unit Price	Annual Revenue (\$M)	Per MSCF of Feed Gas
Methane Residue Gas	158.50	MMSCFD	3.021 \$/MSCF 3 \$/MBTU	\$167,590		1.511 \$/MSCF 1.5 \$/MBTU	\$83,795	
CO ₂	0.00	MMSCFD	0 \$/MSCF	\$0		0 \$/MSCF	\$0	
Sulfur	37.80	LT/d	130 \$/LT	\$1,720		70 \$/LT	\$926	
Helium	0.00	MMSCFD	0 \$/MSCF (I)	\$0		0 \$/MSCF (I)	\$0	
Steam								
400 psia	4250	lb/hr	2 \$/M-lb	\$71		1 \$/M-lb	\$36	
55 psia	1150	lb/hr	1.5 \$/M-lb	\$14		0.75 \$/M-lb	\$7	
Total Annual Gross Revenue				\$169,396	\$2.42		\$84,764	\$1.21
Total Operating Costs				\$4,477	\$0.06		\$4,477	\$0.06
Capital Investment				\$74,697			\$74,697	
Payout Period (yrs)				5.0			5.0	
Annual Capital Charge				\$14,939	\$0.21		\$14,939	\$0.21
Total Annual Costs				\$19,416	\$0.28		\$19,416	\$0.28
Net Annual Revenue (exclusive of raw gas costs)				\$149,979	\$2.14		\$65,348	\$0.93
<hr/>								
Breakeven Allowance for								
Raw Gas Cost								
(\$/MSCF)					\$2.14			\$0.93
(\$/MBTU)					\$2.61			\$1.14
On-Stream days per year	350							
	95.9%							
Total MMSCF gas processed	70,000							
Raw Gas HHV (BTU/SCF)	822							
Residue Gas HHV (BTU/SCF)	1007							

Table IV-11
Annual Revenues
CNG Claus Process, LaBarge Composition Case

	Rate		High Prices			Low Prices		
			Unit Price	Annual Revenue (\$M)	Per MSCF of Feed Gas	Unit Price	Annual Revenue (\$M)	Per MSCF of Feed Gas
Methane Residue Gas	37.18	MMSCFD	2.94 \$/MSCF	\$38,219		1.47 \$/MSCF	\$19,110	
			3 \$/MBTU			1.5 \$/MBTU		
CO2	55.17	MMSCFD	1 \$/MSCF	\$19,310		0.7 \$/MSCF	\$13,517	
Sulfur	378.00	LT/d	110 \$/LT	\$14,553		60 \$/LT	\$7,938	
Helium	1.10	MMSCFD	50 \$/MSCF (i)	\$19,250		40 \$/MSCF (ii)	\$15,400	
Steam								
400 psia	42500	lb/hr	2 \$/M-lb	\$714		1 \$/M-lb	\$357	
55 psia	11500	lb/hr	1.5 \$/M-lb	\$145		0.75 \$/M-lb	\$72	
Total Annual Gross Revenue				\$92,191	\$1.32		\$56,394	\$0.81
Total Operating Costs				\$16,464	\$0.24		\$16,464	\$0.24
Capital Investment				\$120,059			\$120,059	
Payout Period (yrs)				5.0			5.0	
Annual Capital Charge				\$24,012	\$0.34		\$24,012	\$0.34
Total Annual Costs				\$40,476	\$0.58		\$40,476	\$0.58
Net Annual Revenue (exclusive of raw gas costs)				\$51,715	\$0.74		\$15,918	\$0.23
<hr/>								
Breakeven Allowance for Raw Gas Cost								
(\$/MSCF)					\$0.74			\$0.23
(\$/MBTU)					\$3.08			\$0.95
On-stream days per year :	350							
	95.9%							
Total MSCF gas processed	70,000							
Raw Gas HHV (BTU/SCF)	240							
Residue Gas HHV (BTU/SCF)	979							

V. OVERVIEW OF GAS TREATMENT TECHNOLOGY

The general categories of gas treating process technology which might be considered applicable for separating acid gases from the target low quality gas containing significant acid gas concentrations include absorption, distillation, and membranes. Adsorption processes using solid sorbents, generally reserved for polishing operations or guard beds where small quantities of acid gas must be removed, are not considered applicable to removal of bulk acid gases.

Sulfur recovery technology generally falls within two major categories: direct liquid phase oxidation and Claus technology. Both of these sulfur recovery technologies are discussed, with emphasis on Claus technology because it and its various modifications are essentially the only candidates for large scale recovery of commercially acceptable sulfur from hydrogen sulfide.

The Gas Research Institute has an ongoing program in sulfur recovery research. As a basis of comparison for new gas treatment and sulfur recovery technologies, GRI has selected the amine/Claus combination to represent "state-of-the-art." GRI further comments that *"it would be unlikely for any technology to have an advantage over this tried-and-true (amine/Claus) technology provided it is applicable, which is typically for situations requiring modest sulfur recovery, i.e., about 97%, and at large tonnages, i.e., of the order 25 TPD of sulfur."* [19]

A. Acid Gas Removal Technology

Separation of acid gases from low quality natural gas is required to produce pipeline specification methane. A secondary separation of acid gases, more difficult and little appreciated, is often required to yield concentrated hydrogen sulfide for conversion to elemental sulfur and carbon dioxide of sufficient purity for available markets or atmospheric venting. While much process technology is under development or available commercially [20], the overwhelming majority of acid gas removal technology in commercial service is based on amine solvents which chemically absorb acid gases. Physical solvents are also used, and are effective for high acid gas partial pressures. They include cold methanol (Rectisol) and Selexol. Other processes applicable to *bulk* acid gas removal from low quality natural gas include those based on distillation and perhaps membranes. These three classes of acid gas removal processes, solvents, distillation, and membranes, are discussed briefly.

Other technologies and processes which can remove hydrogen sulfide and/or carbon dioxide from low quality gas are commercially available, or are in various stages of development; they include electrochemical and biologic treatments. Some have been developed for small well-defined niches in the gas processing market, but none are believed competitive for removal of bulk acid gases in large scale gas processing plants. They are beyond the scope of this brief review. (The CFZ/CNG-Claus process has targeted the relatively large quantity of low quality gas containing >10% CO₂ which may also contain either or both of H₂S and N₂ and which cannot be economically up-graded by other known processes.)

A.1. Solvent Based Absorption Processes

Absorption processes are characterized by a recirculating solvent that counter-currently contacts the gas being treated in an absorber tower. Acid gases are removed by chemical reaction with a (chemical) absorbent, or by physically dissolving in a (physical) absorbent. Absorbed acid gases are separated from the solvent in a step called solvent regeneration. Absorbed acid gases are separated from the solvent by application of heat (chemical solvents) and by inert gas stripping or pressure reduction (physical solvents). Absorption processes are used to treat low quality natural gas containing a broad range of acid gas concentrations, including concentrations up to 60-70% acid gases. Chemical solvent processes are generally more suitable for the fine removal of moderate concentrations of acid gases. Physical solvent processes are generally more suitable for the bulk removal of very large amounts of acid gas (carbon dioxide).

A.1.1. Chemical Solvents

amines

Conventional sour gas treating technology was developed to remove relatively minor amounts of H₂S (several mole % or less) from natural gas otherwise meeting pipeline requirements. This technology is based in large part on a class of absorbent compounds known as amines which chemically bind with H₂S to remove it from natural gas. Properly regenerated amine based solvents can be continuously recycled to remove H₂S to pipeline specs of 4 ppm or less. However, if the concentration of H₂S increases above several mole%, or if CO₂ is present, the economics of gas cleaning with amine based solvents become less attractive due to increased solvent flows, vessel sizes and energy consumption to regenerate and pump solvent.

Amine solvents readily co-absorb both H₂S and CO₂, and thus these processes are limited to acid gas removal where contaminant compositions fall within two distinct ranges: 1) CO₂ is essentially the only impurity; or 2) if H₂S and CO₂ are both present, the CO₂:H₂S ratio doesn't exceed about 11:1. This ratio is dictated because the off-gas from amine solvent regeneration, without further processing, must be sufficiently concentrated in H₂S (>20%) to be treated in a conventional Claus sulfur recovery unit.

Basic chemistry of the process can be represented by reactions occurring when solutions of primary amine are used to absorb CO₂ and H₂S. Each alkanolamine has at least one hydroxyl group and one amino group; the hydroxyl group serves to reduce the vapor pressure and increase the water solubility, while the amino group provides the alkalinity in water solutions to cause the absorption of acid gases. The absorbed acid gases form chemical compounds with the amines which have appreciable vapor pressures under normal conditions. Because the vapor pressures of these compounds increase rapidly with temperature, the amine solvent can be regenerated by the application of heat to strip the absorbed gases from the solution.

The absorber operates at the temperature and pressure of the raw gas stream being processed. The pressure normally ranges from 75 psia to over 1000 psia, although the use of DEA has minimum pressure limitations (about 250 to 300 psig). The gas stream enters the bottom of the absorber and flows upward, contacting the amine solution which flows downward from the top. Continuous intimate contact between the gas and the solution is

attained by packing or trays in the absorber. The counter-current flow results in efficient extraction of the acid gases because the regenerated amine solution contacts the lowest acid gas concentration.

Contaminants other than H_2S and CO_2 in the raw gas introduce two concerns: how well will the amine solvent remove these contaminants from the raw gas, and will the amine solvent subsequently regenerate with respect to these contaminants. Carbonyl sulfide and carbon disulfide generally react with amines to yield well-defined compounds which are not regenerable by heat; diethanolamine is one exception. Mercaptans do not form compounds similar to those of H_2S and CO_2 , and are therefore removed only to the extent of their (quite small) physical solubility in the amine solution. The presence of appreciable mercaptan requires additional processing steps to insure their removal.

Low quality natural gas treatment costs vary with many conditions, some of which are site specific. An example has been calculated based on a GRI study [1,21-24] of a New Mexico site using the amine MDEA to remove H_2S . The raw gas flow of 20 MMSCFD contains 18.2% CO_2 and 1.68% H_2S . A five year simple payback on capital, and sulfur sales at \$100/LT (LT = long ton, 2,240 pounds) are assumed. The amine treater produces a 20% H_2S feed to a conventional Claus sulfur recovery unit (SRU). The H_2S removal cost is about 41¢/MMBtu of pipeline gas produced.

The cost to lower the CO_2 concentration in the above example to 2% pipeline quality by amine treatment is estimated at 27¢/MMBtu (taking into account that the H_2S treatment also lowers the CO_2 concentration from 18% to 13% in the H_2S -free sweet gas). Thus, an estimated cost to remove both H_2S and CO_2 from a typical low quality natural gas to produce pipeline specification gas is 68¢/MMBtu, a large fraction of today's wellhead price of \$2.00 to \$2.50/MCF for natural gas.

The conventional amine-based sour gas treatment processes discussed above do not remove nitrogen from the natural gas and they do not produce a sweetened product gas which is conditioned for cryogenic nitrogen rejection, the only current economic approach for separating N_2 and CH_4 [25].

carbonate solutions

Solutions of sodium and potassium carbonate have found application for removal of H_2S and CO_2 from raw gases, particularly synthesis gases. Carbonate solutions probably predate amine solvents as evidenced by their early use for CO_2 recovery from flue gas for the manufacture of dry ice [26]. Development of processes based on carbonate solutions has been driven primarily by the need to remove bulk carbon dioxide from synthesis gases while taking maximum advantage of synthesis gas conditions of high CO_2 partial pressure and high temperature.

Perhaps the most widely recognized process based on carbonate solutions is the Hot Potassium Carbonate, or Benfield, Process. Many improvements to the basic process were made by Benson and Field, original innovators of the process at the U.S. Bureau of Mines. They include the development of solution additives to lower costs and improve gas purities, improvements in energy economy, and advances in partially selective removal of H_2S in the presence of CO_2 [27,28,29]. Contaminants present in gas streams in small concentrations are removed in different degrees by hot potassium carbonate solutions, including carbonyl

sulfide, carbon disulfide, and mercaptans; the long term fate of such contaminants in the process is less discussed [30]. Much of the comparative operating data reported in the literature on the Benfield Process has been gained in treatment of ammonia synthesis gas and crude hydrogen, rather than natural gas.

chemical solvents to treat LaBarge low quality gas

To treat LaBarge reservoir gas, chemical solvents require large amounts of regeneration energy due to the high acid gas content of LaBarge gas. Estimates of the required regeneration energy for chemical solvents such as MEA, DEA, and hot potassium carbonate range from 800 to 1,900 million Btu/hr [32]. For comparison, the higher heating value of the methane in the feed gas is 1,700 million Btu/hr. *Chemical solvent regeneration would require from half to 100% of the product methane be burned to supply the necessary thermal energy.*

A.1.2. Physical Solvents

Physical solvents rely on the solubility of acid gases for their removal from gas streams. There are many physical solvents each with special properties and niches in the gas treatment industry; two representative processes (solvents) which have achieved commercial acceptance are discussed here. Cold methanol is a public domain solvent available for use by one skilled in the art, while Selexol® is a proprietary solvent sold by Union Carbide.

cold methanol

Cold methanol is an effective solvent for carbon dioxide and hydrogen sulfide, especially at low temperatures and elevated pressures; trace contaminants are also absorbed by cold methanol, although less aggressively than CO₂ and H₂S [31]. Carbon dioxide is readily released from solution by pressure reduction, but fine regeneration with respect to H₂S requires steam heating of the methanol, resulting in large temperature swings from absorber to regenerator. There is evidence that although cold methanol is effective in first pass removal of trace contaminants, they tend to build up in the circulating methanol and permit eventual breakthrough of the contaminants with the product gas [4]

Cold methanol offers lower energy consumption than chemical absorbents, reasonable removal of trace impurities, and a dehydrated product gas. The primary disadvantages of cold methanol are its complex flow scheme and significant vaporization losses of methanol even at low temperatures.

The process is generally known by the name Rectisol, developed in Germany by Lurgi Gesellschaft. The original patents on the use of cold methanol for acid gas removal have long since expired, and gas clean-up processes using cold methanol are in general in the public domain. For example, final CO₂ removal in the CNG-Claus conceptual process is accomplished with cold methanol (see Section IV).

Selexol®

The Selexol process, a physical absorption process, is used widely for bulk CO₂ removal from natural and synthesis gases, and for selective removal of H₂S from a variety of raw gases. Selexol solvent is the dimethyl ether of polyethylene glycol, available from Union Carbide. Selexol has several attractive features, including greater solubility for H₂S

than CO₂ (making selective absorption possible), good solubilities for common natural gas contaminants (COS is troublesome, however, and Selexol flows must be increased over that needed for H₂S removal to insure COS removal), good stability and non-reactivity with gas constituents, and low vapor pressure. In process configuration, Selexol is similar to most physical absorption processes: acid gas absorption at high pressure, solvent regeneration by steam heating and pressure reduction, lean solvent recirculation to the primary absorber.

During a Northwest Pipeline feasibility study performed by Stearns Rogers and completed in 1982, the Selexol solvent process emerged as the most economical well-proven technology which could be used to selectively remove acid gases from the unique high acid gas mixtures produced from the LaBarge reservoir of low quality natural gas [32]. Selexol is used by Exxon at Schute Creek, Wyoming, to treat about 0.5 BCF (billion cubic feet) of raw gas produced from the LaBarge reservoir. The emergence of an EOR carbon dioxide market, together with helium and sulfur by-products, made production of the LaBarge gas economic. Selexol does not remove all COS at Schute Creek, and COS is a troublesome contaminant, especially of the vent CO₂.

The major plant power consumers are the carbon dioxide product compressors necessary to produce 2000 psia EOR CO₂. The high CO₂ partial pressure in the raw gas is destroyed in the step of solvent regeneration by pressure let-down. In this case, the Selexol flowsheet is somewhat complicated by the necessity to make the maximum amount of CO₂ from the Selexol regeneration available at elevated pressures. By comparison, the CFZ-CNG process produces a large fraction of the separated CO₂ as liquid at triple-point pressure, reducing process energy requirements if the CO₂ is to be reinjected as high pressure 2000 psia EOR CO₂.

Stearns & Rogers describes use of the Selexol solvent to treat 200 MMSCFD of LaBarge gas [32]. The capital cost is reported as \$210 million (in 1983 \$'s). Selexol refrigeration, and CO₂ compressors represented almost 50% of the total plant cost. The estimated cost in 1991 dollars (using the *Chemical Engineering* plant cost index) is about \$249 million. Capital cost of the CFZ-CNG at 200 MMSCFD scale is estimated to be about \$138 million, about 42% less than the anticipated cost of Selexol. The amount of methane and sulfur product are about the same for the two designs (Stearns 1983 and Acirion's conceptual design presented herein) but the Selexol design produces nearly twice as much EOR grade carbon dioxide, 131 versus 76 MMSCFD. The operating costs estimated by Stearns Rogers were about \$21.1 million (1983) per year based on 350 days per year operation at peak summer atmospheric conditions.

A.2. Distillation Processes

Distillation processes, to a large extent, do not use external agents to effect separation of methane from acid gases. If the acid gases are intended for any commercial use, these processes are generally more efficient because they produce the acid gases at pressure, usually in liquid form. However, distillative separation of methane from acid gases requires low temperatures, and thus refrigeration, with its own attendant compression and heat exchange requirements, is essential to each process configuration.

The major barrier to simple distillative separation of methane from carbon dioxide to produce high purity methane is the formation of solid carbon dioxide. Much process

development effort has been expended to insure that conditions favorable to the formation of solid carbon dioxide are avoided in the cryogenic distillation of methane and carbon dioxide; two processes which take this preventive approach are discussed briefly. Another development effort discussed fully herein (Exxon's CFZ process) has taken the opposite approach, and actually provides for the formation and melting of solid carbon dioxide within a single distillation column separating methane from carbon dioxide.

A.2.1. Dual Pressure Distillation Process

Schianni [33] discusses a dual pressure distillation process conceived to remove bulk carbon dioxide from methane, and permit economic exploitation of low quality natural gases. The process, developed by Snamprogetti SpA, Milano, Italy, consists essentially of two columns operating at different pressure levels. The first high pressure column produces a carbon dioxide bottoms product, and a methane-CO₂ overhead product which feeds a second, lower pressure column. This second column produces an overhead methane product of pipeline quality, and a methane-CO₂ bottoms product which is recycled to the first column. Pressures of the two columns must be carefully selected and controlled to avoid solid CO₂ formation.

Schianni compared the dual pressure distillation process against what he considered the best available process for bulk CO₂ removal, the Selexol process. The estimated grass roots plant cost and annual operating cost were each about 2/3 of the corresponding estimated cost for Selexol. No public domain information has been found regarding further development or commercialization of the dual pressure concept for low quality natural gas treatment.

A.2.2. Ryan-Holmes Processes

The Ryan/Holmes technology is a collection of distillation separations based on the use of extractive agents employed to enhance the volatility of key components, to raise operating temperatures in the separation of methane and carbon dioxide, and to prevent the formation of solid carbon dioxide. The most common extractive agent is a butanes-plus mixture of hydrocarbons. A number of patents have issued for this technology [34-38], and several commercial installations have been licensed [39,40]. The technology is currently offered by Process Systems International, Westborough, Massachusetts; a primary market application appears to be CO₂ recovery from enhanced oil recovery project.

Although the Ryan/Holmes approach uses no chemicals external to the raw natural gas system, a mixture of C₄+ hydrocarbons is deliberately added to the distillation process which must be separated downstream for recycle within the process. Thus, to make the initial methane-acid gas separation, at least two columns are necessary in order to recover and recycle additive. Additional columns as desired, also using the extractive hydrocarbon agents, can be employed to separate relatively pure carbon dioxide, and to achieve high recovery of ethane. The fate of trace contaminants present in natural gas systems in Ryan/Holmes type processes is not well documented, and may be of some concern given the additional required separation of extractive agent. If the feed gas does not contain any C₄+ hydrocarbons, as is the case for the LaBarge gas, then they must be added as makeup for losses within the process. The cost of the hydrocarbon makeup can make the gas processing costs prohibitively high.

A.2.3. Exxon Controlled Freeze Zone Process

The Exxon Controlled Freeze Zone (CFZ) Process is unique in that it deliberately encourages formation of solid carbon dioxide and in a single distillation column producing methane overhead and carbon dioxide bottoms marches directly through portions of the methane-CO₂ phase diagram where other distillative processes fear to tread [5,6,7]. CFZ provides the methane-CO₂ separation capability for one embodiment of the proposed process for treating low quality natural gas, and as such, is discussed in greater detail in Section III.A.1.

A.3. Membranes

Membranes separate gaseous components by selective movement or permeation of the components across the membrane. The various gas species in contact with the membrane dissolve in the membrane material and are transported across the membrane at different rates under externally imposed partial pressure gradients. Thus membrane separations are pressure driven. The high pressure product is called the residue; the low pressure product which preferentially moves through the membrane is called the permeate. Recovery of residue or permeate decreases as purity requirements are increased.

Currently, nitrogen separation from ambient air is the most widespread application of membranes, but CO₂ removal from low quality natural gas is also practiced, reducing CO₂ content to less than 2%. However, membranes alone are probably not able to meet the H₂S pipeline specification of 4 ppm, especially at low H₂S partial pressures in the feed gas [41]. For production of pipeline quality gas, hybrid membrane-amine processes are being examined, membranes being used to remove the major portion of acid gases, with final processing to pipeline quality being accomplished by conventional absorption technologies.

Membranes are similar to solvent based absorption processes in that they substantially degrade the often high partial pressure of CO₂ available in the raw gas. Thus, CO₂ recompression is necessary for membrane recovery of CO₂ in CO₂-enhanced oil recovery applications where a number of membrane systems are now in service.

B. SULPHUR RECOVERY TECHNOLOGY

Due to the unique nature of the upstream separation processes, there was no specific precedent to utilize an existing process or technology for removing and/or recovery the sulphur from the H₂S rich stream. Therefore an extensive review of all existing commercial and many experimental processes was completed.

Due to the wide range of potential processes it was necessary to categorize them. Following are the categories used for this evaluation and a description of the most significant features of the category. Where available, there is a simplified process flow diagram for each process. The following process descriptions are presented here in summary form in order to give an overview of the alternative technologies. More detailed process descriptions are available in various references [42,43]

B.1. Commercial Sulphur Recovery Processes

This group of processes represents those processes which have been installed in commercial scale sulphur recovery facilities. Many of them have found successful commercial application and represent the best available current technology.

B.1.1. Conventional Claus Process

This process is a non-proprietary process which is used extensively in the natural gas and crude oil industries to produce elemental sulphur from H_2S laden feed streams. A typical process flow diagram is shown in Appendix F. The main process steps are the partial oxidation of H_2S with air in a free flame reaction furnace followed by a series of condensers to remove sulphur product and catalytic gas phase reactors.

In the first reaction step, the acid gas feed is partially oxidized with air in the free flame reaction furnace. Typically, the reaction stoichiometry is controlled so that 1/3 of the H_2S is oxidized. In this first reaction step a significant amount of elemental sulphur is formed directly (typically from 50 to 75 percent of the total). The reaction furnace effluent contains a mixture of acid gases that contain H_2S and SO_2 in a 2:1 ratio. This gas stream is subsequently reacted over a solid catalyst in order to further promote the reaction to elemental sulphur. The sulphur product is removed at intermediate points in the process via condensation in order to promote the reaction closer to equilibrium values.

Table V-1 shows the conventional Claus process is capable of processing a wide range of acid gas feeds. The process has been proven commercially viable at production rates of 5 to 2000 LTD. The recovery performance capability of a conventional Claus plant is practically limited to approximately 99 percent of the total inlet sulphur [44].

Table V-1
Conventional Modified Claus Process

Process	Conventional Claus
Commercially Available	Yes
Acid Gas Composition (mole % H_2S)	15 - 100
Practical Recovery Efficiency (%)	≈ 99.0

B.1.2. Gas Phase - Claus Continuation Processes

These processes include a number of licensed technologies which are used extensively in the natural gas and oil processing industry. Following is a comprehensive listing of the available processes and the owner of the process license :

- III CBA - Amoco
- III Maxisulf - Davy McKee
- III MCRC - Delta Projects
- III Sulfreen - SNEA / Lurgi

Figures in Appendix F show the typical process flow diagram for each of the processes (except Maxisulf). While the process flow scheme varies somewhat between the various systems, the fundamental process idea is the same. In each case the Claus plant effluent or "tail gas" is further reacted over a solid catalyst in the gas phase in order to continue the $\text{H}_2\text{S}/\text{SO}_2$ reaction. The overall reaction yield is enhanced by allowing the reaction to occur at a temperature below the sulphur condensation point. This effectively removes the sulphur product from the reaction phase in order to closer approach the reaction equilibrium. The lower reaction temperatures also enhance the overall sulphur yield.

These processes are utilized strictly as an extension to a conventional Claus plant in order to improve the overall recovery efficiency. As a group they are practically limited to an overall sulphur recovery efficiency of approximately 99.5 percent. These processes have been installed on Claus systems which process a wide variety of feed streams over a wide range of inlet acid gas flow rates. Table V-2 summarizes the capability of each of the Claus continuation processes.

Table V-2
Claus Continuation Processes

Process	CBA	Maxisulf	MCRC	Sulfreen
Commercially Available	Yes	Yes	Yes	Yes
Acid Gas Composition (mole % H_2S) ¹	15 - 100	15 - 100	15 - 100	15 - 100
Feed Gas Composition (mole % $\text{H}_2\text{S} + \text{SO}_2$) ²	> 3.0	> 3.0	> 3.0	> 3.0
Practical Recovery Efficiency (%)	< 99.5	< 99.5	< 99.5	< 99.5

¹ Based on installation downstream of a conventional Claus process.

² Actual feed gas composition to the specific process.

B.1.3. Gas Phase - Direct Catalytic Oxidation Processes

These processes are individually licensed and have also been used extensively in existing facilities. The general process principle is to promote the direct oxidation of H_2S to elemental sulphur over a specialized solid catalyst. This process group includes the following processes:

- Catasulf - BASF AG
- MODOP - Mobil Oil Corp.
- Selectox - Unocal Science and Technology, Ralph M. Parsons
- SuperClaus - Comprimo B.V.

Process flow diagrams for these processes are given in Appendix F. The Catasulf and Selectox processes can be used on a stand-alone basis in applications with a specific acid gas

feed composition. In this configuration both of these processes have recovery capabilities similar to the conventional modified Claus process.

The MODOP and SuperClaus processes would typically be installed downstream of a conventional Claus plant. In general they promote the H₂S oxidation reaction closer to the equilibrium value to give overall recovery efficiencies up to 99.5 percent. Table V-3 summarizes the capabilities of these processes.

Table V-3
Gas Phase - Direct Catalytic Oxidation Processes

Process	Catasulf ¹	MODOP ²	Selectox ¹	SuperClaus ²
Commercially Available	Yes	Yes	Yes	Yes
Acid Gas Composition (mole % H ₂ S) ³	na	15 - 100	5 - 100	15 - 100
Feed Gas Composition (mole % H ₂ S) ⁴	5 - 15	> 3.0	5 - 100	> 3.0
Practical Recovery Efficiency (%)	97.5	< 99.6	< 99.5	< 99.0

¹ Based on stand alone installation.

² Based on installation downstream of a conventional Claus process.

³ Acid gas composition to upstream conventional Claus plant.

⁴ Feed gas composition to specific process.

B.1.4. Liquid Phase - Direct Catalytic Oxidation Processes

This process group includes those processes which utilize a liquid phase catalytic agent to remove the H₂S from the process gas stream and then promote direct oxidation of the H₂S to elemental sulphur. These processes are only capable of treating sour streams with H₂S. If other sulphur compounds are present, as in typical Claus plant tail gases, all of the sulphur components must first be converted to H₂S. This is normally accomplished by use of a hydrogenation reactor system. Appendix F contains the simplified process flow diagrams for each of the processes.

▣ LoCat - ARI Technologies

▣ Stretford - Ralph M. Parsons Co.

▣ SulFerox - Shell Oil Co.

▣ Sulfint - Le Gaz Integral Enterprise

▣ Sulfolin - Linde AG.

All of the processes have been successfully installed in a commercial application and are usually installed as a tail gas clean up process on a conventional Claus plant. The exception to this type of configuration is where the raw sour gas stream has a very low H₂S content in which case they can be used directly.

As a group they are capable of achieving sulphur recovery efficiencies of nearly 100 percent and often the vent gas streams can be vented directly to atmosphere. They have also shown commercial feasibility as tail gas clean-up processes on sulphur recovery units with a wide range of acid gas feed compositions and flow rates. Table V-4 summarizes each of these processes.

Table V-4
Liquid Phase - Direct Catalytic Oxidation Processes

Process	LoCat ¹	Stretford ²	SulFerox ²	Sulfint ²	Sulfolin ²
Commercially Available	Yes	Yes	Yes	Yes	Yes
Acid Gas Composition (mole % H ₂ S) ³	15 - 100	15 - 100	15 - 100	15 - 100	15 - 100
Feed Gas Composition (mole % H ₂ S) ⁴	5 - 15	> 3.0	> 3.0	> 3.0	> 3.0
Practical Recovery Efficiency (%)	99.9+	99.9+	99.9+	99.9+	99.9+

¹ Based on stand alone installation.

² Based on installation downstream of a conventional Claus process.

³ Acid gas composition to upstream conventional Claus plant.

⁴ Feed gas composition to specific process.

B.1.5. Liquid Phase Claus Processes

These processes utilize the concept of promoting the continued reaction of H₂S and SO₂ in the presence of a liquid catalyst after the sour gases have been absorbed into the liquid reagent. In general, these processes are single stage, two phase absorber/reactors which produce clean elemental sulphur directly off the bottom of the reaction vessel. This group of processes includes:

- AquaClaus - Stauffer Chemical Company
- ASR Sulphoxide Process - Alberta Sulphur Research
- IFP Process - Institute Francais du Petrol
- RSRP Process - Alberta Energy Company
- Townsend Process

Table V-5 summarizes the capability of these processes. Process flow diagrams for the AquaClaus, IFP and RSRP processes are shown in Appendix F.

B.2. Non-Commercial Sulphur Recovery Processes

Since the advent of the conventional modified Claus process, various process developers have attempted to alter the configuration or operating conditions in order to

improve the recovery capability. In general they have not shown commercial feasibility because the suggested advantages have not been sufficient for commercial applications. They are listed below with references for more detailed descriptions:

- III High Pressure Claus Process for Sulfur Recovery - US patent 4,684,514
- III Low Temperature Converter and Feedline Arrangement, Canadian patent 1,126,932
- III U.S.B.M Citrate Process - WR seminar manual
- III Pritchard CLEANAIR - WR seminar manual

While each of these processes has demonstrated some additional benefits to the original Claus process, they do not have any substantial advantages which could be applied to the CNG Claus system.

Table V-5
Liquid Phase - Claus Continuation Processes

Process	Aqual Claus ¹	ASR Sulphoxide ¹	IFP ¹	RSRP ²	Townsend ¹
Commercially Available	Yes	No	Yes	No	No
Acid Gas Composition (mole % H ₂ S)	up to 5.0 (est.)	up to 3.0 (est.)	15 - 100 ¹	> 5.0	15 - 100 ¹
Practical Recovery Efficiency (%)	99+	99.9	99.3	99.9+	unknown

¹ Based on installation downstream of a conventional Claus process.

² Based on stand-alone installation.

B.3. Other Processes

Though not commonly used in the sour natural gas or oil industries, there are other commercial processes which are capable of producing usable end products from H₂S. A listing of the process and their associated end products is as follows:

- III ATS - Coastal States Gas - ammonium thiosulphate
- III Thoroughbred 101 - Chiyoda - gypsum
- III Saarberg/Holter - Saarberg/Holter GmbH - gypsum

These processes are not commonly used in industrial applications.

B.4. Supplementary H₂S Removal Processes

In current industrial gas processing plants, the traditional sulphur recovery technologies are often supplemented with a process which ensures very high overall recovery efficiencies and very low total sulphur emissions. These processes do not directly recover sulphur from the sour gas, but rather, capture all of the sulphur compounds from the tail gas for return to the upstream sulphur recovery process. Therefore, the sulphur compounds are ultimately recycled to extinction.

These processes are based on one of two possible technologies ; conversion of all sulphur species to H₂S followed by H₂S capture for return to the upstream SRU, or oxidation of all sulphur compounds to SO₂ followed by SO₂ capture for return to the upstream SRU. Following is a list of the identified processes :

B.4.1. H₂S Based Processes

- III BSR/MDEA - R.M. Parsons
- III Flexsorb - Exxon
- III SCOT - Shell
- III Trencor M - Trentham Corp.

Appendix F shows a typical process flow diagram for each process except the Trencor M process. The main process steps for this group of technologies are :

1. Conversion of all sulphur compounds in the SRU tail gas to H₂S in a hydrogenation reactor in the presence of excess hydrogen.
2. Process gas cooling and water removal by heat exchange and direct contact water quench.
3. Removal of H₂S from the dry tail gas by selective absorption in an direct amine contactor.
4. Removal of the concentrated H₂S from the rich amine via steam stripping.
5. Recycle of the concentrated H₂S .

Most of these processes have demonstrated commercial experience on a wide range of SRU tail gas streams. The proven removal efficiency in these units results in overall recovery efficiencies in excess of 99.9 percent. Table V-6 summarizes the capabilities of these processes.

Table V-6
H₂S Capture and Recycle Processes

Process	BSR/MDEA	Flexsorb	SCOT	Trencor M
Commercially Available	Yes	Yes	Yes	No
Acid Gas Composition (mole % H ₂ S) ¹	15 - 100	15 - 100	15 - 100	15 - 100
Practical Recovery Efficiency (%) ²	99.9+	99.9+	99.9+	99.9+

¹ Based on installation downstream of a conventional Claus process.

² Overall recovery capability of combined system.

B.4.2. SO₂ Based Processes

- ▮ Clintox - Linde AG
- ▮ LUCAS Process - Lurgi
- ▮ UCAP - Union Carbide
- ▮ Wellman Lord - Davy McKee

Process flow diagrams for the Clintox and Wellman-Lord processes are shown in Appendix F. The main process steps for this group of technologies are:

1. Conversion of all sulphur compounds in the SRU tail gas to SO₂ via oxidation.
2. Process gas cooling and water removal by heat exchange and direct contact water quench.
3. Removal of SO₂ from the dry tail gas by selective absorption in a gas-liquid contactor.
4. Separation of the SO₂ rich absorbent material.
5. Recycle of the concentrated SO₂.

Of these processes, only the Wellman-Lord and Clintox processes have demonstrated commercial success. In general they are capable of meeting all of the specified criteria when operated in conjunction with a upstream Claus SRU. However, the cost and performance of these processes is no better than the much more experienced H₂S systems and therefore they are generally not favored for industrial applications.

VI. CONCLUSIONS

A technical and economic evaluation of the CFZ, CNG and high pressure Claus processes integrated to treat 200 MMSCFD of low quality natural gas has been completed. The technical feasibility of the proposed integrated process has been established. The processes are economically attractive and superior to conventional sweetening processes over the range of product prices assumed in this study. The CNG process is economically superior to the CFZ/CNG process at lower EOR carbon dioxide prices.

Low quality natural gas processing with the integrated CFZ/CNG Claus process is feasible for low quality natural gas containing 10% or more of CO₂, *and any amount of H₂S*. The CNG Claus process requires a minimum CO₂ partial pressure in the feed gas of about 100 psia (15% CO₂ for a 700 psia feed gas). The process is well suited for handling a variety of trace contaminants usually associated with low quality natural gas and Claus sulfur recovery.

The integrated process can produce high pressure carbon dioxide at purities required by end use markets, including food grade CO₂. A major objective of the Optional Program Research Plan is demonstration of triple point crystallization of CO₂ to purify CO₂ at the rate of 25 tons/day. The ability to economically co-produce high pressure CO₂ as a commodity with significant revenue potential frees process economic viability from total reliance on pipeline gas, and extends the range of process applicability to low quality gases with relatively low methane content.

Gases with high acid gas content and high CO₂ to H₂S ratios can be economically processed by the CFZ/CNG Claus and CNG Claus processes. The large energy requirements for regeneration make chemical solvent processing prohibitive. The cost of Selexol physical solvent processing of the LaBarge gas is significantly greater than the CFZ/CNG Claus and CNG Claus processes.

CIIG's high pressure Claus sulfur recovery process is economically superior to conventional Claus processes, and permits essentially 100% sulfur recovery. Western Research studied the thermodynamic and mechanical aspects of the new process and concluded the processing scheme is feasible. However, since there is little experimental data relating to sulfur recovery at high pressure, further development efforts as proposed in the accompanying Research Plan are deemed necessary to confirm several key operating parameters.

VII. REFERENCES

1. Hugman, R.H., E.H. Vidas, and P.S. Springer, "Chemical Composition of Discovered and Undiscovered Natural Gas in the Lower 48 United States," GRI Contract 5088-222-1745, GRI 90/0248, November, 1990.
2. Meyer, Howard S., Presentation, Chemical Process Research Department, Gas Research Institute, October 3, 1990.
3. Kohl, Arthur, and Fred Riesenfeld, *Gas Purification*, Gulf Publishing Company, Houston, 1974.
4. Rosseau, R.W., et al, "Regeneration of Physical Solvents in Conditioning Gases from Coal," in *Acid and Sour Gas Treating*, Stephen A. Newman, Ed., Gulf Publishing Company, Houston, 1985, p 131.
5. Valencia, J. A. and Denton, R. D., "Method and Apparatus for Separating Carbon Dioxide and Other Acid Gases from Methane by the Use of Distillation and a Controlled Freeze Zone", U.S. Patent 4,533,372, August 6, 1985.
6. Victory D. J., and J. A. Valencia, "The CFZ Process: Direct Methane-Carbon Dioxide Fractionation," *Hydrocarbon Processing*, 66 (5) 1987.
7. Haut, R. C., et. al., "Development and Application of the Controlled Freeze Zone Process," SPE Production Engineering, p 265, August 1989.
8. Donnelly, H. G. and Katz, D. L., "Phase Equilibria in the Carbon Dioxide - Methane System", *Ind & Eng Chem*, Vol. 46, No. 3, p. 511, March 1954.
9. Brown, W.R., et al, "Gas Separation Process," U.S. Patent 4,270,937, June 2, 1981.
10. Brown, W.R., et al, "Gas Separation Process," U.S. Patent 4,581,052, April 8, 1986.
11. Brown, W.R., et al, "Gas Separation Process," U.S. Patent 4,609,388, September 2, 1986.
12. Brown, W.R., Cook, W.J., et al, "Crystallization Process," U.S. Patent 4,623,372, November 18, 1986.
13. Brown, W.R., et al, "Triple-Point Crystallization Separates and Concentrates Acid Gases," presented at AIChE Spring National Meeting, March 27-31, 1983, Houston, Texas.
14. Siwajek, L.A., et al, "CNG Acid Gas Removal Process," Final Technical Report, US Department of Energy, Morgantown Energy Technology Center, Contract DE-AC21-83MC20230, July, 1986.
15. Hise, Ralph E., and Jeffrey Cook, "Sulfur Recovery Process," US Patent 5,021,232, June 4, 1991.
16. Hyne, J.B., "Design and Chemistry of Front End Reaction Furnaces," *Canadian Gas Journal*, March-April 1972, pp 12-16.
17. Siwajek, L.A., Brown, W.R., et. al., "CNG Gas Separation Process," Final Technical Report, Gas Research Institute Contract 5086-222-1429, September 1988.
18. Kurata, Fred, "Solubility of Solid Carbon Dioxide in Pure Light Hydrocarbons and Mixtures of Light Hydrocarbons," Gas Processors Association, Research Report RR-10, February, 1974.
19. Leppin, Dennis, James M Evans, and Kevin Krist, "Gas Research Program in Sulfur Recovery Research," GRI Liquid Redox Sulfur Recovery Conference, May 5-7, 1991, Austin, Texas, Proceedings, GRI-91/0188, June 1991, p 389.
20. Echterhoff, Linda, "State of the Art of Natural Gas Processing Technologies," Topical Report Task 3 Volume I, GRI Contract No. 5088-221-1753, March, 1991, p.389.
21. Hoffman, E J, "Subquality Natural Gas, The Resource and Its Potential," GRI Report issued under GRI Contract No 5082-260-0666, January, 1986.
22. Dalrymple, D.A., Skinner, F.D. and Meserole, N.P., "Investigation of US Natural Gas Reserves, Demographics and Gas Treatment Processes", GRI-91/0019, January 1991.

23. Meyer, H.S., Leppin, D. and Savidge, J.L., "GRI's Gas Processing R&D Program", 69th Annual GPA Meeting, Phoenix, March 12-14, 1990.
24. Smelzer, S.C., Bucklin, R.W., Ramanathan, R.V., Lamprecht, D.G. and Fowler, R., "HiGEE Tech. Development Program Phase I", Topical Report, January 1988 - August 1988, GRI-89/0002.
25. Meyer, Howard S, "Chemical Process Research Technology Brief: Small-Scale Gas Processing Research," Chemical Process Research Department R&D Status Report, Gas Research Institute, December 1989, pp 77-79.
26. Quinn, E.L. and C.L. Jones, *Carbon Dioxide*, Americal Chemical Society Monograph Series, New York, Reinhold Publishing Corporation, 1936.
27. Benson, H.E., J.H. Field, and R.M. Jameson, *Chem Engr Progress*, 50:356, 1954.
28. Benson, H.E., J.H. Field and W.P. Haynes, *Chem Engr Progress*, 52:433, 1956.
29. The Benfield Corporation, *The Way to Low Cost Scrubbing of CO₂ and H₂S from Industrial Gases*, 1971.
30. Parrish, R.W., and H.B. Heilson, "Synthesis Gas Purification Including Removal of Trace Contaminants by the BENFIELD Process," Paper presentee at the National Meeting of the American Chemical Society, Division of Ind & Eng Chem, Los Angeles, March, 1974.
31. Hochgesand, G., *Ind Eng Chem*, 62(7):37-43, 1970.
32. Johnson, John E., and Arthur C. Homme, Jr., "Selexol Solvent Process Reduces Lean, High-CO₂ Natural Gas Treating Costs," Paper 65c, AIChE 1983 Summer National Meeting, Denver, August 29, 1983.
33. Schianni, G.C., "Cryogenic Removal of Carbon Dioxide from Natural Gas," Inst Chem Eng Symp, Ser n 44, 1976, Natl Gas Process and Util, Eur Fed of Chem Eng Meet, 171st, Dublin, Ireland, April 6-9, 1976, Pub by Inst of Chem Eng, Rugby, England, 1976, Sess 1, p 50-55.
34. Holmes, Arthur S., and James M. Ryan, "Distillative Separation of Carbon Dioxide from Light Hydrocarbons," U.S. Patent 4,350,511, September 21, 1982.
35. Ryan, James M., and Arthur S. Holmes, "Distillative Separation of Carbon Dioxide from Hydrogen Sulfide," U.S. Patent 4,383,841, May 17, 1983.
36. Holmes, Arthur S., and James M. Ryan, "Distillative Separations of Gas Mixtures Containing Methane, Carbon Dioxide and Other Components," U.S. Patent 4,462,814, July 31, 1984.
37. O'Brien, John V., "Distillative Separation of Methane and Carbon Dioxide," U.S. Patent 4,383,842, May 17, 1983.
38. Holmes, Arthur S., and James M. Ryan, "Cryogenic Distillative Separation of Acid Gases from Methane," U.S. Patent 4,318,723, March 9, 1982.
39. Ryan, J.M., and F.W. Schaffert, "CO₂ Recovery by the Ryan/Holmes Process," *Chemical Engineering Progress*, October 1984, pp 53-56.
40. "Review of Selected Acid Gas Removal Processes (Draft), The M.W. Kellogg Company, Gas Research Institute Job 6440-01, October 1984.
41. Changela, M.K. and G.J. Reading, "Technical Evaluation of Hybrid Membrane/DEA Modeling," Gas Research Institute Topical Report, GRI-91/0093 (1991).
42. Paskall, H.G., and J.A. Sames, "Sulfur Recovery," Text from Western Research Sulphur Recovery Seminar.
43. Gas Process Handbook '90, *Hydrocarbon Processing*, April, 1990.
44. Paskall, H.G., "Capability of the Modified-Claus Process," A Final Report to the Department of Energy and Natural Resources of the Province of Alberta, March, 1979.

45. Cole, Eric T, et al, "Evaluation of the Exxon CFZ and CNG-Claus Process for the Treatment of Subquality Natural Gas," Final Report # GRI 92/0251, Contract 5090 -222-2011, June 26, 1992.
46. Guthrie, Kenneth M., *Process Plant Estimating Evaluation and Control*, Craftsman Book Company of America, Solana Beach, CA, 1974.
47. Commodities, "As Natural-Gas Prices Soar, Many Analysts Say 7-Year Supply Glut, Bearish Market May Be Over," *Wall Street Journal*, May 3, 1993.
48. Thiel, Klaus F., "Sulfur Recovery by the Claus Process," M.S. Thesis, Case Institute of Technology, 1974.

APPENDIX A

GLOSSARY

of

ABBREVIATIONS AND DEFINITIONS

CFZ	Controlled Freeze Zone
CNG	Consolidated Natural Gas Company
EOR	Enhanced Oil Recovery
EPR	Exxon Production Research Company
GRI	Gas Research Institute
HHV	Higher Heating Value
HP	High Pressure
HPSRU	High Pressure Sulfur Recovery Unit
MSCF	Thousand Standard Cubic Feet
MSCFD	Thousand Standard Cubic Feet per Day
LNG	Liquefied Natural Gas
LT	Long Ton (2,240 lbs)
Mlb	Thousand Pounds Mass
MMSCF	Million Standard Cubic Feet
MMSCFD	Million Standard Cubic Feet per Day
MMBTU	Million British Thermal Units
NRU	Nitrogen Rejection Unit
ppb	Parts per Billion
ppm	Parts per Million
SRU	Sulfur Recovery Unit
TCF	Trillion Cubic Feet
TPC	Triple Point Crystallizer
\$M or \$K	Thousands of Dollars US
\$MM	Millions of Dollars US

APPENDIX B

PROCESS FLOW DIAGRAMS and MATERIAL BALANCES

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Figure B.1-1
CFZ Section Process Flow Diagram
LaBarge Composition Case

Table B.1-1
CFZ Section Material Balance
LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/24 Version 386|C2.10 Case Name ACR-EXXF.SIM
 Time 13:38:53 Prop Pkg PR

CFZ Section <> Labarge Composition

Stream	1	2	3	4
Vapour frac.	1.0000	1.0000	0.1104	0.9240
Temperature F	60.0000	60.0000	-50.0000	-74.9416
Pressure psia	1060.0000	1040.0001	1000.0000	600.0000
Molar Flow lbmole/hr	21960.9399	21956.3111	21956.3111	2423.4082
Helium mole frac.	0.005989	0.005991	0.005991	0.042434
Nitrogen mole frac.	0.074868	0.074884	0.074884	0.277136
Methane mole frac.	0.204639	0.204683	0.204683	0.465853
Ethane mole frac.	0.000100	0.000100	0.000100	0.000068
CO2 mole frac.	0.663830	0.663970	0.663970	0.201895
H2S mole frac.	0.049912	0.049923	0.049923	0.012576
COS mole frac.	0.000450	0.000450	0.000450	0.000038
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.000211	0.000000	0.000000	0.000000

Stream	5	6	7	8
Vapour frac.	0.1338	0.0001	1.0000	1.0000
Temperature F	-57.7361	-149.7168	-149.7205	41.2743
Pressure psia	600.0000	600.0000	600.0000	600.0000
Molar Flow lbmole/hr	19532.9024	18205.3392	6110.6022	11488.4925
Helium mole frac.	0.001469	0.002837	0.021526	0.000000
Nitrogen mole frac.	0.049791	0.143154	0.268912	0.000472
Methane mole frac.	0.172280	0.853856	0.709513	0.044825
Ethane mole frac.	0.000104	0.000000	0.000000	0.000216
CO2 mole frac.	0.721299	0.000153	0.000050	0.900088
H2S mole frac.	0.054556	0.000000	0.000000	0.054189
COS mole frac.	0.000501	0.000000	0.000000	0.000209
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	9	10	11	12
Vapour frac.	0.0000	1.0000	1.0000	1.0000
Temperature F	41.2743	-10.0000	118.5465	100.0000
Pressure psia	600.0000	570.0000	268.9474	263.9474
Molar Flow lbmole/hr	15845.7111	6110.6022	7499.1541	7499.1541
Helium mole frac.	0.000000	0.021526	0.000000	0.000000
Nitrogen mole frac.	0.000061	0.268912	0.000000	0.000000
Methane mole frac.	0.010004	0.709513	0.001000	0.001000
Ethane mole frac.	0.000139	0.000000	0.004000	0.004000
CO2 mole frac.	0.919999	0.000050	0.000000	0.000000
H2S mole frac.	0.069174	0.000000	0.000000	0.000000
COS mole frac.	0.000624	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.995000	0.995000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Table B.1-1 (continued)
CFZ Section Material Balance
LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/24 Version 386/C2.10 Case Name ACR-EXXF.SIM
 Time 14:25:23 Prop Pkg PR

CFZ Section <> Labarge Composition

Stream	13	14	15	16
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	-44.1766	-60.0000	-60.0000	-130.0000
Pressure psia	243.9474	148.9384	148.9384	143.9383
Molar Flow lbmole/hr	7499.1541	4821.3660	2181.1940	4821.3660
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.001000	0.000728	0.000728	0.000728
Ethane mole frac.	0.004000	0.004096	0.004096	0.004096
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.995000	0.995176	0.995176	0.995176
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	17	18	19	20
Vapour frac.	0.0000	0.9999	1.0000	1.0000
Temperature F	-155.0000	-154.7464	36.6964	140.8417
Pressure psia	14.9348	14.9348	143.9383	194.0000
Molar Flow lbmole/hr	4400.0011	4821.3660	7499.1541	18099.1503
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000275	0.000728	0.001000	0.000000
Ethane mole frac.	0.004246	0.004096	0.004000	0.001000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.995479	0.995176	0.995000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.995000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.004000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	21	22	23	24
Vapour frac.	0.2185	0.0000	0.0000	0.0000
Temperature F	100.0998	52.0592	40.0000	40.0000
Pressure psia	189.0000	184.0000	78.4887	78.4887
Molar Flow lbmole/hr	18099.1503	18099.1503	496.2690	623.5660
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.001000	0.001000	0.000883	0.000883
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.995000	0.995000	0.994998	0.994998
i-Butane mole frac.	0.004000	0.004000	0.004119	0.004119
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Table B.1-1 (continued)
CFZ Section Material Balance
LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc

Date 93/05/24 Version 386/C2.10 Case Name ACR-EXXF.SIM

Time 14:37:10 Prop Pkg PR

CFZ Section <> LaBarge Composition

Stream	25	26	27	28
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	40.0000	-10.0000	-10.0000	-10.0000
Pressure psia	78.4887	31.3333	31.3333	31.3333
Molar Flow lbmole/hr	16122.8609	6155.9052	572.5410	6686.5792
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000883	0.000519	0.000519	0.000519
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.994998	0.994829	0.994829	0.994829
i-Butane mole frac.	0.004119	0.004652	0.004652	0.004652
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	29	30	31	32
Vapour frac.	0.0000	0.0000	0.9999	1.0000
Temperature F	-40.0000	-40.0000	-39.5706	-2.1588
Pressure psia	16.1298	16.1298	16.1298	31.3333
Molar Flow lbmole/hr	1935.5130	3966.7509	6686.5792	16122.8609
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000351	0.000351	0.000519	0.000883
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.994677	0.994677	0.994829	0.994998
i-Butane mole frac.	0.004972	0.004972	0.004652	0.004119
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	33	34	35
Vapour frac.	1.0000	0.0000	0.0000
Temperature F	64.8089	-47.0000	28.0045
Pressure psia	78.4887	490.0000	480.0000
Molar Flow lbmole/hr	18099.1503	9225.7278	9225.7278
Helium mole frac.	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000
Ethane mole frac.	0.001000	0.000000	0.000000
CO2 mole frac.	0.000000	0.999996	0.999996
H2S mole frac.	0.000000	0.000004	0.000004
COS mole frac.	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000
Propane mole frac.	0.995000	0.000000	0.000000
i-Butane mole frac.	0.004000	0.000000	0.000000
H2O mole frac.	0.000000	0.000000	0.000000

Figure B.1-2
 TPC Section / Labarge Composition Case
 with EOR Carbon Dioxide Recovery / CFZ-CNG Process

Figure B.1-3
TPC Section / LaBarge Composition Case
Detail of Heat Exchanger HX-1

Table B.1-2
TPC Section Material Balance
LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/24 Version 386/C2.10 Case Name TPCEXX10.SIM
 Time 17:38:39 Prop Pkg PRSV

TPC SECTION <> LABARGE COMPOSITION <> CFZ-CNG PROCESS

Stream	1	2	3	4
Vapour frac.	0.0000	0.0000	0.0453	0.0000
Temperature F	40.3679	-51.0770	-62.2315	-66.9663
Pressure psia	600.0000	580.0000	100.0000	100.0000
Molar Flow lbmole/hr	15842.3988	15842.3988	15842.3988	4449.9999
Methane mole frac.	0.010005	0.010005	0.010005	0.000000
CO2 mole frac.	0.920189	0.920189	0.920189	0.999996
H2S mole frac.	0.069188	0.069188	0.069188	0.000004
Nitrogen mole frac.	0.000038	0.000038	0.000038	0.000000
Ethane mole frac.	0.000139	0.000139	0.000139	0.000000
COS mole frac.	0.000442	0.000442	0.000442	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Stream	5	6	7	8
Vapour frac.	0.9998	1.0000	0.0000	1.0000
Temperature F	-55.2589	-60.4279	-55.7207	-50.5785
Pressure psia	100.0000	100.0000	100.0000	100.0000
Molar Flow lbmole/hr	1049.8400	3636.4016	18829.5704	182.0511
Methane mole frac.	0.000007	0.043590	0.000010	0.001044
CO2 mole frac.	0.928487	0.923014	0.923834	0.918380
H2S mole frac.	0.070396	0.000018	0.075661	0.075399
Nitrogen mole frac.	0.000211	0.017944	0.000004	0.000505
Ethane mole frac.	0.000839	0.000588	0.000094	0.004593
COS mole frac.	0.000039	0.000000	0.000396	0.000079
Hydrogen mole frac.	0.000021	0.014847	0.000000	0.000000
Stream	9	10	11	12
Vapour frac.	0.0000	0.0000	0.0000	1.0000
Temperature F	-53.5510	-55.2590	-45.8000	-54.1968
Pressure psia	100.0000	100.0000	105.0000	100.0000
Molar Flow lbmole/hr	10880.7398	2500.0000	728.8817	11481.3533
Methane mole frac.	0.000001	0.000000	0.000000	0.000001
CO2 mole frac.	0.791481	0.892639	0.316213	0.895930
H2S mole frac.	0.207515	0.107030	0.681447	0.103668
Nitrogen mole frac.	0.000000	0.000002	0.000330	0.000022
Ethane mole frac.	0.000238	0.000218	0.000000	0.000273
COS mole frac.	0.000765	0.000111	0.001980	0.000104
Hydrogen mole frac.	0.000000	0.000000	0.000031	0.000002
Stream	13	14	15	17
Vapour frac.	1.0000	0.0015	0.0915	0.0000
Temperature F	17.7259	-30.1030	-55.2589	-55.2589
Pressure psia	176.0000	171.0000	100.0000	100.0000
Molar Flow lbmole/hr	11481.3533	11481.3533	11481.3533	10431.3545
Methane mole frac.	0.000001	0.000001	0.000001	0.000000
CO2 mole frac.	0.895930	0.895930	0.895930	0.892653
H2S mole frac.	0.103668	0.103668	0.103668	0.107018
Nitrogen mole frac.	0.000022	0.000022	0.000022	0.000002
Ethane mole frac.	0.000273	0.000273	0.000273	0.000216
COS mole frac.	0.000104	0.000104	0.000104	0.000111
Hydrogen mole frac.	0.000002	0.000002	0.000002	0.000000

Table B.1-2 (continued)
TPC Section Material Balance
LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/24 Version 386|C2.10 Case Name TPCEXX10.SIM
 Time 17:38:39 Prop Pkg PRSV

TPC SECTION <> LABARGE COMPOSITION <> CFZ-CNG PROCESS

Stream	18	19	20	21
Vapour frac.	0.5638	1.0000	0.0000	1.0000
Temperature F	-45.8000	126.1200	-40.0929	14.5500
Pressure psia	105.0000	110.0000	100.0000	95.0000
Molar Flow lbmole/hr	1670.9100	1670.9100	2628.2675	2628.2675
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.486394	0.486394	0.299620	0.299620
H2S mole frac.	0.441496	0.441496	0.697010	0.697010
Nitrogen mole frac.	0.038691	0.038691	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000002	0.000002
COS mole frac.	0.001107	0.001107	0.003367	0.003367
Hydrogen mole frac.	0.032312	0.032312	0.000000	0.000000
Stream	23	24	26	28
Vapour frac.	0.0000	1.0000	1.0000	0.0000
Temperature F	-55.2590	-45.8000	-67.1554	-66.9662
Pressure psia	100.0000	105.0000	78.0000	100.0000
Molar Flow lbmole/hr	7941.7722	942.0283	2892.5361	12781.6238
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.892639	0.618069	0.999997	0.999996
H2S mole frac.	0.107030	0.255837	0.000003	0.000004
Nitrogen mole frac.	0.000002	0.068373	0.000000	0.000000
Ethane mole frac.	0.000218	0.000000	0.000000	0.000000
COS mole frac.	0.000111	0.000432	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.057289	0.000000	0.000000
Stream	29	31	32	33
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	-66.9662	-63.5385	20.0000	41.9694
Pressure psia	100.0000	500.0000	480.0000	2010.0000
Molar Flow lbmole/hr	8331.6237	8331.6237	8331.6237	8331.6237
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.999996	0.999996	0.999996	0.999996
H2S mole frac.	0.000004	0.000004	0.000004	0.000004
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Stream	34	35	36	37
Vapour frac.	1.0000	1.0000	1.0000	1.0000
Temperature F	-66.1900	14.0000	-78.8963	14.0000
Pressure psia	78.0000	73.0000	21.0000	16.0000
Molar Flow lbmole/hr	6528.9380	6528.9380	6528.9380	6528.9380
Methane mole frac.	0.024278	0.024278	0.024278	0.024278
CO2 mole frac.	0.957120	0.957120	0.957120	0.957120
H2S mole frac.	0.000011	0.000011	0.000011	0.000011
Nitrogen mole frac.	0.009994	0.009994	0.009994	0.009994
Ethane mole frac.	0.000328	0.000328	0.000328	0.000328
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.008269	0.008269	0.008269	0.008269

Table B.1-2 (continued)
TPC Section Material Balance
LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/24 Version 386/C2.10 Case Name TPCEXX10.SIM
 Time 17:38:39 Prop Pkg PRSV

TPC SECTION <> LABARGE COMPOSITION <> CFZ-CNG PROCESS

Stream	T5	T6	T7	T11
Vapour frac.	1.0000	0.0000	0.0000	0.0000
Temperature °	-81.1753	-68.7387	-81.0000	-58.0000
Pressure psia	51.0000	75.0000	78.0000	94.0000
Molar Flow lbmole/hr	12924.8152	26200.0003	55400.0005	12742.7645
Methane mole frac.	0.000047	0.000000	0.000001	0.000033
CO2 mole frac.	0.882156	0.998793	0.791480	0.881639
H2S mole frac.	0.116347	0.001200	0.207516	0.116932
Nitrogen mole frac.	0.000012	0.000000	0.000000	0.000005
Ethane mole frac.	0.001211	0.000002	0.000238	0.001163
COS mole frac.	0.000226	0.000005	0.000765	0.000228
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	T13	T15	T18	T21
Vapour frac.	1.0000	0.0000	0.0000	0.0000
Temperature F	-69.8000	-80.7330	-70.0000	-68.7526
Pressure psia	73.0000	105.0000	78.0000	75.0000
Molar Flow lbmole/hr	1050.0000	10880.7398	55400.0005	26200.0003
Methane mole frac.	0.000000	0.000001	0.000000	0.000000
CO2 mole frac.	0.996984	0.791481	0.995969	0.999996
H2S mole frac.	0.003000	0.207515	0.004000	0.000004
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000004	0.000238	0.000001	0.000000
COS mole frac.	0.000012	0.000765	0.000030	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	T25	T26	2A	2B
Vapour frac.	1.0000	1.0000	0.0000	0.0000
Temperature F	-67.1530	-62.0751	-23.5247	-51.0000
Pressure psia	78.0000	78.0000	590.0000	580.0000
Molar Flow lbmole/hr	10499.9996	1050.0000	8682.7859	2564.9154
Methane mole frac.	0.000000	0.000000	0.010005	0.010005
CO2 mole frac.	0.999997	0.996984	0.920189	0.920189
H2S mole frac.	0.000003	0.003000	0.069188	0.069188
Nitrogen mole frac.	0.000000	0.000000	0.000038	0.000038
Ethane mole frac.	0.000000	0.000004	0.000139	0.000139
COS mole frac.	0.000000	0.000012	0.000442	0.000442
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	2C	2D	2AA	31A
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	-40.0000	-45.0000	-55.6437	-54.9000
Pressure psia	590.0000	590.0000	580.0000	490.0000
Molar Flow lbmole/hr	2247.5540	2348.1436	8682.7859	8331.6237
Methane mole frac.	0.010005	0.010005	0.010005	0.000000
CO2 mole frac.	0.920189	0.920189	0.920189	0.999996
H2S mole frac.	0.069188	0.069188	0.069188	0.000004
Nitrogen mole frac.	0.000038	0.000038	0.000038	0.000000
Ethane mole frac.	0.000139	0.000139	0.000139	0.000000
COS mole frac.	0.000442	0.000442	0.000442	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Figure B.1-4
High Pressure Claus Section
Labarge Composition Case

Table B.1-3
HPSRU Section Material Balance
LaBarge Composition Case

SRU SECTION - LABARGE COMPOSITION

Stream		Acid Gas	O2/ Air	RF Out	Cond Out
Stream		1	2	3	4
Temperature	F	50	50	2006	300
Pressure	Psia	95	95	94	93
Molar Flow	Lbmole/hr	2630	595	3300	2750
Methane	mol frac.	0	0	0	0
CO2	mol frac.	.300	0	.208	.249
H2S	mol frac.	.697	0	.205	.246
Nitrogen	mol frac.	0	.1	.018	.022
COS	mol frac.	.003	0	.001	.002
Hydrogen	mol frac.	0	0	.027	.032
H2O	mol frac.	0	0	.322	.387
SO2	mol frac.	0	0	.017	.019
CO	mol frac.	0	0	.035	.042
Oxygen	mol frac.	0	.9	0	0
S2 (vapor)	mol frac.	0	0	.167	.0002
S1 (liquid)	mol frac.	0	0	0	0

Stream		Sulfur Product	RGG Feed	RGG CH4	RGG O2
Stream		5	6	7	8
Temperature	F	300	432	50	50
Pressure	Psia	93	92	92	92
Molar Flow	Lbmole/hr	1100	2750	12.5	14
Methane	mol frac.	0	0	1	0
CO2	mol frac.	0	.249	0	0
H2S	mol frac.	0	.246	0	0
Nitrogen	mol frac.	0	.022	0	.1
COS	mol frac.	0	.0	0	0
Hydrogen	mol frac.	0	.	0	0
H2O	mol frac.	0	.	0	0
SO2	mol frac.	0	.0	0	0
CO	mol frac.	0	.042	0	0
Oxygen	mol frac.	0	0	0	.9
S2 (vapor)	mol frac.	0	.0002	0	0
S1 (liquid)	mol frac.	1	0	0	0

Stream		Hyd in	Hyd out	WHE out	Quench out
Stream		9	10	11	12
Temperature	F	500	775	275	69
Pressure	Psia	92	91	90	89
Molar Flow	Lbmole/hr	2790	2730	2730	1970
Methane	mol frac.	0	0	0	0
CO2	mol frac.	.250	.298	.298	.485
H2S	mol frac.	.243	.270	.270	.440
Nitrogen	mol frac.	.022	.022	.022	.035
COS	mol frac.	.002	.001	.001	.001
Hydrogen	mol frac.	.040	.020	.020	.032
H2O	mol frac.	.381	.388	.388	.004
SO2	mol frac.	.020	0	0	0
CO	mol frac.	.041	.001	.001	.002
Oxygen	mol frac.	0	0	0	0
S2 (vapor)	mol frac.	.0002	0	0	0
S1 (liquid)	mol frac.	0	0	0	0

Table B.1-3 (continued)
HPSRU Section Material Balance
LaBarge Composition Case

SRU SECTION - LABARGE COMPOSITION

		DeHi	Comp	TPC sect.	Regen
		Out	out	feed	Gas
Stream		13	14	15	16
Temperature F		85	150	126	126
Pressure	Psia	85	115	115	115
Molar Flow	Lbmole/hr	1960	1960	1670	290
Methane	mol frac.	0	0	0	0
CO2	mol frac.	.487	.487	.487	.487
H2S	mol frac.	.442	.442	.442	.442
Nitrogen	mol frac.	.035	.035	.035	.035
COS	mol frac.	.001	.001	.001	.001
Hydrogen	mol frac.	.032	.032	.032	.032
H2O	mol frac.	0	0	0	0
SO2	mol frac.	0	0	0	0
CO	mol frac.	.002	.002	.002	.002
Oxygen	mol frac.	0	0	0	0
S2 (vapor)	mol frac.	0	0	0	0
S1 (liquid)	mol frac.	0	0	0	0
		HP	HP	LP	HP
		steam	steam	steam	steam
Stream		17	18	19	20
Temperature F		445	445	287	445
Pressure	Psia	400	400	55	400
Molar Flow	Lbmole/hr	3070	230	640	230
H2O	mol frac.	1	1	1	1
		HP	Quench	Quench	Quench
		steam	Water	Water	Water
Stream		21	22	23	24
Temperature F		445	247	100	65
Pressure	Psia	400	89	90	90
Molar Flow	Lbmole/hr	60	8000	6000	2000
CO2	mol frac.	0	4×10^{-4}	4×10^{-4}	4×10^{-4}
H2S	mol frac.	0	1×10^{-3}	1×10^{-3}	1×10^{-3}
H2O	mol frac.	1	.9986	.9986	.9986
		Byprod.	EOR	EOR	
		Water	CO2	CO2	
Stream		25	26	27	
Temperature F		100	42	48	
Pressure	Psia	89	2010	2000	
Molar Flow	Lbmole/hr	1060	8330	8330	
CO2	mol frac.	0	1	1	
H2S	mol frac.	10^{-6}	4×10^{-6}	4×10^{-6}	
H2O	mol frac.	1	0	0	

Figure B.2-1

CFZ Section Process Flow Diagram Generic Composition Case

Table B.2-1
CFZ Section Material Balance
Generic Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/25 Version 386/C2.10 Case Name ACR_GRIW.SIM
 Time 9:56:24 Prop Pkg PR

CFZ Section <> Generic Composition

Stream	1	2	3	4
Vapour frac.	0.9997	1.0000	1.0000	1.0000
Temperature F	90.0000	90.0000	-53.4000	-61.2024
Pressure psia	715.0000	695.0001	665.0000	600.0000
Molar Flow lbmole/hr	21995.6408	21960.9421	21960.9421	21960.9421
Methane mole frac.	0.803724	0.804994	0.804994	0.804994
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.189699	0.189998	0.189998	0.189998
H2S mole frac.	0.005000	0.005008	0.005008	0.005008
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.001578	0.000000	0.000000	0.000000

Stream	5	6	7	8
Vapour frac.	0.0000	1.0000	1.0000	0.0000
Temperature F	-122.8565	-122.8567	40.9521	40.9522
Pressure psia	600.0000	600.0000	605.0000	605.0000
Molar Flow lbmole/hr	47705.9293	17675.8154	4074.8220	4285.1280
Methane mole frac.	0.996042	0.997731	0.044565	0.009999
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.003949	0.002265	0.934523	0.964392
H2S mole frac.	0.000009	0.000004	0.020912	0.025608
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	9	10	11	12
Vapour frac.	1.0000	1.0000	1.0000	1.0000
Temperature F	-81.8590	60.0000	111.9660	100.0000
Pressure psia	590.0000	570.0000	268.9474	263.9474
Molar Flow lbmole/hr	17675.8154	17675.8154	9285.2230	9285.2230
Methane mole frac.	0.997731	0.997731	0.001000	0.001000
Ethane mole frac.	0.000000	0.000000	0.004000	0.004000
CO2 mole frac.	0.002265	0.002265	0.000000	0.000000
H2S mole frac.	0.000004	0.000004	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.995000	0.995000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.000000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Table B.2-1 (continued)
CFZ Section Material Balance
Generic Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acirion Technologies Inc
 Date 93/05/24 Version 386/C2.10 Case Name ACR_GRIW.SIM
 Time 16:06:23 Prop Pkg PR

CFZ Section <> Generic Composition

Stream	13	14	15	16
Vapour frac.	0.0000	0.0000	0.9999	1.0000
Temperature F	-112.8567	-130.0000	-129.6380	140.0622
Pressure psia	243.9474	31.3907	31.3907	194.0000
Molar Flow lbmole/hr	9285.2230	8798.7298	9285.2230	6996.0920
Methane mole frac.	0.001000	0.000550	0.001000	0.000000
Ethane mole frac.	0.004000	0.004101	0.004000	0.001000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.995000	0.995349	0.995000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.995000
i-Butane mole frac.	0.000000	0.000000	0.000000	0.004000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	17	18	19	20
Vapour frac.	0.1768	0.0000	0.0000	0.0000
Temperature F	100.0871	52.0018	40.0000	40.0000
Pressure psia	189.0000	184.0000	78.4889	78.4889
Molar Flow lbmole/hr	6996.0920	6996.0920	505.9560	770.6150
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.001000	0.001000	0.000883	0.000883
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.995000	0.995000	0.994998	0.994998
i-Butane mole frac.	0.004000	0.004000	0.004119	0.004119
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	21	22	23	24
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	40.0000	-10.0000	-10.0000	-10.0000
Pressure psia	78.4889	31.3333	31.3333	31.3333
Molar Flow lbmole/hr	5390.0505	604.2120	707.5570	3173.0220
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000883	0.000519	0.000519	0.000519
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.994998	0.994830	0.994830	0.994830
i-Butane mole frac.	0.004119	0.004651	0.004651	0.004651
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Table B.2-1 (continued)
CFZ Section Material Balance
Generic Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/24 Version 386/C2.10 Case Name ACR_GRIW.SIM
 Time 16:16:26 Prop Pkg PR

CFZ Section <> Generic Composition

Stream	25	26	27	28
Vapour frac.	0.0000	0.9999	1.0000	1.0000
Temperature F	-40.0000	-39.5707	0.9681	64.0115
Pressure psia	16.1298	16.1298	31.3333	78.4889
Molar Flow lbmole/hr	2889.3215	3173.0220	5390.0505	6996.0920
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000351	0.000519	0.000883	0.001000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
Ethylene mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.994677	0.994830	0.994998	0.995000
i-Butane mole frac.	0.004972	0.004651	0.004119	0.004000
H2O mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	29	30
Vapour frac.	0.1546	0.9150
Temperature F	-51.1073	-41.3263
Pressure psia	100.0000	100.0000
Molar Flow lbmole/hr	4010.4201	4010.4201
Methane mole frac.	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000
CO2 mole frac.	0.646108	0.646108
H2S mole frac.	0.353892	0.353892
Ethylene mole frac.	0.000000	0.000000
Propane mole frac.	0.000000	0.000000
i-Butane mole frac.	0.000000	0.000000
H2O mole frac.	0.000000	0.000000

Figure B.2-2
TPC Section Process Flow Diagram
Generic Composition Case

Table B.2-2
TPC Section Material Balance
Generic Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/03/25 Version 386|C2.10 Case Name LQNG5X10.SIM
 Time 17:15:17 Prop Pkg PRSV

TPC Section <> Generic Composition

Stream	1	2	3	4
Vapour frac.	0.0000	0.0000	0.1138	0.0000
Temperature F	39.2261	-30.0000	-59.7417	-68.75403
Pressure psia	600.0000	590.0000	100.0000	100.0000
Molar Flow lbmole/hr	4291.9000	4291.9000	4291.9000	3394.5471
Methane mole frac.	0.010000	0.010000	0.010000	0.000000
CO2 mole frac.	0.964392	0.964392	0.964392	0.999998
H2S mole frac.	0.025608	0.025608	0.025608	0.000002
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	5	6	8	9
Vapour frac.	1.0000	0.0000	0.0000	0.0000
Temperature F	-57.4297	-40.1231	-56.1150	-76.0031
Pressure psia	100.0000	100.0000	100.0000	105.0000
Molar Flow lbmole/hr	3459.9203	261.5740	7380.9034	3255.2695
Methane mole frac.	0.012405	0.000000	0.000000	0.000000
CO2 mole frac.	0.984255	0.299340	0.954741	0.897436
H2S mole frac.	0.000018	0.700022	0.045257	0.102561
Nitrogen mole frac.	0.001791	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000638	0.000001	0.000002
Hydrogen mole frac.	0.001532	0.000000	0.000000	0.000000

Stream	11	17	18	19
Vapour frac.	0.0000	1.0000	1.0000	1.0000
Temperature F	-63.2051	-64.3120	11.9672	-77.5843
Pressure psia	100.0000	75.0000	70.0000	21.0000
Molar Flow lbmole/hr	3255.2695	4190.5913	4190.5913	4190.5913
Methane mole frac.	0.000000	0.010242	0.010242	0.010242
CO2 mole frac.	0.897436	0.987000	0.987000	0.987000
H2S mole frac.	0.102561	0.000015	0.000015	0.000015
Nitrogen mole frac.	0.000000	0.001479	0.001479	0.001479
COS mole frac.	0.000002	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.001264	0.001264	0.001264

Stream	20	21	22	24
Vapour frac.	1.0000	0.2527	1.0000	1.0000
Temperature F	11.9672	-60.0000	-11.3582	-1.3582
Pressure psia	16.0000	105.0000	97.5000	110.0000
Molar Flow lbmole/hr	4190.5913	162.7000	261.5740	162.7000
Methane mole frac.	0.010242	0.000000	0.000000	0.000000
CO2 mole frac.	0.987000	0.482999	0.299340	0.482999
H2S mole frac.	0.000015	0.445309	0.700022	0.445309
Nitrogen mole frac.	0.001479	0.038099	0.000000	0.038099
COS mole frac.	0.000000	0.001024	0.000638	0.001024
Hydrogen mole frac.	0.001264	0.032569	0.000000	0.032569

Table B.2-2 (continued)
TPC Section Material Balance
Generic Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc

Date 93/05/25 Version 386/C2.10

Case Name LQNG5X10.SIM

Time 11:48:24

Prop Pkg PRSV

TPC Section <> Generic Composition

Stream	25	26	27	29
Vapour frac.	1.0000	0.0000	1.0000	1.0000
Temperature F	137.5891	-60.0000	-60.0000	85.0000
Pressure psia	115.0000	105.0000	105.0000	85.0000
Molar Flow lbmole/hr	162.7000	121.5889	41.1112	162.7000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.482999	0.457534	0.558313	0.482999
H2S mole frac.	0.445309	0.540162	0.164774	0.445309
Nitrogen mole frac.	0.038099	0.000904	0.148107	0.038099
COS mole frac.	0.001024	0.001321	0.000144	0.001024
Hydrogen mole frac.	0.032569	0.000079	0.128661	0.032569

Stream	30	T5	T6	T7
Vapour frac.	1.0000	1.0000	1.0000	0.0000
Temperature F	74.6037	-76.3764	-8.0558	-57.0000
Pressure psia	95.0000	60.0000	102.7832	97.7832
Molar Flow lbmole/hr	261.5740	3339.8872	3339.8872	3339.8872
Methane mole frac.	0.000000	0.000023	0.000023	0.000023
CO2 mole frac.	0.299340	0.932799	0.932799	0.932799
H2S mole frac.	0.700022	0.067139	0.067139	0.067139
Nitrogen mole frac.	0.000000	0.000038	0.000038	0.000038
COS mole frac.	0.000638	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000001	0.000001	0.000001

Stream	T8	T12	T13	T17
Vapour frac.	0.0000	1.0000	0.0000	1.0000
Temperature F	-76.3764	-68.7511	-68.7448	-69.8214
Pressure psia	78.0000	75.0000	75.0000	73.0000
Molar Flow lbmole/hr	14550.0005	730.6709	6889.0004	2764.0000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.897437	0.999998	0.999400	0.998500
H2S mole frac.	0.102561	0.000002	0.000600	0.001500
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000002	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	T18	T22	T25	T26
Vapour frac.	1.0000	0.0000	1.0000	0.0000
Temperature F	-61.6776	-68.7526	-67.1530	-67.1340
Pressure psia	78.0000	75.0000	78.0000	78.0000
Molar Flow lbmole/hr	2764.0000	6889.0004	2764.0000	14550.0005
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.998500	0.999998	0.999998	0.998000
H2S mole frac.	0.001500	0.000002	0.000002	0.002000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Figure B.2-3
High Pressure Claus Section
Generic Composition Case

Table B.2-3
HPSRU Section Material Balance
Generic Composition Case

SRU SECTION - GENERIC COMPOSITION

		Acid Gas	O2/ Air	RF Out	Cond Out
Stream		1	2	3	4
Temperature	F	75	50	2006	300
Pressure	Psia	95	95	94	93
Molar Flow	Lbmole/hr	262	60	328	275
Methane	mol frac.	0	0	0	0
CO2	mol frac.	.300	0	.208	.249
H2S	mol frac.	.699	0	.205	.246
Nitrogen	mol frac.	0	.1	.018	.022
COS	mol frac.	.001	0	.001	.002
Hydrogen	mol frac.	0	0	.027	.032
H2O	mol frac.	0	0	.322	.387
SO2	mol frac.	0	0	.017	.019
CO	mol frac.	0	0	.035	.042
Oxygen	mol frac.	0	.9	0	0
S2 (vapor)	mol frac.	0	0	.167	.0002
S1 (liquid)	mol frac.	0	0	0	0

		Sulfur Product	RGG Feed	RGG CH4	RGG O2
Stream		5	6	7	8
Temperature	F	300	432	50	50
Pressure	Psia	93	92	92	92
Molar Flow	Lbmole/hr	109	275	1.3	1.4
Methane	mol frac.	0	0	1	0
CO2	mol frac.	0	.249	0	0
H2S	mol frac.	0	.246	0	0
Nitrogen	mol frac.	0	.022	0	.1
COS	mol frac.	0	.002	0	0
Hydrogen	mol frac.	0	.032	0	0
H2O	mol frac.	0	.387	0	0
SO2	mol frac.	0	.020	0	0
CO	mol frac.	0	.042	0	0
Oxygen	mol frac.	0	0	0	.9
S2 (vapor)	mol frac.	0	.0002	0	0
S1 (liquid)	mol frac.	1	0	0	0

		Hyd in	Hyd out	WHE out	Quench out
Stream		9	10	11	12
Temperature	F	500	775	275	69
Pressure	Psia	92	91	90	89
Molar Flow	Lbmole/hr	278	272	272	195
Methane	mol frac.	0	0	0	0
CO2	mol frac.	.250	.298	.298	.485
H2S	mol frac.	.243	.270	.270	.440
Nitrogen	mol frac.	.022	.022	.022	.035
COS	mol frac.	.002	.001	.001	.001
Hydrogen	mol frac.	.040	.020	.020	.032
H2O	mol frac.	.381	.388	.388	.004
SO2	mol frac.	.020	0	0	0
CO	mol frac.	.041	.001	.001	.002
Oxygen	mol frac.	0	0	0	0
S2 (vapor)	mol frac.	.0002	0	0	0
S1 (liquid)	mol frac.	0	0	0	0

Table B.2-3 (continued)
HPSRU Section Material Balance
Generic Composition Case

		DeHi	Comp	TPC sect.	Regen
		Out	out	feed	Gas
Stream		13	14	15	16
Temperature	F	85	150	126	126
Pressure	Psia	85	115	115	115
Molar Flow	Lbmole/hr	195	195	165	30
Methane	mol frac.	0	0	0	0
CO2	mol frac.	.487	.487	.487	.487
H2S	mol frac.	.442	.442	.442	.442
Nitrogen	mol frac.	.035	.035	.035	.035
COS	mol frac.	.001	.001	.001	.001
Hydrogen	mol frac.	.032	.032	.032	.032
H2O	mol frac.	0	0	0	0
SO2	mol frac.	0	0	0	0
CO	mol frac.	.002	.002	.002	.002
Oxygen	mol frac.	0	0	0	0
S2 (vapor)	mol frac.	0	0	0	0
S1 (liquid)	mol frac.	0	0	0	0

		HP	HP	LP	HP
		steam	steam	steam	steam
Stream		17	18	19	20
Temperature	F	445	445	287	445
Pressure	Psia	400	400	55	400
Molar Flow	Lbmole/hr	305	23	63	23
H2O	mol frac.	1	1	1	1

		HP	Quench	Quench	Quench
		steam	Water	Water	Water
Stream		21	22	23	24
Temperature	F	445	247	100	65
Pressure	Psia	400	89	90	90
Molar Flow	Lbmole/hr	7	795	596	199
CO2	mol frac.	0	4×10^{-4}	4×10^{-4}	4×10^{-4}
H2S	mol frac.	0	1×10^{-3}	1×10^{-3}	1×10^{-3}
H2O	mol frac.	1	.9986	.9986	.9986

		Byprod.	CFZ	CFZ
		Water	BTTMS.	BTTMS.
Stream		25	26	27
Temperature	F	100	41	41
Pressure	Psia	89	600	590
Molar Flow	Lbmole/hr	106	4285	4285
Methane	mol frac.	0	.010	.010
CO2	mol frac.	0	.965	.965
H2S	mol frac.	10^{-6}	.025	.025
Nitrogen	mol frac.	0	0	0
COS	mol frac.	0	0	0
Hydrogen	mol frac.	0	0	0
H2O	mol frac.	1	0	0
SO2	mol frac.	0	0	0
CO	mol frac.	0	0	0
Oxygen	mol frac.	0	0	0
S2 (vapor)	mol frac.	0	0	0
S1 (liquid)	mol frac.	0	0	0

Figure B.3-1

Table B.3-1
H₂S and Bulk CO₂ Removal Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrlon Technologies Inc
 Date 93/05/26 Version 386|C2.10 Case Name CNG1Q10.SIM
 Time 16:35:00 Prop Pkg PRSV

H₂S + BULK CO₂ REMOVAL SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	1	2	3	4
Vapour frac.	1.0000	0.1542	1.0000	0.0000
Temperature F	60.0000	-30.0000	-67.0044	-64.5263
Pressure psia	1040.0000	1000.0000	700.0000	700.0000
Molar Flow lbmole/hr	21959.9991	21959.9991	7615.0092	2410.0000
Helium mole frac.	0.005999	0.005999	0.017301	0.000000
Methane mole frac.	0.204979	0.204979	0.590186	0.000000
Nitrogen mole frac.	0.073493	0.073493	0.211939	0.000000
CO ₂ mole frac.	0.664934	0.664934	0.180494	0.999996
H ₂ S mole frac.	0.049995	0.049995	0.000002	0.000004
COS mole frac.	0.000500	0.000500	0.000000	0.000000
Ethane mole frac.	0.000100	0.000100	0.000078	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	5	6	7	8
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	47.6370	57.6347	93.5784	-38.0789
Pressure psia	700.0000	700.0000	700.0000	695.0001
Molar Flow lbmole/hr	18487.4311	33270.8393	1971.4095	33270.8393
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000011	0.000219	0.000000	0.000219
Nitrogen mole frac.	0.000002	0.000001	0.000000	0.000001
CO ₂ mole frac.	0.772604	0.882098	0.290510	0.882098
H ₂ S mole frac.	0.226181	0.116983	0.703654	0.116983
COS mole frac.	0.001101	0.000596	0.005827	0.000596
Ethane mole frac.	0.000101	0.000104	0.000008	0.000104
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	9	10	11	12
Vapour frac.	0.0000	1.0000	1.0000	0.0150
Temperature F	-61.3576	-81.3673	9.7170	-58.0000
Pressure psia	95.0000	50.0000	102.1408	97.1408
Molar Flow lbmole/hr	16274.7221	15301.7551	15301.7551	15301.7551
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000160	0.000630	0.000630	0.000630
Nitrogen mole frac.	0.000461	0.000489	0.000489	0.000489
CO ₂ mole frac.	0.864690	0.874535	0.874535	0.874535
H ₂ S mole frac.	0.133930	0.123467	0.123467	0.123467
COS mole frac.	0.000329	0.000316	0.000316	0.000316
Ethane mole frac.	0.000384	0.000512	0.000512	0.000512
Hydrogen mole frac.	0.000047	0.000050	0.000050	0.000050

Table B.3-1 (continued)
H₂S and Bulk CO₂ Removal Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acirion Technologies Inc
 Date 93/05/26 Version 386/C2.10 Case Name CNGLQ10.SIM
 Time 16:35:00 Prop Pkg PRSV

H₂S + BULK CO₂ REMOVAL SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	13	14	15	16
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	-81.6673	-81.6673	-76.2866	-67.1425
Pressure psia	78.0000	50.0000	705.0000	78.0000
Molar Flow lbmole/hr	55623.5478	18490.8887	18490.8887	26328.4786
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000011	0.000011	0.000011	0.000000
Nitrogen mole frac.	0.000002	0.000002	0.000002	0.000000
CO ₂ mole frac.	0.772544	0.772543	0.772543	0.998793
H ₂ S mole frac.	0.226241	0.226241	0.226241	0.001200
COS mole frac.	0.001101	0.001101	0.001101	0.000006
Ethane mole frac.	0.000101	0.000101	0.000101	0.000001
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Stream	17	18	19	20
Vapour frac.	1.0000	1.0000	0.0000	0.0000
Temperature F	-68.7084	-63.8873	-68.7000	-67.1559
Pressure psia	75.0000	78.0000	78.0000	78.0000
Molar Flow lbmole/hr	10568.4746	10568.4746	55623.5478	26328.4786
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO ₂ mole frac.	0.996984	0.996984	0.995963	0.999996
H ₂ S mole frac.	0.003000	0.003000	0.004000	0.000004
COS mole frac.	0.000015	0.000015	0.000037	0.000000
Ethane mole frac.	0.000001	0.000001	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Stream	21	22	23	24
Vapour frac.	0.6250	1.0000	1.0000	1.0000
Temperature F	-67.1579	-67.1530	-67.1530	-69.8435
Pressure psia	78.0000	78.0000	78.0000	73.0000
Molar Flow lbmole/hr	26328.4786	16453.9998	10568.4746	5886.3595
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO ₂ mole frac.	0.999996	0.999996	0.999996	0.999997
H ₂ S mole frac.	0.000004	0.000004	0.000004	0.000003
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Table B.3-1 (continued)
H₂S and Bulk CO₂ Removal Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/26 Version 386/C2.10 Case Name CNGLO10.SIM
 Time 16:35:00 Prop Pkg PRSV

H₂S + BULK CO₂ REMOVAL SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	25	26	27	28
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	-69.8435	-64.5265	-64.5265	47.6347
Pressure psia	73.0000	700.0000	700.0000	695.0001
Molar Flow lbmole/hr	9873.6402	9873.6402	6058.6402	6058.6402
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO ₂ mole frac.	0.999996	0.999996	0.999996	0.999996
H ₂ S mole frac.	0.000004	0.000004	0.000004	0.000004
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	29	30	31	32
Vapour frac.	0.0000	0.0000	1.0000	1.0000
Temperature F	72.6121	-64.5265	85.0000	204.5693
Pressure psia	2000.0001	700.0000	85.0000	165.5475
Molar Flow lbmole/hr	6058.6402	1405.0000	1670.9100	1670.9100
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.038691	0.038691
CO ₂ mole frac.	0.999996	0.999996	0.486400	0.486400
H ₂ S mole frac.	0.000004	0.000004	0.441501	0.441501
COS mole frac.	0.000000	0.000000	0.001107	0.001107
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.032300	0.032300

Stream	33	34	35	36
Vapour frac.	1.0000	0.6100	0.0000	1.0000
Temperature F	100.0000	-25.0000	-25.0000	-25.0000
Pressure psia	160.5475	155.5475	155.5475	155.5475
Molar Flow lbmole/hr	1670.9100	1670.9100	651.6549	1019.2551
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.038691	0.038691	0.000500	0.063109
CO ₂ mole frac.	0.486400	0.486400	0.311143	0.598449
H ₂ S mole frac.	0.441501	0.441501	0.686242	0.285028
COS mole frac.	0.001107	0.001107	0.002057	0.000500
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.032300	0.032300	0.000058	0.052914

Table B.3-1 (continued)
H₂S and Bulk CO₂ Removal Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/05/26 Version 386/C2.10 Case Name CNGLQ10.SIM
 Time 16:35:00 Prop Pkg PRSV

H₂S + BULK CO₂ REMOVAL SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	37	38	39	40
Vapour frac.	1.0000	1.0000	1.0000	1.0000
Temperature F	-63.1365	-69.4979	47.6347	-47.8845
Pressure psia	95.0000	73.0000	68.0000	20.0000
Molar Flow lbmole/hr	1451.2886	7337.6483	7337.6483	7337.6483
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.004859	0.000961	0.000961	0.000961
Nitrogen mole frac.	0.044305	0.008763	0.008763	0.008763
CO ₂ mole frac.	0.912498	0.982691	0.982691	0.982691
H ₂ S mole frac.	0.000079	0.000018	0.000018	0.000018
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.001092	0.000216	0.000216	0.000216
Hydrogen mole frac.	0.037167	0.007351	0.007351	0.007351

Stream	41	42
Vapour frac.	1.0000	1.0000
Temperature F	47.6347	-12.2548
Pressure psia	15.0000	95.0000
Molar Flow lbmole/hr	7337.6483	2623.0644
Helium mole frac.	0.000000	0.000000
Methane mole frac.	0.000961	0.000000
Nitrogen mole frac.	0.008763	0.000124
CO ₂ mole frac.	0.982691	0.295636
H ₂ S mole frac.	0.000018	0.699329
COS mole frac.	0.000000	0.004891
Ethane mole frac.	0.000216	0.000006
Hydrogen mole frac.	0.007351	0.000015

Figure B.3-2
Final CO₂ Removal Section
CNG Claus Process / LaBarge Composition Case

Table B.3-2
Final CO₂ Removal Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acirion Technologies Inc
 Date 93/05/26 Version 386/C2.10 Case Name DWGCNG4.SIM
 Time 18:02:9 Prop Pkg SRK-Zero_HC_K

FINAL CO₂ REMOVAL SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	201	203	204	205
Vapour frac.	1.0000	0.0000	0.0237	0.0000
Temperature F	-79.4691	-39.9982	-38.8682	-38.8682
Pressure psia	680.0000	700.0000	220.0000	220.0000
Molar Flow lbmole/hr	6032.8104	7732.1996	7732.1996	7553.0425
Methane mole frac.	0.712283	0.025709	0.025709	0.010022
Nitrogen mole frac.	0.265883	0.001347	0.001347	0.000100
Methanol mole frac.	0.000011	0.795139	0.795139	0.813997
CO ₂ mole frac.	0.000032	0.177766	0.177766	0.175880
Helium mole frac.	0.021791	0.000038	0.000038	0.000001

Stream	206	207	208	209
Vapour frac.	1.0000	0.1053	1.0000	1.0000
Temperature F	-38.8682	-61.6404	-20.0000	-71.7138
Pressure psia	220.0000	20.0000	20.0000	18.0000
Molar Flow lbmole/hr	179.1573	7553.0425	1349.9999	2753.3481
Methane mole frac.	0.687084	0.010022	0.199998	0.125079
Nitrogen mole frac.	0.053934	0.000100	0.799992	0.392376
Methanol mole frac.	0.000093	0.813997	0.000000	0.000105
CO ₂ mole frac.	0.257264	0.175880	0.000010	0.482438
Helium mole frac.	0.001624	0.000001	0.000000	0.000002

Stream	210	211	212	213
Vapour frac.	0.0000	0.0000	1.0000	1.0000
Temperature F	-80.8077	-78.8515	43.0000	45.0000
Pressure psia	20.0000	690.0000	215.0000	15.0000
Molar Flow lbmole/hr	6149.6946	6149.6946	179.1573	2753.3481
Methane mole frac.	0.000213	0.000213	0.687084	0.125079
Nitrogen mole frac.	0.000065	0.000065	0.053934	0.392376
Methanol mole frac.	0.999702	0.999702	0.000093	0.000105
CO ₂ mole frac.	0.000020	0.000020	0.257264	0.482438
Helium mole frac.	0.000000	0.000000	0.001624	0.000002

Stream	214	215	216
Vapour frac.	1.0000	1.0000	1.0000
Temperature F	-40.3600	-24.8400	-10.0000
Pressure psia	675.0000	670.0000	665.0000
Molar Flow lbmole/hr	6032.8104	6032.8104	6032.8104
Methane mole frac.	0.712283	0.712283	0.712283
Nitrogen mole frac.	0.265883	0.265883	0.265883
Methanol mole frac.	0.000011	0.000011	0.000011
CO ₂ mole frac.	0.000032	0.000032	0.000032
Helium mole frac.	0.021791	0.021791	0.021791

Figure B.3-3
Ethane + Propane Refrigeration Section
CNG Claus Process / Labarge Composition Case

Table B.3-3
Ethane + Propane Refrigeration Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/04/27 Version 386/C2.10 Case Name CNGPRINT.SIM
 Time 16:02:44 Prop Pkg PRSV

C2 + C3 REFRIGERATION SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	101	102	103	104
Vapour frac.	1.0000	1.0000	1.0000	1.0000
Temperature F	-39.9992	23.2635	9.3994	0.0018
Pressure psia	16.0632	38.3805	38.3805	38.3805
Molar Flow lbmole/hr	5952.9841	5952.9841	14871.7074	6278.6362
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	1.000000	1.000000	1.000000	1.000000

Stream	105	106	107	108
Vapour frac.	1.0000	1.0000	1.0000	1.0000
Temperature F	-0.0000	77.7008	71.6583	50.0015
Pressure psia	38.3805	92.7353	92.7353	92.7353
Molar Flow lbmole/hr	2640.0868	14871.7074	19073.6567	4201.9490
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	1.000000	1.000000	1.000000	1.000000

Stream	109	110	111	112
Vapour frac.	1.0000	1.0000	0.0000	0.0000
Temperature F	137.3237	120.3898	100.0000	50.0000
Pressure psia	200.2607	195.2607	190.2607	92.7353
Molar Flow lbmole/hr	19073.6567	19073.6567	19073.6567	14871.7084
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	1.000000	1.000000	1.000000	1.000000

Table B.3-3 (continued)
Ethane + Propane Refrigeration Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acirion Technologies Inc
 Date 93/04/27 Version 386/C2.10 Case Name CNGPRINT.SIM
 Time 16:02:44 Prop Pkg PRSV

C2 + C3 REFRIGERATION SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	113	114	115	116
Vapour frac.	0.0000	0.0000	0.0000	0.0000
Temperature F	-0.0000	-0.0000	-0.0000	-5.6939
Pressure psia	38.3805	38.3805	38.3805	34.3805
Molar Flow lbmole/hr	12231.6214	5952.9851	6278.6362	5952.9851
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	1.000000	1.000000	1.000000	1.000000

Stream	117	118	119	120
Vapour frac.	1.0000	0.0000	0.0000	0.0000
Temperature F	-40.0000	-40.0000	-40.0000	-40.0000
Pressure psia	16.0632	16.0632	16.0632	16.0632
Molar Flow lbmole/hr	622.9912	5329.9940	3160.4096	2169.5842
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	0.000000	0.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	1.000000	1.000000	1.000000	1.000000

Stream	121	122	123	124
Vapour frac.	1.0000	1.0000	1.0000	0.0000
Temperature F	-39.9981	-40.0000	17.8307	-30.0000
Pressure psia	16.0632	16.0632	140.2398	135.2398
Molar Flow lbmole/hr	2169.5842	3160.4096	3032.5271	3032.5271
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	0.000000	0.000000	1.000000	1.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	1.000000	1.000000	0.000000	0.000000

Table B.3-3 (continued)
Ethane + Propane Refrigeration Section Material Balance
CNG Claus Process / LaBarge Composition Case

Hyprotech's Process Simulator HYSIM - Licensed to Acrion Technologies Inc
 Date 93/04/27 Version 386/C2.10 Case Name CNGPRINT.SIM
 Time 16:02:44 Prop Pkg PRSV

C2 + C3 REFRIGERATION SECTION <> LABARGE COMPOSITION <> CNG PROCESS

Stream	125	126	127	128
Vapour frac.	0.0000	1.0000	0.0000	1.0000
Temperature F	-70.0000	-77.0000	-77.0000	-76.9967
Pressure psia	130.2398	53.8125	53.8125	53.8125
Molar Flow lbmole/hr	3032.5271	71.0615	2961.4657	2961.4657
Helium mole frac.	0.000000	0.000000	0.000000	0.000000
Methane mole frac.	0.000000	0.000000	0.000000	0.000000
Nitrogen mole frac.	0.000000	0.000000	0.000000	0.000000
CO2 mole frac.	0.000000	0.000000	0.000000	0.000000
H2S mole frac.	0.000000	0.000000	0.000000	0.000000
COS mole frac.	0.000000	0.000000	0.000000	0.000000
Ethane mole frac.	1.000000	1.000000	1.000000	1.000000
Hydrogen mole frac.	0.000000	0.000000	0.000000	0.000000
Propane mole frac.	0.000000	0.000000	0.000000	0.000000

Stream	129
Vapour frac.	1.0000
Temperature F	-76.9951
Pressure psia	53.8125
Molar Flow lbmole/hr	3032.5271
Helium mole frac.	0.000000
Methane mole frac.	0.000000
Nitrogen mole frac.	0.000000
CO2 mole frac.	0.000000
H2S mole frac.	0.000000
COS mole frac.	0.000000
Ethane mole frac.	1.000000
Hydrogen mole frac.	0.000000
Propane mole frac.	0.000000

A diagram showing a 5x2 grid of zigzag lines. The left side has numbers: 6, 28, 41, 39, 5. The right side has numbers: 8, 27, 40, 38, 15. Arrows point from the numbers to the corresponding zigzag lines.

APPENDIX C

PROCESS POWER REQUIREMENTS

TABLE C-1
OPERATING POWER REQUIREMENTS
CFZ-CNG Labarge Case
200 Million SCFD Unit

	<u>Feed Dehydration</u>	<u>Refrigeration</u>	<u>CFZ</u>	<u>TPC</u>	<u>Claus</u>	<u>Overall Process</u>
EXPANDER/COMPRESSORS						
Expanders						
Plant Feed Expander			-220			
TPC Vent Gas Expander				-1890		
TOTAL (HP)	0	0	-220	-1890	0	-2110
Compressors						
Ethylene Stg 1		4110				
Ethylene Stg 2		2200				
Propane Stg 1		1820				
Propane Stg 2		6470				
Propane Stg 3		7620				
TPC Stg 1				3220		
TPC Stg 2				320		
Distill. Vapor Recomp				2750		
Crystallizer Inerts				5.0		
Claus Recycle					330	
TOTAL (HP)	0	22220	0	6295	330	28845
PUMPS						
Lean Glycol	27					
Solvent Pump	13					
CFZ Reflux			42			
CFZ Sprays			93			
TPC Slurry Stg1				250		
TPC Slurry Stg2				80		
TPC Carrier Liquid				25		
TPC Melter Stg 1				10		
TPC Melter Stg 2				10		
CO2 Adsorbent				25		
H2S from TPC				50		
Low P CO2 - EOR				290		
High P CO2 - EOR				1340		
Boiler Feed Water					60	
Quench Water					5	
TOTAL (HP)	40	0	135	2080	65	2320
MISC						
Enriched Air					8378	
GRAND TOTAL (HP)	40	22220	135	8375	8773	39543
Available Expander Power Used				1340		1340
Net Power Requirements (HP)	40	22220	135	7035	8773	38203
Electrical (KW)	30	16576	101	5248	6545	28500

TABLE C-2
OPERATING POWER REQUIREMENTS
 Generic Case
 200 MMSCFD Unit

	<u>Feed Dehydration</u>	<u>Refrigeration</u>	<u>CFZ</u>	<u>TPC</u>	<u>Claus</u>	<u>Overall Process</u>
EXPANDER/COMPRESSORS						
Expanders						
TPC Vent Gas Expander				-1170		
TOTAL (HP)	0	0	0	-1170	0	-1170
Compressors						
Ethylene		7570				
Propane Stg 1		860				
Propane Stg 2		2180				
Propane Stg 3		2940				
TPC Stg 1				714		
TPC Stg 2				80		
Claus Compressor					33	
TOTAL (HP)	0	13550	0	794	33	14377
PUMPS						
Lean Glycol	27					
Solvent Pump	13					
CFZ Reflux			124			
CFZ Sprays			180			
TPC Slurry Stg1				42		
TPC Slurry Stg2				20		
TPC Carrier Liquid				10		
TPC Melter Stg 1				5		
TPC Melter Stg 2				5		
CO2 Adsorbent				7		
H2S from TPC				13		
Boiler Feed Water					6	
Quench Water					1	
TOTAL (HP)	40	0	304	102	7	453
MISC						
Enriched Air					838	
GRAND TOTAL (HP)	40	13550	304	896	878	15668
Available Expander Power Used				714		714
Net Power Requirements (HP)	40	13550	304	182	878	14954
Electrical (KW)	30		227	136	655	1047
Fuel (mscfd)		2508				2508

TABLE C-3
OPERATING POWER REQUIREMENTS
CNG Labarge Case
200 Million SCFD Unit

	<u>Feed Dehydration</u>	<u>Refrigeration</u>	<u>MeOH Abs.</u>	<u>TPC</u>	<u>Claus</u>	<u>Overall Process</u>
EXPANDER/COMPRESSORS						
Expanders						
TPC Vent Gas Expander				-2770		
TOTAL (HP)	0	0	0	-2770	0	-2770
Compressors						
Ethane		1070				
Propane Stg 1		14690				
Propane Stg 2		0				
Propane Stg 3		0				
TPC Stg 1				4410		
TPC Stg 2				300		
Claus Recycle					770	
TOTAL (HP)	0	15760	0	4710	770	21240
PUMPS						
Lean Glycol	27					
Solvent Pump	13					
MeOH Absorbent			350			
TPC Slurry Stg1				180		
TPC Slurry Stg2				70		
CO2 Adsorbent				540		
H2S from TPC				1030		
CO2 - EOR				920		
Boiler Feed Water					60	
Quench Water					5	
TOTAL (HP)	40	0	350	2740	65	3195
MISC						
Enriched Air					8378	
GRAND TOTAL (HP)	40	15760	350	7450	9213	32813
Available Expander Power Used		1070				1070
Net Power Requirements (HP)	40	14690	350	7450	9213	31743
Electrical (KW)	30	10959	261	5558	6873	23680

APPENDIX D

EQUIPMENT SPECIFICATIONS and COSTS

TABLE D-1

CAPITAL COSTS - (1992 \$K)
200 MMSCFD Unit
CFZ-CNG Labarge Case

	<u>Feed Dehydration</u>	<u>CFZ</u>	<u>TPC</u>	<u>Claus</u>	<u>Overall Process</u>
DIRECT COSTS					
Major Equipment					
Compressors		\$8,268	\$3,889	\$298	\$12,455
Heat Exchangers		\$3,591	\$3,022	\$541	\$7,154
Pumps		\$539	\$1,405	\$34	\$1,978
Towers		\$2,327	\$521	\$28	\$2,876
Vessels		\$1,728	\$1,637	\$655	\$4,020
Dehydration Unit	\$1,394				\$1,394
Filters		\$1,072			\$1,072
Taxes (4%)	\$56	\$701	\$419	\$62	\$1,238
Freight (1.9%)	\$26	\$333	\$199	\$30	\$588
Total Major Equipment	\$1,476	\$18,559	\$11,092	\$1,648	\$32,775
Other Direct Costs					
Direct Labor	\$570	\$7,171	\$4,286	\$637	\$12,664
Piping, Electrical, and Instr.	\$894	\$11,234	\$6,714	\$997	\$19,839
Insulation, Paint, etc.	\$146	\$1,830	\$1,094	\$162	\$3,231
Structural, Concrete, Earthwork	\$294	\$3,698	\$2,210	\$328	\$6,531
Buildings	\$214	\$2,685	\$1,605	\$238	\$4,743
Taxes (4%)	\$62	\$778	\$465	\$69	\$1,374
Freight (1.0%)	\$15	\$194	\$116	\$17	\$343
Total Other Direct Costs	\$2,195	\$27,590	\$16,490	\$2,450	\$48,724
Total Direct Costs	\$3,671	\$46,149	\$27,582	\$4,097	\$81,499
INDIRECT COSTS					
Fabrication/Installation					
Supervision/Payroll Burden	\$288	\$3,616	\$2,161	\$321	\$6,386
Field Expense	\$402	\$5,060	\$3,024	\$449	\$8,936
Total Fabrication/Install.	\$690	\$8,676	\$5,185	\$770	\$15,322
Engineering Cost					
Contractor	\$628	\$7,901	\$4,722	\$701	\$13,953
Owner	\$278	\$3,493	\$2,088	\$310	\$6,169
Total Engineering Costs	\$906	\$11,394	\$6,810	\$1,012	\$20,122
Contractor Fee/Profit	\$139	\$1,751	\$1,046	\$155	\$3,092
Commissioning					
System Charge/Consumables	\$23	\$285	\$170	\$25	\$503
Supervision & Expenses	\$127	\$1,601	\$957	\$142	\$2,827
Total Commissioning	\$150	\$1,885	\$1,127	\$167	\$3,329
Total Indirect Costs	\$1,886	\$23,706	\$14,168	\$2,105	\$41,865
TOTAL BEFORE CONT.	\$5,557	\$69,855	\$41,750	\$6,202	\$123,364
CONTINGENCY (20 %)	\$1,111	\$13,971	\$8,350	\$1,240	\$24,673
TOTAL INSTALLED COST	\$6,668	\$83,826	\$50,100	\$7,443	\$148,037

TABLE D-1.1

**Equipment Cost
Labarge Gas
CFZ-CNG Process
200 MMSCFD
CFZ Section**

Tag	Service	Quantity	Size	Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C-201/202	C2 Comp-2 stage	1	6310 hp		\$2,723	\$13,027
C-301/302/303	C3 Comp - 3 stage	1	15910 hp		\$5,449	\$26,065
XC-101E	Feed Expander	1	220 hp		\$96	\$457
					<u>\$8,268</u>	<u>\$39,549</u>
Exchangers						
E-101	Feed Cooler	1	92.6 MMBtu		\$864	\$4,132
E-105	CFZ Condenser	1	25.3 MMBtu		\$494	\$2,361
E-106	CFZ Reboiler	1	49.4 MMBtu		\$1,266	\$6,056
E-201	Ethylene Exchanger	1	50.4 MMBtu		\$247	\$1,181
A-201	Ethylene Air Cooler	1	1.6 MMBtu		\$56	\$267
A-301	Propane Air Cooler	1	99.6 MMBtu		\$665	\$3,180
					<u>\$3,591</u>	<u>\$17,177</u>
Pumps						
P-101	CFZ Reflux Pumps	2	42 hp		\$235	\$1,125
P-101	CFZ Spray Pumps	2	93 hp		\$304	\$1,452
					<u>\$539</u>	<u>\$2,577</u>
Towers						
T-101	CFZ Bottom Tower	1	170 K-lbs	750 psi	\$657	\$3,141
T-102	CFZ Top Tower	1	660 K-lbs	750 psi	\$1,671	\$7,991
					<u>\$2,327</u>	<u>\$11,132</u>
Vessels						
V-101	Feed Separator	1	214 K-lbs	1250 psi	\$191	\$914
V-102	CFZ Reflux Accm.	1	327 K-lbs	750 psi	\$880	\$4,209
V-201	C2 HP Econ.	1	36 K-lbs	175 psi	\$135	\$647
V-202	C2 LP Econ.	1	30 K-lbs	175 psi	\$119	\$571
V-301	C3 HP Econ.	1	136 K-lbs	250 psi	\$155	\$743
V-302	C3 MP Econ.	1	107 K-lbs	250 psi	\$135	\$647
V-303	C3 LP Econ.	1	76 K-lbs	250 psi	\$111	\$533
					<u>\$1,728</u>	<u>\$8,265</u>
Dehydration Filters					Subtotal	
					\$16,453	\$78,700
					\$1,394	\$6,666
					<u>\$1,072</u>	<u>\$5,128</u>
					Total	
					<u>\$18,918</u>	<u>\$90,494</u>

TABLE D-1.2

**Equipment Cost
Labarge Gas
CFZ-CNG Process
200 MMSCFD
TPC Section**

Tag	Service	Quantity	Size	Mat. + Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C-TPC-1	TPC stage 1	1	3220 hp		\$1,644	\$7,865
C-TPC-2	TPC stage 2	1	320 hp		\$291	\$1,392
C-1	Dist Comp	1	2750 hp		\$1,461	\$6,987
C-2	TPC Vent	1	5 hp		\$13	\$62
Exp-1	Expander	1	1890 hp		\$480	\$2,295
					<u>\$3,889</u>	<u>\$18,601</u>
Exchangers						
HX-2	Dist. Reb./Cond.	1	64400 ft ²		\$326	\$1,558
HX-3	Lights Reb./Tail Gas Cond.	1	3730 ft ²		\$175	\$839
HX-TPC	TPC Reboiler	1	61030 ft ²		\$1,834	\$8,774
HX-4	CFZ Bttrms. Cooler	1	6360 ft ²		\$90	\$432
HX-5	Feed Cooler	1	4480 ft ²		\$110	\$527
HX-8&10	CFZ Bttrms. Cooler	1	3180 ft ²		\$155	\$743
HX-6&9	CFZ Bttrms. Cooler	1	14000 ft ²		\$331	\$1,582
					<u>\$3,022</u>	<u>\$14,456</u>
Pumps						
P-TPC-S1	TPC Slurry Stage 1	2	250 hp		\$130	\$624
P-TPC-S2	TPC Slurry Stage 2	2	80 hp		\$59	\$281
P-TPC-L2	TPC Liquid Stage 2	2	25 hp		\$27	\$130
P-TPC-M1	TPC Melter Liq Stg 1	2	10 hp		\$17	\$80
P-TPC-M2	TPC Melter Liq Stg 2	2	10 hp		\$17	\$80
P-1	Pump 1 CO2 Absorbent	2	25 hp		\$27	\$130
P-2	Pump 2 TPC H2S	2	50 hp		\$43	\$204
P-3	Pump 3 LP EOR CO2	2	290 hp		\$142	\$680
P-4	Pump 4 HP EOR CO2	2	1340 hp		\$943	\$4,511
					<u>\$1,405</u>	<u>\$6,720</u>
Towers						
T-Lights	Lights Column	1	Dia Ht 6 65'	SS 100 psi	\$272	\$1,302
T-Dist	Dist. Column	1	9 35'	SS 100 psi	\$249	\$1,190
					<u>\$521</u>	<u>\$2,493</u>
Vessels						
F-1	TPC Flasher 1	4	Dia Ht 9.25 6'	SS 50 psi	\$250	\$1,194
F-2	TPC Flasher 2	4	9.25 6'	SS 75 psi	\$250	\$1,194
WM-1	TPC Melter 1	4	9.25 12'	SS 80 psi	\$435	\$2,078
WM-2	TPC Melter 2	4	9.25 12'	SS 80 psi	\$435	\$2,078
SEP-TPC	TPC Flash Sep.	1	7 24'	SS 95 psi	\$86	\$411
SEPR -1	Dist Ovhd Sep.	1	6.5 23'	SS 105 psi	\$96	\$457
SEPR -2	Claus Offgas Sep.	1	3 10'	CS 105 psi	\$6	\$31
SEP-CO2	CO2 Storage	1	6 27'	SS 155 psi	\$81	\$387
					<u>\$1,637</u>	<u>\$7,830</u>
Total					<u>\$10,474</u>	<u>\$50,100</u>

TABLE D-1.3

**Equipment Cost
Labarge Gas
CFZ-CNG Process
200 MMSCFD
SRU Section**

Tag	Service	Quantity	Size	Mat. + Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C-SRU	Claus Recycle Compressor	1	330 hp		\$298	\$1,425
					<u>\$298</u>	<u>\$1,425</u>
Exchangers						
HX-20	Waste Heat Ex.	1	7090 ft ²		\$100	\$479
HX-30	Sulfur Condenser	1	4290 ft ²		\$70	\$336
HX-40	Hydrogenator Preheat	1	3030 ft ²		\$55	\$264
HX-50	Hydrogenator Prod Cool	1	3320 ft ²		\$55	\$264
HX-60	Quench Reboiler	1	380 ft ²		\$15	\$72
HX-70	Air Cooler	1	20600 ft ²		\$205	\$983
HX-80	Air Cooler	1	1880 ft ²		\$40	\$192
					<u>\$541</u>	<u>\$2,589</u>
Pumps						
P-BFW	Boiler Feed Water	2	60 hp		\$27	\$130
P-Quench	Quench Water	2	5 hp		\$7	\$31
					<u>\$34</u>	<u>\$162</u>
Towers						
T-Q	Quench Tower	1	Dia Ht 5.5 20'	100 psi	\$28	\$133
					<u>\$28</u>	<u>\$133</u>
Vessels						
R-1	Claus Reactor	1	Dia Ht 6 17'	100 psi	\$250	\$1,196
R-2	Hydrogenator	1	9 30'	100 psi	\$166	\$796
D-1	Mol Sieve Dryers	2	7 21'	100 psi	\$209	\$998
RGG	Reducing Gas Generator	1		100 psi	\$25	\$120
SEP-S	Sulfur Tank	1	3 11'	100 psi	\$5	\$26
					<u>\$655</u>	<u>\$3,135</u>
Total					<u>\$1,556</u>	<u>\$7,443</u>

TABLE D-2

CAPITAL COSTS (1992 \$K)
200 MMSCFD Unit
Generic Case

	Feed Dehydration	CFZ	TPC	Claus	Overall Process
DIRECT COSTS					
Major Equipment					
Compressors		\$5,737	\$969	\$53	\$6,759
Heat Exchangers		\$2,604	\$1,315	\$108	\$4,027
Pumps		\$718	\$140	\$10	\$868
Towers		\$2,832	\$272	\$10	\$3,114
Vessels		\$2,130	\$571	\$199	\$2,900
Dehydration Unit	\$1,394				\$1,394
Filters		\$1,072			\$1,072
Taxes (4%)	\$56	\$604	\$131	\$15	\$805
Freight (1.9%)	\$26	\$287	\$62	\$7	\$383
Total Major Equipment	\$1,476	\$15,983	\$3,460	\$402	\$21,322
Other Direct Costs					
Direct Labor	\$517	\$5,594	\$1,211	\$141	\$7,463
Piping, Electrical, and Instr.	\$812	\$8,791	\$1,903	\$221	\$11,727
Insulation, Paint, etc	\$148	\$1,598	\$346	\$40	\$2,132
Structural, Concrete, Earthwork	\$221	\$2,398	\$519	\$60	\$3,198
Buildings	\$74	\$799	\$173	\$20	\$1,066
Taxes (4%)	\$62	\$670	\$145	\$17	\$894
Freight (1.0%)	\$15	\$167	\$36	\$4	\$223
Total Other Direct Costs	\$1,849	\$20,018	\$4,333	\$504	\$26,703
Total Direct Costs	\$3,325	\$36,001	\$7,793	\$906	\$48,025
INDIRECT COSTS					
Fabrication/Installation					
Supervision/Payroll Burden	\$295	\$3,197	\$692	\$80	\$4,264
Field Expense	\$221	\$2,398	\$519	\$60	\$3,198
Total Fabrication/Install.	\$517	\$5,594	\$1,211	\$141	\$7,463
Engineering Cost					
Contractor	\$516	\$5,583	\$1,209	\$141	\$7,448
Owner	\$148	\$1,598	\$346	\$40	\$2,132
Total Engineering Costs	\$663	\$7,182	\$1,555	\$181	\$9,580
Contractor Fee/Profit	\$148	\$1,598	\$346	\$40	\$2,132
Commissioning					
System Charge/Consumables	\$23	\$245	\$53	\$6	\$327
Supervision & Expenses	\$127	\$1,378	\$298	\$35	\$1,839
Total Commissioning	\$150	\$1,624	\$351	\$41	\$2,166
Total Indirect Costs	\$1,478	\$15,998	\$3,463	\$403	\$21,341
TOTAL BEFORE CONT.	\$4,803	\$51,999	\$11,256	\$1,309	\$69,366
CONTINGENCY	\$369	\$3,996	\$865	\$101	\$5,330
TOTAL INSTALLED COST	\$5,172	\$55,995	\$12,120	\$1,410	\$74,697

TABLE D-2.1

**Equipment Cost
Generic Gas
CFZ-CNG Process
200 MMSCFD
CFZ Section**

<u>Tag</u>	<u>Service</u>	<u>Quantity</u>	<u>Size</u>	<u>Press.</u>	<u>Equipment Cost, \$K</u>	<u>Installed Cost 1992 \$K</u>
Compressors						
C-201	C2 Comp - 1 stage	1	7570 hp		\$3,122	\$11,582
C-301/302/303	C3 Comp - 3 stage	1	5980 hp		\$2,616	\$9,704
					<u>\$5,737</u>	<u>\$21,286</u>
Exchangers						
E-101	Feed Cooler	1	34.7 MMBtu		\$478	\$1,772
E-105	CFZ Condenser	1	48.1 MMBtu		\$725	\$2,688
E-106	CFZ Reboiler	1	17.4 MMBtu		\$677	\$2,511
E-201	Ethylene Exchanger	1	66.6 MMBtu		\$291	\$1,078
A-201	Ethylene Air Cooler	1	1.3 MMBtu		\$48	\$177
A-301	Propane Air Cooler	1	40.1 MMBtu		\$386	\$1,433
					<u>\$2,604</u>	<u>\$9,659</u>
Pumps						
P-101	CFZ Reflux Pumps	2	124 hp		\$338	\$1,253
P-101	CFZ Spray Pumps	2	180 hp		\$381	\$1,412
					<u>\$718</u>	<u>\$2,665</u>
Towers						
T-101	CFZ Bottom Tower	1	420 K-lbs	750 psi	\$1,179	\$4,375
T-102	CFZ Top Tower	1	650 K-lbs	750 psi	\$1,653	\$6,132
					<u>\$2,832</u>	<u>\$10,507</u>
Vessels						
V-102	CFZ Reflux Accm.	1	924 K-lbs	750 psi	\$1,640	\$6,085
V-202	C2 Economizer	1	102 K-lbs	175 psi	\$251	\$930
V-301	C3 HP Econ.	1	61 K-lbs	250 psi	\$96	\$354
V-302	C3 MP Econ.	1	44 K-lbs	250 psi	\$80	\$295
V-303	C3 LP Econ.	1	31 K-lbs	250 psi	\$64	\$236
					<u>\$2,130</u>	<u>\$7,902</u>
Dehydration Filters					Subtotal	\$14,021
						\$52,019
						\$1,394
						\$5,170
						\$1,072
						\$3,977
Total					\$16,487	\$61,167

TABLE D-2.2

**Equipment Cost
Generic Gas
CFZ-CNG Process
200 MMSCFD
TPC Section**

Tag	Service	Quantity	Size	Mat. + Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C-TPC-1	TPC stage 1	1	714 hp		\$531	\$1,971
C-TPC-2	TPC stage 2	1	80 hp		\$103	\$382
Exp-1	Expander	1	1170 hp		\$335	\$1,242
					<u>\$969</u>	<u>\$3,595</u>
Exchangers						
HX-1	Feed Cooler	1	8120 ft ²		\$267	\$991
HX-4	Reboiler	1	21170 ft ²		\$252	\$934
HX-TPC	TPC Reboiler	1	23690 ft ²		\$707	\$2,625
HX-5	Tail Gas Condenser	1	233 ft ²		\$9	\$34
HX-6	Tail Gas Cooler	1	344 ft ²		\$80	\$296
					<u>\$1,315</u>	<u>\$4,879</u>
Pumps						
P-TPC-S1	TPC Slurry Stage 1	2	42 hp		\$38	\$141
P-TPC-S2	TPC Slurry Stage 2	2	20 hp		\$24	\$89
P-TPC-L2	TPC Liquid Stage 2	2	10 hp		\$18	\$67
P-TPC-M1	TPC Melted Liq Stg 1	2	5 hp		\$12	\$46
P-TPC-M2	TPC Melted Liq Stg 2	2	5 hp		\$12	\$46
P-1	Pump 1 CO2 Absorbent	2	7 hp		\$15	\$56
P-2	Pump 2 TPC H2S	2	13 hp		\$20	\$73
					<u>\$140</u>	<u>\$518</u>
Towers						
T-AS	Lights Column	1	Dia Ht 6 65'	SS 100 psi	\$272	\$1,010
					<u>\$272</u>	<u>\$1,010</u>
Vessels						
F-1	TPC Flasher 1	2	7 6'	SS 60 psi	\$94	\$350
F-2	TPC Flasher 2	2	7 6'	SS 75 psi	\$94	\$350
WM-1	TPC Melted 1	2	7 12'	SS 80 psi	\$164	\$610
WM-2	TPC Melted 2	2	7 12'	SS 80 psi	\$164	\$610
SEPR -2	Claus Offgas Sep.	1	2 7'	SS 105 psi	\$11	\$42
SEP -CO2	CO2 Storage	1	6 12'	SS 100 psi	\$42	\$157
					<u>\$571</u>	<u>\$2,120</u>
Total					<u>\$3,267</u>	<u>\$12,122</u>

TABLE D-2.3

**Equipment Cost
Generic Gas
CFZ-CNG Process
200 MMSCFD
SRU Section**

Tag	Service	Quantity	Size	Mat. + Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C-SRU	Claus Recycle Compressor	1	33 hp		\$53	\$196
					\$53	\$196
Exchangers						
HX-20	Waste Heat Ex.	1	709 ft ²		\$20	\$74
HX-30	Sulfur Condenser	1	429 ft ²		\$14	\$52
HX-40	Hydrogenator Preheat	1	303 ft ²		\$11	\$41
HX-50	Hydrogenator Prod Cool	1	332 ft ²		\$11	\$41
HX-60	Quench Reboiler	1	38 ft ²		\$3	\$11
HX-70	Air Cooler	1	2060 ft ²		\$41	\$152
HX-80	Air Cooler	1	188 ft ²		\$8	\$30
					\$108	\$401
Pumps						
P-BFW	Boiler Feed Water	2	6 hp		\$7	\$25
P-Quench	Quench Water	2	0.5 hp		\$3	\$11
					\$10	\$36
Towers						
T-Q	Quench Tower	1	Dia Ht 2 20'	100 psi	\$10	\$37
					\$10	\$37
Vessels						
R-1	Claus Reactor	1	Dia Ht 3 7'	100 psi	\$120	\$445
R-2	Hydrogenator	1	4 17'	100 psi	\$24	\$87
D-1	Mol Sieve Dryers	2	3 11'	100 psi	\$47	\$173
RGG	Reducing Gas Generator	1		100 psi	\$5	\$19
SEP-S	Sulfur Tank	1	2 7'	100 psi	\$4	\$13
					\$199	\$738
Total					\$380	\$1,408

TABLE D-3

CAPITAL COSTS - (1992 \$K)
200 MMSCFD Unit
CNG Labarge Case

	Feed Dehydration	CNG	Claus	Overall Process
DIRECT COSTS				
<i>Major Equipment</i>				
Compressors		\$8,761	\$562	\$9,323
Heat Exchangers		\$5,701	\$541	\$6,242
Pumps		\$3,198	\$34	\$3,232
Towers		\$1,432	\$28	\$1,460
Vessels		\$1,719	\$655	\$2,374
Dehydration Unit	\$1,394			\$1,394
Filters		\$1,072		\$1,072
Taxes (4%)	\$56	\$875	\$73	\$1,004
Freight (1.9%)	\$26	\$416	\$35	\$477
<i>Total Major Equipment</i>	<u>\$1,476</u>	<u>\$23,174</u>	<u>\$1,927</u>	<u>\$26,578</u>
<i>Other Direct Costs</i>				
Direct Labor	\$570	\$8,954	\$745	\$10,269
Piping, Electrical, and Instr.	\$894	\$14,027	\$1,167	\$16,088
Insulation, Paint, etc.	\$146	\$2,285	\$190	\$2,620
Structural, Concrete, Earthwork	\$294	\$4,618	\$384	\$5,296
Buildings	\$214	\$3,353	\$279	\$3,846
Taxes (4%)	\$62	\$971	\$81	\$1,114
Freight (1.0%)	\$15	\$243	\$20	\$278
<i>Total Other Direct Costs</i>	<u>\$2,195</u>	<u>\$34,451</u>	<u>\$2,865</u>	<u>\$39,511</u>
<i>Total Direct Costs</i>	<u><u>\$3,671</u></u>	<u><u>\$57,625</u></u>	<u><u>\$4,793</u></u>	<u><u>\$66,089</u></u>
INDIRECT COSTS				
<i>Fabrication/Installation</i>				
Supervision/Payroll Burden	\$288	\$4,515	\$376	\$5,178
Field Expense	\$402	\$6,318	\$525	\$7,246
<i>Total Fabrication/Install.</i>	<u>\$690</u>	<u>\$10,834</u>	<u>\$901</u>	<u>\$12,425</u>
<i>Engineering Cost</i>				
Contractor	\$629	\$9,876	\$821	\$11,326
Owner	\$278	\$4,362	\$363	\$5,002
<i>Total Engineering Costs</i>	<u>\$907</u>	<u>\$14,238</u>	<u>\$1,184</u>	<u>\$16,329</u>
<i>Contractor Fee/Profit</i>	\$139	\$2,186	\$182	\$2,507
<i>Commissioning</i>				
System Charge/Consumables	\$23	\$355	\$30	\$408
Supervision & Expenses	\$127	\$1,999	\$166	\$2,292
<i>Total Commissioning</i>	<u>\$150</u>	<u>\$2,354</u>	<u>\$196</u>	<u>\$2,700</u>
<i>Total Indirect Costs</i>	<u><u>\$1,886</u></u>	<u><u>\$29,611</u></u>	<u><u>\$2,463</u></u>	<u><u>\$33,960</u></u>
TOTAL BEFORE CONT.	<u>\$5,557</u>	<u>\$87,237</u>	<u>\$7,255</u>	<u>\$100,049</u>
CONTINGENCY (20 %)	<u>\$1,111</u>	<u>\$17,447</u>	<u>\$1,451</u>	<u>\$20,010</u>
TOTAL INSTALLED COST	<u><u>\$6,669</u></u>	<u><u>\$104,684</u></u>	<u><u>\$8,707</u></u>	<u><u>\$120,059</u></u>

TABLE D-3.1
Equipment Cost
Labarge Gas CNG Claus Process
200 MMSCFD

Tag	Service	Quantity	Size	Mat. + Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C2 Ref	C2 Comp	1	1070 hp		\$720	\$3,442
C3 Ref	C3 Comp - 3 stage	1	14690 hp		\$5,133	\$24,551
C-TPC-1	TPC stage 1	1	4410 hp		\$2,082	\$9,957
C-TPC-2	TPC stage 2	1	300 hp		\$277	\$1,326
Exp	Expander	1	2270 hp		\$550	\$2,633
					<u>\$8,761</u>	<u>\$41,910</u>
Exchangers						
HX-FD	Feed Cooler	1	42620 ft ²	CS	\$371	\$1,775
HX-TPC-FD	TPC Feed Cooler	1	72150 ft ²	Al	\$1,207	\$5,774
HX-TPC-REB	TPC Reboiler	1	85360 ft ²	Al	\$1,358	\$6,495
HX-TAIL	Tail Gas Cooler	1	5820 ft ²	CS	\$92	\$441
HX-REB	Reboiler	1	126230 ft ²	CS	\$794	\$3,796
HX-COND	Condenser	1	14420 ft ²	Al	\$391	\$1,871
HX-C2-SUB	C2 Subcooler	1	4640 ft ²	Al	\$177	\$846
HX-C3-SUB	C3 Subcooler	1	1230 ft ²	CS	\$31	\$148
HX-C3-AIR	C3 Air Cooler	1	61 MMBtu	CS	\$496	\$2,373
HX-C2-C3	C3 Evaporator	1	30050 ft ²	CS	\$785	\$3,753
					<u>\$5,701</u>	<u>\$27,272</u>
Pumps						
P-EOR	CO2 EOR	2	920 hp		\$2,219	\$10,613
P-CO2ABS	CO2 absorbent	2	540 hp		\$243	\$1,163
P-H2S	H2S pump	2	1030 hp		\$385	\$1,841
P-SL1	TPC Slurry 1	2	180 hp		\$97	\$463
P-SL2	TPC Slurry 2	2	70 hp		\$41	\$197
P-MeOH	MeOH Absorbent	2	350 hp		\$168	\$806
P-REFLUX	Reflux	2	20 hp		\$45	\$215
					<u>\$3,198</u>	<u>\$15,298</u>
Towers						
ABS-STR	CO2 abs-stripper	1	Dia Ht 9 100'	CS 700 psi	\$428	\$2,048
TAIL-ABS	Tailgas absorber (top)	1	3.5 40'	SS 155 psi	\$280	\$1,339
TAIL-ABS	Tailgas absorber (bottom)	1	5 10'	SS 155 psi	\$132	\$631
MeOH-STRIP	MeOH Stripper (top)	1	7 10'	SS 20 psi	\$57	\$272
MeOH-STRIP	MeOH Stripper (bottom)	1	5.5 40'	SS 20 psi	\$135	\$648
MeOH-ABS	MeOH absorber	1	5 40'	SS 700 psi	\$400	\$1,913
					<u>\$1,432</u>	<u>\$6,851</u>
Vessels						
F-1	TPC Flasher 1	4	9 6' V	SS 55 psi	\$243	\$1,162
F-2	TPC Flasher 2	4	9 6' V	SS 75 psi	\$243	\$1,162
WM-1	TPC Melted 1	4	9 12' V	SS 78 psi	\$423	\$2,022
WM-2	TPC Melted 2	4	9 12' V	SS 78 psi	\$423	\$2,022
S-REFLUX	Reflux drum	1	5 27' H	SS 700 psi	\$175	\$838
S-MeOH	MeOH Fuel Gas Sep	1	6 21' H	CS 200 psi	\$23	\$112
S-TAIL	Tail Gas Sep	1	3 11' H	CS 155 psi	\$7	\$32
S-C2	C2 Econ.	1	5 17' H	SS 150 psi	\$58	\$276
S-C3L	C3 Low P Econ.	1	6.5 23' H	CS 190 psi	\$27	\$131
S-C3M	C3 Mid. P Econ.	1	8.5 30' H	CS 190 psi	\$44	\$211
S-C3H	C3 High P Econ.	1	9.5 33' H	CS 190 psi	\$53	\$255
					<u>\$1,719</u>	<u>\$8,223</u>
				Subtotal	<u>\$20,812</u>	<u>\$99,554</u>
HP Claus	High Pressure Claus System				\$1,821	\$8,711
Dehydration					\$1,394	\$6,666
Filters					\$1,072	\$5,128
				Total	<u>\$25,099</u>	<u>\$120,059</u>

TABLE D-3.2
Equipment Cost
Labarge Gas
CNG Claus Process
200 MMSCFD
SRU Section

Tag	Service	Quantity	Size	Mat. + Press.	Equipment Cost, \$K	Installed Cost 1992 \$K
Compressors						
C-SRU	Claus Recycle Compressor	1	770 hp		\$562	\$2,690
					\$562	\$2,690
Exchangers						
HX-20	Waste Heat Ex.	1	7090 ft ²		\$100	\$479
HX-30	Sulfur Condenser	1	4290 ft ²		\$70	\$336
HX-40	Hydrogenator Preheat	1	3030 ft ²		\$55	\$264
HX-50	Hydrogenator Prod Cool	1	3320 ft ²		\$55	\$264
HX-60	Quench Reboiler	1	380 ft ²		\$15	\$72
HX-70	Air Cooler	1	20600 ft ²		\$205	\$983
HX-80	Air Cooler	1	1880 ft ²		\$40	\$192
					\$541	\$2,589
Pumps						
P-BFW	Boiler Feed Water	2	60 hp		\$27	\$130
P-Quench	Quench Water	2	5 hp		\$7	\$31
					\$34	\$162
Towers						
			Dia Ht			
T-Q	Quench Tower	1	5.5 20'	100 psi	\$28	\$133
					\$28	\$133
Vessels						
			Dia Ht			
R-1	Claus Reactor	1	6 17'	100 psi	\$250	\$1,196
R-2	Hydrogenator	1	9 30'	100 psi	\$166	\$796
D-1	Mol Sieve Dryers	2	7 21'	100 psi	\$209	\$998
RGG	Reducing Gas Generator	1		100 psi	\$25	\$120
SEP-S	Sulfur Tank	1	3 11'	100 psi	\$5	\$26
					\$655	\$3,135
Total					\$1,821	\$8,708

APPENDIX E

OPERATING COSTS

TABLE E-1

**CFZ - CNG Claus Process
Annual Operating Costs
Labarge Composition Case
200 Million SCFD Unit**

	Requirement	Unit Cost	Annual Cost (\$k)	Cost per MSCF of Plant Feed
Utilities				
Electrical Power	28500 kW	0.03 \$/kW-hr	\$7,182	\$0.103
Chemicals				
Glycol	1500 gal/yr	4.46 \$/gal	\$7	
Amine (pH ctrl.)	300 gal/yr	5.19 \$/gal	\$2	
Drizo Solvent	18000 gal/yr	0.56 \$/gal	\$10	
Oxygen	87550 \$/mo		\$1,051	
Ethylene	4000 lb/yr	0.28 \$/lb	\$1	
<i>Subtotal</i>			\$1,070	\$0.015
Total Utilities & Chemicals			\$8,252	\$0.118
Direct Labor				
Supervision	2080 hr/yr	20 \$/hr	\$42	
Operating	17520 hr/yr	15 \$/hr	\$263	
Maintenance	4380 hr/yr	15 \$/hr	\$66	
<i>Total Direct Labor</i>			\$370	\$0.005
Overhead				
Plant Overhead	125.0% of labor		\$463	
Engineering Support	25.0% of labor		\$93	
General Administrative	25.0% of labor		\$93	
<i>Total Overhead</i>			\$648	\$0.009
Total Labor & Overhead			\$1,018	\$0.015
Maintenance Materials	1.0% of Cap. Invest.		\$1,480	\$0.021
Other Costs				
NRU/He Processing			\$5,250	\$0.075
Local Taxes & Insurance	1.0% of Cap. Invest.		\$1,480	\$0.021
	1.0% of Sales		\$807	\$0.012
Total Other Costs			\$9,018	\$0.129
ANNUAL OPERATING COSTS			\$18,288	\$0.261
* Capital Investment (\$K)	\$148,037			
* On stream days per year :	350			
	95.9%			
* Total MSCF gas processed	70,000,000			

TABLE E-2

**CFZ CNG Claus Process
Annual Operating Costs
Generic Composition Case
200 Million SCFD Unit**

	Requirement	Unit Cost	Annual Cost (\$k)	Cost per MSCF of Plant Feed
Utilities				
Electrical Power	1047 kW	0.05 \$/kW-hr	440	\$0.006
Fuel Gas	2508 mscf/d	Reduces Available Sales Gas		
Chemicals				
Glycol	1500 gal/yr	4.464 \$/gal	7	
Amine	300 gal/yr	5.187 \$/gal	2	
Drizo Solvent	18000 gal/yr	0.56 \$/gal	10	
Oxygen	19600 \$/mo		235	
Ethylene	4000 lb/yr	0.28 \$/lb	1	
Subtotal			255	\$0.004
Total Utilities & Chemicals			694	\$0.010
Direct Labor				
Supervision	2080 hr/yr	20 \$/hr	42	
Operating	17520 hr/yr	15 \$/hr	263	
Maintenance	4380 hr/yr	15 \$/hr	66	
Total Direct Labor			370	\$0.005
Overhead				
Plant Overhead	125.0% of labor		463	
Engineering Support	25.0% of labor		93	
General Administrative	25.0% of labor		93	
Total Overhead			648	\$0.009
Total Labor & Overhead			1,018	\$0.015
Maintenance Materials	1.0% of Cap. Invest.		747	\$0.011
Other Costs				
Local Taxes & Insurance	1.0% of Cap. Invest.		747	\$0.011
	1.0% of sales		1,271	\$0.018
Total Other Costs			2,765	\$0.039
ANNUAL OPERATING COSTS			4,477	\$0.064
* Capital Investment (\$K)	\$74,697			
* On stream days per year :	350			
	95.9%			
* Total MSCF gas processed	70,000,000			

TABLE E-3

**CNG Claus Process
Annual Operating Costs
Labarge Composition Case
200 Million SCFD Unit**

	Requirement		Unit Cost		Annual Cost (\$K)	Cost per MSCF of Plant Feed
Utilities						
Electrical Power	23680	kW	0.03	\$/kW-hr	\$5,967	\$0.085
Chemicals						
Glycol	1500	gal/yr	4.46	\$/gal	\$7	
Amine (pH ctrl.)	300	gal/yr	5.19	\$/gal	\$2	
Drizo Solvent	18000	gal/yr	0.56	\$/gal	\$10	
Methanol	15000	gal/yr	1	\$/gal	\$15	
Oxygen	87550	\$/mo			\$1,051	
Ethylene	4000	lb/yr	0.28	\$/lb	\$1	
<i>Subtotal</i>					\$1,085	\$0.016
Total Utilities & Chemicals					\$7,052	\$0.101
Direct Labor						
Supervision	2080	hr/yr	20	\$/hr	\$42	
Operating	17520	hr/yr	15	\$/hr	\$263	
Maintenance	4380	hr/yr	15	\$/hr	\$66	
<i>Total Direct Labor</i>					\$370	\$0.005
Overhead						
Plant Overhead	125.0%	of labor			\$463	
Engineering Support	25.0%	of labor			\$93	
General Administrative	25.0%	of labor			\$93	
<i>Total Overhead</i>					\$648	\$0.009
Total Labor & Overhead					\$1,018	\$0.015
Maintenance Materials	1.0%	of Cap. Invst.			\$1,201	\$0.017
Other Costs						
NRU/He Processing					\$5,250	\$0.075
Local Taxes & Insurance	1.0%	of Cap. Invst.			\$1,201	\$0.017
	1.0%	of Sales			\$743	\$0.011
<i>Total Other Costs</i>					\$8,394	\$0.120
ANNUAL OPERATING COSTS					\$16,464	\$0.235
<hr/>						
* Capital Investment (\$K)			\$120,059			
* On stream days per year :			350			
			95.9%			
* Total MSCF gas processed			70,000,000			

APPENDIX F

EXISTING SRU TECHNOLOGY PROCESS FLOW DIAGRAMS

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Figure F-1
Process Flow Diagram
Conventional Claus Process

Figure F-2
Amoco CBA Process

Figure F-3
Delta Projects MCRC Process

Figure F-4
SNEA/Lurgi Sulfreen Process

Figure F-5
BASF AG Catasulf Process

Figure F-6
Mobil Oil MODOP Process

Figure F-7
Unocal Selectox Process

Figure F-8
Comprimo BV Superclaus Process

Figure F-9
ARI Technologies LoCat Process

Figure F-10
Ralph M Parsons Stretford Process

Figure F-11
Shell Oil SulFerox Process

Figure F-12
La Gaz Integral Enterprise Sulfint Process

Figure F-13
Linde AG Sulfolin Process

Figure F-14
Institute Francais du Petrol IFP Process

Figure F-15
Alberta Energy Company RSRP Process

END

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4/14/94