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BIOLOGICAL CONVERSION OF SYNTHESIS GAS

Topical Report: Economic Evaluations

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1.0 INTRODUCTION

Synthesis gas, a mixture of primarily CO, H₂ and CO₂, is a major building block in the production of fuels and chemicals. The gas is derived from non-gaseous raw materials such as coal, shale oil, tar sands, heavy residue and biomass. The composition of synthesis gas is dependent upon the raw material used and the gasification process. Coal-derived gas is rich in CO and H₂, with lower concentrations of CO₂ and CH₄ and traces of H₂S and COS (1).

Synthesis gas may be used in the production of both gaseous and liquid fuels. The conventional method of producing gaseous fuel first employs the catalytic shift reaction. In this reaction, CO and H₂O are converted into H₂ and CO₂. Following the removal of excess CO₂ and sulfur contaminants such as H₂S and COS, the H₂ and CO₂ are catalytically methanated to CH₄ and H₂O. After drying, the product gas contains 95-98% CH₄.

Most methanation processes require temperatures of 300-700°C and operating pressures of 3-20 atm (2). The catalysts used for the methanation step include nickel and potassium-based catalysts, as well as mixtures of iron/chromium oxides and zinc/copper oxides (3,4). Poisons for these catalysts include chlorine and sulfur (5,6). In addition, chemical catalytic processes are known to produce by-products such as methanol, formaldehyde and acetic acid (7).

A wide variety of both liquid and gaseous fuels may be produced from synthesis gas using Fischer-Tropsch synthesis including light hydrocarbons (methane, ethane), fuel range hydrocarbon distillates and heavy waxes. The stoichiometry for product formation may be ideally represented by the equation:

However, based on actual yields in a Synthol entrained flow reactor, the overall stoichiometry is better characterized by the empirical relation (8):

As is noted in Equations (1.1) and (1.2), about 2 moles of H_2 are required for every mole of CO .

The H_2/CO ratio in synthesis gas is dependent upon both the grade of coal used in the feed and the type of gasifier employed. Modern gasifiers produce synthesis gas with a H_2/CO ratio as low as 0.6. Therefore, an increase in the H_2 content of synthesis gas is usually necessary prior to Fischer-Tropsch synthesis, which requires coupling water gas shift and gasification.

The utilization of coal to produce liquid fuels by Fischer-Tropsch synthesis is a multi-step process. After gasification at $700\text{--}2000^\circ\text{C}$ (9), the raw gas is quenched and scrubbed to condense tars and oils and to remove dust and water soluble materials from the gas stream such as phenols, chloride, ammonia, cyanide, thiocyanate, and perhaps some H_2S (10). A catalytic water-gas shift conversion at about 225°C is then used to adjust the H_2/CO ratio:

A high H_2/CO ratio is not only necessary to satisfy the stoichiometry of the Fischer-Tropsch synthesis, but is beneficial in controlling carbon deposition and reducing associated catalyst deactivation (11).

Following shift conversion, the gas contains primarily H_2 , CO , CO_2 , H_2O , sulfur gases and possibly methane, although small amounts of higher aliphatic hydrocarbons and light oils are sometimes present. The sulfur is present as H_2S and COS , with smaller amounts of organic sulfides or mercaptans. These "acid gases", including CO_2 , are removed using a variety of processes, including Claus, absorption, or liquid phase oxidation processes. The sulfur

concentration must be reduced to less than 0.1 ppm to protect sensitive downstream catalysts (10). Following purification, the gas is then available for conversion to C₁-C₃₅ hydrocarbons in a Fischer-Tropsch synthesis reactor at 225-365°C.

There are several limitations with water gas shift, purification and Fischer-Tropsch synthesis. Each of these operations are conducted in separate steps, at different conditions requiring intermediate adjustment of temperature or pressure, which results in a very complex and expensive process. Also, these processes are carried out at elevated temperatures and pressures, and thus suffer from low thermal efficiencies. The catalysts used in the Fischer-Tropsch synthesis are particularly sensitive to the presence of sulfur gases or carbon deposition and deteriorate rapidly. Carbon deposition, as was mentioned previously, is enhanced by low H₂/CO ratios in the synthesis gas. Finally, more specific catalysts are required to avoid production of a broad array of chemicals and fuels and to improve raw material yields from Fischer-Tropsch synthesis.

1.1 Biological Synthesis Gas Conversion

Biological processes may be utilized as an alternative to the traditional methods of producing fuels from raw synthesis gas. Biological processes operate at near ambient temperatures, and yields and energy efficiencies are high. Many microorganisms (biocatalysts) are not sensitive to the presence of low concentrations of toxic gas compounds found in synthesis gas (12,13). In addition, biological reactions are very specific, and often a single product is produced as a result of biosynthesis.

A single reaction vessel can often be used to carry out several biological reactions in series or parallel. Therefore, one reactor might be

used for water gas shift, gas purification (sulfur and CO₂ removal) and product formation (Fischer-Tropsch synthesis or methanation) by utilizing a co-culture or mixed culture. As is shown in Figure 1.1, the process for biological production of hydrocarbons from synthesis gas could be very simple. Following gasification, the gas temperature would be reduced to 95-135°F, with the opportunity for heat recovery and improved energy efficiency. The cool gases would then be bubbled through a biological reactor, where a culture of microorganisms would be maintained to produce H₂ and remove sulfur gases and CO₂. Cultures may be developed that utilize CO, CO₂ and H₂ as substrates to produce hydrocarbons, such that adjustment of the H₂/CO ratio in the synthesis gas is not essential. Also, the removal of CO₂ may not be necessary if a liquid product is produced in the reactor. High conversions in the reactor will be necessary to avoid recycle. Therefore, significant potential exists for developing a simple, more economical biological process for hydrocarbon production from coal.

The disadvantages of biological processes include slower reactions and the need for sterile conditions to prevent contamination. When utilizing synthesis gas, sterility is insured by the presence of CO, which is toxic to most microorganisms. Biological reactions involving gaseous substrates may be operated under mass transfer limiting conditions, where mass transfer controls the rate of transport of gaseous reactants and, consequently, reactor size. Reactor rates have been found to be proportional to pressure, and the use of special bioreactors at moderate pressures (150 psia) have resulted in residence times for near complete conversion of synthesis gas of only a few minutes (14,15). Therefore, fast reaction rates and small reactors can be achieved by the appropriate bioreactor design.

Figure 1.1. Comparison of Conventional and biological routes for methane production from coal.

1.2 Brief Statement of Purpose

The purpose of the proposed research is to develop a technically and economically feasible process for biologically producing H_2 from synthesis gas while, at the same time, removing harmful sulfur gas compounds. This research builds upon previous research studies carried out in the University of Arkansas laboratories where the technical and economic feasibility of biologically producing CH_4 from synthesis gas components has been demonstrated. Six major tasks are being studied:

1. Culture development, where the best cultures are selected and conditions optimized for simultaneous hydrogen production and sulfur gas removal;
2. Mass transfer and kinetic studies in which equations necessary for process design are developed;
3. Bioreactor design studies, where the cultures chosen in Task 1 are utilized in continuous reaction vessels to demonstrate process feasibility and define operating conditions;
4. Evaluation of biological synthesis gas conversion under limiting conditions in preparation for industrial demonstration studies;
5. Process scale-up where laboratory data are scaled to larger-size units in preparation for process demonstration in a pilot-scale unit; and
6. Economic evaluation, where process simulations are used to project process economics and identify high cost areas during sensitivity analyses.

The bioconversion studies are built upon previous experimental work performed under DOE contract. A strong foundation in gaseous substrate fermentation with a major emphasis on continuous bioreactor design and operation has been achieved. Although the technical feasibility of biological

methane production from synthesis gas has been demonstrated, several important pieces of information are needed prior to recommending a demonstration unit for this biological conversion. These process development studies can be grouped into three categories:

1. Culture development under industrial conditions;
 2. Investigation of alternate reactor systems to minimize capital costs;
- and
3. Process scale-up measurements.

Culture development studies under industrial conditions involve operating the cultures under less than optimum conditions in order to define operating limits for the cultures. The effects of trace quantities of oxygen in the feed on the strict anaerobes will be quantified. In addition, synthesis gas contaminants other than sulfur gases will be added to the gas stream to measure the effects of these materials on cell growth, substrate consumption and product formation. The effects of various CO/H₂ ratios on product yield and fermentation rates will be measured. Finally, efforts will be directed toward developing a cheap minimal media for synthesis gas fermentation.

An economic evaluation of various reactor alternatives has shown that the use of traditional mass transfer equipment yields high capital costs due to relatively poor gas-liquid mass transfer. A reaction system with mass transfer capabilities that minimize capital costs is thus sought. Various reaction systems are potential candidates and will be tested, including modified trickle-bed reactors and pulse and airlift fermenters. Also, various compounds will be added to the fermentation media in an effort to increase the mass transfer coefficient by changing physical properties.

Scale-up measurements involve several experiments. First, physical properties of the fermentation media will be evaluated under a variety of fermentation conditions. Two physical properties, medium viscosity and surface tension, are thought to change significantly with increasing cell density during the fermentation. Methods for handling these changes during process scale-up will be addressed. SEcondly, the limiting conditions of gas-liquid fermentation need to be defined. The maximum operable cell density and the performance of the cultures under extreme agitation rates, cell densities, operating pressure and high product concentrations will be evaluated. Finally, the scale-up correlations used in the economic analysis will be tested in larger bench-scale continuous reactors prior to recommending orocess demonstration in a PDU.

The production of hydrogen from CO and water can be carried out biologically by the facultative photosynthetic bacterium *Rhodospirillum rubrum* as an alternative to the water gas shift reaction. H₂ production is essentially in adjusting the H₂/CO ratio for further utilization of the syngas. Experiments in the University of Arkansas laboratories have shown that this bacterium is fast growing, exhibits fast CO uptake and is table in the presence of high CO and sulfur gas compositions. *R. rubrum* requires a tungsten light source for growth; however, the actual chemical reaction can be carried out in the dark. Therefore, only a small bioreactor for bacterial growth is required, thereby reducing capital and radiant energy costs.

Sulfur gas removal may be carried out by the anaerobic bacterium *Chlorobium thiosulfatophilum* (16,17). *C. thiosulfatophilum*, a phototropic sulfur bacterium, has the advantage of producing sulfur which could be easily

separated as a solid by-product. It has also been shown to utilize COS in conjunction with the chemical reaction

on *C. thiosulfatophilum* is superior to other sulfur gas-utilizing species such as *Thiobacillus denitrificans*, which produces sulfate as the product in H₂S conversion (18,19).

1.2.1 Purpose

The purpose of this report is to present economic evaluations for H₂ production from synthesis gas by *R. rubrum*. Cases are presented with and without light requirements and in stirred tank and immobilized cell reactors. In addition, economic information is presented for isolate ERIH₂ (from Engineering Resources, Inc.) in the two reactors with and without H₂ recovery.

2.0 ECONOMIC EVALUATIONS

The ultimate test for a biological process is an economic evaluation based upon the technical information gathered. An evaluation of biological water gas shift using *R. rubrum* has been performed to process 50 million standard cubic feet of synthesis gas per day (50 MMSCFD). The gas is assumed to consist of 65 percent (by volume) CO, 20 percent H₂, 4 percent CH₄ and 11 percent CO₂, a typical concentration for coal-derived synthesis gas. The CO conversion is assumed to be 95 percent, the growth yield is 0.4 g cell/g glucose and the operating liquid phase tension is 0.2 atm. A summary of these design assumptions is shown in Table 2.1. Several cases were considered:

- a) CSTRs without light
- b) CSTRs with light

Table 2.1

Design Assumptions

Design Basis: 50 MM SCF/day synthesis gas

Syngas Composition:

CO: 65%
H₂: 20%
CH₄: 4%
CO₂: 11%

CO conversion: 95 percent

Culture: *R. rubrum*

Growth Yield: 0.4 g cell/g Glucose

Specific gas uptake rate: 0.055 gmole/g cell hr

Operating Pressure: 17 atm

Operating P_LCO: 0.1 atm

Table 2.2

Capital Equipment Requirements

Case A: *R. rubrum*, CSTRs, no light

Ractors, 6 CSTRs, 1200 m³, 17 atm
K_{La} = 177 hr⁻¹, 1500 hp for agitation
Cost: \$6,045,000 (1993, delivered)

Medium pumps, 6, 11 hp, reciprocating
Cost: \$176,000 (1993, delivered)

Fixed Capital Investment

3 X \$6,221,000 = \$18,663,000
= \$0.107/MSCF synthesis gas

- c) countercurrent ICRs, no light
- d) countercurrent ICRs with light
- e) improved strain in CSTR
- f) improved strain in ICR.

2.1 Evaluations with *R. rubrum*

Table 2.2 presents the capital costs of the major equipment items for H₂ production from synthesis gas using CSTRs as the reactors of choice. Light requirements are not considered. As is noted, the total capital cost is \$6.221 million and the fixed capital investment is estimated as \$18.663 million or \$0.107/MSCF synthesis gas. Table 2.3 shows the projected economics for H₂ production from syngas, with each operating cost shown in dollars per MSCF synthesis gas. The total H₂ cost is estimated at \$0.672/MSCF synthesis gas. Tables 2.4 and 2.5 show similar results when considering the light requirements for *R. rubrum*. The fixed capital investment increases to \$29.6 million and the H₂ cost increases to \$1.30/MSCF synthesis gas.

Tables 2.6-2.9 show results using an immobilized cell reactor with and without light. Without light the cost of H₂ is \$0.459/MSCF syngas (see Table 2.7, and with light the cost of H₂ is \$1.00/MSCF syngas. It is apparent from these analyses that higher rates of CO utilization would improve the economics and that the light requirements for *R. rubrum* cannot be economically supplied by electricity.

2.2 Evaluations with ERIH₂

Engineering Resources, Inc. (ERI) has identified an organism for water gas shift which has much higher rates of CO utilization (and thus H₂ production) and does not require light for conversion. The economics for this bacterium ERIH₂, in CSTRs and ICRs are shown in Tables 2.10-2.13. In the

Table 2.3

Economic Evaluation

Case A: *R. rubrum*, CSTRs, no light

	\$/MSCF Synthesis Gas
Capital	0.107
Chemicals and Energy	
Electricity	0.193
Chemicals and Steam	<u>0.287</u>
Total	0.480
Labor and Supervision	0.010
Maintenance (4% FCI)	0.043
Taxes and Insurance (3% FCI)	0.032
Depreciation	0.107
Total H ₂ Cost	\$ 0.672/MSCF - Synthesis Gas

Table 2.4

Capital Equipment Requirements

Case B: *R. rubrum*, CSTRs, Light Required

Reactors, 6 CSTRs, 1200 m ³ , 17 atm
K _{La} - 177 hr ⁻¹ , 1500 hp for agitation
Cost = \$6,045,000 (1993, delivered)
Medium Pumps, 6, 11 hp, reciprocating
Cost = \$176,000 (1993, delivered)
Light Source, 18277 KW, Change every half year for 10 years
Cost \$3,655,000
Fixed Capital Investment
3 x 9,876,000 = \$29,628,000
= \$0.169/MSCF Synthesis Gas

Table 2.5

Economic Evaluation

Case B: *R. rubrum*, CSTRs, Light Required

	\$/MSCF Synthesis Gas
Capital	0.169
Chemicals and Energy	
Electricity - Agitation and Pumps	0.193
Electricity - Light	0.526
Chemicals and Steam	<u>0.287</u>
	1.006
Labor and Supervision	0.010
Maintenance (4% FCI)	0.068
Taxes and Insurance (3% FCI)	0.051
Depreciation	0.169
Total H ₂ Cost	\$1.304/MSCF Synthesis Gas

Table 2.6

Capital Equipment Requirements

Case C: *R. rubrum*, Countercurrent ICRs, no light

Reactors, 5, 3.5 m diameter, 20 m High, 10 Columns, 17 atm
1 inch porcelain Berl Saddles

Cost - \$2,106,000

Medium Pumps, 5, 30 hp, reciprocating

Cost - \$263,000

Fixed Capital Investment

3 x 2 \$369,000 = \$7,107,000
= \$0.041/MSCF = Synthesis Gas

Table 2.7

Economic Evaluation

Case C: *R. rubrum*, Countercurrent ICRs, no light

Capital	\$/MSCf Synthesis Gas 0.041
Chemicals and Energy	
Electricity - pumps	0.003
Chemicals and Steam	<u>0.287</u>
Total	0.290
Labor and Supervision	0.010
Maintenance (4% FCI)	0.016
Taxes and Insurance (3% FCI)	0.012
Depreciation	0.041
	\$0.459/MSCF = Synthesis Gas

Table 2.8

Capital Equipment Requirements

Case D: *R. rubrum*, Countercurrent ICRs, with light

Reactors, 5, 3.5 m diameter, 20 m height, 17 atm
1 inch porcelain Berl Saddles

Cost = \$2,106,000

Medium pumps, 5, 30 hp, reciprocating

Cost = \$263,000

Light sources, 18207 KW, Change every half year for 10 years

Cost = \$3,665,000

Fixed Capital Investment

3 X 6,034,000 = \$18,102,000
= \$0.103/MSCF Synthesis Gas

Table 2.9

Economic Evaluation

Case D: *R. rubrum*, Countercurrent ICRs, with light

	\$/MSCF Synthesis Gas
Capital	\$0.103
Chemicals and Energy	
Electricity - Pumps	0.003
Electricity - Light	0.526
Chemicals and Steam	<u>0.287</u>
	0.816
Labor and Supervision	0.01
Maintenance (4% FCI)	0.041
Taxes and Insurance (3% FCI)	0.031
Depreciation	\$0.103
	\$1.001/MSCF Synthesis Gas

Table 2.10

Capital Equipment Requirements

Case E: ERIH2, CSTRs

Reactors, 3 CSTRs, 1200m³, 17 atm
 $K_La = 205 \text{ hr}^{-1}$, 500 hp for agitation
 Cost: \$3,031,000 (1992, delivered)

Medium pumps, 3, 11 hp, reciprocating
 Cost: \$88,000 (1992, delivered)

Fixed Capital Investment
 = 3 x \$3,119,000 = \$9,357,000
 = \$0.053/MSCF synthesis gas

Table 2.11
Economic Evaluation
Case E: ERIH2, CSTRs

	\$/MSCF synthesis gas
Capital	0.053
Chemicals and Energy	
Electricity	0.097
Chemicals and Steam	<u>0.072</u>
Total	0.169
Labor and Supervision	0.010
Maintenance (4% FCI)	0.021
Taxes and Insurance (3% FCI)	0.016
Depreciation	0.053
Total H ₂ Cost	\$0.269/MSCF syngas or \$0.436/MSCF H ₂

Table 2.12
Capital Equipment Requirements
Case F: ERIH2

ICR Columns, 2, 3.2 m dia x 20 m high, 17 atm,
packed with 1 in Berl saddles
Cost: \$1,136,000 (1992, delivered)

Medium pumps, 2, 40 hp, reciprocating pumps
Cost: \$98,000 (1992, delivered)

Fixed Capital Investment
 = 3 x \$1,234,000 = \$3,702,000
 = \$0.021/MSCF synthesis gas

Table 2.13
Economic Evaluation
Case F: ERIH2, ICRs

	\$/MSCF synthesis gas
Capital	0.021
Chemicals and Energy	
Electricity	0.002
Chemicals and Steam	<u>0.072</u>
Total	0.074
Labor and Supervision	0.010
Maintenance (4% FCI)	0.008
Taxes and Insurance (3% FCI)	0.006
Depreciation	0.021
 Total H ₂ cost	 \$0.119/MSCF syngas or \$0.193/MSCF H ₂

CSTR, ERIH2 requires a H₂ production cost of \$0.269/MSCF synthesis gas or \$0.436/MSCF hydrogen. In the ICR, ERIH2 requires a H₂ production cost of \$0.119/MSCF syngas or \$0.193/MSCF H₂.

Hydrogen removal from CO₂ was next considered in the economics. There are a number of processes to remove CO₂ from gas streams. The separation is based on one of the following four principles (Eickmeyer *et al.*, 1978):

- (1) CO₂ is weakly acidic, forming H₂CO₃ when dissolved in water. This permits the use of liquid alkaline solutions, either regenerable or nonregenerable, for absorption of CO₂.
- (2) CO₂ is soluble in water and in many organic liquids. Processes based on solubility as opposed to chemical affinity are also used.
- (3) The molecular size and structure of CO₂ permit it to be selectively absorbed on solid adsorbants, particularly molecular sieves.
- (4) The acidic nature or size and structures of CO₂ in some mixtures permit it to be separated by the use of permeable membranes.

The amount of CO₂ removed by solid adsorbants is small compared to that removed by liquid absorbant systems. Permeable membrane separation is of minor commercial importance (Eickmeyer *et al.*, 1978). For the solvent system, the CO₂ content of the liquid phase increases in direct proportion to the partial pressure of CO₂ in the vapor phase. For the chemical reaction system, the equilibrium pressure increases only slightly with loading of the liquid over the lower range of solution loadings, but curves upward sharply as the loading approaches the stoichiometric limit set by the chemical reaction. It is obvious that at high available partial pressures of CO₂, as in this application, the solvent system is capable of attaining a higher loading of the scrubbing solution and, therefore, will require lower solution circulation rates for the same degree of CO₂ removal.

Another important advantage of the solvent system is a regeneration. The ratio of stripping vapor/solution required to strip the solution to a very low residual CO₂ content will be substantially less for the solvent system than for the chemical system. Thus a stripped solvent system will generally be capable of scrubbing the gas to be treated to a lower level of residual CO₂. Compared to a chemical system, the total energy required for the solvent system per mole of CO₂ removed is generally lower and is composed of power for circulating the solution and, in many cases, for refrigeration since the solubility of CO₂ is greatly increased by operating at less than ambient temperature.

Table 2.14 lists the typical operating conditions of various CO₂ removal processes (Eickmeyer *et al.*, 1978). As discussed above, solvent systems involving sulfinol, selexol, rectisol, purisol, and fluror appear to be most suitable. The feed gas in this application will contain about 45 percent CO₂ and a total pressure of 400 psi, or 180 psi CO₂ partial pressure. The majority of these solvents can also simultaneously remove H₂S, and some solvents such as sulfinol and selexol even remove COS. The processes using these solvents have the advantages of a low solvent circulation rate, low plant costs, low utilities costs, high effectiveness for the removal of COS, CS₂ and HS, low solvent degradation and low corrosion rates (Eickmeyer *et al.*, 1978). The major disadvantage of these processes is the requirement of a royalty. Thus, there is not much published information available to estimate CO₂ removal costs.

In this study, the estimation of CO₂ removal costs is based on the only published material and energy balance information for the sulfinol process (Goar, 1979). Figure 2.2 shows the system flow chart of the sulfinol process

Figure 2.1 Typical Equilibrium Lines for Chemical and Physical Solvents
(Eickmeyer, *et al*)

Figure 2.2 System Flow for Sulfinol Process (Goar)

Table 2.14

Typical Operating Conditions of Various CO₂ Removal Processes

Process	Solution Composition	Net Loading (scf/gal)	CO ₂ in Feed Gas by Volume	Operating Conditions		Utilities	
				Pressure (lb/in ² gauge)	Temperature (°F)	Heat	Power
Catacarb	Varies	3-5.5	2-27	100-1200	100-260	Low	Medium
Benfield	K ₂ CO ₃ + promoter	3-5.5	2-27	100-2000	100-260	Low	Medium
Vetrocoke	Varies	Not known	2-27	100-1200	120-212	Low	Medium
Conventional MEA	15-20%	2.5-3.5	0.5-10.0	25-950	80-130	High	High
High-loaded MEA	30-40%	7	15-25	400-900	130-160	Medium	Medium
Conventional DEA	20-25%	2-3	0.5-0.8	175-500	80-130	High	High
High-loaded DEA	25-27%	5-5.4	2.0-20	175-950	80-130	Medium	Medium
SNPA:DEA	25-30%	5.5-6.2	2.0-30	200-2000	80-130	Medium	Medium
DGA	65-70%	5-7	5.0-20	150-950	80-130	Medium	Medium
Sulfinol	Varies	8-16	2.0-30	300-1000	100-160	Medium	Medium
Selexol	DMPEG	11-15	2.0-45	500-1400	100-160	Low	Low-Medium
Rectisol	Methanol	5-20	2.0-35	300-1200	-5 to -100	Low	Medium
Purisol	NMP	5-10	5.0-35	450-1200	40-105	Low	Medium
Fluor	Propylene carbonate		5.0-53				

*/Depends on solubility and recompression of flash gas

Note: NMP: n-methyl 2-pyrrolidone

DMPEG: dimethyl ether of polyethylene glycol

Sulfinol solvent: sulfolane, di-isopropanol amine

and Table 2.15 lists the material and energy balance information. The calculation will be based on Gas "E", since it has exactly the same CO₂ partial pressure as in this application (18 percent in 1000 psi, or 180 psi). The feed gas flow rate in Table 2.15 is 100 MMSCFD, or 18 MMSCFD of CO₂ load. In this application, the feed gas will be 1.6175 x 50 MMSCFD with 45 percent CO₂, or a 36.4 MMSCFD CO₂ load. Thus, the amount treated is roughly double the published data. The lower total pressure and higher CO₂ concentration in the feed should not affect the process operation much since it is CO₂ that is absorbed. Based on the above discussion, the major operation cost will be for the solution circulation pump energy and steam in the regenerator boiler.

Table 2.16 shows the equipment and their costs for CO₂ removal. As is noted, an absorber, regenerators, reboilers and condensers are required, along with associated pumps. The total fixed capital investment is \$11,847,000 or \$0.068/MSCF of synthesis gas. Table 2.17 shows an economic evaluation for CO₂ removal alone. The total CO₂ removal cost is estimated as \$0.369/MSCF of synthesis gas or \$0.598/MSCF of hydrogen.

The revised economics for H₂ production using ERIH₂ including CO₂ removal are shown in Tables 2.18 and 2.19. The total H₂ cost ranges from \$0.772/MSCF of H₂ for the ICR case to \$1.231/MSCF of H₂ for the CSTR case. Thus, the economic projections indicate that H₂ can be produced biologically for 19.3¢/MSCF without H₂ removal and for 77.2¢/MSCF if H₂ separation by the sulfinol process is employed.

Table 2.15

Typical Circulation and Reboiler System Rates
for Sulfinol Process

Item	Rates for Sulfinol Process				
	Gas "A"	Gas "B"	Gas "C"	Gas "D"	Gas "E"
Contactor pressure, psia	1,000	1,000	1,000	1,000	1,000
Feed-gas temp., °F	110	110	110	110	110
Feed-gas flow (dry), MMSCFD	100	100	100	100	100
Feed-gas composition, mole %					
H ₂ S	0.65	20.1	20.10	51.5	0.10
CO ₂	8.73	2.0	2.00	3.5	18.00
N ₂	2.37	1.4	1.40	8.6	0.70
C ₁	87.90	71.5	63.01	25.8	80.94
C ₂	0.35	2.0	8.43	5.8	0.17
C ₃		1.7	3.71	3.2	0.05
C ₄		1.1	0.82	1.6	0.04
C ₅		0.2	0.53		
Total	100.00	100.00	100.00	100.00	100.00
COS, grains/100 scf	3.0	7.3	7.3	8.4	
RSH, grains/100 scf	2.1	1.5	1.5	3.10	
H ₂ S/CO ₂ ratio	0.0744	10.05	10.05	14.71	0.0056
Sweet gas H ₂ S content, grains/100 scf	<0.25	<0.25	<0.25	<0.25	<0.25
Sweet gas CO ₂ content, mole %	<1.0	<1.0	<1.0	<1.0	<2.0
Sweet gas total sulfur, grains/100scf	<1.0	<1.0	<1.0	<1.0	<1.0
Solution circulation rate, gpm at 100°F	1,483	1,748	1,790	2,366	2,167
Solution net pickup, scf acid-gas/gal	4.39	8.78	8.57	16.14	5.24
Reboiler steam rate, lb/hr	69,740	111,440	125,960	156,010	99,600
Reboiler steam rate, lb/gal solution	0.78	1.06	1.17	1.10	0.77

*Includes 0.02 mole % aromatics. The calculated acid gas produced from Gas "C" could contain 1.0 mole % hydrocarbons, which would contain 0.20 mole % of C₅⁺ fraction, based on the H₂S portion only. This acid gas could be processed in a sulfur plant without any further treatment. The Sulfinol system for Gas "E" was designed to leave 2.0 mole % CO₂ in the sweet gas, while reducing the H₂S content to 0.25 grains/100 scf.

Table 2.16
Capital Costs for CO₂ Removal

Absorber	
9.8 min liquid retention time	
Cost: \$568,000	
Regenerator	
9.8 min liquid retention time	
Cost: \$568,000	
Pumps	
7-300 hp reciprocating pumps	
Cost: \$613,000	
Reboiler	
Total area: 55,000 ft ² (5 units of 11,000 ft ² each)	
Cost: \$1,543,000	
Condenser	
Cost: \$657,000	
Fixed Capital Investment	
- 3 x \$3,949,000 = \$11,847,000 = \$0.068/MSCF synthesis gas	

Table 2.17
Economic Evaluation for CO₂ Removal

	\$/MSCF synthesis gas
Capital	0.068
Chemicals and Energy	
Electricity	0.041
Steam	0.191
Chemicals	<u>0.054</u>
Total	0.286
Labor and Supervision	0.010
Maintenance (4% FCI)	0.003
Taxes and Insurance (3% FCI)	<u>0.002</u>
Total CO ₂ Removal Cost	\$ 0.369/MSCF syngas
	or \$ 0.598/MSCF H ₂

Table 2.18

Economic Evaluation Including CO₂ Removal

Case G: ERIH2, CSTRs Utilized

	\$/MSCF synthesis gas
Capital	0.021
Chemicals and Energy	0.455
Labor and Supervision	0.020
Maintenance (4% FCI)	0.024
Taxes and Insurance (3% FCI)	0.018
Depreciation	0.121
Total H ₂ Cost	\$ 0.759/MSCF syngas or \$ 1.231/MSCF H ₂

Table 2.19

Economic Evaluation Including CO₂ Removal

Case H: ERIH2, ICRs Utilized

	\$/MCSF synthesis gas
Capital	0.089
Chemicals and Energy	0.360
Labor and Supervision	0.020
Maintenance (4% FCI)	0.004
Taxes and Insurance (3% FCI)	0.003
Depreciation	0.021
Total H ₂ Cost	\$ 0.476/MSCF syngas or \$ 0.772/MSCF H ₂

3.0 CONCLUSIONS

Based upon the results of these analyses, it is seen that the light requirements for *R. rubrum* cannot be economically satisfied with electricity. Furthermore, higher specific CO uptakes rates are desirable. The bacterium ERIH₂, isolated at Engineering Resources, Inc., does not require light and has very high specific uptakes rates. The economic evaluation with ERIH₂ indicates that H₂ recovery and for 77¢ MSCF if H₂ separation by the sulfinol process is employed.

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