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ALTERNATIVE PROCESS SCHEMES FOR COAL CONVERSION

PROGRESS REPORT NO. 3
JUNE 1 - AUGUST 31, 1979

October 1979

PROCESS SCIENCES DIVISION
DEPARTMENT OF ENERGY AND ENVIRONMENT

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ALTERNATIVE PROCESS SCHEMES FOR COAL CONVERSION

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October 1979

**Work performed for
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Summary

An absorption/stripping process calculation using propane as the absorption solvent for separation of methane from hydrogen and carbon monoxide has been performed. Detailed material and energy balances for the process as well as the dimensions of the absorber and the stripper are reported. Other major pieces of equipment such as heat exchangers, pumps, and compressors were evaluated in order to determine the equivalent electrical energy of the process as approximately 13550 cal(e)/gm-mole methane produced. The purity of methane in the final stream is 96% by volume at 100°F and 1000 psi. The present process appears to be a potential working process for methane separation in large quantity.

Introduction

In coal gasification processes such as the flash hydropyrolysis or the Exxon catalytic process, methane is the major desired end product. However, methane is usually produced along with hydrogen and/or carbon monoxide. To obtain pipeline grade methane, it is necessary to develop economical methods of separating methane from hydrogen and/or carbon monoxide mixtures. The hydrogen and carbon monoxide are then recycled in the process. Several separation technologies such as absorption/stripping, cryogenic, clathrate formation have been examined [1]. In this report, a method of absorption/stripping using propane as the solvent is presented. Absorption/stripping is a well-known chemical engineering operation and can be readily designed and constructed on a large scale. Therefore, the present method can be considered as a potentially useful method.

In the present work, a feed gas mixture of 40% CH₄, 45% H₂, and 15% CO at 100°F and 500 psi is to be separated. A production capacity of 250×10^6 scf/day of CH₄ is assumed. The methane is delivered to the pipeline at 100°F and 1000 psi. For the application of absorption/stripping, it is desirable to utilize a solvent which has a high solubility for methane. Table 1 shows some solubility data for methane in water and in various organic solvents. It is seen that methane has a higher solubility in propane than in any of the other solvents listed. Therefore, propane was chosen as the solvent for the absorption/stripping separation process investigated in this report.

Operating conditions for the absorption and stripping towers are important design parameters for the process. Due to vapor pressure and entrainment, propane will be present in the effluent gas streams from both the absorber and stripper. Usually this quantity of propane is not recovered and is considered an economic loss. The amount of propane in the gas phase is mainly dependent on the operating temperature and pressure of the towers.

Figure 1 shows the vapor pressure of some of the relevant compounds as a function of temperature. At -146°C and -120°C, the vapor pressure of propane is about 0.1 mm Hg and 1 mm Hg respectively. Since the

operating pressures of the absorber and stripper are about 500 psi and 487 psi respectively, the mole fractions of propane in the outlet gas streams of the absorber and stripper are about 6.75×10^{-6} and 1×10^{-4} respectively. When the absorber operates at higher temperature as shown in Table 2, propane losses increase. When the absorber operates at -146°C , the propane loss is equivalent to 0.0850% of the cost of methane assuming the cost of methane and propane is the same. This low percentage of propane loss was considered acceptable. Hence it was determined to operate the absorber at -146°C and the stripper at -120°C .

Process Description

A schematic process flow sheet is shown in Fig. 2. Inlet gas, a mixture of methane, hydrogen and carbon monoxide at 100°F and 500 psi (stream 1) is successively cooled to -140°C by the outlet gas stream from the absorber and some recycle gas. The absorber is a packed column of 1-in. berl saddles with 50% void fraction. The rich liquid from the bottom of the absorber is heat exchanged with the bottom liquid from the stripper. The stripper is also a packed column with 2-in. berl saddles. The dissolved methane, hydrogen, and carbon monoxide is stripped out by heating at the bottom of the stripper. The outlet gas stream from the stripper is heated in a heat exchanger by a recycle gas stream and is further compressed to produce the final methane product at 100°F and 1000 psi.

To perform a further detailed process calculation on this multi-component absorption and stripping process, vapor liquid equilibrium data for methane, hydrogen, and carbon monoxide is needed. Equilibrium K-values of methane and propane are obtained from DePriester [2]. Solubility data for hydrogen in propane is obtained from Trust and Kurata [3]. Solubility of carbon monoxide in propane is obtained from Prausnitz and Shair [4].

The absorber calculation is based on 1 lb-mole of gas stream entering into the bottom of the absorber. The entering liquid, free of absorbing gases, is fed at a rate of 1.5 lb-mole/lb mole of gas. For 95% methane removed, a preliminary estimate of methane absorbed is 0.38 lb-mole, carbon monoxide absorbed is 0.13 lb-mole, hydrogen absorbed is 0.1 lb-mole.

The absorber is operated under isothermal conditions of -146°C . The liquid to gas mass ratio at the top of the tower is 3.85 while at the bottom it is 2.11 with an average value of 2.98. The key component is methane. Using a method of average absorption factor which for methane is 6.478 and 95% methane removal, it is possible to determine the number of theoretical transfer units to be 2. With these conditions, methane in the outlet gas stream of the absorber is 0.002 lb-mole, carbon monoxide is 0.13 lb-mole and hydrogen 0.44 lb-mole. The molar concentrations of the absorber effluent gas are CH_4 , 3.5×10^{-3} ; CO , 0.227; H_2 , 0.769; C_3H_8 , 6.75×10^{-6} . The molar concentrations in the liquid phase are CH_4 , 0.101; CO , 0.005; H_2 , 0.003; C_3H_8 , 0.891.

Absorber design requires the calculation of the quantity $\frac{L}{G} \sqrt{\frac{\rho_G}{\rho_L}}$ which is 0.79 for the present case. At flooding conditions

$$\frac{G^2 \left(\frac{a_p}{\epsilon} \right) (\mu_L)^{0.2}}{\rho_c \rho_G \rho_L} = 0.05$$

and therefore assuming 60% of flooding conditions is 2920 lb/hr ft^2 . Since it is desired to produce $250 \times 10^6 \text{ scf/day}$ methane, the inlet gas flow into the absorber is $2.35 \times 10^6 \text{ lb/hr}$.

From this and the gas mass velocity, it is determined that 3 towers are needed, each one with a diameter of 18.5 ft. The tower height is determined by the height of a transfer unit which is 20.7 ft. Therefore, the height of the packed tower is 41.4 ft. Other detailed tower characteristics are given in Table 3.

The stripping tower is operated at -120°C and 487 psi and the molar ratio of liquid to gas is 9.7. Again taking methane as the key component, and 95% methane stripped, about 0.0962 lb mole of methane, 0.0045 lb-mole of carbon monoxide, 0.002 lb-mole of hydrogen are stripped. The preliminary estimates of the liquid to gas ratio is 26.7 at the top and 24.0 at the bottom with an average of 25.4. The average stripping factor for methane is 21.2 and two transfer units are needed. Under these conditions 0.0046 lb-mole of methane, 0.0032 lb-mole of carbon monoxide,

and 0.00036 lb-mole of hydrogen remain in the bottom liquid phase. In the bottom outlet stream of the liquid phase, mole fractions of the chemical species are CH_4 , 0.0051; CO , 0.0036; H_2 , 0.0004; C_3H_8 , 0.991. The outlet mole fraction of the gas phase from the stripper is CH_4 , 0.960; CO , 0.019; H_2 , 0.021; C_3H_8 , 0.0001.

To design the stripper, again methane is chosen as the key component.

$\frac{L}{G} \sqrt{\frac{\rho_G}{\rho_L}}$ is taken to be 7.81. At flooding

$$\frac{G^2 \left(\frac{a_p}{\epsilon} \right) (\mu_L)^{0.2}}{g_c \rho_g \rho_L} = 0.0018.$$

Therefore, the gas mass velocity is 1372 lb/hr ft^2 for 60% flooding condition with a gas mass flow of $1.28 \times 10^6 \text{ lb/hr}$. It is determined that 3 towers are required each one 17.6 ft in diameter. The height of a transfer unit is a 11.4 ft and the total height of a packed tower is 22.8 ft. Pressure drop in the tower is 1.4 lb/in^2 . A summary of the composition of various streams is given in Table 4 and stripper characteristics are given in Table 3.

A thermal energy balance of the process is performed for the coolers, for streams 1 and 2 in Fig. 2 and the heat exchanger between streams 5 and 7. The heat input for the heat exchanger between the absorber and the stripper is considered to be negligible. The electrical power input to the compressor is 5820 horsepower. The thermal energy load for the heater between streams 13 and 9 in Fig. 2 is $2.2 \times 10^7 \text{ cal/min}$ and for the cooler at the outlet gas stream from the absorber is $5.06 \times 10^7 \text{ cal/min}$. A further breakdown of the thermal energy requirement of the process is given in Table 5. Detailed calculations are given in the Appendix. Combining the thermal energy and the pumping and compressing power required for the process as shown in Fig. 2, the total thermal energy required is 33875 cal/g-mole which is about 12.9 times higher than the minimum ideal separation energy required which is calculated to be $2621 \text{ cal/g-mole CH}_4$.

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Appendix

Energy Calculation for the Separation of Methane from Process Gasification Stream Using Propane Absorption/Stripping

The energy calculations for the process are based on electrical and thermal energy. It is assumed that the heat exchanger between the absorber and the stripper operates ideally so no net heat loss or gain will be realized in this unit. The electrical energy requirements are for (1) the pumps between the absorber and the stripper, (2) the inlet gas blower, (3) the compressor at the end for the product methane, and (4) pumping power for fluid transport. Thermal energy is supplied for (1) the heater between streams 9 and 13, and (2) the cooler at the outlet of the absorber. The calculations will be performed separately in the following sections.

A. Pumping Power Required to Transport Fluid from Bottom of Absorber to Top of Stripper

$$\begin{aligned} \text{Liquid flow rate} &= 8.48 \times 10^8 \text{ lb/day} = \frac{8.48 \times 10^8 \text{ lb/day} \times 7.48 \text{ gal/ft}^3}{36.52 \text{ lb/ft}^3 \times 24 \text{ hr/day}} \\ &= 7.24 \times 10^6 \text{ gal/hr} = 1.21 \times 10^5 \text{ gal/min} \end{aligned}$$

$$\text{Density of liquid} = 36.52 \text{ lb/ft}^3$$

Height of stripping tower packing is 22.8 ft. Consider heading of the tower is 20% additional. Then the tower height ΔZ is 27.4 ft. The fluid is transported by 30 pipes and the diameter of the pipe is about 0.25 ft. The energy balance for each pipe is

$$\frac{g}{g_c} \Delta Z + 0.8 W_e + \frac{2fu^2L}{g_c D} = 0 \quad (\text{A-1})$$

$$\text{Velocity } u \text{ in each pipe} = \frac{8.48 \times 10^8 \times 4}{24 \times 36.52 \times \pi \times 30 \times 0.25^2} = 6.56 \times 10^5 \text{ ft/hr} = 182.4 \text{ ft/sec}$$

$$Re = \frac{uD\rho}{\mu} = \frac{182.4 \text{ ft/sec} \times 0.25 \text{ ft} \times 36.52 \text{ lb/ft}^3}{1.2 \times 10^{-2} \times \frac{36.52}{62.4} \times 2.42} = 9.8 \times 10^4$$

Relative roughness factor and friction factor are taken to be 0.0006 and 0.004, respectively. Therefore, the frictional loss term is

$$lw_f = \frac{2 \times 0.004 \times 182.42 \times 37.4}{32 \times 0.25} = 1236 \text{ ft lb}_f/\text{lb}$$

From Eq. (A-1), we can calculate the loss per set of pipes (10 pipes) connecting each absorber and stripper is

$$W_s = \frac{-(1236 \times 10 + 27.4)}{0.8} = 15484 \text{ ft lb}_f/\text{lb}$$

The total electrical power required for the 30 pipes is

$$P = \frac{15484 \text{ ft-lb}_f/\text{lb} \times 8.48 \times 10^8 \text{ lb/day} \times 5.05 \times 10^{-7} \text{ IP hr/ft-lb}_f}{24 \text{ hr/day}} = 2.76 \times 10^5 \text{ IP}$$

B. Pumping Power Required to Transport Fluid from Bottom of Stripper to Top of Absorber

$$\begin{aligned} \text{Liquid flow rate} &= 8.12 \times 10^8 \text{ lb/day} = \frac{8.12 \times 10^8 \text{ lb/day} \times 7.48 \text{ gal/ft}^3 \times 10^6 \text{ gal/hr}}{36.52 \times \text{lb/ft}^3 \times 24 \text{ hr/day}} \\ &= 1.15 \times 10^5 \text{ gal/min} \end{aligned}$$

Again we take 20% additional height for the height of the stripper which has a packing height of 41.4 ft so the total height of the stripper is 50 ft. Let's apply the same mechanical energy balance Eq. (A-1) to calculate the pumping power here. Velocity of the fluid is

$$U = \frac{8.12 \times 10^8 \times 4}{24 \times 36.52 \times \pi \times 30 \times 0.25^2} = 6.29 \times 10^5 \text{ ft/hr} = 174.9 \text{ ft/sec}$$

$$Re = \frac{uD\rho}{\mu} = \frac{174.9 \times 0.75 \times 36.52 \text{ lb/ft}^3}{1.2 \times 10^{-2} \times \frac{36.52}{62.4} \times 6.72 \times 10^{-4}} = 3.38 \times 10^8$$

Using the relative roughness factor and friction factor to be 0.0006 and 0.004 again. The frictional loss in the pipe is

$$lw_f = \frac{2 \times 0.004 \times 174.9^2 \times 60}{32.2 \times 0.25} = 1824 \text{ ft-lb}_f/\text{lb}$$

So the total energy required per set of pipe (10 pipes) connecting each stripper and absorber is

$$W_s = \frac{-(1824 \times 10 + 50)}{0.8} = 22863 \text{ ft-lb}_f/\text{lb}$$

The total pumping power is

$$P = \frac{22863 \text{ ft-lb}_f/\text{lb} \times 8.12 \times 10^8 \text{ lb/day} \times 5.05 \times 10^{-7} \text{ IP hr/ft-lb}_f}{24 \text{ hr/day}}$$

$$P = 3.9 \times 10^5$$

C. Inlet Gas Blower

Incoming gas to absorber is $2.32 \times 10^7 \text{ ft}^3/\text{day}$. Main flow rate at absorber is 2920 lb/hr ft^2 at a density $\rho_g = \frac{P_{M_{av}}}{Z_{av} RT}$, $T = 230^\circ\text{F} = 127^\circ\text{K}$

$$\text{CH}_4: T_r = \frac{127}{191} = 0.665, P_r = \frac{500}{14.7 \times 45.8} = 0.7426, Z_c = 0.56$$

$$\text{CO} : T_r = \frac{127}{133} = 0.955, P_r = \frac{500}{14.7 \times 34.5} = 0.986, Z_c = 0.32$$

$$\text{H}_2 : T_r = \frac{127}{33.3} = 3.814, P_r = \frac{500}{14.7 \times 12.8} = 2.657, Z_c = 0.98$$

$$Z_{av} = 0.4 \times 0.56 + 0.15 \times 0.32 + 0.45 \times 0.98 = 0.71$$

$$M_{av} = 0.4 \times 16 + 0.15 \times 28 + 0.45 \times 2 = 11.50$$

$$\rho_g = \frac{500 \times 11.5}{0.71 \times 10.73 \times 230} = 3.28 \text{ lb/ft}^3$$

$$\frac{G}{\rho_g} = \frac{2970 \text{ lb/hr ft}^2}{3.28 \text{ lb/ft}^3} = 889.2 \text{ ft/hr} = 0.25 \text{ ft/sec}$$

Number of tubes in absorber = n

$$U = \frac{2.32 \times 10^7 \text{ ft}^3/\text{day} \times 4}{24 \text{ hr/day} \times \pi \times n D^2 \text{ ft}^2} = \frac{1.231 \times 10^6}{n D^2} \text{ ft/hr} = \frac{342}{n D^2} \text{ ft/sec}$$

$$\text{Re} = \frac{u D \rho}{\mu} = \frac{342}{n D^2} \times \frac{D \times 3.28 \text{ lb/ft}^3}{0.007 \text{ cp} \times 6.72 \times 10^{-4}} = \frac{2.38 \times 10^8}{n D}$$

$$lw_f = \frac{2fLU^2}{g_c D} = \frac{2 \times 0.004 \times 10 \text{ ft}}{32.2 D} \times \left(\frac{342}{nD^2} \right)^2 \frac{\text{ft}^2}{\text{sec}^2} = \frac{291}{n^2 D^5} \text{ ft-lb}_f/\text{lb}$$

let $n = 30$, $D = 0.25 \text{ ft}$

Frictional loss/tube = $330.63 \text{ ft-lb}_f/\text{lb}$

$$\frac{g}{g_c} \Delta Z + lw_f + \eta W_s = 0 \quad (\text{A-2})$$

gas flow = $2.32 \times 10^7 \text{ ft}^3/\text{day}$ ($= 5.64 \times 10^7 \text{ lb/day}$)

Pressure drop in tower = 9 in $\text{H}_2\text{O}/\text{ft}$ depth, $\Delta Z = 41.4 \text{ ft}$

$$\text{Total pressure drop} = \frac{9 \times 41.4 \text{ ft H}_2\text{O}}{12} \frac{g}{g_c} = 31.1 \text{ ft-lb}_f/\text{lb}$$

Applying Eq. (A-2) to one set of pipe (10 pipes) for incoming gas into the absorber, one gets

$$(41.4 + 31.1) + 330.63 \times 10 + \eta W_s = 0$$

$$W_s = - \frac{31.1 + 41.4 + 3306.3}{0.8} = - 4223.5 \text{ ft-lb}_f/\text{lb}$$

$$\text{Energy required} = \frac{5.64 \times 10^7 \frac{\text{lb}}{\text{sec}}}{24 \text{ hr/day} \times 3600 \text{ sec/hr}} \times \frac{4223.5}{550} = 5012.8 \text{ IP}$$

D. Compressor at End of Product Methane Between Streams 7 and 8

$$P = \frac{3.03 \times 10^{-5}}{k-1} P_1 q_{f_{m_1}} \left[\left(\frac{P_2}{P_1} \right)^{\frac{k-1}{k}} - 1 \right] \quad (\text{A-3})$$

where $k = \frac{C_p}{C_v} = 1.31$ for methane

For the present case

$$P = \frac{3.03 \times 10^{-5} \times 1.31}{1.31-1} \times 379 \text{ lb/in}^2 \times 144 \text{ in}^2/\text{ft}^2$$

$$\times \frac{4.64 \times 10^6 \text{ ft}^3/\text{day}}{24 \text{ hr/day} \times 60 \text{ min/hr}} \left[\left(\frac{1000}{485} \right)^{\frac{.31}{1.31}} - 1 \right]$$

$$P = 5820 \text{ IP}$$

If the compressor is 80% efficient, then the power required = 7275 IP.

E. Pumping Power for Fluid Transport

Pumping power is required to transport fluid in the streams 9, 10, and 13. Gas flow rate and density in these streams are $7.1 \times 10^6 \text{ lb/day}$ and 0.44 lb/ft^3 , respectively. Using a 0.25 ft diameter pipe, velocity in the pipe is

$$\frac{7.1 \times 10^6 \text{ lb/day} \times 4}{30 \times 0.44 \text{ lb/ft}^3 \times 24 \times 3600 \times \pi \times .25^2} = 127 \text{ ft/sec}$$

$$Re = \frac{uD\rho}{\mu} = \frac{127 \times 2.5 \times 0.44}{(.0183 \times .23 + .009 \times .77) \times \frac{.44}{62.4} \times 6.72 \times 10^{-4}} = 2.65 \times 10^8$$

Relative roughness factor and friction factor are again taken to be 0.0006 and 0.004, respectively. Assuming the total transport line as 600 ft, frictional loss in fluid transport is

$$lw_f = \frac{2fu^2L}{g_c D} = \frac{2 \times 0.004 \times 127^2 \times 600}{32.2 \times 0.25} = 9617 \text{ ft-lb}_f/\text{lb}$$

With 80% efficiency of the pump, total energy loss is $1.2 \times 10^4 \text{ ft-lb}_f/\text{lb}$ and the pumping power required is

$$P = \frac{1.2 \times 10^4 \text{ ft-lb}_f/\text{lb} \times 7.1 \times 10^6 \text{ lb/day} \times 5.05 \times 10^{-7} \text{ IP-hr/ft-lb}_f}{24 \text{ hr/day}} = 1796 \text{ IP}$$

F. Thermal Energy Required in Heat Exchanger Between Streams 13 and 9

$$\begin{aligned} Q &= 7.1 \times 10^6 \text{ lb/day} \times 7 \text{ Btu/lb-mole } ^\circ\text{F} / 7.89 \text{ lb/lb-mole } (110-90) ^\circ\text{F} \\ &= 1.13 \times 10^8 \text{ Btu/day} = 1.3 \times 10^8 \text{ Btu/day} \times 252 \text{ cal/Btu} / 24 \text{ hr/day} \times 60 \text{ min/hr} \\ &= 2.2 \times 10^7 \text{ cal/min} \end{aligned}$$

Thermal energy supplied to the heat exchanger

$$= \frac{2.2 \times 10^7 \text{ cal/min} \times 24 \text{ hr/day} \times 60 \text{ min/hr} \times 16 \text{ g/g-mole}}{3.06 \times 10^7 \text{ lb/day} \times 454 \text{ g/lb}} \\ = 36.5 \text{ cal(t)/g-mole}$$

G. Energy Required to the Cooler in Absorber Outlet Stream

$$(-Q) = 2.23 \times 10^7 \text{ lb/day} \times 7 \text{ Btu/lb-mole } ^\circ\text{F} / 9.73 \text{ lb/lb-mole } (-156 + 146) \times 1.8 \\ = 2.89 \times 10^8 \text{ Btu/day} \\ = 2.89 \times 10^8 \text{ Btu/day} \times 252 \text{ cal/Btu} / 24 \text{ hr/day} \times 60 \text{ min/hr} = 5.06 \times 10^7 \text{ cal/min}$$

Energy supplied to the cooler

$$= \frac{5.06 \times 10^7 \text{ cal/min} \times 24 \text{ hr/day} \times 60 \text{ min/hr} \times 16 \text{ g/g-mole}}{3.06 \times 10^7 \text{ lb/day} \times 454 \text{ g/lb}}$$

It is also interesting to report the theoretical minimum energy requirement for the present process

$$W_{\min, T} = RT \sum_j x_j F \ln \frac{P_2}{x_j F P_1}$$

For the present process, composition temperature and pressure of incoming gas are: CH_4 , 0.4; H_2 , 0.45; CO , 0.15; 100°F and 500 psi respectively. Temperature and pressure of final product gases are at 100°F and 1000 psi. For complete separation

$$W_{\min, T} = 1.98 \text{ Btu/lb-mole } ^\circ\text{Rx560}^\circ\text{R} [0.4 \ln \frac{1000}{0.4 \times 500} + 0.45 \ln \frac{1000}{.45 \times 500} + \\ 0.15 \ln \frac{1000}{.15 \times 500}] = 1889 \text{ Btu/lb-mole} \times 252 \text{ cal/Btu} / 454 \text{ g-mole/lb-mole} \\ = 1049 \text{ cal/mole gas} / 0.4 \frac{\text{mole CH}_4}{\text{mole gas}} = 2621 \frac{\text{cal}}{\text{mole CH}_4}$$

Table 1

The Solubility of Methane in Water and In Organic Solvents
at 25° and at Pressures Up to 140 Atmospheres

Pressure, atm	cc CH ₄ (at 25° and 760 mm) dissolved in 1.0 cc solvent at pressure					
	20	40	60	80	100	140
Water	0.9	1.2	1.8	2.0	2.6	3.3
Methanol ⁽¹⁾	9.0	19.0	29.5	41.0	55.0	--
Ethanol	7.5	14.5	21.5	29.5	38.0	48.0 (120 atm)
Propane	41	83	125	174	200 (90 atm)	--
Butane	37	71	107	144	--	--
Pentane	33	63	97	129	164	--
Hexane	30	59	89	120	--	--
Octane	24	51	80	110	140	--
Cyclohexane	15	33	56	83	--	--
Benzene	11	24	41	59	80	--
Heavy Naphtha	10	20	31	43	--	--
Gas Oil	8	16	23	32	41	62

(1) The solubility of methane in n-propanol, isopropanol, n-butanol and iso-butanol is, within experimental error, the same as in methanol.

Table 2

Propane Losses at Different Absorber Temperatures
 (Stripper Temperature = -120°C)
 Plant Capacity: 250×10^6 SCF/Day

T($^{\circ}\text{C}$)	lb/day loss of Propane			% of Propane Cost Loss Assuming Methane and Propane Cost to be the Same
	Absorber	Stripper	Absorber + Stripper	
- 50	2.79×10^6			
- 60	1.58×10^6			
- 80	4.81×10^5			
-100	1.11×10^5	8415	1.19×10^5	1.133
-128.9	4.77×10^3	8415	1.319×10^4	0.126
-146	4.77×10^2	8415	8892	0.085

Table 3

Summary of Dimensions and Characteristics of
Absorbers and Strippers

		<u>Absorbers</u>
Diameter of Absorber	=	18.5 ft
Number of Absorber	=	3
Height of Packed Tower	=	41.4 ft
Pressure Drop in Packed Tower	=	13.4 psi
Wall Thickness of Tower	=	2.98 in.
		<u>Strippers</u>
Diameter of Stripper	=	17.6 ft
Number of Stripper	=	3
Height of Packed Tower	=	22.8 ft
Pressure Drop in Packed Tower	=	1.4 psi
Wall Thickness of Tower	=	2.35 in.

Table 4

Temperature, Pressure, Flow Rate, and Mole Fraction of Different Numbered Streams in the Flow Diagram

Stream Number	Temp °F	Pressure psi	Flow Rate $\frac{\text{lb}}{\text{day}}$ $\left(\frac{\text{ft}^3}{\text{day}}\right)$	Mole Fraction			
				CH ₄	H ₂	CO	C ₃ H ₈
1	100	500	5.64×10^7 (2.32×10^7)	0.4	0.45	0.15	0
2	-230.8	500	5.64×10^7 (9.5×10^6)	0.4	0.45	0.15	0
3	-230.8	487	2.23×10^7 (8.68×10^6)	0.0035	0.769	0.227	6.75×10^{-6}
4	-230.8	500	8.48×10^8	0.1013	0.0025	0.0051	0.891
5	-184	485	3.06×10^7	0.96	0.021	0.019	0.0001
6	-184	487	8.12×10^8	0.0051	0.0004	0.0036	0.999
7	100	485	3.06×10^7	0.96	0.021	0.019	0.0001
8	100	1000	3.06×10^7 (4.64×10^6)	0.96	0.021	0.019	0.0001
9	110	487	7.1×10^6	0.0035	0.769	0.227	6.75×10^{-6}
10	-110	487	7.1×10^6	0.0035	0.769	0.227	6.75×10^{-6}
11	-100	487	1.77×10^8	0.0035	0.769	0.227	6.75×10^{-6}
12	90	487	2×10^8	0.0035	0.769	0.227	6.75×10^{-6}
13	90	487	7.1×10^6	0.0035	0.769	0.227	6.75×10^{-6}

Table 5

Energy Requirement for the Separation of Methane
by Absorption/Stripping Using Propane

Electrical Energy Requirement

Compressor Power (Horsepower)	= 5820
Pumping Power Required from Bottom of Absorber to Top of Stripper (Horsepower)	= 2.76×10^5
Pumping Power Required from Bottom of Stripper to Top of Absorber (Horsepower)	= 3.9×10^5
Pumping Power of Inlet Gas Blower	(Horsepower) = 5013
Estimated Pumping Power for Fluid Transport	(Horsepower) = 1796

Total Pumping Power/g-mole methane produced	cal(e)/g-mole = 13502
Equivalent Total Thermal Pumping Power/g-mole CH ₄	cal(t)/g-mole = 33755

Thermal Energy Requirement

Thermal Energy Supplied to Heater Between Streams 13 and 9	cal(t)/g-mole = 36.5
Thermal Energy Load to Cooler in Absorber Outlet	cal(t)/g-mole = 84.1

Total Thermal and Equivalent Electrical Energy	cal(t)/g-mole = 33875
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Figure 1. Vapor pressure of selected compounds versus $1/T$

Figure 2. Flow diagram for methane separation by absorption/stripping method.

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