American Institute of Chemical Engineers

STUDENT CONTEST PROBLEM

1982

345 East 47 Street • New York, New York 10017

CONTEST PROBLEM

1982

AMERICAN INSTITUTE OF CHEMICAL ENGINEERS STUDENT CHAPTERS

Open Only to Undergraduates or Those
Without a Degree in Chemical Engineering

DEADLINE FOR MAILING

Solution must be postmarked not later than midnight, June 1, 1982

RULES OF THE CONTEST

Solutions will be graded on (a) substantial correctness of results and soundness of conclusions, (b) ingenuity and logic employed, (c) accuracy of computations, and (d) form of presentation. Accuracy of computations is intended to mean primarily freedom from mistakes; extreme precision is not necessary.

It is to be assumed that the statement of the problem contains all the pertinent data except for those readily available in handbooks and similar reference works. The use of textbooks, handbooks, journal articles, and lecture notes is permitted. In cases where there is disagreement in the data reported in the literature, the values given in the statement of the problem have been chosen as being most nearly applicable.

The problem is not to be discussed with any person whatever until June 1, 1982. This is particularly important in cases where neighboring institutions may not begin the problem until after its completion by another chapter. Submission of a solution for the competition implies adherence to the foregoing condition.

A period of not more than thirty consecutive days is allowed for completion of the solution. This period may be selected at the discretion of the individual counselor, but in order to be eligible for an award a solution must be postmarked not later than midnight, June 1, 1982.

The finished report should be submitted to the chapter counselor within the thirty-day period. There should not be any variation in form or content between the solution submitted to the chapter counselor and that sent to the AIChE office. The report should be neat and legible, but no part need be typewritten.

The solution should be accompanied by a letter of transmittal giving only the contestant's name, school address, home address, and student chapter, lightly attached to the report. This letter will be retained for identification by the Secretary of the Institute. The solution itself must bear no reference to the student's name or institution by which it might be identified. In this connection, graph paper bearing the name of the institution should be avoided.

Each counselor should select the best solution or solutions, not to exceed two, from his chapter and send these by registered mail to

J. C. Forman, Secretary
American Institute of Chemical Engineers
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AIChE STUDENT CONTEST PROBLEM

1982

Methanation Unit Design

INTRODUCTION

The present energy shortage has generated a great deal of interest in producing substitute natural gas (SNG) from coal. This gas can be distributed through the natural gas pipeline network and burned in existing equipment without modification.

The major processing sequence of converting coal to this so-called "substitute natural gas" (SNG) is shown in Figure 1. It starts with coal gasification (partial oxidation), where coal reacts with oxygen and steam to form a raw gas rich in CO and H_2 and having a higher heating value of 200 to 400 Btu/scf. Before it can be upgraded into pipeline-quality gas in the methanation unit, the raw gas is cooled, sent to the shift converter to produce sufficient H_2 for the methanation reaction, and then sent to the gas purification unit where sulfur compounds plus some CO_2 are removed. In the methanation unit, H_2 and CO are converted to methane. Any excess CO_2 may have to be removed so that the remaining gas will have a higher heating value comparable to that of natural gas (about 1,000 Btu/scf). Finally, the gas is compressed, dehydrated, and then delivered to the pipeline.

Your assignment is to develop a process design for the methanation unit in this SNG plant.

THEORY

Methanation of coal-derived synthesis gases involves catalytic conversion of carbon oxides and hydrogen according to the following exothermic reactions:

$$CO + 3H_2 = CH_4 + H_2O \text{ (CO methanation)}$$

$$CO_2 + 4H_2 = CH_4 + 2H_2O \text{ (}CO_2\text{ methanation)}$$

The CO_2 methanation most probably proceeds through an endothermic intermediate state as shown below:

$$CO_2 + H_2 = CO + H_2O \text{ (reverse shift)}$$

The methanation process is usually carried out in two steps: a bulk methanation step and a cleanup methanation step. Most of the carbon oxides are methanated in the bulk methanation step. The product gas from the bulk reactors generally does not meet the heating value and CO content specifications of substitute natural gas. This is because the extent of methanation is limited by the large amount of water by-product. To achieve additional methanation, the gas from the bulk methanator is cooled to condense out the water, then reheated for reaction in the cleanup methanator. The cleanup methanation step uses a lower temperature than the bulk methanation step because this increases the equilibrium constant.

For the purpose of this study, the bulk and cleanup methanators are adiabatic, fixed bed reactors containing alumina- or silica-supported nickel oxide catalysts. Other systems, such as direct-cooled tubular or fluid bed reactors, have been proposed, but have not been commercially proven. There is an upper temperature limit as well as a lower temperature limit at which the methanation catalyst can operate. Below the lower temperature limit, the reactions will not initiate. Above the upper temperature limit, the catalyst sinters and loses its activity. Owing to the high heat of reaction, there is a potential for a large temperature rise in the methanator. The temperature rise can be controlled either by recycling cooled product gas or by injecting steam into the methanator. Both the recycled gas and the steam act as heat carriers for the heat released during methanation. Multiple reactors in series are often suggested for the bulk methanation step. This allows interreactor cooling, which reduces the quantity of recycled gas or steam injection.

Two empirical factors are specified for commercial reactor designs: the "space velocity" and "approach temperature difference." Space velocity is defined as the gas rate in standard cubic feet per hour divided by catalyst volume. Approach temperature difference, or approach ΔT, indicates the degree of departure from chemical equilibrium. The approach ΔT is the difference between the actual reactor outlet temperature and the equilibrium temperature corresponding to the composition of the reactor effluent. The approach ΔT varies with catalyst activity for a given space velocity. At start-of-run, the fresh catalyst is very active and the reaction proceeds to a close approach to equilibrium conversion, or small approach ΔT. The catalyst activity declines with continued use. This results in less conversion or a larger approach ΔT at end-of-run. The approach ΔT usually recommended for CO methanation is 10°F at start-of-run and 50°F at end-of-run.

The reverse shift reaction is considerably faster than CO methanation. Therefore, it has a closer approach to equilibrium in the methanators. The approach ΔT for this reaction is usually assumed to be zero at start and end-of-run.

PROCESS DESCRIPTION

A flow diagram for a methanation plant is shown in Figure 2. Note that this flow diagram does not include all equipment and/or heat exchange services, and the flow sequence shown is not necessarily optimum. It is provided for guidance only.

Purified synthesis gas enters a guard reactor for removal of the last traces of sulfur compounds by reaction with ZnO to protect the methanation catalyst from poisoning. ZnO in the guard reactor is periodically replaced. The minimum temperature required to operate the guard reactor is 450°F.

The sulfur-free feed gas from the zinc oxide reactor then enters the bulk methanator system. Effluent gas from each reactor in the bulk methanation system is cooled before entering the next reactor. Part of the effluent gas from the last bulk methanator is compressed and recycled to join the fresh feed, while the rest of the gas is further cooled to condense out the water. After water separation in the knockout drum, the gas is reheated and sent to the cleanup methanator. The effluent gas from the cleanup methanator is then compressed and dried in a dehydration unit before delivery to the pipeline.

The dehydration unit consists of an absorber and a regenerator. In the absorber, the gas is dehydrated by a glycol solution. The glycol solution is regenerated by steam-stripping and recycled to the absorber.

PROBLEM SCOPE

Your company is considering installation of a coal-based SNG plant and you are a member of the study team assigned to this project. Other team members have studied and selected gasification, syngas shift, and purification sections. Your assignment is to develop an optimum methanation plant design based on the feed gas composition and product specifications listed in Table 1.

Many processing options for a methanation plant will produce a workable design. Any number of bulk methanators with interreactor cooling for heat removal may be used, along with steam injection and/or product gas recycle for temperature moderation.

In this case, previous studies indicate that steam injection need not be considered. The principal focus is on the tradeoff between the number of bulk methanators with interreactor cooling versus the recycle gas rate. A system consisting of one bulk methanator will require a very high recycle gas rate. On the other hand, a series of bulk methanators with interreactor cooling will require a lower recycle gas rate, but will entail higher reactor capital costs and a higher system pressure drop. Thus, there is a best case from the standpoint of capital and operating cost.

In addition, optimization of the various heat exchange services is important. Various process streams may be heated with steam or hot process gas and they may be cooled by steam generation, cool process gas, boiler feed water, or cooling water.

The criterion for selecting the process alternatives should be the lowest present worth of capital cost plus operating costs over 20 years, which yields a mimimum 17 percent rate of return on incremental investment.

Equipment should be designed to handle start-of-run operation with a fresh catalyst and end-of-run operation with reduced catalyst activity. Equipment necessary for startup of the plant should also be included.

Prepare your report in the format described in the Final Report Format. Use the information provided in the Design Guidelines and the Economics Guidelines for your study.

FINAL REPORT FORMAT

A. Cover Letter or Transmittal Document

B. Introduction — Give a concise statement of the problem, covering background and objectives.

C. Summary — Submit a brief description of work performed in your evaluation, and your conclusions and/or recommendations.

D. Technical Information:

1. A process flow diagram of a methanation plant showing the methanation reactors, drums, compressors, heat exchanger arrangement, special equipment for startup and shutdown (if any), and major process controls. The flow diagram should show temperature (°F) and pressure (psia) for each stream.

Assign numbers and names to all equipment and label streams with numbers on the flowsheet. These numbers should be used to identify equipment and streams in any table and discussion.

2. A material balance table showing the following for each stream:

- Temperature (°F)
- Pressure (psia)
- Total flow rate (lb/hr and lb-mole/hr)
- Component flow rate (lb-mole/hr)

The process flow diagram and material balance table should show a consistent case (start-of-run or end-of-run), whichever case controls most of the equipment design.

3. An equipment list for the plant. The list should include the following information for each equipment item in addition to its name and number.

Equipment Type	Minimum Required Information
Compressors	• Pressure and temperature at inlet and outlet • Efficiency • Energy consumption, in brake horsepower • Fluid composition and properties
Heat exchangers	• Duty • Area • Stream temperatures • Temperature difference (corrected) • Pressure at inlet and outlet • Fluid composition and properties • Exchanger type and materials of construction
Pressure vessels	• Inside diameter • Cylindrical section length or height • Shell thickness (inches) • Type of heads • Materials of construction • Catalyst volume and type • Operating pressure, design pressure and temperature
Steam turbine drivers	• Brake horsepower • Rates, conditions, and enthalpies of steam and condensate
Motor drivers	• Horsepower

For each equipment item, determine whether its size is set by the start-of-run or end-of-run conditions as discussed in the theory section. Indicate whether start-of-run or end-of-run controls the equipment size and provide an explanation.

4. A utilities summary showing users (or generators) of utilities, tabulated by equipment. Show the rate of consumption (or generation) for each piece of equipment for both start-of-run and end-of-run.

- Steam (1,000 lb/hr) at each pressure level
- Electricity (kW)
- Cooling water circulation (gpm)
- Boiler feed water (gpm)
- Steam condensate (gpm)
- Steam generator blowdown (gpm)

5. A cost summary for the selected design, listing the capital cost of each piece of equipment and the total installed plant cost. Also include an operating cost summary broken down into utilities, catalyst and chemicals, labor, and maintenance.

6. An explanation for your choice of design. Submit a summary table showing the following for each case examined.

- Number of reactors
- Recycle compressor horsepower
- Total capital costs
- Annual operating cost in current dollars
- 20-year present value of capital and operating costs

E. Appendix

The appendix of the report should include an explanation of all assumptions made and calculation method used. Calculations, graphs, tables, etc., necessary for a complete understanding of your solution should be included.

DESIGN GUIDELINES

Flow rate, composition, and properties of the feed and product gases are specified in Table 1. Other design-related information is in Tables 2 to 5.

For the methanator mass and heat balance calculations, actual reactor outlet gas composition versus actual reactor outlet temperature is provided in Table 3. These compositions were computed using the equilibrium constant data from Table 2 and the indicated feed pressures and approach ΔT as in Table 3. At a given reactor outlet temperature and pressure, the reactor outlet composition is a function of the atomic ratios, percent inerts, and approach ΔT. Since atomic ratios are unaffected by chemical reactions, recycling some of the product gas to the feed does not affect the atomic ratios as long as no material is removed from the recycle stream. Tables 3a and 3b can be used for bulk methanators with and without recycle. Removal of water from bulk methanator products changes the atomic ratio. Tables 3c through 3f provide data for the cleanup methanation calculations. Neglect pressure effects for our range of operation. Linear interpolations of these tables will give valid results.

For this study, neglect corrections to gas enthalpy due to pressure.

Compressors are conventional centrifugal type. The maximum discharge temperature is 450°F. The adiabatic efficiency is 75 percent. Maximum compression ratio is 3.5.

Compressors can be driven by either electric motors or steam turbines. Motor efficiency is 90 percent. Maximum moisture content for the steam turbines at isentropic exhaust conditions is 16 percent. Efficiency of steam turbines is given in Figure 5.

In determining equipment operating pressures, use 0.5 psi pressure drop for the connecting piping between two pieces of equipment.

Use the following general guidelines for vessel design:

- Maximum diameter is 16 feet because of shop fabrication and transportation limitations
- For vessel wall thickness and weight calculations, see Reference 1; pp. 6-91 to 6-105. Use same allowable stress for low alloy steel as carbon steel.
- Design pressure (psig) = (1.1 × operating pressure in psia) − 14.7,
  or
  50 psi + operating pressure in psig, whichever is greater
- Design temperature = 50°F + operating temperature (600°F, minimum)
- ⅛ inch for corrosion allowance
- Total vessel weight = 1.1 × (weight of cylindrical section + weight of heads). This factor accounts for nozzles, flanges, skirts, and other parts of the vessel.

Use the following for general basis for reactor design:

- Calculate the catalyst bed pressure drop for the methanator and ZnO guard reactors using the correlation given by Leva for incompressible fluids (Reference 1, p. 5-52). Both the catalyst shape factor and the modified friction factor are 1.0. Use the gas density at reactor outlet temperature for the average density.
- Assume the pressure drop for nozzles, distributor, and supports is equivalent to 3 feet of bed height
- All reactors should be axial downflow vertical vessels

The methanator sizing basis and configuration are as follows:

- Reactor space velocity is 15,000 scfh of total reactor feed gas per cubic foot of catalyst
- Minimum catalyst bed depth is 0.5 × bed diameter
- If hemispherical heads are used, the catalyst bed and supports may extend into the vessel heads up to 15 percent of the vessel diameter. If ellipsoidal heads are used, the catalyst bed may not extend into the heads.

The ZnO reactor sizing basis and configuration are as follows:

- Assume that 1 lb of ZnO adsorbs 0.15 lb of H_2S
- Set amount of ZnO for 2-year life, minimum
- Use minimum ZnO bed depth of 0.5 × bed diameter
- Parallel reactors may be used to minimize pressure drop.

In designing condensate knockout drums:

- Use vertical cylindrical vessels
- Size the drum to settle out water droplets greater than 300 microns. See reference 1, pp. 5-61 to 5-62
- Allow 5 minutes of liquid residence time above the bottom tangent line
- Provide 3 feet above the inlet nozzle to the top tangent line for disengagement space

For all heat exchange equipment, assume the following:

- The thermal resistance of the tube wall is 0.0004 hr-ft²-°F/Btu
- The fouling factor on each side of the tube is 0.001 hr-ft²-°F/Btu for gas and BFW services, 0.0005 for steam generation and condensation, and 0.002 for cooling water
- Gas side Δp = 5 psi

For the waste heat boilers, assume the following:

- The boilers are of the kettle type
- Blowdown is 2 weight percent of boiler feedwater
- $h_{boiling}$ = 1,000 Btu/hr-ft²-°F (referred to outside tube area)
- h_{gas} = 125 Btu/hr-ft²-°F (referred to outside tube area)

For the gas-gas heat exchangers, assume the following:

- h_{gas} = 125 Btu/hr-ft²-°F (referred to outside tube area)

For the water-gas exchangers, assume the following:

- Water is on the tube side
- Maximum cooling water temperature rise is 40°F
- h_{water} = 500 Btu/hr-ft²-°F (referred to outside tube area)
- h_{gas} = 100 Btu/hr-ft²-°F (referred to outside tube area)

For selection of material of construction, use the following criteria for gas streams:

Temperature, °F	Mol% H_2	Material
>600	all	low alloy steel
<600	<5	carbon steel
<600	>5	low alloy steel

Use carbon steel for the shells of waste heat boilers and water-cooled gas coolers.

The dehydration unit does not need to be taken into account in this study because it is essentially the same for all alternative cases.

ECONOMIC GUIDELINES

Alternative designs are to be compared on the basis of the 20-year present value of capital costs plus operating costs. This method is often used in making comparisons of different designs.

No alternative design that produces less than a 17 percent rate of return on incremental capital cost should be chosen. The rate of return on incremental investment is defined as follows: (Reference 2, Chapter 9)

$$ROR = \frac{S}{I} \times 100\%$$

where

ROR = rate of return on incremental investment
I = incremental capital cost
S = present worth of savings (over 20 years) resulting from an incremental investment of I dollars

The present worth of operating, or direct, costs is calculated using the following assumptions:

- Time value of money = 11%/year
- Cost escalation rate = 8%/year
- Interest compounded annually
- Expenses paid annually, at end of year
- Estimated plant life = 20 years (zero salvage value)
- 330 operating days per year

To estimate the installed capital cost for the methanation plant, you may use a "factored estimate." A factor is applied to the purchased cost of each major equipment item (vessels, reactors, compressors, exchangers, etc.) to account for costs associated with: (1) Piping, insulation, instruments, electrical supply, foundations, etc., and (2) Engineering, construction labor, and supervision. Based on experience, a factor of 4 is applicable to this plant.

Operating costs for utilities, catalysts, and chemicals can be computed directly for each alternative design. Other annual operating costs are assumed as:

- Maintenance: 4% of installed capital cost
- Operating labor: $600,000/yr regardless of design case
- Insurance, property taxes, administration: 6% of installed capital cost

Purchase costs of major equipment are given in Figures 3 through 5. Utility costs are given in Table 6. Catalyst costs are given in Table 4.

NOMENCLATURE

MM	= 10^6
HHV	= higher heating value; heat of combustion of 1 scf of gas to gas products and liquid water at 60°F and 1 atmosphere
scf	= standard cubic feet at 60°F and 1 atmosphere. 1 lb mole = 379.5 scf
scfd	= standard cubic feet per day
scfh	= standard cubic feet per hour
h	= film heat-transfer coefficient based on outside tube area, Btu/hr-ft²-°F

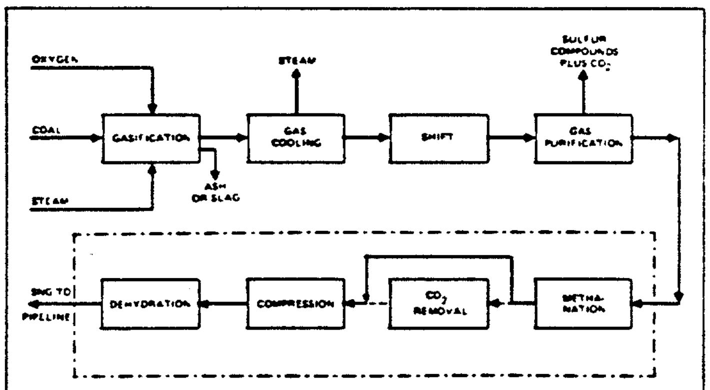
Table 1

FEED AND PRODUCT SPECIFICATIONS

Item	Feed	Product
Gas composition, mol %		
CH_4	15.60	
CO	16.50	0.1 mol % max
CO_2	4.15	3.0 mol % max
H_2	63.40	5.0 mol % max
N_2	0.35	
H_2O	NIL	4 lb/MMscf max
H_2S	0.1 ppm	—
Rate	390 MMscfd	
Pressure	360 psia	1,000 psia
Temperature	65°F	120°F max
HHV, Btu/scf		950 min

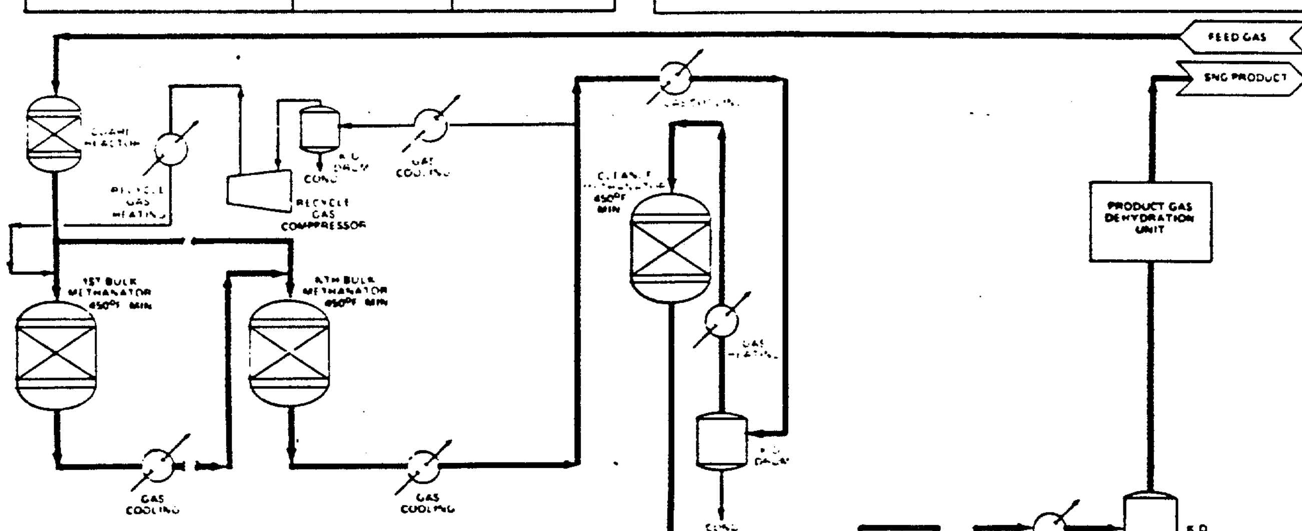
REFERENCES

1. Perry, R. H., and C. H. Chilton, *Chemical Engineers' Handbook*, 5th Ed., McGraw Hill, New York, 1973.
2. Peters, M. S., and K. D. Timmerhaus, *Plant Design and Economics for Chemical Engineers*, 3rd Ed., McGraw-Hill, 1980.
3. United Catalyst Inc., *Physical and Thermodynamic Properties of Elements and Compounds*.
4. Moeller, F. W., H. Roberts, B. Britz, "Methanation of Coal Gas for SNG," *Hydrocarbon Processing*, April 1974.
5. Allen, D. W., and W. H. Yen, "Methanator Design and Operation," *Chemical Engineering Progress*, January 1973.



NOTE —·— DEFINES THE SCOPE OF PROCESS DESIGN FOR THIS PROBLEM

Figure 1 TYPICAL PROCESSING SEQUENCE FOR COAL CONVERSION TO SNG (SUBSTITUTE NATURAL GAS)



NOTES

- K.O. DRUM - KNOCKOUT DRUM
- MINIMUM REACTOR INLET TEMPERATURE IS 450°F (SEE TABLE 4)
- COND - CONDENSATE
- USE THIS FLOW SHEET AS A GUIDANCE BUT NOT AS A LIMITATION TO YOUR OPTIMIZATION EFFORT. FOR EXAMPLE, HEAT EXCHANGERS CAN BE ADDED OR ELIMINATED FROM THE GIVEN FLOW SHEET

Figure 2

PROCESS FLOW DIAGRAM
METHANATION OF COAL DERIVED
SYNTHESIS GAS FOR THE SNG PRODUCTION

Table 2
GAS PROPERTIES

Temperature, °F	Water (gas) H_2O			Hydrogen (gas) H_2		Nitrogen (gas) N_2		Carbon Monoxide (gas) CO			Carbon Dioxide (gas) CO_2			Methane (gas) CH_4		
	Mol. wt = 18.015 b p = 212°F m p = 32°F d_{l1} = 62.3 lb/cu ft at 68°F T_c = 705.4°F P_c = 3206 psia			Mol. wt = 2.0159 b p = −423 F m p = −434.6 F* d_{l1} = 4.42 lb/cu ft at −423 F T_c = −400 F P_c = 188 psia		Mol. wt = 28.013 b p = −320.5 F m p = −346.0 F* d_{l1} = 50.4 lb/cu ft at −320.4 F T_c = −232.8 F P_c = 492.3 psia		Mol. wt = 28.011 b p = −312.7 F m p = −337.1 F* d_{l1} = 50.8 lb/cu ft at −319 F T_c = −218.2 F P_c = 514 psia			Mol. wt = 44.010 b p = −109.3 F (sublimes) m p = −69.2 F (5.2 atm) d_{l1} = 68.7 lb/cu ft at −34.6°F T_c = 88 F P_c = 1073 psia			Mol. wt = 16.043 b p = −258.7 F m p = −296.5 F* d_{l1} = 25.9 lb/cu ft at −263.2 F T_c = −116.5 F P_c = 637 psia		
	$H° - H°_0 + \Delta H°_{f0}$ Btu/lb mole	$\log_{10} K_f$	$C°_p$ Btu/lb mole °F	$H° - H°_0$ Btu/lb mole	$C°_p$ Btu/lb mole °F	$H° - H°_0$ Btu/lb mole	$C°_p$ Btu/lb mole °F	$H° - H°_0 + \Delta H°_{f0}$ Btu/lb mole	$\log_{10} K_f$	$C°_p$ Btu/lb mole °F	$H° - H°_0 + \Delta H°_{f0}$ Btu/lb mole	$\log_{10} K_f$	$C°_p$ Btu/lb mole °F	$H° - H°_0 + \Delta H°_{f0}$ Btu/lb mole	$\log_{10} K_f$	$C°_p$ Btu/lb mole °F
32	−98,890	43.9218	8.001	3,334	6.814	3,417	6.959	−45,546	25.8214	6.959	−165,509	75.4005	8.596	−24,834	10.0893	8.296
60	−98,667	41.4322	8.015	3,526	6.871	3,612	6.960	−45,351	24.6811	6.962	−165,265	71.3461	8.768	−24,605	9.3263	8.440
77	−98,530	40.0470	8.026	3,643	6.887	3,730	6.960	−45,233	24.0479	6.965	−165,115	69.0915	8.870	−24,462	8.8985	8.537
100	−98,346	38.3051	8.039	3,802	6.905	3,890	6.962	−45,073	23.2517	6.968	−164,909	66.2567	9.010	−24,264	8.3547	8.665
200	−97,538	32.1369	8.122	4,496	6.953	4,587	6.976	−44,375	20.4419	6.990	−163,979	56.2342	9.565	−23,370	6.4122	9.306
300	−96,720	27.5823	8.231	5,193	6.977	5,286	7.004	−43,674	18.3757	7.032	−162,997	48.8521	10.058	−22,402	4.9584	10.033
400	−95,890	24.0794	8.359	5,891	6.987	5,988	7.047	−42,968	16.7917	7.091	−161,969	43.1866	10.499	−21,362	3.8206	10.811
500	−95,048	21.3002	8.498	6,590	6.993	6,696	7.107	−42,255	15.5379	7.167	−160,899	38.6999	10.891	−20,241	2.9006	11.604
600	−94,192	19.0410	8.646	7,289	7.002	7,410	7.180	−41,534	14.5200	7.256	−159,792	35.0585	11.243	−19,041	2.1393	12.389
700	−93,319	17.1669	8.799	7,991	7.014	8,133	7.262	−40,804	13.6770	7.352	−158,652	32.0451	11.560	−17,764	1.4996	13.149
800	−92,431	15.5878	8.958	8,694	7.030	8,863	7.351	−40,063	12.9668	7.450	−157,481	29.5103	11.846	−16,413	0.9543	13.877
900	−91,526	14.2380	9.120	9,400	7.052	9,603	7.440	−39,313	12.3590	7.548	−156,283	27.3460	12.111	−14,991	0.4837	14.569
1000	−90,612	13.0713	9.288	10,107	7.080	10,352	7.530	−38,554	11.8341	7.643	−155,060	25.4793	12.344	−13,501	0.0726	15.227
1100	−89,669	12.0520	9.456	10,817	7.112	11,109	7.618	−37,784	11.3750	7.733	−153,814	23.8507	12.560	−11,947	−0.2896	15.850
1200	−88,714	11.1545	9.627	11,530	7.151	11,875	7.705	−37,007	10.9707	7.820	−152,550	22.4188	12.754	−10,331	0.6117	16.438
1300	−87,741	10.3574	9.800	12,248	7.193	12,650	7.785	−36,221	10.6113	7.901	−151,266	21.1483	12.930	− 8,658	0.9002	17.000
1400	−86,753	9.6453	9.971	12,970	7.241	13,433	7.861	−35,427	10.2895	7.977	−149,964	20.0147	13.088	− 6,929	1.1581	17.513
1500	−85,750	9.0050	10.139	13,697	7.292	14,223	7.932	−34,626	9.9997	8.044	−148,648	18.9965	13.234	− 5,163	1.3901	17.995
1600	−84,727	8.4261	10.305	14,429	7.344	15,021	8.000	−33,818	9.7373	8.109	−147,318	18.0768	13.366	− 3,349	1.6008	18.444
1700	−83,687	7.9003	10.467	15,167	7.399	15,825	8.063	−33,004	9.4990	8.167	−145,975	17.2425	13.489	− 1,471	1.7930	18.878
1800	−82,631	7.4197	10.626	15,911	7.455	16,634	8.120	−32,184	9.2800	8.223	−144,618	16.4805	13.599	439	1.9693	19.279
1900	−81,562	6.9800	10.779	16,660	7.512	17,449	8.175	−31,359	9.0793	8.274	−143,253	15.7840	13.701	2,397	−2.1313	19.648
2000	−80,477	6.5758	10.928	17,414	7.570	18,269	8.224	−30,529	8.8944	8.322	−141,879	15.1434	13.797	4,372	2.2799	19.985
2100	−79,378	6.2028	11.061	18,174	7.629	19,093	8.271	−29,694	8.7236	8.366	−140,492	14.5530	13.885	6,394	2.4169	20.306
2200	−78,264	5.8574	11.210	18,939	7.685	19,922	8.313	−28,856	8.5651	8.406	−139,100	14.0072	13.960	8,434	2.5438	20.595

Reference 3

*triple point

Notes:
1. The standard state of gaseous elements and compounds shown in this table is that of an ideal gas at atmospheric pressure.
2. $(H° - H°_0 + H°_{f0}) = h^1$ is a combined ideal gas enthalpy and heat of formation term that can be used to compute the heat of reaction at any given temperature.
   $\Delta h^T = (\Sigma x_i h^1_i)$ products $- (\Sigma x_i h^1_i)$ reactants where x_i is the mole fraction of component i; Δh^T is the heat of reaction at any given temperature.
3. K_f is the equilibrium constant of formation from the elements. $\log_{10} K_f$ is zero for an element.
4. $C°_p$ is the ideal gas specific heat at constant pressure. $(C_p)_{mixture} = \Sigma x_i C_{pi}$ where x_i is the mole fraction of component i; C_{pi} is the specific heat of component i.

Table 3a

BULK METHANATION — REACTOR EFFLUENT COMPOSITIONS

PRESSURE (PSIA)	APPROACH (DEGF)	ATOM RATIO (C/H)	ATOM RATIO (C/O)
320.0	10.0	.192	1.462

COMPONENT MOL PERCENT

	H2O	H2	N2	CO	CO2	CH4	TOTAL
FEED COMPOSITION	0.00	63.40	.35	16.50	4.15	15.60	100.00

REACTOR EFFLUENT TEMPERATURE AND COMPOSITION

T-(DEGF)	H2O	H2	N2	CO	CO2	CH4	TOTAL
760	35.42	5.23	.57	.03	2.38	56.37	61.67
770	35.24	5.50	.57	.04	2.44	56.22	61.76
780	35.01	5.82	.57	.04	2.52	56.04	61.86
790	34.79	6.15	.56	.05	2.59	55.86	61.97
800	34.57	6.47	.56	.06	2.66	55.68	62.08
810	34.34	6.80	.56	.06	2.74	55.49	62.18
820	34.12	7.13	.56	.07	2.81	55.31	62.29
830	33.89	7.45	.56	.08	2.88	55.13	62.40
840	33.67	7.78	.56	.09	2.96	54.94	62.51
850	33.40	8.17	.56	.11	3.04	54.72	62.64
860	33.14	8.56	.56	.12	3.12	54.50	62.77
870	32.87	8.95	.56	.14	3.21	54.28	62.91
880	32.60	9.34	.56	.15	3.29	54.05	63.04
890	32.33	9.74	.55	.17	3.37	53.83	63.18
900	32.06	10.15	.55	.20	3.46	53.59	63.32
910	31.74	10.61	.55	.22	3.55	53.33	63.49
920	31.50	10.96	.55	.24	3.62	53.12	63.61
930	31.19	11.43	.55	.27	3.71	52.85	63.78
940	30.86	11.92	.55	.31	3.80	52.56	63.96
950	30.57	12.34	.55	.34	3.88	52.31	64.12
960	30.28	12.78	.54	.38	3.96	52.06	64.28
970	29.94	13.28	.54	.43	4.05	51.75	64.47
980	29.60	13.80	.54	.48	4.14	51.45	64.66

Table 3b

BULK METHANATION — REACTOR EFFLUENT COMPOSITIONS

PRESSURE (PSIA)	APPROACH (DEGF)	ATOM RATIO (C/H)	ATOM RATIO (C/O)
320.0	50.0	.192	1.462

COMPONENT MOL PERCENT

	H2O	H2	N2	CO	CO2	CH4	TOTAL
FEED COMPOSITION	0.00	63.40	.35	16.50	4.15	15.60	100.00

REACTOR EFFLUENT TEMPERATURE AND COMPOSITION

T-(DEGF)	H2O	H2	N2	CO	CO2	CH4	TOTAL
760	34.41	6.70	.56	.05	2.73	55.56	62.15
770	34.15	7.08	.56	.06	2.81	55.35	62.27
780	33.93	7.40	.56	.06	2.88	55.17	62.38
790	33.70	7.72	.56	.07	2.96	54.98	62.49
800	33.41	8.16	.56	.08	3.05	54.74	62.63
810	33.18	8.48	.56	.09	3.13	54.56	62.74
820	32.92	8.87	.56	.11	3.21	54.34	62.87
830	32.66	9.25	.56	.12	3.29	54.12	63.00
840	32.35	9.69	.55	.14	3.39	53.87	63.15
850	32.09	10.08	.55	.16	3.47	53.65	63.29
860	31.82	10.48	.55	.17	3.55	53.42	63.43
870	31.51	10.93	.55	.20	3.65	53.17	63.59
880	31.24	11.33	.55	.22	3.73	52.93	63.73
890	30.97	11.74	.55	.25	3.81	52.70	63.88
900	30.65	12.20	.55	.28	3.90	52.43	64.05
910	30.29	12.74	.54	.31	4.00	52.11	64.24
920	30.01	13.16	.54	.35	4.08	51.87	64.40
930	29.64	13.70	.54	.39	4.18	51.54	64.60
940	29.35	14.15	.54	.43	4.25	51.28	64.77
950	29.01	14.65	.54	.48	4.34	50.98	64.96
960	28.66	15.17	.54	.54	4.43	50.66	65.17
970	28.31	15.71	.54	.59	4.51	50.34	65.38
980	28.00	16.18	.53	.65	4.59	50.05	65.57

Note: Product composition is shown in mole percent. The "total" amount is the moles of product per 100 miles of feed.

Table 3c

CLEANUP METHANATION — REACTOR EFFLUENT COMPOSITIONS

PRESSURE (PSIA)	APPROACH (DEGF)	ATOM RATIO (C/H)	ATOM RATIO (C/O)
300.0	10.0	.243	7.800

COMPONENT MOL PERCENT

	H2O	H2	N2	CO	CO2	CH4	TOTAL
FEED COMPOSITION USED AS BASIS -	.56	14.42	.81	.22	5.01	78.99	100.00

REACTOR EFFLUENT TEMPERATURE AND COMPOSITION

T-(DEGF)	H2O	H2	N2	CO	CO2	CH4	TOTAL
560	7.91	.65	.87	.00	1.85	88.71	92.99
570	7.88	.72	.87	.00	1.86	88.67	93.02
580	7.85	.78	.87	.01	1.88	88.62	93.05
590	7.81	.85	.87	.01	1.89	88.58	93.08
600	7.78	.91	.87	.01	1.91	88.53	93.11
610	7.73	.99	.87	.01	1.92	88.47	93.15
620	7.69	1.07	.87	.01	1.94	88.42	93.19
630	7.65	1.16	.87	.01	1.96	88.36	93.23
640	7.60	1.24	.87	.01	1.98	88.30	93.27
650	7.55	1.34	.87	.02	2.00	88.23	93.31
660	7.50	1.44	.86	.02	2.02	88.16	93.36
670	7.44	1.54	.86	.02	2.05	88.08	93.41
680	7.39	1.64	.86	.03	2.07	88.01	93.46
690	7.33	1.76	.86	.03	2.09	87.92	93.52
700	7.27	1.87	.86	.03	2.12	87.85	93.57
710	7.21	1.99	.86	.04	2.14	87.76	93.63
720	7.13	2.13	.86	.05	2.17	87.65	93.70
730	7.08	2.24	.86	.05	2.19	87.57	93.76
740	7.00	2.39	.86	.06	2.22	87.47	93.83
750	6.92	2.54	.86	.07	2.25	87.36	93.91
760	6.85	2.69	.86	.08	2.28	87.24	93.98
770	6.77	2.85	.86	.09	2.31	87.13	94.06
780	6.70	2.99	.86	.10	2.33	87.02	94.14
790	6.61	3.17	.86	.12	2.37	86.88	94.23

Table 3d

CLEANUP METHANATION — REACTOR EFFLUENT COMPOSITIONS

PRESSURE (PSIA)	APPROACH (DEGF)	ATOM RATIO (C/H)	ATOM RATIO (C/O)
300.0	10.0	.236	5.176

COMPONENT MOL PERCENT

	H2O	H2	N2	CO	CO2	CH4	TOTAL
FEED COMPOSITION USED AS BASIS -	5.72	13.67	.77	.21	4.75	74.89	100.00

REACTOR EFFLUENT TEMPERATURE AND COMPOSITION

T-(DEGF)	H2O	H2	N2	CO	CO2	CH4	TOTAL
560	12.91	.83	.82	.00	1.80	83.64	93.45
570	12.87	.91	.82	.00	1.81	83.59	93.49
580	12.83	.98	.82	.00	1.83	83.54	93.52
590	12.77	1.07	.82	.00	1.85	83.48	93.56
600	12.72	1.16	.82	.01	1.87	83.42	93.61
610	12.67	1.25	.82	.01	1.89	83.36	93.65
620	12.61	1.36	.82	.01	1.92	83.28	93.70
630	12.56	1.45	.82	.01	1.94	83.22	93.74
640	12.50	1.57	.82	.01	1.97	83.14	93.80
650	12.43	1.69	.82	.01	2.00	83.06	93.86
660	12.36	1.81	.82	.01	2.02	82.97	93.92
670	12.29	1.93	.81	.02	2.05	82.89	93.98
680	12.21	2.09	.81	.02	2.09	82.79	94.05
690	12.41	2.21	.81	.02	2.11	82.70	94.11
700	12.05	2.37	.81	.03	2.15	82.59	94.19
710	11.97	2.51	.81	.03	2.18	82.50	94.26
720	11.88	2.68	.81	.04	2.22	82.38	94.34
730	11.79	2.84	.81	.04	2.25	82.27	94.42
740	11.68	3.03	.81	.05	2.29	82.13	94.52
750	11.59	3.19	.81	.05	2.33	82.02	94.60
760	11.51	3.36	.81	.06	2.36	81.91	94.68
770	11.39	3.57	.81	.07	2.41	81.76	94.79
780	11.29	3.75	.81	.08	2.44	81.63	94.88
790	11.18	3.96	.81	.09	2.48	81.48	94.98

Note: Product composition is shown in mole percent. The "total" amount is the moles of product per 100 miles of feed.

Table 3e

CLEANUP METHANATION — REACTOR EFFLUENT COMPOSITIONS

PRESSURE (PSIA)	APPROACH (DEGF)	ATOM RATIO (C/H)	ATOM RATIO (C/O)
300.0	50.0	.243	7.800

COMPONENT MOL PERCENT

	H2O	H2	N2	CO	CO2	CH4	TOTAL
FEED COMPOSITION USED AS BASIS -	.56	14.42	.81	.22	5.01	78.99	100.00

REACTOR EFFLUENT TEMPERATURE AND COMPOSITION

T-(DEGF)	H2O	H2	N2	CO	CO2	CH4	TOTAL
560	7.74	.98	.87	.01	1.92	88.48	93.14
570	7.69	1.06	.87	.01	1.94	88.43	93.18
580	7.65	1.14	.87	.01	1.96	88.37	93.22
590	7.60	1.23	.87	.01	1.98	88.31	93.26
600	7.55	1.34	.87	.01	2.01	88.23	93.31
610	7.50	1.42	.86	.01	2.02	88.17	93.35
620	7.45	1.52	.86	.02	2.05	88.10	93.40
630	7.40	1.62	.86	.02	2.07	88.03	93.45
640	7.33	1.75	.86	.02	2.10	87.94	93.51
650	7.27	1.85	.86	.02	2.12	87.86	93.56
660	7.20	1.99	.86	.03	2.15	87.77	93.63
670	7.14	2.11	.86	.03	2.18	87.68	93.69
680	7.08	2.23	.86	.04	2.20	87.59	93.75
690	7.00	2.37	.86	.05	2.23	87.49	93.82
700	6.93	2.51	.86	.05	2.26	87.39	93.89
710	6.85	2.66	.86	.06	2.29	87.28	93.96
720	6.78	2.81	.86	.07	2.32	87.17	94.03
730	6.70	2.96	.86	.08	2.35	87.06	94.11
740	6.62	3.11	.86	.09	2.37	86.95	94.19
750	6.53	3.30	.86	.10	2.41	86.80	94.29
760	6.46	3.44	.86	.12	2.43	86.69	94.36
770	6.36	3.63	.85	.13	2.47	86.55	94.46
780	6.28	3.80	.85	.15	2.49	86.42	94.55
790	6.19	3.98	.85	.17	2.52	86.28	94.65

Table 3f

CLEANUP METHANATION — REACTOR EFFLUENT COMPOSITIONS

PRESSURE (PSIA)	APPROACH (DEGF)	ATOM RATIO (C/H)	ATOM RATIO (C/O)
300.0	50.0	.236	5.176

COMPONENT MOL PERCENT

	H2O	H2	N2	CO	CO2	CH4	TOTAL
FEED COMPOSITION USED AS BASIS -	5.72	13.67	.77	.21	.4.75	74.89	100.00

REACTOR EFFLUENT TEMPERATURE AND COMPOSITION

T-(DEGF)	H2O	H2	N2	CO	CO2	CH4	TOTAL
560	12.68	1.25	.82	.00	1.90	83.36	93.65
570	12.62	1.35	.82	.01	1.92	83.29	93.70
580	12.56	1.45	.82	.01	1.94	83.22	93.75
590	12.50	1.56	.82	.01	1.97	83.15	93.80
600	12.43	1.68	.82	.01	2.00	83.07	93.85
610	12.37	1.80	.82	.01	2.02	82.99	93.91
620	12.30	1.92	.81	.01	2.05	82.90	93.97
630	12.21	2.07	.81	.01	2.09	82.80	94.04
640	12.14	2.21	.81	.02	2.12	82.71	94.11
650	12.06	2.34	.81	.02	2.15	82.61	94.17
660	11.97	2.50	.81	.02	2.18	82.51	94.25
670	11.89	2.65	.81	.03	2.22	82.40	94.32
680	11.79	2.84	.81	.03	2.26	82.28	94.41
690	11.70	2.99	.81	.03	2.29	82.17	94.49
700	11.61	3.15	.81	.04	2.33	82.06	94.57
710	11.49	3.37	.81	.05	2.38	81.91	94.68
720	11.39	3.56	.81	.05	2.42	81.77	94.77
730	11.28	3.76	.81	.06	2.46	81.64	94.87
740	11.17	3.95	.81	.07	2.50	81.50	94.97

Note: Product composition is shown in mole percent. The "total" amount is the moles of product per 100 miles of feed.

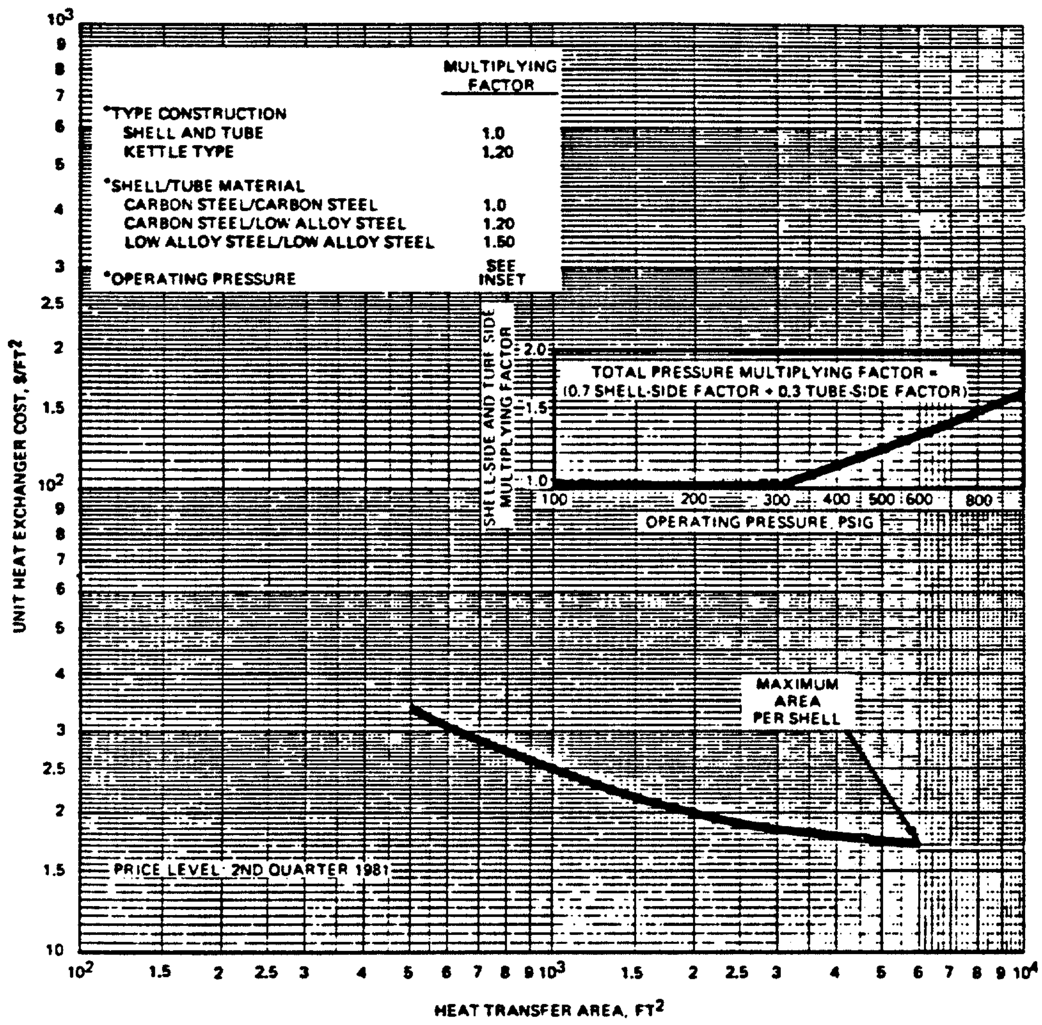


Figure 3 UNIT HEAT EXCHANGER COST VS HEAT TRANSFER AREA

Table 4

CATALYST PROPERTIES

Item	Bulk Methanation	Cleanup Methanation	ZnO
Form	Tablet	Tablet	Extrusion
Size	1/4" ¢ x 1/4"	1/4" ¢ x 1/4"	3/16" ¢ x 3/16"
Operating temperature range, °F	450-900	450-850	450-800
Void fraction	0.4	0.4	0.4
Life, yr	2	2	-
Bulk density, lb/ft^3	60	50	70
Cost, \$/ft^3	400	520	110

Table 5

UTILITIES AVAILABLE

Type	Condition
Steam: High pressure	600 psig, saturated
Medium pressure	150 psig, saturated
Low pressure	50 psig, saturated
Boiler feedwater	250°F
Cooling water	80°F
Ambient air	90°F

Table 6

UTILITY COSTS

Electricity	$0.06/kWh
600 psig steam, sat'd	$8.00/1,000 lb
150 psig steam, sat'd	$7.00/1,000 lb
50 psig steam, sat'd	$6.00/1,000 lb
Cooling water circulation	$10/gpm/yr
Steam turbine condensate credit (100°F)	$0.20/1,000 lb
Condensate credit (250°F)	$1.00/1,000 lb
Boiler feedwater (250°F)	$1.00/1,000 lb

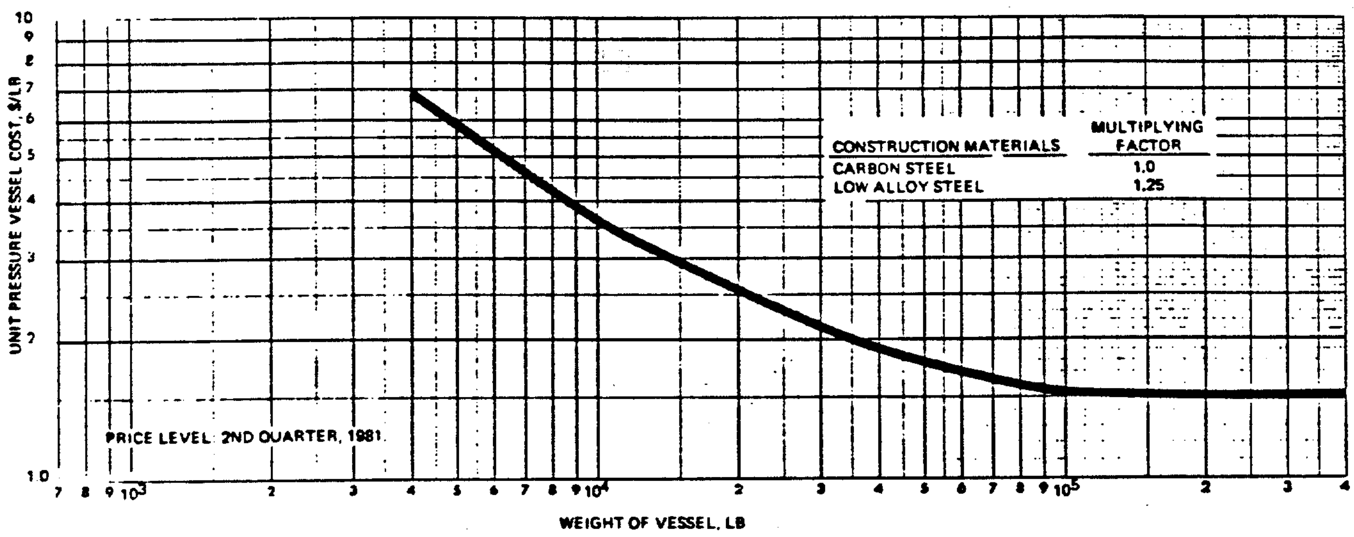
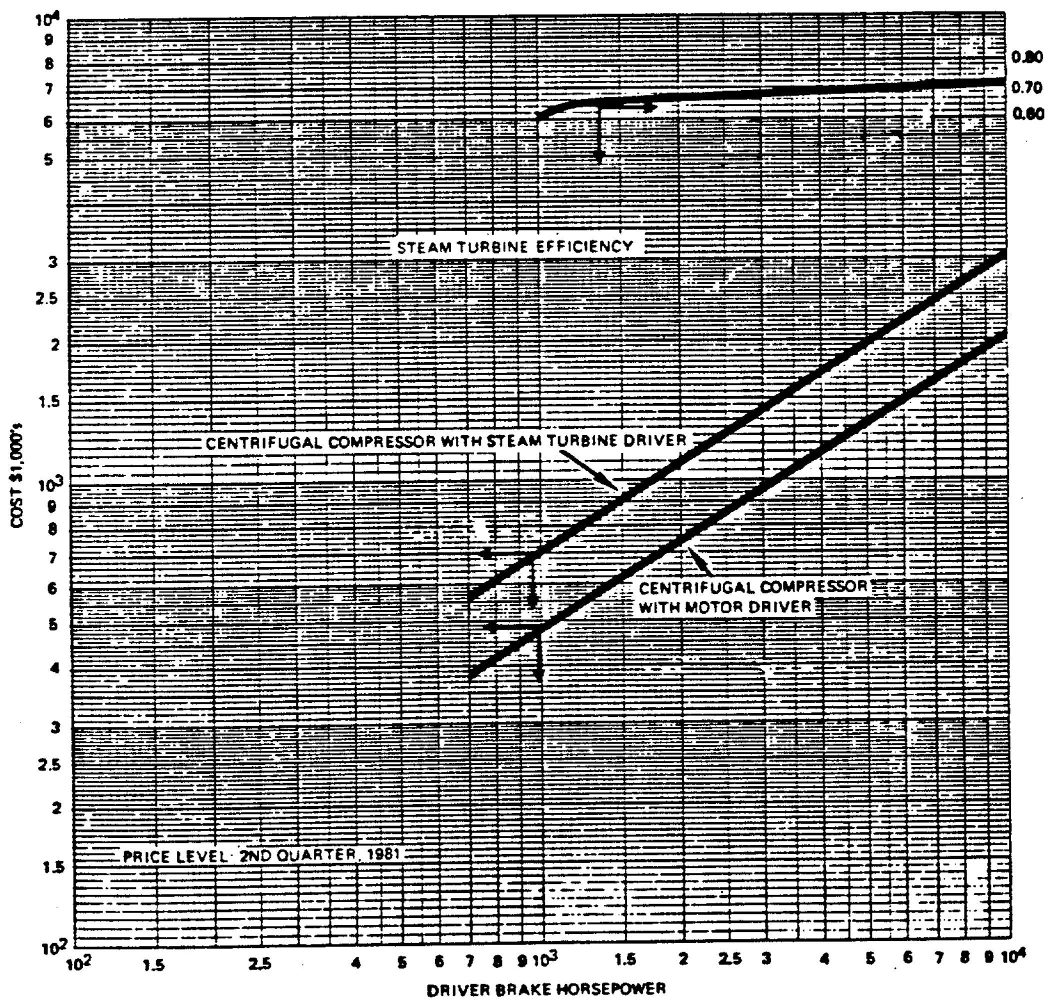


Figure 4 UNIT PRESSURE VESSEL COST VS VESSEL WEIGHT



Note: Maximum Compressor Capacity is 100,000 ACFM at Suction.

Figure 5 COST OF COMPRESSORS, INCLUDING DRIVER AND AUXILIARY EQUIPMENT; STEAM TURBINE EFFICIENCY