



**American Institute of
Chemical Engineers
STUDENT CONTEST
PROBLEM
1987**

DESIGN DATA (including simplifying assumptions)

Material Balance

The material balance should be based on 2.5×10^6 gram-moles per hour of feed gas.

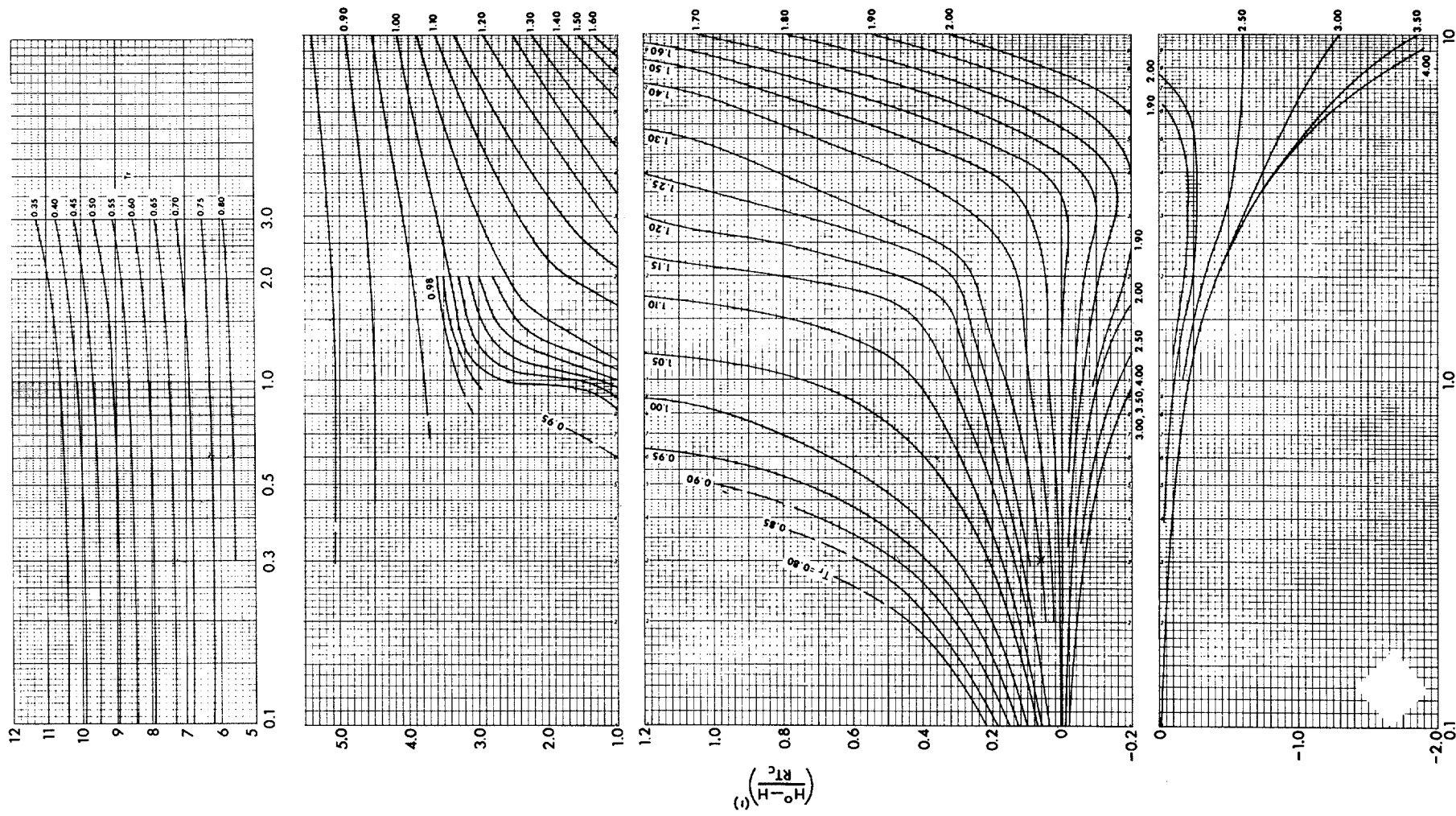
General Design Information and Guidelines

Data or data sources supplied below provide the only basis for the calculations and evaluation of solutions:

- Steam tables for water and steam are given in pages 3-238 through 3-244 in the Chemical Engineers' Handbook [1].
- Circulating cooling water is available at 30°C and may be heated to a maximum of 40°C;
- The following tables in the Chemical Engineers' Handbook [1], may be used to obtain:
 - Heats of Combustion: Table 3-207
 - Thermodynamic Properties of Organic Compounds for cooling or heating:
 - Table 3-222 n-Butane, Saturated
 - Table 3-223 n-Butane, Superheated
 - Table 3-236 Ethane, Saturated
 - Table 3-237 Ethane, Superheated
 - Table 3-256 Methane, Saturated
 - Table 3-257 Methane, Superheated
 - Table 3-272 Propane, Saturated

Source: Reproduced by permission of Gas Processors Suppliers Association, "Engineering Data Book," 9th Edition, 1972

Figure 4: Effect of Pressure on Enthalpy (Correction for Real Fluids)



Where:

- C_c = allowance for corrosion, in. (use 0.125 in. for carbon steel, zero for stainless steel)
- E_j = efficiency of joints expressed as a fraction (use $E_j = 1.0$)
- P = maximum allowable internal pressure (design pressure), psig
- r_1 = inside radius of the shell, before corrosion allowance is added, in.
- S = maximum allowable working stress, psi
- t = minimum wall thickness, in.

Notes:

- Use the same allowable stress (13,700 psi) for carbon steel and stainless steel.
- Design pressure = $1.1 \times$ operating pressure
- Total vessel weight = $1.2 \times$ (weight of cylindrical section). This factor accounts for heads, nozzles, flanges, skirts, and other parts of the vessel.

l) For sizing distillation reflux drums use horizontal cylindrical vessels. Size the drum for 10 minutes liquid residence time when 60% of the volume is filled with liquid. Assume a length to diameter ratio of 3.

m) For sizing separators use equation 18-148 on page 18-75 of the Chemical Engineers' Handbook [1]. Use a vertical cylindrical vessel and allow 5 minutes liquid residence above the bottom tangent line. Provide 1.5 meters above the inlet nozzle to the top tangent line for disengagement space.

n) For each pump assume a pump efficiency of 60% and an electric motor efficiency of 90%. Provide a spare for each pump service.

Expander and Compressor

- Assume that the inlet temperature to the expander is minus 29°C for all cases. Also assume that expander inlet pressure is 7,000 kPa for all cases.
- Assume an adiabatic expander efficiency of 80% for the expander. Assume an adiabatic compressor efficiency of 75% for both the compressor driven by the electric expander and the compressor driven by the electric motor. Assume an electric motor efficiency of 92%.
- Provide cooling of the discharge of the expander/compressor discharge so that the inlet temperature of the sales gas compressor is 45°C.
- Adjust the expander-driven compressor discharge pressure to match power available from the expander.
- For an initial approximate estimate of expander temperature drop use the following formula:

$$T_2 - T_1 = T_1[(P_2/P_1)^{(K-1)/K} - 1] \alpha_{ad}$$

Where:

- α_{ad} = adiabatic efficiency (use $\alpha_{ad} = 0.80$)
- T_1 = suction temperature, K
- T_2 = discharge temperature, K
- P_1 = suction pressure, kPa
- P_2 = discharge pressure, kPa
- K = adiabatic exponent (C_p/C_v)

Assume:

$$(K - 1)/K = 0.25 \text{ for initial approximation}$$

Temperature Range, °C

Carbon steel +343 to -45
304 stainless steel -46 to -255

- Use the equation shown below to calculate pressure vessel wall thickness for cylindrical shells. Calculate vessel weight using a metal specific gravity of 7.83 (7830 kg/m³ = 489 pounds/ft³ = 0.283 pounds/in.³). Recommended design equation for cylindrical vessels under internal pressure:

$$t = \frac{Pr_i}{SE_j - 0.6P} + C_c$$

Physical Properties

Relevant transport properties may be obtained for the pure components from the Chemical Engineers' Handbook [1]. Critical properties of the pure compounds are found in Table 1.

Vapor-Liquid Equilibrium Data

High pressure (7,000 kPa) K-value data ($K_i = Y_i/X_i$) for methane, ethane, propane, and n-butane are found in Table 2. These data should be used for performing flash calculations at or near 7,000 kPa upstream or downstream of the demethanizer. Low pressure K-value data are found in Tables 3, 4 and 5 at pressures of 1,034, 2,069, and 2,759 kPa respectively and may be used in flash calculations near these pressures. Many methods exist to use K-values to evaluate single-stage vapor-liquid equilibrium and partitioning; methods suggested by Holland [3] or Treybal [4] are acceptable procedures.

For this problem, distillation calculations for the demethanizer and de-ethanizer should utilize the isobaric T-H-X-Y data for the binaries methane-ethane and ethane-propane which are tabulated in Tables 6 through 12. Other data should be interpolated or extrapolated from these data.

General Distillation Column Design Assumptions

- At the client's request, use sieve (perforated) trays for the demethanizer column and for the de-ethanizer column.
- Assume 61 cm tray spacing.
- While a 0.7 kPa pressure drop across each tray in each column is typical, assume *no pressure drop* across the tower to simplify the separation calculations.
- Figure 18-10 in the Chemical Engineers' Handbook [1] may be used to determine the flood velocity for sieve tray calculations. Use 80% of flood as a maximum velocity. For these calculations, surface tensions of liquid hydrocarbons are summarized in Table 13.
- Assume 80% active tray area.
- Assume the following overall column efficiencies for calculating actual number of trays:
 - Demethanizer — 55%
 - De-ethanizer — 60%

The number of calculated theoretical and actual trays should each be rounded up to the nearest integer.

- Assume two meters above the top tray and $(0.06 \times \text{number of trays} + 2)$ meters below the bottom tray when calculating column height.
- Assume kinetic and potential energy terms are negligible when calculating the energy balances.
- Use only the two key components to calculate each distillation separation. Any special theory to perform the separation calculations should be discussed, otherwise a specific literature citation is adequate.
- For the demethanizer, the cold expander outlet (vapor and liquid) enters onto the top tray. The chilled liquid from the chilled gas separator enters onto an intermediate tray to be optimally defined. *The number of theoretical trays is fixed at ten for the demethanizer.*
- For the de-ethanizer, feed is located at the optimum tray. Reflux ratio is fixed at $1.15 \times$ minimum reflux

ratio, thus the total number of trays may vary between cases.

- Use all of the stream components for calculating condenser and reboiler duty and for determining top and bottom column temperatures. Use the enthalpy data base discussed in section titled "Curl Pitzer Data Base."

Equipment Pressure Drop

Assume the following pressure drops:

- Heat exchangers
 - Shell side: negligible
 - Tube side: negligible
- Distillation columns: negligible
- Separators and drums: negligible
- Piping: negligible

Note:

For detailed design of equipment and piping, pressure drops would be considered. For this preliminary evaluation, these pressure drops are to be neglected.

Heat Transfer Coefficients

The following overall heat transfer coefficients may be assumed:

System	Overall Heat Transfer Coefficient $W/(m^2 \cdot K)$	Btu/(h · ft ² · °F)
Shell & Tube Exchangers		
Feed/Cold Gas Exchanger	340	60
Feed Gas Chiller	450	80
Sales Gas Cooler	450	80
Sales Gas Intercooler	280	50
Demethanizer Reboiler	450	80
De-ethanizer Reboiler	800	140
De-ethanizer Condenser	680	120

Conversion Factor: $W/(m^2 \cdot K) \times 0.17623 = \text{Btu}/(\text{h} \cdot \text{ft}^2 \cdot \text{°F})$

Curl-Pitzer Data Base

The Curl-Pitzer enthalpy and entropy departure data base [2] based on corresponding states should be used for the expander calculations and for calculating the inlet and outlet enthalpies of the tower streams; this is needed to calculate the tower condenser or reboiler duties. For these stream enthalpies, conversion from a pseudocomponent to an actual component mixture will frequently require resetting a dew or bubble point temperature; this should be done first using the low pressure K-values, and then the stream enthalpy is estimated using reduced temperature and pressure properties.

The molar-weighted average (also known as Kay's rule) of ideal gas enthalpy, critical temperature, critical pressure, and acentric factor should be used to estimate the pseudocritical properties, enthalpy departures and real enthalpy of the stream mixture.

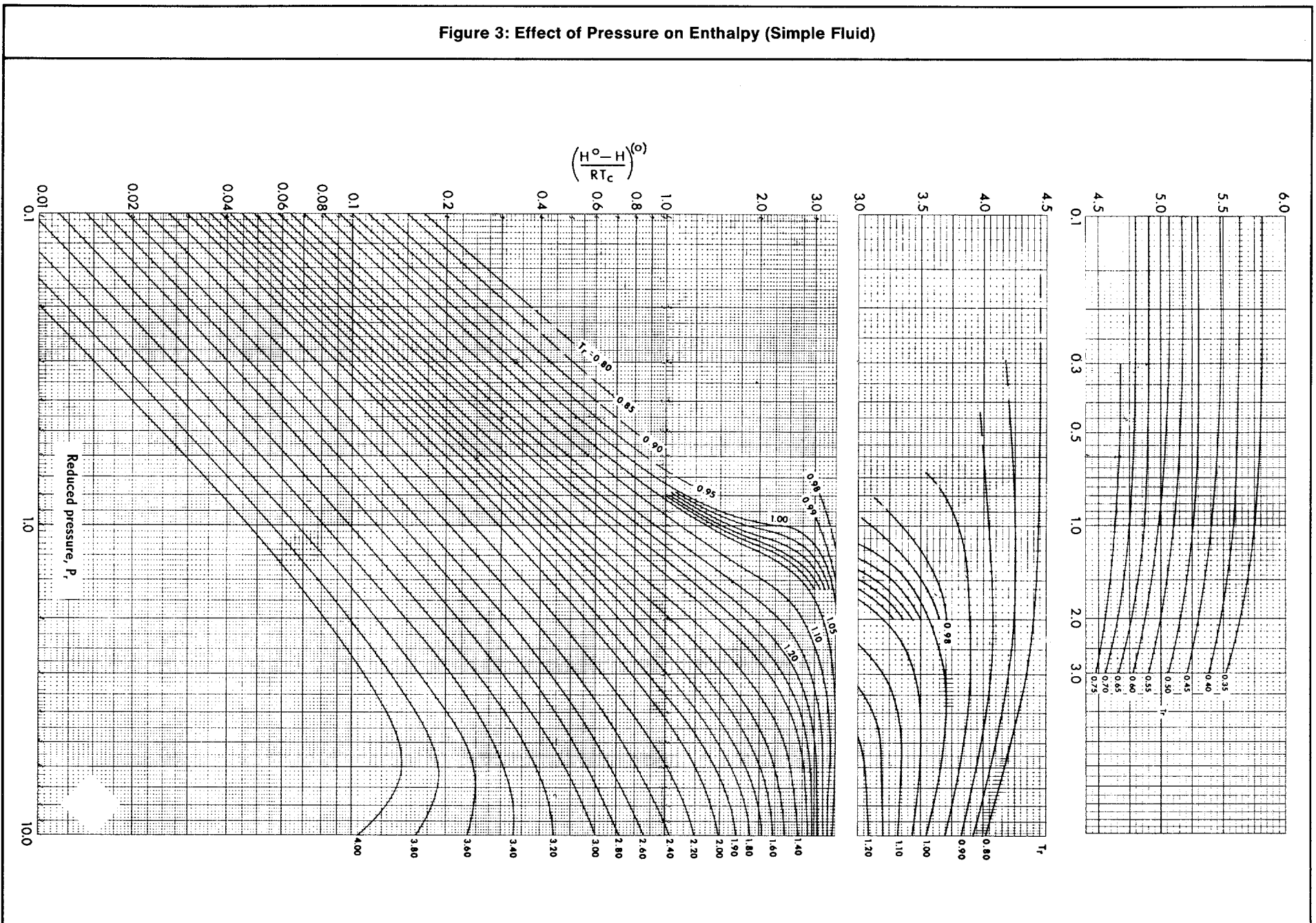
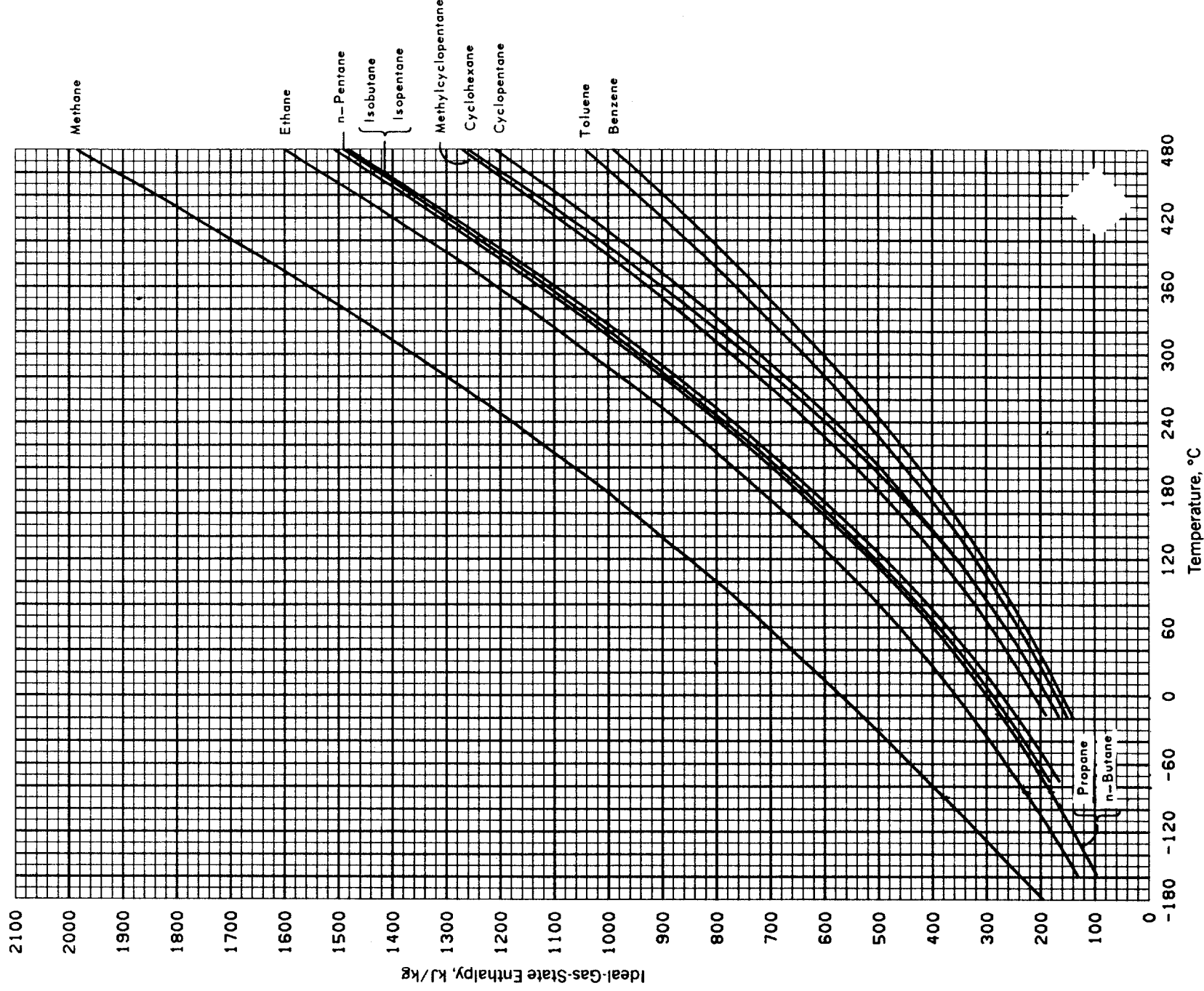


Figure 3: Effect of Pressure on Enthalpy (Simple Fluid)

Figure 2: Ideal-Gas-State Enthalpy of Pure Components



Source: Reproduced by permission of Gas Processors Suppliers Association, "Engineering Data Book," 9th Edition, 1972

With an enthalpy datum of zero gaseous enthalpy at 0 K and 0 kPa, the enthalpy of a mixture, H , is given by:

$$\left(\frac{H^\circ - H}{RT_c}\right)_{\text{mix}} = \left(\frac{H^\circ - H}{RT_c}\right)^{(1)} + \omega_m \left(\frac{H^\circ - H}{RT_c}\right)^{(1)}$$

Where:

- 1) H° , the ideal gas enthalpy for the mixture, is estimated from molar weighted pure component enthalpies found in Figure 2;
- 2) The simple fluid enthalpy departure term, i.e., $[(H^\circ - H)/RT_c]^{(0)}$ is found in Figure 3, as a function of the mixture reduced temperature and pressure.
- 3) The real fluid enthalpy departure term, i.e., $[(H^\circ - H)/RT_c]^{(1)}$ is found in Figure 4 as a function of the mixture reduced temperature and pressure.

Capital Costs

General

The unit equipment prices shown in these sections should be used to estimate purchased equipment costs. The data should be plotted on log-log graph paper to obtain interpolated values.

The fixed plant capital cost should be estimated by multiplying the sum of the purchased equipment costs by a Lang factor of 3.5.

Heat Exchangers

(Carbon steel floating head)

Surface Square Feet	Dollars Per Square Foot		Cost Multiplier for 304 Stainless Steel Tubes
	1,000 lb/in. ² to 450 lb/in. ² Pressure	450 lb/in. ² and lower Pressure	
100	60	50	1.15
300	37	31	1.28
600	28	23	1.38
1,000	23	20	1.45
2,000	18	15	1.58
3,000	16	13	1.64
4,000	14	12	1.70

Entropy, S , with a datum of an ideal gas at 101.325 kPa and 0 K is calculated from:

$$\left[\frac{(S^\circ - S)}{R}\right]_{\text{mix}} = \left[\frac{(S^\circ - S)}{R}\right]^{(0)} + \omega \left[\frac{(S^\circ - S)}{R}\right]^{(1)} + \ln(P/P_{\text{ref}})$$

Where:

- 1) the ideal gas entropy for the mixture is calculated from:
 $S^\circ_{\text{mix}} = \sum Y_i S_i^\circ - R \sum (Y_i \ln Y_i)$ where the pure component, ideal gas entropy, S_i° , is given in Figure 5;
 - 2) P_{ref} is 101.325 kPa; P is system pressure, kPa;
 - 3) the simple fluid departure function, $[(S^\circ - S)/R]^{(0)}$ is given by Figure 6;
 - 4) the real fluid departure function, $[(S^\circ - S)/R]^{(1)}$ is given by Figure 7;
 - 5) ω = acentric factor of mixture;
 - 6) R = gas constant
- Since entropies will be needed in the expander calculations, a Mollier diagram for methane is provided in Figure 8 to assist in the initial guesses for the adiabatic calculations used with the Curl-Pitzer data base.
 For the methane rich sales gas, Figure 8 should be used for all compressor calculations.

ECONOMIC DATA

The economic data supplied in this section are to be used for the economic evaluation. Non-SI units may appear in these tables because prior experience is with pre-

vious English speaking clients. All costs are based on 1987 dollars. All evaluations should be done in terms of constant 1987 dollars; that is, ignore inflation.

Column Trays

Tray Costs — 304 stainless steel sieve trays (includes shop installation).

Tray Diameter, Feet	Dollars per Tray
2.0	300
3.0	460
5.0	860
7.0	1,470

Vessels and Columns

Weight, Pounds	Dollars/Pound Carbon Steel
600	7.00
1,000	6.00
2,000	4.50
4,000	3.50
6,000	3.00
10,000	2.30
20,000	1.80
40,000	1.45

Multiply carbon steel cost by 1.6 to obtain cost for 304 stainless steel.

**Pumps — Carbon Steel
(with Electric Motor Drivers)**

Horsepower	Dollars/Horsepower
5	680
10	450
30	230
50	170
70	130

**Sales Gas Centrifugal Compressor with
Electric Motor Drive**

Brake Horsepower	Dollars/Horsepower
500	390
1,000	355
1,500	330
2,000	320
3,000	315

Combined, Integrated Expander-Compressor

Horsepower	Dollars/Horsepower
500	440
1,000	420
1,500	400
2,000	390
3,000	380

Gas Processing Costs

Feed and Product Costs and Credits in 1987 Dollars

Costs	
Natural Gas Feed	\$2.85/GJ (lower heating value)
Product Credits	
Sales Gas to Pipeline	\$2.85/GJ (lower heating value)
Ethane Product	\$0.18/kilogram
Propane & Butane Stream	\$0.20/kilogram
Purchased Utility Costs	

The costs shown below are the total cost of purchasing utilities from a nearby plant.

Circulating Cooling Water	\$0.025 per cubic meter
Electricity	\$0.05/kWh
Steam (1,000 kPa) (includes condensate return credit)	\$7.70/thousand kilograms
Propane Refrigerant @ -34°C	\$24.85/GJ
@ -5°C	\$9.50/GJ

(Note: k = 10³, M = 10⁶, G = 10⁹)

Basis for Discounted Cash Flow Analysis

The following assumptions are to be used for calculating the discounted cash flow rate of return on equity capital:

Item	Basis
Cost estimate	1987 dollars
Common equity capital	100% (zero debt)
Rate of inflation (escalation)	0% per year
Plant schedule	—Start design at the beginning of 1988 —Start commercial plant operation at 100% design flow at the beginning of 1990
Schedule of fixed capital cost expenditures	—1988 50% —1989 50%
Operating plant life	15 years (1990 to 2004)
Method of depreciation	Straight line for 15 years (starting in the beginning of 1990 and terminating at the end of 2004)
Combined federal and state income tax rate	50%
Salvage value	zero
Organization & startup expense	5% of fixed plant capital cost (50% in 1988 and 50% in 1989)
Annual onstream time	8,400 hours/year (350 days/year)
Land cost	5% of fixed plant capital cost (100% at beginning of 1988. Recovered as cash in 2004).
Working capital	10% of the following: total annual value of products minus cost of feed gas and utilities. (One time expenditure at beginning of 1990. Recovered as cash in 2004).
Annual maintenance expense	3% of fixed plant capital cost
Annual property taxes and insurance	2% of fixed plant capital cost
Operating labor (2 operators/shift, total 8 operators)	\$320,000/year
Payroll charges	\$96,000/year
General & administrative expense	1.0% of fixed plant capital cost

Section 25 of the Chemical Engineer's Handbook [1] explains the method for calculating discounted cash flow rate of return. The discounted cash flow rate of return will be the interest rate at which the cumulative sum of the annual net discounted cash flows over 17 years (2 years design and construction plus 15 years operation) equals zero. Interest is compounded annually. The working capital and land cost are not depreciated and are not deductible as an expense. They are considered liquid assets that are recovered as cash at the end of the last year of operation. Assume that organization and startup costs can be treated as a business expense.

Figure 1: Process Flow Diagram of Gas Processing Plant for Ethane Recovery

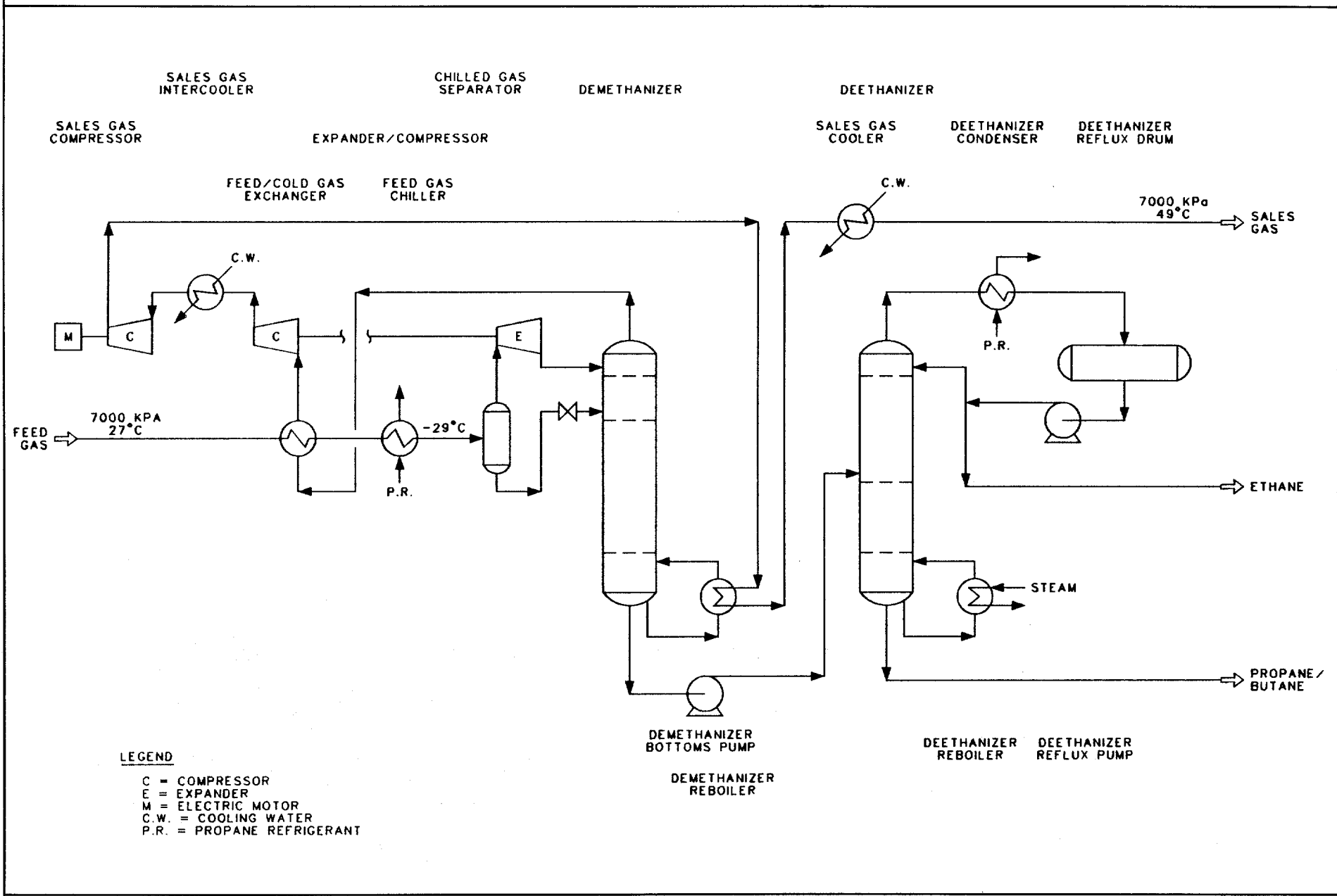


TABLE 11. METHANE-ETHANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 2,758 kPa

Temp., K	Liquid Phase Methane Mole Fraction	Liquid Enthalpy, kJ/kg	Vapor Phase Methane Mole Fraction	Vapor Enthalpy, kJ/kg
174.5	1.0	-16.5	1.0	264.5
176.8	0.95	—	0.994	—
179.0	0.90	—	0.986	—
183.5	0.80	-88.2	0.977	288.3
194.6	0.60	-120.8	0.941	309.5
211.9	0.40	-120.9	0.855	327.5
224.7	0.30	-105.8	0.760	330.1
240.9	0.20	-77.9	0.594	322.7
259.8	0.10	-38.0	0.329	304.5
269.5	0.05	—	0.168	—
274.3	0.025	—	0.084	—
278.9	0.0	8.67	0.0	282.6

*Datum: ideal gas at zero K, zero kPa

TABLE 12. ETHANE-PROPANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 2,758 kPa

Temp., K	Liquid Phase Ethane Mole Fraction	Liquid Enthalpy, kJ/kg	Vapor Phase Ethane Mole Fraction	Vapor Enthalpy, kJ/kg
279.0	1.0	9.7	1.0	287.2
281.3	0.95	—	0.979	—
283.8	0.90	—	0.957	—
288.9	0.80	13.7	0.907	295.1
300.4	0.60	28.0	0.777	306.8
313.8	0.40	51.3	0.592	317.7
321.2	0.30	66.6	0.473	322.8
329.2	0.20	84.2	0.333	327.7
337.5	0.10	104.1	0.175	332.4
341.8	0.05	—	0.089	—
343.9	0.025	—	0.045	—
346.1	0.0	127.0	0.0	341.0

*Datum: ideal gas at zero K, zero kPa

TABLE 13. SURFACE TENSION OF LIQUID, PURE HYDROCARBONS [2]

Temperature, °C	-120	-100	-50	0	50	100
	Surface Tension (mN/m)*					
Methane	6.5	2.5	—	—	—	—
Ethane	22	18.5	10	3.5	—	—
Propane	27	24	16.5	9.5	4	—
n-Butane	—	—	21	14.7	9	4

*1 mN/m = 1 dyne/cm

Notes

- Where dash (—) is shown in Tables 6 through 12, values are not available and should be estimated by interpolation.
- Where dash (—) is shown in Table 13, values are not applicable.

REPORT FORMAT

The final report should include the following:

1. Cover letter
2. Introduction: give a concise statement of the problem, covering background and objectives.
3. Summary: Give a brief review of the work done and methods used.
4. Conclusions: List the major conclusions in decreasing order of importance. The conclusions must follow logically from the summary (i.e., new material is not introduced here).
5. Recommendations: List actions, goals, and other recommendations, in decreasing order of importance. The recommendations must follow logically from the conclusions.
6. Process Appendix: Present the process flow diagram with the related material balance and equipment list with design sizes or duties where appropriate.
7. Calculational Appendix: Present the details of the performed work and include appropriate data, any special theory, calculations, assumptions, diagrams, references, etc. for complete and stand-alone documentation.

CONVERSION FACTORS AND ABBREVIATIONS

Conversion Factors for SI Units

British thermal unit (Btu) $\times 1055.1 =$ joules (J)
 Btu/hour $\times 0.29307 =$ watts (W)
 Btu/lbm °F $\times 4.1868 =$ kilojoules/kilogram-Kelvin (kJ/kg K)
 calorie (cal) $\times 4.1868 =$ joules (J)
 centipoise (cP) $\times 0.001 =$ pascal-second (Pa·S)
 degrees Fahrenheit (°F): $5/9 (°F - 32) =$ degrees Celsius (°C)
 dyne/centimeter (dyne/cm) $= 0.001$ newton/meter (N/m)
 gallons (gal) $\times 0.003785 =$ cubic meters (m³)
 horsepower (hp) $\times 745.7 =$ watts (W)
 pound-force (lbf) $\times 4.4482 =$ newtons (N)
 pound-force/square inch (lbf/in.²) $\times 6894.8 =$ pascal (Pa)

pound-mass (lbm) $\times 0.45359 =$ kilograms (kg)
 pound-mass/cubic foot (lbm/ft³) $\times 16.0185 =$ kilograms/cubic meter (kg/m³)
 degrees Rankine (°R) $\times 5/9 =$ Kelvin (K)
 tons refrigeration $\times 3.51685 =$ kilowatts (kW)
 tons refrigeration $\times 12,000 =$ Btu/hour

Additional conversion factors may be obtained from Section I of the Chemical Engineers' Handbook [1].

Abbreviations

G	= giga (10 ⁹)
J	= joules
k	= kilo (10 ³)
Pa	= Pascals
M	= mega (10 ⁶)
m	= milli (10 ⁻³)
m	= meter
kW	= kilowatts
kWh	= kilowatt hours
bhp	= brake horsepower
in.	= inch
ft	= feet
h	= hour
min	= minute
K	= Kelvin
cm	= centimeter
lb	= pound

REFERENCES

1. Perry, R. H., D. W. Green, and J. O. Maloney, "Perry's Chemical Engineers' Handbook," Sixth Edition, McGraw Hill, New York (1984).
2. "SI Engineering Data Book," Adapted From "Engineering Data Book," Ninth Edition, Gas Processors Suppliers Association, Tulsa (1980).
3. Charles D. Holland, "Fundamentals of Multicomponent Distillation," McGraw-Hill Book Co., New York, p. 17 ff (1981).
4. Robert E. Treybal, "Mass Transfer Operations," Third Edition, McGraw-Hill Book Co., New York, p. 366 ff, (1980).

APPENDIX: FIGURES AND TABLES

TABLE 1. PURE COMPONENT CRITICAL PROPERTIES

	Temperature, K	Pressure, kPa	Volume, cc/gmole	Acentric Factor
Methane	190.63	4,617	99	0.010
Ethane	305.43	4,884	148	0.099
Propane	369.82	4,250	203	0.152
n-Butane	425.16	3,797	255	0.201

TABLE 2. HIGH PRESSURE K-VALUE DATA

Pressure: 7,000 kPa					
Temperature: at values shown below					
	-73.3°C	-51.1°C	-28.9°C	-6.7°C	15.6°C
Methane, K-value	1.4	1.55	1.68	1.82	2.10
Ethane, K-value	0.33	0.41	0.50	0.61	0.77
Propane, K-value	0.079	0.12	0.175	0.25	0.35
n-Butane, K-value	0.021	0.044	0.075	0.105	0.153

TABLE 3. LOW PRESSURE K-VALUE DATA

Pressure: 1,034 kPa							
Temperature: at values shown below							
	-101.1°C	-73.3°C	-45.6°C	-17.8°C	10.0°C	37.8°C	65.6°C
Methane, K-value	2.3	4.7	7.3	10.5	13	15.8	18
Ethane, K-value	0.072	0.26	0.66	1.35	2.4	3.75	5.4
Propane, K-value	0.0044	0.029	0.11	0.29	0.65	1.2	2.0
n-Butane, K-value	2.7×10^{-4}	0.003	0.018	0.068	0.18	0.40	0.75

TABLE 4. LOW PRESSURE K-VALUE DATA

Pressure: 2,069 kPa							
Temperature: at values shown below							
	-101.1°C	-73.3°C	-45.6°C	-17.8°C	10.0°C	37.8°C	65.6°C
Methane, K-value	1.18	2.35	3.80	5.30	6.70	8.10	9.40
Ethane, K-value	0.05	0.165	0.39	0.78	1.35	2.05	2.82
Propane, K-value	0.0035	0.0195	0.070	0.182	0.38	0.67	1.18
n-Butane, K-value	0.00025	0.0025	0.0125	0.044	0.112	0.235	0.43

TABLE 5. LOW PRESSURE K-VALUE DATA

Pressure: 2,759 kPa							
Temperature: at values shown below							
	-101.1°C	-73.3°C	-45.6°C	-17.8°C	10.0°C	37.8°C	65.6°C
Methane, K-value	0.94	1.8	2.85	3.95	5.1	6.1	7.1
Ethane, K-value	0.046	0.15	0.35	0.66	1.1	1.65	2.2
Propane, K-value	0.0035	0.018	0.063	0.155	0.34	0.57	0.90
n-Butane, K-value	2.5×10^{-4}	2.4×10^{-3}	0.0115	0.039	0.10	0.205	0.36

TABLE 6. METHANE-ETHANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 1034 kPa

Temp., K	Liquid Phase Methane Mole Fraction	Liquid Enthalpy,* kJ/kg	Vapor Phase Methane Mole Fraction	Vapor Enthalpy,* kJ/kg
149.6	1.0	-140.0	1.0	273.8
151.0	0.95	—	0.999	—
152.3	0.90	—	0.997	—
155.1	0.80	-194.6	0.994	285.9
162.1	0.60	-225.5	0.984	299.8
173.3	0.40	-235.6	0.955	317.6
182.3	0.30	-230.0	0.917	327.0
195.6	0.20	-212.4	0.826	331.2
215.7	0.10	-176.0	0.581	316.0
235.2	0.05	—	0.340	—
242.1	0.0	-119.2	0.182	—

*Datum: ideal gas at zero K, zero kPa

TABLE 7. METHANE-ETHANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 1,379 kPa

Temp., K	Liquid Phase Methane Mole Fraction	Liquid Enthalpy,* kJ/kg	Vapor Phase Methane Mole Fraction	Vapor Enthalpy,* kJ/kg
156.2	1.0	-111.0	1.0	278.0
157.7	0.95	—	0.998	—
159.3	0.90	—	0.996	—
162.5	0.80	-168.8	0.992	290.1
170.4	0.60	-200.4	0.977	305.5
183.0	0.40	-208.5	0.937	324.1
193.0	0.30	-200.7	0.887	332.2
207.4	0.20	-180.3	0.774	333.0
227.5	0.10	-142.5	0.510	315.3
239.5	0.05	—	0.286	—
245.7	0.025	—	0.149	—
251.9	0.0	-89	0.0	282.9

*Datum: ideal gas at zero K, zero kPa

TABLE 8. METHANE-ETHANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 1,724 kPa

Temp., K	Liquid Phase Methane Mole Fraction	Liquid Enthalpy,* kJ/kg	Vapor Phase Methane Mole Fraction	Vapor Enthalpy,* kJ/kg
161.7	1.0	-85.0	1.0	278.0
163.5	0.95	—	0.997	—
165.2	0.90	—	0.995	—
168.7	0.80	-146.2	0.989	291.8
177.4	0.60	-178.2	0.969	308.8
191.4	0.40	-184.4	0.918	327.6
202.2	0.30	-174.5	0.856	334.3
217.3	0.20	-151.8	0.724	332.4
237.3	0.10	-113.2	0.453	313.7
248.6	0.05	—	0.246	—
254.4	0.025	—	0.127	—
260.1	0.0	-61.1	0.0	281.0

*Datum: ideal gas at zero K, zero kPa

TABLE 9. METHANE-ETHANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 2,068 kPa

Temp., K	Liquid Phase Methane Mole Fraction	Liquid Enthalpy,* kJ/kg	Vapor Phase Methane Mole Fraction	Vapor Enthalpy,* kJ/kg
166.5	1.0	-61.5	1.0	273.6
168.4	0.95	—	0.996	—
170.3	0.90	—	0.993	—
174.1	0.80	-125.7	0.985	291.9
183.7	0.60	-157.9	0.961	310.3
198.8	0.40	-162.1	0.898	329.0
210.4	0.30	-150.4	0.824	334.4
226.0	0.20	-125.8	0.678	330.2
245.7	0.10	-86.6	0.406	311.3
256.5	0.05	—	0.215	—
261.8	0.025	—	0.109	—
267.1	0.0	-37	0.0	286.0

*Datum: ideal gas at zero K, zero kPa

TABLE 10. METHANE-ETHANE TEMPERATURE-COMPOSITION-ENTHALPY VLE DATA AT 2,413 kPa

Temp., K	Liquid Phase Methane Mole Fraction	Liquid Enthalpy,* kJ/kg	Vapor Phase Methane Mole Fraction	Vapor Enthalpy,* kJ/kg
170.7	1.0	-39	1.0	263.7
172.8	0.95	—	0.995	—
174.9	0.90	—	0.991	—
179.0	0.80	-106.5	0.981	290.7
189.4	0.60	-138.9	0.951	310.4
205.7	0.40	-141.1	0.877	328.9
217.9	0.30	-127.6	0.792	332.8
233.8	0.20	-101.3	0.635	326.9
253.1	0.10	-61.7	0.365	308.3
263.4	0.05	—	0.189	—
268.4	0.025	—	0.095	—
273.3	0.0	-12	0.0	288.1

*Datum: ideal gas at zero K, zero kPa