1959
Problem
Optimization Calculations for an Ethylene Purification Unit

FOREWORD

A committee from the Humble Oil & Refining Company, Baytown, Texas—B. B. Ashby; Irving Leibson; W. N. Lyster; T. W. Pickel; R. L. Seldomridge; R. J. Smouse, chairman; and C. L. Umholtz—prepared and judged this year's contest. Sixty-one solutions were received, and the judges considered "the caliber of the papers very high. Most of the solutions indicated a solid grasp of the fundamental technical methods necessary in working the problem."

The first prize, the A. McLaren White, of $200 was awarded to J. R. Griffin, University of Illinois; the second prize of $100, the A. E. Marshall, to L. G. Rossa, Purdue University; the third prize of $50 to W. J. Lee, Georgia Institute of Technology; and honorable mention to T. W. Carroll, Yale University; J. C. Kennedy, Tulane University, and J. L. Schlechte, Lamar State College. The first-prize-winning solution follows on page 11.

A. I. CH. E. STUDENT CONTEST PROBLEM FOR 1959

TO THE CONTESTANT:

In the final designing of a new process unit following the initial justification, optimization calculations are usually made to determine the values of process variables which will produce the desired product most economically. Since there is a specific time by which this study must be completed, the engineer working on the problem must plan the utilization of his work carefully.

This requires that he first analyze the problem, separating it into stages of calculation procedure. He then determines the amount of time he may spend on each stage. From this time analysis, he determines the depth or detail of his calculations.

The proper presentation of the final solution is very important. The report must contain a summary of the important conclusions for the quick understanding of management, plus a detailed section showing how these conclusions were reached. In an appendix should be given the calculations, including the figures and references used. Accuracy is of great importance, but almost equally valuable are clarity and good organization.

The subject of this problem is the optimization of the design of a distillation unit to purify ethylene from an ethane-ethylene mixture. The general process flow has been determined and the initial justification developed. Before the detailed process design can be made, the column pressure at which the ethylene product can be produced most economically must be determined. Steven O. Boss, Chief Process Engineer, has called you, James R. Engineer, in to outline the problem.

CONFERENCE NOTES

Present: Steven O. Boss, James R. Engineer

Mr. Boss: Jim, our marketing department some time ago determined that there is going to be quite a demand for ethylene in the near future. In the past we have sent the ethane and lighter streams to the fuel-gas system; however, our engineers have looked at this ethylene demand and the probable selling price for ethylene and have determined that purification of ethylene looks profitable. In their justification of the project they found that the process of distillation, with a heat pump in place of an overhead condenser, seemed the best from our standpoint. Owing to the many factors which influenced the optimum, they did not determine the optimum tower pressure, but before we start on the final design for construction of the unit, we need to know the tower pressure at which we can operate most economically.

Since you have just joined the company, I imagine you will need a rundown on the details of the process. I've asked Slade Rhule, one of our experienced design engineers, to fill you in.

I'd like to have your conclusions in a formal report, so we can inform our plant management quickly and get down to the process design. Be sure to include a summary and append your detailed calculations.

CONFERENCE NOTES

Present: Slade Rhule, James R. Engineer

Mr. Rhule: Jim, first I want to show you the flow diagram for our proposed ethylene purification unit (Figure 1). The feed to the tower comes

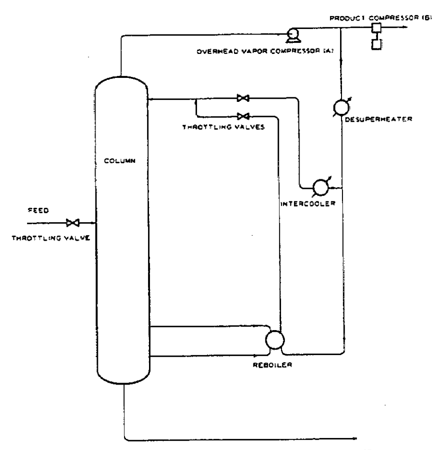

Figure 1. Flow diagram.

from our deethanizer, a tower which produces ethane and lighter as overhead product. Ahead of the deethanizer, most of the methane and noncondensible gases have been removed in our demethanizer. The feed consists of an ethane-ethylene stream, 41,500 lb./hr. 32.5 weight % ethylene. The feed stream, which has been cooled to a saturated liquid at 290 lb./sq. in. abs. in a feed precooler, has slight amounts of methane, propane, and carbon dioxide, but we shall neglect these in the problem.

Our marketing department has specified to our customers an ethylene product of 98 weight % purity; therefore, the overhead vapor must be of this composition. The overhead vapor stream is compressed in the overhead vapor compressor, A. After compression, the ethylene product stream is removed. Since the ethylene product is to be delivered to our pipe line at 500 lb./sq. in. abs. we may need a product compressor, B.

As you can see from the flow diagram, the remainder of the overhead vapor stream is the reflux stream. We need to remove heat from this stream to obtain liquid as reflux.

Now we also need heat for the reboiler. By the proper choice of compression, we can use the overhead vapor or reflux stream to furnish the heat for reboiling.

Mr. Engineer: Mr. Rhule, I'm not sure I understand that. Would you go over it again?

Mr. Rhule: Perhaps a look at these enthalpy-pressure charts will help (Figures 2 and 3). Let's assume that our column, or splitter, as we call it, is operating at 80 lb./sq. in. abs. Since the bottoms stream to the reboiler is almost pure ethane, we see from the chart that the equilibrium temperature is about -58°F. at this pressure.

If we have compressed the overhead vapor stream to 212 lb./sq. in. abs., we find that the equilibrium temperature for ethylene is -40°F.; therefore, we have a differential temperature of 18°F. across the reboiler and can reboil the bottoms while we condense the ethylene reflux. These figures are only approximate, since we are assuming pure streams, but that's the idea.

Mr. Engineer: I think I understand that now. But I don't see why we have those two other exchangers in the overhead vapor-reflux stream.

Mr. Rhule: The amount of heat needed in the reboiler is independent of the heat content of the compressed overhead vapor-reflux stream, as you will see by making a heat balance around the tower. Therefore, we remove the extra heat not needed in the reboiler with these two exchangers. There are several levels of refrigeration available for these exchangers. We use two exchangers since we can use a higher temperature refrigerant in the desuperheater, than we can in the intercooler. This results in savings in refrigerant costs. It may be possible that one of these exchangers can be eliminated at higher tower pressures.

You have probably noticed that we have three throttling valves on the unit. These can be assumed to be constant-enthalpy throttling valves.

I think that about flanges up the process. There are a few more points. The ethane out the bottom will be moved by tower pressure through exchangers to our fuel-gas system. For this reason, as well as some other considerations, we have set 80 lb./sq. in. abs. as the minimum tower pressure to be investigated. We have determined that a 14°F. differential temperature from condensing vapor to boiling liquid across the reboiler condensing section is optimum, and we shall design for a 97 weight % recovery of the ethylene in the feed; that is, 3 weight % of the ethylene will be lost in the bottoms.

Our large overhead vapor compressor will be a centrifugal compressor driven by a steam turbine. The compressor is 65% efficient. This means that actual work input required equals isentropic work input required divided by 0.65. Assume that all this inefficiency enters the compressed vapor as heat. The driver efficiency is included in the cost data.

The product compressor will be a reciprocating compressor with an electric motor driver. It also is 65% efficient owing to its small size. The electric motor driver is 90% efficient.

You may use a U based on outside tube area of 95 B.t.u./(hr.)(sq.ft./°F.) for condensing vapor to boiling liquid in the reboiler and intercooler and 65 B.t.u./(hr.)(sq.ft./°F.) for desuperheating vapor to boiling liquid in all three exchangers. The boiling liquid is shell side in all cases. These numbers include the dirt factor and are based on log mean differential temperatures.

There is one other factor which must be taken into account. The standard material of construction for noncorrosive service, carbon steel, can be used only at temperatures above -20°F. Between -20° and -50°F. we use killed carbon steel, and below -50°F. we use 3 1/2% nickel steel.

Mr. P. I. Dough, our cost estimator, has the cost figures you will need. I suggest you get together with him soon to get those data.

Mr. Engineer: We skipped over the column itself. What about calculating it?

Mr. Rhule: The tower itself can be sized from the Souders-Brown relationship for allowable vapor velocity given on page 597 of Perry's handbook (9). Assume maximum and minimum flow rates through the tower to be plus and minus 20% of the design rate. The number of plates required for the desired separation can be estimated by the McCabe-Thiele method or any comparable one. Plate efficiency can be estimated from the O'Connell correlation given on page 614 of Perry (9). (Viscosity is in centipoise.) Here are the equilibrium data (Table 1). I should recommend that you use a 24-in. plate spacing.

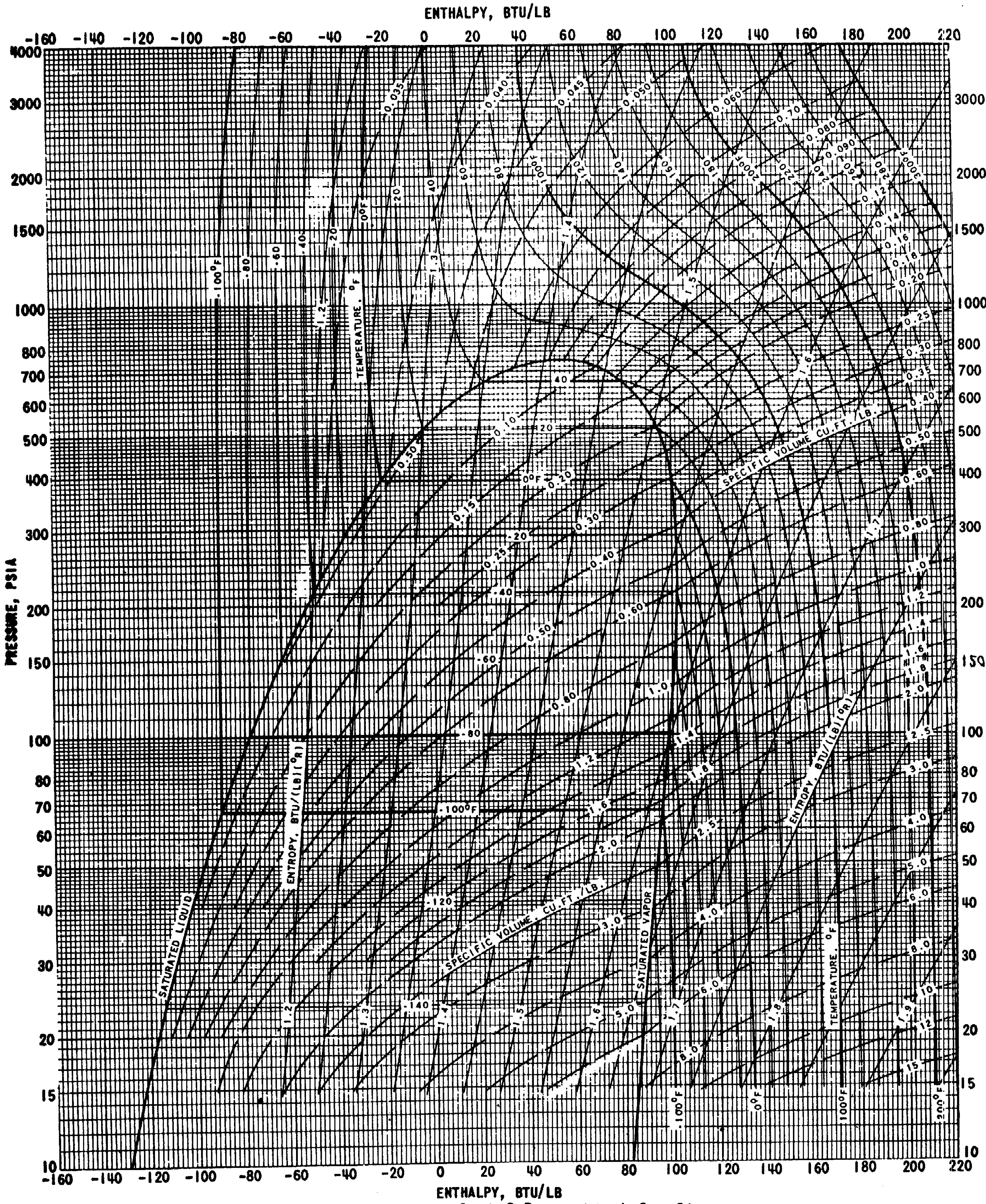

Figure 2a. Mollier chart for ethylene. From W. C. Edmister, Petroleum Refiner (June, 1958), reproduced by permission.

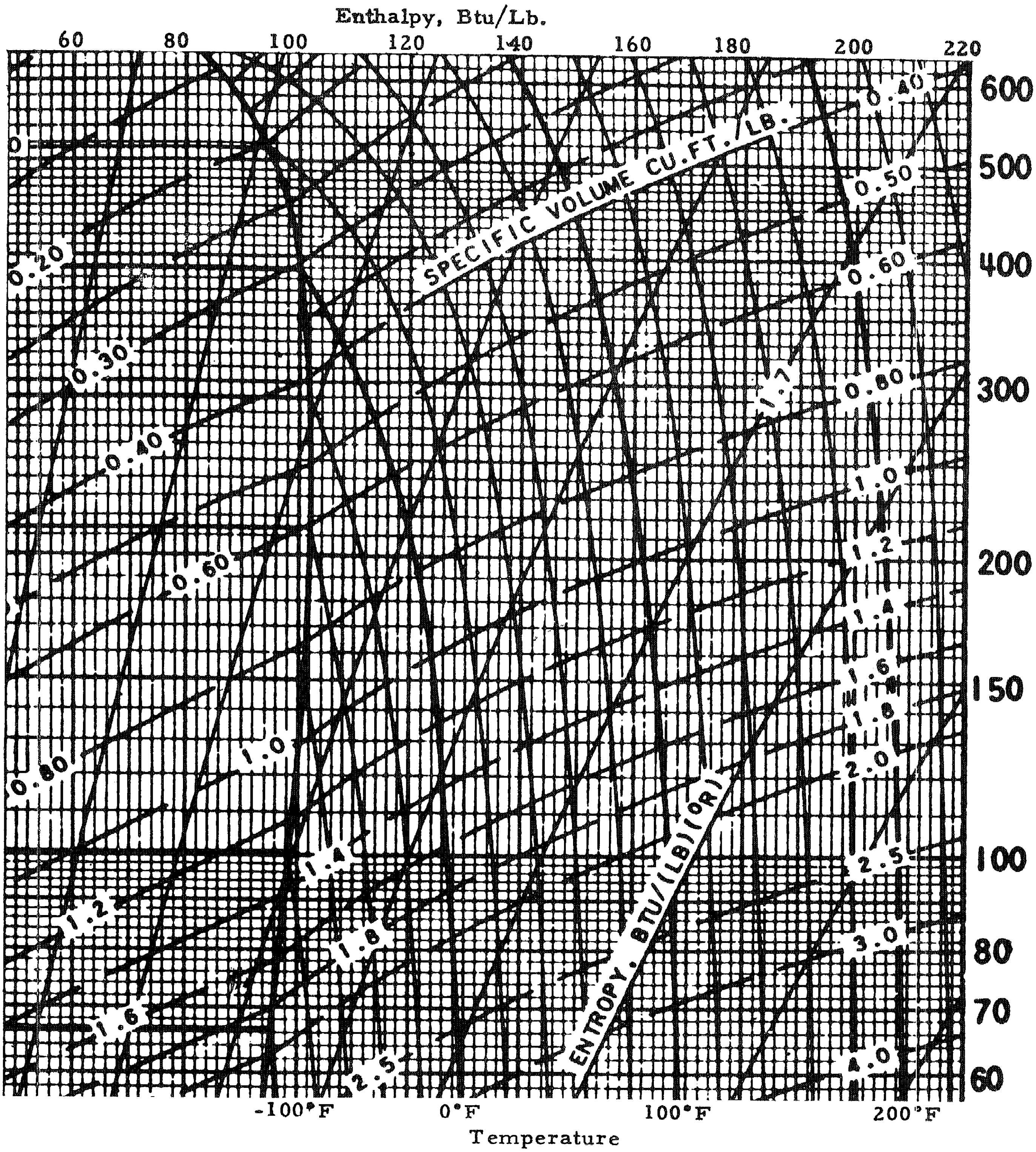

Figure 2b. Mollier chart for ethylene. From W. C. Edmister, Petroleum Refiner (June, 1958), reproduced by permission.

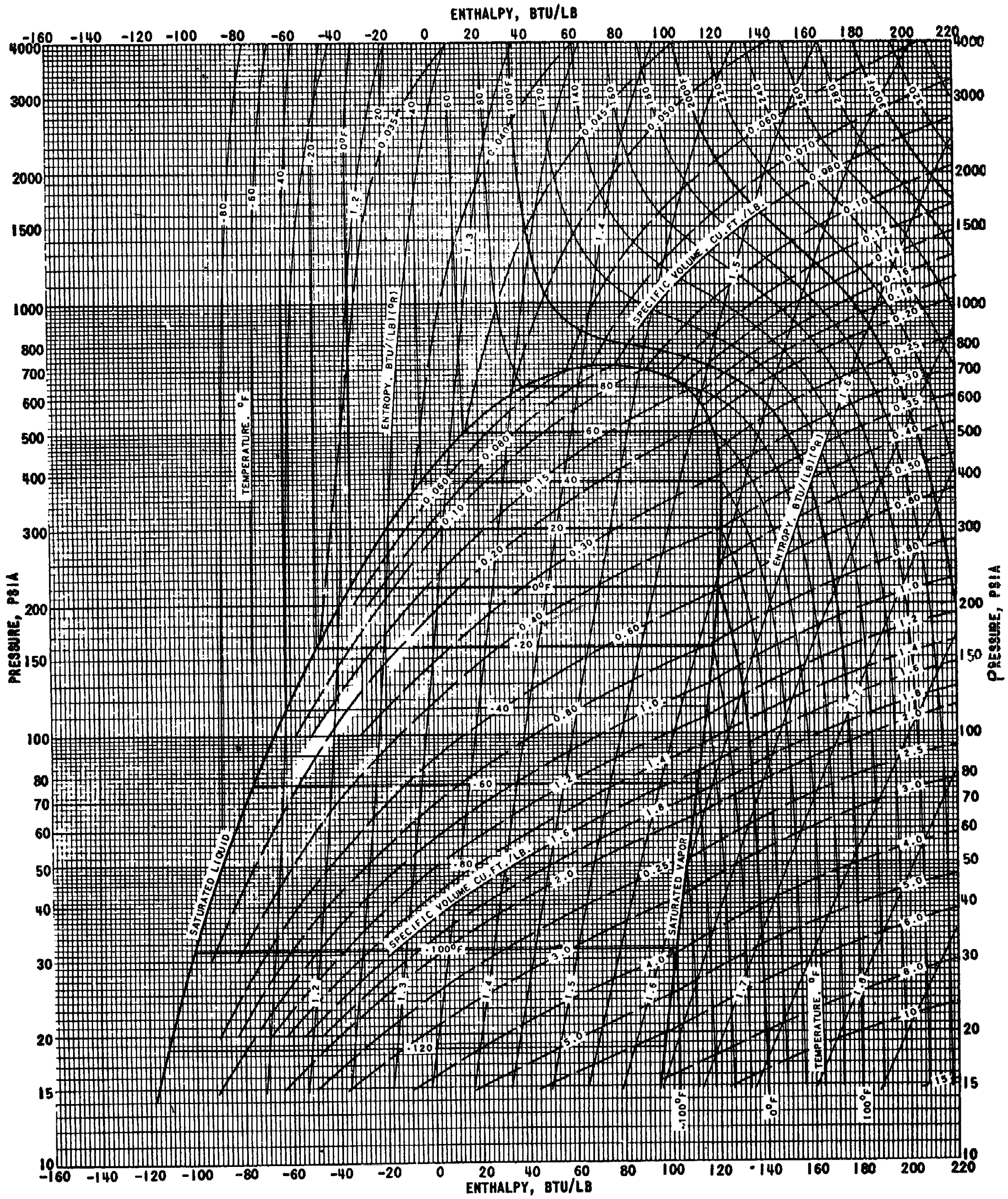

Figure 3. Mollier chart for ethane. From W. C. Edminster, Petroleum Refiner (June, 1958), reproduced by permission.

TABLE 1. RELATIVE VOLATILITY FOR ETHYLENE-ETHANE MIXTURES AS A FUNCTION OF LIQUID COMPOSITION AND SYSTEM PRESSURE

$$\text{Relative volatility}_{\text{ethylene, ethane}} = \frac{K_{\text{ethylene}}}{K_{\text{ethane}}}$$

P, lb./sq.in.abs.	80	100	125	150	200	250
Mole fraction						
0.0+	1.960	1.889	1.807	1.758	1.668	1.600
0.1	1.950	1.883	1.803	1.753	1.664	1.600
0.2	1.939	1.872	1.793	1.745	1.660	1.592
0.3	1.908	1.842	1.770	1.727	1.642	1.576
0.4	1.872	1.808	1.736	1.694	1.611	1.550
0.5	1.848	1.773	1.709	1.665	1.588	1.533
0.6	1.809	1.750	1.686	1.648	1.575	1.526
0.7	1.790	1.731	1.670	1.635	1.562	1.512
0.8	1.769	1.712	1.652	1.608	1.546	1.500
0.9	1.748	1.680	1.617	1.580	1.512	1.465
1.0-	1.698	1.640	1.574	1.540	1.473	1.444

Based on data from G. H. Hanson, (5).

Design for a liquid seal of 2 to 4 in. on the tray. For the optimization procedure, use an actual reflux ratio of 1.3 times the minimum.

There is one other thing we should like to have done, Jim. As you may know, most of our equipment design is done in the Design Engineering Department. This department is staffed mainly by mechanical engineers, who are well able to design mechanical equipment such as exchangers, pumps, and compressors. However, the design layout of the trays in our tower is a job for a chemical engineer; therefore, we should like you to give us a design for the plates in the tower at the optimum pressure. We plan to use perforated plates. In case you're not familiar with perforated-plate design, this design sheet should help you (Appendix A). We generally use 10-gauge sheet metal for the plates with 3/16-in.-diameter circular perforations.

CONFERENCE NOTES

Present: Peter I. Dough, James R. Engineer

Mr. Engineer: Mr. Dough, Mr. Rhule said that you would be able to help me get the data I need for estimating the cost of the new ethylene unit. It looks like I'll need data for exchangers, compressors, and towers, plus associated equipment such as instrumentation and piping.

Mr. Dough: I have some references here, Jim, which should help you. In addition, we have some of our own methods, which I'll give you.

We normally try to use a standard exchanger size in all our designs. This is less expensive initially and generally leads to more economical maintenance. This table (Table 2) gives size,

TABLE 2.

STANDARD EXCHANGER SIZES

Nominal shell, I.D., in.	Area outside tubes, sq.ft.	Cost/sq.ft. of surface outside tubes, $
15	280	8.50
18	460	7.00
20	615	6.25
22	765	5.70
24	950	5.23
26	1,155	4.83
28	1,380	4.51
30	1,640	4.22
32	1,870	4.00
34	2,180	3.77
36	2,405	3.61
38	2,640	3.46
40	3,050	3.29
42	3,425	3.13
44	3,755	3.04
46	4,165	2.92
48	4,540	2.81

These costs and areas are for 3/4-in. O.D. tubes, 1-in.sq. pitch, 16-ft. tubes.

TABLE 3. PRESSURE FACTORS FOR TOWER & EXCHANGER SHELLS AND EXCHANGER TUBES

P, lb./sq.in. gauge	Shell factor	Tube factor
0	1.000	1.000
50	1.011	1.004
100	1.028	1.013
150	1.051	1.026
200	1.078	1.042
250	1.109	1.060
300	1.144	1.083
350	1.181	1.109
400	1.219	1.138
450	1.260	1.169
500	1.301	1.202

Tables 2 and 3 are based on data from Max S. Peters, (10).

outside tube area, and cost per square foot of area for each of the standard exchangers. This table (Table 3) gives the factors by which exchanger costs must be multiplied when the exchanger operates at pressures greater than atmospheric. You'll note that there is a separate factor for the shell and the tubes. Finally the cost must be multiplied by a factor for the material of construction. These factors are

Carbon steel	1.00
Killed carbon steel	1.43
3 1/2% nickel steel	2.12

I believe Mr. Rhule mentioned the allowable temperature ranges of each type of material.

As for tower costs, we use curve 3 on page 70 of Aries and Newton (1). While the curve is for bubble-cap towers at 1955 costs, we find that it is now about right for perforated-plate columns at present-day costs. This cost curve is for an atmospheric-pressure column. To correct the cost figure for tower pressure, you can use the shell-pressure factor I gave you for exchangers. For this particular tower, assume that the cost of the shell at atmospheric pressure, including nozzles and manways, is 70% of the total tower cost. Since the temperature will decrease up the tower, the top and bottom may be made of different materials. The number of trays of each material must be determined, and the proper material cost factor must be used. Both tray and tower shell material will vary with temperature level.

We estimate centrifugal-compressor costs based upon the shaft brake horsepower of the compressor by the following formula, $\$ = 1{,}800\ (\text{b.hp.})^{0.48}$. Driver costs are estimated from this graph (Figure 4). Since only certain parts

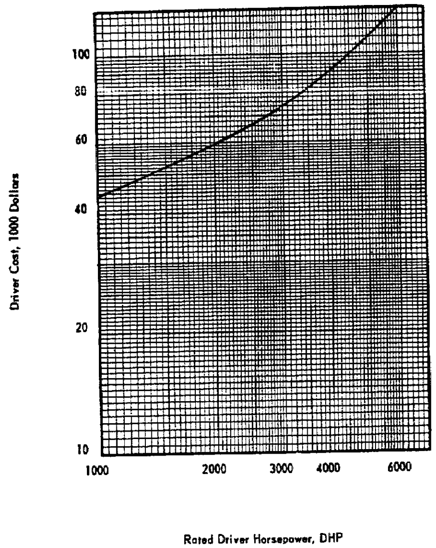

Figure 4. Steam-turbine costs.

of the compressor will touch the incoming cold fluid, these material cost factors will be used:

Carbon steel	1.00
Killed carbon steel	1.29
3 1/2% nickel steel	1.43.

Of course, the driver will not be included in the material-cost-factor correction, since it will not touch the cold fluid.

The product compressor will be a reciprocating compressor with an electric motor driver. Cost of the combined compressor-driver is given by $\$ = 650\ (\text{b.hp.})^{0.7}$.

Well, Jim, I believe that takes into account the cost of the major equipment items. Now the unit has to be controlled; therefore, we need instrument costs. You can get these from Aries and Newton, page 96 (1). Multiply by 1.3 to get present-day costs.

Mr. Engineer: Mr. Dough, I notice that we haven't included piping or insulation. It seems as though these could play a pretty big part, both in calculation and in the final cost.

Mr. Dough: They do, Jim, and an accurate job of estimating would require a plant layout. But for this particular problem, I've looked into the piping and insulation costs to see whether we could simplify your work. From our previous experience in cost estimating, we have found that piping can be fairly accurately expressed as a function of equipment cost. However, the insulation costs will also be a function of the temperatures. Since the operating temperatures are functions of tower pressure, we can determine insulation cost as a function of tower pressure. For the ethylene unit, I believe that you can estimate piping cost as 80% of the difference between total equipment cost and the cost of compressors and drivers. Insulation costs can be estimated as $\left[20\,(1.1)^{100/P}\right]\%$ of the piping cost, where P is tower pressure in pounds per square inch absolute. Compressors and drivers were subtracted from our equipment costs since our previous data were for standard distillation towers with overhead condensers.

Just to give us a little leeway, we figure contingency as 10% of direct plant cost. Direct plant cost includes cost of equipment, piping and insulation, erection and installation, engineering, and contractor's fee. Erection and installation is 30% of equipment cost, engineering is 20% of equipment cost, and contractor's fee is 15% of equipment cost. Fixed-capital investment equals direct plant cost plus contingency.

Since we want production of ethylene at the lowest cost per pound of product, you will need to calculate the production costs, including labor, depreciation, maintenance, property taxes, utilities, and our minimum acceptable return on our investment.

Slade Rhule tells me that we will need only 1/3 man/hr. for the ethylene unit as this unit will be a part of our large light-gas unit, and four operators will be able to handle the whole light-gas unit. You can figure labor plus benefits costs \$5/man-hr.

Steam for the turbine driver on the overhead vapor compressor will cost 41.75¢/100 hp.-hr., based on shaft horsepower to the compressor. Electricity costs 1.5¢/kw.-hr.

REFRIGERATION COSTS

Refrigerant	Temperature level, °F.	Cost, \$/year/1,000,000 B.t.u./hr.
Ethylene	-150	16,600
Ethylene	- 90	14,000
Propane	- 34	10,600
Propane	0	8,600
Propane	18	7,000
Propane	60	3,700

Don't forget that our bottoms product will be a saturated liquid and can also be used as a refrigerant. We pick up a utilities credit for the available refrigeration in this stream. You can use interpolation of our refrigeration costs to get the credit value.

Depreciation equals 10% of fixed capital investment (F.C.I.) Maintenance equals 5% of F.C.I. Property tax is 1.5% of F.C.I. Our minimum acceptable return is 30% of F.C.I.

Well, Jim, that about covers all you will need as far as the economics go. From this you should be able to figure what it will cost us at various pressures on a cents-per-pound basis to purify our ethylene. Naturally we're looking for the pressure at which production cost, cents per pound of product, is the least.

TABLE 4. THERMODYNAMIC PROPERTIES OF SATURATED ETHYLENE

Temperature, °F.	Pressure, lb./sq.in.abs.	Volume, cu.ft./lb.		Heat content above saturated liquid at -154.7°F., B.t.u./lb.	
		Liquid	Vapor	Liquid	Vapor
-100	65.58	0.03122	1.879	32.8	217.2
- 80	100.98	0.03179	1.732	45.2	219.9
- 60	148.45	0.03308	0.857	57.9	221.9
- 40	206.30	0.03468	0.593	70.8	222.8
- 20	289.91	0.03662	0.419	84.7	222.7
0	388.03	0.03912	0.301	100.3	221.0
+ 20	507.88	0.04292	0.212	119.6	216.6
+ 40	654.74	0.05035	0.139	148.7	206.2

Table 4 based on data from Robert York, Jr., and E. F. White., Jr. (11).

TABLE 5. THERMODYNAMIC PROPERTIES OF SATURATED ETHANE

TABLE 5. THERMODYNAMIC PROPERTIES OF SATURATED ETHANE

Temperature, °F.	Pressure, lb./sq.in.abs.	Volume, cu.ft./lb.		Heat content above 0°R., B.t.u./lb.	
		Liquid	Vapor	Liquid	Vapor
-100	31.32	0.03029	3.830	186.2	386.8
- 80	50.34	0.03108	2.451	198.2	391.2
- 60	77.02	0.03199	1.638	210.5	395.2
- 40	113.1	0.03303	1.127	223.2	398.6
- 20	159.9	0.03422	0.7983	236.3	401.7
0	219.7	0.03570	0.5754	250.3	403.9
20	294.0	0.03754	0.4198	265.1	404.9
40	385.0	0.03990	0.3062	281.0	404.5
60	494.2	0.04358	0.2164	299.3	401.3

Table 5 based on data from C. H. Barkelew., et al. (2).

TABLE 6. VISCOSITY CHART

(SATURATED LIQUID)

Temperature, °F.	$\text{Viscosity}_{\text{ethylene}}$, centipoise	$\text{Viscosity}_{\text{ethane}}$, centipoise
-244	0.425	0.481
-208	0.270	0.322
-172	0.188	0.227
-136	0.141	0.162
-100	0.118	0.124
- 64	0.098	0.098
- 28	0.082	0.082
+ 8	0.070	0.070
+ 44	0.058	0.058
+ 62	0.052	0.052

Table 6 based on data from S. F. Gerf., and G. I. Galkov, (4).

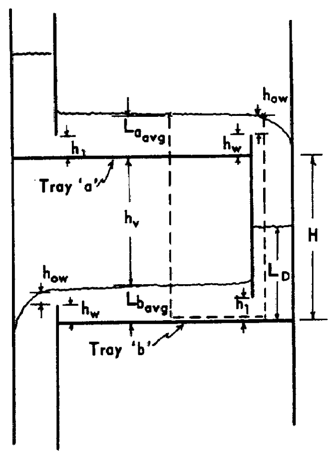

Figure 5. Schematic flow diagram.

APPENDIX A*

Design Sheet for Perforated Trays

The following equations are applicable to the design of perforated plates:

$$L_D = \left[P_{dc} + L_{b_{avg}} + h_v\left(\frac{\rho_V}{\rho_L}\right) + P_D \right. \qquad (1)$$

$$\left. + L_{a_{avg}}\right]\frac{\rho L}{\rho L-\rho V} - \left[H + L_{a_{avg}}\right]\frac{\rho_V}{\rho_L-\rho_V}$$

$$L_{b_{avg}} = h_{OW} + f\,h_W \qquad (2)$$

$$h_{OW} = \left(\frac{Q}{2.98\,L_W}\right)^{2/3} \qquad (3)$$

$$P_{dc} = 0.057\left(\frac{Q}{A_1}\right)^2 \qquad (4)$$

$$P_D = 0.186\left(\frac{u_o}{C_o}\right)^2 \frac{\rho V}{\rho L} \qquad (5)$$

Notation (See Figure 5)

L_D = height of vapor-free liquid in downcomer, in.

P_{dc} = liquid pressure drop through the clearance area between the downcomer and the plate, in. of vapor-free liquid

$L_{b_{avg}}$ = static liquid-seal height on average section of tray b, in.

h_v = hydrostatic head of the vapor between the trays, in. of vapor

ρ_V = density of vapor at conditions of plate, lb./cu.ft.

ρ_L = density of vapor-free liquid at plate conditions, lb./cu.ft.

P_D = dry-tray pressure drop through tray a, in. of vapor-free liquid

$L_{a_{avg}}$ = static liquid-seal height on average section of tray a, in.

H = tray spacing, in.

h_{OW} = head over weir for tray b, in. of vapor-free liquid.

h_W = height of the weir above tray b, in.

f = aeration factor defined as the density of aerated liquid divided by that for vapor-free liquid (Assume f = 1 for design.)

Q = vapor-free liquid flowing over weir, gal./min.

L_W = weir length, in.

A_1 = clearance area between the downcomer and the tray, sq.in.

u_o = linear vapor velocity through the perforations at flowing conditions, ft./sec.

C_o = coefficient of discharge

t = plate thickness, in.

D_o = perforation diameter, in.

Procedure

A. Clear liquid rate in downcomer should not exceed 0.4 ft./sec. From this, calculate downcomer area using maximum liquid flow rate.

B. Calculate crest over the outlet weir using Equation (3) and maximum liquid flow rate.

C. Set outlet weir height by Equation (2), letting $L_{b_{avg}}$ = 4.0 in.

D. Solve for pressure drop for liquid flow through clearance area A by Equation (4). Recommended minimum clearance distance is 1 1/2 in.; maximum is 4.0 in.

E. Solve for maximum permissible dry-tray pressure drop using Equation (1) and setting L_D equal to one-half tray spacing.

F. Solve for perforation area required by trial and error using Equation (5). Use Figure 6 to find C_o.

*This Appendix was abstracted from Petroleum Refiner (February, 1957).

APPENDIX A
(Continued)

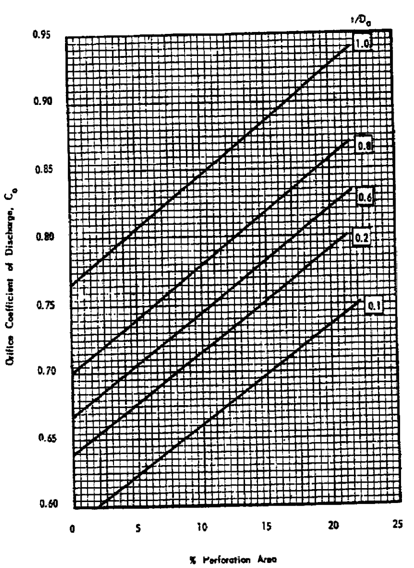

Figure 6. Orifice coefficient of discharge vs. percentage of perforation area.

G. Recheck for weepage at minimum flow by calculating liquid-seal height and dry-tray pressure drop. Compare with Figure 7 to see whether dry-tray pressure drop is high enough to prevent weepage.

H. Calculate triangular hole spacing, using 3/16-in. holes.

$$\text{Spacing} = \left(\frac{\text{area/perforation}}{\text{required perforation area}} \times \frac{\text{area of tray available for perforation}}{0.866} \right)^{1/2}$$

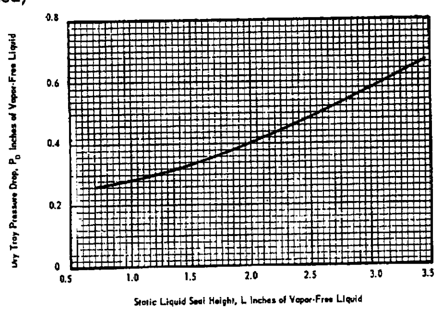

Figure 7. Weepage correlation.

I. Draw final tray layout.

Notes

1. Hydraulic gradient for perforated trays is usually assumed to be negligible.
2. P_{dc} should be less than 1 in. of vapor-free liquid; however, downcomer clearance distance must not be greater than seal depth at minimum flow conditions, so that a positive liquid seal on the downcomer will be provided.
3. Use vertical downcomers.
4. The pitch to diameter ratio of perforations should be greater than 2.0 but less than 5.0.
5. Perforations should be at least 4 in. from the projected edge of the downcomer, and from the outlet weir; and no closer than 3 in. to the tower wall.
6. See Leibson, et al. (7). for details of the method above.

APPENDIX B

Additional Notes and Assumptions

1. The saturated liquid and vapor lines on a constant-pressure enthalpy-composition linear diagram can be assumed to be straight lines.
2. The enthalpies of ethane and ethylene in a mixture are additive in the liquid phase; that is, the molal enthalpy of the mixture equals the sum of the products of the molal enthalpies of the pure components by their molefractions.
3. The minimum temperature difference that can be obtained between the cold fluid and the hot fluid in the exchangers is $10^{\circ}F$.

APPENDIX B
(Continued)

4. The unit will have a service factor of 95%; that is, it will operate 8,322 hr./yr.
5. Neglect pressure drops through lines, exchangers, and the tower.
6. A certain percentage of the total reflux stream will flash as the stream is throttled. Only the portion remaining as liquid will be useful as reflux to the tower. This portion can be assumed to be the same composition as the overhead vapor stream. The percentage flashed can be calculated by heat balances around the throttling valves.
7. To calculate overhead vapor enthalpies,

   A. Add 25 B.t.u./mole to pure-ethylene-vapor enthalpies to get overhead-vapor enthalpies.

   B. Add 20 B.t.u./mole to pure-ethylene-liquid enthalpies to get liquid-reflux enthalpies.

   C. These figures are sufficiently accurate for use at all pressures encountered since the vapor composition will remain constant and the specific heat changes are not extreme.
8. To get vapor and condensing temperatures of the overhead vapor stream, it will be sufficiently accurate to assume that the stream is pure ethylene.
9. To get the temperature of the bottoms stream to the reboiler and the bottoms product stream, it will be sufficiently accurate to assume that the streams are pure ethane.

LITERATURE CITED

1. Aries, R. S., and R. D. Newton, "Chemical Engineering Cost Estimation," McGraw-Hill Book Company, Inc., New York (1955).
2. Barkelew, C. H., J. L. Valentine and C. O. Hund, Trans. Am. Inst. Chem. Engrs., 43, No. 6, 25 (1947).
3. Brown, G. G., et al., "Unit Operations," John Wiley and Sons, Inc., New York (1950).
4. Gerf, S. F., and G. I. Galkov, Zhur. Tekh. Fiz., 2, 801 (1941).
5. Hanson, G. H., et al., Chem. Eng. Progr. Symposium Ser. 6, 49, p. 37 (1953).
6. Kern, D. Q., "Process Heat Transfer," 2 ed., McGraw-Hill Book Company, Inc., New York (1955).
7. Leibson, Irving, et al., Petroleum Refiner, 36, No. 2, p. 127 (1957).
8. McCabe, W. L., and J. C. Smith, "Unit Operations of Chemical Engineering," McGraw-Hill Book Company, Inc., New York (1956).
9. Perry, John H., editor, "Chemical Engineer's Handbook," 3 ed., McGraw-Hill Book Company, Inc., New York (1950).
10. Peters, Max S., "Plant Design and Economics for Chemical Engineers," McGraw-Hill Book Company, Inc., New York (1958).
11. York, Robert, and E. F. White, Trans. Am. Inst. Chem. Engrs., 40, No. 1, p. 227 (1944).