American Institute of Chemical Engineers

STUDENT CONTEST PROBLEM

1967

345 East 47 Street
New York, New Yorl

CONTEST PROBLEM

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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 15, 1967

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 15, 1967. 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 30 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 15, 1967.

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 A.I.Ch.E. 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

Mr. F. J. Van Antwerpen, Secretary
American Institute of Chemical Engineers
345 East 47 Street
New York, New York 10017

AMERICAN INSTITUTE OF CHEMICAL ENGINEERS
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STATEMENT OF PROBLEM

Creating a commercial chemical or petroleum plant is usually a multistaged affair. The project is screened at several levels, the market, technical, economic, and other factors being examined in increasingly greater detail. At each screening the decision to be faced is whether to continue development of the project. At some point in this process an engineering design is called for on which the decision to commercialize the project will be based. Finally the working designs from which the plant is actually constructed must be prepared. This year's student problem is a design at one of the intermediate stages in the evolution of a project. This is not the first process design which has been prepared for this project, nor will it be the last. In fact, part of the problem that you must face is an assessment of the adequacy of some of the conclusions reached in the earlier work.

In this problem we are asking you to prepare a design from which we will decide whether or not to commercialize. This design has a number of requirements. The engineering must be fairly accurate—not too many approximations may be made. On the other hand, every engineering design has a time limit. You will not be able to investigate every possible alternative in detail. Part of the basis for judging your solution will be the degree of emphasis you have put on the important engineering decisions to be made. Economics will play an important part in this problem since the objective is to support a proper commercial decision. You will encounter some instances where it is necessary to arrive at economically optimum process conditions and equipment size. The problem of balancing fixed investment costs against operating costs will be encountered more than once.

The subject of the problem is the manufacture of nitration-grade benzene by the thermal dealkylation of nitration-grade toluene. This is a high-temperature, noncatalytic process in which toluene and hydrogen are reacted to form benzene and methane, with minor amounts of by-product. Excess hydrogen must be used to repress side reactions and coke formation. The reaction products must be separated, by-products rejected, unreacted toluene recovered and recycled, and the benzene product clay treated and distilled to the proper level of purity.

This is an actual problem. The commercialization decision is a real one that has been faced by petroleum and chemical companies in recent years. The market conditions of supply and demand cause fluctuations in the prices of benzene and toluene. Obviously, the expected future of this price differential is a major factor in such a decision. Although you are not being asked to do market forecasts, you should explore the effect of benzene-toluene price spread on profitability of the plant.

The problem will be presented to you as a series of meetings (or notes of meetings) between the process engineer (yourself) and various other members of your company. You are being asked to determine the economic feasibility of such a plant at the company's St. Louis refinery, which requires that you prepare an economically optimum process design of a battery limits plant. In summary, your management expects you to do the following:

1. Prepare a process flow diagram with key instrumentation and a general plot plan of this new facility.
2. Summarize the following:
    - Material balance
    - Utility and labor requirements
    - Equipment specifications (sufficient to price major equipment)
    - Installed equipment cost
    - Fixed and working capital requirement
    - Manufacturing costs as function of plant throughput
3. Evaluate the profitability of the project and state your conclusions and recommendations to management.
4. Furnish supporting calculations.
5. State simplifying assumptions and cite references.

Good luck!

WORK ASSIGNMENT

To: Process Engineer
From: Process Supervisor
Subject: Benzene from Toluene at St. Louis Refinery

Please prepare an economic evaluation of a plant to manufacture nitration-grade benzene by thermal dealkylation of 2,000 barrels per stream day (B/SD) of nitration-grade toluene. The process design should be developed in sufficient detail so that capital and manufacturing costs may be estimated.

Available laboratory data may be obtained from the Process Development Division. You may also wish to consult with the following Engineering Division specialists: Heat Transfer, Distillation, Economics.

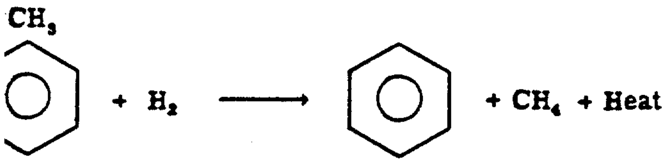
MEETING

Present: Process Supervisor (P.S.)
Process Engineer (P.E.)

.S.: You are probably aware that the Marketing epartment is selling all the benzene that the reinery can supply at prices ranging from 24 to 28¢/ al. and could increase sales if additional benzene ere available. On the other hand, we have excess itration-grade toluene, which is currently blended nto premium gasoline and is valued at 16¢/ gal. his toluene could be thermally hydrodealkylated benzene through a process developed by our R&D epartment.

.E.: What do you mean by thermal hydrodealkyltion?

.S.: Toluene and hydrogen react at elevated temeratures by splitting the methyl group away from luene to produce benzene and methane as shown elow:



.E.: How far has R&D developed the process?

.S.: The process development engineer has inrmed me that several experimental runs have been ade and further work is planned. He has made a reliminary economic evaluation and developed some seful relationships which will enable you to get arted.

.E.: Since the experimental work is still going on, hy not wait until we have all the data before we art?

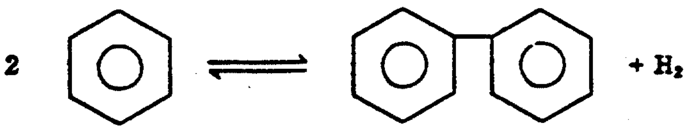
.S.: We could sell the benzene today if we had it. he purpose of your economic evaluation is to present management with sufficient information so that e company may decide whether to build the hydroealkylation unit now. Your work will proceed in rallel with R&D, and if necessary we can make me last-minute adjustments before the presentan to management, which is scheduled for one onth from today. I suggest that you visit Process evelopment to obtain the present R&D thoughts and ta.

EETING

resent: Process Development Engineer (P.D.E.)
Process Engineer (P.E.)

E.: Process Engineering has been asked to recnmend whether or not to go ahead with the construction of a commercial toluene dealkylation unit. I'd like to learn, first, some chemistry of dealkylation and, second, what information you have developed.

P.D.E.: You have already discussed the basic reaction which produces benzene. Some undesirable side reactions reduce the selectivity of the reaction. The main one, and the only one that you should consider in the purification of the product, is diphenyl production, which is represented as follows:



Benzene and toluene can react to form methyl diphenyl, and 2 moles of toluene can form dimethly diphenyl. These two resulting materials will, however, dealkylate to form diphenyl, and we see only trace quantities in the reactor effluent at high conversion. You can disregard these two by-products. But under certain conditions a multi-ring condensation can occur which leads to "coke" formation, and that is trouble.

We define conversion and selectivity quantitatively as follows:

Conversion per pass is the percentage of disappearance of total toluene feed in one pass through the reactor. Selectivity is the net mole percentage of the toluene disappearance that appears as new benzene in the reactor effluent.

The difference between 100 mole percent disappearance and the percentage of selectivity should be considered to be benzene which reacted to produce diphenyl. The coke make will be negligible at the high hydrogen partial pressure we will maintain.

P.E.: I think I understand the chemistry. Please tell me about the information you have developed.

P.D.E.: About 6 months ago we started a program to evaluate the dealkylation of toluene in our existing dealkylation pilot unit. We use a thermal process. There are many catalytic processes now available in the industry. However, our preliminary work indicates that there is no economic incentive to use present-day catalysts, since they offer no significant reduction in operating temperature or improvement in selectivity.

We have determined the kinetics and selectivity vs. conversion relationships. Also, we did a highspot economic calculation based on our early runs. The reaction is 1 1/2 order, and the kinetic expression we arrived at is

$$r_T = -k\,[T]\,[H_2]^{1/2}$$

where

$$k = 6.3 \times 10^{10} \left(\frac{\text{g. mole}}{\text{liter}}\right)^{-\frac{1}{2}} \text{sec.}^{-1} \times \exp\left[\frac{-52{,}000 \text{ cal./g. mole}}{RT}\right]$$

This essentially confirms the data arrived at by others, including Silsby and Sawyer; Betts, Popper, and Silsby; and Zimmerman and York (1).

P.D.E.: Our rough economic calculations (Table 1) were made only for our early runs, which were at 50% conversion. I'll show you what we came up with, but when you get around to making your more detailed investment estimate and economic calculations, you should consult our economics specialist.

P.E.: Do you think that 50% conversion is an optimum?

P.D.E.: I seriously doubt it. Here's what the conversion-selectivity data look like.

	1	2	3	4	5
Conversion/pass, mole %	50	60	70	75	85
Selectivity, mole %	99	98.5	97.7	97	93

As you can see, the conversion-selectivity curve is fairly flat at low conversions; so you don't sacrifice much selectivity until you hit the 75 to 85% conversion range. The payout shown in Table 1 does not look particularly attractive, but I think that at the optimum balance between conversion and selectivity we may be in an acceptable range for our management to approve. As a starter, I would sug-

TABLE 1. TOLUENE DEMETHYLATION

Preliminary Economics

Basis: 2,000 B/SD toluene charge, 5 : 1 hydrogen to oil ratio.
50 mole percent hydrogen recycle gas purity, 50% conversion, 99% selectivity.
Excluding recovery plant losses of toluene and benzene.

Fixed capital investment for battery limits (onsite) plant			$2.15 million
			¢/gal. of fresh toluene charge
Mill cost			
Variable costs (excluding net charge)			
Chemicals		0.03	
Utilities		1.65	
Total variable costs			1.68
Net charge cost			
Toluene @ 16¢/gal.		16.00	
Hydrogen @ 45¢/1,000 std. cu. ft. pure		2.75	
Total charge cost		18.75	
Diphenyl @ 6¢/gal.		(0.04)	
Fuel gas @40¢/MM B.t.u. net heating value		(1.53)	
Total by-product credit		(1.57)	
Net charge cost			17.18
Fixed costs			
Labor, two men/shift		0.34	
Supervision		0.09	
Maintenance	3% fixed capital	1.09	
Depreciation	10% fixed capital		
Local taxes and insurance	1% fixed capital		
General overhead		0.16	
Total fixed costs			1.68
Total mill cost			20.54
Net receipts from benzene sales @ 26¢/gal.			21.52
Net operating income before taxes			0.98
Federal income tax @ 50%			0.49
Net income after taxes			0.49
Depreciation			0.78
Cash flow			1.27
Cash flow payout, yr.			6.1

gest that to establish this "optimum" conversion level, you assume the plant investment varies with total toluene feed rate to the six-tenths power. This will then permit you to calculate the investment-dependent costs. Similarly, we have established that the utility costs will vary according to the following relationships:

Estimated Utility Costs
(For preliminary calculations only)

Basis: 2,000 B/SD toluene charge, 5:1 H_2 to oil feed ratio
50% conversion per pass, 50 mole % H_2 recycle gas purity

Conversion, %	50	60	67	70	75	85
Total utility cost, ¢/gal. fresh toluene	1.65	1.48	1.38	1.35	1.30	1.19

Assume that labor, supervision, hydrogen, and chemical costs will not vary with conversion over the range considered. Use 26¢/gal. benzene for this suboptimization.

P.E.: What design conditions have you established?

P.D.E.: On the basis of our pilot-plant runs, we think that you should not go below a 5:1 hydrogen-to-oil mole ratio at the reaction inlet and definitely not below 40% hydrogen in the recycle gas. This is because of selectivity and coking problems we observed. As you know, nitration-grade toluene contains some minor amounts of impurities which will be removed provided a plug-flow type of reactor is used. Therefore, we recommend that you design a plug-flow reactor with a length-to-diameter ratio of at least 6:1. For all practical purposes, incipient reaction takes place at 1,150°F. We've done all our work at a 500 lb./sq. in. abs. reactor pressure.

P.E.: Do you have any thermodynamic data that might be of use to me?

P.D.E.: Yes, here's a list of enthalpies and heats of formation from API Project 44:

Heats of Formation, H°_f (Kcal./g. mole)

Cpd.	800°K.	900°K.	1,000°K.
Benzene	15.510	15.100	14.818
Toluene	6.399	5.895	5.564
Methane	-20.823	-21.166	-21.430

Heat Content (Enthalpy), $(H^\circ - H^\circ_0)$, (B.t.u./lb.)

	1,100°F.	1,200°F.	1,300°F.
Benzene	545.2	606.2	668.9
Toluene	574.1	638.2	704.1
Methane	1,048.4	1,149.1	1,253.3
Hydrogen	5,361.9	5,715.5	6,071.3

MEETING

Present: Process Supervisor (P.S.)
Process Engineer (P.E.)

P.S.: I've reviewed the notes of your meeting at R&D and have drawn a block diagram of the process (Figure 1) which you can use to start the design.

Toluene fresh feed is mixed with recycled, unconverted toluene and hydrogen-rich recycle gas. From what Process Development has told you, we want 5 moles of hydrogen per mole of "oil" in the gross reactor feed.

P.E.: Do you mean toluene when you refer to oil?

P.S.: By oil, we mean the sum of benzene, toluene, and recycle diphenyl in the total (or gross) reactor feed. There are 2,000 B/SD of toluene available as fresh feed. We will assume that the plant stream factor is 90%; that is, the plant is on stream at full capacity 90% of the time.

Returning to the flow diagram, the gross feed is heated to at least 1,150°F. before being introduced into the reactor. The reactor effluent is cooled and flashed in the phase separator. A portion of the gas produced is vented to purge methane from the system and to keep the unit at constant pressure. Makeup hydrogen is added to the recycle portion of the separator gas, compressed, and fed to the reactor.

P.E.: At what pressure should the phase separator operate?

P.S.: The reactor should be operated at 500 lb./sq. in. abs. The pressure drop through effluent heat exchangers will be approximately 35 lb./sq. in. I would set the separator pressure at 450 lb./sq. in. gauge. The makeup hydrogen is available at 465 lb./sq. in. gauge and 100°F. The maximum amount of gas available is 5 million std. cu. ft./stream day (measured at 60°F.), and the composition is 95 mole % hydrogen and 5% methane. Use the following equilibrium K values for 450 lb./sq. in. gauge and 100°F. in the phase separator:

Hydrogen	86.1
Methane	12.4
Benzene	0.013
Toluene	0.005
Diphenyl	0.0001

Equations for calculating this equilibrium flash are given in reference 7. Wayne Edmister discusses the flash, including a recycle stream specifically, in reference 8.

P.E.: Are there any other design criteria that would be helpful?

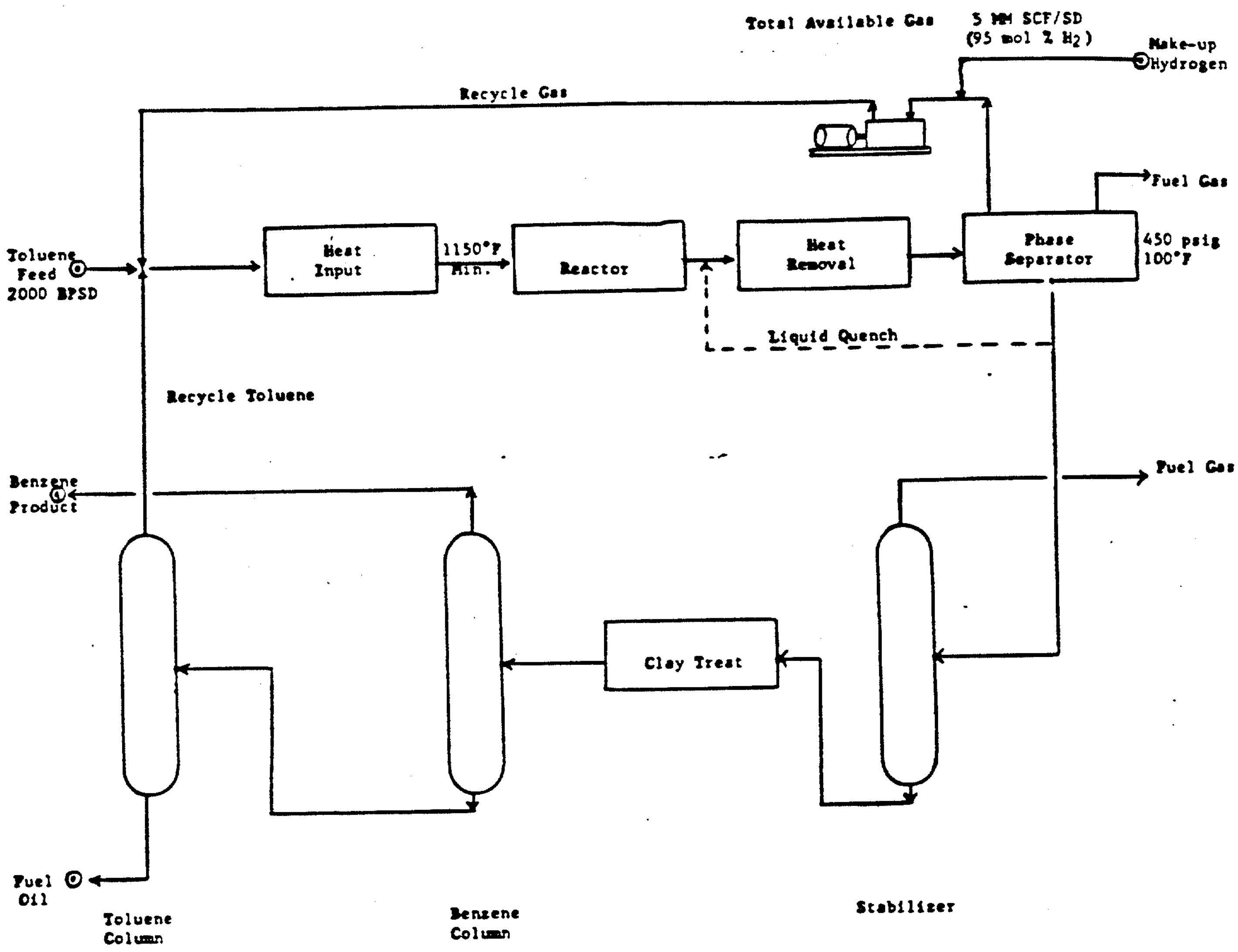


Figure 1. Benzene from Toluene, Block Flow Diagram.

P.S.: Yes, the Metallurgical Department has recommended that we use an internally insulated reactor with a maximum outlet temperature of 1,300°F. Use 6 in. of internal insulation, which will reduce the design temperature of the steel shell to 900°F. The department also recommended that the effluent be quenched to 1,150°F. before entering a heat exchanger.

P.E.: How is the effluent quenched?

P.S.: Let's assume that separator liquid can be used. Recycling this liquid will not affect the flash composition.

P.E.: Shall I prepare heat content curves for the gross reactor feed and gross reactor effluent?

P.S.: No, assume that the specific heat of the gross feed or gross effluent, including phase changes, is 0.9 B.t.u./lb./°F. I don't think we finished the flow diagram. The liquid stream from the phase separator is stabilized to remove hydrogen and methane, clay treated to remove certain contaminents, and fractionated to recover the benzene product from unreacted toluene. The toluene from the benzene column is distilled to reject diphenyl formed in the reactor.

P.E.: I think that I understand the block flow diagram well enough to get started, but I am not sure what steps are required to prepare a complete economic evaluation.

P.S.: For this evaluation, I think we will want to proceed as follows:

1. Using the block flow diagram as a guide, prepare a process flow diagram containing all the major pieces of equipment.
2. Calculate the design material balance. However, in order to determine the material balance, you may have to "suboptimize" some sections of the process. For example, one of the first decisions of this type to be made is the toluene conversion per pass.

3. Make the necessary heat-balance calculations.

4. After the heat and material balances have been finished, the sizes of the major equipment items can be determined.

5. The next step in your evaluation is a capital-cost estimate, which is followed by an estimate of operating cost. At this point we usually decide how the various economic data should be presented to management. For this assignment, I would prefer that you work out these details with our economic specialist.

MEMORANDUM

To: Process Engineer
From: Heat Transfer Specialist

As we discussed yesterday, you may use the following information in preparing your design for the 2,000 B/SD toluene dealkylation unit.

Fired Heater

Since the high outlet temperature of this heater and the hydrogen content of the charge will require alloy tubes (type 316L stainless steel), a double-fired vertical-tube arrangement can be used to minimize the alloy tube surface. For a single-row tube arrangement on 1.8 tube diameter spacing, an average radiant transfer rate of 16,000 B.t.u./hr./sq. ft. may be used for this application. The maximum permissible radiant extraction from the hot flue gas can then be 50% of the fired heat available. Air preheat need not be considered. A pressure drop allowance of 70 lb./sq. in. will be adequate for the reactor-feed heater coil.

Heat Exchangers

The approximate design overall heat transfer coefficients for the shell and tube heat exchanger equipment are as follow:

	B.t.u./hr./sq. ft./°F.
Feed-effluent exchangers	60
Stabilizer condenser	90
Benzene-toluene condenser	130
Other liquid coolers or condensers	50

Reboilers (condensing steam available at 450 lb./sq. in. gauge) limiting flux is 12,000 B.t.u./hr./sq. ft.

For all process fluids a 5 lb./sq. in. pressure-drop allowance may be used for each exchanger on both the shell and tube side. The reboilers will be of the kettle type. Up to 500°F., the material of construction may be carbon steel; above that temperature type 316L stainless steel should be used throughout.

Feed-effluent service should be considered to conserve heat. In this service shell pressure and the use of 20-ft. tubes 1 in. O.D. on 1 1/4 in.-sq. pitch will limit the maximum bundle size to about 2,700 sq. ft. All other units can be 3/4 in. O.D. tubes on 1 in. sq. pitch, 20-ft. bundle length with a maximum area of 4,600 sq. ft./shell.

Cooling water is available and should be used in preference to air cooling. Design cooling-water temperature is 90°F. with a maximum return temperature of 120°F. While U-tube, fixed-tube sheet, or multi-shell-pass types may not be used because of the requirements for mechanical cleaning, it should be noted that floating-head units, either single-pass or two-pass tubes with an internal expansion joint, are acceptable.

Do not calculate detailed design for two-phase flow in exchangers. Cost factors for shell and tube units and for fired equipment are available from our economics specialist.

MEETING

Present: Process Engineer (P.E.)
Distillation Specialist (D.S.)

D.S.: Distillation columns are required to effect the main separation of reactor products. Specifications for the principal process streams in the product-recovery section of the plant are as follows:

1. Stabilizer overhead gas
   Maximum benzene loss
   0.5% of benzene product
2. Benzene product spec. (nitration grade)
   Maximum toluene content
   0.03 vol. %
   (Assume that no more than 0.0001 mole % of benzene product is methane.)
3. Recycle toluene
   Maximum benzene content
   4.0 mole % of toluene in liquid recycle
   Maximum diphenyl content
   25 mole % of net diphenyl make
4. Toluene gas plant losses
   Use a 1% maximum total loss of net toluene feed in a combination of the phase-separator fuel gas, stabilizer overhead gas, and toluene tower bottoms.

We have a special situation when considering the stabilizer and the toluene towers. Existing towers are available for reuse which will be adequate in these applications.

First, let's consider the stabilizer. Column T-101, which is 7 ft. 6 in. in diameter by 65 ft. straight sides and contains twenty-four valve-type trays, will separate methane from the benzene product. Use a 2.0:1 operating reflux to overhead ratio and take the overhead off to fuel as a vapor.

P.E.: What product distribution should be assumed, and what operation pressure do you recommend?

D.S.: We can operate at 150 lb./sq. in. gauge at the bottom of the relocated stabilizer, because this separation is not difficult and most of the reboiler duty will serve to raise the temperature of the bottoms, which must eventually be heated to 450°F. to meet your clay-treating requirements. The design temperature and pressure you will be dealing with are within the ASME code stamp limitations on this vessel.

The distillate to bottoms (D/B) ratios you should achieve are as follows:

Stabilizer-Distillate/Bottoms Mole Ratio

Hydrogen	∞
Methane	50,000.0
Benzene	0.005
Toluene	0.001
Diphenyl	0

P.E.: What equilibrium data should I use to establish the bubble points, dew points, etc., before heat balancing the stabilizer?

D.S.: Use the following data, which have been taken from the Winn Chart (2) at a 5,000 lb./sq. in. gauge convergence pressure.

Winn Equilibrium Constants at 150 lb./sq. in. gauge and t°F.

t	100	150	200	300	400	500
Hydrogen	180	165	150	120	89	66
Methane	32	34	35	34	31	27
Ethane	3.5	5.0	6.5	9.2	11.3	12.0
Benzene	0.028	0.073	0.165	0.53	1.22	2.10
n-Heptane	0.015	0.042	0.102	0.36	0.92	1.65

We have used reference points for hydrogen and methane in the presence of benzene. Equilibrium constants for toluene and diphenyl can be obtained from the Winn Chart data by applying Hadden and Grayson's (3) calculation based on the following relationship:

$$K_{\substack{\text{toluene}\\ \text{or diphenyl}}} = K_{\text{n-heptane}} \Big/ \left(\frac{K_{\text{ethane}}}{K_{\text{n-heptane}}}\right)^{b}$$

where b is 0.08 for toluene and 1.29 for diphenyl.

Assume that the liquid feed from the phase separator flashes isenthalpically into the stabilizer with less than 1% vaporization.

P.E.: Are there any special considerations for clay treating the product?

D.S.: The total liquid yield from the stabilizer is clay treated to produce a high-quality benzene. Consider the clay-treating towers to be a package design with a purchased cost of $45,000 for this plant independent of conversion per pass. You must provide for pumping, heating, and cooling the liquid as required.

The operating temperature of the clay treater is 450°F., and the stabilizer bottoms should pass through the treater as a liquid at an inlet pressure of 300 lb./sq. in. gauge.

Clay consumption is 0.5 lb. clay/bbl. benzene product. Replacement clay costs $65/ton.

You must completely design the benzene tower. Use the stabilizer bottoms as feed and the stream specs from the start of our discussion to establish the material balance around the tower.

Use an operating pressure of 10 lb./sq. in. gauge at the bottom of the column. This low pressure will maximize the relative volatility of the benzene/toluene system short of using a complicated vacuum system. Assume Raoult's and Dalton's laws apply and determine the optimum number of theoretical trays and reflux ratio using the McCabe-Thiele (4) graphical method.

Vapor-pressure data for benzene and toluene which has been calculated from the Antoine equation and constants given in API Research Project 44 are as follow:

Vapor Pressure, Pure Component

°F.	Benzene, lb./sq. in. abs.	Toluene, lb./sq. in. abs.
100	3.225	0.104
150	9.180	3.360
200	21.719	8.805
250	44.672	19.721
300	—	39.114

Enthalpy data for the design of this part of the plant are given as follow:

Enthalpy, B.t.u./lb.

Temp., °F.		60	100	200	300	400	500
Hydrogen		1,748	1,885	2,229	2,574	2,920	3,267
Methane	(V)	260	281	337	397	462	532
Methane	(L)	182	204	260	321	387	459
Benzene	(V)	274	284	313	347	385	427
Benzene	(L)	87	105	148	197	250	310
Toluene	(V)	280	290	321	357	397	441
Toluene	(L)	103	117	160	209	267	335
Diphenyl	(V)	—	295	334	375	424	473
Diphenyl	(L)	—	140	186	237	296	355

Use valve-type trays at an overall efficiency of 60% for single-pass operation at a 24 in. tray spacing. To determine the tower diameter, calculate the vapor loading at the top and bottom trays using the following criteria:

V_s max. allowable vapor velocity through total tower cross-sectional area, ft./sec.

$$V_s = 0.17\sqrt{\frac{\rho L - \rho V}{\rho V}}$$

where ρV = operating vapor density, lb./cu. ft.
ρL = operating liquid density, lb./cu. ft.

Existing column T-123, which is 6 ft. 0 in. in diameter by 31 ft. straight sides and contains twelve valve-type trays, will separate toluene from diphenyl according to the product specifications. Use a 1.5:1 operating-reflux—to—overhead ratio and totally condense the overhead. The pressure at the bottom of the column will be set so that the boiling point of the bottoms is 400°F. (50°F. less than available steam temperature). The ASME code stamp for this vessel has a design pressure of 75 lb./sq. in. gauge and a design temperature of 650°F. To simplify the heat-balance calculations, assume that the dew point of the overhead vapor is 275°F. and that the boiling point of this stream is 250°F.

You might use the following guidelines for miscellaneous equipment:

1. Recycle compressor. Estimate the adiabatic horsepower requirements and use 75% efficiency to calculate the brake horsepower. Provide two compressors with each handling 60% of the recycle gas.
2. Pumps. Size all pumps for 120% of the operating gallons per minute and a Δp estimated by using allowable equipment Δp's already stated plus a 15% allowance for control and lines losses in Bernoulli's equation.
3. Vessels. Use 2:1 elliptical heads on all vessels. Calculate the thickness of the shell and heads on the reactor, phase separator, and benzene tower to establish their cost on our charts. Use carbon steel at a thickness of 1/2 in. for the shell and heads of the stabilizer reflux drum.

For other reflux drums and liquid-feed storage drum consider the shell and heads to be 1/4-in.-thick carbon steel. Make these vessels horizontal with a length-to-diameter ratio of about 3.

Size the phase separator and other drums, using whichever of the following criteria that is controlling:

Liquid loading: 10 min. of liquid holdup when half full

Vapor loading:

Allowable design vapor velocity, ft./sec.

$$= 0.17\sqrt{\frac{\rho L - \rho V}{\rho V}}$$

Provide one day of holdup in the liquid-feed storage drum.

MEMORANDUM

To: Process Engineer
From: Economics Specialist
Subject: Design Factors and Cost Information for Toluene Demethylation Cost Estimate

As we discussed yesterday, I am forwarding to you the design and cost factors you will need in preparing your cost estimate and the associated economics. The estimate you have been asked to submit is based on major-equipment-purchased costs, from which you will factor the overall plant costs.

Vessel Design

1. Table 2 provides the formulas for calculating the wall thickness required for cylindrical shells and ellipsoidal heads along with the allowable stress values for the recommended materials of construction at the design temperatures.
2. In designing your pressure vessels, you should select a design pressure of 10% or 25 lb./sq. in. gauge greater than the maximum expected operating pressure (whichever is greater).
3. In all vessels and towers except the reactor, which is lined, add 1/8-in. corrosion allowance to the calculated thickness.
4. Since the plate for rolling your vessels comes in standard thicknesses, you should round upward any nonstandard wall thickness according to the following schedule of mill availability:

1/16-in. increments from 3/16 to 1/2 in. inclusive
1/8-in. increments from 5/8 to 2 in. inclusive
1/4-in. increments from 2¼ to 3 in. inclusive

Major-Equipment-Purchased Costs Factors

1. Reactor (SA-287 Gr. B, 1 Cr-½Mo Steel)
   a. On the basis of the reactor configuration you determine, calculate the required wall thickness for the shell and heads. Assume that the 6-in. thickness of internal insulation will limit the inside metal wall temperature to 900°F. From Figures 2 and 3 you can determine the approximate weights of the shell, heads, and supporting skirt.
   b. Add 25% to this weight for connections and miscellaneous factors.
   c. Price at $1.00/lb. (1966).
   d. Add $4.00/sq. ft. for the castable refractory used to insulate the reactor internally.
2. Distillation Columns (SA-285 Gr. C Carbon Steel)

Although you have to design only one tower, the other two existing and being available, you should nevertheless determine the cost for all three. This is done for two reasons: first, you can assume that the estimated cost will cover the relocation expenses, and, second, because the total plant cost is to be factored from the major equipment items,

TABLE 2. THICKNESS OF SHELLS AND HEADS UNDER INTERNAL PRESSURE

(Ref.: ASME Boiler and Pressure Vessel Code, Section VIII, Unfired Pressure Vessels, 1965)

Cylindrical Shells (Paragraph UG-27)

$$t = \frac{PR}{SE - 0.6P}$$

where t = minimum required thickness of shell plates, exclusive of corrosion allowance, in.
P = design pressure, lb./sq. in.
R = inside radius of shell course under consideration, before corrosion allowance is added, in.
S = maximum allowable stress value, lb./sq. in.
E = joint efficiency

(Note: For butt joints attained by double welding and examined by fully radiographing, E = 1.00)

Ellipsoidal Heads (Paragraph UG-32)

The required thickness of a dished head of semiellipsoidal form, in which half the minor axis (inside depth of the head minus the skirt) equals one fourth of the inside diameter of the head skirt, shall be determined by

$$t = \frac{PD}{2SE - 0.2P}$$

where D = inside diameter of head skirt, in.
E = joint efficiency
(For seamless heads, use E = 1)

Other terms same as above

Maximum Allowable Stress Values in Tension in Pounds per Square Inch

	Towers and vessels	Reactor
Material and specification number	SA-285	SA-387
Grade	C	B
Nominal composition	Carbon steel	1 Cr – ½ Mo
Minimum tensile, lb./sq. in.	55,000	60,000
Maximum allowable stress, lb./sq. in.	13,750*	13,100†

*For metal temperature not exceeding 650°F.
†For metal temperature not exceeding 900°F.

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your base, without the two towers, would be low. For purposes of calculating return on investment and payout, assume that the towers, while existing, have never been used and that therefore the full depreciation will be picked up by the toluene demethylation plant.

a. Once again using Figures 2 and 3, determine the weight of the shell, heads, and skirt.
b. Add 20% for connections, base ring, tray supports, etc.
c. Price at $0.40/lb. (1966).
d. Trays
   (1) Calculate the total tray area.
   (2) Price at $6.00/sq. ft. (1966).

3. Miscellaneous Surge Tanks (Carbon Steel, Horizontal, No Internals)

a. Except for the vessels which have specific pressure requirements, use the following table of diameter vs. wall thickness to set the thickness:

Vessel diameter, in.	Minimum wall thickness,* in.
48	1/4
48–72	5/16
72–96	5/8
96–120	7/16
120–144	1/2

*Including ⅛-in. corrosion allowance

b. Using Figures 2 and 3, determine the weight of the shell and heads.
c. Add 15% for connections, saddles, etc.
d. Price at $0.50/lb. (1966).

4. Shell and Tube Heat Exchangers

a. Determine the 1958 cost from Chilton (5,

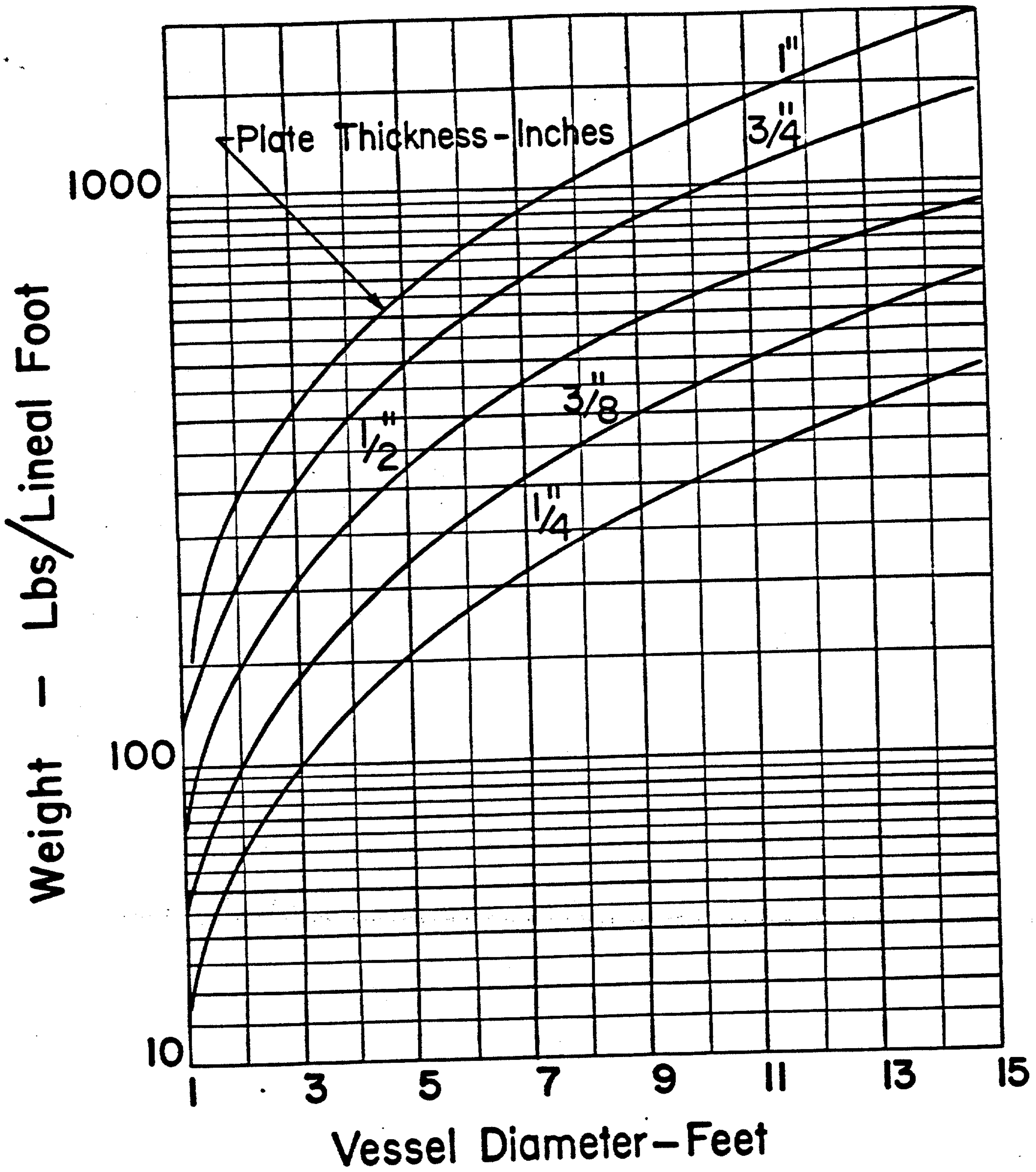

Figure 2. Approximate Weight of Cylindrical Vessel Shell. [Reproduced with permission from Petrol. Refiner, 39, 120 (July, 1960).]

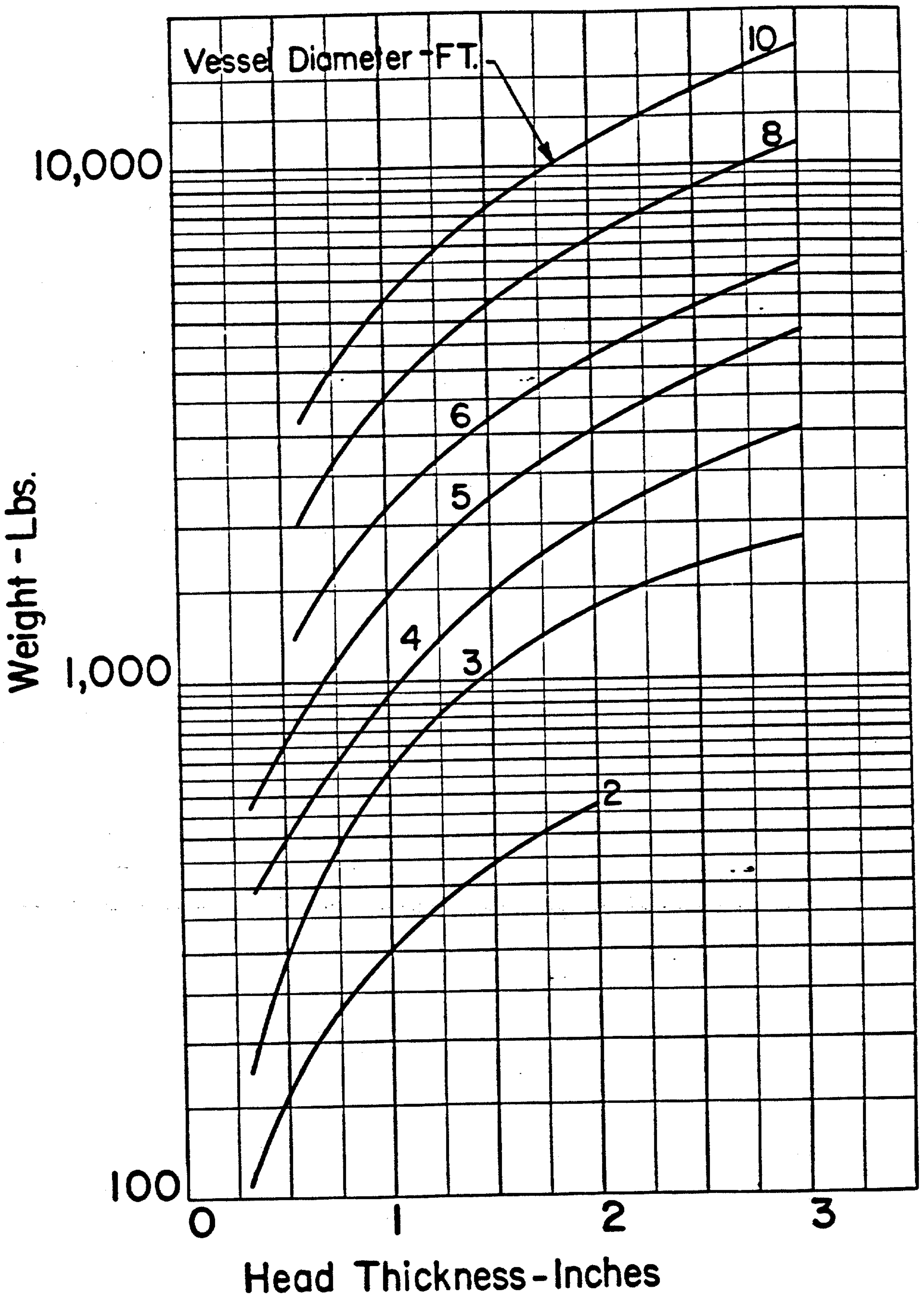

Figure 3. Approximate Weight of Elliptical Heads. [Reproduced by permission from Petrol. Refiner, 39, 120 (July, 1960).]

TABLE 3. COST FACTORS

Hourly labor	$4.00/man-hr.
Labor overhead	35% of hourly labor
Supervision	25% of hourly labor
Supervision overhead	35% of supervision
Depreciation	10% of investment/yr.
Maintenance	3% of investment/yr.
Taxes and insurance	1% of investment/yr.

TABLE 4. UTILITIES—FULL COST*

Fuel	$0.40/million B.t.u.
Steam	0.75/1,000 lb.
Cooling tower water (including makeup)	0.02/(1,000 gal./min. circulated)
Electric power	0.01/kw.-hr.

*Costs include all costs associated with utilities including payout of investment.

p. 119) for the calculated area; adjust the cost for other factors as indicated.

b. Update to 1966 cost as follows:

$$\frac{\text{230 (estimated 1966 Nelson heat exchanger index)}}{\text{181 (1958 Nelson heat exchanger index—composite)}}$$

5. Kettle-type Reboiler
   a. Determine the 1953 cost from Chilton (5, p. 134) for the calculated area; adjust the cost for shell and tube material factors, if necessary, by factors of Figure 3.
   b. Update to 1966 cost as follows:

$$\frac{\text{230 (estimated 1966 Nelson heat exchanger index)}}{\text{176 (1953 Nelson heat exchanger index—composite)}}$$

6. Compressor (Multistage, Reciprocating)
   a. Determine the base cost from Chilton (5, p. 56).
   b. Since this value is an installed cost, multiply by 0.5 to exclude foundations, electricity, instrumentation, building materials, painting, piping, installation labor, etc. The remaining cost is intended to cover only the compressor, driver, coolers, lubricating equipment, and other directly associated items.
   c. Update to 1966 cost as follows:

$$\frac{\text{260 (estimated 1966 Nelson compressor index)}}{\text{100 (1946 Nelson base year)}}$$

7. Fired Heater
   a. Determine the base installed cost from Chilton (5, p. 57) and factor by 0.8 for similar reasons to these described in item 6.
   b. Update to 1966 cost as follows:

$$\frac{\text{230 (estimated 1966 Nelson heat exchanger index)}}{\text{100 (1946 Nelson base year)}}$$

8. Pumps (Horizontal, Centrifugal, Carbon Steel)
   a. On the basis of flow and head requirements, calculate the required horsepower.
   b. Use the following table to determine 1961 cost:

Size	Unit	Purchased Cost (6) January 1961	Size Range	Exponent
0.75	hp.	660	0.75–5	0.08
5	hp.	770	5–25	0.22
25	hp.	1,200	25–50	0.66

   c. Update to 1966 cost as follows:

$$\frac{\text{260 (estimated 1966 Nelson pump index)}}{\text{229 (1961 Nelson pump index)}}$$

Total Plant Cost, Operating Costs and Economic Criteria

1. The total battery-limits (or on-site) costs may be estimated as 4.0 times the major-equipment-purchased costs.

2. The off-sites costs may in this case be estimated as 15% of the on-site cost since the required investment in facilities for steam generation, electrical distribution, and cooling towers has been included in the unit costs factors. (See Table 4.)

3. Operating cost factors are given in Tables 3 and 4.

4. The economic criteria you should calculate for the appropriation request are return on initial investment (ROI), based on the sum of fixed and working capital (which you will have to estimate), and cash flow payout (CFPO).

5. Examine the sensitivity of these economic criteria to benzene selling prices over the range of 24 to 28¢/gal. Do not revise your conversion optimization calculations as a result of this sensitivity check. Simply temper your recommendations to management.

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