(B)—PART OF PROCESS STEAM FURNISHED DIRECT FROM BOILER

RANKINE CYCLE

(where "x" = lbs./hr. of process steam bled from turbine)

Part of the steam (109,500 — x) #/hr. is furnished directly to the process from the boiler (after throttling) and the remaining process steam is bled from the turbines which also generate the necessary amount of power. By Equation 1 (Appendix I) the steam rate to turbines is:

$$M_t = \frac{(45{,}430{,}000) + (x)(H_p - H_c)}{(H_b - H_c)}$$

The total steam rate is:

$$\frac{(45{,}430{,}000)}{(H_b - H_o)} + (109{,}500) + (x)\left[\frac{(H_p - H_c)}{(H_b - H_c)} - 1\right]$$

Differentiating with respect to the variable amount bled from the turbine, (x):

$$\frac{d(M_{total})}{d(x)} = \frac{(H_p - H_c)}{(H_b - H_c)} - 1$$

Since $(H_p - H_c)$ is always less than $(H_b - H_c)$ the right hand side of the equation is negative, showing that the total steam rate decreases as the amount of steam bled from the turbine increases. The total steam rate is obviously a minimum when $x = 109{,}500$ #/hr.; in other words, when no steam is reduced to process conditions directly from the boiler.

Similarly, the steam rate to the condenser is:

$$M_{cond.} = M_t - (x)\,;\quad \frac{d(M_{cond.})}{d(x)} = \frac{d(M_t)}{d(x)} - 1 = \text{negative}$$

showing that the condenser capacity is decreased as the fraction bled from the turbine is increased.

The heat rate will be correspondingly increased if steam is reduced directly to process from boiler conditions, due to increase in total steam rate.

CONTEST PROBLEM

1938

STUDENT CHAPTERS—AMERICAN INSTITUTE OF CHEMICAL ENGINEERS

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

COMMERCIAL PRODUCTION OF DIPHENYL

STATEMENT

In order to obtain preliminary operating cost estimates it is desired to work out the process design of a commercial plant for the production of 8 tons of crude diphenyl per operating day by the pyrolysis of benzene.

As a basis for the design, laboratory pilot plant experiments were carried out in an apparatus designed rapidly to preheat benzene vapor to the desired reaction temperature substantially without decomposition and discharge it into a reactor tube maintained at a constant temperature and pressure. The reactor tube was 37.5 inches long and had an internal diameter of 0.500 inches. The partially decomposed charge was withdrawn from the end of the reactor and cooled rapidly to prevent further decomposition. The liquid mixture was analyzed for benzene, diphenyl, and higher polybenzenes. It will be assumed that the higher polybenzenes are all triphenyl (diphenyl benzene). The following data* were obtained.

In developing and applying reaction velocity equations for reactions of this type it is suggested that graphical differentiations and step-by-step graphical integration be employed rather than attempting rigorous analytical solutions. It may be assumed that the standard molal heat of reaction for the formation of triphenyl from benzene and diphenyl is the same as that for the formation of diphenyl from benzene. The specific heat of diphenyl vapor may be taken as 97% of that of benzene at the same temperature and that of triphenyl vapor as 95% of that of

* The data in the following table are selected for illustrative purposes, and do not necessarily correspond to the results obtained in any actual pilot plant.

benzene. Liquid triphenyl may be assumed to have the same specific heat as liquid diphenyl.

The commercial plant is to produce crude diphenyl containing not over 15% triphenyl directly from the reaction. For construction of the heater and reactor an alloy is available which is supplied in the form of tubes having inside diameters of 1 or 2 inches and lengths up to 20 feet. These tubes are capable of satisfactory operation at metal temperatures not in excess of 1375° F. With heat input rates less than 10,000 B. t. u. per sq. ft. per hr., differences in metal temperature between the inside and outside of the tube may be neglected. To minimize tar formation difficulties the pressure at the outlet of the reactor should be 15 lb. per sq. in. gauge.

REACTION TEMPERATURE: 1265°F. Pressure, 1.0 Atm.

Feed rate liters/hr.	Composition of reaction mixture, % by wt.		
	C_6H_6	$C_{12}H_{10}$	$C_{18}H_{14}$
0.846	83.0	14.6	2.4
0.423	70.8	22.4	6.8
0.282	62.5	26.2	11.3
0.212	56.8	27.8	15.4
0.141	50.3	29.2	20.5
0.121	48.6	29.4	22.0
.106	47.4	29.4	23.2
.0282	44.7	29.4	25.9
.0141	44.7	29.4	25.9
REACTION TEMPERATURE: 1265°F. Pressure, 0.5 Atm.			
0.282	86.6	11.9	1.5
REACTION TEMPERATURE: 1400°F. Pressure, 1.0 Atm.			
7.75	84.3	13.7	2.0

In preliminary process cost estimates of this type which are generally carried out for the purpose of predicting commercial feasibility and making overall comparisons of different types of operations, it is common practice to make arbitrary assumptions in establishing design bases for incidental items of equipment where the resultant deviation from the most economical design could not be sufficient to affect significantly the overall economics of the process. For example, in this design, it may be assumed that heat exchange is economical where minimum temperature differences greater than 50° F. are involved.

However, no heat exchanger transferring less than 200,000 B. t. u. per hr. should be considered, and the maximum temperature in any heat exchanger should not exceed 900° F.

The following overall heat transfer coefficients may be used for exchangers and coolers:

High temperature vapor-liquid exchangers	30	B. t. u./hr./sq. ft./°F.
Condensing vapor-hydrogen cooler	70	
Benzene vapor condenser	90	
Liquid coolers	50	

With these coefficients, pressure drops through the exchangers and coolers may be neglected.

Fractional distillation shall be carried out with a ratio of hot reflux to net overhead product equal to 0.2 to produce bottoms containing 0.1 mol % of overhead and overhead product containing 0.1 mol % bottoms. In the design of the fractionating column the effect of the triphenyl in the bottom product may be neglected.

In the design of the furnace the temperature of combustion gases in contact with tubes should not be in excess of 2000° F., and the temperature of the flue gases leaving the furnace should be 400° F. higher than the temperature of the entering feed.

In more refined design calculations for actual construction, these assumptions could be replaced by the result of detailed calculations of optimum economic values.

The following basic unit costs shall be used:

Utilities:

Steam, 125 lbs./sq. in. gauge pressure	$ 0.25 per 1000 lbs.
Cooling water, 85° F.	$ 0.05 per 1000 gal.
Natural gas, 1000 B. t. u./cu. ft.	$ 0.30 per 1000 cu. ft.
Electricity	$ 0.008 per K. W. hr.

Labor:

Plant operators	$ 8.00 per 8 hr. day
Operators' helpers	$ 6.00 per 8 hr. day
Foreman	$ 10.00 per 8 hr. day

APPROXIMATE EQUIPMENT COSTS:

The following unit costs may be used as approximately correct for the type and scale of equipment used in this plant. All figures include the cost of erection, foundations, design and engineering charge, and contractor's profit.

HEATER AND REACTOR: $32.00 per internal square foot of tube surface.

FRACTIONATING COLUMNS: $45.00 per cubic foot of volume for operation at atmospheric pressure plus $9.25 per square foot of bubble deck.

HEAT EXCHANGERS: The price per square foot of surface may be taken as inversely proportional to the total area of the unit, ranging from $12.00 per external square foot for a 100 sq. ft. exchanger to $6.50 per square foot for a 400 sq. ft. exchanger.

COOLERS AND CONDENSERS: $14.00 per square foot external area.

ACCUMULATOR AND SEPARATOR VESSELS: $48.00 per cubic foot at atmospheric pressure. For higher pressures the cost is increased in proportion to the pressure.

PUMPS: For the relatively small rates involved motor driven centrifugal pumps may be used. For handling hot liquids up to 500° F. with a discharge pressure up to 100 lbs. per sq. in. a pump having a capacity of 15 gallons per minute costs $800.00. An efficiency of 50% may be assumed.

For handling liquids at a temperature below 300° F. the following pumps are available.

Max. Capacity g. p. m.	Discharge Pressure Lbs./sq. in.	Price	Efficiency
8.0–10	100	$450.00	35%
6.0–8.0	30	300.00	25%
2.0–4.0	30	300.00	20%

The costs and efficiencies of the required pumps may be estimated from these data.

PIPING, VALVES, AND INSTRUMENTS: Add 40.0% to the cost of the remainder of the equipment to cover all except automatic control instruments. Where automatic control is desired, costs should be added from the following schedule:

Automatic liquid level controller	$250.00
Automatic temperature recorder and controller	475.00
Automatic rate of flow recorder and controller	450.00
Automatic pressure recorder and controller	200.00
CONTROL HOUSE	$3,800.00

FIXED CHARGES: An annual charge of 35% of the total estimated cost of the plant shall be taken to include depreciation, maintenance, taxes, insurance, and obsolescence.

On the bases given above, prepare a complete process design and calculate the direct manufacturing cost of producing crude diphenyl from benzene costing $0.16 per gallon assigning no value to the hydrogen made as a by-product. The process design should include the following:

1. A complete flow diagram of the process including a brief description of the operation, and indicating the instruments and controllers specified.
2. A complete heat and weight balance of the entire process.
3. Specification of the size, number, and arrangement of the tubes in the furnace.
4. Specification of the areas of all heat exchangers, coolers, and condensers.
5. Specification of the number of plates, plate spacing, and cross-sectional area of the fractionating column.

It is intended that in solving this problem, emphasis be placed on proper design of the reaction system with secondary attention to the auxiliary separating equipment which has only a secondary influence on the overall economics of the process. Energies of activation of both the forward and reverse reactions are preferably calculated in order to derive time-conversion curves for the selected reaction temperature. In the design of a process of this type, it is justifiable to work out the heater and reaction system in considerable detail to insure against inoperability and the necessity of adopting radically different and more expensive operating conditions or types of equipment which might greatly affect the overall economics.

METHOD OF PRESENTATION

The solution should be presented in the form of a report summarizing the overall results and conclusions, and including at least samples of all detailed calculations. The form of the report should be such that the economic conclusions may be analyzed readily by a non-technical executive but should also include complete technical data to permit verification of all results. Where data from the literature are used, the references should be cited. All assumptions and approximations should be stated clearly and the justifications for them explained.

Reports will be graded on (a) conclusions reached, (b) accuracy of computations and (c) form of presentation.

It is to be assumed that the above statement of the problem contains all the data available and your instructor is not to be consulted in regard to doubtful points. The problem is not to be discussed with any person whatever until after March 15, 1938. This is particularly important in cases where neighboring institutions may not begin the problem until after its completion by another chapter. The use of text books, handbooks, journal articles, steam tables, and lecture notes is permitted. Submittal of a report for the competition implies adherence to the above conditions.

A period of not more than 21 consecutive days is allowed for completion of the solution. This period may be selected at the discretion of the individual counsellor, but a solution must be postmarked not later than midnight March 15, 1938, in order to be eligible. Each 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 Chairman of the Committee on Student Chapters. The solution itself must bear no reference to the student's name or institution by which it might be identified. Each counsellor should select the best solution, or solutions, from his chapter, not to exceed two in number, and send these registered mail to Professor Joseph C. Elgin, Princeton University, Princeton, New Jersey.

A.I.Ch.E. Annual Student Competition

FIRST PRIZE WINNING SOLUTION

Contest Problem, 1938, Student Chapters, A.I.Ch.E.

A PROCESS DESIGN FOR THE COMMERCIAL PRODUCTION OF DIPHENYL FROM BENZENE

BY ROBERT EGBERT, Cooper Union Chapter
Cooper Union, New York, N. Y.

INTRODUCTION

In order to obtain preliminary operating cost estimates for the commercial production of 8 tons of crude diphenyl containing not more than 15% triphenyl per day, a process design has been worked out. The design was based on pilot plant data, concerning the kinetics of the pyrolysis reactions of benzene. The crude diphenyl is separated from unreacted benzene in a fractionating column operated to produce a product containing less than .1 mol% of benzene.

CONCLUSIONS

The cost of producing a pound of crude diphenyl containing 14.5% triphenyl, is $.0360.

An operating day of 24 hours, employing three shifts of labor and a year composed of 300 operating days was chosen. On this basis the costs can be distributed as follows using one day as a basis:

Fixed charges	$45.09
Labor	138.00
Raw materials	361.50
Utilities	31.40
Total	$573.82
Product 16,000 lbs./day	
Unit cost	$0.0360

SUMMARY

I. Costs

Equipment costs are as follows:

1 benzene boiler	$1,435.00
1 heater	8,600.00