

PROBLEM

To the Contestant:

Time is an important element in industry. To the process engineer designing a new plant, it is particularly important. No profit can be made from a new plant until its product is sold. The design, fabrication, and construction are necessary steps but require time. The process designer is between two fires. He must design a plant low in both investment and operating cost, but he cannot optimize the design as completely as he may desire as he must also save time. He very often will have to complete the design on a date arbitrarily set for him. In this problem the contestant is the designer who must produce the best design in a given time. He must make calculations where required and, equally important, avoid making calculations that waste time. It is possible to spend an unreasonable amount of time on this problem, but it is also possible to reach a valid conclusion in a reasonable time.

The subject of this problem is the economical process design of a reactor system for a catalytic reformer. This is a process used to convert low-octane naphthas (heavy gasoline) to high-octane gasolines. The conversion occurs in the presence of hydrogen over a catalyst, usually platinum on alumina, at high temperatures and pressures. A typical fixed-bed reactor system for a catalytic reformer is shown in Figure 1.

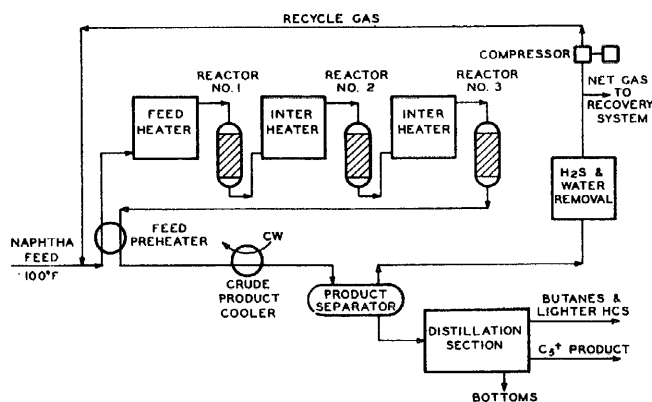


Fig. 1. Typical reactor system, catalytic reformer.

Three reactors are shown, although the number varies from two to five. More information on catalytic reforming is given in the appendix.

STATEMENT OF PROBLEM

The contestant is to design the optimum reactor system of a catalytic reformer whose capacity is 10,000 bbl./of naphtha feed operating day. A written report is required which gives the recommended reactor system and the supporting information for the recommendations. A flow sheet (similar to Figure 1) showing the number of reactors and interheaters, catalyst volumes and arrangement, and principal temperatures and pressures is to be included. One simple sketch of a typical reactor is also required. The report should clearly state what assumptions were made and what short-cuts were used. Calculations are to be included in an appendix with sample calculations detailed but the results tabulated.

The information required to solve this problem is presented in two sections, Process Information and Economic Evaluation. Most of the information is presented graphically to facilitate calculations. A description of catalytic reforming is included as an appendix. Reference to it is not required for the solution of the problem but will increase the contestant's understanding of catalytic reforming. The information contained in all the sections represents roughly the information the process designer in industry would be given to solve a similar problem. Some simplifications in pressure drop and cost data have been made to facilitate the solution. The actual data presented are reasonably representative of actual operations. In some cases, however, the curves indicate a completeness of data that is seldom available in industrial practice.

The design is to be evaluated for lowest initial and operating cost with maximum profit. These factors are combined into one called "pay-out period," which is the time required for the net income from the new plant to equal the initial investment.

One of the charts (Figure 8) shows types of reactors that have been used in commercial plants. The contestant is not limited to these types of reactors and may develop other designs.

These should be indicated in the simple sketch which is required. There is obviously no time for the contestant to show ingenuity by the design of mechanical details of construction. There is also no way to estimate relative costs of other designs. For the purpose of this problem, the cost curves provided may be used. The use of a tubular reactor with catalyst in tubes surrounded by a heating medium would appear attractive from a reaction-rate or heat-economy standpoint. However, this type has not been used commercially, and adequate consideration of tubular reactors is beyond the scope of this problem.

PROCESS INFORMATION

Feed Naphtha

Type of composition:	Weight %	Average molecular weight
Paraffins	65	130
Naphthenes	20	120
Aromatics	15	110
	100	

Octane rating: A measure of the antiknock quality determined in a laboratory engine by use of the F-1 method.

F-1 octane number of the clear naphtha	35
F-1 octane number with 3 ml. of tetraethyl lead antiknock fluid added to each gallon of naphtha	50

A. S. T. M. distillation: A simple batch laboratory distillation at atmospheric pressure universally used in gasoline testing.

	°F.
Initial boiling point	170
5 vol. % distilled over at	230
10	255
30	280
50	300
70	320
90	345
95	365
End point	385

Other properties:

A. P. I. gravity - specific gravity measured on scale developed by American Petroleum Institute, °A. P. I.	55
Average molecular weight	125
Critical temperature, °F.	625

Quantity: 10,000 bbl. (42 gal. at 60°F.) per operating day
Inlet temperature, °F. 100

Recycle Gas

This gas is 90% hydrogen by volume, and the remainder is assumed to be methane. The ratio of recycle gas to feed is 8 moles of hydrogen to 1 mole of naphtha feed. This ratio results in a vapor mixture in the reactors having a specific heat expressed as 1.2 B. t. u. / (°F. lb. of naphtha feed).

Reactor Inlet Condition

Maximum of 700 lb./sq. in. gauge pressure and 950°F. temperature.

Heat of Reaction

The endothermic heat of reaction is plotted against the reciprocal of L. H. S. V. for various average reactor temperatures and for average reactor pressures of 500 and 700 lb./sq. in.

gauge; L. H. S. V. is liquid hourly space velocity, a widely used reaction criterion in the petroleum industry. It is the hourly volume of naphtha feed corrected to 60°F. divided by the volume of the catalyst bed in consistent units. Its reciprocal (1/L. H. S. V.) is nearly proportional to the time of contact of the feed with catalyst in the reactor at any one specified recycle rate. However, strictly speaking, the liquid never contacts the catalyst as liquid or without the recycle stream added. Figure 2 shows the heat of reaction vs. 1/L. H. S. V. for an average reactor pressure of 500 lb./sq. in. gauge and average

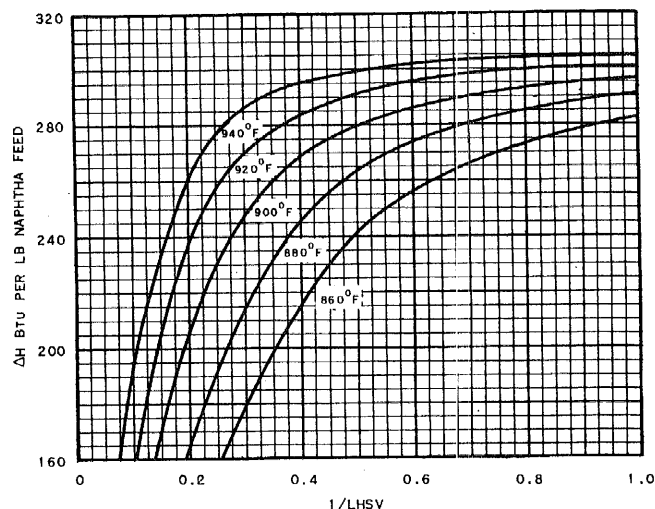


Fig. 2. Endothermic heat of reaction vs. space time (1/LHSV) at 500 lb./sq. in. gauge average reactor pressure for various average reactor temperatures.

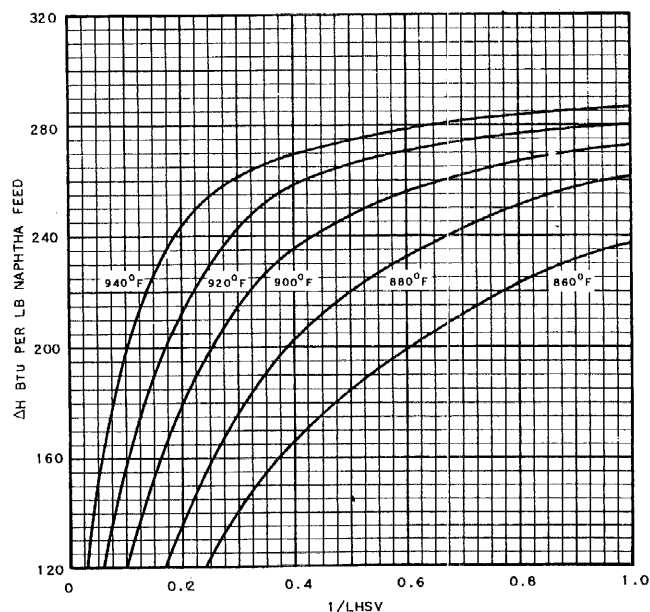


Fig. 3. Endothermic heat of reaction vs. space time (1/LHSV) at 700 lb./sq. in. gauge average reactor pressure for various average reactor temperatures.

reactor temperatures from 860° to 940°F. Figure 3 shows the same information at an average reactor pressure of 700 lb./sq. in. gauge. The curves given are based on laboratory and commercial data for the feed naphtha and are not valid for the product from the first or subsequent reactors.

Product and Product Quality

The amount and quality of product vary with reactor conditions. The amount is measured as liquid volume of C₅+ product and expressed as a percentage yield over liquid naphtha feed. The C₅+ designation refers to pentanes and heavier hydrocarbons, meaning that the lighter hydrocarbons (methane, ethane, propanes, and butanes) have been removed in the distillation section. (See Figure 1.) The product quality used in this problem is the antiknocking quality expressed as octane number. The octane number used is measured by the F-1 method with 3 ml. of tetraethyl lead fluid added to each gallon of C₅+ product. In Figure 4 the endothermic heat of reaction (ΔH) and octane number of the product (F-1 + 3 ml. T. E. L.) are plotted against liquid volume per cent yield of C₅+ product for an average reactor pressure of 500 lb./sq. in. gauge. In Figure 5 these are plotted at an average reactor pressure of 700 lb./sq. in. gauge. Again, this information is for the feed naphtha and not the product from the first or subsequent reactors.

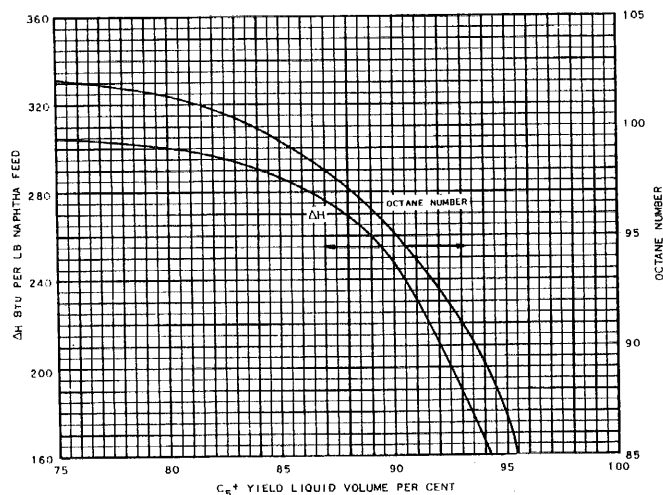


Fig. 4. C₅+ product yield vs. H and octane number (F-1 + 3 ml. TEL); 500 lb./sq. in. gauge reactor pressure.

Temperature Drop in Reactors

The temperature drops rather sharply in the first part of the catalyst bed, but the rate of drop decreases toward the end. The relationship between temperature drop and total catalyst volume is shown in Figure 6. The percentage of the total temperature drop that takes place in one plant of any number of reactors is plotted

against percentage of the total catalyst volume passed through. For example, if the total temperature drop is 200°F. and the first reactor contains 50% of the total catalyst volume, the temperature drop through the first reactor is 75% of the total, or 150°F. If there were only two reactors, the temperature drop through the second (with the last 50% of the total catalyst volume) would be 50°F. This curve is concerned only with the cumulative temperature drop that occurs in the reactors and does not include the temperature rise that occurs in the interheaters.

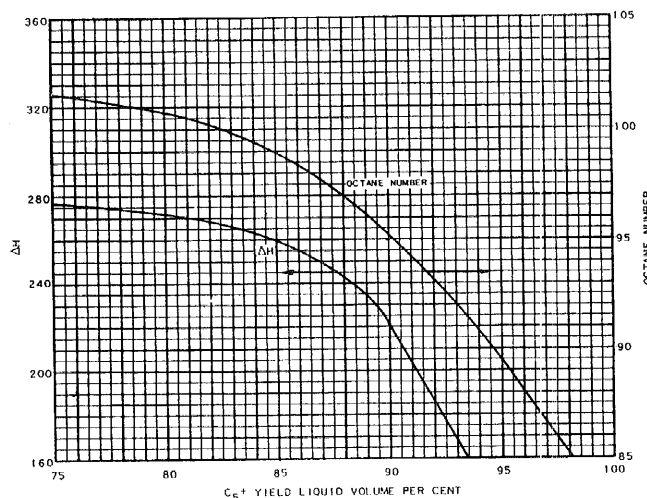


Fig. 5. C₅+ product yield vs. H B. t. u./lb. naphtha feed and octane number (F-1 + 3 ml. TEL); 700 lb./sq. in. gauge reactor pressure.

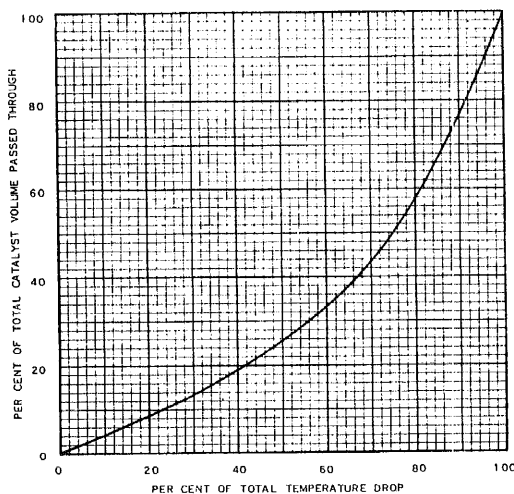


Fig. 6. Temperature-drop catalyst volume relationship.

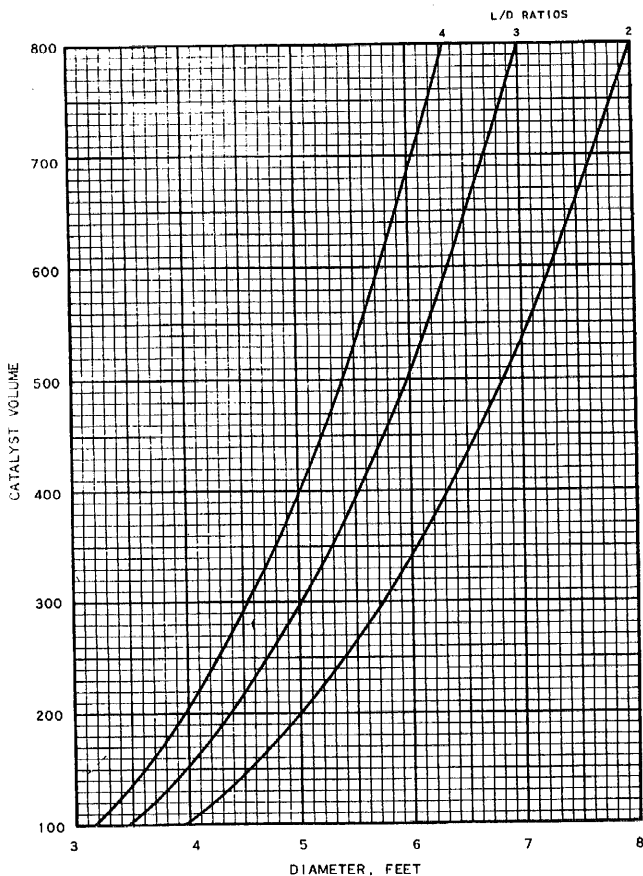


Fig. 7. Volume vs. diameter and length of cylindrical catalyst beds.

Type of Reactors

Several types of fixed-bed arrangements are possible in catalytic reformer reactors. Because of the operating pressure, all types are contained in cylindrical vessels, which usually have length-to-diameter ratios between 2 and 4. The catalyst volume contained in cylindrical catalyst beds is plotted in Figure 7 against bed diameter for length-to-diameter ratios of 2, 3, and 4. The usual types of reactors are shown in the attached Figure 8. The contestant is urged to develop any novel ideas of his own. A sketch of the reactor selected as optimum should be included to show the more important dimensions and any means included to ensure adequate vapor distribution if required.

Pressure Drop through Reactor System

All pressure drops will be assumed except the pressure drops through the reactor vessel and catalyst bed:

1. Through the heater and connected piping - 15 lb./sq. in. gauge reactor-heater combination.
2. On recycle line from product separator to naphtha inlet line - 20 lb./sq. in. gauge. (This is pressure drop only and does not, of course, include the pressure increase from compression.)

The pressure drops in the reactor vessel proper include:

1. Entrance and exit losses. (Assume a 12-in. I. D. pipe into and out of each reactor.)
2. Distribution baffles (It is very important that the flow of reaction mix be uniform across

TYPE	DESCRIPTION	REMARKS
I DOWNFLOW VERTICAL	 DISTRIBUTION BAFFLE CATALYST BED CATALYST SUPPORT GRID AND SCREEN	
II RADIAL FLOW	 ANNULAR ENTRANCE CHAMBER CATALYST BED BED SUPPORT COLLECTOR PIPE	COST IS SAME AS TYPE I FOR SAME CATALYST VOLUME. MUCH LESS PRESSURE DROP THROUGH CATALYST BED FOR SAME CATALYST VOLUME THAN TYPE I. ANNULAR SPACE ABOUT 2-3 INCHES ACROSS.
III DOWNFLOW HORIZONTAL	 CATALYST BED BED SUPPORT	COST IS CONSIDERABLY GREATER THAN TYPE I, BUT VERY LOW PRESSURE DROP.
IV	SAME AS TYPE I BUT TWO OR MORE BEDS IN SAME PRESSURE VESSEL	FOR SAME CATALYST VOLUME COST IS ABOUT SAME, BUT VESSEL LIKELY TO BE TOO HEAVY FOR EASY SHOP FABRICATION AND FIELD INSTALLATION.

Fig. 8. Types of fixed-bed reactors.

the entire catalyst bed so that each filament of the mix will have the same time-temperature history. If the pressure drop across the catalyst bed is too low, poor distribution of the reaction mix may result. This would require correction by baffles, screens, etc., to assure good distribution through the catalyst bed. Similarly, the catalyst support or outlet must be designed to assure uniform flow of the reaction mix through the catalyst bed.

Catalyst bed: The pressure drop through a catalyst bed is given by

$$\frac{\Delta P}{L} = 1.13 u^{1.73} \left(\frac{P}{T}\right)^{0.73}$$

$\frac{\Delta P}{L}$ = pressure drop in lb./sq. in. (ft.) of catalyst bed passed through

u = superficial velocity through space occupied by catalyst bed, ft./sec.

P = absolute pressure, lb./sq. in.

T = absolute temperature, °R.

This equation was derived from Brown* for one of the types and sizes of catalyst used in catalytic reformer reactors.

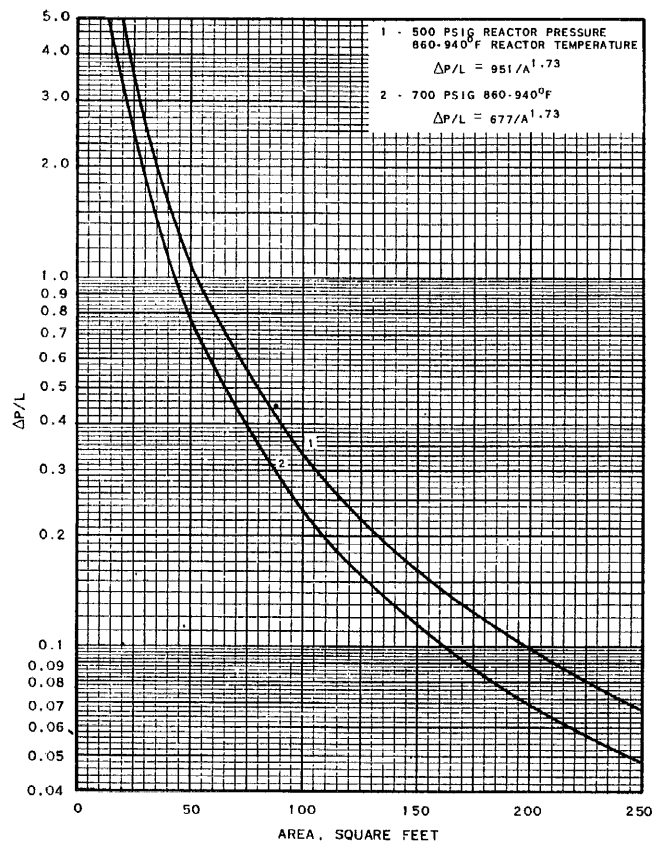


Fig. 9. Pressure drop through catalyst beds P/L lb./sq. in. (ft.) of bed depth vs. cross-sectional area, normal to flow, sq. ft.

*Brown, G. G., "Unit Operations," Chap. 16, John Wiley and Sons, Inc., New York ().

For the feed and recycle rate in this problem, the pressure drop per foot of catalyst bed passed through is plotted in Figure 9 against cross-sectional area of catalyst bed perpendicular to direction of flow.

Support or holddown screens: As the catalyst pellets are about 1/8 to 3/16 in. in their principal dimensions, the catalyst bed will have to be supported or restricted by screens. These are in turn supported by heavy structural members. The pressure drop through the screens and structural members will be neglected.

Collector pipe (radial flow reactors): The pipe must contain holes through which the reaction mix passes to leave the reactor. The total pressure drop through these holes is two and one-half velocity heads (velocity through hole). The collector pipe is 16 in. I.D. in all cases.

Miscellaneous Information

1. Temperature of feed out of feed preheater is 100°F. less than reaction mix into preheater.
2. No heat losses are assumed.
3. Drop in pressure through reactors is assumed not to affect conversion in subsequent reactors.

ECONOMIC EVALUATION

The analysis of the reactor alternatives will be limited to the major variable investment and operating cost items. Several minor items of cost are neglected, and simple curves of cost for the major items are given. The minor cost elements that are neglected in this analysis are as follows:

1. Feed preheater: The heat load on this unit will vary slightly with yield of C_5+ product.
2. Crude product cooler: The heat load on this unit will vary slightly with yield of C_5+ product.
3. Recovery system and distillation section: The investment and operating cost of these sections will vary somewhat with C_5+ product yield.
4. Naphtha feed pump: There would be some variation in operating and investment cost for varying reactor inlet pressures, but these will be neglected.

The choice between reactor alternatives will be based on minimum payout period. This period is defined as

$$\text{Payout period} = \frac{\text{total investment costs}}{\text{annual income} - \text{annual operating costs}}$$

Total investment costs, annual income, and annual operating costs will be estimated as outlined in the following section. No payout periods longer than two years will be considered. The optimum design will be that having the minimum payout period, less than two years.

The operating factor (percentage of time operating) for the catalytic reformer is 90%.

Investment costs

Simple curves are given for the cost of the following major equipment items.

A. Reactor costs: Reactor costs are plotted in Figure 10 as total installed cost (vessel, internals, field installation, including foundations but not catalyst) vs. catalyst volume in cubic feet. Several simplifications have been made in order to reduce reactor costs to this simple curve, but it approximates commercial practice. The reactor vessel has been assumed to be cylindrical and designed for 750 lb./sq. in. gauge maximum operating pressure at 950°F., according to the latest revision of the A. S. M. E. Code for Unfired Pressure Vessels.

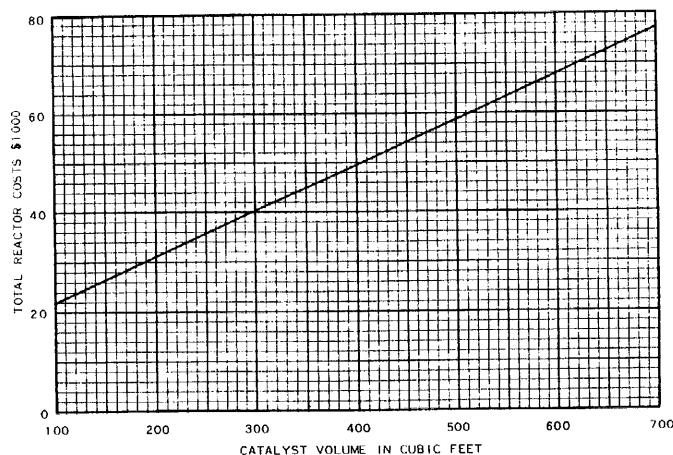


Fig. 10. Reactor costs vs. catalyst volume.

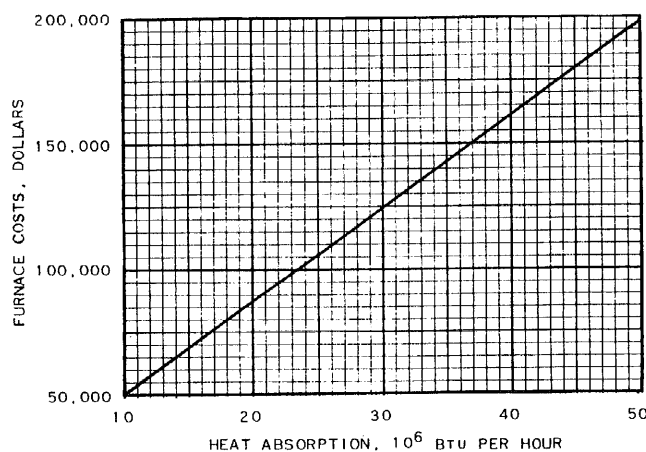


Fig. 11. Furnace costs vs. heat absorption, B. t. u. /hr.

B. Furnace costs: Furnace costs are plotted in Figure 11 as a total installed cost vs. heat absorbed in British thermal units per hour. The costs include a furnace complete with low-alloy tubes and headers for the heating element, gas burners, firebox, all structural elements, including foundation and individual stacks. The costs are reasonably accurate in the range of 10 to 50 million B. t. u. /hr. absorption.

C. Recycle gas compressor: The installed cost of a gas-driven reciprocating compressor for circulation of recycle gas is plotted in Figure 12 vs. horsepower rating of the gas engine. The selection of a gas-engine drive is based on the excess process gas produced by the catalytic reformer. The costs include the cost of installation but exclude the cost of compressor buildings and auxiliary equipment, which will not vary significantly with compressor size. These latter costs are to be disregarded.

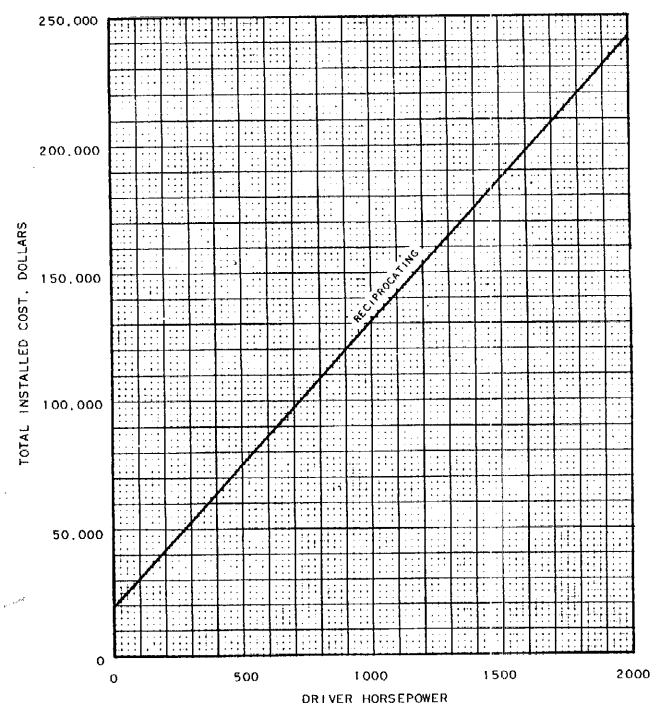


Fig. 12. Compressor costs vs. driver-horsepower installed cost.

D. Piping cost: For each piping hookup between a reactor and a furnace, a cost of \$3,000 is to be included.

E. Other costs: In order to account for contractor's overhead and profit, minor material, and various other variable cost elements, the total costs from A to D, inclusive, are to be increased by 50%.

F. Total investment cost: The sum of items A through E gives the total costs that will be affected by the variables in this problem. However, this sum is not the total cost of the catalytic reformer required for a 10,000-bbl./operating day feed rate. To the sum of items A through E is to be added \$2,200,000 to give the total cost of the catalytic reformer.

Annual Income

The only annual income that will be considered is the increase in value of the C₅+ product because of its increase in octane number. In Figure 13 the value of product less cost of feed in cents divided by gallons of product is plotted

against the octane number of the product (F-1 plus 3 ml. T. E. L. fluid).

Operating Costs

A. Catalyst cost: The cost of catalyst will be treated as an operating cost rather than as a combination of investment and operating cost. The catalyst cost will vary with severity of reforming, being higher as severity increases. "Severity" in this problem will be measured as octane number of product; the higher the severity the higher the octane number of the product.

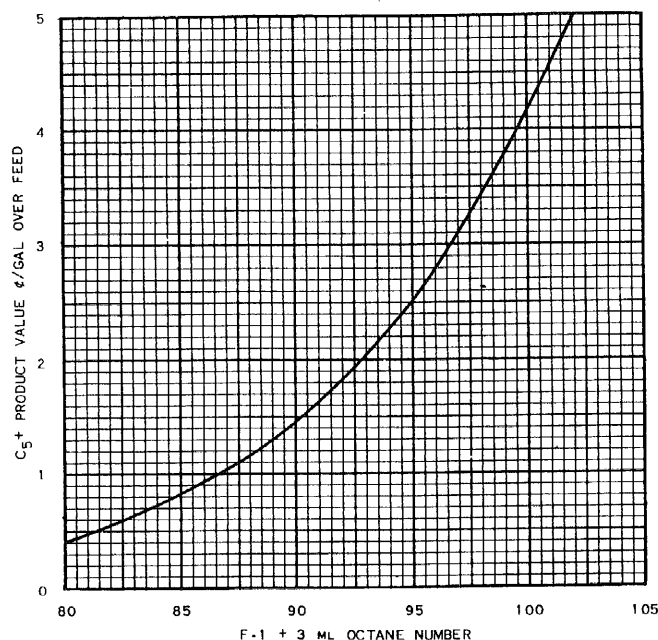


Fig. 13. Product value over feed value of 1 gal. of product (C₅+) over 1 gal. of feed C₅+ product.

However, the higher the octane number is, the lower will be the product yield. In Figure 14 the catalyst cost as cents per barrel of product made is plotted against octane number of product.

B. Compressor operating cost: A gas-engine-driven compressor will require 12 cu. ft. of gas (1,050 B. t. u./cu. ft.)/theoretical horsepower hour. The compressor will probably use process gas from the catalytic reformer, but a heat equivalent amount of natural gas will be used elsewhere within the refinery where the catalytic reformer is located. The cost of gas is 25 cents/million B. t. u.

C. Maintenance costs: Maintenance costs are to be estimated at 10% of total investment cost per year.

D. Furnace-fuel cost: The feed heater and interheaters will similarly burn process gas but are to be charged for the fuel burned at 25 cents/million B. t. u. Because of the high inlet temperatures, the furnaces have a low efficiency, 55%.

E. Other costs: In order to account for other costs, which will not vary appreciably, \$250,000/

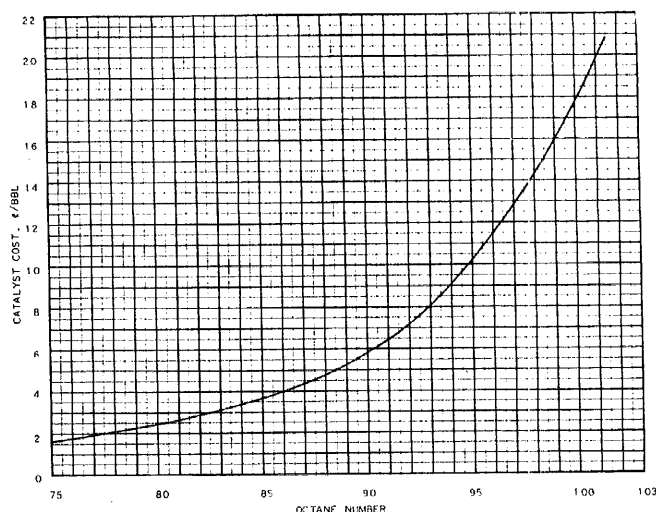


Fig. 14. Catalyst cost in cents per barrel of product vs. octane number of product (F-1 + 3 ml. TEL).

year is to be added to the sum of the operating costs from items A through D. These other costs include taxes, operating labor, laboratory service, etc.

APPENDIX

Catalytic Reforming

Catalytic reforming is a process that converts low-octane gasoline stocks to high-octane fractions. The conversion occurs in the presence of a catalyst, usually platinum on alumina, at pressures from 200 to 900 lb./sq. in. gauge and temperatures from 800° to 950°F. A hydrogen-rich recycle stream is combined with the vaporized feed to suppress polymerization of thermally cracked products and thus avoid coke deposition. Five principal chemical reactions occur during the conversion.

1. Dehydrogenation of naphthenes to form aromatics: An example of this reaction would be the removal of six hydrogen atoms from cyclohexane (C₆H₁₂) to form benzene (C₆H₆).

2. Dehydrocyclization of paraffins to form naphthenes and aromatics: An example of this reaction would be the removal of two hydrogen atoms from normal hexane to form cyclohexane, which is then dehydrogenated to benzene.

3. Hydrocracking of high-molecular-weight paraffins to form lower paraffins: An example of this reaction would be the splitting of dodecane (C₁₂H₂₆) into two hexane molecules (C₆H₁₄) requiring 1 molecule of hydrogen.

4. Isomerization of Naphthenes: An example of this reaction would be the isomerization of methylcyclopentane to cyclohexane, which would then go to benzene. (See item 1.) Some isomerization of paraffins to other paraffins does occur but is of less importance than the example given.

5. Desulfurization: Sulfur in the feed, usually present as disulfides and mercaptans, is removed in the presence of hydrogen to form hydrocarbons and H₂S (which is removed later from the recycle gas).

The heart of a catalytic reformer is the reactor system. Most of the plants built in the United States contain fixed-bed multiple reactors with intermediate heaters to supply the endothermic heat of reaction. In most of the fixed-bed reactors the catalyst is not regenerated in place. Units with this type of reactor are named Houdriformers, Platformers, and Sovaformers. Some of the fixed-bed reactors contain catalyst that is regenerated in place. Units of this type are named Catformers, Hydroformers, Sinclair-Baker Catalytic Reformers, and Ultraformers. There are some catalytic reformers with reactors of the moving-bed or fluidized types. These are called Fluid Hydroformers, Orthoformers, Thermoformers, and Hyperformers.

The growth of catalytic reforming has been more rapid than any of the other preceding major processes - catalytic cracking, thermal cracking, or thermal reforming. Most of its growth has occurred during the last five years and has been caused by the demand for more gasoline of higher octane rating. Its total capacity as gasoline produced has already exceeded thermal reforming and will soon pass thermal cracking. Within less than five years it may exceed catalytic cracking capacity. Also, many refineries now include catalytic reformers which process heavy thermally or catalytically cracked gasolines for sulfur reduction, as well as octane improvement. A more complete discussion of catalytic reforming will be found in a special section of *Petroleum Processing* for August, 1955.

JUDGES' COMMENTS

Judging of the sixty-eight solutions submitted for the 1956 problem proved to be an interesting and rather difficult assignment. There was a wide variety of approaches used and ideas presented. The solutions were of over-all high quality, and the principal numerical results agreed closely.

The problems were screened to select about twenty that were well above the average. These solutions were audited by one reviewer to assure consistency among those doing the screening. The selected group was then reviewed carefully by two reviewers to select the eight best. These were then critically reviewed by two additional engineers who had not previously participated in the grading. The final selections were the consensus of the four engineers who had reviewed each of the eight solutions carefully.

Comparison of the sixty-eight solutions revealed some interesting variations of interpretations. Different methods were used for averaging reactor temperature and pressure. Equal temperature drop in each reactor was considered best by the judges. Almost all the contestants quickly realized that the reactor temperature should be as high as possible. Not so many saw immediately that the product should be above 100 octane number. Fewer still made a simple analysis to prove this. Summaries contained most of

the important numerical results but a few completely omitted the octane number or buried the payout period in the appendix. Few numerical errors were found, although one contestant slipped the decimal two places to the right and another two to the left. Most payout periods were less than 0.9 year. The contestants showed good understanding of the solution of a chemical engineering problem of a type frequently met by young engineers.

SOLUTION

SUMMARY OF RECOMMENDATIONS

The recommended reactor system for the catalytic reforming operation under consideration is one containing three radial-flow reactors operating at 500 lb./sq. in. gauge and 908°F.

The optimum total catalyst volume of these beds is 1,848 cu. ft., corresponding to a Liquid Hourly Space Velocity (L. H. S. V. of 1.267 at 10,000 bbl./day of naphtha feed. From the process information given, this L. H. S. V. will give a heat of reaction of 297 B.t.u./lb. of naphtha feed. The yield under these conditions is 8,180 bbl./day of C₅+ gasoline at an octane rating of 100.85.

This reactor system will have a payout period of 0.776 year, or somewhat more than 9 months; that is, the catalytic reformer profits will pay for the cost of investment in this time.

The flow sheet (Figure 1) gives the recommended reactor system, together with the principal temperatures and pressures. A sketch of the recommended reactor itself is given in Figure 2. Details of the method and approach to the design of the reactor system follow.

DISCUSSION

The catalytic reforming operation under consideration uses a nonregenerative platinum catalyst in fixed-bed reactors. The design of a reactor system for this catalytic reforming operation is primarily an economic problem, the aim being to find that reactor system which will have the lowest payout period. It is interesting to note, however, that in this design the optimum reactor system economically is also close to an optimum in quality, the octane rating of the C₅+ gasoline product being over 100.

There are eight variable economic items in this design, all of which can be classified in three categories: income, investment, and operating costs. These three can be converted to one item called the "payout period" (total investment divided by income minus operating costs), the time required for the catalytic reformer to pay for the cost of investment. I propose to discuss the method used in this design by considering separately each of the variable economic items.

An examination of the data showed that three items - catalyst costs per year and fuel costs per year, which are both operating costs, and