

SECTION A

TRANSACTIONS OF

AMERICAN INSTITUTE

OF

CHEMICAL ENGINEERS

Issued Bi-monthly



FEBRUARY 25, 1941

Section A:

Technical Papers & Discussions

Pages 1-207

PUBLICATION ADDRESS

American Institute of Chemical Engineers, 15 North 7th Street
Philadelphia, Pa.

EXECUTIVE OFFICE

American Institute of Chemical Engineers
The Chemists Bldg., 50 East 41st Street
New York, N. Y.

DISTILLATION EQUIPMENT IN THE OIL REFINING INDUSTRY

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Presented at the New Orleans, La., Meeting, December 2, 3, 4, 1948.

ABSTRACT

A broad field of application for distillation equipment is found in the oil refining industry. In a typical complete modern refinery the investment for distillation facilities, including distillation sections of other processing equipment, represents approximately 64% of the total investment for processing units. During 1939 the investment made for new processing equipment in oil refineries in the U. S. was approximately \$110,000,000, of which about 45% was for distillation units and distillation sections of other units.

Essentially, there are twelve distinct applications of distillation facilities in oil refineries which fall in two classifications:

- A. Those involving the separation of a small number of pure components in which case theoretical methods are used for determining the operating temperatures and pressures and the fractionation.
- B. Those involving the separation of wide boiling range products in which case empirical methods are used for the calculations on the fractionating system.

A comparison is made between test results and calculated results on a typical atmospheric crude distillation unit which shows very close agreement between the two.

IMPORTANCE OF DISTILLATION EQUIPMENT

In recent years the importance of distillation equipment in the oil refining industry has been overshadowed by the intense interest in the development of new chemical and catalytic processes which have found application in all phases of the refining field. All available facilities have been concentrated on these new developments probably to the detriment of advancement in knowledge in the fundamental process of distillation. Unfortunately, when the art of continuous distillation using tubular fired heaters was in its infancy, the thermal cracking process was developed and from that time up to the time of the development of the modern processes referred to above the refiner's interest was concerned mainly with the cracking process. These circumstances have undoubtedly resulted in a retarded advancement in distillation equipment design and operation.

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This is a regrettable fact since distillation is actually the backbone of the refining industry. Especially is this true when it is considered that practically every process requires distillation equipment for the separation of its products and many require distillation for the special preparation of their feed stock. As an example, the major portion of a thermal cracking unit consists of the fractionating equipment for the separation of the products of cracking and the cycle stock. Even in this process the efforts were concentrated on the cracking reaction and equipment and only recently has the importance of the distillation section been fully recognized and given detailed study.

The cost of a typical complete refinery designed to process 20,000 barrels per day of light crude for the production of motor gasoline, refined oil, lubricating oil, asphalt, and fuel oil, and costing approximately \$11,000,000, would break down into the following approximate percentages:

	Percent
Distillation Units	15.5
Distillation sections of other processing units	15.0
Other processing units (excluding distillation sections)	17.0
Tankage	15.5
Utilities	19.0
Transfer and loading pumps and lines	7.5
Others, incl. shops, bldgs., docks, etc.	10.5
Total	100.0

It will be noted that, considering the processing equipment only, approximately 64 per cent of the cost is for distillation facilities.

A rough review of the literature has indicated that during the year of 1939 the total cost of new processing equipment constructed in oil refineries in the United States amounted to approximately \$110,000,000, distributed in about the following percentages:

	Percent
Distillation Units	13
Distillation sections of other processing units	32
Other processing units (excluding distillation sections)	55
Total	100

Here, again, the large percentage of money outlay for distillation units is evident. This is somewhat less impressive than the figures for a complete refinery probably because a large number of U. S. refineries have already modernized their distillation units. In addition,

the figure includes thirteen catalytic cracking units in which plants the major cost is for the cracking section.

APPLICATION

Essentially, distillation is used in the following refinery processes:

1. Atmospheric distillation of crude—usually for the production of reduced crude for cracking.
2. Two-stage atmospheric-vacuum distillation of crude for the production of lubricating oil distillates and asphalts.
3. Vacuum rerunning of lubricating oil distillates and cylinder stocks.
4. Atmospheric rerunning of treated naphtha, refined oil, and heating oil.
5. Stabilization of cracked naphtha and absorption naphtha.
6. Debutanization or splitting of stabilized naphtha.
7. Depropanization of straight run naphtha.
8. In gas absorption plants for the separation of the absorption naphtha from the lean oil.
9. In thermal cracking, for the separation of the products of cracking and cycle stock.
10. In catalytic cracking and similar plants, for the vaporization of the feed stock and the separation of the products of cracking.
11. In solvent treating and solvent dewaxing, for the separation of the solvent from each product.
12. In polymerization, alkylation and similar plants, for the separation of the products of reaction and cycle stock.

The above services fall into two general classifications, (A) Those involving the separation of a small number of pure components, which includes Items 5, 6, 7, 8, 11, and 12, and (B) Those involving the separation of wide boiling range products, which includes Items 1, 2, 3, 4, 9, and 10. In the equipment design entirely different principles are used for each of the two classifications.

For classification (A), the composition of the feed and products by pure components is known and the separation is treated on a theoretical basis using fugacities for the individual components and determining the compositions of the various streams and operating temperatures and pressures by Raoult's law. Similarly, the separation desired is defined as a maximum allowable percentage of contamination of each product with a pure component desired to be removed in a

different product. In order to obtain this separation, the reflux ratio and number of plates required are determined by plate to plate calculations or by the use of the McCabe-Thiele diagram. In this classification excellent separation is usually desired and this is secured by using bottoms reboiling. Usually, the operating pressure of the fractionating tower is between 100 lbs./sq.in.ga. and 250 lbs./sq.in.ga. Yard steam at a pressure of 125 lbs./sq.in.ga. is normally used as a heating medium in the reboiler, although in some instances it is necessary to use a fired reboiler.

For classification (B), the feed and products are of wide boiling range and empirical methods are used in the design of the fractionating equipment. In this classification, the atmospheric distillation of crude is the most important process because of the larger size, the greater complication, and the greater number of units for this process than any of the others. The following discussion will, therefore, deal mainly with the design and operation of crude distillation units.

CRUDE DISTILLATION UNITS

These units range in size from below 1,000 barrels per day to as high as 50,000 barrels per day, although the size commonly used varies from 15,000 to 30,000 barrels per day. They are eclipsed in size by three two-stage atmospheric-vacuum units constructed in Persia, the first two of which were designed for 65,000 B/D but actually processed as high as 85,000 B/D, and the third of which was designed to operate at above 100,000 B/D. Normally, the atmospheric distillation units are designed to cut the following products:

- Overhead —Light straight run naphtha.
About 250° F. final boiling point.
- 1st Sidestream —Heavy naphtha for reforming.
About 400° F. final boiling point.
- 2nd Sidestream—Heating oil distillate.
About 620° F. final boiling point.
- 3rd Sidestream—Gas oil for cracking.
About 700° F. final boiling point.
- Bottoms —Reduced crude for cracking.

In many cases such a unit would be made integral with the cracking equipment. Such units are called combination units. The heavy naphtha, gas oil, and reduced crude would be pumped directly from

the distillation section to the cracking section without being cooled and the cold crude would be preheated by exchange with the hot vapor and products from the cracking section. This results in a compact unit of low initial and operating cost.

To visualize the size of a 30,000 barrel per day crude pipe still designed for four sidestreams and 30% bottoms with good fractionation and high heat recovery, the following tabulation might be helpful.

Plot area (all exchangers on ground, single tier) 130' × 330'
or equivalent

Fractionating Tower

5'-6" Dia. × 12', 4 plates + 16' Dia. × 54', 19 plates + 12'
Dia. × 43', 15 plates

Stripping Towers

3-4' Dia. × 13', 4 plates and
1-5' Dia. × 14', 4 plates

Exchangers, Condensers, and Coolers

34 shells, 36,000 sq. ft. commercial shell and tube,
6,000 sq. ft. worm surface

Pumps

(Including spares) 23
Total capacity 12,000 G.P.M.

Furnace

40' × 55' × 14' high
100,000,000 B.t.u./Hr. heat fired

Such a unit, if designed for a sour crude requiring a considerable amount of alloy materials, would cost approximately \$1,350,000. The cost of this unit could be greatly reduced by making certain sacrifices in the performance. The magnitude of such reductions would be approximately as follows:

1. Elimination of alloy materials and reduction of corrosion allowance \$140,000
2. Elimination of spare tube bundles, double valves, and other equipment not absolutely required..... 90,000.

Thus, if the simplest possible unit still meeting the process requirements were constructed, the cost could be reduced to approximately \$1,120,000. Naturally, the cost could be still further reduced by decreasing the preheat, fractionation, number of sidestreams, and per cent of feed taken as overhead and sidestream cuts.

DESIGN OF CRUDE DISTILLATION UNITS

As mentioned above, the design of equipment in Classification B is empirical but it is based upon correlations determined from laboratory experimentation and plant tests. The correlations used in the design of crude distillation fractionating towers are presented herein. To demonstrate the application of these correlations to the design of a unit, the results of a five-day test on a typical distillation unit will be compared with calculated conditions.

The equipment tested was an atmospheric crude pipe still designed to process 16,000 barrels per day of a 25° A.P.I. crude taking

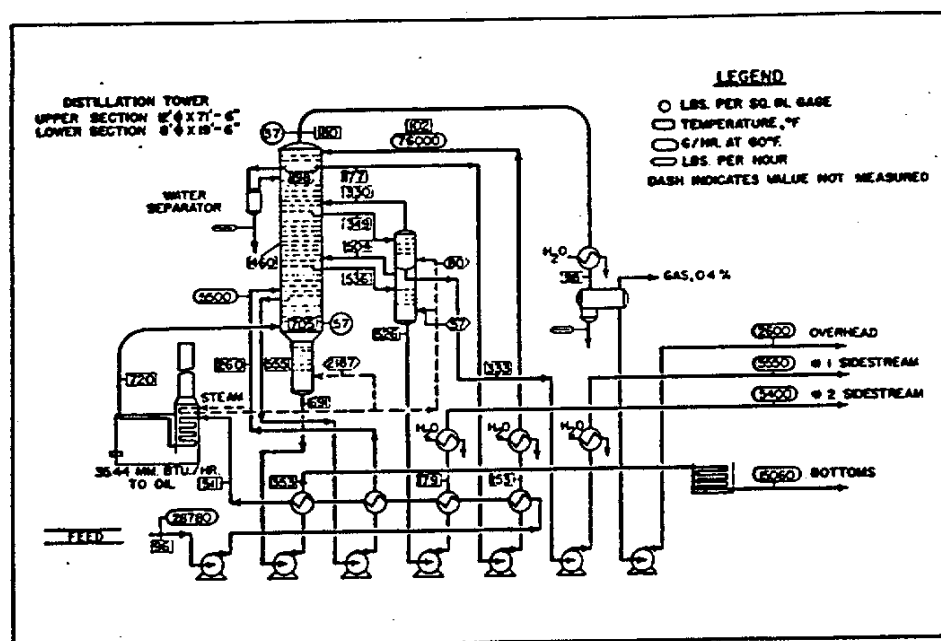


FIG. 1. Flow Plan.

naphtha overhead, water white and gas oil streams and reduced crude bottoms. Figure 1 is a simplified flow plan of the unit showing important temperatures, pressures and quantities averaged over the test period. Figure 2 shows the distillation curves of composite samples of the feed and products for the same period. Table I presents additional physical properties of the various streams.

Let it now be assumed that it is desired to design the fractionating tower of a pipe still to process the same amount of crude of the same quality as in the test and taking the same yields of products. To reduce the amount of detail in the calculations, let it further be assumed that the pressure throughout the tower will be the same as in the

test and that the flash zone temperature will be 705° F. and the bottoms steam quantity 2187 lbs. per hour, both as obtained in the test. The calculations then will involve:

1. The determination of the percentage of the feed stock vaporized at the flash zone and by bottoms steam.

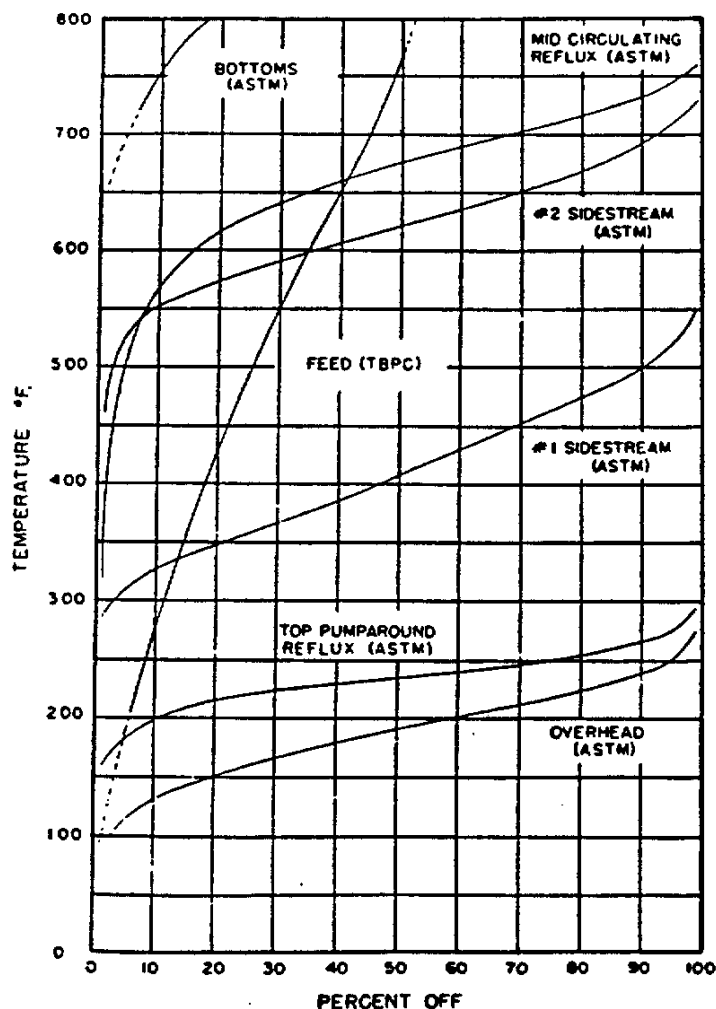


FIG. 2. Distillation Curves of Feed and Products.

2. The determination of the temperatures throughout the tower.
3. A heat balance around the tower.
4. A calculation of the maximum vapor velocity in the tower.
5. A calculation of the fractionation obtainable with the number of plates available in the tower on test.

TABLE I.—QUANTITIES AND PHYSICAL CHARACTERISTICS OF FEED AND PRODUCTS

	Vol. %	G/H	°API	#/H	M.W.	Mols/H.	Boiling Points, °F.		
							5%	50%	95% Gap °F.
Feed	100	28,780	24.1	218,000	280	780	—	—	—
Loss*	0.1	29	—	—	—	—	—	—	—
H ₂ O	0.1	29	—	242	18	13.4	—	—	—
Gas	0.4	116	—	500	55	9.1	—	—	—
Overhead	9.0	2,600	70.0	15,197	96	158	—	190	250
S. S. #1	19.3	5,550	41.9	37,710	159	237	310	405	520
S. S. #2	18.8	5,400	28.4	39,790	247	161	530	620	—
Bottoms	52.3	15,100	12.3	124,000	600	207	—	—	—
Top Pumparound Reflux		76,000	59.8	468,000	105	—	—	235	—
Mid Circulating Reflux		5,500	26.8	40,900	—	—	—	675	—

* Assumed.

PERCENTAGE OF FEED STOCK VAPORIZED AT THE FLASH ZONE

To calculate the percentage of the feed stock vaporized at the point of feed introduction recourse is made to a correlation for predicting the flash curve of a stock, the atmospheric true boiling point or A.S.T.M. distillation of which is known. A flash curve is a plot of per cent vaporized versus temperature and shows for a given pressure the percentage of a given stock that will vaporize at any temperature when the vapors are in equilibrium with the liquid residue. An empirical method of predicting the flash curve from

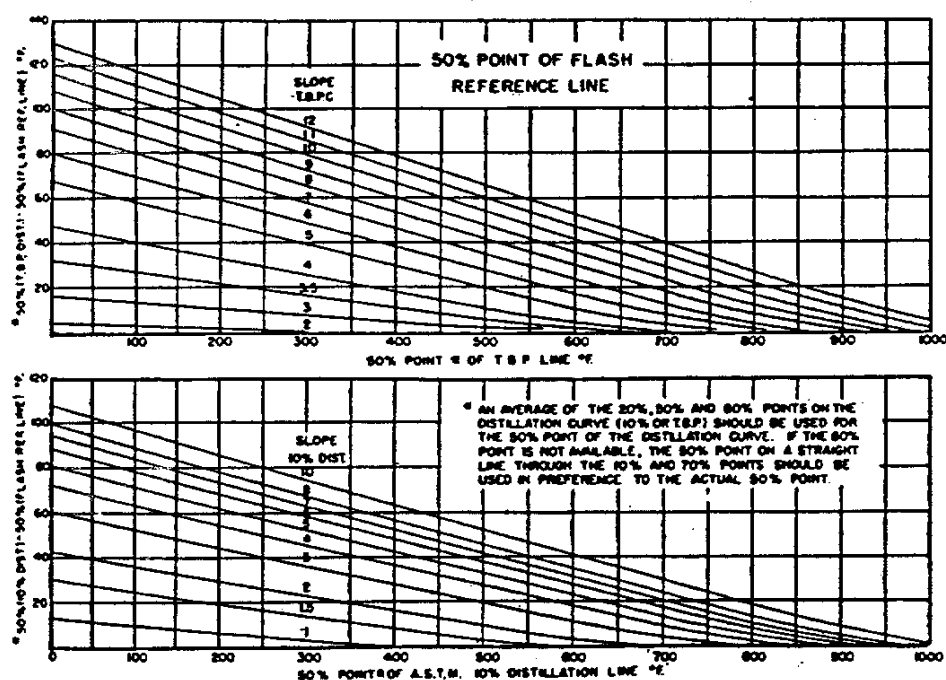


FIG. 3. Flash Vaporization Chart—50% Point.

either the true boiling point or A.S.T.M. distillation was published by Piroomov and Beiswenger in 1929.¹ This has been modified slightly from time to time and probably the most recent and complete correlation is expressed in the attached figures 3 through 5.

These charts will be used to construct the flash curve of the feed stock from its true boiling point curve which is shown in Figure 2 and duplicated in Figure 6. The first step in this procedure is to draw the distillation reference line which is a straight line through the 10% and 70% points of the distillation curve. For the case in hand the 10% point is at 273° F. and the 70% point is determined by extrapolation to be 963° F. The slope of this line is there-

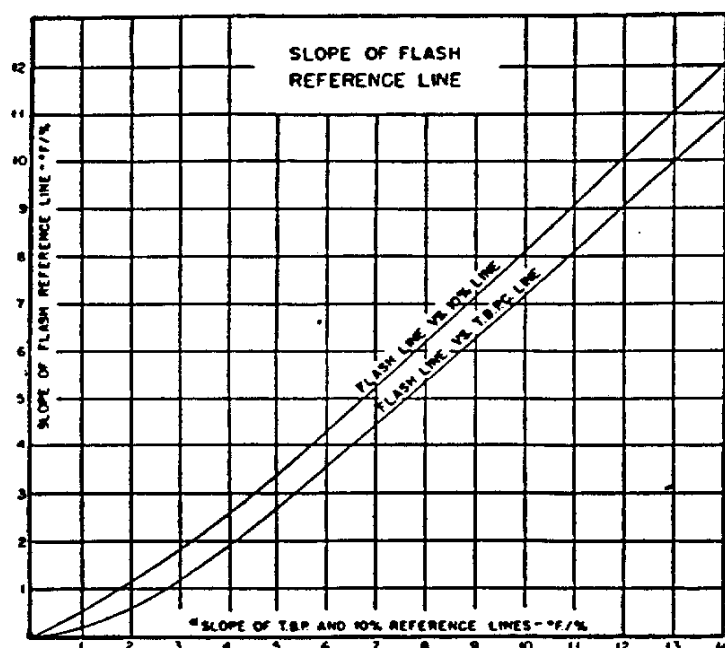


FIG. 4. Flash Vaporization Chart—Slope.

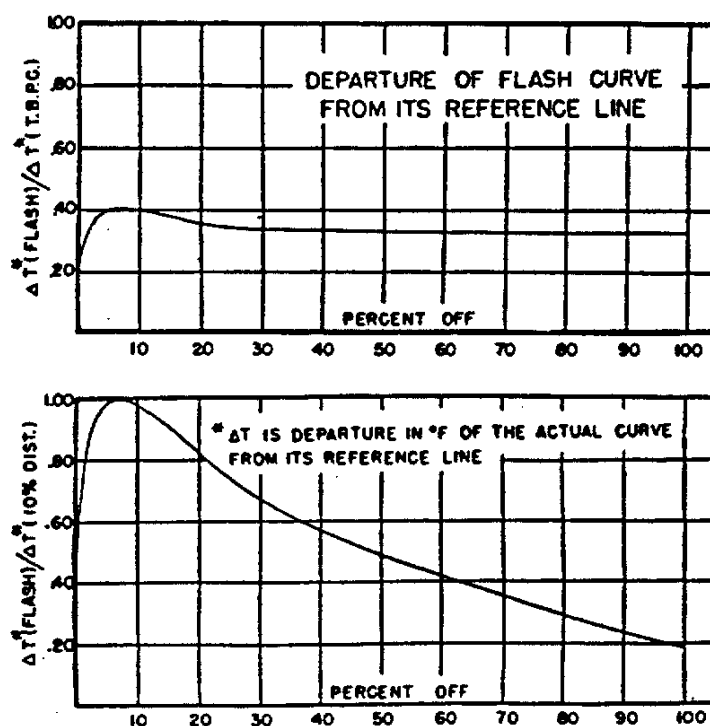


FIG. 5. Flash Vaporization Chart—Curvature.

for $\frac{963 - 273}{60} = 11.5^\circ \text{ F./}\%$. The 50% point on this line is seen to be 737° F. As noted on Figure 3, it is preferable to use the average of the 20%, 50%, and 80% points as the 50% point of the distillation curve. Here, again, by extrapolating the distillation curve as a straight line to obtain the 80% point, we have $\frac{425 + 757 + 1060}{3} = 747^\circ \text{ F.}$

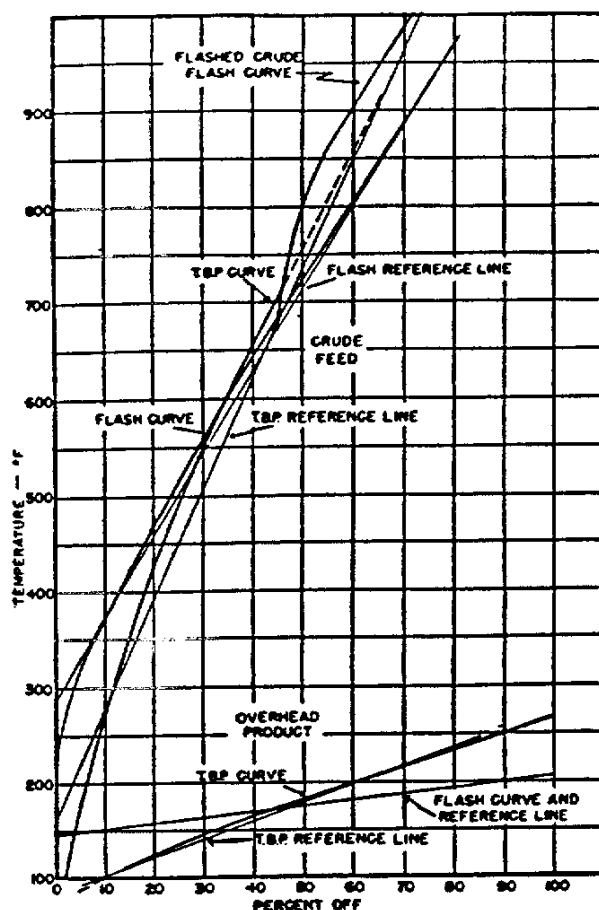


FIG. 6. Flash Curves of Crude Feed, Flashed Crude and Overhead.

Reading from the upper set of curves on Figure 3, for 747° and 11.5 slope the 50% point on the flash reference line will be 31° F. lower than 747° F. or 716° F. From Figure 4, the slope of the flash reference line, using the lower curve (since we are using the true boiling point curve as a basis) is seen to be $8.6^\circ \text{ F./}\%$. These two conditions permit us to construct the flash reference line which is a straight line through the 10% and 70% points on the flash curve. To determine

the curvature of the flash curve the upper curve in Figure 5 is used. For instance, to find the 3% point: it is seen from Figure 6 that the 3% point on the distillation curve is 130° F. and on the distillation reference line is 195° F., giving a difference of 65° F. Reading at the 3% point on Figure 5, the difference for the flash curve will be 0.37 of this amount, or 24° F. The 3% point on the flash reference line is 314° F. and therefore the 3% point on the flash curve will be $314 - 24 = 290^{\circ}$ F. In a similar manner sufficient points can be determined to define the curve.

The curve thus drawn is for a pressure of one atmosphere. Since in practically all cases the pressure will be other than one atmosphere, some method of reconstructing the flash curve for other pressures must be used. Brown and Katz² presented a method of doing this in 1935. The method is quite laborious, however, and it has been found that for the pressures normally encountered in distillation equipment the older method of displacing the flash curve parallel to itself at a higher or lower temperature (depending on whether the pressure is higher or lower than atmospheric) as determined from a Cox vapor pressure chart is sufficiently accurate. The temperature at the point of intersection of the flash and true boiling point reference lines should be used for this extrapolation.

Under the conditions of the test the pressure at the flash zone was 5.7 lbs./sq.in.g.a., or 20.4 absolute. Allowing for the water in the feed say the partial pressure of oil was 20 lbs./sq.in. The point of intersection of the flash and distillation reference lines is 650° F. (Fig. 6.) From a Cox chart the 20 lbs. flash curve will be 27° F. above the atmospheric flash curve. Therefore, at the flash zone temperature of 705° F. and 20 lbs., or $705 - 27 = 678^{\circ}$ F. and atmospheric pressure, the per cent vaporized is seen to be 44.5% from the flash curve of the feed on Figure 6. This will leave 55.5% of liquid flowing to the stripping section of the tower.

BOTTOMS STRIPPING

To determine the amount of this liquid that will be vaporized by means of the stripping steam in the bottom of the tower, it will be necessary to construct the flash curve of this liquid (called the initial bottoms). Obviously, in the design of a distillation tower, the distillation curve of the initial bottoms is not known. It could be assumed that the distillation curve of such a stock is the same as that portion of the distillation curve of the whole crude, which is known, corre-

sponding to the yield of the initial bottoms. This would evidently be in error since the liquid is in equilibrium with the vapors and, accordingly, must contain appreciable quantities of these lighter hydrocarbons.

Perhaps the most direct method of predicting the atmospheric flash curve of a reduced crude (or any reduced stock) which at the same time is reasonably accurate is the following:

- (1) Construct an atmospheric flash curve for the original crude.
- (2) Determine the number of mols of both original crude and reduced crude per given volume of original crude.
- (3) If all but one drop of original crude were flashed at one atmosphere, let it be assumed that this last drop would be in equilibrium with the vapors corresponding to the yield of reduced crude at a partial pressure of 1 atm. $\times$ mols reduced crude/mols original crude.
- (4) Extrapolate the 50% point of that portion of the flash curve corresponding to the yield of reduced crude from the partial pressure computed by (3) to 1 atm.
- (5) If the reduced crude has been stripped of light ends, its atmospheric flash curve is drawn through the extrapolated point parallel to the flash curve of the original crude between the abscissas corresponding to the yield of reduced crude.
- (6) If the reduced crude has not been stripped of light ends, a smooth curve is drawn from the split point on the flash curve of the original crude to the 20% point on the flash curve constructed by (5) to approximate the front end of the flash curve of the reduced crude. Establishment of the 20% point as the point above which unstripped light ends cease to affect the reduced crude flash curve is, of course, entirely arbitrary but, at the same time, fairly representative.

While the method outlined above is empirical to a large extent, it does have some theoretical justification. If all but one drop of reduced crude were flashed, this last drop of liquid would be in equilibrium with the reduced crude vapors at one atmosphere. It is then assumed that if 100% original crude were flashed at one atmosphere, the last drop of liquid would have the same composition as the last drop of reduced crude. The basis for this assumption is that the temperature difference between the boiling range of the last drop and that of the vapors removed in reducing the crude is

usually so great that these vapors could be considered the equivalent of steam or gas insofar as the equilibrium relations of the last drop is concerned. Therefore, the last drop of liquid would be in equilibrium with the reduced crude vapors at 1 atm. if 100% reduced crude were flashed and at a partial pressure of 1 atm. $\times$ mols reduced crude/mols original crude if 100% original crude were flashed.

Then, to locate the flash curve of the reduced crude, the 50%* point on the portion of the flash curve of the original crude corresponding to the yield of reduced crude is extrapolated on a vapor pressure chart from a partial pressure of 1 atm. $\times$ mols reduced crude/mols original crude up to 1 atm. If the reduced crude has been stripped of light ends, the flash curve of the reduced crude is drawn through this point parallel to that portion of the flash curve of the original crude between 0% to 100% on the yield of reduced crude. If there has been no stripping, the front end (0% to 20%) must be approximated by a smooth curve, since the initial flash point of the reduced crude must coincide with the split point on the flash curve of the original crude.

Making the flash curves of reduced crudes parallel to the flash curves of their original crudes was originally suggested by Piroomov and Beiswenger and appears to be fully justified by their data. However, instead of extrapolating the 50% point on a vapor pressure chart, they recommended a more complicated method involving the approximate Clapeyron equation.

Applying this method to the 55.5% initial bottoms for the case under consideration, we find from Table I that the total mols of feed is 780 mols/hr. The 55.5% initial bottoms is made up of 207 mols/hr. of final bottoms (52.3%) and 3.2% of an intermediate product estimated to be 24° A.P.I. and 300 mol. wt., giving $\frac{28,780 \times .032 \times 7.6}{300}$

23 mols/hr. The 44.5 – 100% section of the atmospheric flash curve of the crude feed will, therefore, be the flash curve of the initial bottoms, when corrected for front end curvature, at a pressure of $14.7 \times \frac{207 + 23}{780} = 4.35$ lbs. abs. The 50% point on this section of

* To be consistent with the procedure outlined for shifting the flash curve with variation in pressure, the intersection of the flash and T.B.P. reference lines should be used instead of the 50% point. However, this is more complicated, and the difference between the two methods is within the accuracy of the extrapolation, since the intersection is very close to the 50% point for stocks having comparatively high boiling ranges.

the flash curve is $44.5\% + \frac{100 - 44.5}{2} = 82.3\%$ on crude, which, from Figure 6, is seen to be 990°F . From a Cox chart we find that 990°F . and 4.35 lbs. is equivalent to 1090°F . and 14.7 lbs. Therefore, in Figure 6 the atmospheric flash curve of the initial bottoms is drawn through the point 1090°F . and 82.3% and is made parallel to the flash curve of the crude feed. A smooth curve is drawn from the point 44.5% and 680°F . which lies on the flash curve of the crude feed and the 20% point on the atmospheric flash curve of the initial bottoms just constructed. This corrects for front end curvature due to the presence of light hydrocarbons.

In calculating the percentage stripped from this liquid by the stripping steam, it is assumed that the initial bottoms is flashed in the presence of stripping steam at the pressure existing on the top stripping plate and at the exit temperature of liquid from this plate. Actually, the lower stripping plates also vaporize some of the liquid but the total vapors must pass through the top plate in the presence of the initial bottoms. This results in a somewhat greater concentration of these intermediate fractions on the top plate which changes the flash curve of the total liquid and vapor in this zone. However, sufficient data are not available to evaluate this change, although in cases where a large percentage of liquid is stripped out the indications are that some correction factor must be applied to the calculated quantity of steam.

From Figure 1, the test shows 5.7 lbs. ga. above the top stripping plate and 691°F . as the temperature of the tower bottoms. Let us assume, therefore, that the temperature leaving the top stripping plate is 695°F . Try 5% on the crude vaporized by bottoms steam. This would be

$$\frac{28.780 \times .05 \times 7.6}{300} = 36.5 \text{ mols/hr.}$$

The bottoms steam was 2,187 lbs./hr., or 122 mols/hr., and the partial pressure of oil vapors would be

$$(14.7 + 5.7) \times \frac{36.5}{122 + 36.5} = 4.7 \text{ lbs.}$$

The correction to the flash curve for this pressure is 90°F . Therefore, checking the per cent off on the atmospheric flash curve of the flashed crude at $695 + 90$, or 785°F ., it is seen to be 49%. $49 - 44.5$ gives 4.5% flashed by bottoms steam which closely checks the original

assumption. This would give 51% bottoms whereas the actual bottoms is 52.3%. Therefore, 1.3% on the crude must be overflowing from the plate above the flash zone. This will be checked by a heat balance.

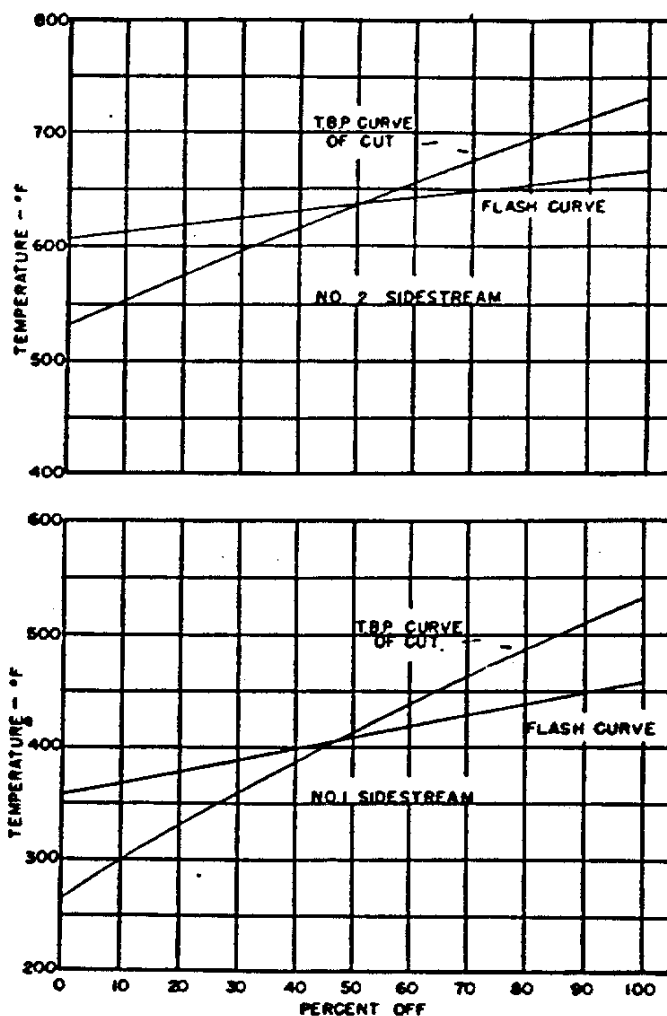


FIG. 7. Flash Curves of Nos. 1 and 2 Sidestreams.

NO. 2 SIDESTREAM TEMPERATURE

To calculate the temperature of the No. 2 sidestream drawoff, reference is made to the flash curve of this product shown in Figure 7. In this case it is necessary to use that section of the true boiling point curve of the feed corresponding to the cut under consideration to determine the flash curve of the product. This product must be completely vaporized below the sidestream drawoff plate. Accordingly, the 100% point on the flash curve is used. To determine the

partial pressure of the product plus reflux vapors, both of which are of the same composition, the lighter vapors are considered to be inert.

The total pressure at this point is approximately 5.0 lbs. ga. and the partial pressure is $(14.7 + 5.0) \times \frac{161}{700} = 4.5$ lbs. abs., assuming negligible reflux below drawoff plate and where 700 = total mols of vapors passing this point, including steam.

The temperature correction for this pressure is 90° F., and the endpoint on the atmospheric flash curve is 665° F., giving a vapor temperature of $665 - 90 = 575^\circ$ F.

To check if there is any reflux below the No. 2 sidestream drawoff plate, the heat removed to this point is:

$$\begin{array}{rcl}
 \text{Products} & -(218,000 - 124,000) & (704 - 575) \times .64 \\
 \text{Overflash} & -28,780 \times .013 \times 7.6 \times [(695 - 665) \times .77 + 86] & \\
 \text{Steam} & -(2187 + 242) (695 - 575) \times .47 & \\
 \hline
 & & 8,203,000
 \end{array} = 7,750,000 \text{ B.t.u./Hr.}$$

where 704° is the average temperature of the product vapors made up of 44.5% at 705° and 3.2% at 695° F.,
 695° is the temperature of vapors from the bottom of the tower, and
 665° is the temperature of liquid overflow from the plate above the flash zone.

less heat removed by mid-circulating reflux (see Fig. 1 for temperatures)

$$\begin{array}{rcl}
 5500 \times 7.44 \times (555 - 260) \times .64 & & = 7,710,000 \\
 \text{plus radiation } 25 \times 12 \times \pi \times (640 - 50) & & \\
 \times .2 & & = 110,000 \\
 \hline
 & & 7,820,000
 \end{array}$$

where 640° is the average tower temperature in this section, and
 50° is the average air temperature.

Heat removed by reflux below No. 2 S.S. = 383,000 B.t.u./Hr.

Quantity of reflux =

$$\begin{array}{rcl}
 383,000 \div [(575 - 536) \times .74 + 96] & = & 3,050 \text{ lbs./hr.} \\
 \text{at 220 M. W.} & = & 14 \text{ mols/hr.}
 \end{array}$$

where 536° is the estimated temperature of reflux from the #2 S.S. drawoff plate.

which checks the assumption of negligible reflux below this plate.

The vapor temperature drop per plate between this section and No. 1 sidestream drawoff is approximately 11° F. per plate. This would give a vapor temperature above the drawoff plate of $575 - 11 = 564^{\circ}$ F.

The heat removed by reflux on the drawoff plate is:

No. 2 sidestream condensed and cooled to
stripper outlet temperature (taken
from test for brevity)

$$39790 \times [(575 - 526) \times .73 + 98] = 5,320,000$$

Lighter Products

$$(37,710 + 15,197 + 500)(575 - 564) \times .62 = 365,000$$

Steam

$$(2,187 + 242)(575 - 564) \times .47 = 13,000$$

Reflux below plate

$$383,000$$

Radiation for this plate

$$\text{Negligible}$$

$$6,081,000 \text{ B.t.u./hr.}$$

and the reflux quantity is

$$6,081,000 \div [(564 - 516) \times .74 + 98] = 45,500 \text{ lbs./hr.}$$

at 210 M.W.

$$= 217 \text{ mols/hr.}$$

where 516° F. is the estimated temperature of reflux entering the plate.

Now to determine the temperature of liquid product being withdrawn it is necessary to determine the initial point on the flash curve of the product at the existing partial pressure of hydrocarbon vapors. Since the liquid is in equilibrium with the total vapors passing through the plate, it will contain considerable low boiling hydrocarbons. Some of these are stripped out in the sidestream stripping column and some remain in the final product. Test data have been correlated to permit using the initial point on the flash curve of the cut on the true boiling point curve of the feed and applying a correction factor to the temperature thus determined to allow for the above factors and arrive at the true drawoff temperature. This correlation is presented in Figure 8.

The partial pressure of the oil vapors considering steam and gas as inert is

$$(14.7 + 5) \left(\frac{217 + 158 + 237}{217 + 158 + 237 + 144} \right) = 15.9\# \text{ abs.}$$

where 144 is the total mols of steam and gas present. The temperature correction for this pressure is 8° F., and the initial point on the

atmospheric flash curve of the cut is 607° F., giving 615° F. on the 15.9 lb. flash curve. From Figure 7 the initial point on the cut on the true boiling point curve is 530° F. and from Figure 8 the correction for 530° F. initial point on the cut on the true boiling point curve is 74° F. The temperature of the liquid leaving the plate will, therefore, be

$$615 - 74 = 541^{\circ} \text{ F.}$$

which checks the test temperature of 536° F.

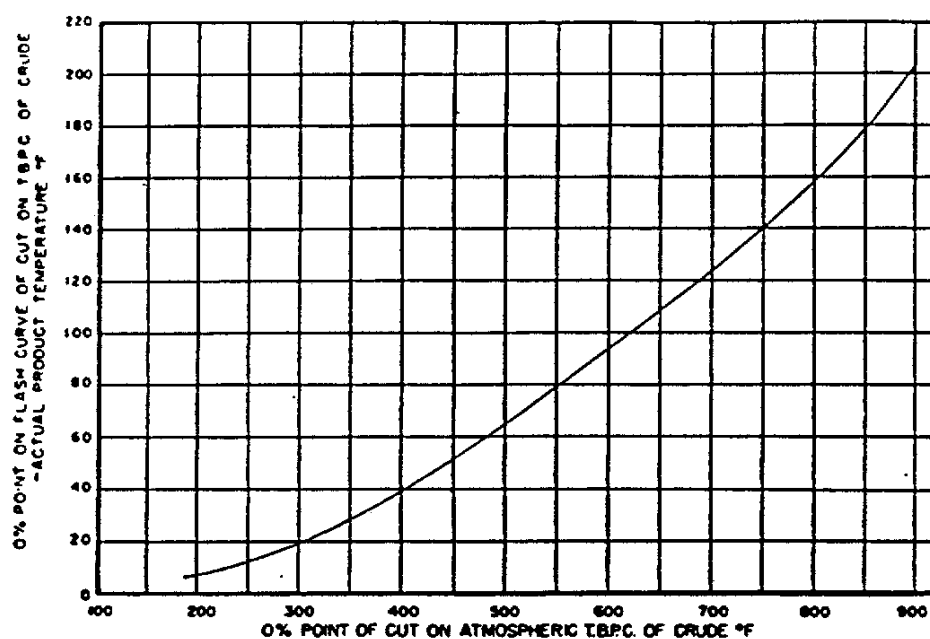


FIG. 8. Temperature Correction for Calculating Sidestream Drawoff Temperature—Crude Distillation Towers.

NO. 1 SIDESTREAM TEMPERATURE

To check the temperature of the No. 1 sidestream drawoff, the flash curve of this cut is shown in Figure 7. The partial pressure of the product plus reflux vapor is

$$(14.7 + 4.5) \times \frac{237 + 360}{902} = 12.7 \text{ lbs. abs.}$$

where 360 = assumed mols of reflux below drawoff plate
and 902 = total mols of vapors passing this point including reflux and steam.

The temperature correction for this partial pressure is 10° F. and the endpoint on the atmospheric flash curve is 457° F., giving a vapor temperature of 457 - 10, or 447° F.

To check the assumed quantity of reflux, the heat removed to this point is:

Reflux below No. 2 S.S. drawoff = 383,000 B.t.u./Hr.
Heat from No. 2 S.S.

$$39,790 \times [(575 - 526) \times .73 + 98] = 5,320,000$$

Light Products

$$(37,710 + 15,197 + 500) (575 - 447) \times .58 = 3,970,000$$

Steam

$$(2,187 + 242 + 57) (575 - 447) \times .47 = 149,000$$

9,822,000

Less radiation

$$25 \times 12 \times \pi \times (500 - 50) \times .2 = 85,000$$

9,737,000

Quantity of reflux

$$9,737,000 \div [(447 - 349) \times .68 + 108] = 56,000 \text{ lbs./hr.}$$

$$\text{at a mol wt. of 135} = 360 \text{ mols/hr.}$$

$$\text{at 6.75 lbs./gal.} = 8,300 \text{ gals./hr.}$$

where 349° is the estimated temperature of reflux from the #1 S.S. drawoff plate.

The vapor temperature drop per plate between this section and the top of the tower is approximately 33° F. This gives a vapor temperature above the drawoff plate of 447 - 33 = 414° F. The heat removed by reflux on the drawoff plate is:

No. 1 sidestream condensed and cooled to stripper outlet temperature (taken from test for brevity)

$$37,710 \times [(447 - 333) \times .675 + 106.5] = 6,920,000 \text{ B.t.u./Hr.}$$

Lighter Products

$$(15,197 + 500) (447 - 414) \times .565 = 290,000$$

Steam

$$(2,187 + 242 + 57) (447 - 414) \times .47 = 40,000$$

Reflux

9,737,000

Radiation for one plate

Negligible

16,987,000 B.t.u./Hr.

and the reflux quantity is

$$16,987,000 \div [(414 - 325) \times .68 + 112] = 98,500 \text{ lbs./hr.}$$

$$\text{at 145 M.W.} = 680 \text{ mols/hr.}$$

where 325 is the estimated temperature of reflux entering the plate.

The partial pressure of the oil vapors is therefore

$$(14.7 + 4.5) \left(\frac{158 + 680}{158 + 680 + 147} \right) = 16.4 \text{ lbs. abs.}$$

where 147 is the mols per hour of steam and gas.

The temperature correction for this pressure is 12° F. and the initial point on the atmospheric flash curve is 358° F., giving 370° F. on the 16.4 lb. flash curve.

From Figure 7, the initial point on the cut on the true boiling point curve is 265° F. and from Figure 8 the correction factor will be 14° F. The temperature of the liquid leaving the plate will be

$$370 - 14 = 356^\circ \text{ F.}$$

which checks the test temperature of 349° F.

TOP OF TOWER TEMPERATURE

The total pressure at the top of the tower is $14.7 + 3.7 = 18.4$ lbs. abs. The total steam in the tower is 2,566 lbs. per hr., or 142 mols/hr.

Therefore, the partial pressure of the oil vapors is $18.4 \times \frac{167.1}{167.1 + 142} = 9.95$ lbs. abs. The endpoint on the atmospheric flash curve of the overhead oil vapors as shown in Figure 6 is 205° F. and the temperature correction for a partial pressure of 9.95 lbs. abs. is 22° F., giving a calculated top of tower temperature of 183° F. which checks the test temperature of 180° F. The endpoint on the flash curve is used since the product must exist as a vapor at the top of the tower. The partial pressure of water vapors is $18.4 - 9.95 = 8.45$ lbs. abs., and the vapor pressure of water at 183° F. is 8.02 lbs. abs., so that a small quantity of steam will be condensed in the tower.

The total heat removed by the top pumparound reflux as determined from the test (see Figure 1) is:

468,000 × (177 - 102) × .542	= 19,000,000 B.t.u./Hr.
by heat balance	
Reflux below No. 1 S.S. drawoff	= 9,737,000 B.t.u./Hr.
Heat from No. 1 S.S.	
37,710 × [(447 - 333) × .675 + 106.5]	= 6,920,000
Overhead and Gas	
(15,197 + 500) (447 - 180) × .505	= 2,120,000
Steam	
2,566 × (447 - 180) × .47	= 320,000
	19,097,000
Less radiation	
21.5 × 12 × π × (310 - 50) × .2	= 42,000
	19,055,000

which checks the test heat removal and thereby gives a check on the amount of products vaporized at the flash zone and in the bottom of the tower. This shows that the calculated 1.3% of overflow from the plate above the flash zone is approximately correct.

It is of interest to note that in all cases where the lowest sidestream in a tower is fractionated, there must be *some* overflow from the plate above the flash zone since the purpose of the fractionation is to remove the entrained and vaporized high boiling materials. In this particular case the overflow is 1.3% on the feed or 2.7% on the total overhead products. The No. 2 sidestream was well fractionated as indicated by the nearly straight line distillation as well as the fact that the endpoint by A.S.T.M. distillation was 730° F. as compared with a cut point of 735° F. on the true boiling point curve of the feed. The fractionation, however, was not as good as true boiling fractionation and it is estimated from other data available that to simulate true boiling fractionation another 2.5% on the feed would have to be fractionated out. This would give a total of 3.8% on the feed or 8.4% on the reduced amount of total overhead as entrained plus vaporized high boiling materials. In the design of new facilities, therefore, it is well to provide for at least 4% overflash.

FRACTIONATION

To determine the number of plates and reflux ratio required for a given degree of fractionation or vice versa, an empirical correlation which was obtained from plant and laboratory tests is used. Figure 9 covers this correlation for fractionation between the overhead product and top sidestream of a column and Figure 10 is for fractionation between two sidestreams. The degree of fractionation is measured as the gap or overlap between the 95% point of the lighter product and the 5% point of the heavier product, both being by A.S.T.M. distillation (distillate plus loss). The factor "F" is equal to the product of the reflux ratio and the number of plates between the two products under consideration. The reflux ratio is defined as the liquid volume ratio measured at 60° F. of the hot reflux below the drawoff plate of the lighter product to the sum of the products lighter than the heavier product. This means that the reflux ratio and number of plates are interchangeable for a given fractionation. Theoretically, this holds quite accurately for the degree of fractionation normally desired and the number of plates and reflux ratios normally

used, *i.e.*, 5 to 10 and 1 to 5, respectively. The numbers on the curves refer to the separation in ° F. between the 50% point of the heavier product and the 50% point of the sum of the lighter products, both measured on the A.S.T.M. distillation curves of the individual products. The correlations are on this basis since in the majority of the

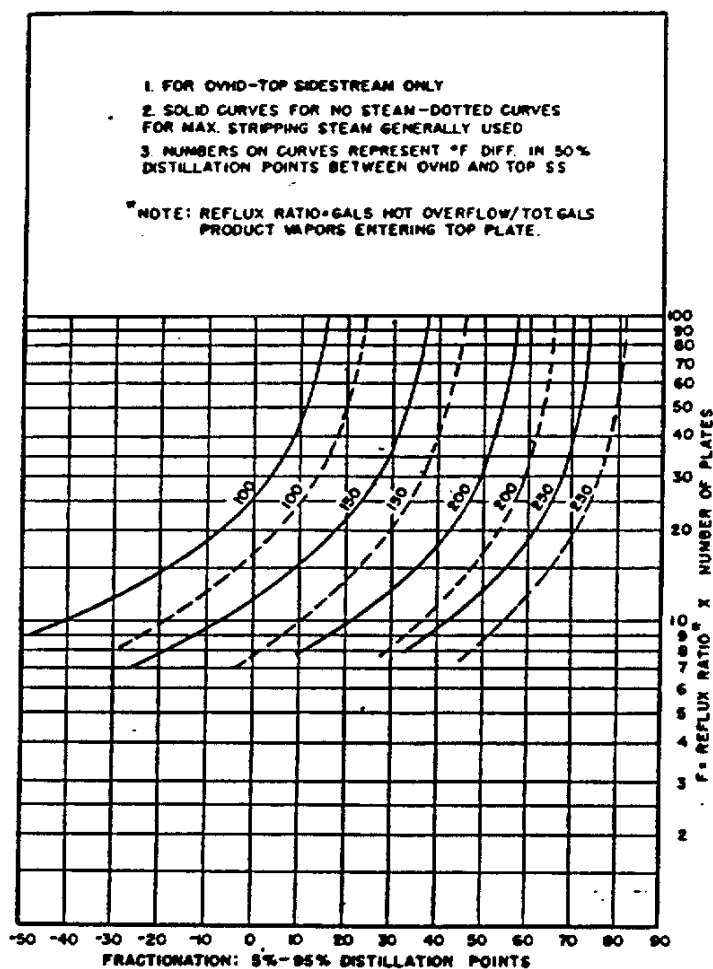


FIG. 9. Fractionation Curves for Overhead to Top Sidestream Products.

tests the true boiling point curves of the feed were not available. In the design of a unit, however, the reverse is true, *i.e.*, the true boiling point curve of the feed is known but the A.S.T.M. distillations of the products are not known and it is necessary to use the 50% points of the cuts as determined from the true boiling point curve of the feed. This gives about a 5° F. greater gap than can actually be expected.

To calculate the fractionation between the overhead and No. 1 sidestream, reference is made to Figure 1.

The number of plates between cuts, considering the two top pump-around reflux plates as equivalent to one fractionating plate, is eight.

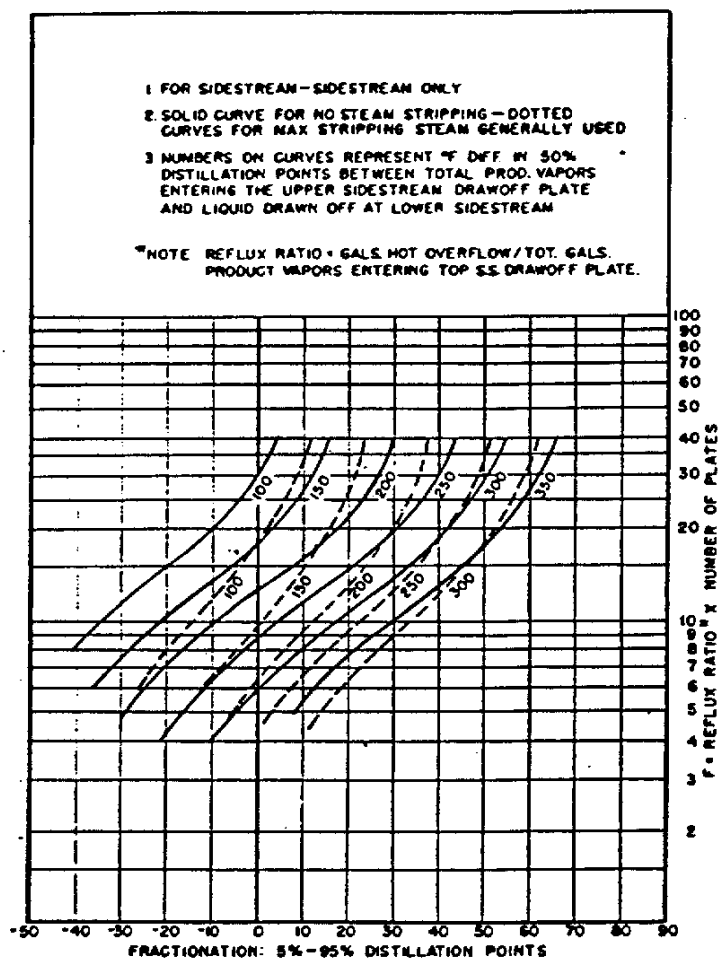


FIG. 10. Fractionation Curves for Sidestream to Sidestream Products.

The hot reflux below the lower pumparound reflux plate removes 19,055,000 B.t.u./hr. less the heat corresponding to the temperature drop through the two reflux plates of

$$\begin{array}{rcl}
 \text{Overhead} & & \\
 (15,197 + 500)(198 - 180) \times .445 & = & 126,000 \\
 \text{Steam} & & \\
 2,566 \times (198 - 180) \times .47 & = & 22,000 \\
 \hline
 & & 148,000 \text{ B.t.u./Hr.}
 \end{array}$$

or 18,907,000 B.t.u./Hr.

The hot overflow is therefore

$$\begin{aligned} 18,907,000 \div [(198 - 177) \times .575 \\ + 138] &= 126,000 \text{ lbs./hr.} \\ \text{at 105 M. W.} &= 1,200 \text{ mols/hr.} \\ \text{at 6.16 lbs./gal.} &= 20,500 \text{ gals./hr.} \end{aligned}$$

The overhead product plus gas is 2,716 gals./hr., so that the reflux ratio is $20,500 \div 2,716 = 7.6$.

$$F = 8 \times 7.6 = 61$$

The 50% point of the No. 1 sidestream cut is 415° F. and of the overhead, 185° F.

$$\text{The separation} = 415 - 185 = 230^\circ \text{ F.}$$

Reading from Figure 9 and interpolating, the gap would be

No steam stripping	65° F. gap
Complete steam stripping	
(0.2 lbs./gal.)	74° F. gap

The steam used in the test was $\frac{80}{5550} = .014$ lbs./gal. which would

give $65 + 9 \times \frac{.014}{.2} = 66^\circ \text{ F.}$ Subtracting 5° due to the use of the true boiling point curve for the 50% points of the cuts gives 61° F. gap which checks the test result of $310 - 250 = 60^\circ \text{ F.}$

To determine the fractionation between the No. 1 sidestream and No. 2 sidestream, the amount of hot reflux below the No. 1 sidestream has been calculated to be 8,300 gals./hr., and the reflux ratio is

$$8,300 \div (116 + 2,600 + 5,550) = 1.0$$

The number of plates between No. 1 and No. 2 sidestreams is 10.

$$F = 10 \times 1 = 10$$

The 50% point of the total product vapors entering the first sidestream drawoff plate is obtained from the true boiling point curve as 345° F. (14.3% point on crude feed) and the 50% point of the No. 2 sidestream cut is 635° F.

$$S = 635 - 345 = 290^\circ \text{ F.}$$

From Figure 10, the gap would be

No steam stripping	14° F. gap
Complete steam stripping	31° F. gap

The steam used in the test was $\frac{57}{5400} = .01$ lbs./gal., which would give $14 + 17 \frac{.01}{2} = 15^\circ \text{ F.}$ Subtracting 5° F. due to the use of the 50% point of the cuts on the true boiling point curve instead of the A.S.T.M. curves gives a 10° F. gap which checks the test results of $530 - 520 = 10^\circ \text{ F.}$ gap.

VAPOR VELOCITIES

The maximum vapor velocity will occur just below the top pump-around reflux circuit. At this point the quantity of vapors is:

reflux	126,000 lbs./hr.	1200 mols/hr.
overhead	15,197	158
gas	500	9
steam	2,566	142
Total	144,263	1509,
		giving 95.5
		Mol. Wt.

$$\text{Vapor volume} = \frac{1509 \times 359}{3600} \times \frac{(460 + 198)}{492} \times \frac{14.7}{18.4}$$

$$= 1610 \text{ cu. ft./sec.}$$

$$\text{Vapor velocity} = \frac{1610}{12^2 \times .785} = 1.43 \text{ ft./sec.}$$

The allowable vapor velocity is then estimated from the entrainment formula

$$v_a = 6.65 \sqrt{\frac{T \times S}{M \times P}}$$

when v_a = allowable velocity in ft./sec.

T = vapor temperature in $^\circ \text{ R.}$

S = liquid specific gravity

M = vapor molecular weight

P = operating pressure—in inches of Hg. absolute

6.65 = factor for maximum velocity with high fractionating efficiency when using 2' 6" plate spacing

$$\text{Allowable velocity } v_a = 6.65 \sqrt{\frac{(460 + 198) \times .74}{95.5 \times 18.4 \times 2.03 \times 1.09}}$$

$$= 2.35 \text{ Ft./sec.}$$

where 2.03 is a conversion factor for lbs./sq.in. to inches Hg. pressure and 1.09 is the coefficient of expansion of the liquid.

From the above it will be seen that the tower is considerably larger in diameter than is required. This fact was realized at the time the unit was constructed.

GENERAL

Methods are available for calculating the temperature of the circulating reflux streams leaving the tower as a function of the vapor temperatures above and below the reflux plates, the heat transfer rates on the plates, the plate surface, and the total heat removed.

Methods are also available for calculating the amount of steam required in steam strippers, the number of plates, the percentage of the product stripped out and the flash and distillation increase in the stripped products.

Test data correlations also permit the prediction of the yield of a product of given characteristics from a feed stock, the true boiling point curve of which is known.

Calculations on the above have not been included in this Paper due to the detail required.

SUMMARY

A summary of the test and calculated results is as follows:

	Calculated	Test
No. 2 sidestream drawoff temperature, °F.	541	536
No. 1 sidestream drawoff temperature, °F.	356	349
Top of tower temperature, °F.	183	180
Heat removal by top reflux, MM B.t.u./Hr.	19.1	19.0
Gap between 95% point of overhead and 5% point of No. 1 sidestream on A.S.T.M. distillation, °F.	61	60
Gap between 95% point of No. 1 sidestream and 5% point of No. 2 sidestream on A.S.T.M. distillation, ° F.	10	10

ACKNOWLEDGMENT

Acknowledgment is made to Mr. J. B. Maxwell, of the Standard Oil Development Company, who developed the attached curves on flash vaporization.

Acknowledgment is also made to Mr. J. S. Clarke, of the Standard Oil Development Company, who developed the attached curves on fractionation.

LITERATURE CITED

1. Piroomov and Beiswenger, *A.P.I. Bulletin* 10, No. 2, 1929.
2. Brown and Katz, *Industrial and Engineering Chemistry*, 25, 1935, pp. 1373-1385.

DISCUSSION

A. J. Hoiberg (Lion Oil Refining Co., El Dorado, Arkansas): I would like to ask the significance of the negative numbers on the last chart.

J. W. Packie: That was the overlap in degrees Fahrenheit between the 95 per cent point of the light product and the 5 per cent point of the heavy product. In other words, if the 95 per cent point of the light product were 300 degrees, and the 5 per cent point of the heavy product were 280 degrees, then the gap would be minus 20 degrees.

ENGINEERING
DESIGN SEMINAR
CRUDE UNIT DESIGN
(CLASS NOTES)

FALL 1991

CRUDE UNIT DESIGN (CLASS NOTES)

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CUSTOMER ENGINEERING DESIGN SEMINAR
SUBJECT CRUDE UNIT DESIGN

PAGE NO. I-1
DATE 9-20-91
PROJECT
BY E N MILLER

INTRODUCTION .

THIS SEGMENT OF THE ENGINEERING DESIGN SEMINAR IS DEVOTED TO THE BASIC PROCESS DESIGN OF A CRUDE UNIT. THE SCOPE OF THIS DESIGN HAS BEEN LIMITED TO THE ESSENTIAL COMPONENTS OF THE DESIGN SO THAT WE CAN COMPLETE THE DESIGN CALCULATIONS WITHIN THE TIME AVAILABLE. THIS SECTION WILL BE OF PRIMARY INTEREST TO PROCESS ENGINEERS. FOR THOSE OF YOU WHOSE PRIMARY INTERESTS LIE ELSEWHERE, WE HOPE TO PROVIDE SOME INSIGHT INTO HOW THE DATA WHICH BECOME THE BASIS FOR EQUIPMENT SPECIFICATION AND DESIGN ARE GENERATED. FOR THOSE OF YOU WHO ARE MORE ACCUSTOMED TO DEALING WITH THE DEFINED CHEMICAL AND HYDROCARBON SPECIES COMMON TO THE CHEMICAL AND PETROCHEMICAL INDUSTRIES, WE HOPE THAT YOU BECOME A LITTLE MORE APPRECIATIVE OF THE PROBLEMS FACED BY ENGINEERS DEALING WITH BOILING RANGE MIXTURES CHARACTERISTIC OF THE PETROLEUM REFINING INDUSTRY. 'BOILING OIL' IS NOT PARTICULARLY GLAMOROUS BUT NECESSARY.

THE CRUDE DISTILLATION TOWER AND ITS AUXILIARIES IS THE 'KEYSTONE' UNIT IN ALL REFINERIES. ITS PURPOSE IS TO FRACTIONATE CRUDE OIL INTO VARIOUS FRACTIONS WHICH BECOME FEED STOCKS FOR DOWNSTREAM PROCESSING UNITS AND/OR PRODUCTS. AS CAN BE SEEN IN THE ATTACHED BLOCK FLOW OF A REFINERY, THE NUMBER OF PRODUCTS AND PROCESSES ORIGINATING FROM THE CRUDE TOWER ARE MANY. THIS SLIDE WAS DEVELOPED TO SHOW THE DIVERSITY OF UOP'S TECHNOLOGY. A SIMILAR BLOCK FLOW DIAGRAM CAN BE DEVELOPED FOR A WIDE RANGE OF PETROCHEMICALS BASED ON FEEDS ORIGINATING IN THE REFINERY. THE ATTACHED BLOCK FLOW EMPHASIZES PETROCHEMICALS THAT CAN BE MADE USING UOP TECHNOLOGY.

FROM A DISTILLATION PERSPECTIVE, THE CRUDE DISTILLATION TOWER REPRESENTS A DISTILLATION CONFIGURATION CHARACTERISTIC OF HYDROCARBON SEPARATION IN A NUMBER OF REFINERY PROCESS UNITS. MOST COMMON ARE THE MAIN COLUMN IN A

INTRODUCTION.

FCC UNIT AND THE PRODUCT FRACTIONATOR IN A HYDROCRACKING UNIT. THE SAME TECHNIQUES DISCUSSED FOR THE CRUDE TOWER DESIGN CAN AND ARE BEING USED TO DESIGN OTHER "CRUDE TYPE" FRACTIONATING TOWERS.

WHEN WE TALK ABOUT "CRUDE TYPE" DISTILLATION SYSTEMS, WE GENERALLY REFER TO A DISTILLATION SYSTEM WHERE ALL OF THE HEAT REQUIRED TO VAPORIZE THE DISTILLATE PRODUCTS IS PUT INTO THE FEED AS OPPOSED TO PUTTING PART OF THE HEAT IN THE FEED AND THE REMAINDER VIA A REBOILER AT THE COLUMN BOTTOMS. MOST "CRUDE TYPE" FRACTIONATORS MAKE MULTIPLE PRODUCTS WITH ONE OR MORE SIDE PRODUCT DRAWS (GENERALLY LIQUID) BETWEEN THE OVER HEAD AND BOTTOMS PRODUCTS. THE SIDE CUT PRODUCTS ARE SENT TO STRIPPERS (EITHER STEAM STRIPPERS OR REBOILED STRIPPERS) FOR LIGHT ENDS REMOVAL WHERE THE STRIPPED VAPOR IS RETURNED TO THE TOWER. THE SIDE CUT STRIPPERS ARE AN INTEGRAL PART OF THE CRUDE TOWER. OFTEN EXCESS HEAT IS REMOVED FROM THE TOWER AT BUT NOT RESTRICTED TO INTERMEDIATE PRODUCT DRAW TRAYS VIA PUMP AROUND EXCHANGE. INTERMEDIATE HEAT REMOVAL INCREASES THERMAL EFFICIENCY OF UNIT AND HELPS BALANCE VAPOR AND LIQUID LOADS IN THE TOWER.

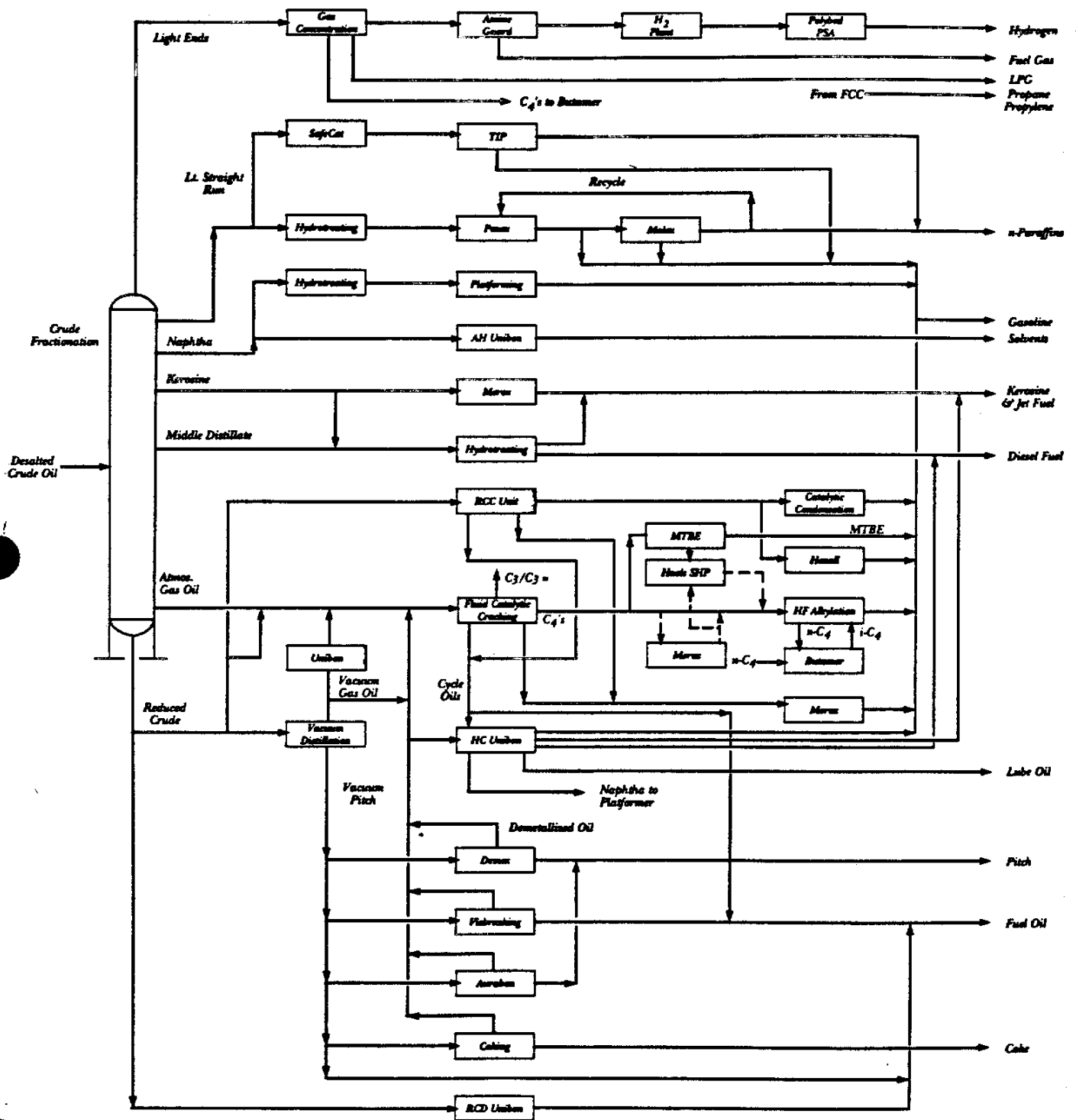
THE CALCULATIONAL TECHNIQUES PRESENTED USE THE TRADITIONAL APPROACH TO CRUDE DISTILLATION. THESE TECHNIQUES DATE BACK TO THE EARLY 1940'S. THE CLASSIC PAPER ON CRUDE COLUMN DESIGN IS THAT BY PACHIE 1943. WE HAVE MADE COPIES FOR THOSE OF YOU WHO ARE INTERESTED. ALTHOUGH IN SOME RESPECTS IT APPEARS DATED, THE RESULTS ARE STILL VALID AND USEFUL. THESE KINDS DATA, REFINED WITH OPERATING INFORMATION FROM COMMERCIAL UNITS, ARE USED TO INTERPRET RESULTS GENERATED WITH CURRENT GENERATION OF PROCESS SIMULATORS. "RIGOROUS" CALCULATIONAL ALGORITHMS WHICH ALLOW TRAY BY TRAY EQUILIBRIUM SIMULATION OF A CRUDE

INTRODUCTION

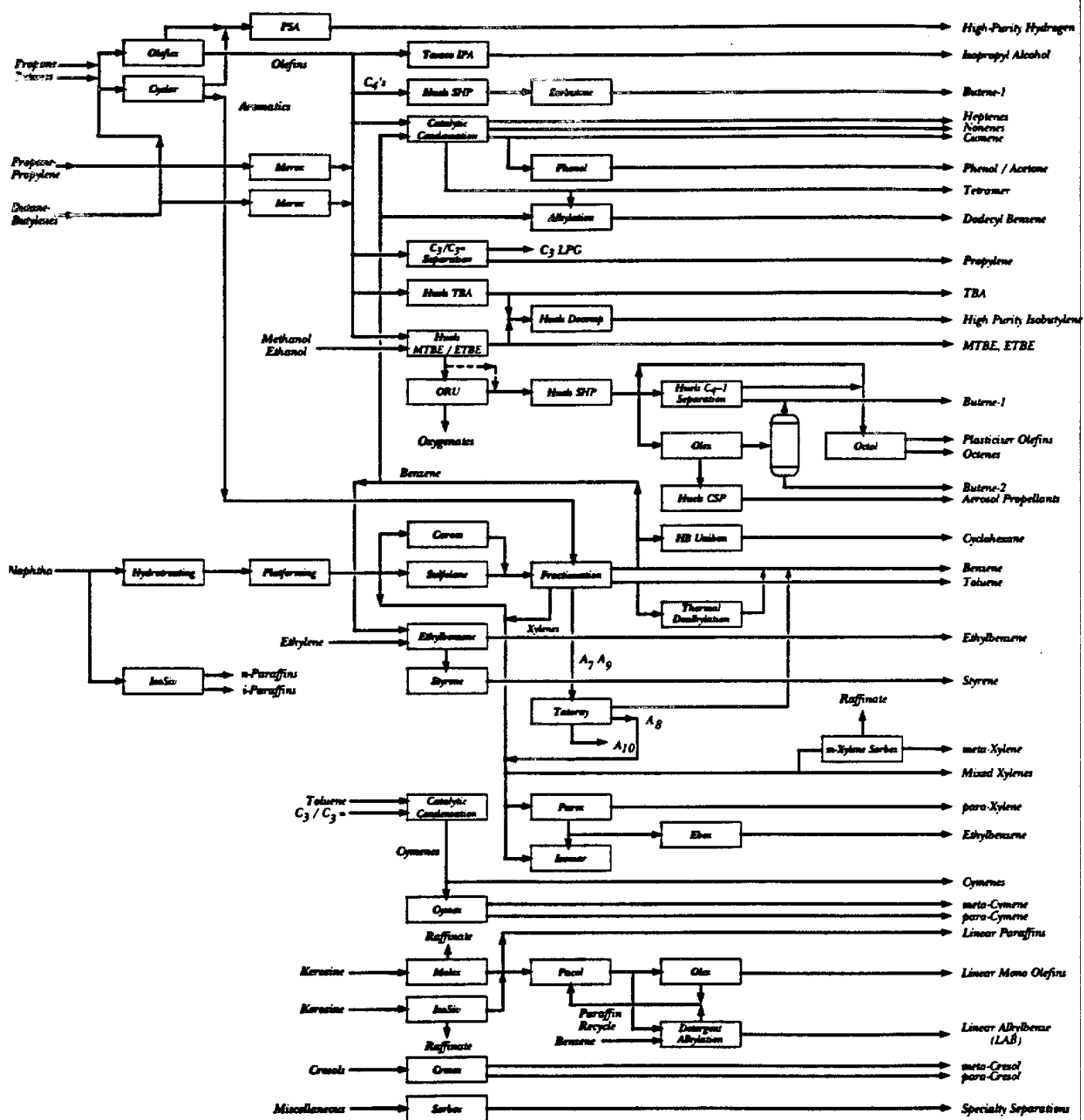
TYPE DISTILLATION COLUMN HAVE BEEN IN EXISTENCE FOR LESS THAN 10 YEARS. AT THE END OF THIS SESSION, TIME PERMITTING, WE WILL DEMONSTRATE USE OF THESE CALCULATIONAL TOOLS IN DESIGN.

THE APPROACH THAT WE WILL USE IS THAT OF INTERACTIVE STUDENT PARTICIPATION. THESE SESSIONS ARE STRUCTURED AS DESIGN WORKSHOPS EMPHASIZING INFORMATION PRESENTED IN PREVIOUS SEMINAR SESSIONS. IN ADDITION, AS A PART OF THE CRUDE UNIT DESIGN EXERCISE, INFORMATION WHICH ARE DESIGN INPUTS FOR TOPICS PRESENTED IN LATER SEGMENTS OF THIS SEMINAR IS GENERATED. WE WILL BUILD THE DESIGN IN A SERIES OF LOGICAL STEPS SIMILAR TO THOSE WHICH A PROCESS ENGINEER WOULD USE IN ACTUAL DESIGN. AT EACH STAGE WE WILL DISCUSS THE OBJECTIVES, DO THE CALCULATIONS, AND THEN REVIEW THE RESULTS.

eight

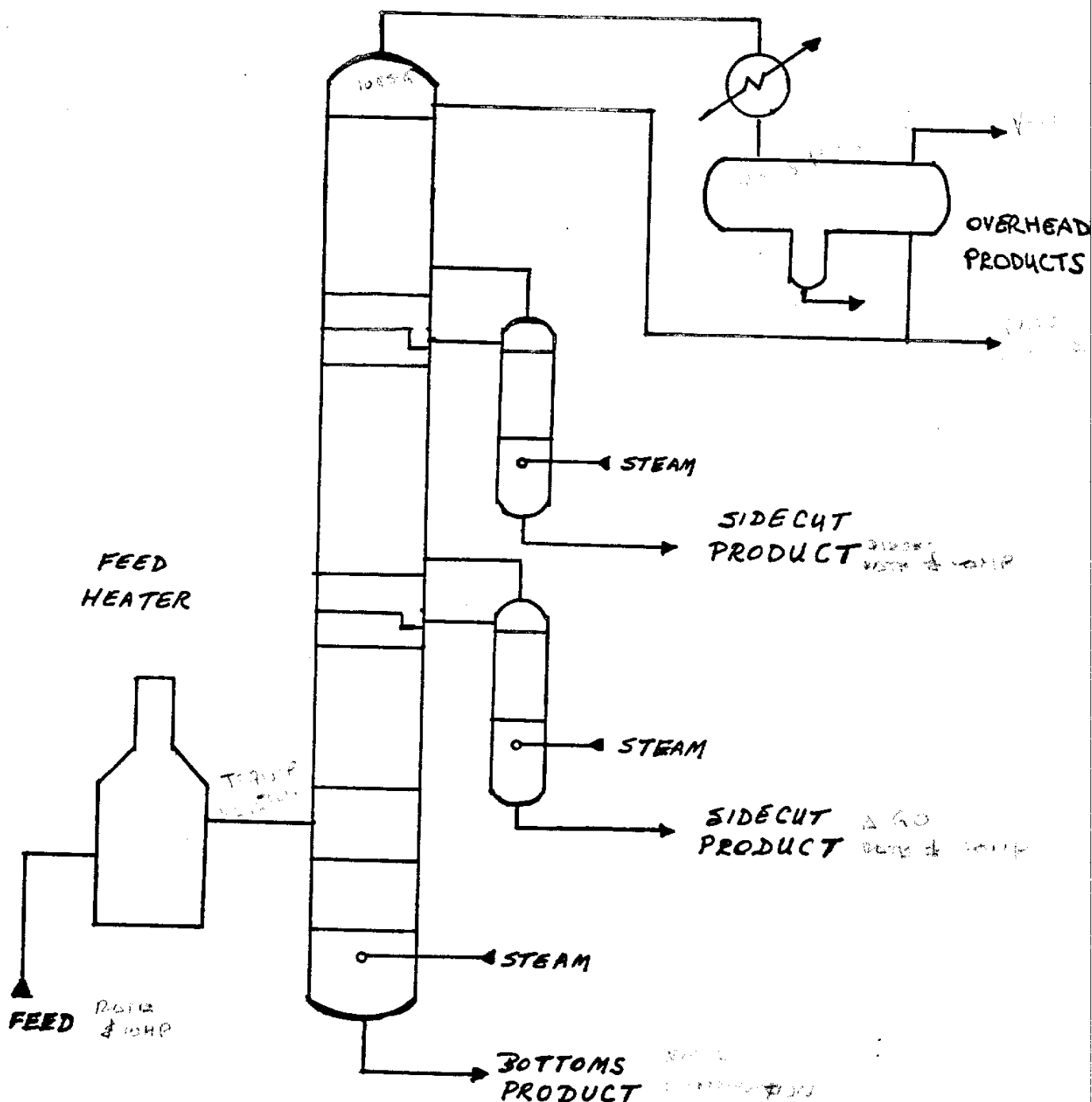


PETROCHEMICAL PROCESSES

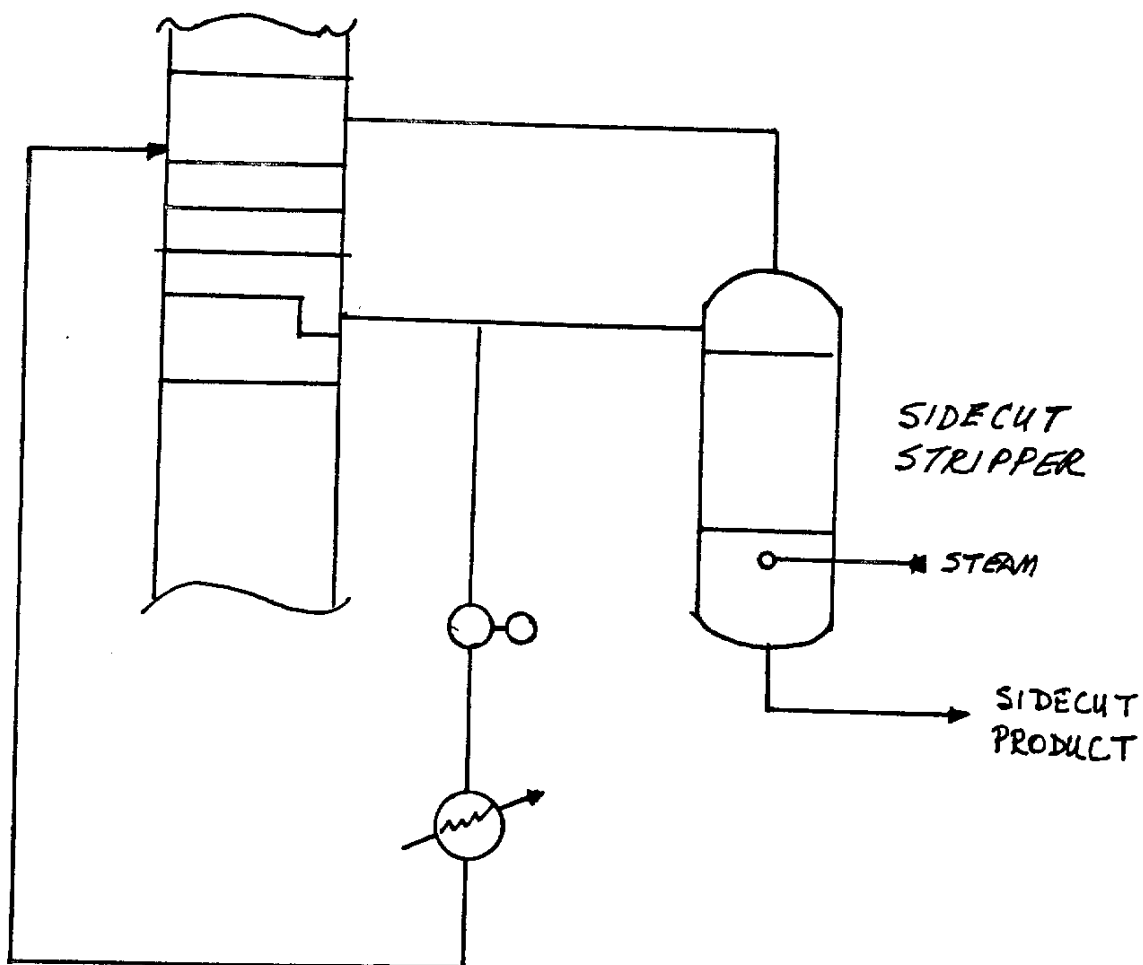


Note: Other applications of UOP Processes, Process Plants and Equipment described in this brochure but not shown on these diagrams include: PolyBed PSA, Cryogenic Plants, Benfield Process, Molecular Sieve Driers, Monirex Systems, Membrane Plants, High-Flux Tubing, and Fractionation Trays.

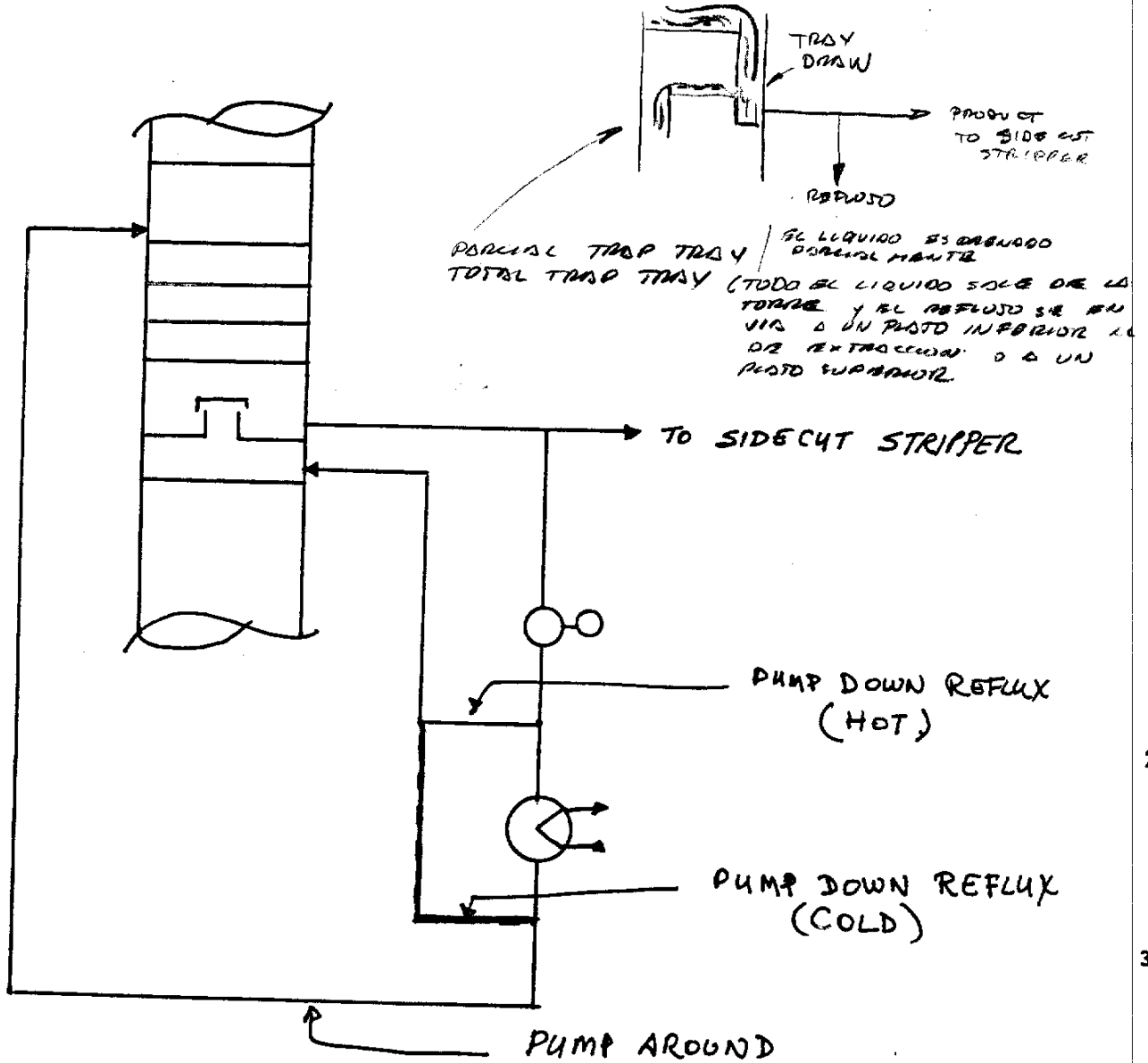
INTRODUCTION
CRUDE TYPE DISTILLATION



INTRODUCTION
CRUDE TYPE DISTILLATION
PUMP AROUND HEAT REMOVAL.



INTRODUCTION.
CRUDE TYPE DISTILLATION.
PUMP DOWN REFLUX / HEAT REMOVAL
TIPOS DE REFLUJOS.



SE EL REFLUJO SE EN VIA A UN PUNTO INFERIOR SE TIENE UN MOTOR CONTROL DEL REFLUJO A LOS PUNOS INFERIORES, INTERNO

CRUDE UNIT DESIGN BASIS.

THE CRUDE DISTILLATION UNIT WILL BE DESIGNED TO PROCESS 100,000 BPSD OF LIGHT ARABIAN CRUDE. THE DESIGN SHOULD BE BASED ON THE CHEVERON OIL NOVEMBER 1965 CRUDE ASSAY DATA. THE FOLLOWING PRODUCTS, WITH A NOMINAL TBP CUT POINT AS INDICATED, WILL BE PRODUCED:

- 1) DEBUTANIZED NAPHTHA - TBP CUTPOINT 375 °F (190 °C).
- 2) JET FUEL - TBP CUTPOINT 540 °F (282 °C).
- 3) DIESEL FUEL - TBP CUTPOINT 625 °F (329 °C).
- 4) ATMOSPHERIC GAS OIL - TBP CUTPOINT 700 °F (377 °C).
- 5) ATMOSPHERIC RESIDUAL OIL.

FROM THIS CRUDE WE DO NOT EXPECT TO PRODUCE A CRUDE COLUMN OVERHEAD NET GAS. FOR THIS REASON, A NET GAS COMPRESSOR WILL NOT BE REQUIRED. THE DESIGNER IS TO CONFIRM THAT THIS IS THE CASE AND TO DISCUSS THE PROS AND CONS OF THIS DECISION.

AS INDICATED, THE FULL RANGE NAPHTHA PRODUCT WILL BE DEBUTANIZED. THE BUTANE AND LIGHTER COMPONENTS OF THE CRUDE WILL BE RECOVERED AS A LIQUID FOR FURTHER PROCESSING IN A DOWNSTREAM LPG RECOVERY UNIT. THE BUTANE CONTENT OF THE NAPHTHA WILL BE LIMITED TO 1 MOL % MAXIMUM AND THE PENTANE CONTENT OF THE DEBUTANIZER ~~OVERHEAD~~ OVERHEAD LIQUID WILL BE LIMITED TO 1 MOL % MAXIMUM.

CRUDE DESALTING WILL BE REQUIRED. DISCUSSIONS WITH DESALTING EQUIPMENT SUPPLIERS HAVE INDICATED THAT THE DESALTER SHOULD BE DESIGNED TO OPERATE AT 250 °F (121 °C). DESALTER WATER REQUIREMENTS WILL BE 5 VOL % OF CRUDE RATE. FOR DESIGN PURPOSES, PRESSURE DROP ACROSS CRUDE/DESALTER WATER MIXER WILL BE 25 PSI (1.8 BAR).

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGN.PAGE NO. II-2DATE 5-20-91

PROJECT

BY EN MILLERCRUDE UNIT DESIGN BASIS.

FOR DESIGN PURPOSES, CRUDE IS AVAILABLE FROM STORAGE AT 60°F. BATTERY LIMIT CONDITIONS FOR THE PRODUCTS FROM THE UNIT ARE LISTED IN TABLE 1.

IT IS ANTICIPATED THAT THE FOLLOWING UTILITIES WILL BE AVAILABLE IN THE PROCESS AREA AS REQUIRED:

COOLING WATER SUPPLY TEMPERATURE 90°F (32°C)
COOLING WATER RETURN TEMPERATURE 120°F (49°C) MAX

STEAM.

150 PSIG (10.5 BAR) SATURATED.
50 PSIG (3.5 BAR) SATURATED.

FUEL GAS 50 PSIG (3.5 BAR)

FUEL OIL AS REQUIRED

POWER AS REQUIRED - 60 Hg.

STEAM PRESSURES CAN BE CONSIDERED AS MINIMUM AVAILABLE AT PROCESS BATTERY LIMITS FOR PROCESS DESIGN PURPOSES.

DESIGN AIR TEMPERATURE FOR AIR FIN CONDENSERS/COOLERS WILL BE 95°F (35°C)

THE FOLLOWING COSTS OF UTILITIES CAN BE USED FOR EVALUATING COST/PAYBACK BENEFITS OF INCREMENTAL CAPITAL INVESTMENT. A SIMPLE PAYOUT ANALYSIS CAN BE USED WITH THE MAXIMUM PAYOUT PERIOD BEING TWO YEARS.

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT DESIGNPAGE NO. II-3DATE 9-22-91PROJECT ENAMILLERBY ENAMILLER

CRUDE UNIT DESIGN BASIS.

UTILITY VALUES:

FUEL OIL.	\$ 3.50 / MM BTU/HR
POWER	35 ¢ / KW
STEAM.	
150 PSIG	\$ 3.10 / 1000 LBS/HR.
50 PSIG	\$ 2.50 / 1000 LB/HR.
COOLING WATER	\$ 1.60 / 1000 GPM.

ADDITIONAL BASIS INFORMATION WILL BE PROVIDED AS REQUIRED AS THE BASIC PROCESS DESIGN IS DEVELOPED.

THIS DESIGN BASIS HAS BEEN SUMMARIZED IN THE FOLLOWING OVERHEAD SLIDE.

CRUDE UNIT DESIGN BASIS.
TABLE 1.
BATTERY. LIMIT CONDITIONS.

FEEDS	PRESSURE		TEMPERATURES.	
	PSIG	(BAR)	°F	(°C).
CRUDE OIL	0	0	60	(16)
PROCESS WATER	50	3.5	100	(38)

PRODUCTS.

DEBUTANIZED NET OVHD. LIQ.	200	(14)	TO BE DETERMINED	
DEBUTANIZED NAPHTHA.	50	(3.5)	100	(38)
JET FUEL	50	(3.5)	100	(38)
DIESEL FUEL	50	(3.5)	120	(49)
ATMOSPHERIC GAS OIL	50	(3.5)	120	(49)
REDUCED CRUDE	TO BE DETERMINED.		175	(80)

CRUDE UNIT DESIGN BASIS.

CRUDE UNIT CAPACITY: 100 000 BPSD

CRUDE TYPE: LIGHT ARABIAN.
(USE CHEVRON OIL NOV 1965 ASSAY DATA)

PRODUCTS

- 1) LIGHT-ENDS FROM DEBUTANIZER.
 - a) MAX PENTANES - 1 MOL %
- 2) DEBUTANIZED NAPHTHA.
 - a) MAX BUTANES - 1 MOL %
 - b) NOMINAL TBP CUTPOINT 375°F (190°C)
- 3) JET FUEL
 - a) NOMINAL TBP CUTPOINT 540°F (282°C)
- 4) DIESEL FUEL
 - a) NOMINAL TBP CUTPOINT 625°F (329°C)
- 5) ATMOSPHERIC GAS OIL
 - a) NOMINAL TBP CUTPOINT 700°F (371°C)
- 6) ATMOSPHERIC RESIDUAL OIL

UTILITIES

- 1) COOLING WATER AVAILABLE AT 90°C (32°C).
- 2) COOLING WATER ΔT MAX = 30°F (17°C)
- 3) DESIGN AIR TEMPERATURE 95°F (35°C)
- 4) STEAM.
 - 150 PSIG SATURATED (10.5 BAR)
 - 50 PSIG SATURATED (3.5 BAR)
- 5) FUEL GAS AVAILABLE AT 50 PSIG (3.5 BAR).

CRUDE UNIT DESIGN.
PRODUCT YIELDS.
(OVERALL MATERIAL BALANCE)

APPROACH:

1. CONSTRUCT BLOCK FLOW FROM DESIGN BASIS.
2. OBTAIN PRODUCT YIELDS FROM CRUDE BREAKUP
3. RATIONALIZE INCONSISTENCIES IN DATA.
4. PRODUCE OVERALL MATERIAL BALANCE.

NOTE:

IN THE CRUDE BREAKUP DISCUSSION, THE CLASS PROBLEM INVOLVED BREAKING THE SAME CRUDE INTO SIMILAR FRACTIONS. USE THIS AS YOUR STARTING POINT.

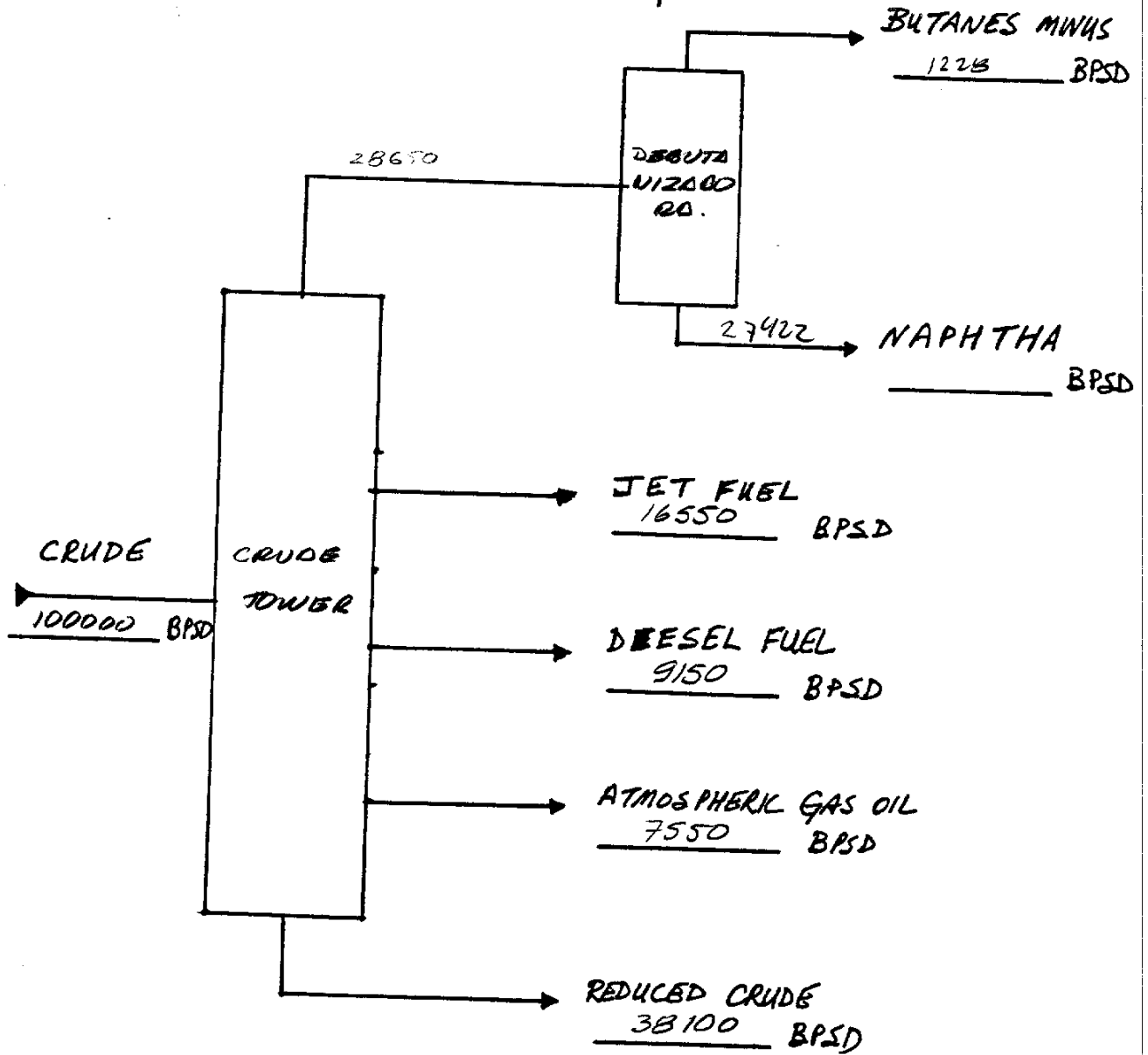
HINTS:

SMALLER PIECES CAN BE SUMMED RATHER THAN RECUR THE CRUDE

WHERE IS IMBALANCE?

CHECK CRUDE UOP K VALUES.

CRUDE UNIT DESIGN
PRODUCT YIELDS.
BLOCK FLOW DIAGRAM.



1°

CRUDE UNIT DESIGN
PRODUCT YIELDS

DE LOS DATOS DEL CRUDE BREAK-UP (PROBLEMA DE TRAMPOL)
ADJUST CLASS PROBLEM SOLUTION TO FIT CRUDE UNIT DESIGN BASIS.

1. C6-375°F NAPHTHA: CONSTRUCT FROM SUM OF LSR GASOLINE AND NAPHTHA. NOMINAL TBP CUTOPOINT SAME.

	API	UOP K	MW	BPSD	LB/HR	MOL/HR	LV%
LSR NAPHTHA	70.6	12.35	95	10830	110640	1164.63	10.83
NAPHTHA	53.3	11.97	125	14150	158085	1264.68	14.15
C6+ NAPHTHA	60.4	12.13	111	24980	268725	2429.31	24.98

$$\text{AVERAGE MOL WT} = \frac{268725}{2429.31} = 110.6 \Rightarrow 111$$

API WEIGHT AVERAGE.

$$API_{MIX} = \left(\frac{110640}{268725} \right) (70.6) + \left(\frac{158085}{268725} \right) (53.3) = 60.42$$

UOP K WEIGHT AVERAGE.

$$UOP K_{MIX} = \left(\frac{110640}{268725} \right) (12.35) + \left(\frac{158085}{268725} \right) (11.97) = 12.13$$

NOTE IN ORIGINAL "K" CALCULATED FROM CHARTS B-2 & B-3 A LITTLE LOW.
COULD HAVE ADJUSTED CABP UP SLIGHTLY.

$$\text{CABP OF } 310^{\circ}\text{F WITH } 53.3 \text{ API} \Rightarrow K = 11.97$$

2. ADJUST DIESEL 50% POINT TO ALLOW A BETTER UOP K MATCH. FIRST TRIAL 600 → 590°F

$$API = 35 \quad \text{CABP} \approx 590^{\circ}\text{F} \quad UOP K = 11.95 \approx 11.93$$

$$\text{SLOPE} = 0.89 \quad VABP = 590 + 1.5 = 591.5$$

$$MABP = 591.5 - 3.5 = 588 \Rightarrow MW = 253$$

A.

CRUDE UNIT DESIGN PRODUCT YIELDS.

ADJUST CLASS PROBLEM SOLUTION TO FIT CRUDE DESIGN BASIS.

3. ADJUST SLOP WAX 50% POINT TO ALLOW A BETTER OOPK MATCH.
FIRST TRIAL $980 \rightarrow 960^\circ\text{F}$

$$\begin{aligned} \text{API} &= 20.7 & \text{CASP} &\approx 960^\circ\text{F} & \text{UOP K} &= 12.05 & \approx 12.05 \text{ OK} \\ \text{SLOPE} &= 1.0 & \text{UASP} &= 960 + 1 & 961^\circ\text{F} \\ \text{MASP} &= 961 - 4 = 957^\circ\text{F} & \Rightarrow \text{MW} &= 550 \end{aligned}$$

4. CREATE REDUCED CRUDE BY SUMMING BOTTOMS FRACTIONS.

COMPONENT.	API	UOP K	MW	BPSD	LB/HR	MOLE/HR	LV%.
LVGO	31.4	11.90	274.9	4600	58295	212.06	4.60
4VGO	23.5	11.96	435	15300	203800	468.51	15.30
* SW	20.7	12.05	550	2200	29845	54.26	2.20
ASPHALT	8.5	11.41	580	16000	235955	406.82	16.00
REDUCED CRUDE	17.51	11.72	462.4	38100	527895	1141.65	38.10.

* ADJUSTED. $AV \text{ MOL WT} = \frac{527895}{1141.65} = 462.4$

API WEIGHT AVERAGE.

$$\begin{aligned} \text{API} &= \left(\frac{58295}{527895} \right) (31.4) + \left(\frac{203800}{527895} \right) (23.5) + \left(\frac{29845}{527895} \right) (20.7) + \left(\frac{235955}{527895} \right) (8.5) \\ &= 17.51. \end{aligned}$$

UOP K WEIGHT AVERAGE.

$$\begin{aligned} \text{UOP K} &= \left(\frac{58295}{527895} \right) (11.90) + \left(\frac{203800}{527895} \right) (11.96) + \left(\frac{29845}{527895} \right) (12.05) + \left(\frac{235955}{527895} \right) (11.41) \\ &= 11.72 \end{aligned}$$

NOTE: WE DID NOT ADJUST PROPERTIES OF DIESEL SLUMP INTO LVGO.

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGNPAGE NO. III - 4ADATE 9-20-91

PROJECT

BY E N MILLER

CRUDE UNIT DESIGN PRODUCT YIELDS

ADJUST CLASS PROBLEM SOLUTION TO FIT CRUDE DESIGN BASIS.

→ 5. DETERMINE UOP K FOR CRUDE USING API AND VISCOSITY DATA.

VISC AT 100°F CS = 5.5 API = 34.5 ⇒ UOP K ≈ 11.9-12.0
130°F CS = 9.0

AT 210°F CS = 2.3

AT 122°C CS = 4.35

6. SUMMARIZE RESTRUCTURED CRUDE BREAKUP.

	API	UOP K	MW	BPSD	LB/HR	MOL/HR	LV%
CRUDE	34.5	11.90	211.6	100000	1243700	5876.91	100

ESTIMATED YIELDS

C3		14.7	44.094	170	1260	28.56	0.17
IC4		13.81	58.12	170	1395	24.03	0.17
NC4		13.50	58.12	1060	9040	155.49	1.06
ICS		13.05	72.146	730	6655	92.24	0.73
NCS		13.03	72.146	1470	13535	187.62	1.47
CP		11.11	70.13	70	765	10.93	0.07
→ NAPHTHA.	60.4	12.13	111	24980	268725	2429.31	24.98
JET FUEL	42.8	11.97	187	16550	196035	1048.32	16.55
DIESEL	35.0	11.93	253	9150	113460	448.46	9.15
AGO	29.5	11.87	312	7550	96815	310.30	7.55
RESID	33.5	11.72	462.4	38100	527895	1141.65	38.10
Σ	35.6	11.92	210.3	100,000	1235580	5876.91	100.00

IN BALANCE IN MASS FLOW = $\frac{1243700 - 1235580}{1243700} \times 100 = \underline{\underline{0.65\%}}$

CUSTOMER ENGINEERING DESIGN SENWAR.SUBJECT CRUDE UNIT DESIGNPAGE NO. III-5A.DATE 9-20-91

PROJECT

BY E. N. MILLER

CRUDE UNIT DESIGN PRODUCT YIELDS.

REFERRING TO RESTRUCTURED CRUDE BREAKUP:

ALTERNATIVES DATA FOR 200 A 1212 CRUDE W/ BALANCE
RESTRUCTURED.
A NUMBER OF OPTIONS ARE AVAILABLE TO FORCE A WEIGHT
BALANCE. THE API'S OF ALL THE C6+ PRODUCTS CAN
BE ADJUSTED DOWN BY 0.5 API NUMBER. THIS WILL
ACCOUNT FOR ABOUT 50% OF THE INBALANCE. FURTHER
ADJUSTMENTS IN PRODUCTS CAN BE MADE OR THE API OF
THE FEED CAN BE INCREASED. THIS SITUATION WILL
ALWAYS ARISE BECAUSE THE APPROACHED USED FORCES
A VOLUME BALANCE CREATING A SMALL MASS INBALANCE.
THERE IS A 1-2% VOLUME EXPANSION OF PRODUCTS WHEN
A CRUDE IS FRACTIONATED.

FOR THIS DESIGN PROBLEM, WE WILL ADJUST THE CRUDE
PROPERTIES TO MATCH PRODUCTS.

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGN.PAGE NO. III-6ADATE 9-20-91

PROJECT

BY ENMILLERCRUDE UNIT DESIGN
PRODUCT YIELDS.

SPLIT OF CRUDE COLUMN OVERHEAD INTO BUTANES MINUS FRACTION AND DEBUTANIZED NAPHTHA IN A DEBUTANIZING COLUMN. WITH THE LIGHT ENDS YIELD FROM THE CRUDE BREAK-UP MAKE A SPLIT IN THE DEBUTANIZER CONSISTENT WITH THE DESIGN BASIS. ASSUME CRUDE TOWER OVERHEAD TOTALLY CONDENSING.

	FLOWS IN LB MOL/HR.		
	DEBUTANIZER FEED	OVERHEAD	BOTTOMS.
C3	28.56	28.56	-
iC4	24.03	24.03	-
NC4	155.49	128.29	27.20
iC5	92.24	1.81	90.43
NC5	187.62	-	187.62
CP	10.93	-	10.93
NAPHTHA.	2429.31	-	2429.31
TOTALS	2928.18	182.69	2745.49

BASIS: 1 MOL % MAX C4'S IN BOTTOMS.

1 MOL % MAX C5'S IN OVERHEAD.

MAXIMUM BUTANE CONTENT OF BOTTOMS ASSUMING ALL PENTANES IN BTMS.

$$C_4 = (0.01)(2720.1) = 27.20$$

PENTANE CONTENT OF OVERHEAD.

$$C_5 = (0.01)(208.08 - 27.2) = 1.81$$

CHECK DISTRIBUTION

$$C_4'S \text{ IN BTMS} = \frac{27.2}{2745.49} \times 100 = 0.99\% < 1.0\% \text{ OK.}$$

$$C_5'S \text{ IN OVHD} = \frac{1.81}{182.69} \times 100 = 0.99\% < 1.0\% \text{ OK.}$$

WE HAVE SELECTED NC4/IC5 AS KEY COMPONENTS TO MAKE SPLIT IN COLUMN. WE NEED TO CONFIRM THE VALIDITY OF THIS ASSUMPTION LATER.

CRUDE UNIT DESIGN
OVER ALL MATERIAL BALANCE
(MIL / HR)

	CRUDE	CA MINUS	NAPHTHA	JET FUEL	DIESEL	AGO	RESID
C3		28.56					
IC4		24.03					
MC4		128.29	27.20				
IC5		1.81	90.43				
MC5			187.62				
RP			10.93				
NAPHTHA			2429.31				
JET FUEL				1048.32			
DIESEL					448.46		
AGO						310.30	
RESID.							1141.65
TOTAL	5876.91	182.69	2745.49	1048.32	448.46	310.30	1141.65
CB/HR	1235380	10240	291135	196035	113460	96815	527895
MW	210.3	57.05	106.04	187	253	312	462.4
BPSD	100000	1228.	27422	16550	9150	7550	38100
API	36.6		63.0	42.8	35.0	29.5	17.5
MMSCFD							

BTH'S = 38.1 %
LIFT = 61.9 %

CRUDE UNIT DESIGN FLASH ZONE CONDITIONS.

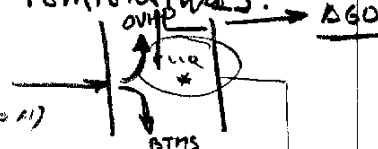
1. CONSTRUCT FLASH CURVES USING PROCEDURES OUTLINED IN SECTION ON FLASH CURVES.

PLOT 0 PSIG "PACIFIC" FLASH CURVE.
 15 PSIG FLASH CURVE
 20 PSIG FLASH CURVE.

2. ESTIMATE FLASH ZONE (HEATER OUTLET) TEMPERATURES.

- a) USE 3 % VOL OVERFLASH.

$$\% \text{VOL} = (100 - 38.1) + 3 = 64.9 \text{ (INCLUDED BE OVERFLASH)}$$



REMEMBER OVERFLASH IS STATED AS A VOL % OF CRUDE AND IS IN ADDITION TO NET PRODUCTS VAPORIZED. (TYPICAL 3-5% VOL OF OVERFLASH IN UNDERTOP OF CRUDE)

* OVERFLASH FROM HEATER REFLUID BUTTER TO COLUMN 1 COLUMN.

- b) IUS WOULD LIKE TO KEEP THIS TEMPERATURE AT OR BELOW 730°F.

PRACTICAL MINIMUM OPERATING PRESSURE AT FLASH ZONE IS ABOUT 15 PSIG.

3. LOOK AT SLOPE OF EFV CURVE AT 15 PSIG USING FOCAL POINT METHOD PREVIOUSLY OUTLINED.

4. ONE SET OF OPERATING DATA FOR THIS CRUDE INDICATES A TRANSFER TEMPERATURE OF 685°F AT 59.5 VOL % LIFT AND 15 PSIG FLASH ZONE TEMPERATURE.

5. DISCUSS CHOICE OF FLASH ZONE CONDITIONS FOR CLASS DESIGN PROBLEM.

- a) REVISIT CONSTRUCTION OF FLASH CURVES & LOCATE NEW DATA
- b) 2030

CRUDE UNIT DESIGN.
FLASH ZONE.

USING PROCEDURES OUTLINED IN SECTION ON FLASH CURVES

- a) PLOT TBP DISTILLATION CURVE
b) FIND DISTILLATION REFERENCE CURVE

$$10\% \quad 204.4^{\circ}\text{F}$$

$$70\% \quad 790^{\circ}\text{F}$$

$$\text{SLOPE} = \frac{(790 - 204.4)}{(70 - 10)} = 9.76 \quad (\text{TBP CURVE})$$

- c) SLOPE OF FLASH REFERENCE CURVE (FROM PD 168-1)
SLOPE = 6.9

- d) CALCULATE AVERAGE MIDPOINT OF TBP CURVE
20% = 300°F 50% = 586.7°F 80% = 928.3°F

$$\text{MIDPOINT} = \frac{20 + 30 + 50}{3} = \frac{300 + 586.7 + 928.3}{3} = 605$$

FROM PD-167

$$\begin{aligned} 50\% \text{ POINT ON FLASH REFERENCE CURVE} \\ = 605 - 39 = 566^{\circ}\text{F} \end{aligned}$$

- e) DRAW FLASH REFERENCE LINE

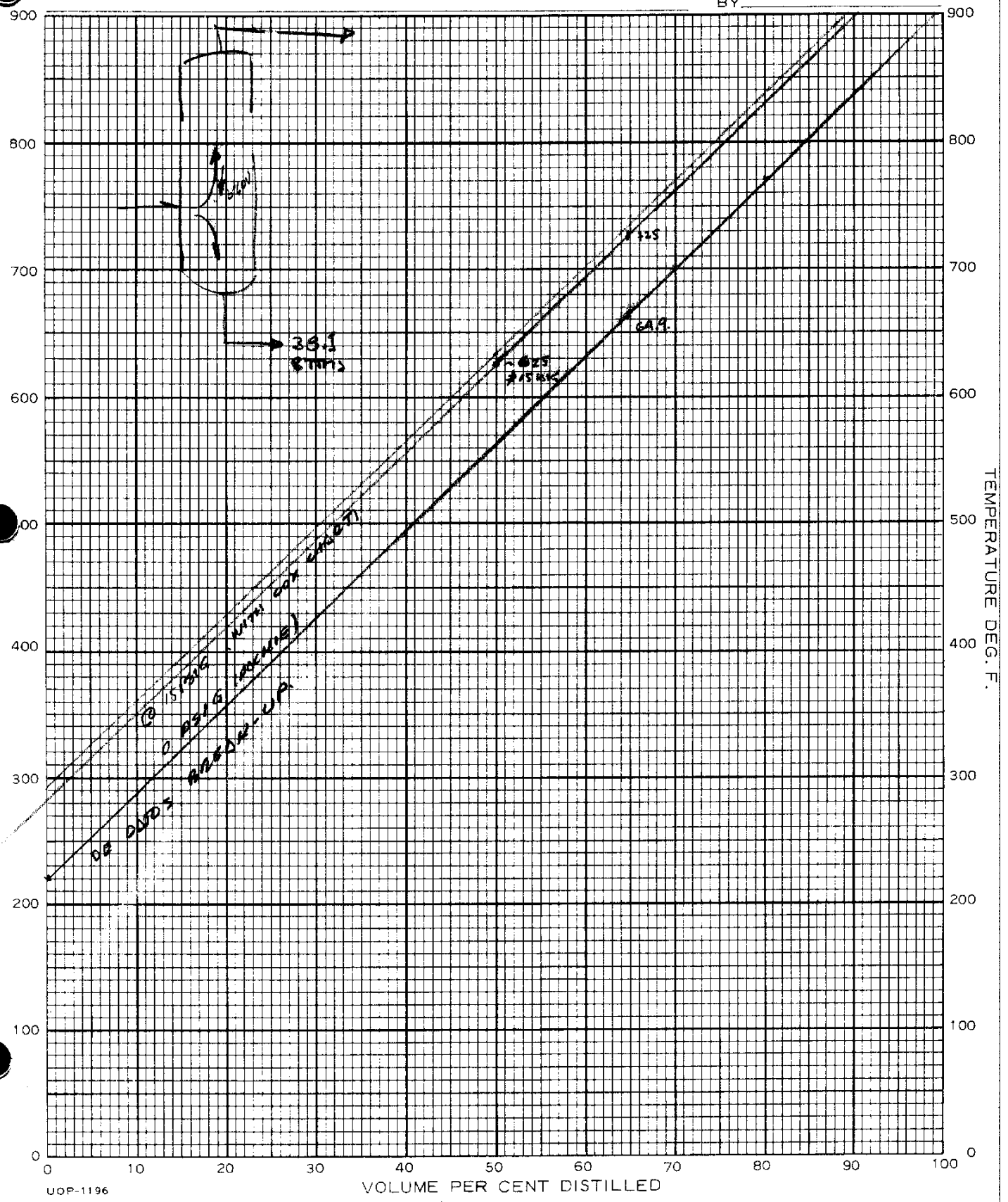
$$\frac{566 - t_0}{50 - 0} = 6.9 \quad t_0 = 566 - (50)(6.9) = 221^{\circ}\text{F}$$

$$\frac{t_{100} - 566}{100 - 50} = 6.9 \quad t_{100} = 566 + (6.9)(50) = 911^{\circ}\text{F}$$

DISTILLATION CHART

CUSTOMER _____
 SUBJECT _____

PAGE NO. _____
 DATE _____
 PROJECT _____
 BY _____



CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGNPAGE NO. 1V-2ADATE 9-20-91

PROJECT

BY EN MILLER

CRUDE UNIT DESIGN. FLASH ZONE.

- 5) INTERSECTION OF DISTILLATION AND FLASH REFERENCE CURVES IS AT 505°F.

ESTIMATE FLASH CURVE AT 15 PSIG AND 20 PSIG USING COX CHART (CHART 4-3) VOL % = 41

$$\begin{aligned} \text{AT } 15 \text{ PSIG} &= 29.7 \text{ PSIA} & T &= 570^\circ\text{F} \\ 20 \text{ PSIG} &= 34.7 \text{ PSIA} & T &= 585^\circ\text{F} \end{aligned}$$

15 PSIG FLASH CURVE

$$\frac{570 - t_0}{41 - 0} = 6.9 \quad t_0 = 570 - (6.9)(41) = 287^\circ\text{F}$$

$$\frac{t_{100} - 570}{100 - 41} = 6.9 \quad t_{100} = 570 + (6.9)(59) = 977^\circ\text{F}$$

20 PSIG FLASH CURVE

$$\frac{585 - t_0}{41 - 0} = 6.9 \quad t_0 = 585 - (6.9)(41.0) = 302^\circ\text{F}$$

$$\frac{t_{100} - 585}{100 - 41} = 6.9 \quad t_{100} = 585 + (6.9)(59) = 992^\circ\text{F}$$

- 8) ESTIMATE FLASH ZONE (HEATER OUTLET) TEMPERATURES.

$$\text{REQUIRE \% LIFT} = 100 - 38.1 = 61.9\%$$

$$\text{USE } 3\% \text{ OVERFLASH} = 3.0\%$$

$$\text{TOTAL LIFT} = 64.9\%$$

$$\text{AT } 20 \text{ PSIG } T \approx 750^\circ\text{F} \quad \text{AT } 15 \text{ PSIG } T \approx 730^\circ\text{F}$$

THIS IS ON HIGH SIDE FOR THIS CRUDE BUT IN GENERAL PALKIE PREDICTS HIGH. IN FINAL ANALYSIS WE MAY NEED TO SHIFT CUT POINT FOR AGO PRODUCT.

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN.

PAGE NO. 1V-34
 DATE 2-20-91
 PROJECT _____
 BY ENRIQUEL

CRUDE UNIT DESIGN. FLASH ZONE

- 2) LOOK AT SLOPE OF EFV CURVES USING FOCAL POINT METHOD PREVIOUSLY OUTLINED.

AT 40% VAPORIZATION ON 0-PSIG EFV CURVE
 TEMPERATURE = 497°F.

FOR NORMAL PARAFFIN WITH NBP = 497°F $\log K \approx 12.6$
 FROM CHART B-2 $API = 49.5$

FROM CHART D-2 $P_c \approx 225 \text{ PSI}$ $T_c \approx 795^\circ\text{F}$
 * *

ADD 150°F TO $T_c = 945^\circ\text{F}$.

BY THIS METHOD 15 PSIG EFV CURVE DATA POINTS ARE

0 % - 295°F	40% - 560°F	70% - 750°F
10 % - 315-370°F.	50% - 620°F	80% - 820°F
30 % - 500°F	60% - 680°F	84% - 835°F

AT DESIGN LIFT ESTIMATED TEMPERATURE
 = 715°F.

- 3) PLOT COMMERCIAL DATA POINT.
 $\% \text{ LIFT} = 59.5$ $T = 685^\circ\text{F}$ $P = 15 \text{ PSI}$

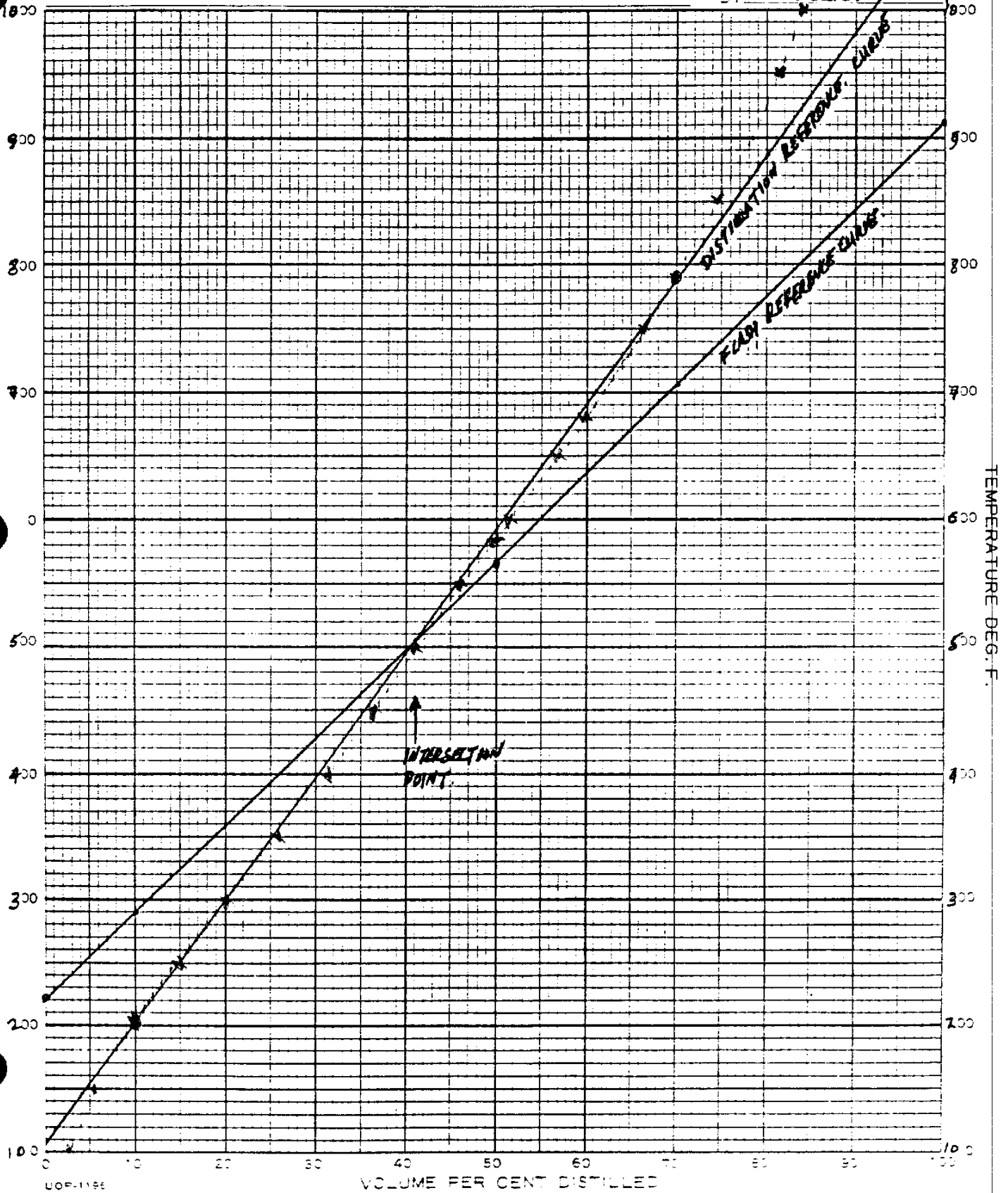
THIS AGREES WITH 15 PSIG FLASH CURVE
 ADJUSTED FOR CURVATURE.

USE FOR THIS DESIGN $T = 715^\circ\text{F}$
 $P = 15 \text{ PSI}$.

DISTILLATION CHART

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT ARABIAN LIGHT CRUDE
FLASH CURVES

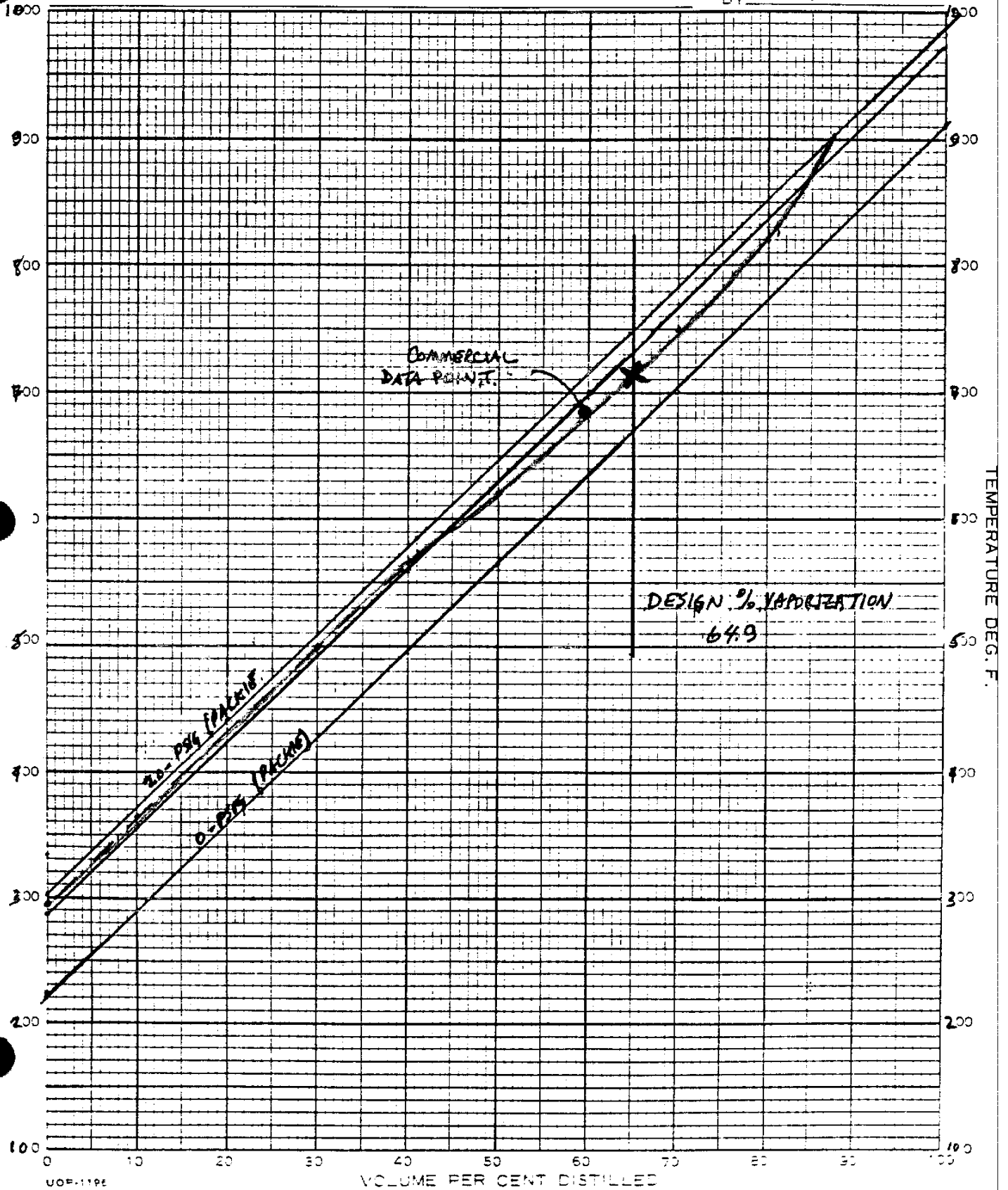
PAGE NO. IV-4A
 DATE 8-20-51
 PROJECT
 BY EN MILLER



DISTILLATION CHART

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT ARABIAN LIGHT CRUDE
CRUDE UNIT EFF CURVES

PAGE NO. 1V-5A
 DATE 9-20-91
 PROJECT _____
 BY ENA



TEMPERATURE DEG. F.

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN

PAGE NO. V-1
 DATE 9-20-91
 PROJECT _____
 BY EN MILLER

CRUDE UNIT DESIGN CRUDE COLUMN OVERHEAD

INVESTIGATE BRIEFLY CASE FOR ADDING A SPIKE TO CRUDE FOR DESIGN PURPOSES.

RECALL BASIS FOR THIS CALCULATION WAS 5 PSIG RECEIVER AT 100 °F. EXPECTED RECEIVER GAS IS ABOUT 0.5 VOL% IF CRUDE INITIAL ADJUSTMENT: 2 TIMES METHANE AND ETHANE CONTENT
 3 TIMES PROPANE CONTENT OF ASSAY

FROM OUR INITIAL YIELD ESTIMATE WORK NET OVERHEAD WITHOUT SPIKE WAS

	VOL % CRUDE	BPSD	LB/HR	MOLE/HR	MOLE FRACTION
C3	0.17	1070	1260	28.56	9.75×10^{-3}
IC4	0.17	170	1395	24.03	8.21×10^{-3}
NC4	1.06	1060	9040	155.49	5.31×10^{-2}
ICS	0.73	730	6655	92.24	3.15×10^{-2}
MCS	1.47	1470	13535	187.62	6.41×10^{-2}
CP	0.07	70	765	10.93	3.73×10^{-3}
NAPHTHA	24.98	2498	268725	2429.31	0.82961
				<u>2928.38</u>	

NO METHANE AND ETHANE IN CRUDE. FOR SPIKE USE 0.25 VOL% ETHANE.

EQUILIBRIUM DICTATES THAT $\sum y_i = \sum x_i K_i = 1.0$ FOR BP LIQUID.

WE CAN USE K_i CHARTS FOR PURE COMPONENTS. BUT STILL NEED TO DETERMINE K VALUE FOR NAPHTHA COMPONENT.

HOW DO WE DO THIS?

USE FUGACITY COEFFICIENT APPROACH
(GENERALIZED FUGACITY CHARTS)

DETERMINE IF OVERHEAD IS AT BUBBLE POINT AT 100°F 15 PSIG WITHOUT AND WITH SPIKE.

CUSTOMER EN ENGINEERING DESIGN SEMINAR
SUBJECT CRUDE UNIT DESIGN.

PAGE NO. V-2
DATE 9-20-91
PROJECT _____
BY EN MILLER

CRUDE UNIT DESIGN.
CRUDE COLUMN OVERHEAD

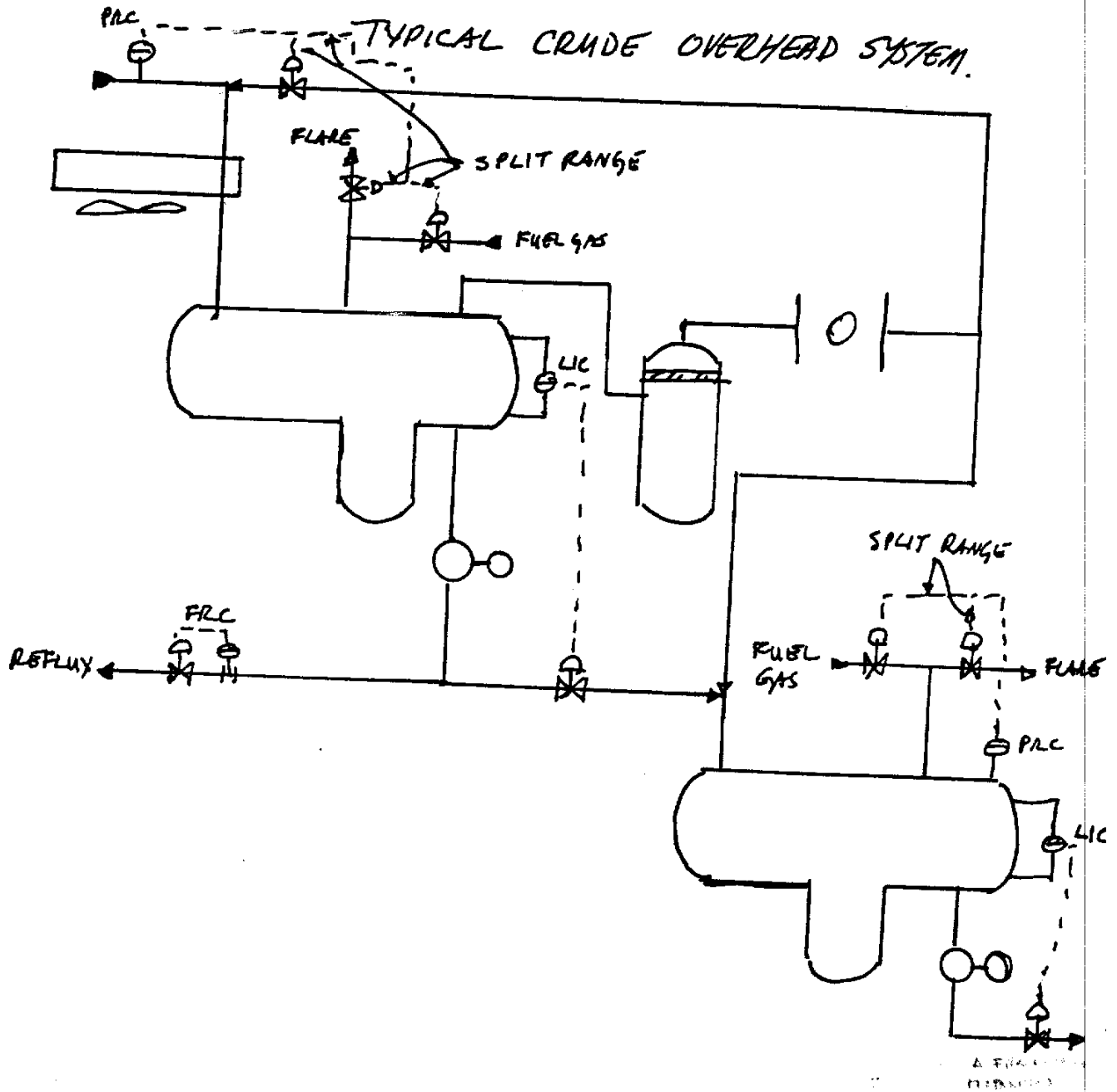
IN GENERAL PURPOSE FOR INCLUDING LIGHT ENDS SPIKE
CAN BE SUMMARIZED AS FOLLOWS:

1. ACCOUNT FOR WEATHERING OF ASSAY SAMPLE
2. BASIS FOR NET GAS COMPRESSION DESIGN
(RECIPROCATING MACHINE: MOL WT OF GAS NOT FACTOR)
3. ALLOWS FOR FEED TYPE FLEXIBILITY
4. PROVIDES SUFFICIENT PRESSURE TO ENSURE
CRUDE REMAINS LIQUID PHASE AT DESALTER
AND AT HEATER FRC CONTROL VALVES.

ONCE SPIKE IS ADDED, IT IS CARRIED THROUGH FRACTIONATION
SECTION.

FOR PURPOSE OF CURRENT DESIGN PROBLEM, IGNORE NET
GAS COMPRESSION DESIGN AND ASSUME THAT COLUMN
RECEIVER WILL OPERATE AT BUBBLE POINT OR
SLIGHTLY SUB COOLED.

CRUDE UNIT DESIGN.
CRUDE COLUMN OVERHEAD.



CRUDE UNIT DESIGN CRUDE COLUMN OVERHEAD

DETERMINE EQUILIBRIUM "K" FOR NAPHTHA AT 5PSIG AND 100°F.

WHAT DO WE KNOW ABOUT THIS NAPHTHA FROM PREVIOUS CALCS.

$$API = 60.4$$

$$LWP K = 12.13$$

$$MW = 111$$

$$16P \approx 120-140^\circ F$$

$$EP \approx 321^\circ F$$

$$SLOPE \approx 2.41$$

CHART B-2

CHART B-1

$$CABP \approx 258^\circ F$$

$$VABP = 258 + 6 = 264^\circ F$$

$$MABP = 264 - 16 = 248^\circ F$$

$$MoABP = 264 - 25 = 239^\circ F$$

Se assume que os par
componentes simples (NAPHTHA)
No ABP = NBP.

DETERMINE CRITICAL PROPERTIES FROM CHART D-2.

PSEUDO CRITICAL PRESSURE ≈ 405 PSIA.

PSEUDO CRITICAL TEMPERATURE $\approx 555 + 460 = 1015^\circ F$.

$$T_R = \frac{T}{T_c} = \frac{560}{1015} = 0.55 \quad P_R = \frac{P}{P_c} = \frac{14.7+5}{405} = 0.0494$$

FROM FUGACITY CHART $V_{T,\pi} = 0.925$

NEED VAPOR PRESSURE DATA. $\rightarrow$ USE COX CHART.

DATOS

NBP $\approx 240^\circ F$ AT 14.7 PSIA $\rightarrow VP @ 100^\circ F \approx 0.88$ PSIA.

$$P_R = \frac{p^*}{P_c} = \frac{0.88}{405} = 0.0022$$

$$V_{T,p^*} \approx 0.99 - 1.0$$

DE LA CARTA DE FUGACIDAD

$$K_{T,\pi} = \frac{(V_{T,p^*})(\phi^*)}{V_{T,\pi}} = \frac{(1.0)(0.88)}{0.925} = 0.951$$

$$K_i = \frac{0.951}{\frac{19.7}{(14.7+5)}} = 0.048 \approx 0.05 \quad \text{FOR NAPHTHA @ 5PSIG + 100°F}$$

DID NOT MAKE CORRECTION FOR LIQUID PHASE NON-IDEALITY
"a" CORRECTION.



ENGINEERING DEPARTMENT

CUSTOMER

ENGINEERING DESIGN SEMINAR

SUBJECT

CRUDE UNIT DESIGN

CRUDE COLUMN OVERHEAD

PAGE NO. V-2A

DATE 9-20-91

PROJECT

BY ENHILSEN

NOW WE CAN CHECK ROSEBUD EQUILIBRIUM AT 100°F.

	NET OVERHEAD mol/HZ	100°F K _i	f _i X _i
C3	28.56	7.49	213.91
IC4	24.03	3.30	79.30
NC4	155.49	2.44	379.40
SC5	92.24	1.03	95.00
NCS	187.62	0.80	150.10
CP	10.93	0.52	5.68
NAPHTHA.	2429.31	0.05	121.46
	2928.08		1044.85

< 2928.08

i.e. BELOW BUBBLE POINT

* ADD PROPANE SPIKE i.e. TOTAL BECOMES 3 TIMES ORIGINAL PROPANE

C3 SPIKE	57.12	7.49	427.83
NEW TOTAL	2985.30		1472.68 < 2985.20

** ADD ETHANE SPIKE i.e. 0.25 VOL % ON CRUDE = 250 BRSD.

** HASTA CONSEGUIR EQUILIBRIO
EN LA ZONA DE BURBUJA

$$\text{mol/HZ} = \frac{250}{5.03} = 49.7$$

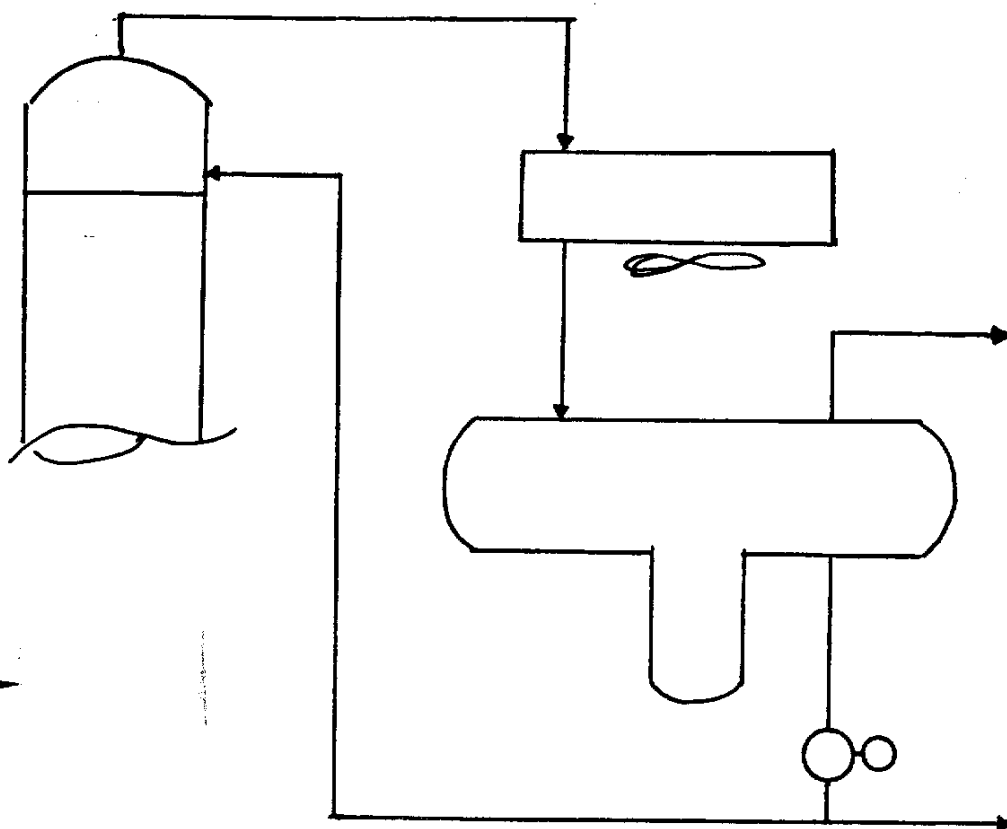
C2 SPIKE	49.7	22.95	1140.62
TOTAL	3035.00		2613.30 < 3034.90

WITH TYPICAL SPIKE CRUDE OVERHEAD RECEIVER STILL NOT TWO PHASE AT TARGET CONDITIONS OF 100°F AND 5 PSIG. ESTIMATED BUBBLE POINT PRESSURE AT 100°F WITH SPIKE IS 2.5 PSIG (17.1 PSIA).

ON THIS BASIS, OPERATION OF CRUDE TOWER OVERHEAD RECEIVER WITH ONLY NET LIQUID MAKE IS NOT AN UNREASONABLE ASSUMPTION.

MOST MODERN CRUDE UNIT DESIGNS INCLUDE PROVISION FOR NET OVERHEAD GAS COMPRESSION. ADDING A LIGHT ENDS SPIKE TO THE CRUDE IS ONE WAY, AS MENTIONED IN THE CRUDE BREAKUP DISCUSSION, TO CREATE A BASIS FOR NET GAS COMPRESSION DESIGN.

CRUDE UNIT DESIGN CRUDE COLUMN OVERHEAD SYSTEM.



DETERMINE OPERATING CONDITIONS FOR CRUDE TOWER OVERHEAD SYSTEM.

WHAT DO WE KNOW?

COMPOSITION OF NET OVERHEAD PRESSURE AT FLASH ZONE

THERE WILL BE LIQUID PHASE WATER IN OVERHEAD RECEIVER. WHY?

RECEIVER HAS NO NET VAPOR PRODUCT (DESIGN BASIS)

APPROACH:

1. PRELIMINARY HYDRAULIC ANALYSIS TO DETERMINE RECEIVER PRESSURE.
2. CALCULATE BUBBLE POINT TEMPERATURE



FOR FIN-FAN

$$Q_p = m \bar{c}_p \Delta T = m_{\text{air}} \dot{c}_p (\Delta T)_{\text{air}} \\ = m_{\text{air}} \dot{c}_p (t_6 - 95)$$

$$Q = U A (\Delta T)_{\text{LM}}$$

OR OR BE ESTIMATE U

CRUDE UNIT DESIGN CRUDE COLUMN OVERHEAD SYSTEM.

OVERHEAD RECEIVER CONDITIONS:

FOR DESIRED LIFT WE DECIDED TO RUN CRUDE FLASH ZONE AT 15 PSIG. IF WE ALLOW 5 PSI PRESSURE DROP FOR TOWER TRAYS AND ADD 5 PSI PRESSURE DROP FOR OVERHEAD SYSTEM, RECEIVER WILL OPERATE AT 5 PSIG.

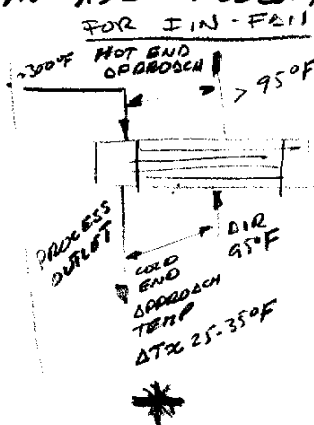
RECALL FROM DESIGN BASIS NOTES THAT DESIGN AIR TEMPERATURE WAS GIVEN AS 95°F. IF WE TARGET A RECEIVER OPERATING TEMPERATURE OF 120-130°F, ALL AIR CAN BE USED FOR CRUDE TOWER OVERHEAD CONDENSING DUTY.

A COLD END APPROACH TEMPERATURE OF 25-35°F IS REASONABLE FOR AIR FIN EXCHANGER DESIGN.

AS IMPLIED IN PREVIOUS DISCUSSIONS, STRIPPING STEAM WILL BE USED IN CRUDE TOWER FOR REMOVAL OF LIGHT ENDS FROM PRODUCTS. THEREFORE THERE WILL BE TWO LIQUID PHASES IN OVERHEAD RECEIVER. EFFECT OF WATER VAPOR SHOULD BE CONSIDERED IN RECEIVER EQUILIBRIUM CALCULATIONS.

TO SIMPLIFY CALCULATIONS, WE HAVE CALCULATED THE IDEAL KTI'S FOR TWO PRESSURES OVER A TEMPERATURE RANGE FOR THE NAPHTHA DEFINED IN THE PROBLEM.

	KTI @ 14.7 PSIA	KTI @ 50 PSIA
100 °F	0.989	1.141
120 °F	1.549	1.630
130 °F	1.960	2.061
140 °F	2.455	2.580
160 °F	3.932	4.472
180 °F	5.796	6.558



CRUDE UNIT DESIGN. CRUDE OVERHEAD SYSTEM.

COMPOSITION OF NET OVERHEAD FROM YIELD CALCULATIONS.

#	S _i	120°F		130°F		160°F	
		K _i	f _i K _i	K _i	f _i K _i	K _i	f _i K _i
C3	28.56	9.97	284.74	11.28	322.16	17.12	488.95
iC4	24.03	4.60	110.54	5.31	127.60	8.42	202.33
NC4	155.49	3.49	542.66	4.08	634.40	6.68	1038.67
iC5	92.24	1.55	142.97	1.85	170.64	3.22	297.01
NC5	187.62	1.22	228.90	1.47	275.80	2.64	495.32
CP	10.93	0.81	8.85	0.993	10.85	1.85	20.22
NAPHTHA	2429.31	0.086	208.92	0.1126	273.54	0.263	638.98
TOTAL	2928.18		1527.58		1814.99		3181.42

> 2928.18.

AT 120°F VP H₂O = 1.69 PSIA
 AT 130°F VP H₂O = 2.22 PSIA.
 AT 160°F VP H₂O = 4.74 PSIA.

AS STATED PREVIOUSLY EXPECTED OPERATING PRESSURE IN RECEIVER WILL BE 5PSIG OR 19.7 PSIA.

AT 120°F EFFECTIVE HYDROCARBON PRESSURE WILL BE
 $\pi = P_{PHC} + P_{PH_2O} = 19.7$.

THERE ARE TWO PHASES
 HC & WATER

$$P_{PH_2O} = P_{H_2O}^{\circ} \left(\text{VAPOR PRESSURE} \right) = 1.69$$

WATER IS LIQUID
#632

$$P_{PHC} = 19.7 - 1.69 = 18.01 \text{ PSIA}$$

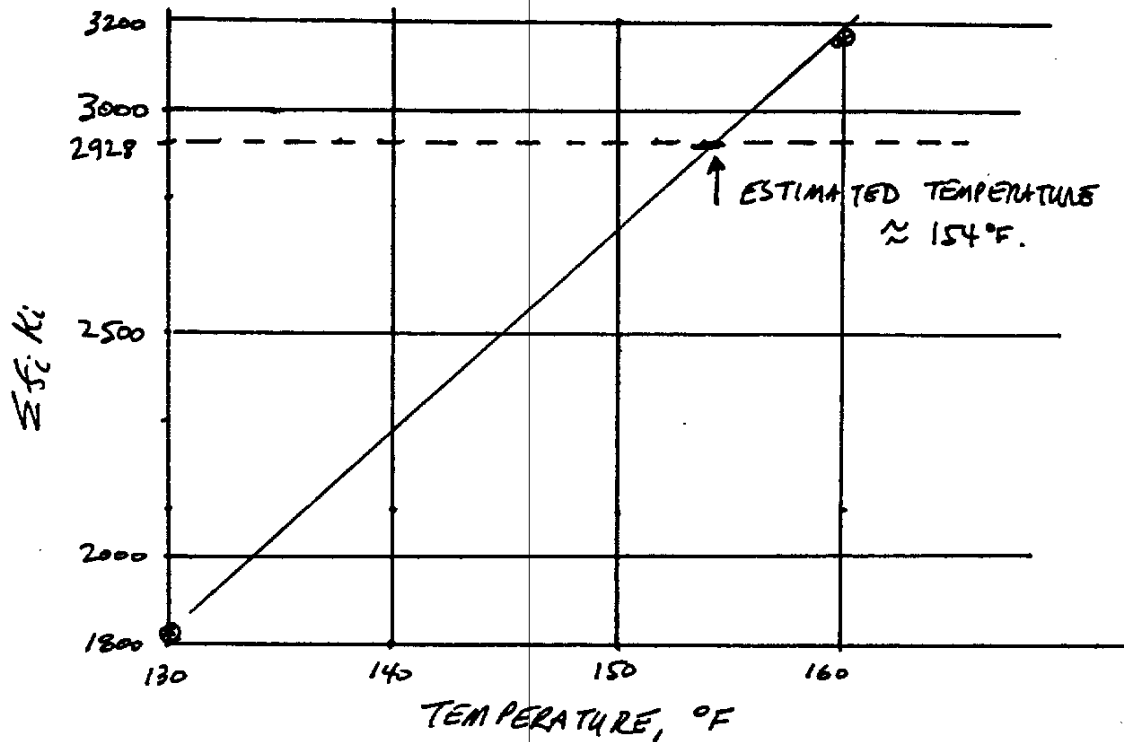
THEREFORE EQUILIBRIUM K VALUES SHOULD BE DETERMINED AT AN EFFECTIVE SYSTEM PRESSURE OF 18.01 PSIA

AT 130°F $P_{PHC} = 19.7 - 2.22 = 17.48 \text{ PSIA}$.

AT 160°F $P_{PHC} = 19.7 - 4.74 = 14.96 \text{ PSIA} \approx 14.7 \text{ PSIA}$

CRUDE UNIT DESIGN
CRUDE OVERHEAD SYSTEM.

EXTRAPOLATE TO ESTIMATE BUBBLE POINT TEMPERATURE.



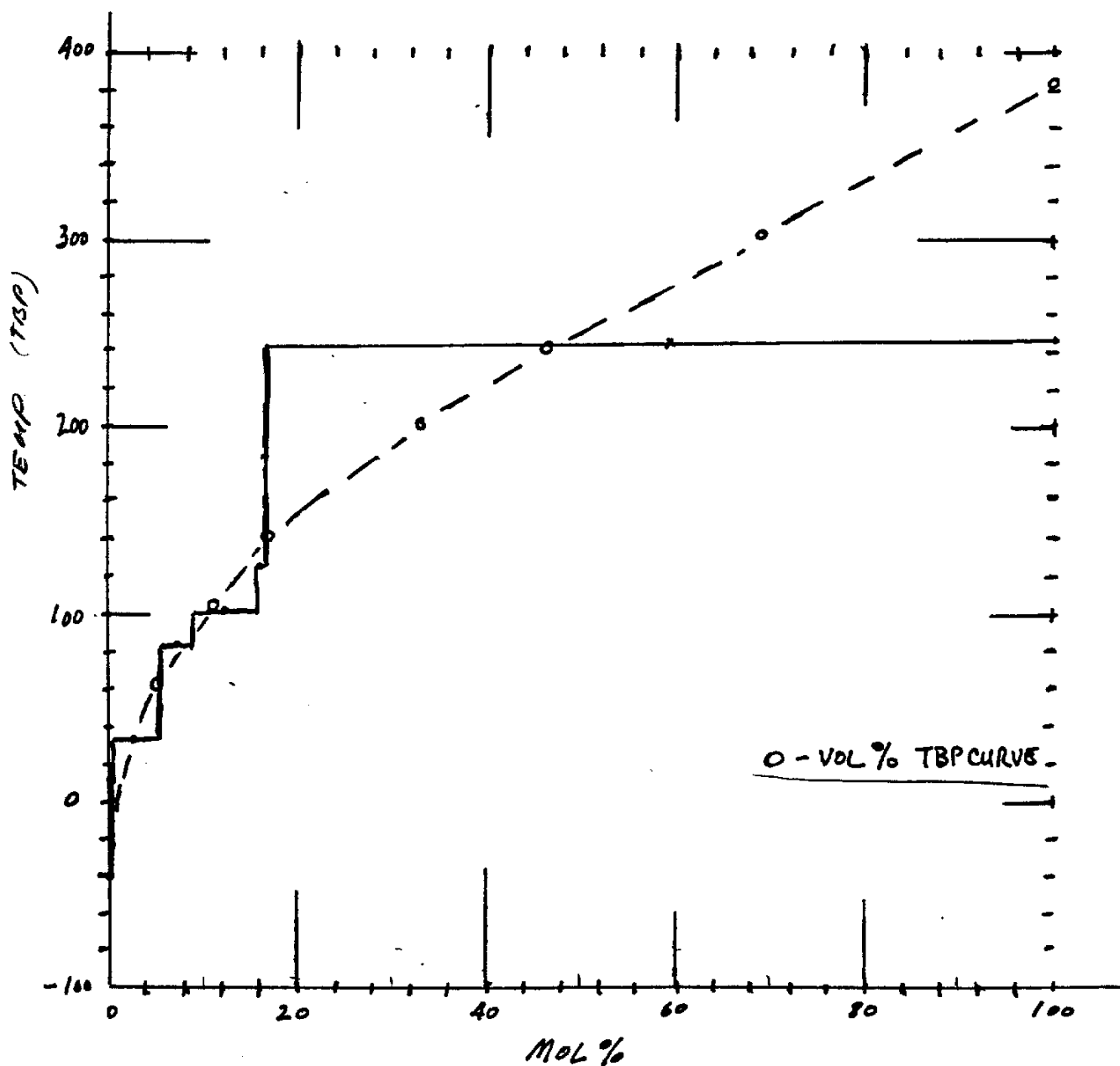
COMPUTER CALCULATION ON THIS BASIS GAVE A CALCULATED TEMPERATURE OF 152°F .

THIS VALUE IS HIGHER THAN WE WOULD HAVE EXPECTED.

WE WILL NOW EXPLORE BRIEFLY SENSITIVITY OF DEW POINT AND BUBBLE POINT TEMPERATURES RELATIVE TO THE CHARACTERIZATION OF THE HYDROCARBON MIXTURE.

CRUDE UNIT DESIGN CRUDE OVERHEAD SYSTEM.

IF WE PLOT ON A MOLAR BASIS THE NET CRUDE TOWER OVERHEAD MATERIAL AS DETERMINED FROM OUR CRUDE BREAKUP, WE OBTAIN THE FOLLOWING PLOT.



CRUDE UNIT DESIGN CRUDE OVERHEAD SYSTEM.

FOR COMPARISON, WE HAVE ALSO PLOTTED THAT PORTION OF THE TBP CURVE OF THE CRUDE ON A VOL% BASIS.

GENERALLY ONE EXPECTS THAT THE BUBBLE POINT OF A MIXTURE TO BE INFLUENCED MORE BY THE FRONT END OF THE DISTILLATION AND THE DEW POINT MORE BY THE BACK END OF THE DISTILLATION.

LET'S RECALCULATE THE CRUDE TOWER OVERHEAD BUBBLE POINT TEMPERATURE REPRESENTING THE NAPHTHA AS TWO PIECES - LSR NAPHTHA (LSR) AND HEAVY NAPHTHA (HVN) - AS WAS DONE IN THE CLASS PROBLEM.

TO DO THIS WE NEED TO DEVELOP PROPERTIES AND K VALUES FOR EACH OF THE CUTS.

FOR THE LSR FRACTION: $API = 70.6$ $UOP K = 12.35$ $MW = 95$
 $CABP \approx 188^\circ F$ $ENGLEER SLOPE \approx 0.9$

FROM B-1, B-2, B-3 CHARTS $VABP \approx 188 + 2 = 190^\circ F$ $MABP \approx 190 - 5 = 185^\circ F$
 $M_oABP \approx 190 - 9 \approx 181^\circ F$

FOR THE HVN FRACTION: $API = 53.3$ $UOP K = 11.97$ $MW = 125$
 $CABP \approx 302$ $ENGLEER SLOPE \approx 0.9375$

FROM B-1, B-2, B-3 CHARTS $VABP \approx 302 + 2 = 304^\circ F$ $MABP \approx 304 - 5 \approx 299^\circ F$
 $M_oABP \approx 304 - 9 \approx 295^\circ F$

USING TECHNIQUES DEMONSTRATED PREVIOUSLY FUGACITY COEFFICIENTS AND K'S AT VARIOUS TEMPERATURES COULD BE GENERATED.

CRUDE UNIT DESIGN CRUDE OVERHEAD SYSTEM

THE FOLLOWING TABLE PROVIDES IDEAL K_{TI} DATA FOR THE HVN FRACTION.

TEMPERATURE, °F	14.7 PSIA K _{TI}	50 PSIA K _{TI}
120	0.43	0.46
130	0.56	0.66
140	0.73	0.86
160	1.20	1.41
180	1.92	2.24

AS A SIMPLIFYING ASSUMPTION TO AID IN THE CALCULATIONS THE LSR K's CAN BE REPRESENTED BY THE COMPOUND 2,2 DIMETHYL PENTANE

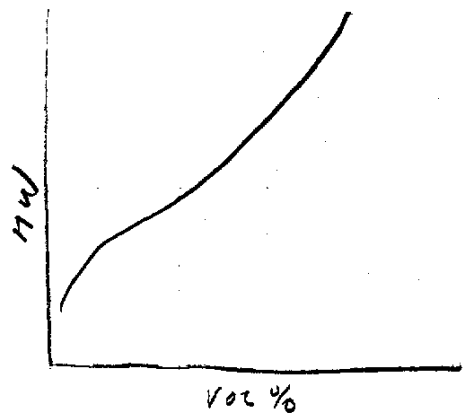
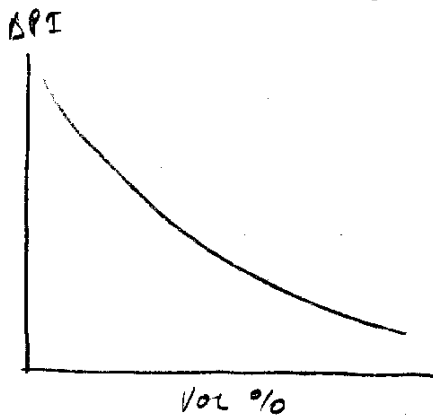
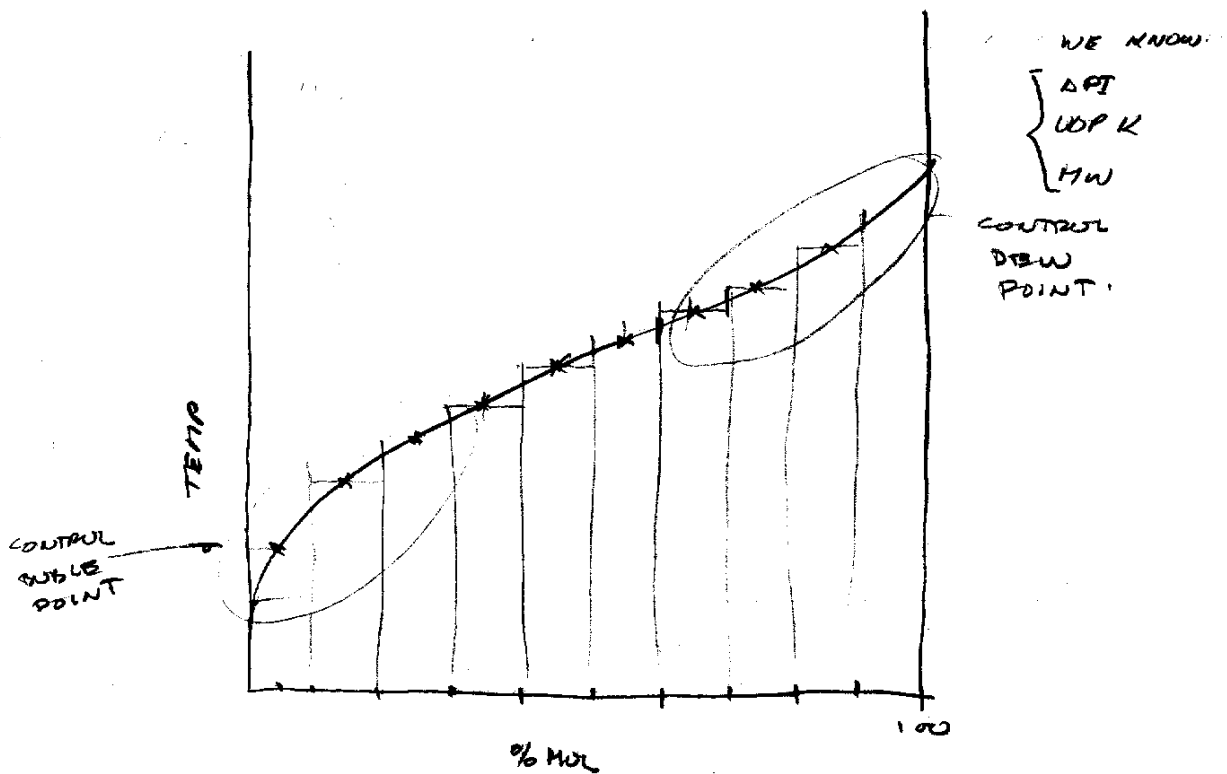
LIGHT STRAIGHT RUN ISOPHITHS

	130°F			150°F		
	f_i	K_i	$f_i K_i$	K_i	$f_i K_i$	
C3	28.56	11.28	322.16	14.72	420.40	
iC4	24.03	5.31	127.60	7.18	172.54	
NC4	155.49	4.08	634.40	5.62	873.85	
iC5	92.24	1.85	170.64	2.65	244.44	
NC5	187.62	1.47	275.80	2.16	405.26	
CP	10.63	0.993	10.55	1.50	15.93	
LSR	1164.63	0.38	442.56	0.61	710.42	
HVN	1264.68	0.032	40.47	0.061	77.15	
	2928.18		2024.18		2919.99	

AT 150°F VP H₂O = 3.72 PSIA $\gamma_p H_c = 19.7 - 3.72 = 15.98$ PSIA.

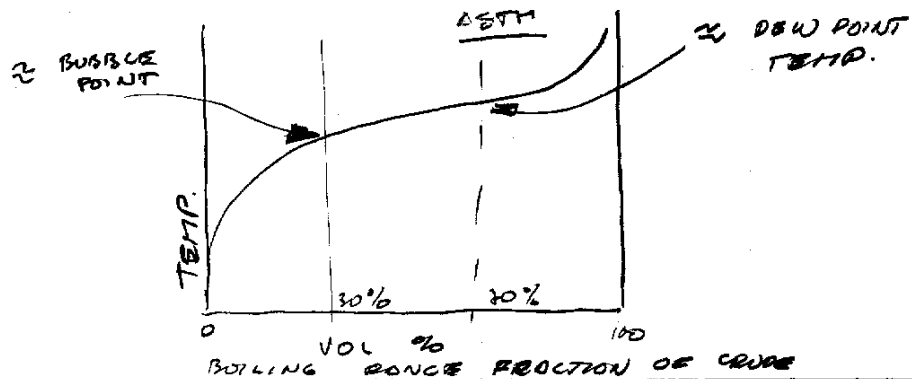
BUBBLE POINT TEMPERATURE $\approx 151^\circ\text{F}$

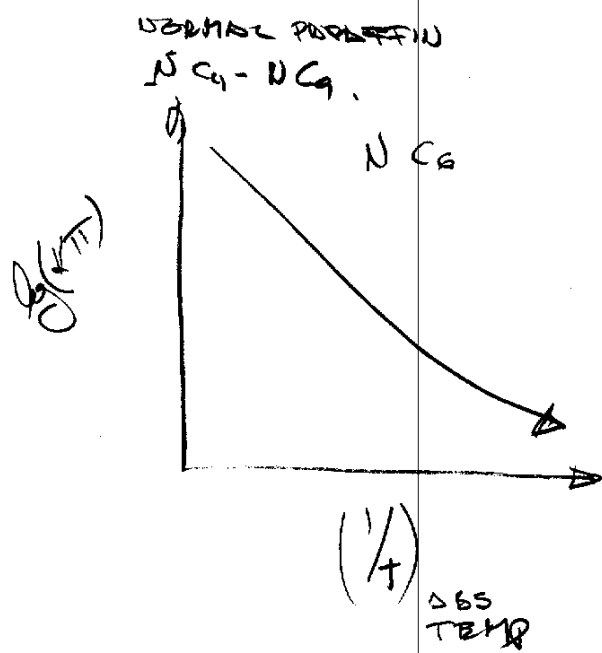
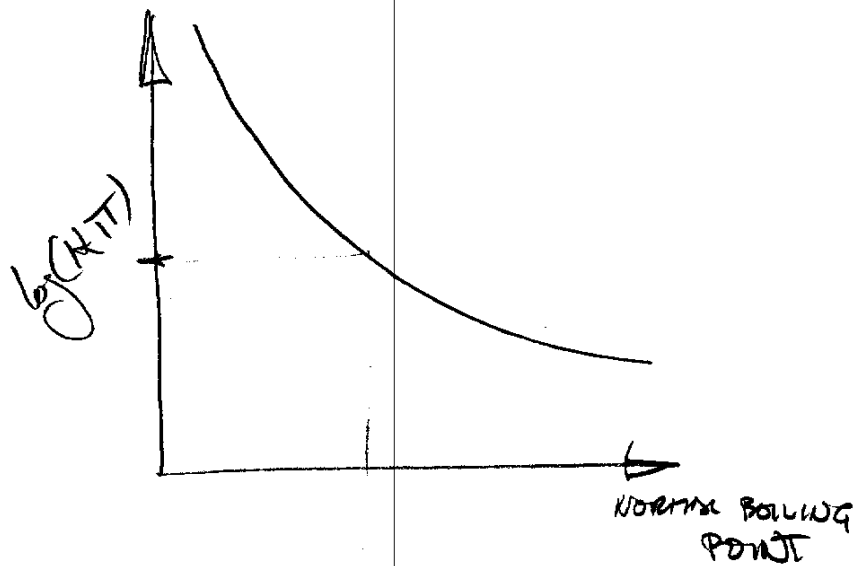
FOR THE PURPOSE OF THIS DISCUSSION FOR DESIGN WE WILL SET THE OPERATING TEMPERATURE AT 140°F.



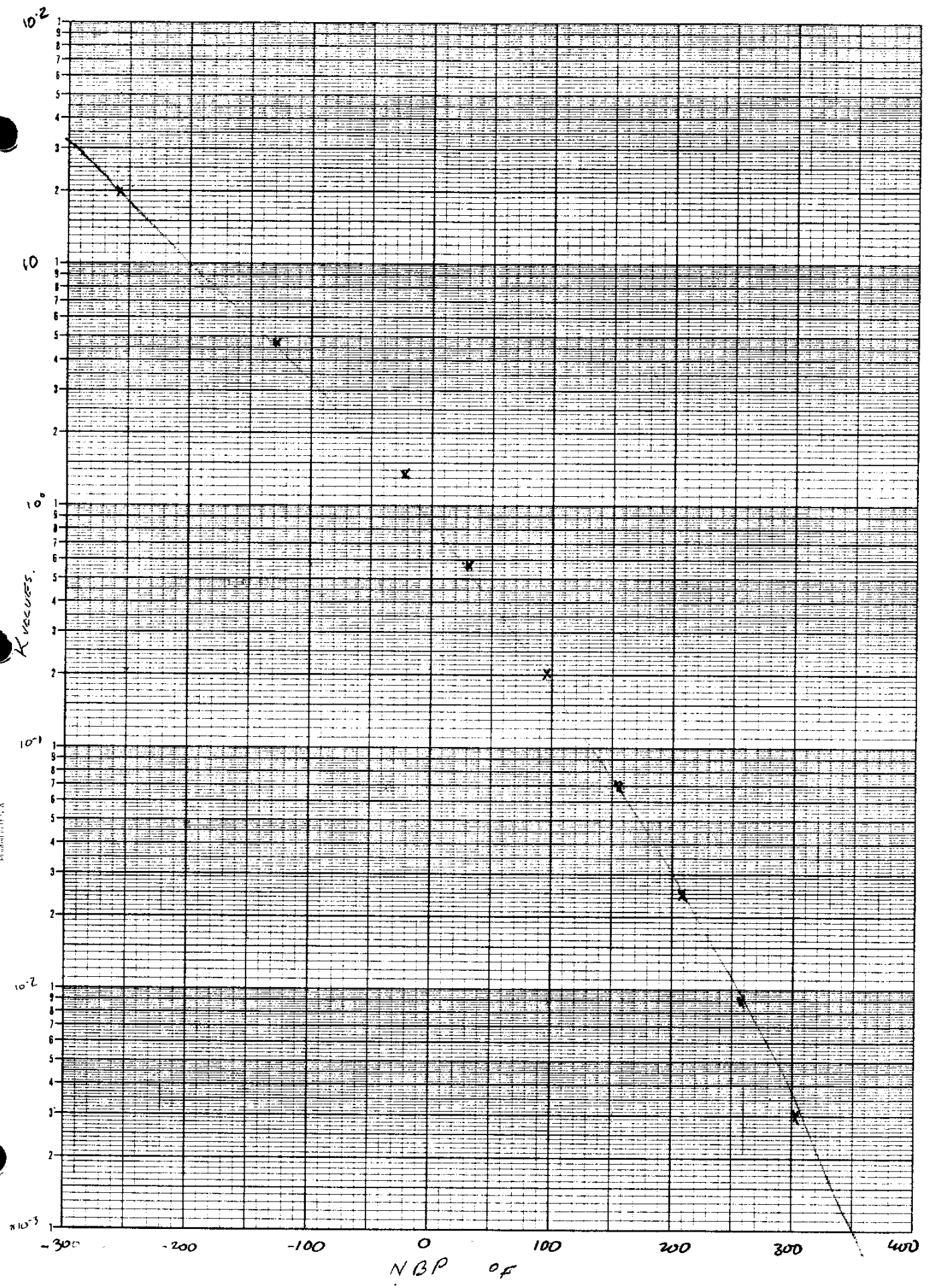
FROM CRUDE BREAKUP WE KNOW ΔPT & MW

EL BUBBLE POINT IS CONTROLLED FOR US FORM OF A CURVE FOR FRACTIONS INDICATED.





GRAPH EXPL (A) GRAPHIC CONTROL CONSTRUCTION
5 CYCLES X 70 DIVISIONS
SEMI LOGARITHMIC
AD 0845 -67

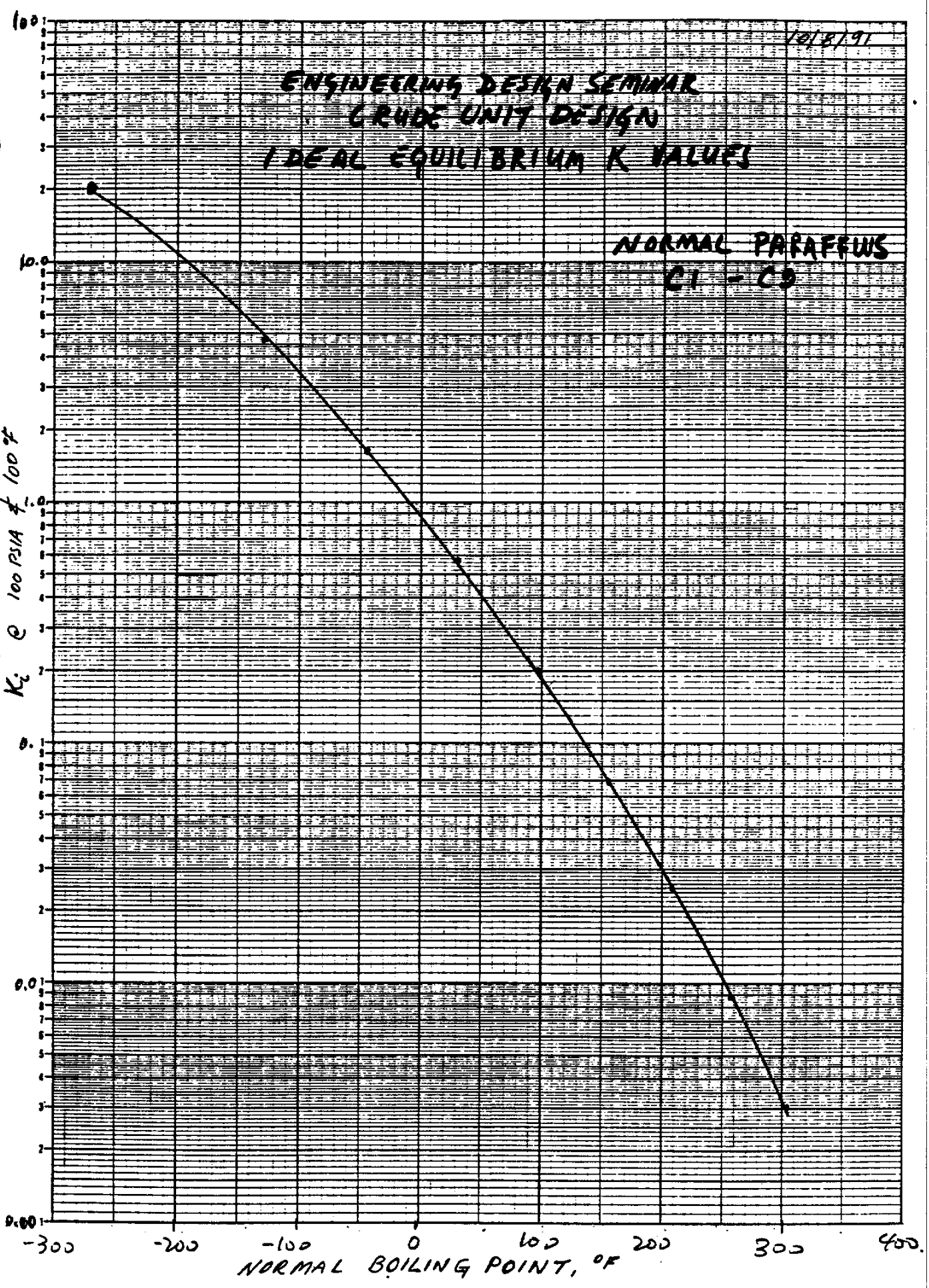


10/8/91

ENGINEERING DESIGN SEMINAR
GRADE UNIT DESIGN
IDEAL EQUILIBRIUM K VALUES

NORMAL PARAFFINS
C1 - C9

$K_i @ 100 \text{ PSIA } \pm 100^\circ \text{F}$



CRUDE UNIT DESIGN DEBUTANIZER CALCULATIONS

DESIGN BASIS REQUIRES NET LIQUID ONLY PRODUCT OVERHEAD FROM THE DEBUTANIZER (I.E. BUBBLE POINT RECEIVER). THE PROBLEM COULD HAVE BEEN SET UP TO YIELD A NET GAS ONLY PRODUCT. THE CALCULATIONAL APPROACH IS THE SAME.

1. COMPOSITION OF OVERHEAD KNOWN - FROM DEFINED SPLIT REQUIREMENTS IN COLUMN DETERMINE FROM MATERIAL BALANCE CALCULATIONS.

2. TEMPERATURE AND PRESSURE UNKNOWN.

a) PICK A TEMPERATURE AND CALCULATE PRESSURE.

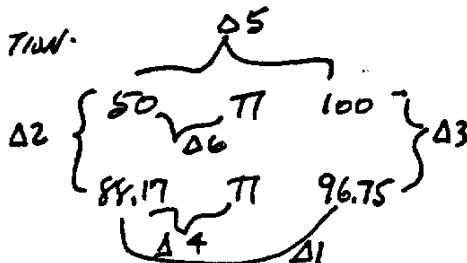
IF WE LOOK AT AIR ONLY IN THIS CONDENSING SERVICE ON COLUMN OVERHEAD, THEN RECEIVER TEMPERATURE APPROXIMATELY $95 + 25 = 120^{\circ}\text{F}$. FIND BUBBLE POINT PRESSURE OF NET OVERHEAD AT THIS TEMPERATURE.

	f_i	100 PSIA K π	$(f_i)(K\pi)$	50 PSIA K π	$(f_i)(K\pi)$
C3	28.56	199.84	5707.43	186.89	5337.58
LC4	24.03	97.815	2343.19	88.064	2116.18
NC4	128.27	74.523	9559.07	67.024	8597.17
iC5	1.81	34.829	63.04	30.287	54.82
	182.69		17672.82		16105.74
			CALC P = 96.75		CALC P = 88.17

LINEAR INTERPOLATION.

ASSUMED

CALCULATED



$$\Delta 6 = \frac{\Delta 5}{1 - \Delta 3/\Delta 2}$$

$$\Delta 4 = \frac{\Delta 1}{1 - \Delta 3/\Delta 2}$$

$$\pi = 50 + \Delta 6 = 50 + 50 / [1 - (-3.25/38.17)] = 96.08$$

$$\pi = 88.17 + \Delta 4 = 88.17 + 8.58 / [1 - (-3.25/38.17)] = 96.08$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN.

PAGE NO. _____
 DATE 10-1-91
 PROJECT _____
 BY E. N. MILLER

CRUDE UNIT DESIGN DEBUTANIZER CALCULATIONS.

TO DATE WE HAVE IGNORED POTENTIAL PRESENCE OF WATER IN THIS DEBUTANIZER OVERHEAD RECEIVER. SINCE STRIPPING STEAM WILL BE USED IN THE CRUDE TOWER, WE WOULD EXPECT THAT THE FEED TO THE DEBUTANIZER WOULD BE WATER SATURATED. FURTHERMORE WE WOULD EXPECT THE BULK OF THIS WATER TO GO OVERHEAD IN THE STABILIZER AND BE CONDENSED AS A SEPARATE PHASE IN THE RECEIVER.

LET'S TEST THIS ASSUMPTION:

FROM SOLUBILITY CHARTS (SEE HC WATER EQUILIBRIUM SECTION)
 ESTIMATED WATER CONTENT IS 0.002 MOL FRACTION.

$$\text{MOL/HR WATER IN FEED} = \left(\frac{2928.18}{0.998} \right) (0.002) = 5.87 \text{ MOL/HR}$$

ASSUME DEBUTANIZER OVERHEAD FOR SOLUBILITY PURPOSES TO BE N-BUTANE.
 AGAIN FROM SOLUBILITY CHARTS WATER IN NET OVHD $\approx$ 0.0007 MOL FRACTION

$$\text{MOL/HR WATER IN NET LIQUID} = \frac{(182.69)}{(0.9993)} (0.0007) = 0.13 \text{ MOL/HR.}$$

Q.E.D. TWO LIQUID PHASES WILL EXIST IN RECEIVER

$$\begin{aligned} \text{AT } 120^\circ\text{F} \quad P_{H_2O} &= 1.692 \text{ PSIA} \Rightarrow T_1 = P_{PHC} + P_{PH_2O} \\ &= 96.08 + 1.69 \\ &= 97.8 \text{ PSIA.} \end{aligned}$$

FOR CALCULATIONAL PURPOSES WE WILL SET RECEIVER PRESSURE AT 100 PSIA $\approx$ 85 PSIG.

OPERATING TEMPERATURE WILL BE 120°F

CRUDE UNIT DESIGN DEBUTANIZER CALCULATIONS

LET'S BRIEFLY LOOK AT DEBUTANIZER BOTTOMS

BASED ON PREVIOUS CALCULATIONS, THE OPERATING PRESSURE OF DEBUTANIZER HAS BEEN SET AT 100 PSIA. NOW DEVELOP PRELIMINARY HYDRAULICS FOR COLUMN AS FOLLOWS:

	PSIG	PSIA.
RECEIVER	85	100
ΔP CONDENSER	5 PSI	
COLUMN TOP	90	105
ΔP COLUMN	5 PSI	
COLUMN BOTTOM	95	110

REVIEW:

1. FRACTIONATION SPLIT SET BY PRODUCT REQUIREMENTS
(BOTTOMS COMPOSITION KNOWN)
2. COLUMN BOTTOMS PRESSURE SET BY SYSTEM HYDRAULICS
(RECEIVER PRESSURE SET IN PREVIOUS CALCULATION)
NO TRAYS, FEED TRAY, THERMAL COND OF FEED.
3. BOTTOMS MATERIAL AT BUBBLE POINT.
⇒ TEMPERATURE IS FIXED.

NOW CALCULATE BUBBLE POINT TEMPERATURE OF BOTTOMS.

	350°F	($Q_i/K_i \pi$)	400°F	
	K π		K π	$Q_i/K_i \pi$
MC4	27.20	442	555	15096
LC5	90.43	285	364	32916.5
NCS	187.62	263	343	64353.7
CP	10.93	220	300	3279.0
NAPH	2429.31	68.8	106.3	258285.7
	2745.49	256680.1		373880.9
		2333.5		3398.9

INTERPOLATE $T \approx 369^\circ\text{F}$

USE 365°F

CRUDE UNIT DESIGN. DEBUTANIZER CALCULATIONS.

ASSUME THIS COLUMN CAN BE REBOILED WITH A HOT STREAM FROM THE CRUDE TOWER. THIS WILL BE A THERMOSIPHON TYPE REBOILER. ASSUME 30 WT% VAPORIZATION ACROSS REBOILER. RETURN PRESSURE IS THAT OF THE TOWER. WE WOULD EXPECT 10 TO 15 °F TEMPERATURE RISE ACROSS THIS REBOILER.

TYPICAL VALUES
PRESSURE LOSS REBOILER
TYPICAL THERMOSIPHON 1

TO MAKE THIS CALCULATION INVOLVES TRIAL & ERROR TO DETERMINE TEMPERATURE ASSOCIATED WITH 30 WT% VAPORIZATION AT 110 PSIA. SOLUTION OF THIS PART OF THE PROBLEM IS LEFT WITH READER.

DEBUTANIZER - TRAYS AND REFLUX.

EXPERIENCE HAS INDICATED THAT IN THIS SERVICE ~30 REAL TRAYS AND 0.5-0.7 R/F (REFLUX/FEED) WILL BE SUFFICIENT TO MAKE THE SPECIFIED SPLIT. HOW DO WE GO ABOUT CONFIRMING THIS? THOSE WHO DOUBT CAN USE THEIR FAVORITE PROCESS SIMULATOR.

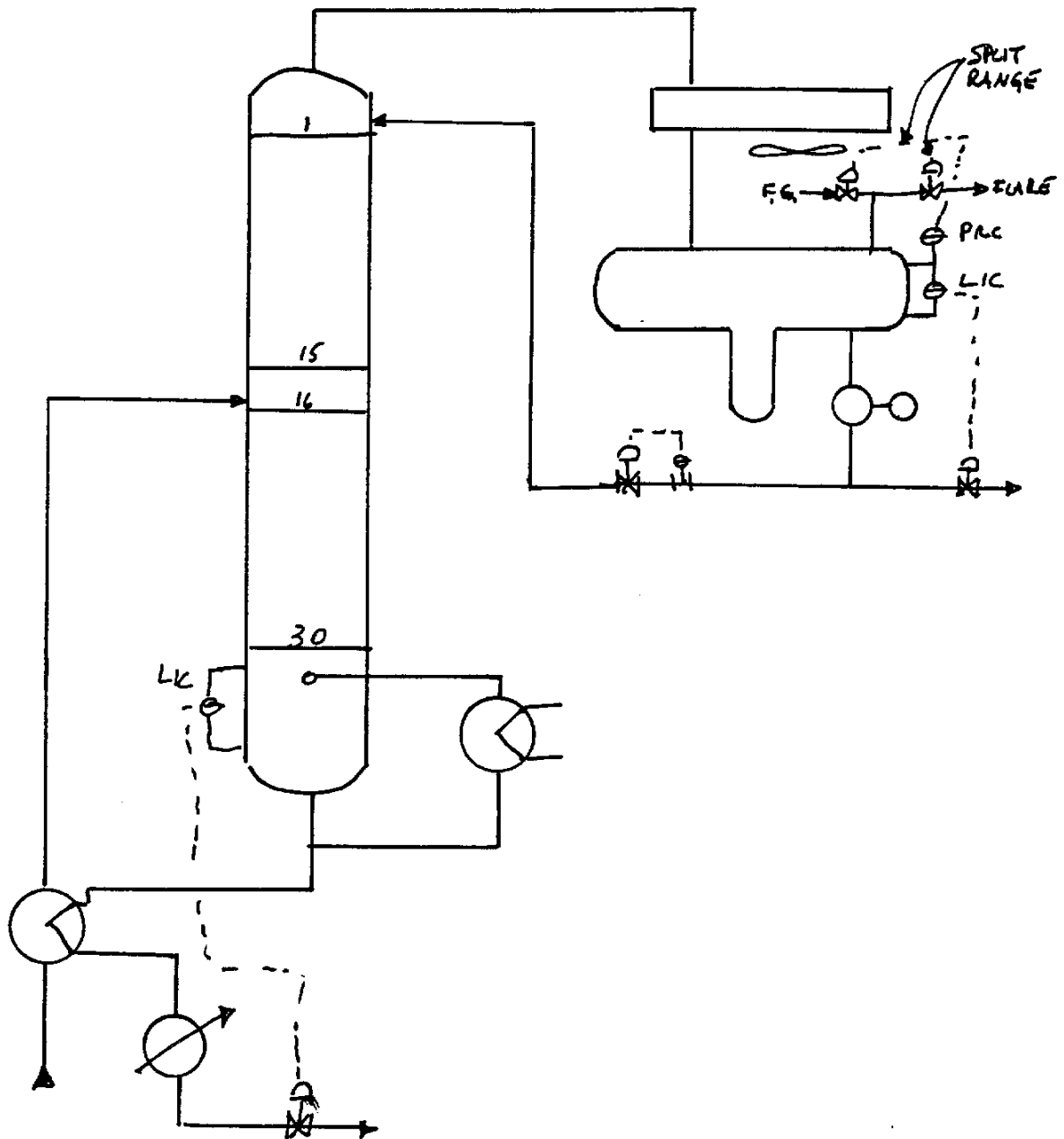
ASSUME:

- ① TRAY EFFICIENCY 70-80% $\Rightarrow$ 24 THEORETICAL TRAYS AT 80%
- ② ASSUME FEED APPROXIMATELY AT BUBBLE POINT (341)
- ③ FEED COLUMN IN ABOUT THE MIDDLE

FOR OUR DESIGN, USE A R/F = 0.5

$$\text{REFLUX RATE} = (2928.18)(0.5) = 1464 \text{ MOL/HR.}$$

CRUDE UNIT DESIGN.
DEBUTANIZER CALCULATIONS.



Hyprotech's Process Simulator HYSIM - Licensed to UOP
 Date 91/9/10 Version 386|C1.50 Case Name ENGRSEM1.SIM
 Time 7:48:22 Prop Pkg PR Column Name DEBUT

Eng. Design Seminar Debutanizer

**** Stage Variables ****

Reflux Ratio 6.41966 = R/D
 Reflux Subcooled to 120.00 F

Stg No	Press Psia	Temp F	Flow Rates (Lbmole/hr)		Duty MMBtu/hr
			Liquid	Vapour	
1	100.0	120.0	1171.3		182.5 L -11.280
2	105.0	138.5	1207.6	1353.7	
3	105.2	145.1	1204.5	1390.0	
4	105.4	149.7	1191.2	1386.9	
5	105.6	154.7	1168.7	1373.7	
6	105.8	161.3	1139.9	1351.1	
7	106.0	169.8	1111.9	1322.4	
8	106.3	179.4	1091.0	1294.3	
9	106.5	188.4	1076.3	1273.4	
10	106.7	196.3	1041.3	1258.8	
11	106.9	207.7	877.6	1223.7	
12	107.1	247.6	631.9	1060.0	
13	107.3	324.6	3484.5	814.3	2928.1
14	107.5	325.4	3501.4	738.9	
15	107.7	326.1	3514.6	755.7	
16	107.9	326.7	3526.0	769.0	
17	108.1	327.3	3536.9	780.4	
18	108.3	327.9	3548.4	791.2	
19	108.5	328.7	3561.5	802.7	
20	108.8	329.6	3577.2	815.9	
21	109.0	330.9	3596.7	831.6	
22	109.2	332.8	3621.5	851.1	
23	109.4	336.0	3655.2	875.9	
24	109.6	342.6	3710.2	909.6	
25	109.8	357.2	3828.8	964.6	
26	110.0	384.3		1083.2	2745.6 L 18.245

EXT REFLUX $R/F = \frac{1171.3}{2928.1} = 0.400$

No TRAYS

30 REFL * 0.80 = 24 THEORETICAL TRAYS.

Hyprotech's Process Simulator HYSIM - Licensed to UOP
 Date 91/9/10 Version 386|C1.50 Case Name ENGRSEM1.SIM
 Time 7:48:22 Prop Pkg PR Column Name DEBUT

Eng. Design Seminar Debutanizer

**** Products ****

Leaving from stage Product Phase Assigned to Stream Name	1 Liquid DEB_OVHD_LIQ	26 Liquid DEB_BOTTS
Temperature - F	120.00	384.31
Pressure - Psia	100.00	110.00
Propane - Mole Frac	0.156509	0.000002
i-Butane - Mole Frac	0.125813	0.000392
n-Butane - Mole Frac	0.707679	0.009605
i-Pentane - Mole Frac	0.008007	0.033063
n-Pentane - Mole Frac	0.001985	0.068202
Cyclopentane - Mole Frac	0.000008	0.003980
c6naph - Mole Frac	0.000000	0.884756
Total Flow - Lbmole/hr	182.45	2745.63

Hyprotech's Process Simulator HYSIM - Licensed to UOP
 Date 91/9/10 Version 386/C1.50 Case Name ENGRSEM1.SIM
 Time 7:58:42 Prop Pkg PR Column Name DEBUT

Engr. Design Seminar - Debutanizer Alternate (70%)

**** Stage Variables ****

Reflux Ratio 9.53401
 Reflux Subcooled to 120.00 F

Stg No	Press Psia	Temp F	Flow Rates		(Lbmole/hr)		Duty MMBtu/hr
			Liquid	Vapour	Feed	Draws	
1	100.0	120.0	1862.8			195.4 L	-17.198
2	105.0	138.7	1934.4	2058.2			
3	105.3	144.6	1935.8	2129.8			
4	105.5	148.4	1923.3	2131.1			
5	105.8	152.2	1895.5	2118.7			
6	106.1	157.5	1850.5	2090.8			
7	106.3	165.5	1778.6	2045.9			
8	106.6	178.4	1556.2	1974.0			
9	106.8	214.7	1080.0	1751.6			
10	107.1	305.4	3747.3	1275.4	2928.1		
11	107.4	307.1	3777.0	1014.6			
12	107.6	308.7	3806.0	1044.4			
13	107.9	310.6	3836.7	1073.3			
14	108.2	312.6	3870.2	1104.0			
15	108.4	314.9	3906.8	1137.5			
16	108.7	317.5	3945.9	1174.2			
17	108.9	320.9	3986.7	1213.2			
18	109.2	326.3	4031.9	1254.0			
19	109.5	336.6	4098.4	1299.2			
20	109.7	356.6	4249.4	1365.7			
21	110.0	387.1		1516.8		2732.7 L	24.594

Hyprotech's Process Simulator HYSIM - Licensed to UOP
 Date 91/9/10 Version 386|C1.50 Case Name ENGRSEM1.SIM
 Time 7:58:42 Prop Pkg PR Column Name DEBUT

Engr. Design Seminar - Debutanizer Alternate (70%)

**** Products ****

Leaving from stage Product Phase Assigned to Stream Name	1 Liquid DEB_OVHD_LIQ	21 Liquid DEB_BOTTS
Temperature - F	120.00	387.11
Pressure - Psia	100.00	110.00
Propane - Mole Frac	0.146165	0.000000
i-Butane - Mole Frac	0.121697	0.000092
n-Butane - Mole Frac	0.727139	0.004909
i-Pentane - Mole Frac	0.003705	0.033489
n-Pentane - Mole Frac	0.001286	0.068566
Cyclopentane - Mole Frac	0.000008	0.003999
c6naph - Mole Frac	0.000000	0.888944
Total Flow - Lbmole/hr	195.39	2732.69

EL USARLO DEPENDE DE LAS ESPECIFICACIONES DE LOS PRODUCTOS.

ES

OPER SPEC:

1% MAX C5⁺ IN OVHD

1% MAX C4⁻ IN BTMS.

OR

0.05% C5⁺ IN OVHD

0.05% C4⁻ IN BTMS.

OR QUE LAS ESPECIFICACIONES REQUERIDAS VARIEN DENTRO

DE UN RANGO I.E. 1% MAX C5⁺ - 0.05% MAX C5⁺ BTMS

Hyprotech's Process Simulator HYSIM - Licensed to UOP
 Date 91/10/9 Version 386|C1.50 Case Name ENGRSEM1.SIM
 Time 5:44:55 Prop Pkg PR Column Name DEBUT

Engr. Design Seminar - Debutanizer Alternate (70%)

**** Stage Variables ****

Reflux Ratio 7.02596
 Reflux Subcooled to 120.00 F

Stg No	Press Psia	Temp F	Flow Rates (Lbmole/hr)				Duty MMBtu/hr
			Liquid	Vapour	Feed	Draws	
1	100.0	120.0	1281.8			182.4 L	-12.200
2	105.0	138.6	1320.9	1464.2			
3	105.2	145.3	1316.8	1503.4			
4	105.5	150.1	1300.7	1499.2			
5	105.7	155.5	1273.4	1483.1			
6	106.0	162.8	1239.3	1455.9			
7	106.2	172.2	1205.1	1421.7			
8	106.4	182.9	1157.1	1387.6			
9	106.7	197.2	981.7	1339.6			
10	106.9	238.8	696.9	1164.2			
11	107.1	321.1	3519.3	879.3	2928.1		
12	107.4	322.0	3537.5	773.7			
13	107.6	322.8	3552.4	791.8			
14	107.9	323.5	3566.1	806.8			
15	108.1	324.3	3580.0	820.5			
16	108.3	325.2	3595.6	834.4			
17	108.6	326.3	3613.9	849.9			
18	108.8	327.8	3636.3	868.3			
19	109.0	329.9	3664.1	890.6			
20	109.3	333.4	3701.2	918.5			
21	109.5	340.7	3760.0	955.5			
22	109.8	356.3	3885.4	1014.3			
23	110.0	384.3		1139.8		2745.6 L	19.174

Hyprotech's Process Simulator HYSIM - Licensed to UOP
 Date 91/10/9 Version 386|C1.50 Case Name ENGRSEM1.SIM
 Time 5:44:55 Prop Pkg PR Column Name DEBUT

Engr. Design Seminar - Debutanizer Alternate (70%)

**** Products ****

Leaving from stage Product Phase Assigned to Stream Name	1 Liquid DEB_OVHD_LIQ	23 Liquid DEB_BOTTS
Temperature - F	120.00	384.30
Pressure - Psia	100.00	110.00
Propane - Mole Frac	0.156522	0.000002
i-Butane - Mole Frac	0.126354	0.000356
n-Butane - Mole Frac	0.707130	0.009647
i-Pentane - Mole Frac	0.007472	0.033099
n-Pentane - Mole Frac	0.002507	0.068167
Cyclopentane - Mole Frac	0.000014	0.003980
c6naph - Mole Frac	0.000000	0.884749
Total Flow - Lbmole/hr	182.43	2745.65



25 East Algonquin Road
Des Plaines, Illinois 60017-5017

HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR

CRUDE UNIT DESIGN

DEBUTANIZER SECTION

PAGE NO. _____

DATE 10-1-91

PROJECT _____

ENGINEERING DEPARTMENT

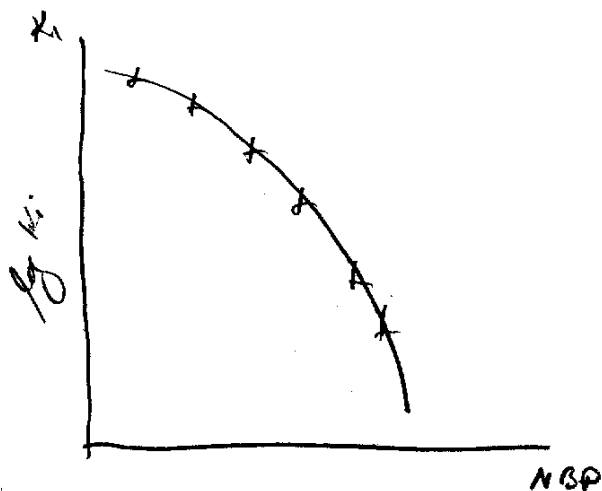
BY _____

NOTE - These data are confidential and the property of UOP, Inc. and shall not be disclosed to others or reproduced in any manner or used for any purpose whatsoever except by written permission or as provided in a signed agreement with UOP, Inc. relating to such data.

STREAM	°API (LB/GAL)	K	MW	FLOWING SP. GR. (LB/CF)	°F	PSIG (MM HG ABS)	BPSD @ 60°F (10 ⁶ SCF/D)	FLOWING GPM (CFS)	MOLS PER HR.	LB/HR	BTU/ LB	10 ⁶ BTU/HR
CRUDE COLUMN OVER TO DECY BTMS EXCH									2926.14	301375		
EXCHANGER DUTY												
FEED TO DEBUTANIZER					330	92			2928.14	301375	161	48.5
VAPORS FR DEBUTANIZER OVERHEAD			56.1		138	90			1646.71	92355	184	17.0
CONDENSER DUTY												13.7
MATERIAL FROM CONDENSER			56.1		120	85			1646.71	92355	36	3.3
REFLUX TO COLUMN			56.1		"				1464	82115	"	2.9
NET LPG TO RECOVERY			56.1		"				18269	10240	"	0.4
LIQUID TO REBOILER					365	95				472905	184	27.0
REBOILER DUTY												19.2
MATERIAL FROM REBOILER					380	95				472905	224	106.2
VAPOR										141870	296	42
LIQUID										331035	194	64.2
DEBUTANIZER BOTTOMS TO EXCHANGE	63.0		106		365	95	2.7422		2745.5	291135	184	53.6
EXCHANGER DUTY												
DEBUTANIZER BOTTOMS TO COOLER	63.0		106			85	2.7422		"	291135		
COOLER DUTY												
DEBUTANIZER BOTTOMS TO STORAGE	63.0		106		100	60	2.7422		"	291135		

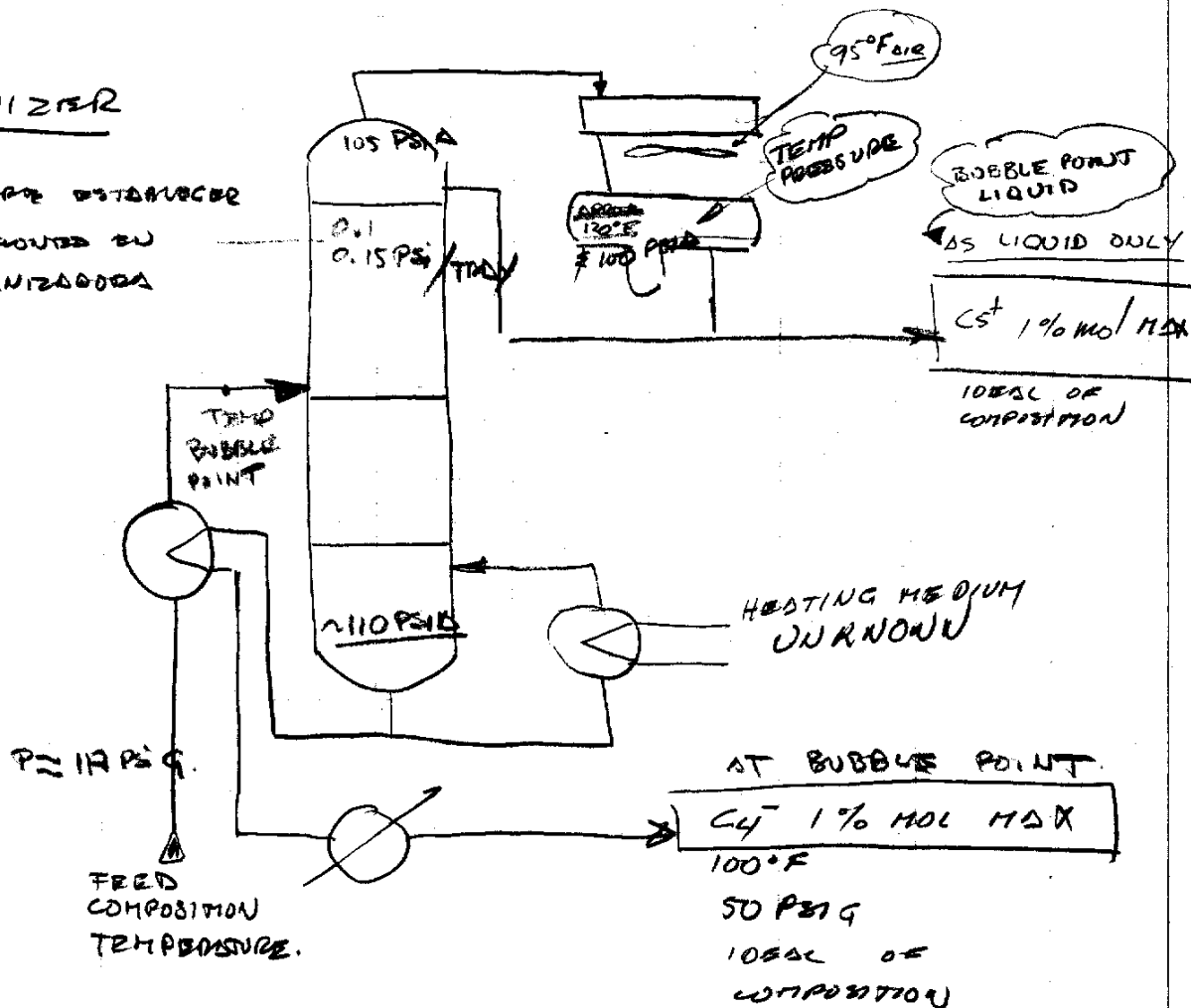
AT 100 PSIA
± 100 °F

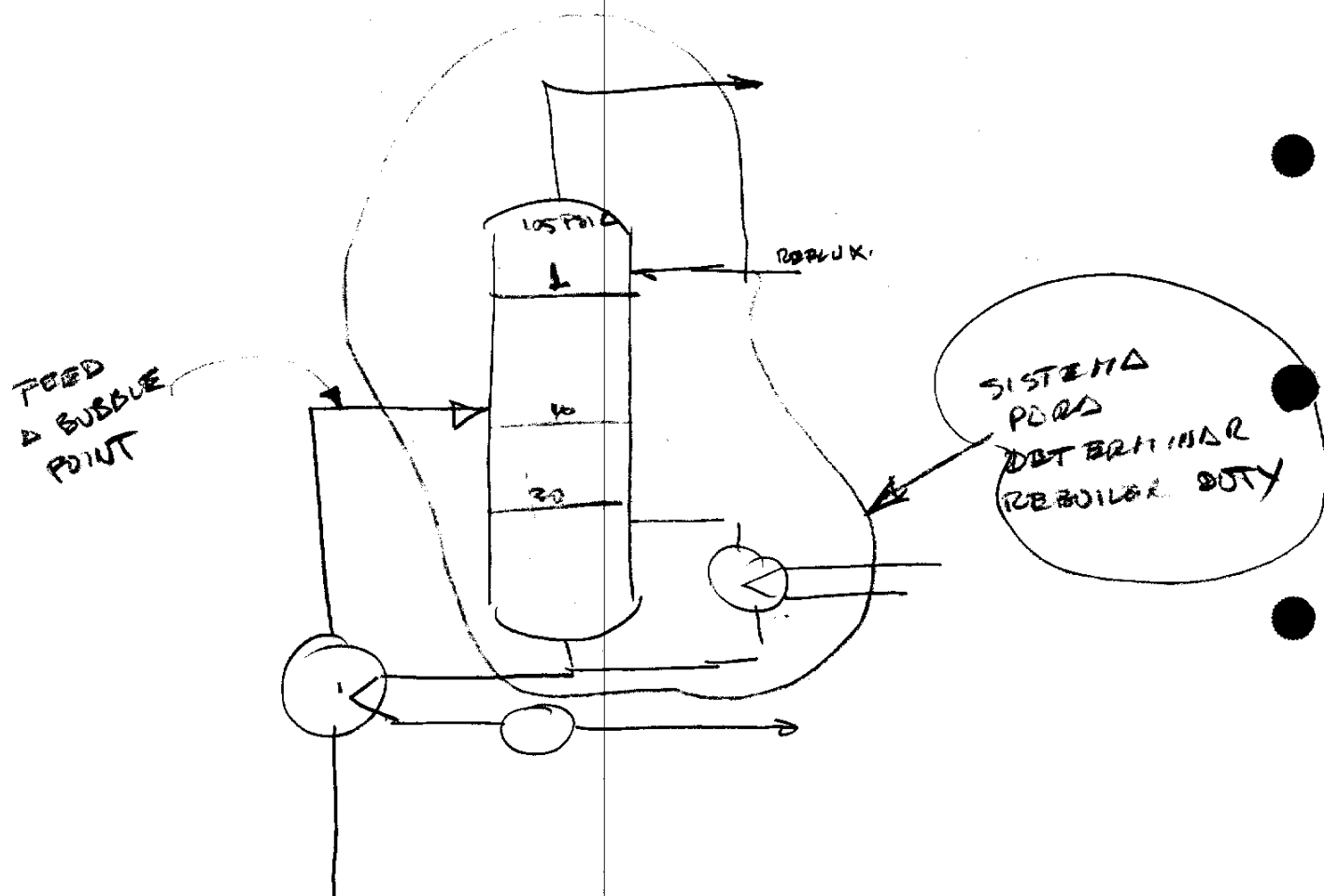
	K_i	NBP °F
C ₁	22.049	-259
C ₂	4.76	-128
C ₃	1.65	-44
NC ₄	0.57	31
NC ₅	0.198	97
NC ₆	0.07	156
NC ₇	0.025	209
NC ₈	0.009	258
UC ₉	0.003	303



DEBUTANIZER

52 REQUIRING DISTILLATION
VES SUBJECTED TO
10 DEBUTANIZATION





DEBUTANIZADOR

PARA SISTEMAS CIMA

- * STIMULO TEMP EN EL ACUMULADOR DE CIMA, SE DEBE DETERMINAR LA PRESION DE OPERACION DEL ACUMULADOR DE CIMA POR PROBLEMAS Y ERROR
CONSIDERANDO SISTEMAS DE CONDENSACION CON CIRQ. ($T \approx 120^\circ\text{C}$)

PARA SISTEMAS FONDOS ESTABLECIDA LA CAIDA DE PRESION A TRAYES DE LA DEBUTANIZADORA Y CONOCIENDO LA COMPOSICION DE FONDOS, SE PUEDE DETERMINAR TEMP DE BURBUJAS.

NECESARIO PARA ESTO

- ① NO DE PLUGS.
- ② REFLUJO REQUERIDO
- ③ UBICACION DE ALIMENTACION
- ④ CONDICION TERMICA DE LA ALIMENTACION
- ⑤ CONDENSER DUTY
- ⑥ REBOILER DUTY & BOUNCE ENERGIA
- ⑦ WASTE HEAT RECOVERY

DEFINIR DE ACUERDO A EXPERIENCIA
UDP

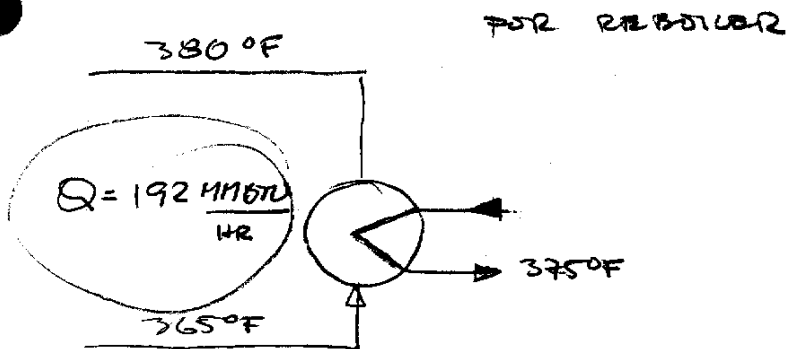
~ 30 REAL TRAYS

$$R/F \approx 0.5 \text{ (REFLUX/FEED)}$$

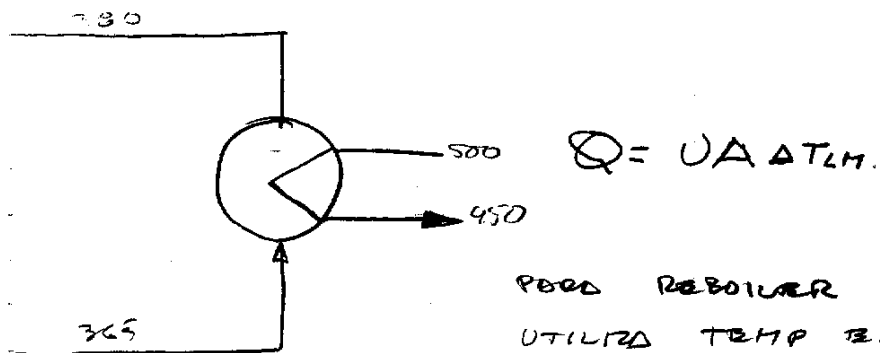
EXTERNAL REFLUX.

FEED TRAY IS MIDDLE

FEED AT BUBBLE POINT



165 PSIA SATURATED STEAM
 ~ 375 °F



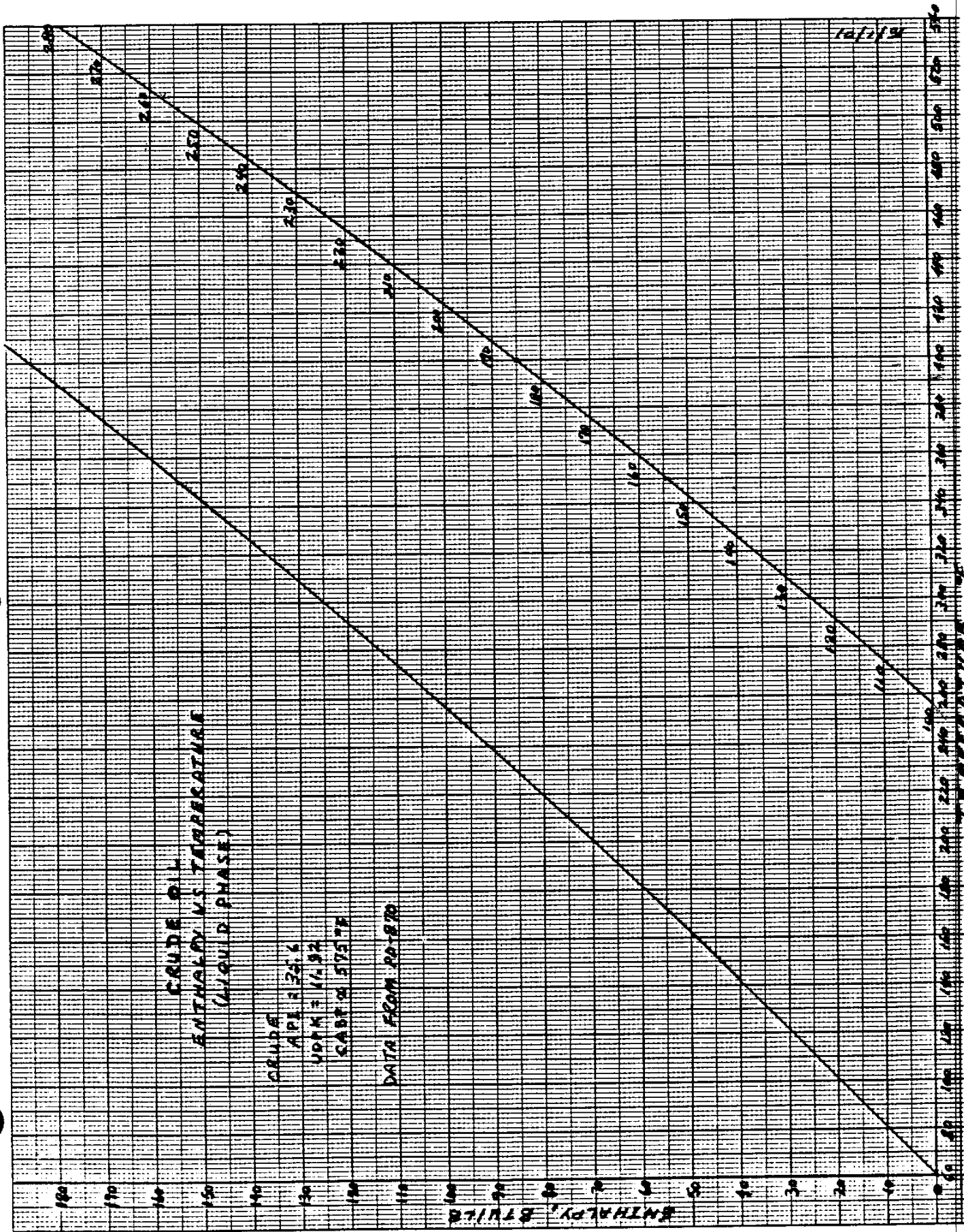
PARA REBOILER CALCULO DE ΔT_{LM} .
 UTILIZA TEMP ENTRADA = TEMP SAÍDA

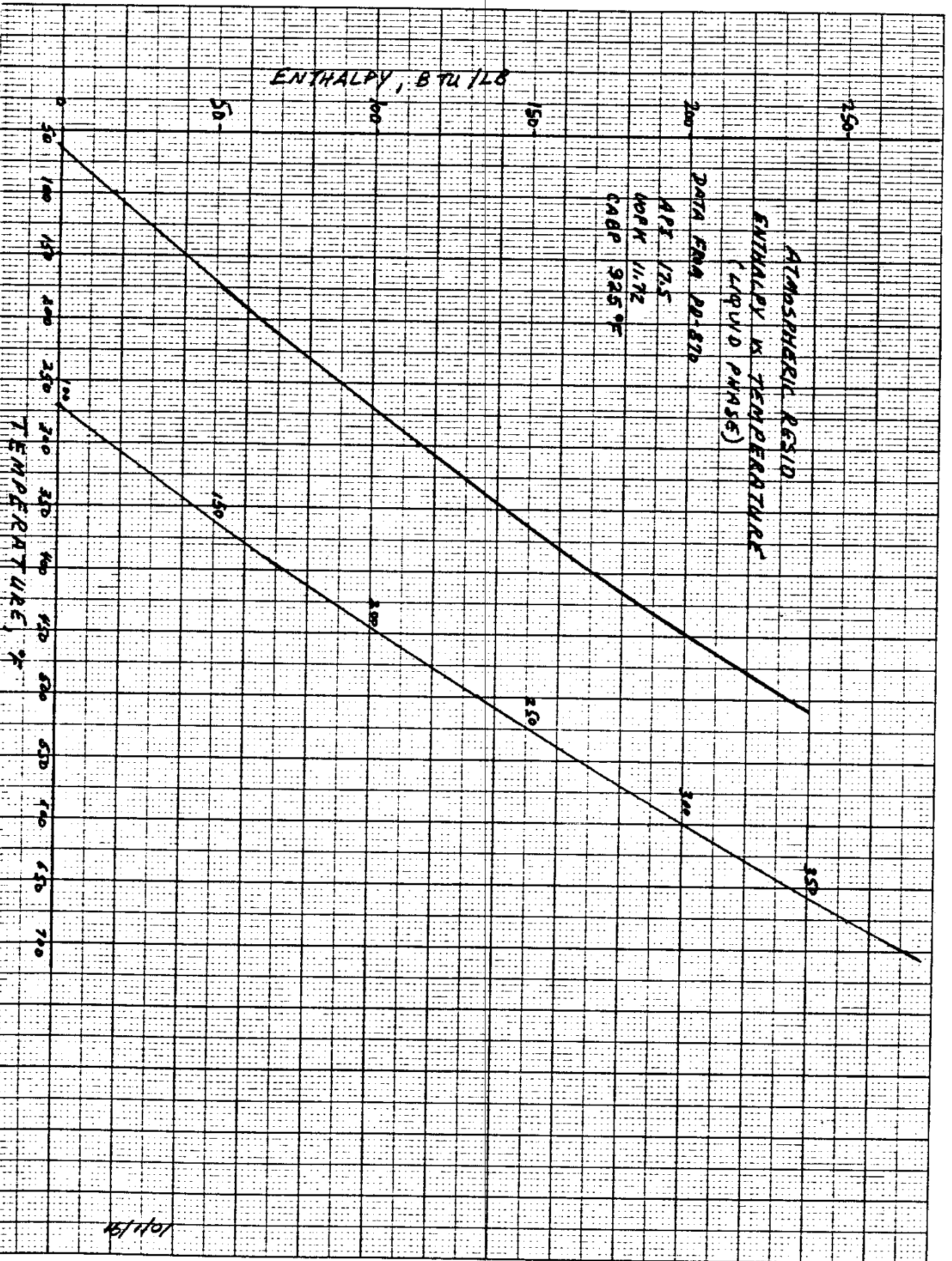
500 → 450 PARA ΔT_{LM} DE REBOILER
 380 ← 380

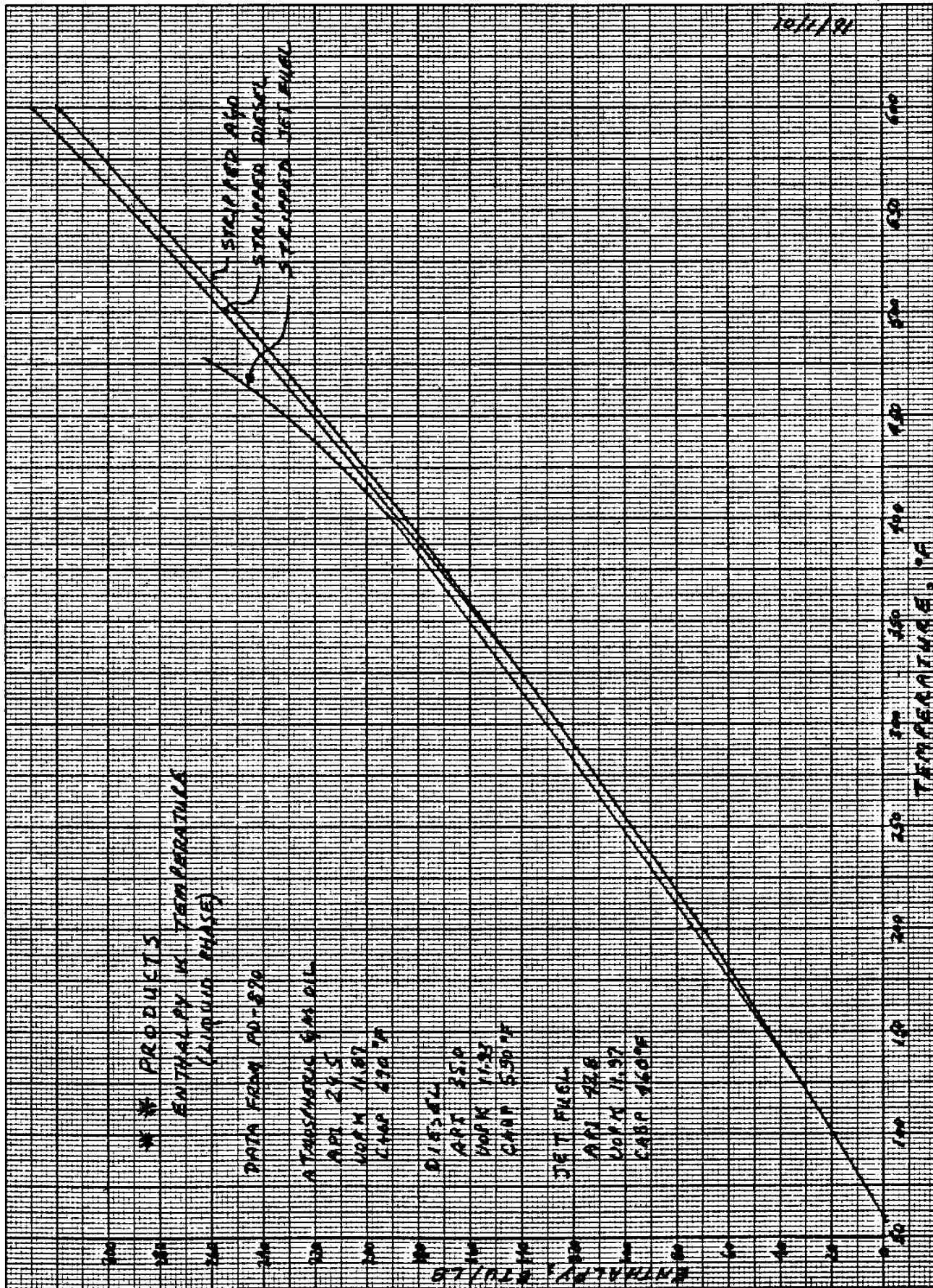
CRUDE OIL ENTHALPY VS. TEMPERATURE (LIQUID PHASE)

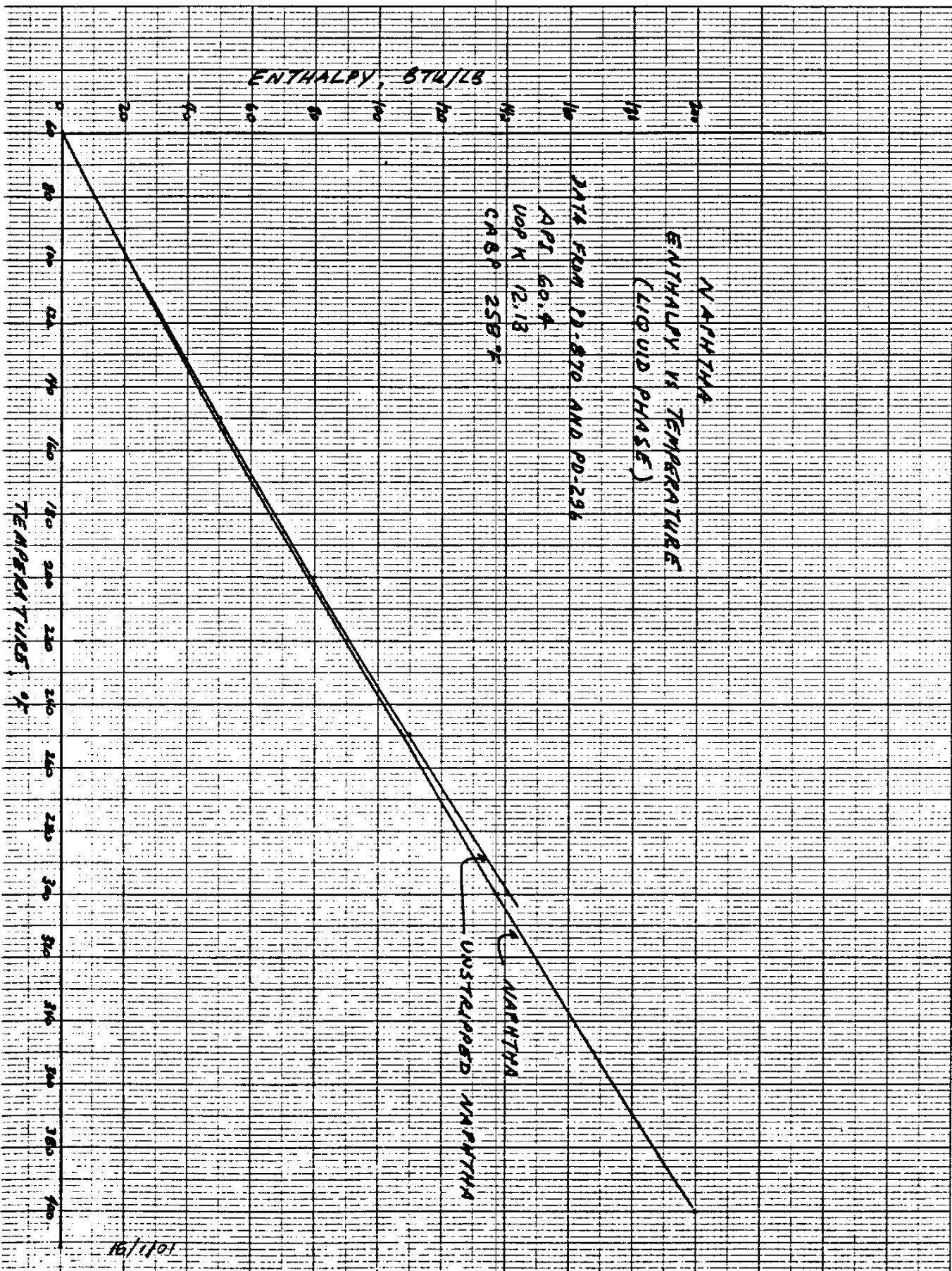
CRUDE
API 235.6
DDM = 11.92
CABP = 5757F

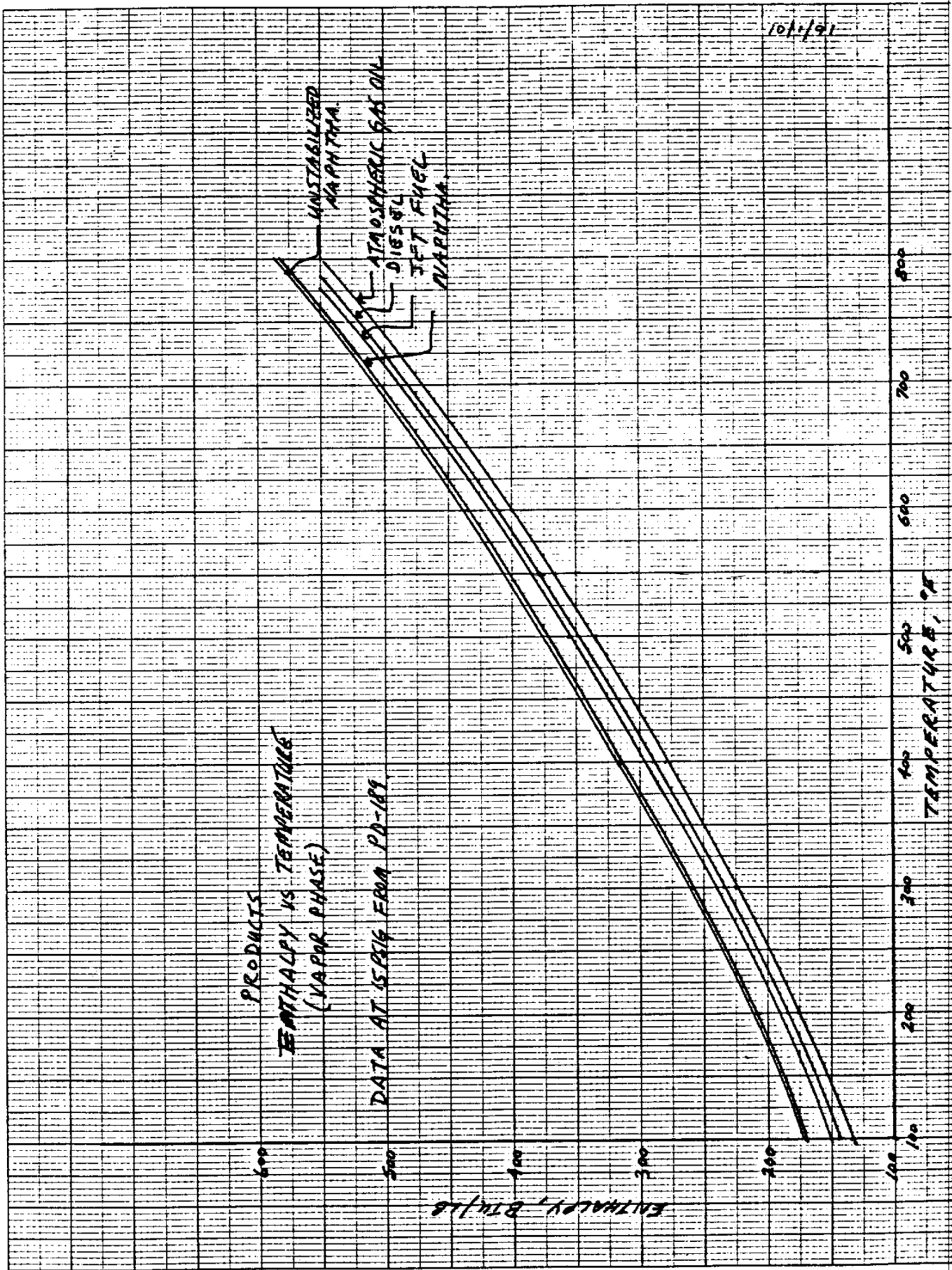
DATA FROM 20870





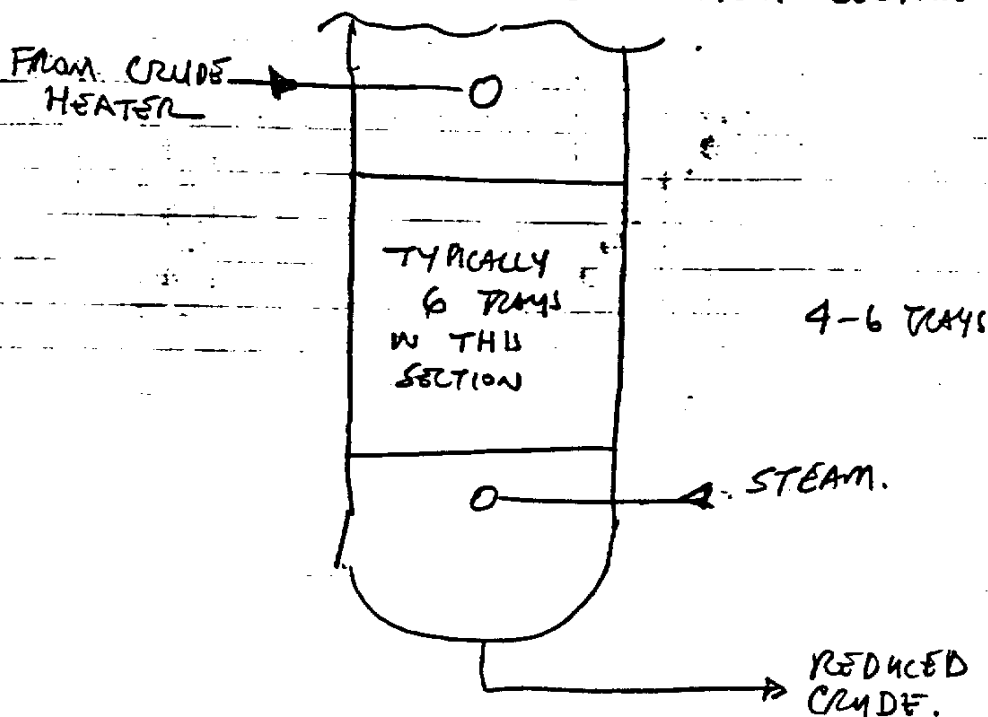






CRUDE TOWER FEED RICHNESS

CRUDE COLUMN BOTTOM SECTION



IN THE STRIPPING SECTION OF THE CRUDE TOWER BELOW THE FLASH ZONE, WE WILL PUT IN ~~6~~ REAL TRAYS. AS WITH THE SIDECUT STRIPPERS, STRIPPING STEAM WILL BE USED TO MINIMIZE LOSS OF GAS OIL PRODUCT TO THE REDUCED CRUDE STREAM.

STEAM RATE TO BOTTOM SECTION FOR DESIGN PURPOSES WILL BE SET AT 10 LBS/BDL OF COLUMN BOTTOMS.

EXPERIENCE INDICATES THAT COLUMN BOTTOMS TEMPERATURE IS 20-40 °F LOWER THAN THE FEED TEMPERATURE TO THE FLASH ZONE. FOR OUR CURRENT CLASS DESIGN PROBLEM WE WILL ASSUME THAT THE BOTTOMS DELTA TEMPERATURE IS 40 °F.

AS YOU RECALL FROM OUR FLASH CALCULATIONS, WE SET THE FEED TEMPERATURE TO THE FLASH ZONE AT 715 °F. THEN COLUMN BOTTOMS TEMPERATURE WILL BE 675 °F

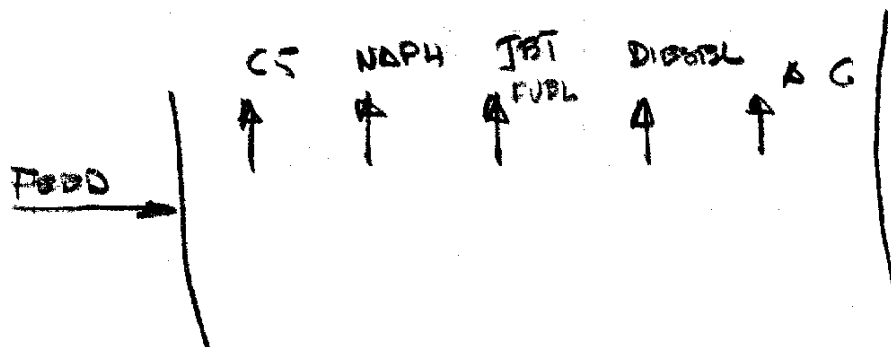
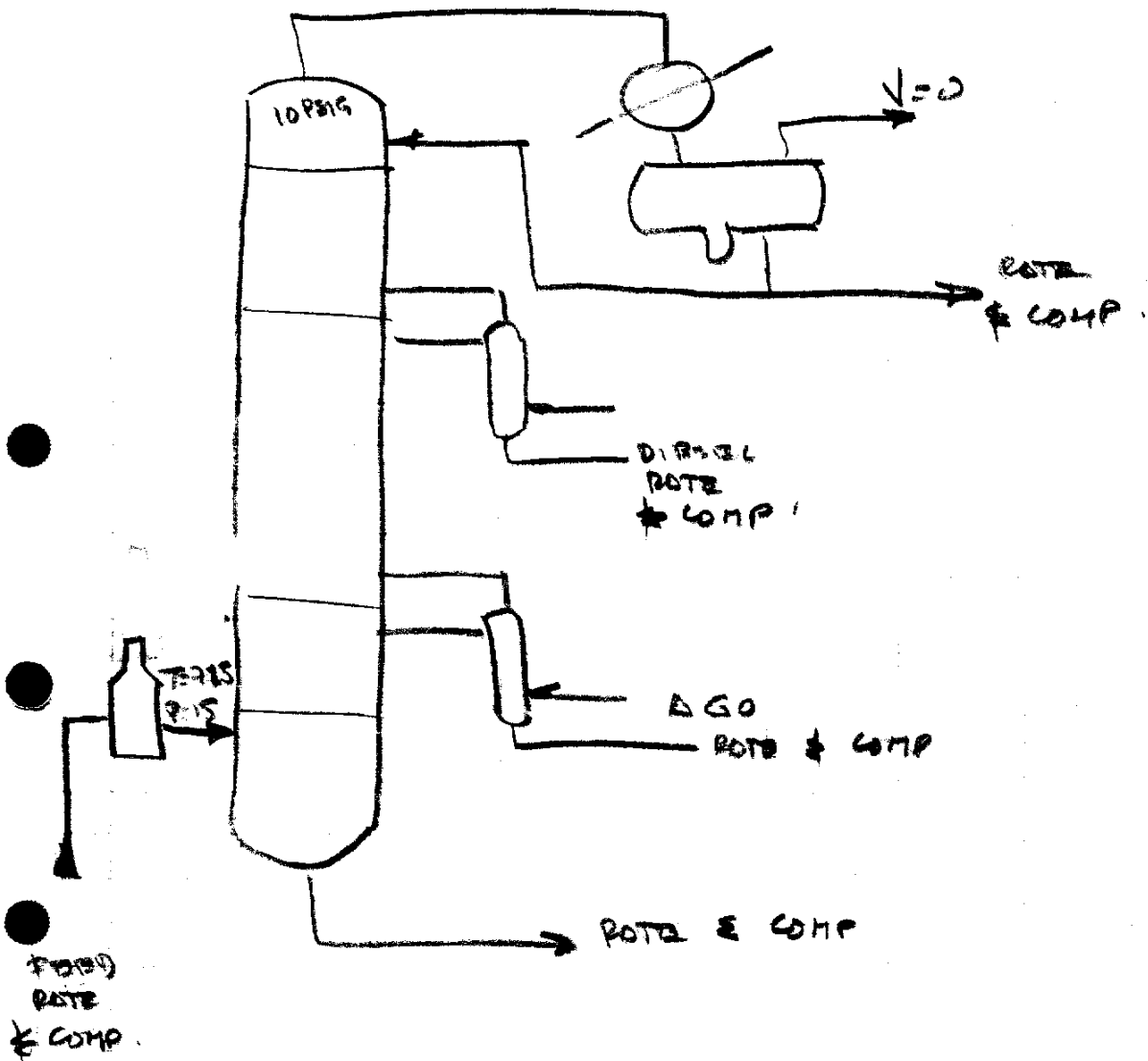
CUSTOMER ENGINEERING DESIGN SEANNAHSUBJECT CAUDE UNIT DESIGNCAUDE TOWER HEAT BALANCEPAGE NO. 12DATE 10-5-91PROJECT EN MILLERBY EN MILLER

NOW CALCULATE STRIPPING STEAM RATE TO THE
BOTTOM OF THE TOWER.

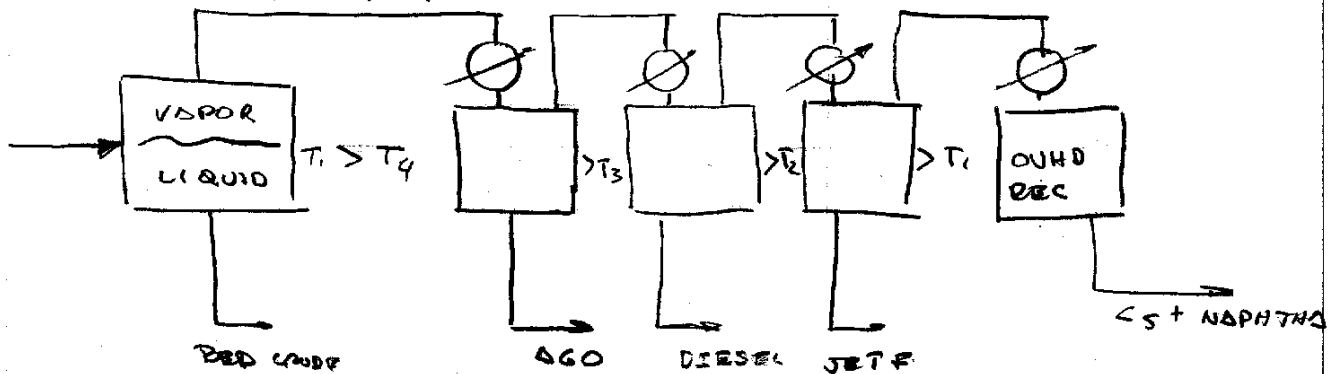
FROM CAUDE BREAKUP: RESID RATE IS 38100 BPD.

$$\text{STEAM TO BOTTOM OF THE TOWER} = \frac{(10)(38100)}{24} = 15875 \text{ LB}$$

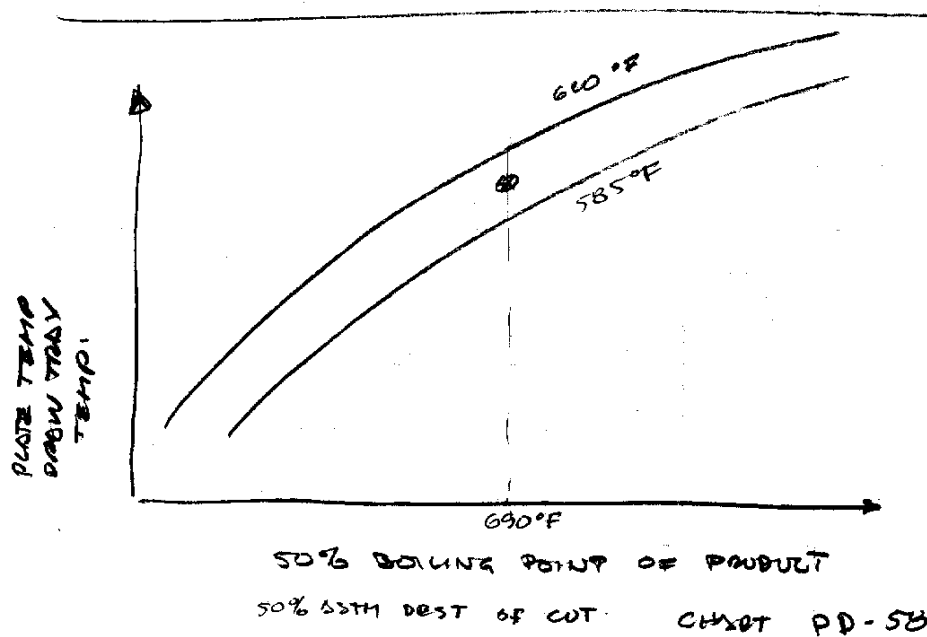
$$\equiv 881.9 \text{ MOL/HR.}$$



EL DISEÑO DE LA PROCESADORA CONSIDERA, VARIOS ETAPOS FLASH DE ACUERDO AL N° DE CORTE, SEGUN EL SIGUIENTE CONCEPTO.



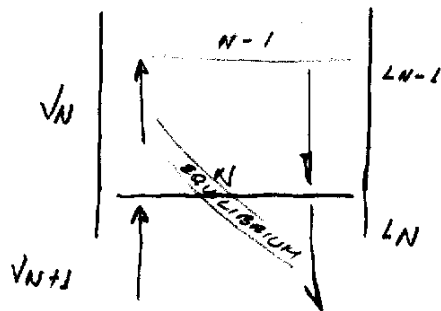
REQUIERIENDOSE CALCULAR BOUNCES DE HOT Y ENERGIA RED CON ETAPOS Y DETERMINAR LA TEMP DE SUDO DE CADA CORTE. (DRAW TDOY TEMPERATURES.)



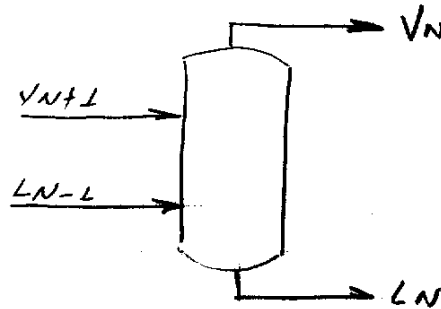
CALCULATION MODEL OF SYSTEM.

① "RIGOROUS" TRAY TO TRAY CALCULATIONS
PROCESS SIMULATORS.

EQUILIBRIUM STAGE CALCULATION



CONSIDERAR EL SIGUIENTE SIST. EQUILIBRIO



NECESARIO

1° MATERIAL BALANCE EQUATION

COMPONENT BALANCE
OVERALL BALANCE

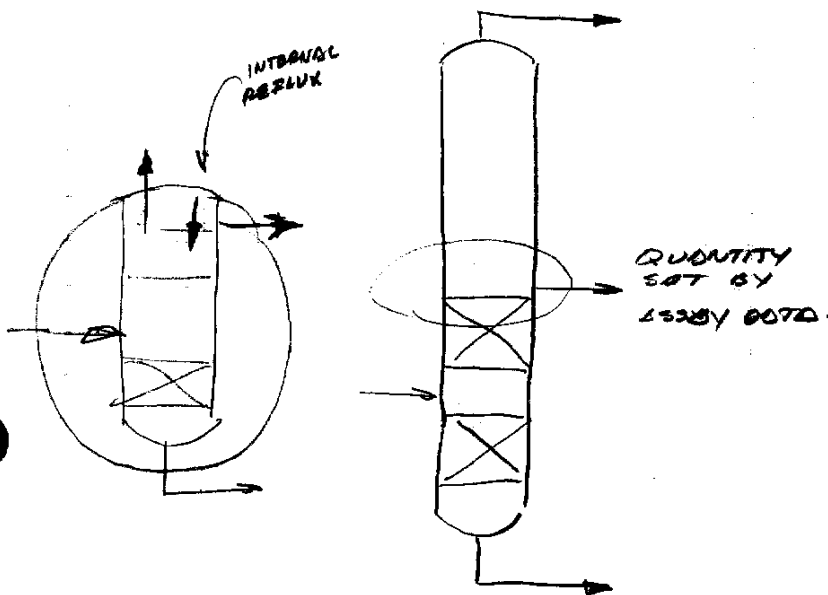
2° ENTHALPY BALANCE EQUATION $H_i = f(T, P, \text{COMPOSITION})$

3° EQUILIBRIUM EQUATION $y_i = K_i x_i$

$K_i = f(T, P, \text{COMPOSITION})$

↑
FIXED.

PARA EL CASO DE MUCHOS COMPONENTES. SE CONSIDERA O LA FRACCIONA
COMO CONO UNO CADA UNO NEGRO PARA EFECTOS DE SIMULACION Y CALCULO



LOS FUNCIONES PARA CALCULAR LAS ENTALPIAS DE MEZCLAS NO SON MUY PRECISOS PARA MEZCLAS.

OVERALL MATERIAL BALANCE

$$CRUDE = REDUCED CRUDE + DGO + DIESEL + JF + NAPHTHA (INCLUDING LIGHTENDS)$$

ENERGY BALANCE

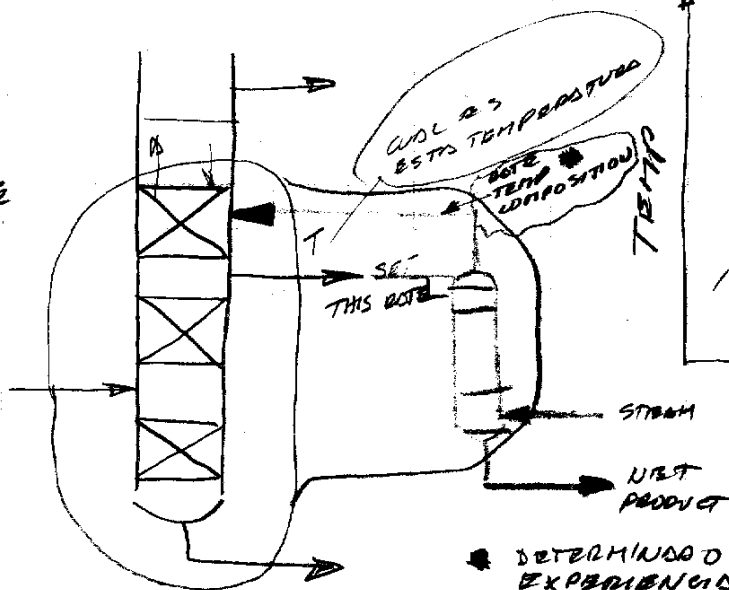
$$(CRUDE)/h_c = (REDUCED CRUDE)/h_{rc} + \overbrace{(NET VAPORS)/H_{NV}}^{EXCESS HEAT} + \Delta Q$$

IF $\Delta Q = 0$, THEN $IR = 0$ (THERE ISN'T ACCUMULATION OF HEAT & MASS).

$$\begin{aligned} \Delta Q &= \Delta Q_1 & IR &= IR_1 \\ \Delta Q_2 &> \Delta Q_1 & IR_2 &> IR_1 \end{aligned}$$

EFFICIENCY OF SEPARATION
DEGREE OF FRACTIONATION
NUMBER OF TRAYS
INTERNAL REFLUX

MASS
ENERGY
BALANCE

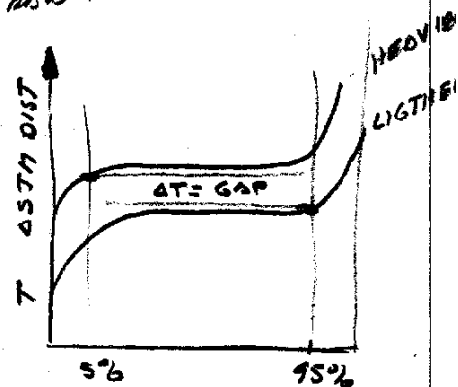
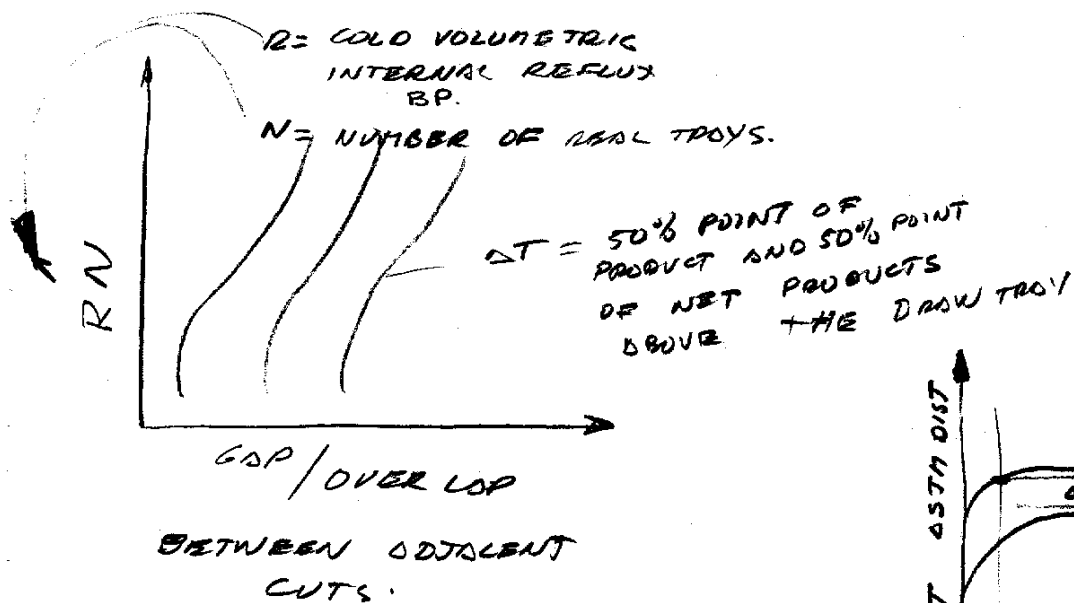


CURVA PREPARADA
DE ACUERDO A DATOS
OPERACIONALES DE
PLANTAS EXISTENTES
PARA ESTABLECER TEMPS
DE LOS CORTE

50% POINT
OF STRIPPED
PRODUCT.

DETERMINADO DE ACUERDO A LA
EXPERIENCIA.

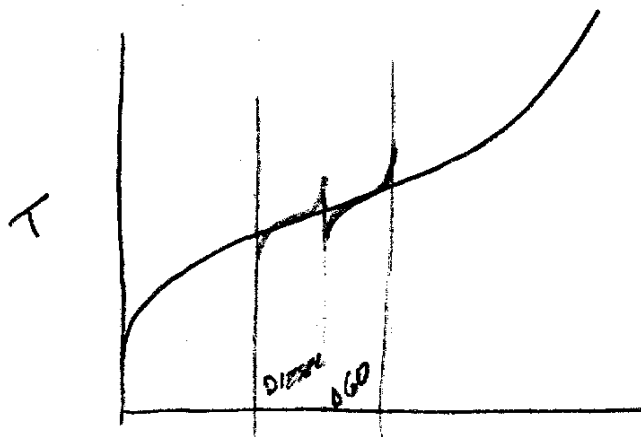
PODO EFECTOS DE CALCULO, LOS GASES QUE SOLEN DEJ. STRIPPER
LATERAL SE CONSIDERAN EL 10% DE LA CARGA DEL STRIPPER



GAP: DIFERENCIA ENTRE 95% Y 5% ENTRE CURVAS ADYACENTES
 DE TEMP.
 OR LIGHT PRODUCT
 HEAVY PRODUCT

PERO UN CAMINO
 EN CURVA DE DESTILACION

OVERLAP



DETERMINE DRAW TRAY TEMPERATURES. WORK FROM BOTTOM TO TOP.

REFERENCE: PD-58. CORRELATION OF DRAW TRAY TEMPERATURE VERSUS 50% BOILING POINT OF STRIPPED PRODUCT BASED PRIMARILY ON COMMERCIAL DATA. AS INDICATED THERE IS A BAND OF TEMPERATURES FOR A SPECIFIC PRODUCT 50% BOILING POINT. SELECTION OF THE MOST APPROPRIATE DRAW TEMPERATURE REVERTS INTO THE REFLM OF 'ENGINEERING JUDGEMENT / EXPERIENCE'. COMMERCIAL DATA FOR SIMILAR CRUDES / CUT RANGES IS ALWAYS OF GREAT VALUE IN PROVIDING DIRECTION. OPERATING PRESSURE OBVIOUSLY HAS AN IMPACT. NUMBER OF CUTS (I.E. NUMBER OF SIDE STREAMS), BOILING RANGE (WIDTH OF CUT) WILL ALSO AFFECT DRAW TRAY TEMPERATURE. DIRECTIONALLY ONE WOULD EXPECT FOR WIDER BOILING RANGE SIDECUT PRODUCTS (FEWER SIDE STREAMS) WOULD REDUCE DRAW TRAY TEMPERATURE.

ONCE THE DRAW TRAY TEMPERATURES ARE DETERMINED, PRUDENT ENGINEERING DESIGN PRACTICE INDICATES AN ANALYSIS OF IMPACT OF THIS TEMPERATURE BEING DIFFERENT ON DESIGN OF EQUIPMENT AND UNIT OPERABILITY. WE WILL COME BACK TO THIS POINT LATER.

LET'S PROCEED WITH AGO DRAW.

FROM CRUDE BREAKUP 50% POINT ON ESTIMATED ASTM IS = 690 °F

DRAW TRAY TEMP FROM PD-58 : 585 - 620 °F

CHOOSE 600 °F AS DRAW TRAY TEMPERATURE.

DIESEL DRAW

FROM CRUDE BREAKUP 50% POINT ON ESTIMATED ASTM IS ^{PD-58} 600 °F

DRAW TRAY TEMP FROM PD-58 : 527 - 569 °F

CHOOSE 550 °F AS DRAW TRAY TEMPERATURE

ENGINEERING DEPARTMENT
CUSTOMER ENGINEERING DESIGN SEMINAR
SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER HEAT BALANCE

PAGE NO. 2
DATE 10-5-91
PROJECT
BY E. MILLER

KEROSENE (JET FUEL) DRAW:

FROM CRUDE BREAKUP 50% ON ESTIMATED ASTM IS 460 °F.

DRAW TRAY TEMPERATURES FROM PD-SR: ⁴¹⁰ 410 - 448
CHOOSE 425 °F AS DRAW TRAY TEMPERATURES.

NOW WE WILL LOOK AT THE HEAT AND MATERIAL BALANCE AROUND EACH OF THE SIDECUT STRIPPERS

IN GENERAL, SIDECUT STRIPPERS IN CRUDE UNITS ARE STEAM STRIPPERS. ON OCCASION, THE UPPER SIDECUT STRIPPERS ARE REBOILED - IN THIS CASE THE JET FUEL SIDECUT STRIPPER IN OTHERS THE HEAVY NAPHTHA SIDECUT. IN OUR CASE ALL SIDECUT STRIPPERS WILL BE STEAM STRIPPERS. EXPERIENCE HAS INDICATED THAT THE FOLLOWING APPROACH IS ADEQUATE FOR STEAM SIDECUT STRIPPER DESIGN.

FOR REBOILED STRIPPERS 70-75%.

1. THE VAPORS STRIPPED OUT OF THE SIDECUT STRIPPER FEED WILL BE 80% OF THE FEED TO THE STRIPPER. I.E., THE STRIPPED PRODUCT WILL BE 90% OF THE UNSTRIPPED FEED.

2. TEMPERATURE OF THE STRIPPER VAPORS WILL BE 5 °F LOWER THAN THE ESTIMATED DRAW TRAY TEMPERATURE.

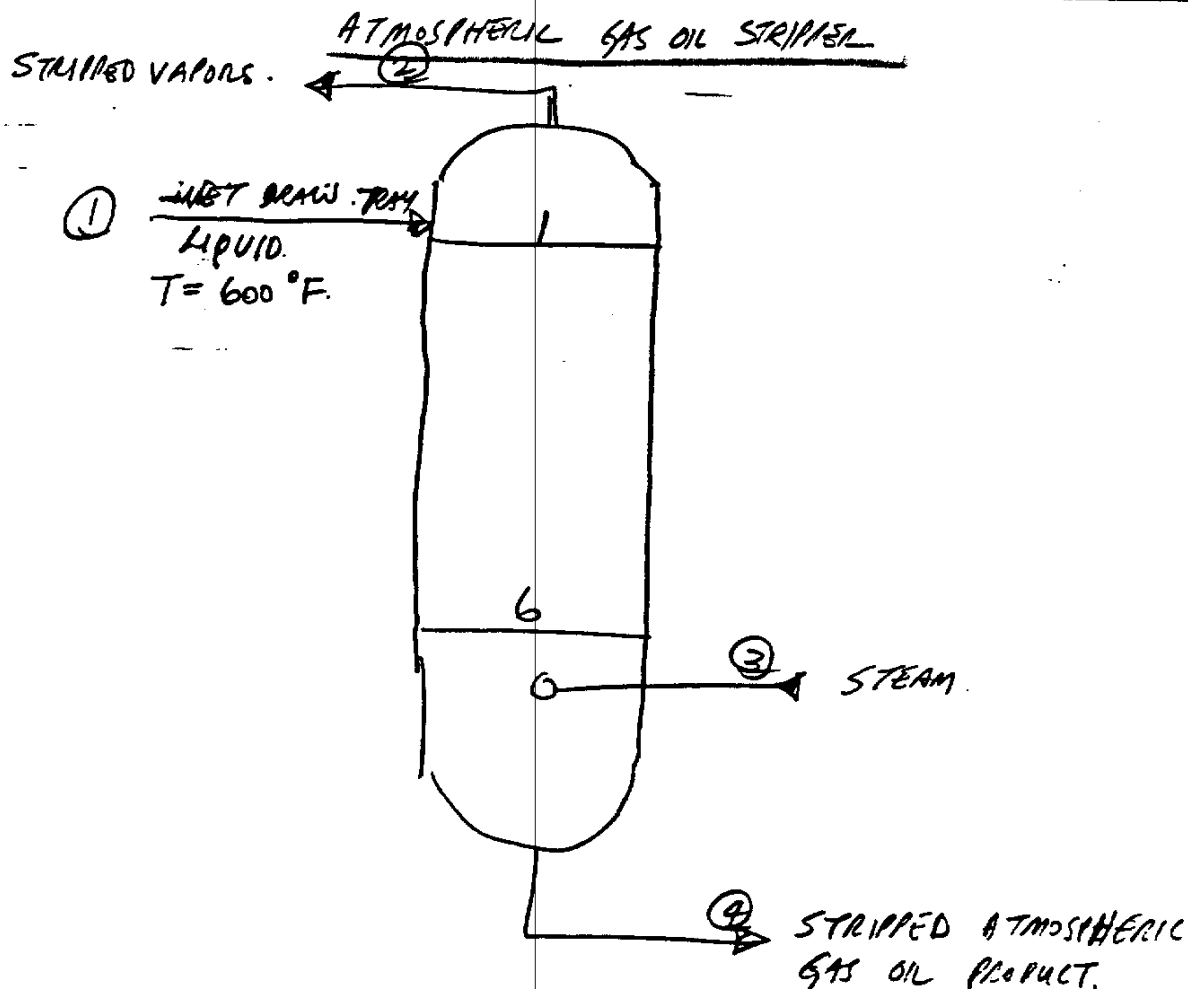
3. THE STRIPPED HYDROCARBON VAPORS CAN BE ADEQUATELY CHARACTERIZED BY ASSUMING THAT THE API OF THIS MATERIAL WILL BE 5 °API HIGHER THAN THE STRIPPED PRODUCT AND WILL HAVE A UOP K ABOUT EQUAL TO THE STRIPPED PRODUCT.

4. A STEAM RATE TO THE STRIPPER FOR DESIGN WILL BE 10 LBS / BBL OF STRIPPER FEED. GENERALLY LOW PRESSURE STEAM IS USED IN THIS SERVICE.

5. TEMPERATURE OF STRIPPED PRODUCT FROM SIDECUT STRIPPER WILL DETERMINED BY HEAT BALANCE.

WE WILL CALCULATE THE HEAT AND MATERIAL BALANCE FOR THE ATMOSPHERIC GAS OIL STRIPPER. THE DIESEL AND JET FUEL STRIPPER CALCULATIONS WILL BE DONE AS CLASS PROBLEMS.

6. TYPICALLY THESE STRIPPERS HAVE 6 REAL TRAYS.

CRUDE TOWER HEAT BALANCE

FROM CRUDE BREAKUP CALCULATIONS FOR STREAM ④

	④	③	②	①
AGO DRAW LIQUID	—	—	—	352.93
STRIPPED HL VAPORS.	—	—	42.63	—
STEAM	—	194.17	194.17	—
AGO PRODUCT	310.30	—	—	—
TOTAL MOL/HR	310.30	194.17	236.80	352.93
LB/HR	96815.	3495	13940	107260.
MOL WT	312	18	58.87	303.9
BPSD.	7550.	—	—	8390.
API	29.5	—	—	33.9
UOP K	11.87	—	—	11.87.
MASCED.	—	—	2.154	—

UOP'S GUIDELINES.

STEAM STRIPPERS

STRIPPED H₂C = 10% WT OF DRAW LIQUID

REFRIGERATED STRIPPER

STRIPPED H₂C = 20 - 25% WT OF DRAW LIQUID.

STEAM RATE = 10 LBS / BBL. OF FEED TO STRIPPER PARA DISEÑO
OPERACIONAL 7-8 UNICAMENTE.

COMPOSITION = $\Delta PI_{\text{VAPOR HC}} = \Delta PI_{\text{STRIPPED PRODUCT}} + 5$

UOPK $\approx$ UOPK PRODUCT.

TEMPERATURES

VAPOR FROM STR = DRAW TRAY - 5°F
TEMP

$$\text{STRIPPED HYDROCARBON VAPORS} = (0.1) \frac{(7550)}{0.9} \\ = 838.88 \Rightarrow 840 \text{ BPD.}$$

$$7550 + 840 = 8390 \text{ BPD}$$

$$\text{STRIPPER HL VAPORS} \quad \text{API} = 29.5 + 5 = 34.5 \\ \text{VOPK} = 11.87.$$

$$\text{FROM CHART B-2} \quad \text{CABP} \approx 580.$$

ENGLER SLOPE SHOULD BE SIMILAR OR SLIGHTLY GREATER THAN ESTIMATED PRODUCT ENGLER OF 1.06.

$$\text{THENCE MABP} \approx 575 \quad \text{FROM CHART B-1}$$

$$\text{MOL WT FROM CHART B-2} \approx 245.$$

$$\text{MASS FLOW} = (12.437)(840) = 10447 \Rightarrow 10445 \text{ LB/HR.}$$

$$\text{MOLAR FLOW} = \frac{10445}{245} = 42.63.$$

$$\text{DRAW TRAY LIQUID} = 7550 + 840 = 8390 \text{ BPD}$$

$$\text{MOLAR FLOW} = 310.3 + 42.63 = 352.93 \text{ mol/HR.}$$

$$\text{MASS FLOW} = 96815 + 10445 = 107260 \text{ LB/HR.}$$

$$\text{MOL WT} = \frac{107260}{352.93} = 303.9.$$

$$\text{PD-475.} \quad \text{API} = \frac{107260}{8390} = 12.7843 \Rightarrow 33.9 \\ \text{VOPK} = 11.87.$$

$$\text{STEAM RATE} = \frac{(8390) \times (10)}{24} = 3495.8 \Rightarrow 3495 \text{ LB/HR.} \\ = \frac{3495}{18} = 194.17 \text{ mol/HR.}$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER HEAT BALANCE

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 DATE 10-5-91
 PROJECT
 BY E. N. MILLER

ENTHALPY BALANCE

STRIPPED VAPOR TEMPERATURE = 595 °F
 STEAM TEMPERATURE = 297.8 °F (FROM STEAM TABLES)

SEE STEAM
TABLES

→ STEAM ENTHALPY. 1179.6 BTU/LB (REFERENCED TO SAT LIQ AT 32 °F)
 REMEMBER TO ADJUST DATA FOR DIFFERENCE IN
ENTHALPY DATUM.

STEAM ENTHALPY = 1179.6 - 27.4 = 1152 BTU/LB (SAT LIQ AT 60 °F)
 CORRECTION DATUM

USING ENTHALPY VS TEMPERATURE CHARTS DEVELOPED FOR
 THIS CLASS DESIGN PROBLEM ENTHALPY OF ATMOSPHERIC
 GAS OIL AND TEMPERATURE CAN BE CALCULATED. ASSUME
 THAT DIFFERENCES IN COMPOSITION OF HC STREAMS HAVE NO
 EFFECT ON ENTHALPY FUNCTION.

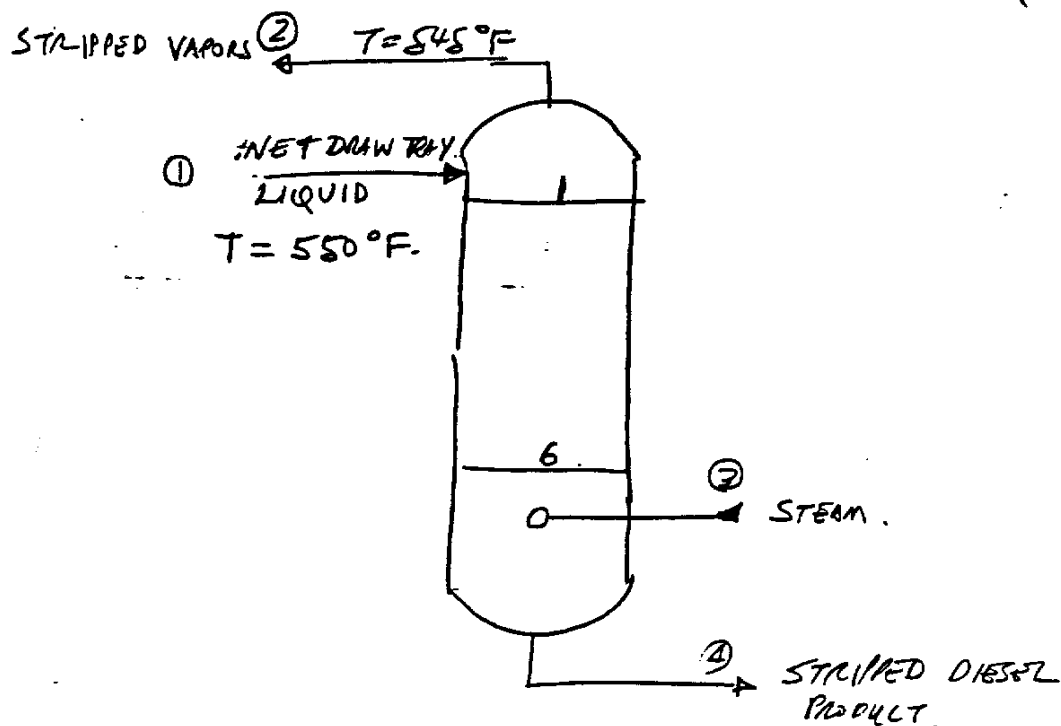
	IN 10 ⁶ BTU/HR	OUT 10 ⁶ BTU/HR
STREAM ①	34.323	
STREAM ②	-	8.685
STREAM ③	4.026	-
STREAM ④	-	-
TOTAL	38.349	8.685

FOR STREAM ③ HC VAPORS AT 595 °F = (395)(10445) = 4.126
 STEAM AT 595 °F = (1354.4)(3495) = 4.559
 (141514) 8.685

ENTHALPY OF ④ = 38.349 - 8.685 = 29.664 × 10⁶ BTU/HR.
 SPECIFIC ENTHALPY = 29.664 × 10⁶ / 96815 = 306.4

STRIPPED AGO TEMPERATURE = 518.2 °F **

SEE CURVA PRODUCTS ENTHALPHY
 TEMPERATURE

DIESEL STRIPPER.

FROM CRUDE BREAKUP CALCULATIONS FOR STREAM ④

	④	②	②	①
DIESEL DRAW LIQ.	-	-	-	509.56
STRIPPED HL VAPORS	-	-	61.1	-
STEAM	-	235.5	235.5	-
DIESEL PRODUCT	448.46	-	-	-
TOTAL MOL/HR	448.46	235.5	296.6	509.56
LB/HR	113460	4240	16520	125740
MOL WT.	253	18	55.7	246.8
BPSD	9150	-	-	10170
API	35.0	-	-	35.5
UOPK	11.93	-	-	11.93
MM SCFD	-	-	2.648	-
Temp, °F	530	-	545	550
Qty. MM Btu/HR	31.442	4.885	10.033	36.590

CRUDE TOWER HEAT BALANCE

$$\text{STRIPPED HYDROCARBON VAPOR} = \frac{(0.1)(9150)}{0.9} = 1016.7 \Rightarrow 1020.0 \text{ BPD.}$$

STRIPPER HC VAPORS

$$\text{API} = 35.0 + 5 = 40$$

$$\text{UOP K} = 11.93$$

CHART D-2

$$\text{CABP} \approx 495$$

$$\text{MABP} \times 4.90$$

$$\text{MW} = 201$$

$$\text{MASS FLOW} = (12.038)(1020) = 12278.8 \Rightarrow 12280 \text{ LB/HR}$$

$$\text{MOLE FLOW} = (12280)/(201) = 61.10$$

$$\text{DRAW TRAY LIQUID} = 1020 + 9150 = 10170 \text{ BPD.}$$

$$\text{MOLE FLOW} = 448.46 + 61.1 = 509.56 \text{ mol/hr}$$

$$\text{MASS FLOW} = 113460 + 12280 = 125740 \text{ LB/HR.}$$

$$\text{MOLE WT} = (125740)/509.56 = 246.76 \Rightarrow 246.8$$

$$\text{API: } \frac{125740}{10170} = 12.3638 \Rightarrow 35.5 \quad \text{UOP K} = 11.93$$

$$\text{STEAM RATE} = \frac{(10170)(107)}{24} = 4237.5 \Rightarrow 4240 \text{ LB/HR} \equiv 235.5 \text{ mol,}$$

ENTHALPY BALANCE

	IN	OUT.
(FROM CHART $\frac{1}{2} T = 550$) $\times$ MASS FLOW		
STREAM ①	106 BTU/HR	106 BTU/HR.
STREAM ②	36,590	10,033
STREAM ③	4,885	?
STREAM ④		
TOTAL	41,475	10,033

$$\text{FOR STREAM ②} \quad \text{HC VAPORS AT } 545^\circ\text{F} = (375)(12280) = 4,605$$

$$\text{STEAM AT } 545^\circ\text{F} = (1280.3)(4240) = 5,428$$

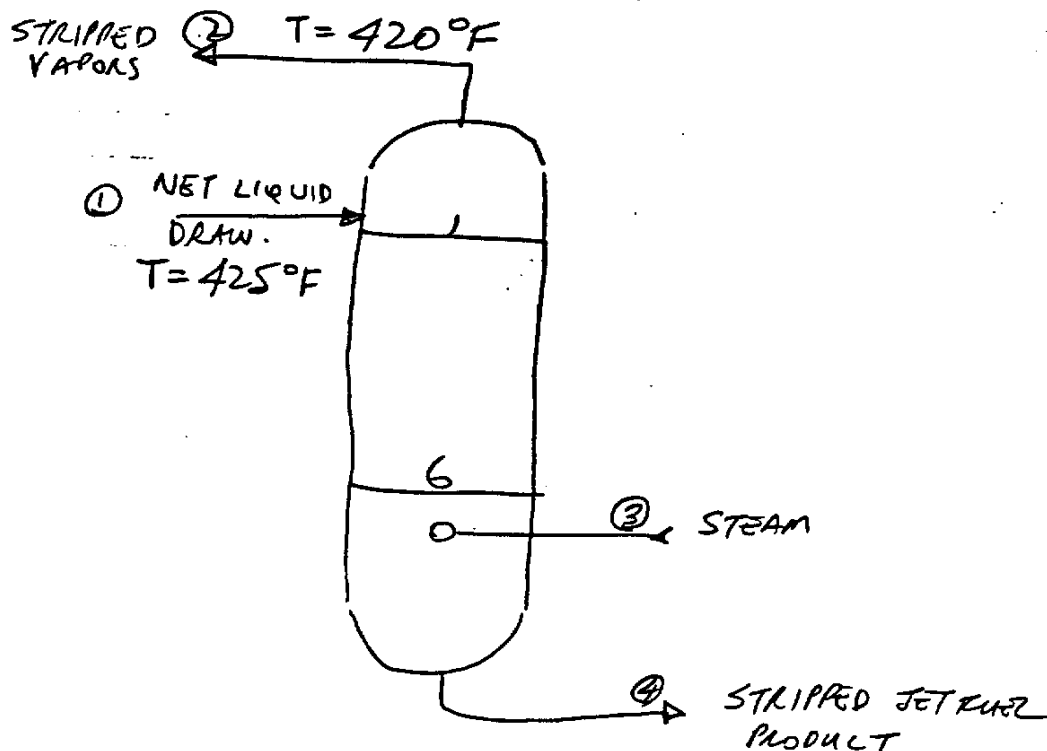
$$10,033$$

$$\text{ENTHALPY OF STREAM ④} = 41,475 - 10,033 = 31,442 \times 10^6$$

$$\text{SPECIFIC ENTHALPY} = 31,442 \times 10^6 / 113460 = 277.1$$

$$\text{STRIPPED DIESEL TEMPERATURE} = 530^\circ\text{F.}$$

JET FUEL STRIPPER



FROM CRUDE BREAKUP CALCULATIONS FOR STREAM ④

	④	③	②	①
JET FUEL DRAW LIQUID	-	-	-	1185.03
STRIPPED HC VAPORS	-	-	136.71	-
STEAM	-	425.5	425.50	-
JET FUEL PRODUCT	1048.32	-	-	-
TOTAL MOL/Hr	1048.32	425.5	562.21	1185.03
LB/Hr.	196035	7660	28850	217225
MOL WT.	187	18	51.3	183.3
BPSD	16550	-	-	18390
API	42.8	-	-	43.3
UOP K.	11.97	-	-	11.97
MM STD.	-	-	5.114	-

CLUDE TOWER HEAT BALANCE

$$\text{STRIPPED HYDROCARBON VAPOR} = \frac{(0.1)(16530)}{0.9} = 1838.9 \Rightarrow 1840 \text{ BPD}$$

$$\text{STRIPPER HC VAPORS: } \text{API} = 42.8 + 5 = 47.8$$

$$\text{UOP } K = 11.97$$

$$\text{CABP} \approx 385 \quad \text{MAB} \approx 380 \quad \text{MW} = 155$$

$$\text{MASS FLOW} = (11.575)(1840) = 2127.6 \Rightarrow 21190 \text{ LB/HR}$$

$$\text{MOLAR FLOW} = 136.71$$

$$\text{DOWN TRAY LIQUID} = 1840 + 16530 = 18390 \text{ BPD}$$

$$\text{MOLAR FLOW} = 136.71 + 1048.32 = 1185.03$$

$$\text{MASS FLOW} = 21190 + 196035 = 217225$$

$$\text{MOL WT} = 217225 / 1185.03 = 183.3$$

$$\text{API: } \frac{217225}{18390} = 118.12$$

$$\text{API} = 43.3 \quad \text{UOP } K = 11.97$$

$$\text{STEAM RATE: } \frac{(18390)(10)}{24} = 7662.5 \Rightarrow 7660 \text{ LB/HR} \approx 425.5 \text{ MOL/HR}$$

ENTHALPY BALANCE

	IN.	OUT.
	10^6 BTU/HR.	10^6 BTU/HR.
STREAM ①	95.617	
STREAM ②		15.85
STREAM ③	8.824	
STREAM ④		?
	<u>54.441</u>	<u>15.85</u>

$$\text{FOR STREAM ②: } \text{HC VAPORS AT } 920^\circ\text{F} = (307)(21190) = 6.505$$

$$\text{STEAM AT } 920^\circ\text{F} = (1220)(7660) = 9.345$$

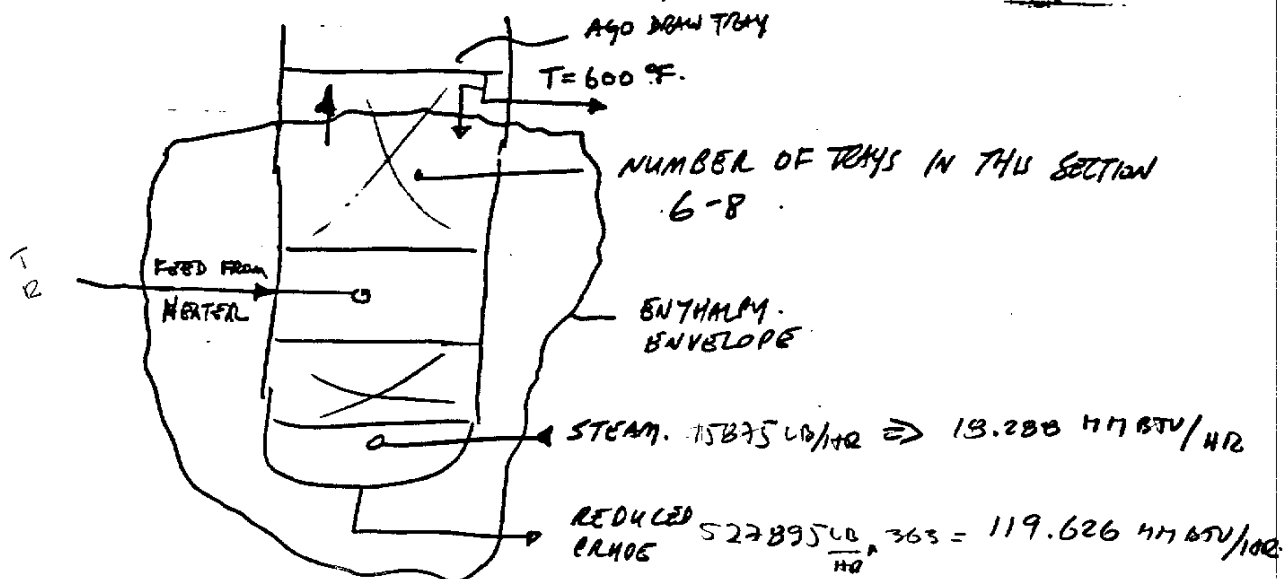
$$15.850$$

$$\text{ENTHALPY OF STREAM ④} = 54.441 - 15.85 = 38.591 \times 10^6$$

$$\text{SPECIFIC ENTHALPY} = 38.591 \times 10^6 / 196035 = 196.86 \text{ BTU/LB.}$$

$$** \text{ STRIPPED 887 FUEL TEMPERATURE} = 408^\circ\text{F}$$

LET'S MAKE A HEAT BALANCE AROUND BOTTOM SECTION OF TOWER BT. THE ATMOSPHERIC GAS OIL DRAW TO DETERMINE EXCESS HEAT RISING IN THE COLUMN AT THIS POINT. THIS WILL TELL US HOW MUCH HEAT IS AVAILABLE FOR INTERNAL REFLEX. MINIMUM EXCESS HEAT AT A DRAW IS 10% OF THE HEAT IN THE NET VAPORS RISING TO THE DRAW TRAY. THIS MINIMUM WILL GUARANTEE THAT WE CAN LIFT THE ATMOSPHERIC GAS OIL UP TO THE DRAW OFF. IN THIS SECTION THE EXCESS HEAT AVAILABLE IS DETERMINED BY FEED OVERFLSH.



TYPICAL TEMPERATURE OF VAPORS TO DRAW TRAY 20-30°F HIGHER THAN DRAW TRAY TEMPERATURE. FOR THIS PROBLEM SET TEMPERATURE OF VAPOR RISING TO DRAW TRAY AT $600 + 25 = 625^\circ\text{F}$.

MATERIAL BALANCE:

VAPOR TO DRAW TRAY CONSISTS OF

- UNSTABILIZED NAPHTHA
- JET FUEL
- DIESEL
- ATMOSPHERIC GAS OIL
- STRIPPING STEAM TO TOWER
- INTERNAL REFLEX FROM AGO DRAW TRAY

DEFINED AS
NET RISING VAPORS.



ENGINEERING DEPARTMENT

CUSTOMER ENGINEERING DESIGN SEMINAR
SUBJECT CRUDE UNIT DESIGN

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BY EN MILLER

EXCESS HEAT

CALCULATION OF HEAT IN CRUDE TO FLASH ZONE. TEMPERATURE IS 715°F. AT 151514 AS DETERMINED FROM OUR FLASH CURVE ANALYSIS. FROM CRUDE BREAKUP MASS FLOW IS 1235380 LB/HR.

VAPOR PORTION OF CRUDE FEED TO TOWER CONSISTS OF THE FOLLOWING:

UNSTABILIZED NAPHTHA (NAPHTHA LIGHTER)	= 301375 LB/HR
JET FUEL	= 196035 LB/HR
DIESEL	= 113460 LB/HR
ATMOSPHERIC GAS OIL	= 96815 LB/HR
OVER FLASH VAPOR	= 39200 LB/HR

WHEN WE DETERMINED THE FLASH ZONE TEMPERATURE WE SET THE VALUE OF THE OVERFLASH AS 310L% IN CRUDE. I.E.
OVERFLASH RATE = (100,000) (0.03) = 3000 BPD

WE NOW NEED TO ESTIMATE PROPERTIES OF THIS MATERIAL:

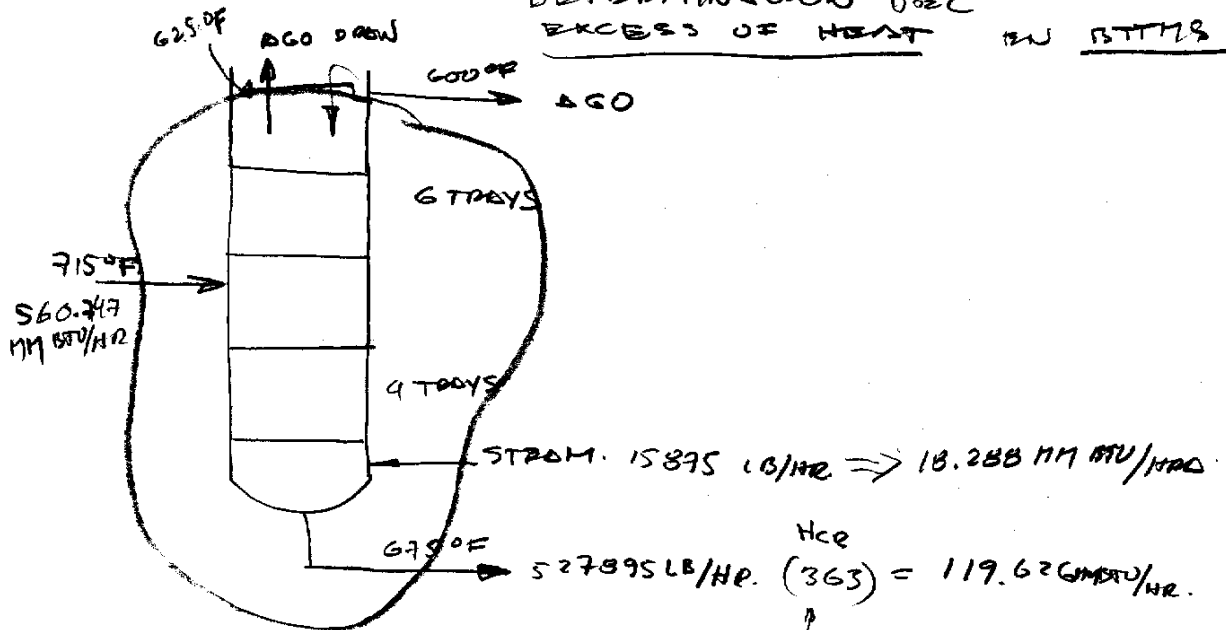
THE UOP K SHOULD BE ABOUT THE SAME AS THE AGO = 11.87
BOILING POINT SHOULD BE A LITTLE LESS THAN THE AGO
END POINT OF 767°F SAY 750°F. API OF 26.5 GIVES A CBP OF 750°F AND MOL WT. $\approx$ 350.

$$\text{MASS RATE} = (13.067)(3000) = 39,200 \text{ LB/HR}$$

LIQUID PORTION OF CRUDE FEED TO TOWER IS NET BOTTOMS MINUS OVER FLASH = 527895 - 39200 = 488695

	LB/HR	BTU/LB AT 70°F	BTU/HR AT 70°F
VAPOR			
UNSTAB NAPHTHA	301375	518	156,112
JET FUEL	196035	500	98,018
DIESEL	113460	492	55,822
AGO	96815	480	46,471
OVERFLASH VAPOR	39200	475	18,620
LIQUID	488695	380	185,704
TOTAL			560,747 MBTU/HR OF FEED
SPECIAL ENTHALPY = 4454 BTU/LB OR			

DETERMINATION OF EXCESS OF HEAT IN REFLUX



H_{CE}
BTU/LB
POSS CRUDO
REDUCED
GRAPHIC
ENTHALPIA
CRUDO REFL

	LB/HR.	625°F BTU/LB	MM BTU/HR
UNSTAB NAPHTHA	301375	455	137.126
JET-FUEL	196035	435	85.275
DIBENZOL	113.460	428	48.561
AGO	96.815	416	40.275
STEAM	15875	1319	20.935
			332.176

EXCESS OF HEAT = FEED STEAM CRUDO VAPOR IN DRAW TRAY
 $= 560.747 + 18.288 - 119.626 - 332.176$
 $= 55.233$ MM BTU/HR

$\% = \frac{55.233}{332.176} \times 100 = 16.63$

IF INTERNAL REFLUX = 0
 $\Rightarrow$ MINIMUM EXCESS HEAT = 10%

FOR RESONABLE FRACTIONATION
 MINIMUM EXCESS HEAT $\approx$ 20% MAX = 25%

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PROJECT

BY E N MILLEREXCESS HEAT CALCULATION

	IN. LB/HR.	OUT LB/HR.	SPECIFIC ENTHALPY BTU/HR. C 625 OUT.	IN MMBTU/HR.	OUT. MMBTU/HR.
CRUDE	1235580			560.747	
STEAM	15875	15875	1319	18.288	20.939
UNSTAB NAPH		301375	455		137.126
JET FUEL		196035	435		85.275
DIESSEL		113460	428		48.561
A60		96815	416		40.275
RESID.		527895	(363)		191.626
				579.035	523.802

$$\text{HEAT AVAILABLE FOR INTERNAL REFLUX} = 579.035 - 523.802 \\ = 55.233 \text{ MMBTU/HR.}$$

$$\text{EXCESS HEAT} = \frac{55.233}{(20.939 + 137.126 + 85.275 + 48.561 + 40.275)} = \frac{55.233}{332.176} \times 100 \\ = 16.63 \% \quad \text{OK.}$$



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EXCESS HEAT

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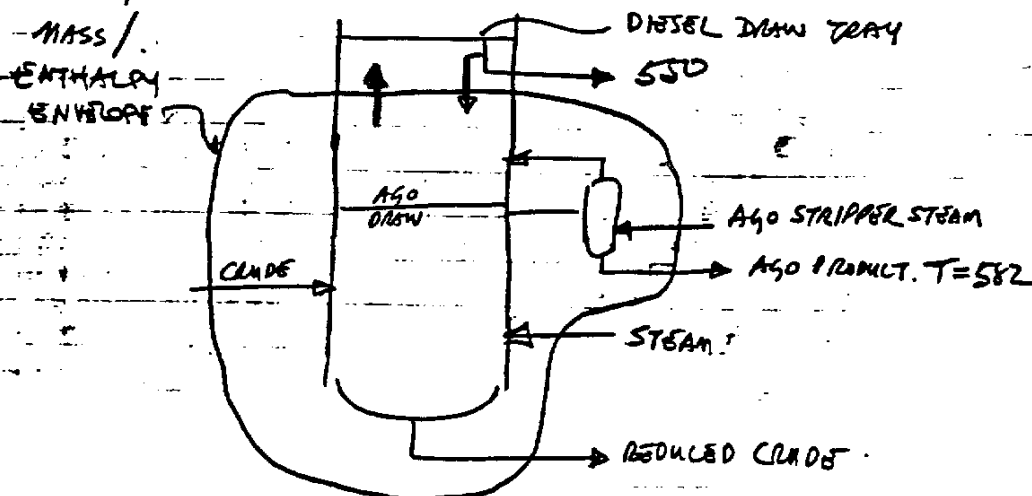
DATE 10-5-91

PROJECT

BY E. MILLER

EXCESS HEAT CONCEPT.

MOVING UP THE TOWER TO THE DIESEL DRAW TRAY.



ASSUME. VAPORS RISING TO DRAW TRAY ARE AT $550 + 25 = 575^{\circ}\text{F}$.
IN SAME FASHION AS BEFORE MAKE BALANCES. NOTE
AGO SIDECUT STRIPPER INSIDE ENVELOPE.

	IN LB/HR	OUT LB/HR	SPECIFIC ENTHALPY @ 575	ENTHALPY IN MM BTU/HR	SAVED OUT MM BTU/HR
CRUDE	1235580			560,767	
COLUMN STEAM	15875	15875	1295	18,288	
AGO STR STEAM	3495	3495	1295	4,026	
UNSTAB NAPK		301375	422		20,554
JET FUEL		196035	400		4,526
DIESEL		113460	395		127,180
AGO		96815			78,414
RESID		527895			44,817
TOTALS				583,081	496,781

$$\text{HEAT AVAILABLE FOR INTERNAL REFLUX} = 583,081 - 496,781 = 86.3 \times 10^6 \text{ BTU/HR.}$$

$$\text{EXCESS HEAT} = \frac{86.3}{(20,554 + 4,526 + 127,180 + 78,414 + 44,817)} = \frac{86.3}{275,491} \times 100 = 31.3\%$$

AS MENTIONED PREVIOUSLY, 10% EXCESS HEAT IS CONSIDERED THE MINIMUM REQUIRED TO ASSURE LIFTING OF ALL OF THE DIESEL PRODUCT. GENERALLY, TO ENSURE GOOD FRICTION BETWEEN PRODUCTS 20% EXCESS HEAT IS OFTEN USED AS A TARGET IN DESIGN.

SINCE THE EXCESS HEAT AVAILABLE AT THE DIESEL DRAW EXCEEDS OUR TARGET OF 20% EXCESS HEAT AT THE DRAW, SOME HEAT CAN BE REMOVED FROM THE SYSTEM VIA A PUMPAROUND EXCHANGE AT THE AGO DRAW. LET'S DETERMINE WHAT THAT HEAT REMOVAL DUTY IS.

FOR 20% EXCESS HEAT OF DIESEL DRAW HEAT REQUIRED FOR INTERNAL REFLEX = $(275.491)(0.2) = 55.098$ MMBTU/HR.

HEAT AVAILABLE FOR PUMPAROUND HEAT REMOVAL = $86.3 - 55.1$
 $= 31.2$ MMBTU/HR.

AT THIS POINT WE WILL SET THE AGO PUMPAROUND HEAT REMOVAL RATE AT 30 MMBTU/HR.. WE MAY NEED TO ADJUST THIS PUMPAROUND DUTY LATER DEPEND ON EFFECTIVENESS OF THE FRACTION IN THIS SECTION RELATIVE TO PRODUCT SPECIFICATIONS AND UTILIZATION OF THIS HEAT IN THE HEAT EXCHANGE NETWORK DESIGN.

THERE SOME OPTIONS AT THIS POINT AVAILABLE RELATIVE TO THE ORDER OF THE CALCULATIONS.

- 1) WE CAN CONTINUE HEAT BALANCE CALCULATIONS UP THE COLUMN AND THEN CHECK PRODUCT QUALITY LATER. THIS MAY REQUIRED A REITERATION OF THE HEAT BALANCE.
- 2) WE CAN CHECK PRODUCT QUALITY AS WE PROCEED UP THE COLUMN.

SINCE WE HAVE NOT SET ANY GAP/OVERLAP SPECIFICATION IN THE GIDERTY PRODUCTS LETS CONTINUE OUR HEAT BALANCE CALCULATIONS UP THE COLUMN.

EXCESS HEAT

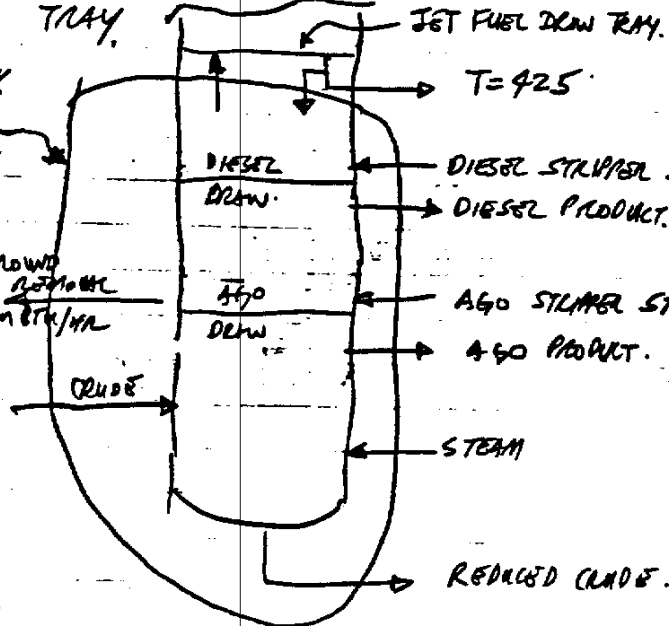
AS A CLASS PROBLEM, DO THE HEAT BALANCE AT THE JET FUEL DRAW TRAY.

- ① ASSUME THAT VAPORS TO TRAY ARE 20°F HIGHER THAN DRAW TRAY TEMPERATURE.
- ② CALCULATE EXCESS HEAT. IF GREATER THAN 20% DETERMINE AMOUNT OF PUMPAROUND HEAT THAT COULD BE REMOVED AT DEESEL DRAW TRAY MAINTAINING 20% MINIMUM EXCESS HEAT AT THE JET FUEL DRAW TRAY.
- ③ REMEMBER WE HAVE REMOVED 30 MM BTU/HR FROM THE SYSTEM WITH AN A-60 PUMPAROUND EXCHANGE. THIS MUST BE ACCOUNTED FOR IN HEAT BALANCE.
- ④ ROUND DOWN AVAILABLE HEAT REMOVAL DUTY TO NEAREST 5 MM BTU/HR.

EXCESS HEAT

JET FUEL DRAW TRAY

JET FUEL DRAW TRAY

MASS/ENTHALPY
ENVELOPE

	IN LB/HR	OUT LB/HR	SPECIAL ENTHALPY @ 495°F.	IN MM BTU/HR	OUT MM BTU/HR @ 445°F
CRUDE	1235580			560,767	
COLUMN STEAM	15875	15875	1232	18,288	19,558
AGO STRIPPED STEAM	3495	3495	1232	4,026	4,306
DIESEL STRIPPED STEAM	4240	4240	1232	4,885	5,224
UNSTAB NAPHTHA		301375	345		103,974
JET FUEL		196035	322		63,123
DIESEL		113460			36,590
AGO		98815			29,664
RESID		527895			196,626
AGO PUMP AROUND					30,000
TOTALS				587,996	484,065

HEAT AVAILABLE FOR INTERNAL REFLUX = 103,901 MM BTU/HR

$$\text{EXCESS HEAT} = \frac{103,901}{(19,558 + 4,306 + 5,224 + 103,974 + 63,123)} = \frac{103,901}{196,215} \times 100$$

$$= 53 \%$$

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EXCESS HEAT

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PROJECT

BY F. A. MILLER

HEAT REMOVAL AVAILABLE AT DIESEL DRAIN VIA PUMP-AROUND EXCHANGE.

$$\text{HEAT REQUIRED FOR INTERNAL REFLUX} = (196.215)(0.2) = 39.243$$

$$\text{HEAT AVAILABLE FOR EXCHANGE} = 103.901 - 39.243 = 64.658 \text{ MBTU/HR}$$

LET'S LIMIT HEAT REMOVAL TO 60 MBTU/HR



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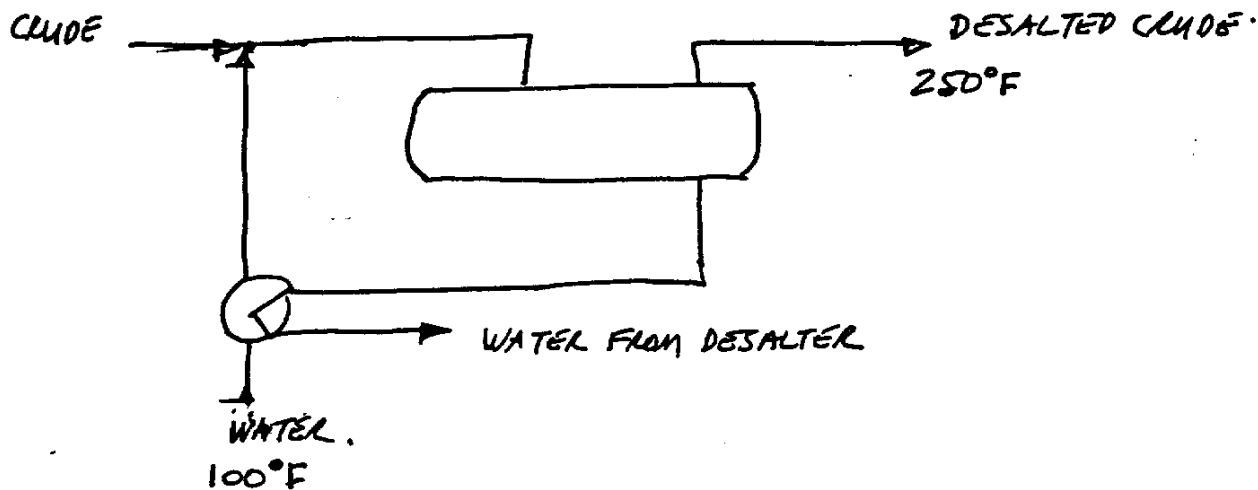
10-5-91

PROJECT

BY

EA MILLER

DETERMINE CRUDE TEMPERATURE TO DESALTER.



SET COLD END APPROACH ON EXCHANGE AT 50°F

CALCULATE CRUDE TEMPERATURE TO DESALTER.

2

ENTHALPY BALANCE

ENTHALPY OF CRUDE TO DESALTER

$$= 119.857 + 3.94 - 1.75$$

$$= 122.04 \text{ MBTU/HR.}$$

$$\text{SPECIFIC ENTHALPY} = \frac{122.04 \times 10^6}{1235580} = 98.77 \text{ BTU/LB.}$$

CRUDE TO DESALTER TEMP = 255°F

PERO 5% ACID OR DIBASIC T = 261°F.



$$m_{H_2O}(h_{150} - h_{100}) = m_{H_c}(H_{250} - H_x)$$

$$H_{250} = 96.5 \text{ BTU/LB.} \quad \text{HYDROCARB}$$

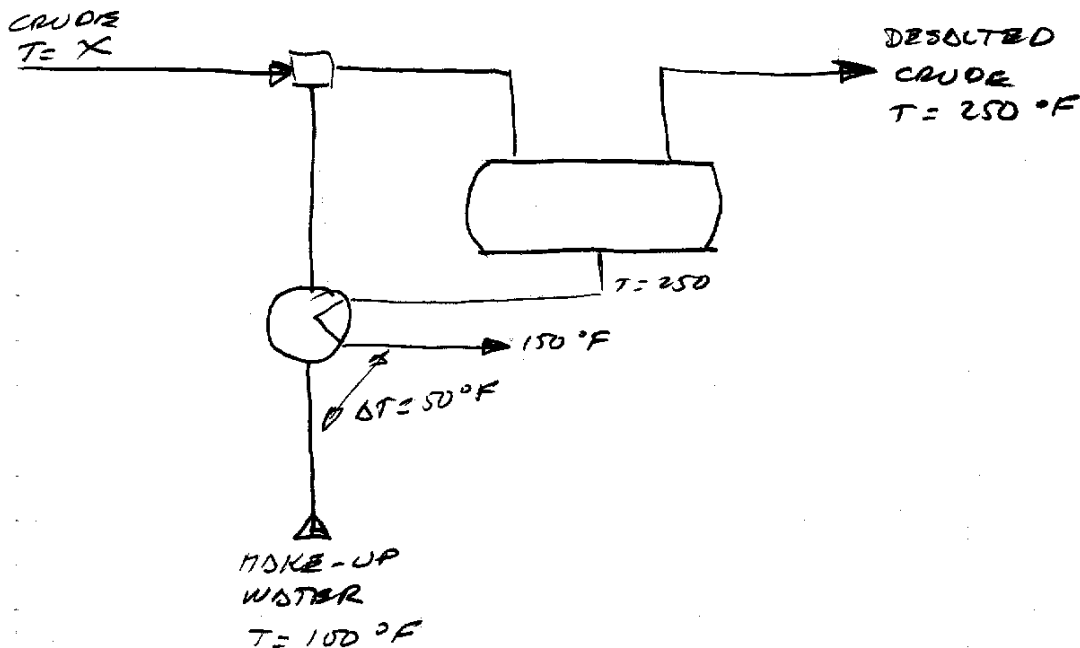
$$h_{100} = 40 \text{ BTU/LB.}$$

$$h_{150} = 90 \text{ BTU/LB.}$$

$$H_x = H_{250} - \frac{m_{H_2O}(h_{150} - h_{100})}{m_{H_c}}$$

$$H_x = 98.9 \quad T = 255^\circ\text{F}$$

DETERMINE CRUDE TEMP TO DESALTER



$$\text{WATER RATE} = 0.03 \times 100000 \text{ BPSD} = 3000 \text{ BPSD}$$

$$= \frac{3000 \times 42.934}{24} = 43785 \text{ LB/HR}$$

TEMP OF DESALT WATER FROM DESALTER 250°F
TEMP OF DESALT WATER AFTER EXCHANGE 150°F

$$\text{EXCHANGER OUT} = 43785 \times 1 \times (250 - 150)$$

MAKEUP WATER TEMP AFTER EXCHANGE $\approx$ 200°F

$$\text{ENTHALPY OF MAKE UP WATER} = 43785 \times 1 (100 - 60) = 1.751$$

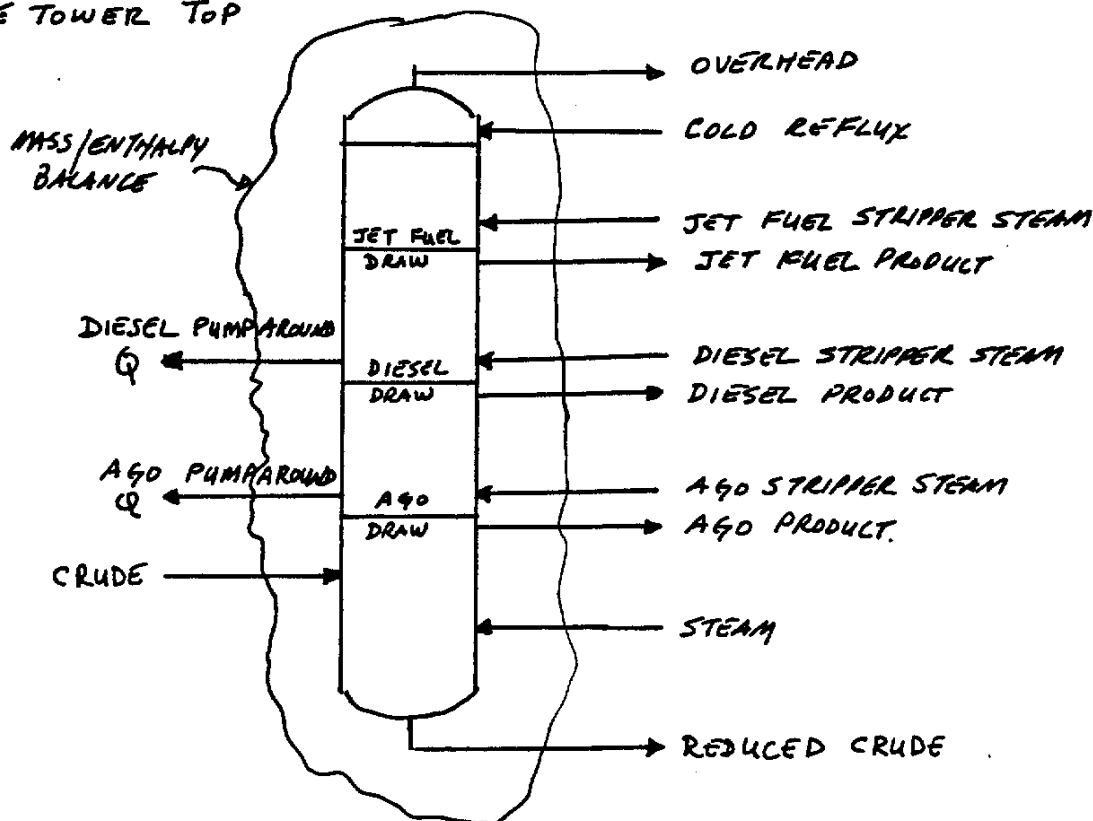
ENTHALPY OF DESALTER WATER AFTER EXCH

$$43785 (150 - 60) = 3.94 \text{ MM BTU/HR}$$

MM BTU/HR

$$\text{ENTHALPY OF DESALTED CRUDE} = 1235580 \left(\frac{17.87 \text{ BTU/LB}}{\text{LB/HR}} \right) = 119.851 \text{ MM BTU/HR}$$

CRUDE TOWER TOP



GOING TO THE TOP OF THE TOWER WHERE THE HEAT BALANCE IS SOMEWHAT DIFFERENT THAN THAT CONSIDERED FOR OTHER SECTIONS. IN THIS CASE THE OVERHEAD PRODUCT IS A VAPOR WITH HEAT REMOVAL BY COLD EXTERNAL REFLUX.

AS MENTIONED IN THE CRUDE BREAKUP A $R/D = 1.0$ MINIMUM (EXTERNAL REFLUX/DISTILLATE) IS REQUIRED. IF THIS REQUIREMENT CAN NOT BE MET THEN HEAT REMOVAL VIA PUMP AROUNDS IN THE COLUMN MUST BE REDUCED.

IN ANY EVENT, THE REFLUX WILL REMOVE AS MUCH HEAT AS REQUIRED TO HEAT BALANCE THE COLUMN.

COLD REFLUX ^{RATIO} HAS ALSO BEEN DEFINED AT THE RATIO OF THE HEAT REMOVED BY THE REFLUX TO THE HEAT IN THE NAPHTHA GOING OVERHEAD. TYPICALLY THIS RATIO IS 1.0 TO 1.5 FOR CRUDE TOWERS.

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LET'S NOW CALCULATE THE TEMPERATURE OF THE OVERHEAD VAPOR FROM THE CRUDE TOWER. FROM OUR PREVIOUS HYDRAULIC ANALYSIS WE SET THE OPERATING PRESSURE AT THE TOP OF THE TOWER AT 10 PSIG. WE ALSO HAD DECIDED TO OPERATE RECEIVER AT CONDITIONS SUCH THAT THE NET OVERHEAD FROM THE COLUMN WOULD BE ALL LIQUID. THE OPERATING CONDITIONS OF THE RECEIVER WERE SET AT 140°F AND 5 PSIG.

FIRST CONSIDER DEW POINT OF THE HYDROCARBON ONLY NOT CONSIDERING PRESENCE OF WATER VAPOR. REMEMBER DEW POINT IS DEFINED AS $\sum y_i / K_i = 1.0$

FROM CRUDE BREAKUP @ 10 PSIG.

		300°F		10 PSIG.	
		K_i	f_i / K_i	250°F	f_i / K_i
C ₃	28.56	24.99	1.14	19.24	1.48
iC ₄	24.03	14.76	1.63	10.91	2.20
nC ₄	155.49	13.03	11.93	9.11	17.07
iC ₅	92.24	7.46	12.37	5.05	18.26
nC ₅	187.62	6.72	27.92	4.45	42.16
CP	10.93	5.49	1.99	3.47	3.15
NAPH	2429.31	1.45	1685.39	0.768	3163.16
	2928.18		173236		3247.50
			< 2928.18		> 2928.18
			ALL VAPOR		TWO PHASE

ESTIMATED DEW POINT TEMPERATURE $\approx 260^\circ\text{F}$

LET'S BREAK NAPHTHA INTO TWO PIECES AND RECALCULATE DEW POINT.

		300°F		250	
		K_i	f_i / K_i	K_i	f_i / K_i
C ₃	28.56	24.99	1.14	19.24	1.48
iC ₄	24.03	14.76	1.63	10.91	2.20
nC ₄	155.49	13.03	11.93	9.11	17.07
iC ₅	92.24	7.46	12.37	5.05	18.26
nC ₅	187.62	6.72	27.92	4.45	42.16
CP	10.93	5.49	1.99	3.47	3.15
LSR	1164.63	2.90	406.60	1.071	681.07
HVN	1264.68	0.67	2049.72	0.287	4406.55
	2928.18		2508.3		5171.94

ESTIMATED DEW POINT TEMPERATURE $\approx 292^\circ\text{F}$

COMPUTER SIMULATION CALCULATION OF DEW POINT WITH TOTAL NAPHTHA
BROKEN INTO 5 PSEUDO COMPONENTS WAS 300°F.

USING 300°F AS TEMPERATURE LETS LOOK AT EXCESS HEAT AT TOP.

	IN LB/HR	OUT LB/HR	SPECIFIC ENTHALPY @ 300°F	IN MM BTU/HR.	OUT. MM BTU/HR.
CRUDE	1235580			560.767	
COLUMN STEAM	15875	15875	1162.8	18.288	18.460
A90 STR. STEAM	3495	3495		4.026	4.064
DIESEL STR. STEAM	4240	4240		4.985	4.930
JET FUEL STR. STEAM	7660	7660		8.824	8.907
UNSTABILIZED NAPHTHA		301375	263		79.262
JET FUEL		196035			38.591
DIESEL		113460			36.590
A90		96815			29.626
RESID.		527895			191.626
A90 PUMP/AROUND					30.0
DIESEL PUMP/AROUND					60.0
				596.79 EXCEPTO REFLUX	502.056

$$\text{EXCESS HEAT} = 596.790 - 502.056 = 94.704 \text{ MM BTU/HR.}$$

$$\text{REFLUX REQUIRED TO REMOVE HEAT} = \frac{94.704 \times 10^6}{(263 - 44)} = 432,438 \text{ LB/HR.}$$

ADDING REFLUX TO OUR BALANCE.

REFLUX 432,438 432,438

$$\begin{array}{r} 19.027 \\ 113.731 \\ \hline \text{TOTAL } 615.817 \equiv 615.817 \end{array}$$

$$\text{REFLUX / DISTILLATE} = \frac{432,438}{301,375} = 1.435$$

IN THIS ANALYSIS SEVERAL THINGS WERE OVERLOOKED OR NEGLECTED. TOTAL OVERHEAD FROM COLUMN INCLUDING REFLUX.

	MOL/HR
H ₂ O	1737.07.
C3	40.98
iC4	34.48
nC4	223.11
iC5	132.35
nC5	269.21
CP	15.68
NAPH	<u>3485.78</u>
TOTAL	5938.66

WHAT PHASE IS THE WATER AT COLUMN TOP?

$$T_1 = 24.7 \text{ PSIA.}$$

$$P_{PH_2O} = \left(\frac{1737.07}{5938.66} \right) (24.7) = 7.22 \text{ PSIA.}$$

$$P_{PHC} = 24.7 - 7.22 = 17.48 \text{ PSIA.}$$

$$VP \text{ H}_2\text{O AT } 300^\circ\text{F} \quad 67.005 \text{ PSIA.}$$

$$P_{PH_2O} < VP_{H_2O} \quad \text{THEN ALL VAPOR.}$$

IS DEW POINT (OVERHEAD VAPOR TEMPERATURE) AFFECTED BY PRESENCE OF STEAM IN VAPOR? HOW MUCH?

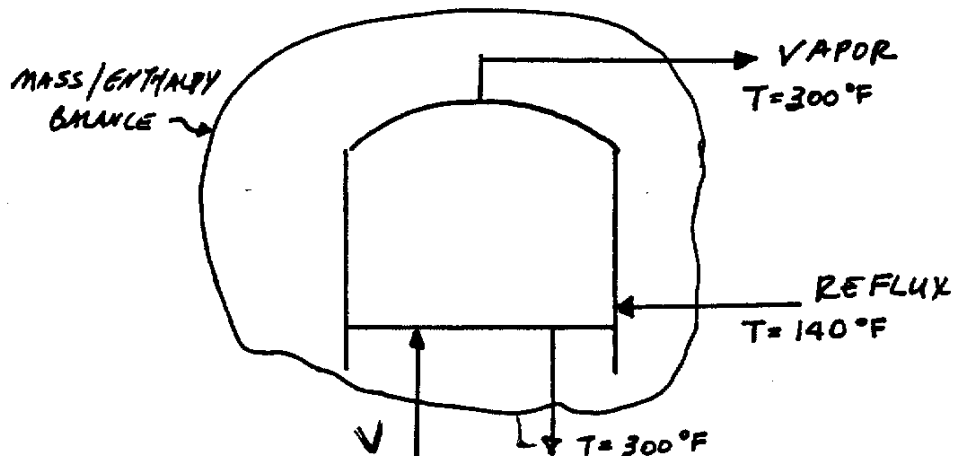
WHAT HAPPENS TO THE REFLUX RATE?

WE WILL LET THE STUDENT ANSWER THOSE QUESTIONS.

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN.
EXCESS HEAT

PAGE NO. 27
 DATE 10-5-91
 PROJECT _____
 BY E N MILLER

DETERMINE THE INTERNAL REFLUX AT THE TOP OF THE TOWER.

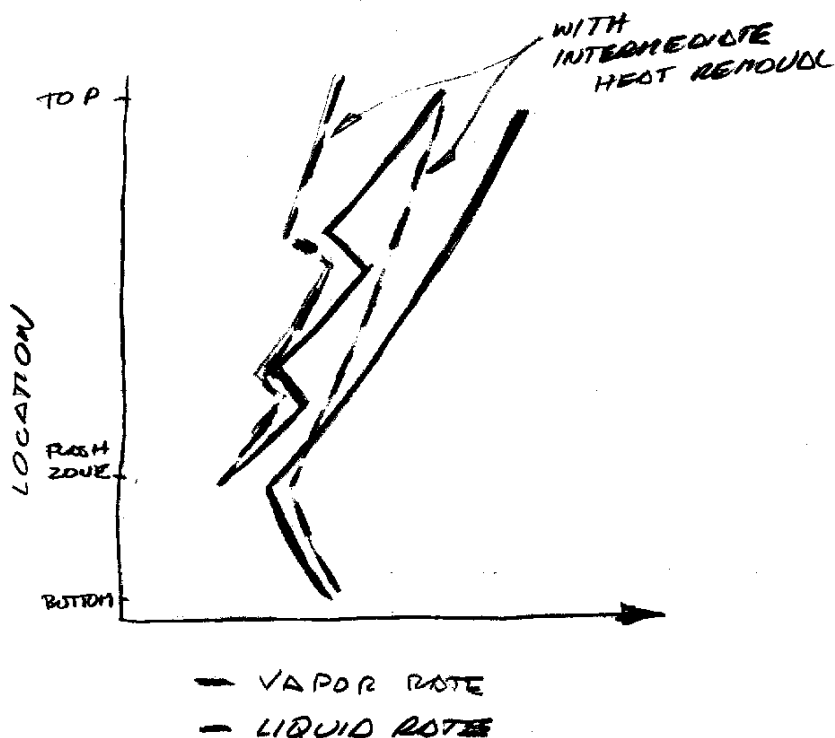


ASSUME TEMPERATURE OF VAPORS TO TOP TRAY = 310°F

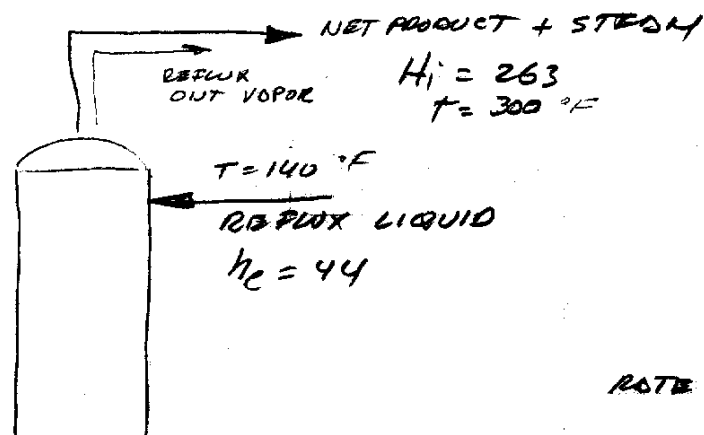
	IN LB/HR	OUT LB/HR	SPECIFIC ENTHALPY IN OUT		IN MMBTU/HR	OUT MMBTU/HR.
STEAM	31270	31270	1167.66	1162.8	36.513	36.361
UNSTABILIZED NAPHTHA	301375	301375	267	263	80.467	79.262
EXT REFLUX (L)	432438		44		19.027	
EXT REFLUX (V)		432438		263		113.371
					136.007	228.994

$$\begin{aligned} \text{HEAT BROUGHT IN BY INTERNAL REFLUX} &= 228.994 - 136.007 \\ &= 92.987 \times 10^6 \text{ BTU/HR} \end{aligned}$$

$$\text{INTERNAL REFLUX} = \frac{92.987 \times 10^6}{267 - 141} = \underline{\underline{737,992 \text{ LB/HR}}}$$



DETERMINACION DEL REFLUJO EXTERNO



$$\text{RATE} = \frac{Q}{H_i - h_e}$$

ENTHALPY BALANCE : IN = OUT

CRUDE + STEAM + AGO STR STM + DIESEL STR STM + JF STR STM + REFLUX (L)

= REDUCE CRUDE + AGO + DIESEL + JF + NAPHTHA + AGO PRADISEA + ALL VAPOR + REFLUX (V)

$$590.79 + \text{REFLUX (L)} = 502.056 + \text{REFLUX (V)}$$

$$596.79 + MR h_L = 502.056 + MR \cdot H_V$$

$$MR (H_V - h_L) = 596.79 - 502.056.$$

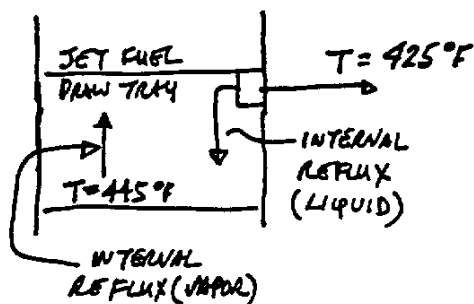
CUSTOMER ENGINEERING DESIGN SEAWAR
 SUBJECT CLUDGE UNIT FRACTIONATION.
EXCESS HEAT

PAGE NO. 28
 DATE 10-5-91
 PROJECT
 BY REN MILLER

CALCULATION OF INTERNAL REFLUX RATE BELOW JET FUEL DRAW.

AFTER REMOVING 60 MM BTU/HR OF HEAT AT DIESEL DRAW WITH A PUMPAROUND EXCHANGER, HEAT AVAILABLE FOR INTERNAL REFLUX (SEE PAGE 21) IS AS FOLLOWS:

$$\text{HEAT AVAILABLE FOR INTERNAL REFLUX} = 103.901 - 60 = 43.901 \text{ MM BTU/HR}$$



FROM ENTHALPY/TEMPERATURE CHARTS

$$\text{ENTHALPY AS LIQUID} = 210 \text{ BTU/LB @ } 425^\circ\text{F}$$

$$\text{ENTHALPY AS VAPOR} = 322 \text{ BTU/LB @ } 445^\circ\text{F}$$

$$\text{INTERNAL REFLUX} = \frac{43.901 \times 10^6}{(322 - 210)} = 391,973 \text{ LBS/HR.}$$

EARLIER WE CALCULATED API OF DRAW TRAY LIQUID AS 43.3

$$\text{INTERNAL REFLUX (BPSD)} = \frac{391,975}{302,650 / 11.812} = \frac{33,184}{25,627} \text{ BPSD}$$

(BELOW JET FUEL DRAW)

CALCULATION OF INTERNAL REFLUX RATE BELOW DIESEL DRAW:

AFTER REMOVING 30 MM BTU/HR OF HEAT AT A40 DRAW WITH A PUMPAROUND EXCHANGER, HEAT AVAILABLE FOR INTERNAL REFLUX (SEE PAGE 15)

$$\text{HEAT AVAILABLE FOR INTERNAL REFLUX} = 86.3 - 30 = 56.3 \text{ MM BTU/HR}$$

$$\text{INTERNAL REFLUX ENTHALPY AS LIQUID} = 311 \text{ BTU/LB AT } 550^\circ\text{F}$$

$$\text{INTERNAL REFLUX ENTHALPY AS VAPOR} = 395 \text{ BTU/LB AT } 575^\circ\text{F}$$

$$\text{INTERNAL REFLUX} = \frac{56.3 \times 10^6}{(395 - 311)} = 670,240 \text{ LBS/HR.}$$

PREVIOUS CALCULATION OF DRAW TRAY LIQUID API (SEE PAGE 6) = 35.5

$$\text{INTERNAL REFLUX (BPSD)} = \frac{670,240}{12.3638} = 54,210 \text{ BPSD}$$

(BELOW DIESEL DRAW)

CUSTOMER

ENGINEERING DESIGN SEMINAR

SUBJECT

CRUDE UNIT FRACTIONATION
EXCESS HEAT

PAGE NO. 29

DATE 10-5-91

PROJECT

BY E.N. MILLER

CALCULATION OF INTERNAL REFLUX RATE BELOW A90 DRAW:

HEAT AVAILABLE FOR INTERNAL REFLUX (SEE PAGE 14) = 55.233 MM BTU/HR

INTERNAL REFLUX ENTHALPY AS LIQUID = 320 BTU/LB @ 600°F

INTERNAL REFLUX ENTHALPY AS VAPOR = 416 BTU/LB @ 625°F

$$\text{INTERNAL REFLUX} = \frac{55.233 \times 10^6}{(416 - 320)} = 575,345 \text{ LB/HR}$$

PREVIOUS CALCULATION OF DRAW TRAY LIQUID API (SEE PAGE 5) = 33.9

$$\text{INTERNAL REFLUX (BPSD)} = \frac{(575,345)}{12.7843} = \underline{\underline{45,004 \text{ BPSD}}}$$

(BELOW A90 DRAW)

CONVERT INTERNAL REFLUX AT TOP OF COLUMN (CALCULATED ON PAGE 27)
TO BPSD

NET OVERHEAD VOLUME RATE = 28650 BPSD

$$\text{INTERNAL REFLUX AT TOP} = \frac{(737,992)}{(301.875)} (28650) = \underline{\underline{70,157 \text{ BPSD}}}$$

DEGREE OF FRACTIONATION.

FOR OUR DESIGN ASSUME THE FOLLOWING ^{FRACTIONATION} TRAYS INSTALLED

TOP TO JET FUEL DRAW = 10

JET FUEL TO DIESEL DRAW = 8

DIESEL TO AGO DRAW = 6

AGO TO FLASH ZONE = 6

FLASH ZONE TO BOTTOM = 4

FOR PUMPAROUNDS ^{DEAL} THREE TRAYS ARE USED FOR
HEAT REMOVAL IN TOWER. THESE ARE CONSIDERED
TO BE ONE FRACTIONATING TRAY.

BASED ON HEAT BALANCE DEVELOPED PREVIOUSLY
DETERMINING GAP/OVERLAP FOR EACH OF THE
PRODUCTS. (UTILIZING PO-171 CHART)
PO-170

$$\text{REFLUX RATIO} = \frac{\text{INTERNAL REFLUX}}{\sum \text{NET PRODUCTS TO DRAW TRAYS}}$$

CUSTOMER ENGINEERING DESIGN SEMINALSUBJECT CLADE UNIT DESIGN
DEGREE OF FRACTIONPAGE NO. 2DATE 10-5-91

PROJECT

BY EN MILLER

COMBINATIONS DEVELOPED BY PALKIE BASED ON COMMERCIAL CLADE TOWER OPERATION CAN BE USED TO DETERMINE THE DEGREE OF FRACTIONATION BETWEEN SIDECUT PRODUCTS AND BETWEEN THE OVERHEAD AND THE FIRST SIDECUT PRODUCT. THE DEGREE OF FRACTIONATION (LINEAN IMPLUITY SPECIFICATION FOR DISTILLATION OF PURE COMPONENT HYDROCARBONS) IS MEASURED IN TERMS OF GAP (OVERLAP) OF THE 95% ASTM DISTILLATION TEMPERATURE OF THE UPPER SIDECUT PRODUCT AND THE 5% ASTM DISTILLATION TEMPERATURE OF THE LOWER SIDECUT. THE GAP HAS BEEN CORRELATED IN TERMS OF A REFLUX RATIO, NUMBER OF TRAYS IN SECTION AND THE DELTA TEMPERATURE DIFFERENCE BETWEEN THE 50% POINT OF THE TOTAL PRODUCTS ~~ENTERING~~ TO UPPER DRAW AND THE 50% POINT OF THE LOWER DRAW PRODUCT. SEE PD-170 AND PD-171. A CALCULATIONAL SHEET IS ATTACHED.

CALCULATE GAP BETWEEN DIESEL AND AGO PRODUCTS

$$\Sigma \text{ PRODUCTS TO DIESEL DRAW} = 28650 + 16550 + 9150 = 54350 \text{ BPSD}$$

$$50\% \text{ OF } \Sigma \text{ PRODUCTS TO DIESEL DRAW} \quad 54350/2 = 27175 \text{ BPSD}$$

$$= 27.18 \text{ VOL}\%$$

$$\text{FROM TBP CURVE TEMPERATURE} = 362^\circ\text{F}$$

$$50\% \text{ OF AGO PRODUCT} \quad T = 660^\circ\text{F} \quad \Delta T = 660 - 362 = \underline{298^\circ\text{F}}$$

$$\text{INTERNAL REFLUX} = 54210 \quad \text{REFLUX RATIO} = \frac{54210}{54350} = 0.997$$

$$\text{NUMBER OF FRACTIONATING TRAYS} = 6 \quad (R)(N) = (6)(0.997) = 5.98$$

$$\text{FROM PD-171} \quad \text{GAP} = 18^\circ\text{F} \quad (\text{CALCULATED})$$

TYPICALLY GAP BETWEEN DIESEL PRODUCT AND AGO PRODUCT IS 0°F .

WHAT IS SIGNIFICANCE OF CALCULATION?

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT DESIGN.DESIGN OF FRACTIONATION.PAGE NO. 3DATE 10-5-91

PROJECT

BY E. N. MILLERCALCULATE GAP BETWEEN JET FUEL AND ~~TOP~~ DIESEL PRODUCTS.

$$\Sigma \text{ PRODUCTS TO JET FUEL DRAW} = 28650 + 16550 = 45200 \text{ BPSD.}$$

$$\begin{aligned} 50\% \text{ POINT } \Sigma \text{ PRODUCTS TO JET FUEL DRAW} &= 45200/2 = 22,600 \text{ BPSD} \\ &= 22.6 \text{ VOL\%} \end{aligned}$$

FROM TBP CURVE TEMPERATURE = 323°F 50% OF DIESEL PRODUCT = 583°F

$$\Delta T = 583 - 323 = \underline{260^{\circ}\text{F.}}$$

$$\text{INTERNAL REFLUX} = \frac{33184}{25626} \text{ BPSD}$$

$$\text{REFLUX RATIO} = \frac{33184}{45200} = \frac{0.734}{0.587}$$

NUMBER OF FRACTIONATING TRAYS = 8

$$R \times N = (8)(0.734) = \underline{5.87}$$

FROM PD-171 GAP = 9°F TYPICALLY GAP BETWEEN ~~CRUDE~~ JET FUEL PRODUCT AND ~~TOP~~ DIESEL PRODUCT IS $10-15^{\circ}\text{F}$ IF WE INCREASE TRAYS TO 10
GAP = 15°F .

$$R \times N = (10)(0.734) = \underline{7.34}$$

CALCULATE GAP BETWEEN UNSTABILIZED NAPHTHA AND JET FUEL

$$\Sigma \text{ PRODUCTS OVERHEAD} = 28650 \text{ BPSD.}$$

$$50\% \text{ POINT } \Sigma \text{ PRODUCTS OVERHEAD} = 14325 \text{ BPSD} = 14.325 \text{ VOL\%}.$$

FROM TBP CURVE TEMPERATURE = 248°F

$$\Delta T = 454 - 248 = \underline{206^{\circ}\text{F}}$$

50% OF JET FUEL PRODUCT = 454°F

$$\text{INTERNAL REFLUX} = 70157 \text{ BPSD}$$

$$\text{REFLUX RATIO} = \frac{70157}{28650} = 2.45$$

NUMBER OF FRACTIONATING TRAYS = 10

$$R \times N = 10 \times 2.45 = 24.5$$

FROM PD-170 GAP = 58°F UOP17-1031A TYPICALLY GAP BETWEEN NAPHTHA AND JET FUEL $20-30^{\circ}\text{F}$

Proj:
By:

CRUDE COLUMN FRACTIONATION

OVHD →

NET OH, BPSD

IR, BPSD

R - IR/NET

NET OH 50% PT, °F

SIDE CUT 50% PT, °F

Δ 50% PT

No. ACTUAL TRAYS

N - No. FRACT. TRAYS

R x N (PD-170)

CALC'D GAP (95-5)

→

Σ PROD TO UPPER DRAW, BPSD

IR, BPSD

R - IR/NET

Σ PROD TO UPPER DRAW 50% PT, °F

LOWER DRAW 50% PT, °F

Δ 50% PT

No. ACTUAL TRAYS

N - No. FRACT. TRAYS

R x N (PD-171)

CALC'D GAP (95-5)

→

Σ PROD TO UPPER DRAW, BPSD

IR, BPSD

R - IR/NET

Σ PROD TO UPPER DRAW 50% PT, °F

LOWER DRAW 50% PT, °F

Δ 50% PT

No. ACTUAL TRAYS

N - No. FRACT. TRAYS

R x N (PD-171)

CALC'D GAP (95-5)

HEAT EXCHANGE NETWORK DESIGN

GENERAL GUIDELINES

1. MINIMUM APPROACH TEMPERATURE = 50°F
2. NUMBER OF SHELLS IN EACH SERVICE = 1-2
3. AVOID CRUDE/TOWER OVERHEAD EXCHANGE
IF USED, OVERHEAD VAPORS MUST LEAVE EXCHANGE ABOVE WATER DEW POINT TEMPERATURE.
(FOR THIS DESIGN WATER DEW POINT $\approx 200^{\circ}\text{F}$)
4. FRACTIONATION REBOILER HEAT REQUIREMENTS SHOULD BE MATCHED FIRST. A CIRCULATING HEAT EXCHANGE STREAM IS ~~RECOMMENDED~~ PREFERRED.
5. DELTA TEMPERATURE FOR PUMPAROUNDS $150-200^{\circ}\text{F}$

FOR OUR DESIGN HOT STREAMS ARE LISTED BELOW:

- | | |
|--------------------------|---|
| 1. CRUDE TOWER OVERHEAD. | $300 \rightarrow 140^{\circ}\text{F}$ |
| 2. JET FUEL PRODUCT. | $408 \rightarrow 100^{\circ}\text{F}$ |
| 3. DIESEL PRODUCT | $530 \rightarrow 120^{\circ}\text{F}$ |
| 4. AGO PRODUCT | $582 \rightarrow 120^{\circ}\text{F}$ |
| 5. REDUCED CRUDE | $675 \rightarrow 175^{\circ}\text{F}$ |
| 6. AGO PUMPAROUND | $Q = 30 \times 10^6 \text{ BTU/HR} \quad 600 \rightarrow ?$ |
| 7. DIESEL PUMPAROUND | $Q = 60 \times 10^6 \text{ BTU/HR} \quad 550 \rightarrow ?$ |

FOR OUR DESIGN COLD STREAMS ARE LISTED BELOW:

1. CRUDE $60^{\circ}\text{F} \rightarrow 715^{\circ}\text{F}$
2. DEBUTANIZER REBOILER $Q = 19.2 \times 10^6 \text{ BTU/HR} \quad 380^{\circ}\text{F}$

FOR PRELIMINARY EVALUATION PURPOSES

- C_p OF PRODUCTS $100-400^{\circ}\text{F}$ RANGE = 0.55
 C_p OF PRODUCTS $400-600^{\circ}\text{F}$ RANGE = 0.78
 C_p OF REDUCED CRUDE $300-700^{\circ}\text{F}$ RANGE = 0.66
 C_p OF REDUCED CRUDE $175-300^{\circ}\text{F}$ RANGE = 0.42
 C_p OF CRUDE $60-260^{\circ}\text{F}$ RANGE = 0.52
 C_p OF CRUDE $260-500^{\circ}\text{F}$ RANGE = 0.625

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGN.PAGE NO. 2DATE 10-8-91

PROJECT

BY E N MILLER.HEAT EXCHANGE NETWORK DESIGN.

$$\begin{aligned}\text{TOTAL HEAT REQUIRED} &= \text{CRUDE} + \text{DEBUTANIZER REBOILER} \\ &= 560.767 + 19.2 = 579.967 \text{ MM BTU/HR}\end{aligned}$$

TOTAL HEAT AVAILABLE.

$$1. \text{ REDUCED CRUDE TO STORAGE} = (527895)(52) = 27.451 \times 10^6 \text{ BTU/HR}$$

$$\Delta Q = \text{REDUCED CRUDE} = (191.626 - 27.451) = 164.175 \times 10^6 \text{ BTU/HR}$$

$$2. \text{ AGO TO STORAGE} = (96815)(29) = 2.808 \times 10^6 \text{ BTU/HR}$$

$$\Delta Q = \text{AGO} = (29.664 - 2.808) = 26.856 \times 10^6 \text{ BTU/HR}$$

$$3. \text{ DIESEL TO STORAGE} = (113460)(29) = 3.290 \times 10^6 \text{ BTU/HR.}$$

$$\Delta Q = \text{DIESEL} = (36.59 - 3.29) = 33.300 \text{ DIESEL.}$$

JET FUEL TO STORAGE

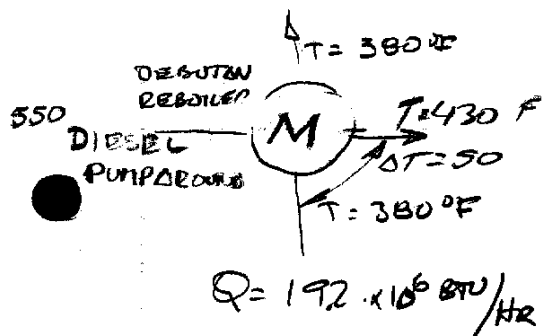
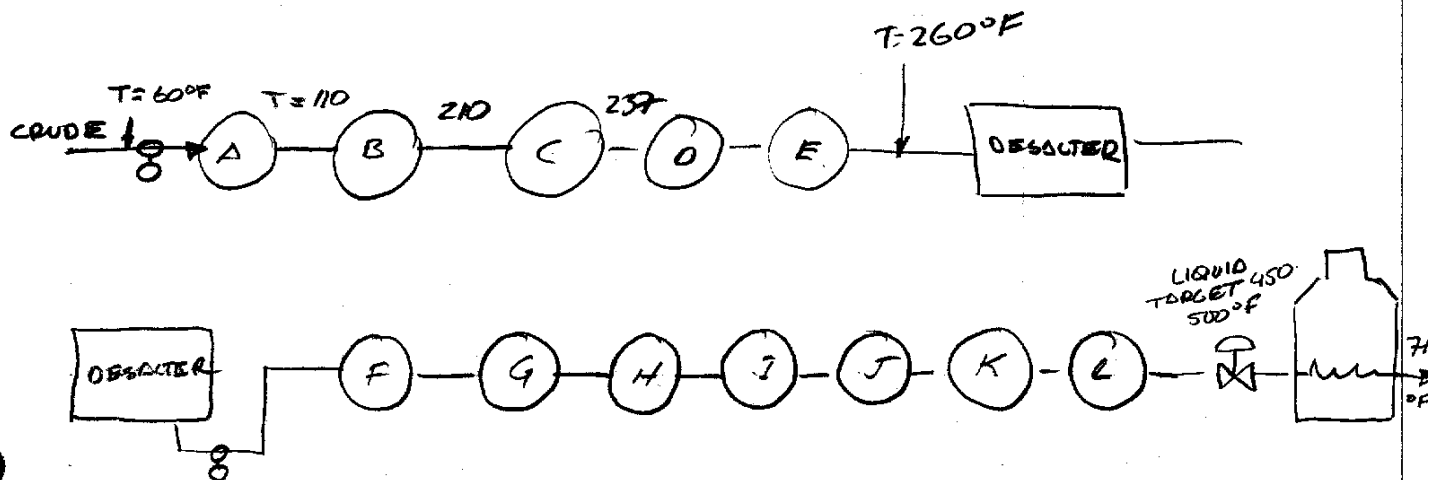
$$+ (38.591 - 3.921) = 34.67 \text{ JF}$$

OVERHEAD

$$\begin{aligned}\text{STEAM OVERHEAD} &= 30.761 - 90.6 \times 31270 \\ &= 33.941\end{aligned}$$

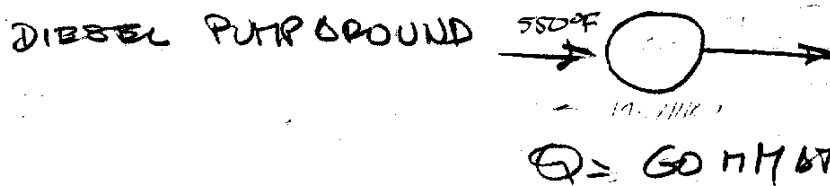
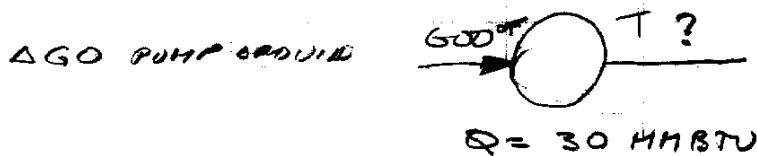
$$\begin{aligned}\text{HYC} &= (79.262 + 113.71) - (19.027 + 44 \times 301375) \\ &= 160.684\end{aligned}$$

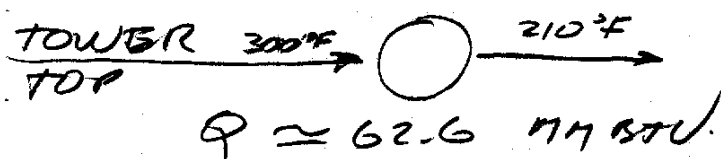
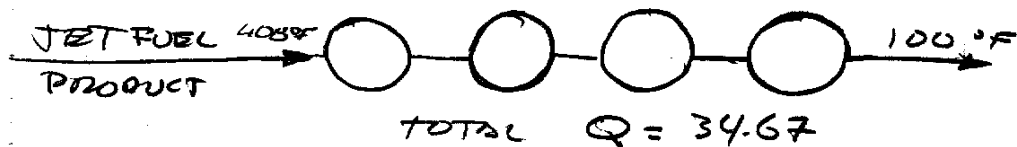
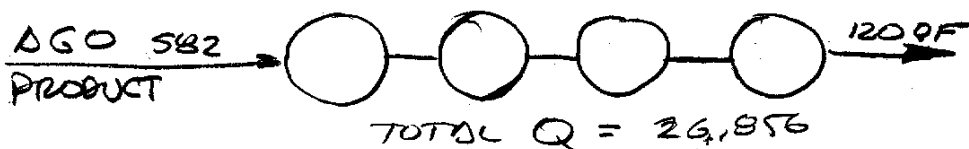
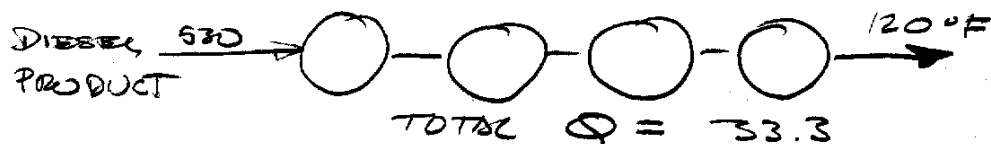
<u>TOTAL HEAT AVAILABLE</u>	164.175	REDUCED CRUDE
	26.856	AGO STORAGE
	33.300	DIESEL STORAGE
	34.67	JET FUEL
	30.00	AGO PUMP AROUND
	60.00	DIESEL PUMP AROUND
	+ 194.525	OVERHEAD
	<u>543.526.</u>	



HEN GENERAL GUIDELINES
 ΔT APPROACH 50°F MIN
 MATCH FRACTIONATOR REBOILER FIRST
 PREFER PUMP AROUND
 MAX ΔT PUMP AROUND ≈ 150-200°F

COOLING TRAIN





CRUDE $60^{\circ}\text{F} - 260^{\circ}\text{F}$

$$M_{CP} = 0.51 \times 1235580 = 6.3014 \times 10^5$$

CRUDE $260 - 500^{\circ}\text{F}$

$$M_{CP} = 7.72 \times 10^5$$

REDUCED CRUDE $700 - 300$

$$M_{CP} =$$

REDUCED CRUDE $300 - 175$

$$M_{CP} =$$

600 - 400

AGO $m_{cp} = 0.755 \times 10^5$

DIESEL $m_{cp} = 0.885 \times 10^5$

JET-FUEL $m_{cp} = 1.529 \times 10^5$

400 - 100

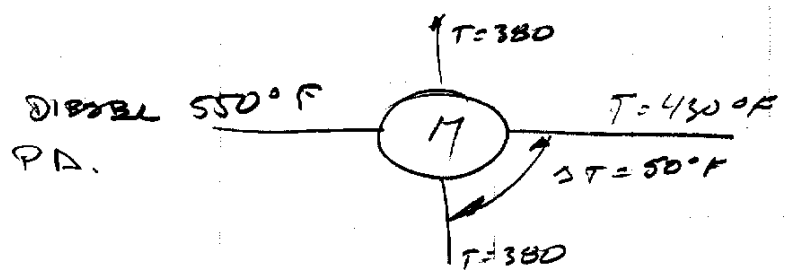
AGO $m_{cp} = 0.533 \times 10^5$

DIESEL $m_{cp} = 0.624 \times 10^5$

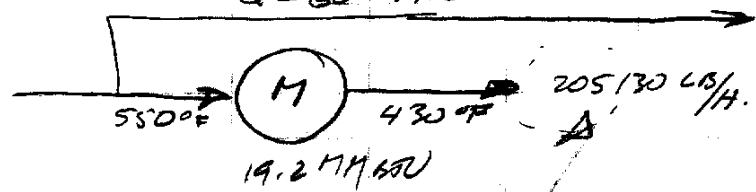
JET FUEL $m_{cp} = 1.078 \times 10^5$

1° PARA EL RECORRER DELEGAMOS UNA LONGITUD EN UN RANGO DE TEMPERATURA ADECUADO

EN ESTE CASO DIESEL PUMPROUND.



DIESEL PUMPROUND $Q = 60 - 19.2 = 40.8 \text{ MMSTU}$



$Q = m_{cp}(\Delta T)$

$19.2 \times 10^6 = m_{cp} \cdot 0.78 \cdot 120$

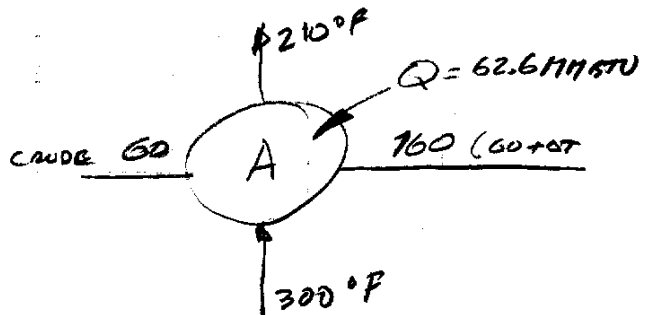
$m = 205130 \text{ LB/H.}$

- A CONTINUACION CORRIENTE DE MAYOR TEMP. POR EL CIRCUITO DE CARGA. DESPUES DE LA DESACUORDA. DE ACUERDO A SU TEMPERATURA.

EXCHG STREAM
 L = CRUDO REDUCIDO
 K = 560 PD.
 J = 560 PROD.
 I = DIESEL PL
 H = DIESEL PROD.

ANTES DE LA DESACUORDA
 POR EL LADO PROD.

A = TOWER TOP
 COVERED

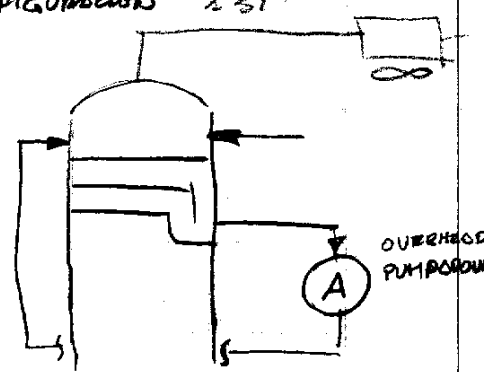


$$\Delta T = \frac{Q}{M C_p}$$

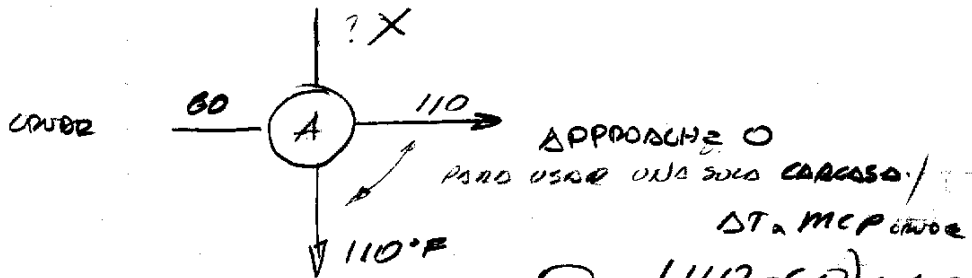
$$(\Delta T = \frac{62.6 \times 10^6 \text{ MMSTU}}{6.3 \times 10^5} \cdot \frac{\text{TOP TOW.}}{M C_p (\text{CRUDE})})$$

$$\Delta T = 100.$$

PODS ESTE CASO SE TENDRÁ
 UNA CONFIGURACION ASI

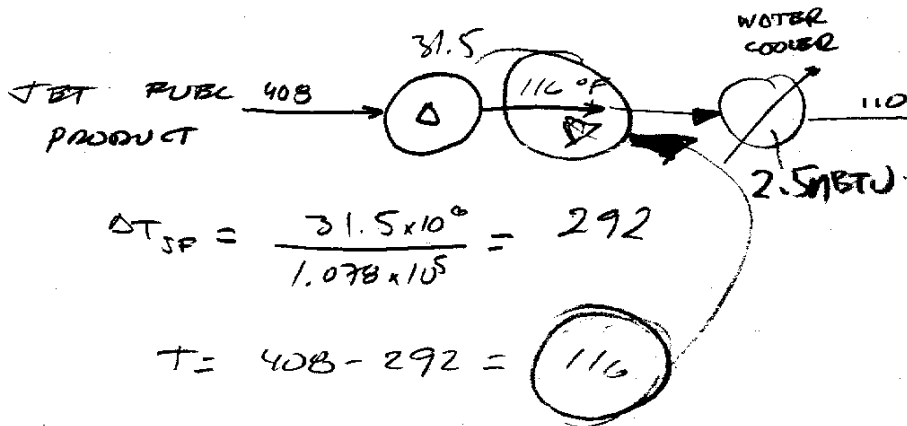


USANDO JET FUEL
A EN LUGAR DE OVERHEAD. PRA



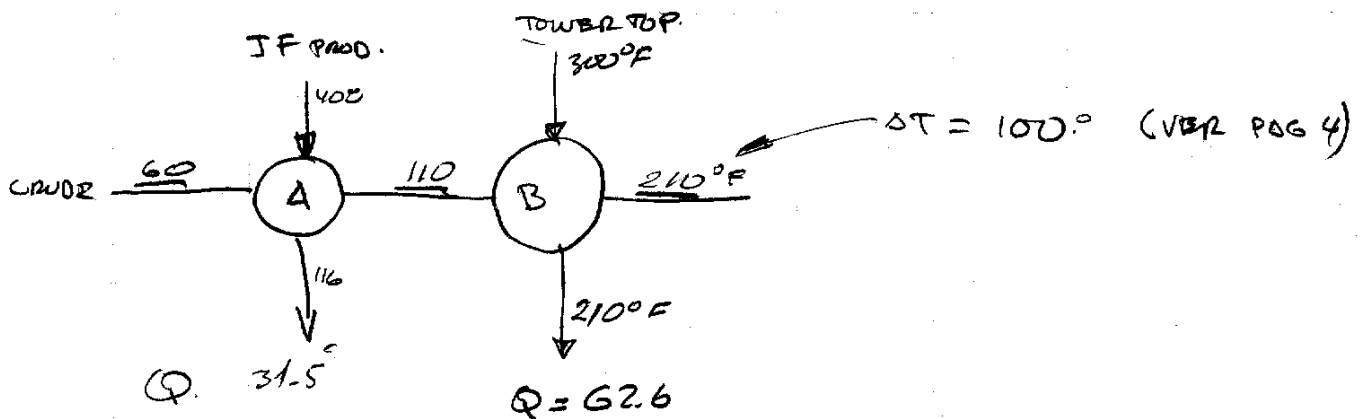
$$Q = (110 - 60) \times 6.3 \times 10^5$$

$$Q = 31.5 \times 10^6 \text{ BTU} \approx \text{IGUAL A LO CARBON TERMINAL DISPONIBLE EN J-F}$$

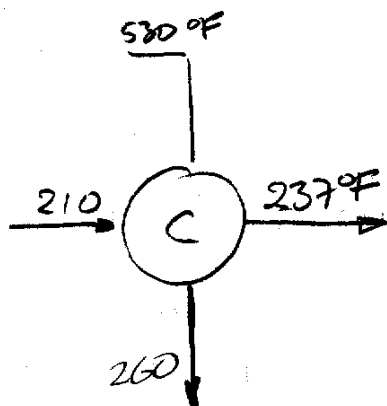


$$\Delta T_{JP} = \frac{31.5 \times 10^6}{1.078 \times 10^5} = 292$$

$$T = 408 - 292 = 116$$



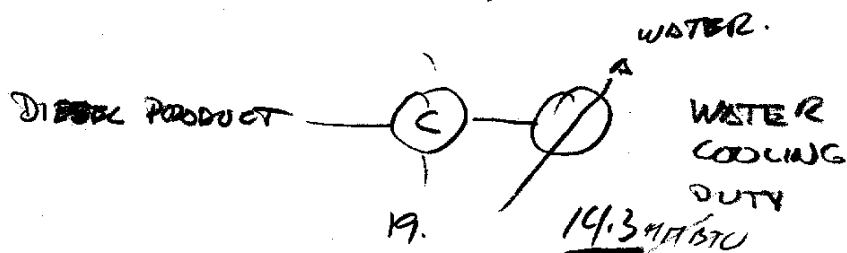
PARA EXCH C. SELECCIONAR DIESEL PRODUCT



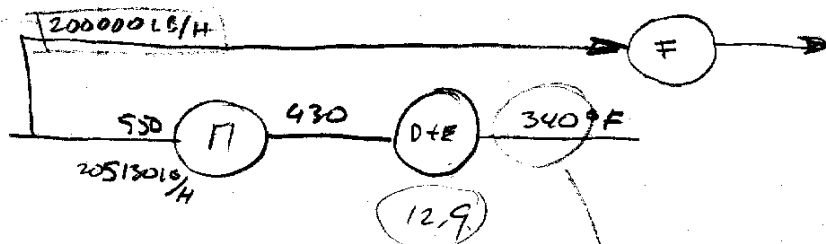
530. 260
250 210

$$Q = (530 - 260) \times 0.7 \times 10^5$$

$$Q = 19.10^6 \text{ BTU/Hr.}$$

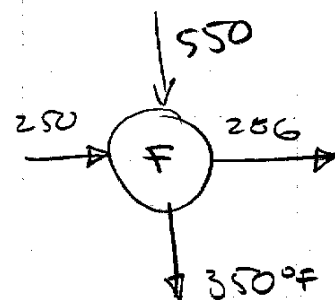


PARA D & E SE REQUIERE 12.9 MBTU/HRS. SE UTILIZAN UN SOLO INTERCAMBIADOR, SE DEBERIA DECIDIR QUE CORRIENTE USAR PARA ESTE. SE RECOMIENDA LA CORRIENTE DE DIESEL PROVENIENTE DEL RESERVOIR DE LA DEBUT



$$\Delta T = \frac{12.9}{205180 \times 0.7} = 90$$

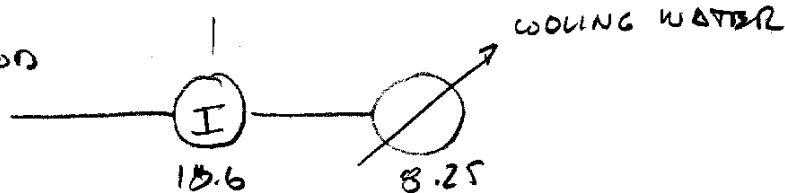
$$M = \frac{27.9}{(550 - 350) 0.7} = 200000 \text{ LB/H.}$$

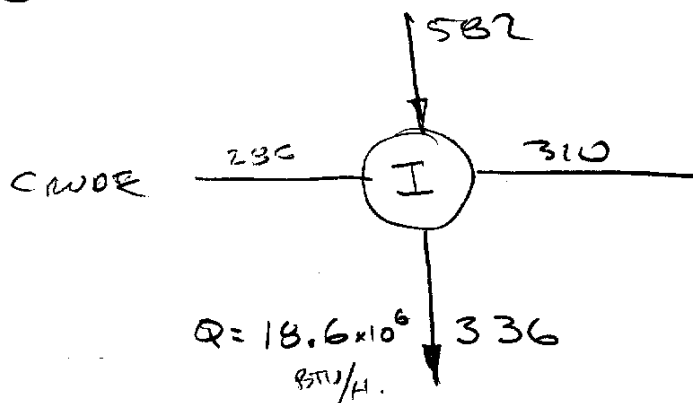


AGO PROD

$$Q = mcp (582 - 336) = 18.6$$

AGO PROD



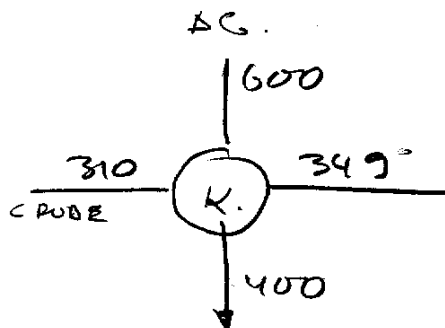


$$\Delta T = \frac{18.6 \times 10^6}{7.72 \times 10^5}$$

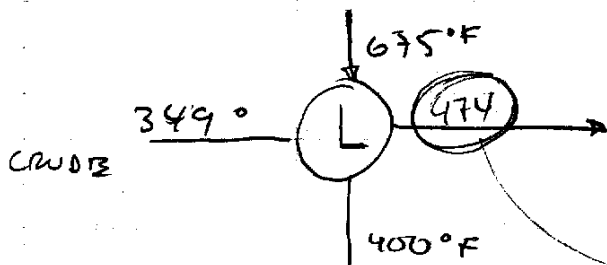
AGO PUMP



$$Q = 30 \times 10^6 \text{ BTU/H.}$$



REDUCED CRUDE



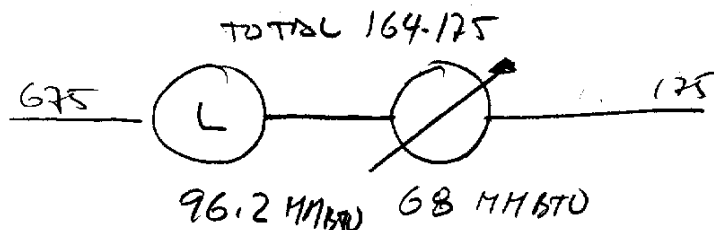
$$Q = 3.5 \times 10^5 (275) = 96.2 \text{ MM BTU/H.}$$

$$Q = 96.2 \text{ MM BTU/H.}$$

$$\Delta T = \frac{96.2 \times 10^6}{7.72 \times 10^5} = 124.6^\circ \text{F}$$

REDUCED CRUDE

675	474
400	349

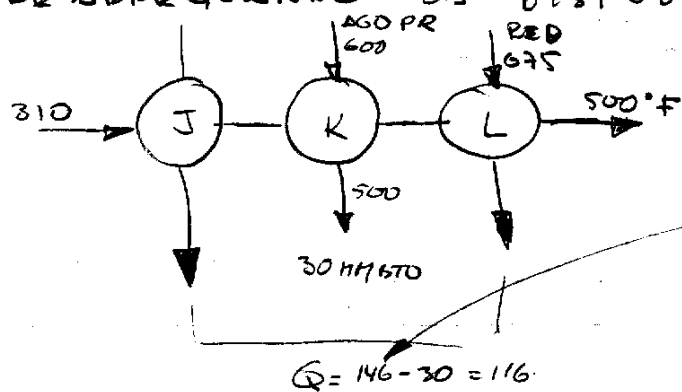


WATER COOLING TOWER

14.3
2.5
18.256
68.
93.00 MM BTU

HEAT REMOVED 126.03
27.09
18.60
30.00
96.20
298.7 MM BTU ~ 75%
93.
391.7

SE PUEDE RECUPERAR CALOR DEL CRUDO REDUCIDO
68 MM BTU a 400°F
REARREGLANDO LA DISPOSICION



REDA CRUDO DE 310 a 500
NECESITO

$$Q = 7.72 \times 10^5 (500 - 310)$$

$$Q = 146. \text{ MM BTU}$$

HEAT EXCHANGE NETWORK DESIGN

GENERAL GUIDELINES

1. MINIMUM APPROACH TEMPERATURE = 50°F
2. NUMBER OF SHELLS IN EACH SERVICE = 1-2
3. AVOID CRUDE/TOWER OVERHEAD EXCHANGE
IF USED, OVERHEAD VAPORS MUST LEAVE
EXCHANGE ABOVE WATER DEW POINT TEMPERATURE.
(FOR THIS DESIGN WATER DEW POINT $\approx 200^{\circ}\text{F}$)
4. FRACTIONATION REBOILER HEAT REQUIREMENTS
SHOULD BE MATCHED FIRST. A CIRCULATING
HEAT EXCHANGE STREAM IS ~~RECOMMENDED~~ PREFERRED.
5. DELTA TEMPERATURE FOR PUMPAROUNDS $150-200^{\circ}\text{F}$

FOR OUR DESIGN HOT STREAMS ARE LISTED BELOW:

1. CRUDE TOWER OVERHEAD	$300 \rightarrow 140^{\circ}\text{F}$	
2. JET FUEL PRODUCT	$408 \rightarrow 100^{\circ}\text{F}$	
3. DIESEL PRODUCT	$530 \rightarrow 120^{\circ}\text{F}$	
4. AGO PRODUCT	$582 \rightarrow 120^{\circ}\text{F}$	
5. REDUCED CRUDE	$675 \rightarrow 175^{\circ}\text{F}$	
6. AGO PUMPAROUND	$Q = 30 \times 10^6 \text{ BTU/HR}$	$600 \rightarrow ?$
7. DIESEL PUMPAROUND	$Q = 60 \times 10^6 \text{ BTU/HR}$	$550 \rightarrow ?$

FOR OUR DESIGN COLD STREAMS ARE LISTED BELOW:

1. CRUDE	$60^{\circ}\text{F} \rightarrow 715^{\circ}\text{F}$	
2. DEBUTANIZER REBOILER	$Q = 19.2 \times 10^6 \text{ BTU/HR}$	380°F

FOR PRELIMINARY EVALUATION PURPOSES

- C_p OF PRODUCTS $100-400^{\circ}\text{F}$ RANGE = 0.55
 C_p OF PRODUCTS $400-600^{\circ}\text{F}$ RANGE = 0.78
 C_p OF REDUCED CRUDE $300-700^{\circ}\text{F}$ RANGE = 0.66
 C_p OF REDUCED CRUDE $175-300^{\circ}\text{F}$ RANGE = 0.42
 C_p OF CRUDE $60-260^{\circ}\text{F}$ RANGE = 0.52
 C_p OF CRUDE $260-500^{\circ}\text{F}$ RANGE = 0.625

HEAT EXCHANGE NETWORK DESIGN.

$$\begin{aligned}\text{TOTAL HEAT REQUIRED} &= \text{CRUDE} + \text{DEBTANIZER REBILDER} \\ &= 560.767 + 19.2 = 579.967 \text{ MM BTU/HR}\end{aligned}$$

TOTAL HEAT AVAILABLE.

$$1. \text{ REDUCED CRUDE TO STORAGE} = (527895)(52) = 27.451 \times 10^6 \text{ BTU/HR}$$

$$\Delta H = \text{REDUCED CRUDE} = (191.626 - 27.451) = 164.175 \times 10^6 \text{ BTU/HR}$$

$$2. \text{ AGO TO STORAGE} = (96815)(29) = 2.808 \times 10^6 \text{ BTU/HR}$$

$$\Delta H = \text{AGO} = (29.664 - 2.808) = 26.856 \times 10^6 \text{ BTU/HR}$$

$$3. \text{ DIESEL TO STORAGE} = (113460)(29) = 3.290 \times 10^6 \text{ BTU/HR.}$$

$$\Delta H = \text{DIESEL} = (36.590 - 3.29) = 33.3 \times 10^6 \text{ BTU/HR}$$

$$4. \text{ JET FUEL TO STORAGE} = (196035)(20) = 3.921 \times 10^6 \text{ BTU/HR.}$$

$$\Delta H = \text{JET FUEL} = (38.591 - 3.921) = 34.67 \times 10^6 \text{ BTU/HR.}$$

$$5. \text{ DIESEL PUMPAROUND} \quad \Delta H = 60 \times 10^6 \text{ BTU/HR.}$$

$$6. \text{ AGO PUMPAROUND} \quad \Delta H = 30 \times 10^6 \text{ BTU/HR.}$$

FOR CRUDE TOWER OVERHEAD LIMIT HEAT REMOVAL TO CONDENSER
SIDE OUTLET TEMPERATURE TO 210°F (ABOUT 10°F ABOVE WATER DEW POINT)

FOR ESTIMATING PURPOSES, WE WILL ASSUME HYDROCARBON CONDENSING
CURVE IS LINEAR FROM 300°F TO 140°F. THIS IS NOT EXACTLY TRUE
AND WE WILL NEED TO CONSTRUCT OVERHEAD CONDENSING CURVE
LATER.

$$\text{SPECIFIC ENTHALPY OF HC AT 210°F} = 143 \text{ BTU/LB.}$$

$$\text{FRACTION CONDENSED} = \frac{(300 - 210)}{(300 - 140)} = \frac{90}{160} = 0.5625 \quad \begin{array}{l} \text{VAPOR ENTHALPY AT 210} = 217 \text{ BTU/LB} \\ \text{LIQ ENTHALPY AT 210} = 85 \end{array}$$

$$\Delta H \text{ HYDROCARBON} = (301375 + 432438)(263 - 143) = 88.06 \times 10^6 \text{ BTU.}$$

$$\Delta H \text{ STEAM} = (31270)(1162.8 - 1122.3) = 1.266 \times 10^6 \text{ BTU/HR}$$

$$\text{TOTAL} = \underline{\underline{89.326 \times 10^6 \text{ BTU/HR}}}$$

HEAT EXCHANGER NETWORK

TO MAKE INITIAL NETWORK DESIGN SIMPLER WE WILL CALCULATE THE PRODUCT OF MASS FLOW $\times C_p$ FOR EACH OF STREAMS.

FOR TEMPERATURE RANGE 60-260°F FOR CRUDE.

$$mC_p = (1235580)(0.51) = 6.3 \times 10^5$$

FOR TEMPERATURE RANGE 260-500°F FOR CRUDE.

$$mC_p = (1235580)(0.625) = 7.72 \times 10^5$$

FOR TEMPERATURE RANGE 100-400°F FOR PRODUCTS (AGO, DIESEL, JET FUEL)

$$\text{AGO } mC_p = (96815)(0.55) = 0.53 \times 10^5$$

$$\text{DIESEL } mC_p = (113460)(0.55) = 0.624 \times 10^5$$

$$\text{JET FUEL } mC_p = (196035)(0.55) = 1.078 \times 10^5$$

FOR TEMPERATURE RANGE 400-600°F FOR PRODUCTS (AGO, DIESEL, JET FUEL)

$$\text{AGO } mC_p = (96815)(0.78) = 0.751 \times 10^5$$

$$\text{DIESEL } mC_p = (113460)(0.78) = 0.885 \times 10^5$$

$$\text{JET FUEL } mC_p = (196035)(0.78) = 1.53 \times 10^5$$

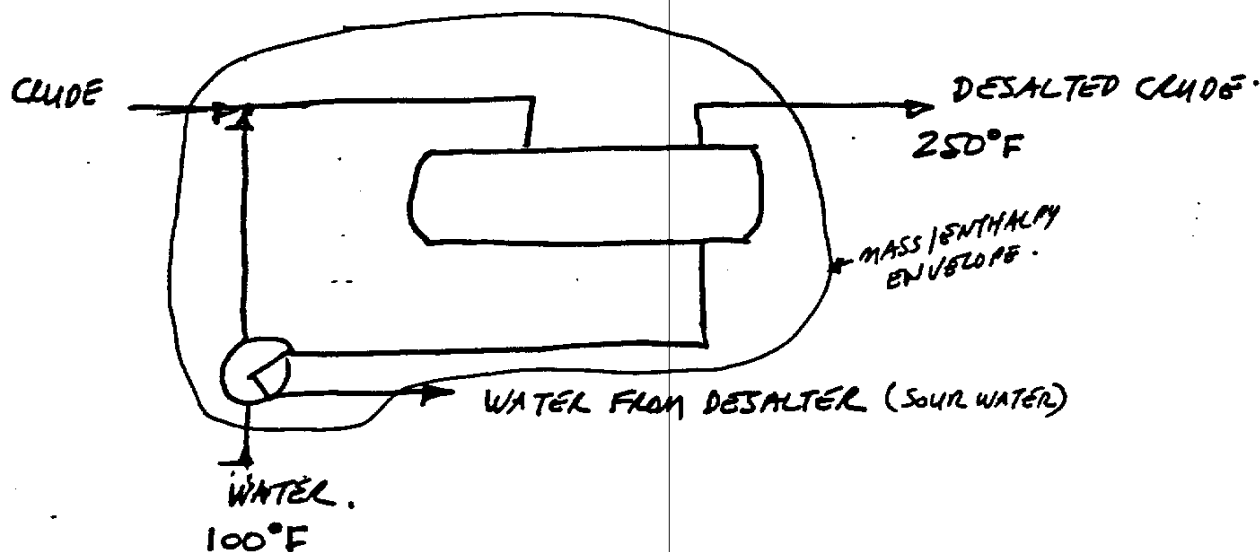
FOR TEMPERATURE RANGE 175-300°F FOR REDUCED CRUDE.

$$mC_p = (527845)(0.42) = 2.217 \times 10^5$$

FOR TEMPERATURE RANGE 300-700°F FOR REDUCED CRUDE

$$mC_p = (527845)(0.66) = 3.484 \times 10^5$$

(DETERMINE CRUDE TEMPERATURE TO DESALTER.



SET COLD END APPROACH ON EXCHANGE AT 50°F

CALCULATE CRUDE TEMPERATURE TO DESALTER.

① DESIGN BASIS WATER RATE = 5 VOL% ON CRUDE = 5000 BPSD
 MAKEUP WATER = $\frac{(5000)(42)(834)}{(24)} = 72975 \text{ LB/HR}$

② IGNORE EFFECT OF SALT FROM CRUDE IN WATER FROM DESALTER AND IGNORE WATER INTO DESALTED CRUDE.

③ WATER FROM DESALTER = 72975 LB/HR TEMP = 100 + 50 = 150°F

MASS/ENTHALPY BALANCE

	IN LB/HR	OUT LB/HR	IN MMBTU/HR	ENTHALPY BTU/LB	OUT MMBTU/HR
CRUDE	1235580		X	97	
DESALTED CRUDE		1235580			119.851
WATER	72975		2.919	40	
SOUR WATER		72975		90	6.568
	1308555	1308555	X + 2.19		126.419

SPECIFIC ENTHALPY OF CRUDE IN = $\frac{(126.419 - 2.919) \times 10^6}{1235580} = 99.95 \text{ BTU/LB}$

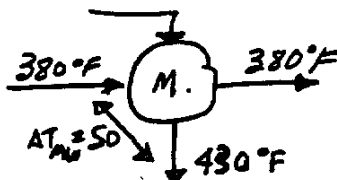
FROM ENTHALPY CURVE TEMPERATURE = 256°F

FOR OUR DESIGN WE WILL SET CRUDE TEMPERATURE TO DESALTER
 AT 260°F

HEAT EXCHANGE NETWORK.

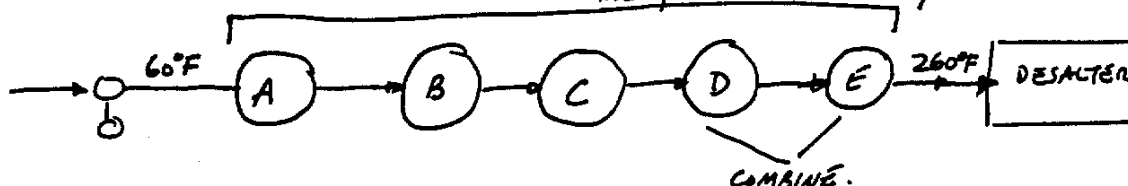
COLD STREAMS REQUIRE HEATING.

- 1 DEBUTANIZER REBOILER
 $Q = 19.2 \times 10^6 \text{ BTU/HR.}$

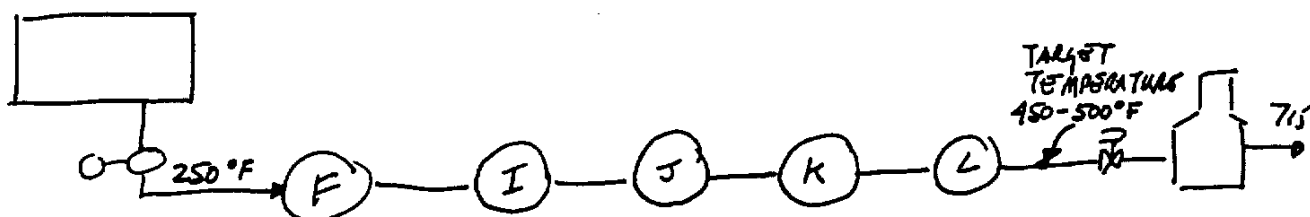


REMEMBER FOR MTD CALCULATION NO CREDIT TAKEN FOR TEMPERATURE RISE ACROSS REBOILER ON PROCESS SIDE.

- 2 CRUDE TO DESALTER TOTAL $Q = 126.03 \times 10^6 \text{ BTU/HR.}$

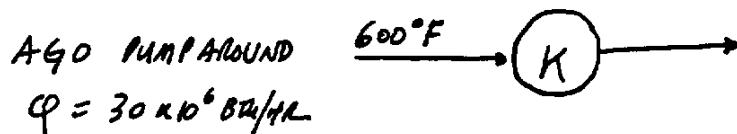


3. CRUDE FROM DESALTER TO HEATER.

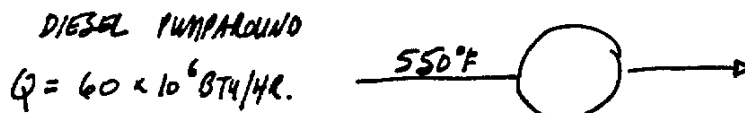


HOT STREAMS AVAILABLE FOR EXCHANGE AND/OR COOLING.

1



2

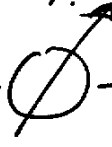


CUSTOMER ENGINEERING DESIGN SEMINAR.
 SUBJECT CRUDE UNIT DESIGN.

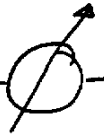
PAGE NO. 6
 DATE 10-10-91
 PROJECT
 BY EN MILLER.

HEAT EXCHANGER NETWORK.

HOT STREAMS AVAILABLE FOR EXCHANGE AND/OR COOLING.

3. RESID PRODUCT 675°F (L) —  —  175°F
 TOTAL Q = $164.175 \times 10^6 \text{ BTU/HR}$

4. AGO PRODUCT 502°F (I) —  120°F
 TOTAL Q = $26.86 \times 10^6 \text{ BTU/HR}$

5. DIESEL PRODUCT 530°F (C) —  120°F
 TOTAL Q = $33.3 \times 10^6 \text{ BTU/HR}$

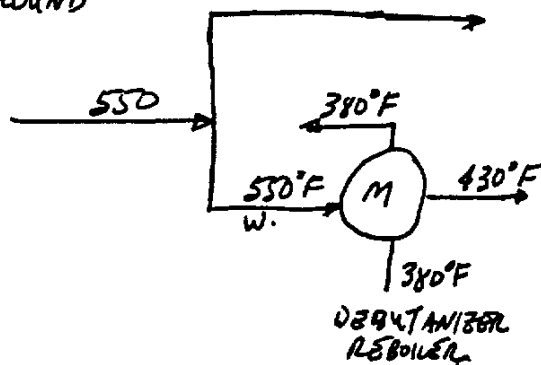
6. JET FUEL 408°F (A) —  100
 TOTAL Q = $34.67 \times 10^6 \text{ BTU/HR}$

7. COLUMN OVERHEAD 300°F (B) — ?
 Q $\approx 62.6 \times 10^6 \text{ BTU/HR}$

HEAT EXCHANGE NETWORK.

FOR OPERATING FLEXIBILITY WE PREFER TO MATCH REBOILERS WITH PUMP AROUND STREAMS. IN THIS CASE WE WILL USE DIESEL PUMP AROUND AS HEATING MEDIA FOR DEBUTANIZER REBOILER. LETS SPLIT PUMP AROUND FLOW BETWEEN TWO SERVICES.

DIESEL PUMP AROUND



- ① SET PUMP AROUND OUTLET TEMPERATURE AT $430^{\circ}\text{F} = 380 + 50 = 430^{\circ}\text{F}$
- ② DUTY 19.2×10^6 BTU/HR.
- ③ CALCULATE PUMP AROUND FLOW THROUGH THIS BRANCH

$$W = \frac{Q}{C_p \Delta T} = \frac{19.2 \times 10^6}{(0.74)(550 - 430)} = 205130 \text{ LB/HR}$$

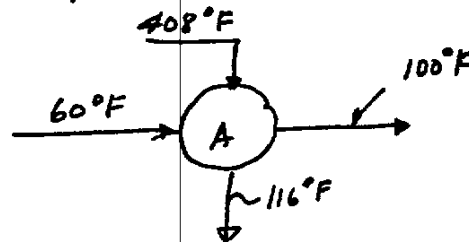
- ④ CHECK EXCHANGER APPROACHES

HOT SIDE	SSD	→	430
COLD SIDE	380	→	380
	170		50

AS A GENERAL APPROACH TO MATCHING REMAINING HEAT SOURCES (HOT STREAMS IN THE CRUDE EXCHANGE TRAIN (COLD STREAM), THE HIGHER LEVEL HEAT WILL BE USED AT THE HOT END OF THE CRUDE EXCHANGE TRAIN AND VICE VERSA. I.E. EXCHANGE HOT AGAINST HOT AND COLD AGAINST COLD. SINCE WE NEED TO HEAT THE CRUDE TO THE DESALTER OPERATING TEMPERATURE USING ONLY EXCHANGE, WE WILL MATCH THE COLD STREAMS FIRST.

HEAT EXCHANGE NETWORK.

IN THE CLASS WE FIRST TRIED TO MATCH THE TOWER OVERHEAD AGAINST THE COLD CRUDE (EXCHANGER A) WITHOUT MUCH SUCCESS. WE THEN DECIDED TO MATCH JET FUEL PRODUCT AGAINST CRUDE IN EXCHANGER A.



TOTAL HEAT AVAILABLE IN JET FUEL PRODUCT STREAM IS 34.67×10^6 BTU/HR. LET'S TRY TO USE IT ~~FOR~~ OUR APPROACH TEMPERATURE CONSTRAINT. I.E. JET FUEL OUTLET TEMPERATURE OF 110°F

$$Q = (1.078 \times 10^5)(408 - 110) = 32.12 \times 10^6 \text{ BTU/HR}$$

THEN CRUDE TEMPERATURE FROM EXCHANGE WILL BE

$$T = 60 + \frac{32.12 \times 10^6}{6.3 \times 10^5} = 60 + 51.0 = 111^\circ\text{F}$$

CHECK APPROACH TEMPERATURES

WE HAVE TEMPERATURE CROSS.

408	110	HOT
116	60	COLD
298	50	ΔT OK.

IN THE CLASS WE DECIDED TO USE 110°F AS CRUDE OUTLET TEMPERATURE WHICH GAVE A Q FOR THIS EXCHANGER OF 31.5 MM BTU/HR . THE JET FUEL PRODUCT OUTLET = 116°F

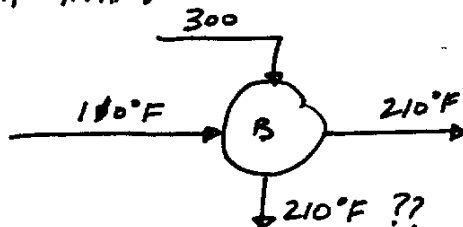
CALCULATE COOLER (ASSUME WATER) DUTY FOR JET FUEL PRODUCT



$$\text{DUTY} = 31.5 + \text{DUTY} = 3.17 = 34.67 \times 10^6 \text{ BTU/HR TOTAL OK.}$$

HEAT EXCHANGE NETWORK

NOW WE WILL MATCH THE TOWER OVERHEAD AGAINST EXCHANGER B IN CRUDE PREHEAT TRAIN.



USE ALL OF DUTY IN OVERHEAD $Q = 62.6 \times 10^6$ BTU/HR. CALCULATE CRUDE SIDE OUTLET TEMPERATURE.

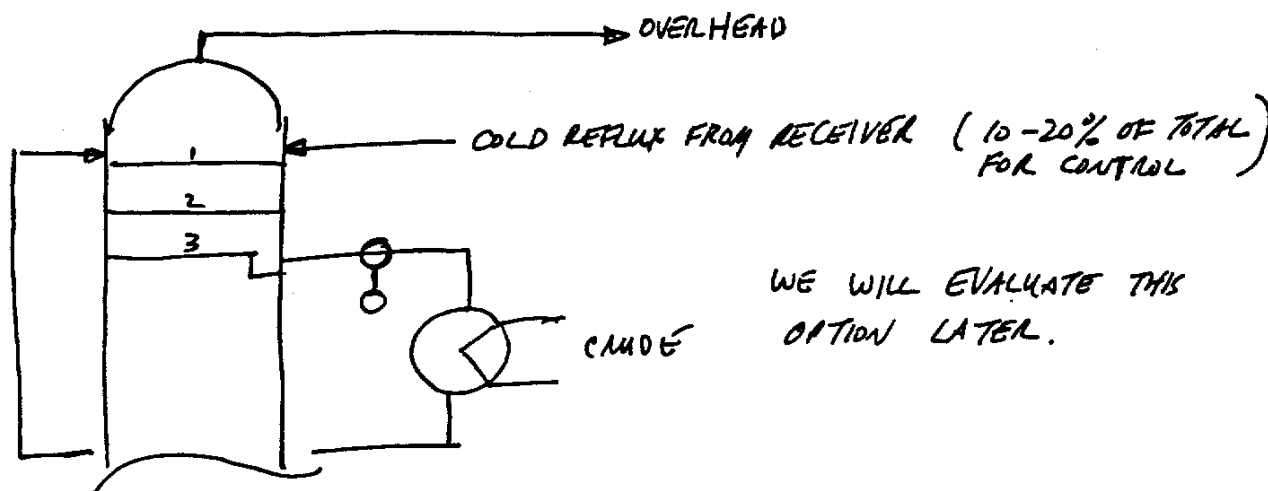
$$T = 180 + \frac{62.6 \times 10^6}{6.3 \times 10^5} = 180 + 99.4 = 210^\circ\text{F}$$

IN THE CLASS OVERHEAD OUTLET TEMP FOR THE DUTY WAS ESTIMATED AT 210°F. THIS IS NOT CORRECT (SEE CALCS ON PAGE 2)

CHECK APPROACH TEMPERATURES
ONE SHELL.

HOT	300 - 210	
COLD	$\frac{210 - 180}{90}$	$\frac{110}{100}$ OK

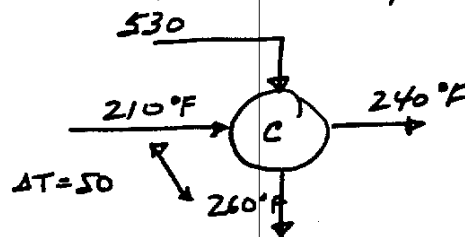
BECAUSE OF LOW AVAILABLE ΔP IN THE OVERHEAD CIRCUIT AND PROBLEMS WITH CRUDE GOING STRAIGHT TO THE OVERHEAD PRODUCT IN CASE WHERE EXCHANGER LEAKS, SOME DESIGNER AVOID USE OF THIS EXCHANGER IN CRUDE PREHEAT TRAIN. MOST PEOPLE PREFER TO RECOVER OVERHEAD HEAT WITH A PUMPAROUND EXCHANGE AT TOP OF TOWER.



WE WILL EVALUATE THIS OPTION LATER.

HEAT EXCHANGE NETWORK.

NOW EXCHANGE DIESEL PRODUCT AGAINST CRUDE IN EXCHANGER C



- ① SET DIESEL PRODUCT FROM EXCHANGER AT 260°F. MINIMUM COLD SIDE APPROACH TEMPERATURE: $210 + 50 = 260$.

- ② CALCULATE DUTY. $Q = (0.7 \times 10^5)(530 - 260) = 19 \times 10^6 \text{ BTU/HR.}$
NOTE WE USED AN INTERMEDIATE VALUE FOR mC_p

- ③ CALCULATE CRUDE SIDE OUTLET TEMPERATURE

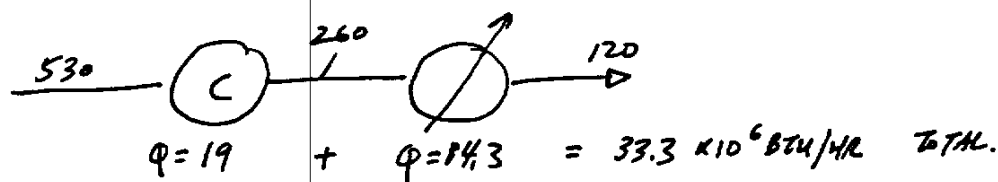
$$T = 210 + \frac{19 \times 10^6}{63 \times 10^5} = 210 + 30 = 240^\circ\text{F}$$

IN THE CLASS WE SAY THIS TEMP WAS 237°F PLS CHECK

- ④ CHECK APPROACH TEMPERATURES.

HOT	530	→	260	
COLD	240		210	
	290		50	OK.

- ⑤ CALCULATE COOLER DUTY FOR DIESEL PRODUCT TO STORAGE.



CALCULATE REMAINING HEAT REQUIRED IN CRUDE EXCHANGE TRAIN TO DESALTER

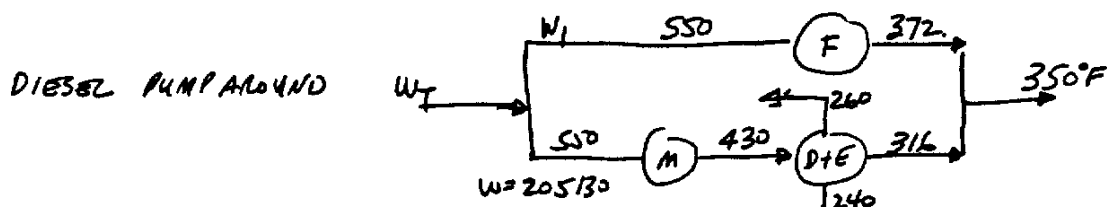
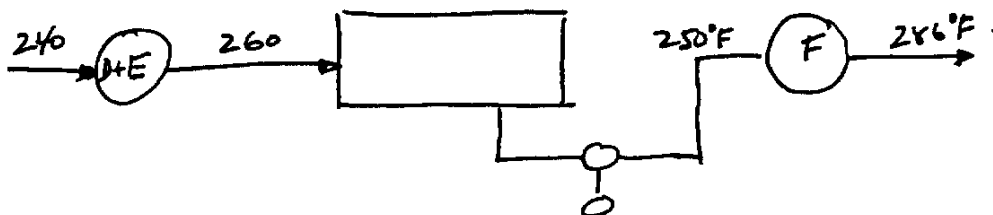
$$Q_{\text{REMAINING}} = 126.03 - 31.5 - 62.6 - 19 = 12.9 \times 10^6 \text{ BTU/HR}$$

LET'S USE ONE EXCHANGE SERVICE TO PICK UP THIS HEAT.

THEREFORE D AND E BECAME ONE EXCHANGER (D+E)

HEATEXCHANGE NETWORK

TO PROVIDE SOME FLEXIBILITY IN THE OPERATING TEMPERATURE OF THE DESALTER WE WILL USE THE DIESEL PUMPAROUND STREAM IN THE CRUDE EXCHANGE TRAIN IMMEDIATELY UPSTREAM AND DOWN STREAM OF THE DESALTER



LOOKING AT EXCHANGER (D+E) $Q = 12.9 \times 10^6 \text{ BTU/HR}$

$$\text{CALCULATE TEMP OF PUMPAROUND FROM EXCH} = 430 - \frac{12.9 \times 10^6}{(205130)(0.55)} = 430 - 114 = 316^\circ\text{F}$$

CHECK APPROACH TEMPERATURES FOR (D+E)

$$\begin{array}{l} \text{HOT } 430 \rightarrow 316 \\ \text{COLD } \frac{260}{90} \leftarrow \frac{240}{76} \end{array} \quad \text{OK}$$

TOTAL DIESEL PUMPAROUND $Q = 60 \times 10^6 \text{ BTU/HR}$

HEAT AVAILABLE FOR EXCH F = $60 - 19.2 - 12.9 = 27.9 \times 10^6 \text{ BTU/HR}$

$$\text{CALCULATE CRUDE SIDE OUTLET TEMP} = 250 + \frac{27.9 \times 10^6}{7.72 \times 10^6} = 250 + 36 = 286^\circ\text{F}$$

SET PUMPAROUND RETURN TEMP AT 350°F CALCULATE TOTAL PUMPAROUND FLOW

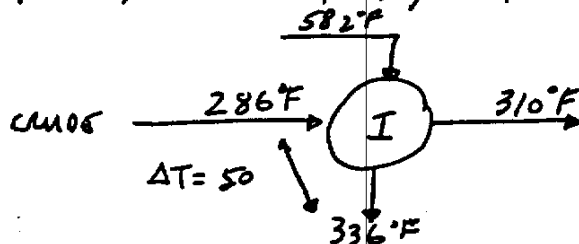
$$W_T = \frac{60 \times 10^6}{0.7 (550 - 350)} = 428570 \text{ LB/HR}$$

$$W_1 = W_T - W = 428570 - 205130 = 223440 \text{ LB/HR}$$

$$\text{TEMP FROM EXCHANGE (F)} = 550 - \frac{27.9 \times 10^6}{(0.7)(223440)} = 550 - 178 = 372^\circ\text{F}$$

HEAT EXCHANGE NETWORK

EXCHANGE AGO PRODUCT AGAINST CRUDE IN EXCHANGER I.



$$\text{CALCULATE AGO OUTLET TEMP} = 286 + 50 = 336^{\circ}\text{F}$$

$$\text{CALCULATE DUTY } Q = (6.756 \times 10^6) (582 - 336) = 18.6 \times 10^6 \text{ BTU/HR}$$

$$\begin{aligned} \text{CALCULATE CRUDE SIDE OUTLET TEMP} &= 286 + \frac{18.6 \times 10^6}{7.72 \times 10^5} = 286 + 24 \\ &= 310^{\circ}\text{F} \end{aligned}$$

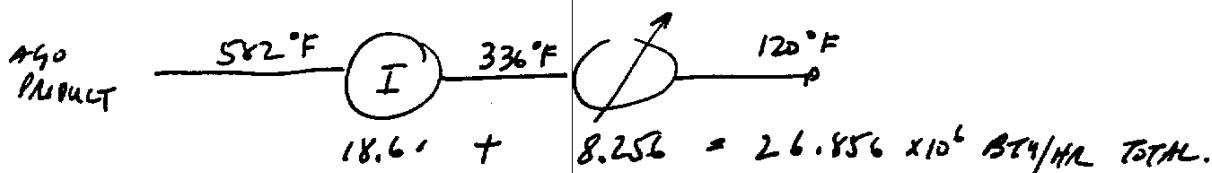
CHECK APPROACH TEMPERATURES

NOW DO

$$\text{HOT} = 582 \rightarrow 336$$

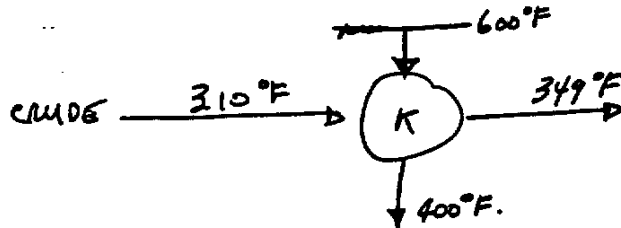
$$\frac{310 - 286}{272} = \frac{24}{50} \quad \text{OK}$$

NOW CALCULATE COOLER DUTY FOR AGO PRODUCT EXCHANGE TRAIN.



HEAT EXCHANGE NETWORK.

NOW USE A50 PUMPAROUND FOR EXCHANGER (K) IN CRUDE TRAIN. BYPASSING EXCHANGER (J) FOR THE MOMENT.



PUMPAROUND DUTY IS 30 MM BTU/HR

CALCULATE CRUDE SIDE OUTLET TEMPERATURE FROM EXCHANGER (K)

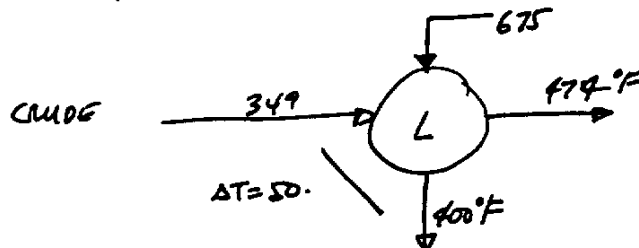
$$T = 310 + \frac{30 \times 10^6}{7.72 \times 10^5} = 310 + 39 = 349^\circ\text{F}.$$

USE A 200°F ΔT FOR A50 PUMPAROUND STREAM.

$$\text{OUTLET TEMP} = 600 - 200 = 400^\circ\text{F}$$

$$\text{CALCULATE PUMPAROUND RATE} = \frac{30 \times 10^6}{(0.78)(600 - 400)} = 192,310 \text{ LB/HR}.$$

ONLY REMAINING STREAM IS REDUCED CRUDE STREAM.



REDUCED CRUDE FROM EXCHANGER TEMPERATURE SET AT $349 + 50 = 400^\circ\text{F}$

$$\text{CALCULATE DUTY} = (3.464 \times 10^5)(675 - 400) = 96.2 \times 10^6 \text{ BTU/HR}$$

CALCULATE CRUDE TEMPERATURE FROM EXCHANGER.

$$T = 349 + \frac{96.2 \times 10^6}{7.72 \times 10^5} = 349 + 125 = 474^\circ\text{F}$$

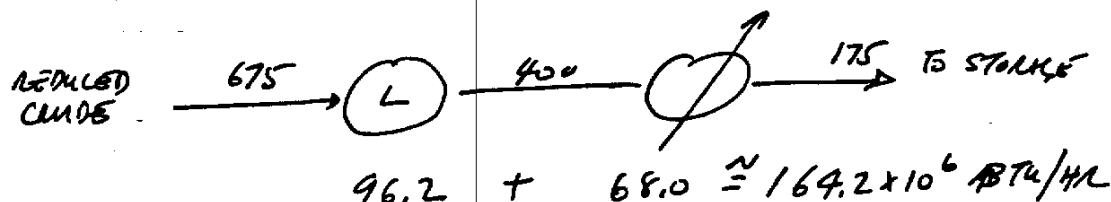
CHECK APPROACHES

$$\begin{array}{r} \text{HOT } 675 \rightarrow 400 \\ 474 \leftarrow 349 \\ \hline 101 \quad 51 \end{array}$$

OK.

HEAT EXCHANGE NETWORK.

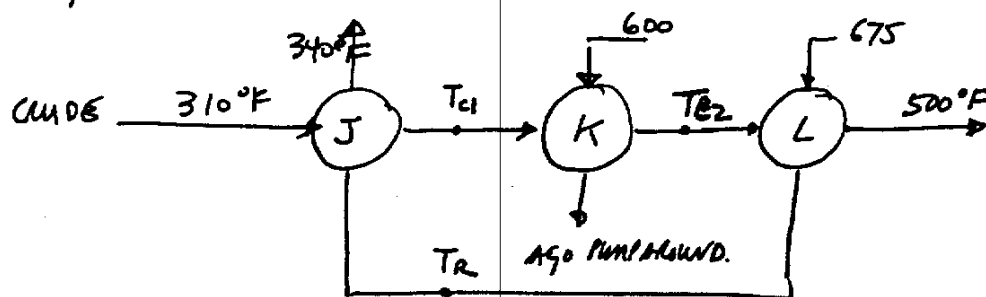
COMPLETING THE REDUCED CLUDE EXCHANGE TRAIN TO DETERMINE REDUCED CLUDE COOLER DUTY.



SUMMARIZING TOTAL HEAT REJECT TO COOLING.

FET FUEL COOLER	= 3.17
DIESEL COOLER	14.30
AGO COOLER	8.30
RESID COOLER	68.00
TOTAL.	$93.77 \times 10^6 \text{ BTU/HR}$

THE LARGEST HEAT REJECTION IS IN RESID COOLER. LOOK AT ALTERNATE RESID EXCHANGE TO IMPROVE HEAT RECOVERY. LETS SPLIT THE RESID EXCHANGE AROUND THE AGO PUMPAROUND EXCHANGE. SET A CLUDE OUTLET TEMPERATURE OF 500°F



$$\text{CLUDE SIOB TOTAL DUTY} = (7.72 \times 10^5)(500 - 310) = 146.68 \times 10^6 \text{ BTU/HR}$$

$$\begin{aligned} \text{DUTY REQUIRED FROM REDUCED CLUDE EXCHANGE (J) + (L)} \\ = 146.68 - 30 = 116.68 \times 10^6 \text{ BTU/HR} \end{aligned}$$

$$\begin{aligned} \text{TEMPERATURE OF REDUCED CLUDE OUT} &= 675 - \frac{116.68 \times 10^6}{3.484 \times 10^5} = 675 - 335 \\ &= 340^\circ\text{F} \end{aligned}$$

HEAT EXCHANGE NETWORK.

LOOK AT APPROX TEMPERATURE ON EXCHANGER (J).

HOT	T_R	340
COLD	T_{C1}	310
	?	30

WE HAVE VIOLATED OUR ΔT OF 50°F GUIDELINE. TO DETERMINE IF THIS EXCHANGE SCHEME IS BETTER THAN THE PREVIOUS SCHEME. A CAPITAL COST VS UTILITY SAVING EVALUATION NEEDS TO BE DONE. LETS FINISH THE CALCULATION.

BY CLASS VOTE IT WAS DECIDED THAT IT WOULD EASIER TO GUESS T_R AND WORK OUT THE REMAINING TEMPERATURES THAN SETUP THE EQUATIONS TO SOLVE THE PROBLEM. THE FINAL ITERATION IS SHOWN BELOW.

$$T_R = 400^\circ\text{F}$$

$$Q_{RC} = (3.5 \times 10^5)(400 - 340) = 21 \times 10^6 \text{ BTU/HR.}$$

CALCULATE CLUDE TEMPERATURE FROM EXCHANGER (J)

$$T_{C1} = 310 + \frac{21 \times 10^6}{7.72 \times 10^5} = 310 + 27 = 337^\circ\text{F}$$

$$\begin{array}{r} 400 \rightarrow 340 \\ 337 \leftarrow 310 \\ \hline 63 \end{array}$$

CALCULATE CLUDE TEMPERATURE FROM EXCHANGER (K)

$$T_{C2} = 337 + \frac{30 \times 10^6}{7.72 \times 10^5} = 337 + 39 = 376^\circ\text{F}$$

$$\begin{array}{r} 600 \rightarrow 400 \\ 376 \leftarrow 337 \\ \hline 224 \end{array}$$

$$\text{NOW DUTY OF EXCHANGER (L)} = 116.68 - 21 = 95.68 \times 10^6 \text{ BTU/HR}$$

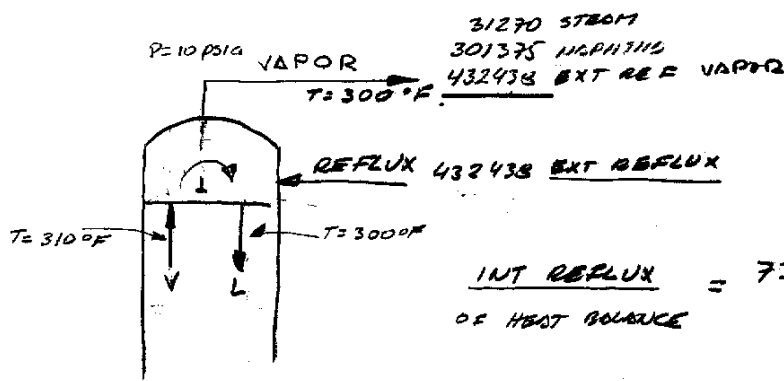
CALCULATE CLUDE TEMPERATURE FROM EXCHANGER (L) AS A CHECK.

$$T = 376 + \frac{95.68 \times 10^6}{7.72 \times 10^5} = 376 + 124 = 500^\circ\text{F}$$

$$\begin{array}{r} 675 \rightarrow 400 \\ 500 \leftarrow 376 \\ \hline 175 \end{array}$$

CHECKS WITH STARTING TEMPERATURE OF 500°F OK.

LB/H



$$\text{INT REFLUX} = 737992 \text{ LB/H}$$

OF HEAT BALANCE

- 1° ASSUME COMPOSITION OF VAPOR TO TRAY 1 EQUAL TO OVHD VAPOR
 - 2° ASSUME COMPOSITION OF LIQUID TO TRAY 1 EQUAL TO REFLUX COLUMN
- LET V & L BE VAPOR & LIQUID LOADINGS AT TOP OF TOWER.
 CALCULATE TOWER DIAMETER
 USE IDEAL GAS LAW TO LOCATE VAPOR VOLUME RATE
 AND EXPANSION FACTORS TO CALCULATE HOT LIQUID RATE.

BALANCE PARA DETERMINAR V

$$31270 + 301375 + 432438 + 737992 = 432438 + V$$

$$V = 1070637$$

$$L = 737992$$

MDL WT OVHD 102.9 OF HC

$$1070637 \text{ LB/HR} = \text{VAPOR LOAD (SOLUBIL ORL PRO)}$$

$$PV = \frac{\text{MASS}}{\text{MDL WT}} \cdot \frac{RT}{\text{MDL WT}}$$

$$\text{VAPOR DENSITY AT TEMPERATURE} = \underline{\underline{0.272}} \text{ LB/FT}^3$$

$$\text{DENSITY} = \frac{\text{MASS}}{\text{VOLUME}}$$

$$\text{DENSITY OF VAPOR} = \frac{P(\text{MDL WT})}{RT}$$

$$= \frac{(10 + 10.7)(102.9)}{(10.7)(310 + 460)}$$

$$= 0.3082$$

ADJUST OF MOL WT FOR PRESENCE OF STEAM.

$$\text{MOL HC} = \frac{1070637}{102.9}$$

$$= 10404.6 \text{ mol}$$

$$\text{MOL STEAM} = \frac{31270}{18}$$

$$= 1737$$

$$\text{TOTAL} = 12141.6$$

$$\Delta V \text{ MOL WT} = \frac{1101903 \text{ LB}}{12141.6 \text{ mol}} -$$

$$= 90.75$$

$$\text{DENSITY} = \frac{24.7 \times 90.75}{10.71 \times 770}$$

$$= \underline{\underline{0.272 \text{ LB/ft}^3}}$$

$$\text{LIQUID LOAD} = 737992 \text{ LB/HR. (SOLUENDO OR PLATO)}$$

$$\text{LIQUID DENSITY} = \frac{36.9}{\text{AT TEMP}}$$

$$\frac{301375 \text{ LB/HR}}{28650 \text{ BPD}} = 10.5192 \text{ LB/bbl.}$$

$$K = 12.5$$

$$\Delta \text{PI} = 64.77$$

$$\text{EXPANSION FACTOR (CHART)} = 0.82$$

$$\text{SPG @ } 300^\circ\text{F} = 0.7214 \times 0.82$$

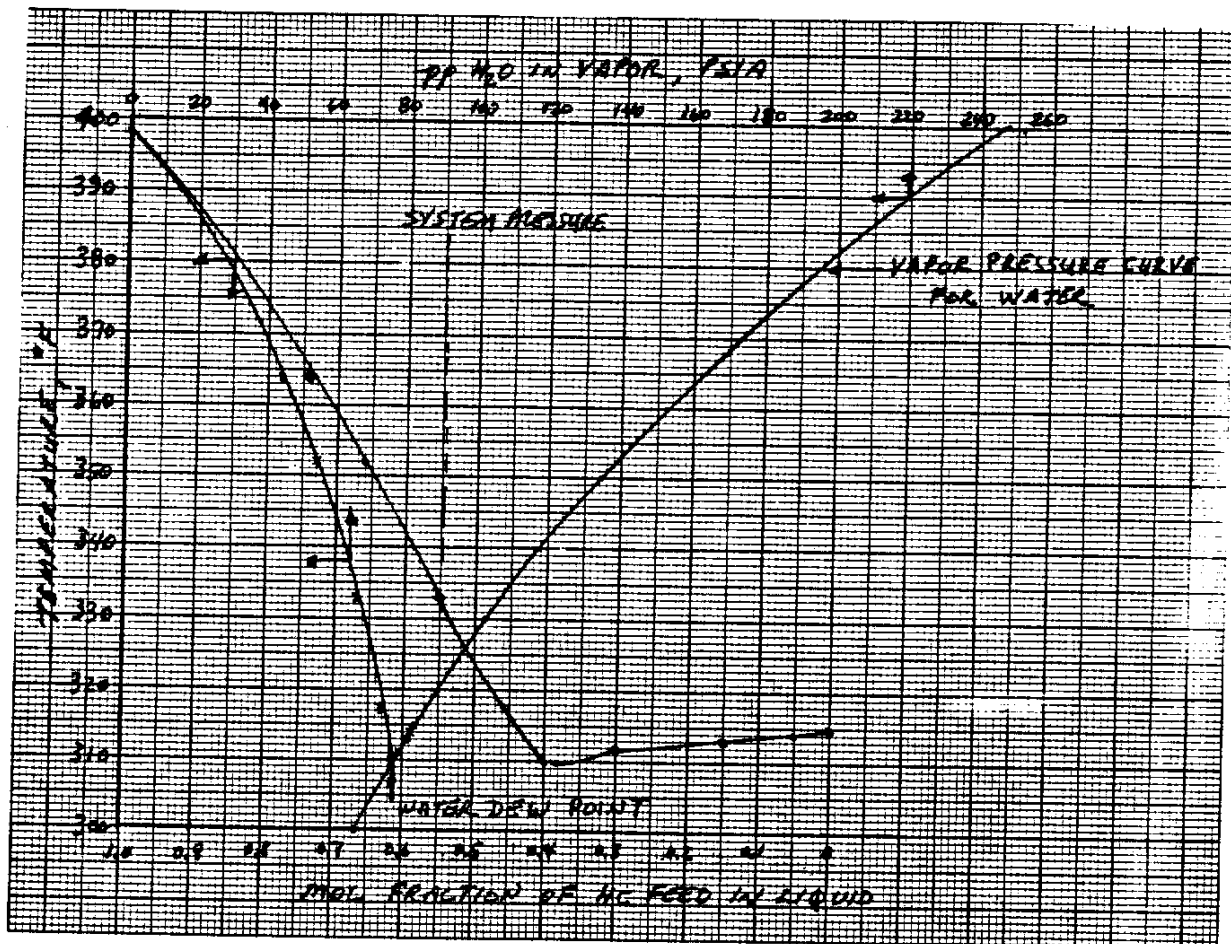
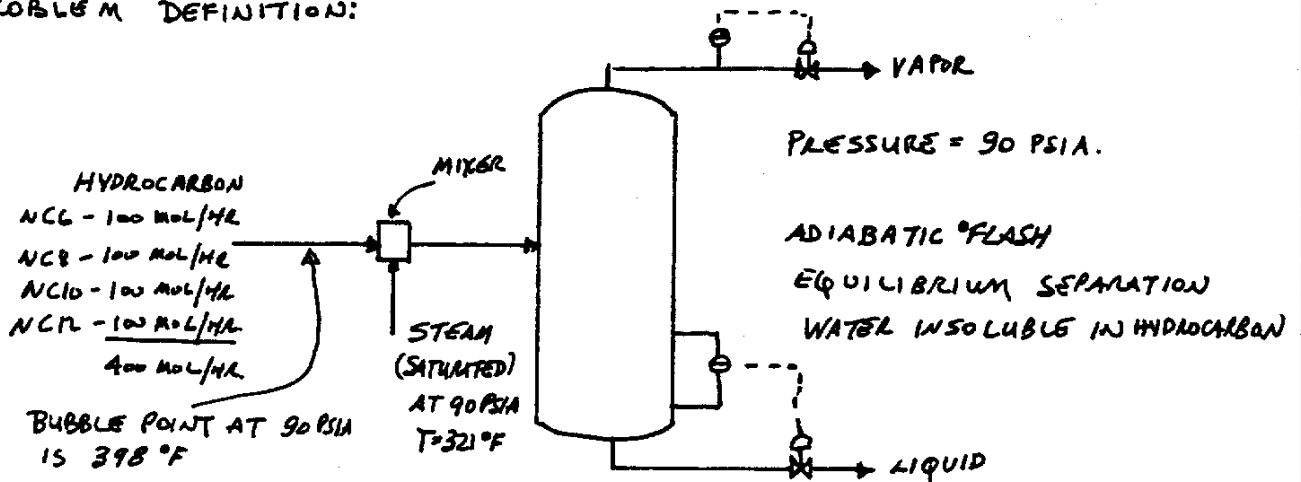
$$= 0.5917$$

$$\text{DENSITY} = 0.5917 \times 62.38$$

$$= \underline{\underline{36.9}}$$

SINGLE STAGE FLASH WITH STEAM

PROBLEM DEFINITION:



AFTER WATER DEW POINT REACHED TEMP & PP H₂O IN VAPOR FOLLOW WATER VAPOR PRESSURE CURVE APPROACHING 90 PSIA AND 321 °F IN LIMIT.

HEAT EXCHANGE NETWORK.

LET'S SEE IF WE CAN FINALIZE FOR THIS DESIGN OUR HEAT EXCHANGE NETWORK. FOR THE REDUCED CRUDE / CRUDE EXCHANGE WE HAS A BASE CASE AND AN ALTERNATE CASE WHICH IMPROVED HEAT RECOVERY FROM REDUCED CRUDE BUT WITH ADDITIONAL SURFACE REQUIRED. DETERMINE IF ADDITIONAL SURFACE CAN BE JUSTIFIED.

ESTIMATE SURFACE REQUIRED FOR BASE CASE DESIGN.

$$U \approx 40.$$

$$DUTY = 96.2 \times 10^6 \text{ BTU/HR (FROM PAGE 13 OF NETWORK DESIGN)}$$

$$\begin{array}{r} 675 \rightarrow 400 \\ 374 \leftarrow 349 \\ \hline 201 \quad 51 \end{array} \quad \begin{array}{l} \frac{275}{125} = 2.2 = R \\ \frac{125}{326} = 0.383 = F \end{array}$$

MIN OF TWO SHELLS IN SERIES.

$$LMTD = 109.37$$

$$F = 0.86$$

$$Q = UA \Delta T_{LM}$$

$$A = \frac{96.2 \times 10^6}{(40)(109.37)(0.86)} = 25570 \text{ ft}^2$$

$$AREA / SHELL = 25570 / 2 = 12785 \text{ FT}^2$$

TYPICAL TUBE CONFIGURATION $\frac{3}{4}$ " OD ON 1 inch square pitch.
BASED ON 20 FT TUBES APPROXIMATE SHELL SIZE 58" ID.

IF WE DIVIDE INTO TWO PARALLEL EXCHANGE TRAINS

$$AREA / SHELL = 6395 \text{ FT}^2 \quad (2 \text{ PARALLEL - 2 SERIES FLOW})$$

$$APPROXIMATE SHELL ID = 42"$$

CUSTOMER ENGINEERING DESIGN SEMINAR.
 SUBJECT CANOE UNIT DESIGN.

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 PROJECT HEAT EXCHANGE NETWORK.
 BY EN MILLER.

LET'S ESTIMATE EXCHANGE SIZE REQUIREMENTS FOR ALTERNATE EXCHANGE SCHEMES.

FOR THE HOT REDUCED CANOE/CANOE EXCHANGE. (SEE PAGE 15)

$$\begin{array}{r} 675 \rightarrow 400 \\ 800 \leftarrow 276 \\ 175 \quad 24 \end{array} \quad \begin{array}{l} \frac{275}{124} = 2.2 \\ \frac{124}{299} = 0.415 \end{array}$$

$$LMTD = 76 \quad F = 0.88.$$

$$A = \frac{95.68 \times 10^6}{(40)(76)(0.88)} = 35766 \text{ ft}^2.$$

$$AREA/SHELL = 35766/3 = 11922 \text{ FT}^2.$$

NEED TO DIVIDE INTO TWO PARALLEL AND 2 IN SERIES.

$$AREA/SHELL = \frac{11922}{2} = 5961 \text{ FT}^2 \text{ (2 PARALLEL - 3 SERIES FLO.)}$$

FOR THE COLD REDUCED CANOE/CANOE EXCHANGE. (SEE PAGE 15).

$$\begin{array}{r} 400 \rightarrow 340 \\ 337 \leftarrow 310 \\ 63 \quad 30 \end{array} \quad \begin{array}{l} \frac{60}{27} = 2.22 = R \\ \frac{27}{90} = 0.3 = P \end{array}$$

$$LMTD = 44.48 \quad F = 0.83$$

$$A = \frac{21 \times 10^6}{(40)(44.48)(0.83)} = 14221 \text{ ft}^2$$

NEED TO DIVIDE INTO TWO PARALLEL AND 1 IN SERIES.

$$AREA/SHELL = 14221/2 = 7111 \text{ FT}^2.$$

NEED MINIMUM OF THREE SHELLS IN SERIES.

$$U \approx 40.$$

$$Q = 95.68 \times 10^6 \text{ BTU/HR.}$$

NEED ONE SHELL IN SERIES.

$$U \approx 40$$

$$Q = 21 \times 10^6 \text{ BTU/HR.}$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT DESIGNPAGE NO. 18DATE 10-17-91PROJECT EN MILLERHEAT EXCHANGE NETWORK

LET'S SUMMARIZE THE TWO REDUCED CRUDE/CRUDE EXCHANGE OPTIONS BASED ON TOTAL AREA REQUIREMENTS.

	BASE CASE.	ALTERNATE CASE.
TOTAL DUTY.	96.2×10^6 BTU/HR	116.68×10^6 BTU/HR
TOTAL AREA.	25570 FT ²	49987 FT ² .

CALCULATE HEATER DUTY FOR TWO CASES.

CRUDE TO COLUMN.	560.747×10^6 BTU/HR	560.747×10^6 BTU/HR.
CRUDE TO HEATER	287.890×10^6 BTU/HR	310.131
HEATER DUTY.	272.857	250.616

$$\Delta \text{DUTY} = 22.241 \times 10^6 \text{ BTU/HR}$$

ASSUME CRUDE HEATER EFFICIENCY OF 90 % .

$$\Delta \text{DUTY FIRED} = \frac{22.241 \times 10^6}{0.90} = 24.71 \times 10^6 \text{ BTU/HR.}$$

EXCHANGER SURFACE COST ABOUT \$36.50 / FT².

$$\begin{aligned} \text{DELTA SURFACE COST} &\approx (49987 - 25570)(36.50) \\ &\approx \$892,000 \end{aligned}$$

FROM DESIGN BASIS FUEL OIL COST WAS \$3.50 MM BTU/HR.

SIMPLE PAYOUT TIME FOR SURFACE

$$= \frac{892,000}{(24.71)(3.50)(8000)} = 1.3 \text{ years.}$$

LOOKS LIKE THE ALTERNATE SCHEME WILL PAYOUT. WHAT WE DID NOT EVALUATE WAS THE EFFECT ON THE CRUDE/AGO PUMPAROUND EXCHANGER SURFACE OF THE ALTERNATE SCHEME. LETS DO THAT NOW.

HEAT EXCHANGE NETWORK.

ESTIMATE SURFACE REQUIREMENTS FOR CRUDE/AGO EXCHANGER FOR BASE CASE OPERATION. (SEE PAGE 13)

U VALUE SHOULD BE BETTER FOR THIS SERVICE TRY $U = 50$.

$$\begin{array}{rcl} 600 \rightarrow 400 & \frac{200}{39} = 5.13 = R. \\ 349 \leftarrow 310 & \frac{39}{290} = 0.135 = P \\ \hline 251 & 90 \end{array}$$

USE ONE SHELL
IN SERIES.

$$LMTD = 156.98$$

$$F = 0.935$$

$$Q = 30 \times 10^6 \text{ BTU/HR}$$

$$A = \frac{30 \times 10^6}{(156.98)(50)(0.935)} = 4090 \text{ ft}^2.$$

FOR ALTERNATE EXCHANGE SCHEME. (SEE PAGE 15)

$$\begin{array}{rcl} 600 \rightarrow 400 & \frac{200}{39} = 5.13 = R. \\ 376 \leftarrow 337 & \frac{39}{263} = 0.148 = P \\ \hline 224 & 63 \end{array}$$

USE ONE SHELL
IN SERIES

$$LMTD = 126.92$$

$$F = 0.9$$

$$Q = 30 \times 10^6 \text{ BTU/HR.}$$

$$A = \frac{30 \times 10^6}{(126.92)(0.9)(50)} = 5253 \text{ ft}^2.$$

$$\text{DELTA SURFACE} = 5253 - 4090 = 1163 \text{ ft}^2$$

$$\text{DELTA COST} \approx (1163)(36.5) = \$42,500$$

$$\text{RECALCULATE PAYOUT} = \frac{(892,000 + 42,500)}{(24.71)(3.5)(8000)} = 1.35 \text{ years.}$$

NOTE: WE HAVE NOT TAKEN CREDIT FOR REDUCTION IN AIR COOLER SURFACE ~~REDUCTION~~ RESULTING FROM REDUCED COOLING DUTY REQUIREMENT ON REDUCED CRUDE STREAM.

CONCLUSION: USE ALTERNATE EXCHANGE SCHEME FOR DESIGN.

HEAT EXCHANGE NETWORK.

THE OTHER OPTION DISCUSSED WAS EXCHANGE OF CRUDE WITH A NAPHTHA PUMP AROUND AT THE TOWER TOP RATHER THAN CRUDE/OVERHEAD EXCHANGE. LETS EVALUATE USE OF A NAPHTHA PUMP AROUND USING THE SAME HEAT REMOVAL - 62.6×10^6 BTU/HR - AS THE BASE CASE EVALUATION. SINCE THIS IS PRIMARILY A HEAT REMOVAL SECTION THE, DRAW TRAY TEMPERATURE SHOULD BE ABOUT 310°F . FOR PROPERTIES WE WILL ASSUME THAT THE PUMP AROUND NAPHTHA IS SIMILAR TO THE NAPHTHA FROM THE SAMPLE PROBLEM CRUDE BREAKUP. UOP $K=12$ $CASP=300^\circ\text{F}$ $\Rightarrow$ $API=54$.

ASSUME PUMP AROUND RETURN TEMPERATURE IS 210°F .
CALCULATE PUMP AROUND RATE.

$$W = \frac{62.6 \times 10^6}{(148-81)} = 934,335 \text{ LB/HR}$$

TOTAL EXCESS HEAT TO BE REMOVED BY REFLUX (EITHER EXTERNAL REFLUX OR PUMP AROUND) REMAINS SAME.

TOTAL EXCESS TO BE REMOVED AT TOP WAS 94.704 MMBTU/HR
(SEE PAGE 25 OF EXCESS HEAT).

EXCESS HEAT TO BE REMOVED BY EXTERNAL REFLUX
 $= (94.704 - 62.6) = 32.104 \times 10^6 \text{ BTU/HR.}$

$$\text{EXTERNAL REFLUX RATE} = \frac{32.104 \times 10^6}{(263-44)} = 146,595 \text{ LB/HR.}$$

$$= 1424.33 \text{ MO/HR.}$$

SINCE THE PARTIAL PRESSURE STEAM IN OVERHEAD VAPOR INCREASES, THE HYDROCARBON DEW POINT TEMPERATURE SHOULD DECREASE.

IN OVERHEAD TOTAL HC = EXTERNAL REFLUX + NET OVERHEAD = $1424.33 + 2928.18 = 4352.51 \text{ MO/HR.}$
 STEAM = 1737.07 MO/HR.
 6089.58

$$PP \text{ HC} = 24.7 - \frac{1737.07}{6089.58}(24.7) = 17.7 \text{ PSIA VS } 17.9 \text{ PSIA}$$

(SEE CORRECTED SHEET 26 OF EXCESS HEAT)

USING COX CHART EST TEMP = 295°F .



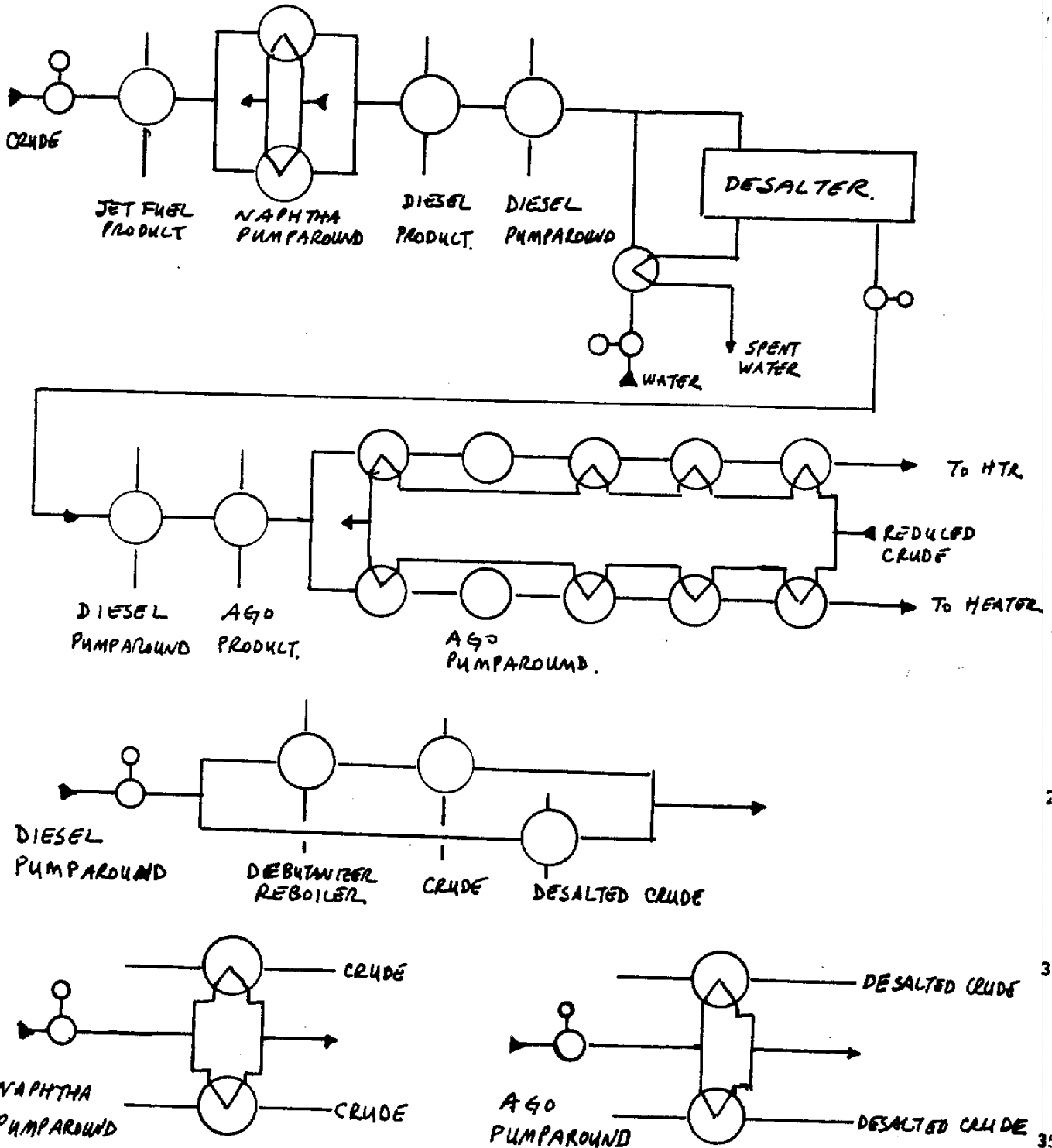
ENGINEERING DEPARTMENT

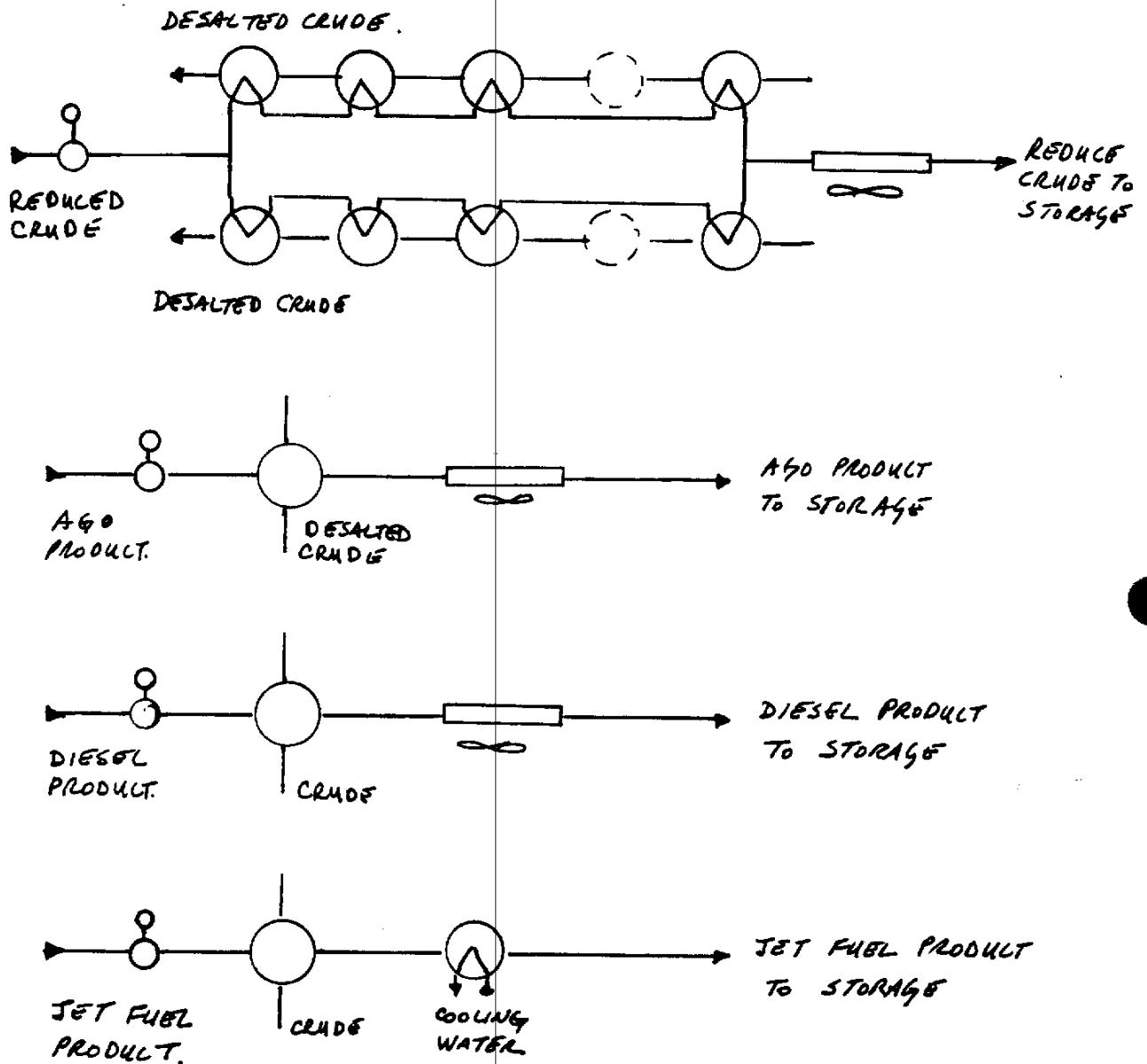
CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT DESIGNPAGE NO. 21DATE 10-17-91

PROJECT

BY E N MILLERHEAT EXCHANGE NETWORK

IN THEORY WE SHOULD REITERATE THE ENTIRE HEAT BALANCE AT THE TOP OF THE TOWER FOR THE LOWER OVERHEAD TEMPERATURE WHICH FOR A CONSTANT PUMPAROUND HEAT REMOVAL RATE WILL INCREASE THE EXTERNAL REFLUX RATE. THIS IN TURN WILL INCREASE THE PARTIAL PRESSURE OF HYDROCARBON IN THE OVERHEAD WHICH IN TURN WILL INCREASE THE OVERHEAD TEMPERATURE BACK TOWARDS 300°F. ETC. SINCE ~~WE~~ WE EXPECT THE CALCULATED TEMPERATURES TO BE NO BETTER THAN $\pm 5^\circ\text{F}$, WE WILL LEAVE THE TOP TEMPERATURE AT 300°F AND NOT REITERATE THE HEAT BALANCE.





PERTINENT DATA FOR THE HEAT EXCHANGE NETWORK HAS BEEN SUMMARIZED ON THE FOLLOWING PAGE. THIS SHOULD BE CONSIDERED PRELIMINARY AND DOES NOT REPRESENT A FINAL DESIGN. HOWEVER THIS IS AS FAR AS WE WILL GO IN THIS CLASS.

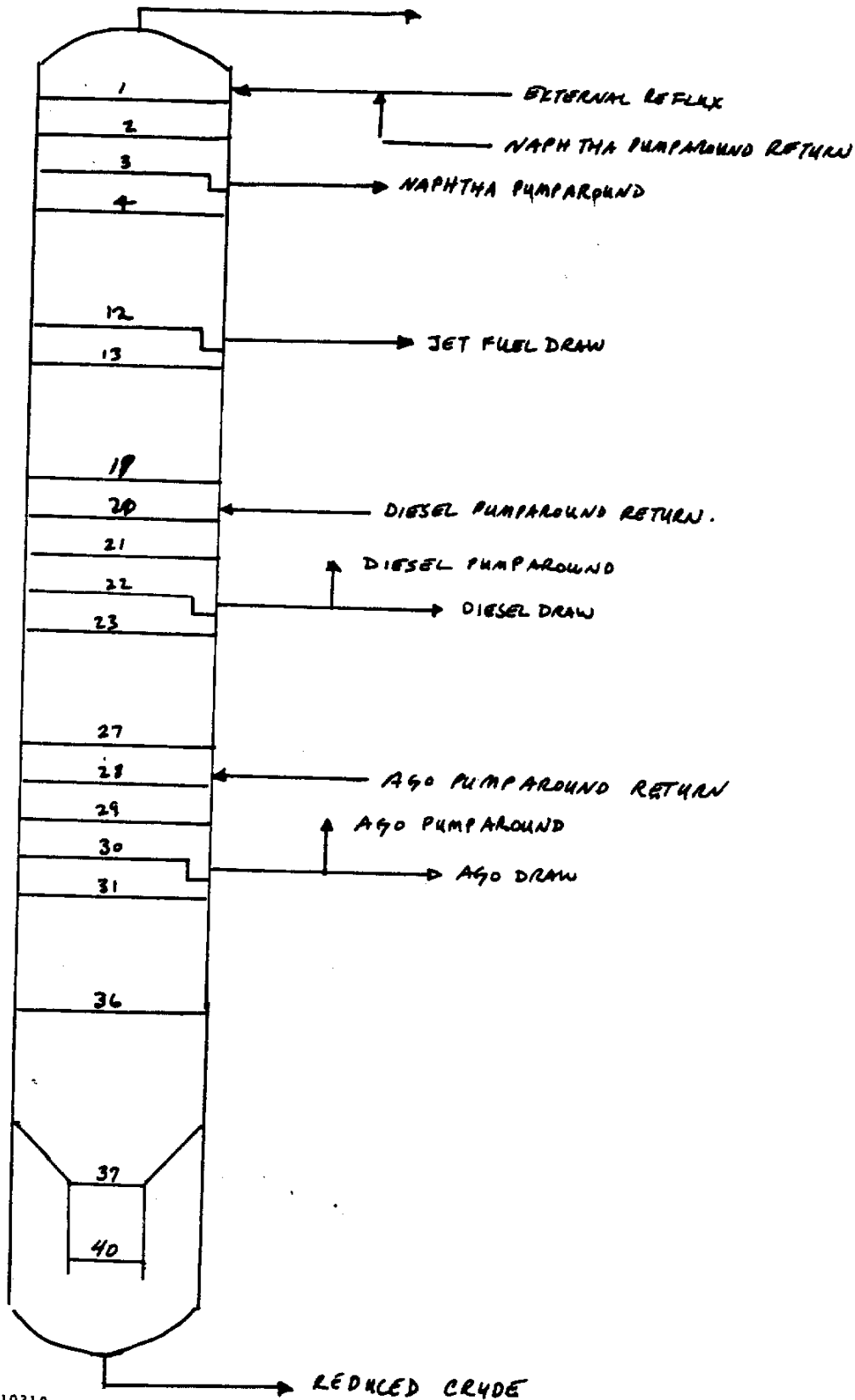
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Project

Customer ENGINEERING DESIGN SEMINAR - CRUDE UNIT DESIGN[illegible]

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER SIZE / FLOW LOADINGS

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 BY EN MILLER



CUSTOMER

ENGINEERING DESIGN SEMINAR

SUBJECT

CRUDE UNIT DESIGN

CRUDE TOWER SIZE/TRAY LOADINGS

PAGE NO.

2

DATE

10-15-91

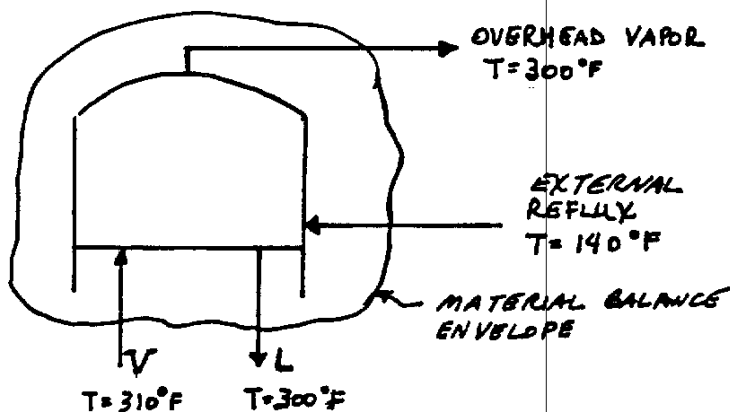
PROJECT

BY EN MILLER

CRUDE TOWER SIZING:

IN ORDER TO SIZE THE CRUDE TOWER WE MUST FIRST CALCULATE TRAY VAPOR AND LIQUID LOADINGS FOR THE VARIOUS SECTIONS OF THE TOWER. WE WILL DO THIS NOW STARTING AT THE TOP OF THE TOWER. SINCE THE HEAT BALANCE WAS DONE AT THE TIME THE "EXCESS" HEAT CALCULATIONS WERE MADE, WE NOW NEED TO COMPLETE THE MATERIAL BALANCE CALCULATIONS AT EACH SECTION AND DETERMINE THE PROPERTIES OF THE LIQUID AND VAPOR.

FOR THE TRAYS BETWEEN THE TOP AND THE JET FUEL DRAW TRAY. (SEE PAGE 27 OF EXCESS HEAT NOTES).

MATERIAL BALANCE

$$V + (\text{REFLUX}) = L + (\text{OVERHEAD VAPOR})$$

$$L \text{ (INTERNAL REFLUX)} = \underline{737992 \text{ LB/HR}}$$

$$\text{REFLUX} = 432,438 \text{ LB/HR}$$

$$\begin{aligned} \text{OVERHEAD VAPOR} &= \text{STEAM} + \text{UNSTABILIZED NAPHTHA} + \text{REFLUX} \\ &= 31270 + 301575 + 432438 = 765083 \text{ LB/HR} \end{aligned}$$

$$V + 432438 = 737992 + 765083.$$

$$\underline{V = 1070637 \text{ LB/HR}}$$

TO DETERMINE THE PROPERTIES OF THE LIQUID AND VAPOR ASSUME THE COMPOSITION OF THE LIQUID AND THE HYDROCARBON PORTION OF THE VAPOR TO BE THE SAME AS THE NET OVERHEAD PRODUCT FROM THE TOWER. THIS IS NOT STRICTLY TRUE BUT ADEQUATE FOR OUR PURPOSES.

CUSTOMER ENGINEERING DESIGN SEMWAR
 SUBJECT CRUDE UNIT DESIGN.
CRUDE TOWER SIZE / TRAY LOADINGS

PAGE NO. 3
 DATE 10-15-91
 PROJECT EN MILLER

PROPERTIES OF NET OVERHEAD FROM CRUDE TOWER ARE SUMMARIZED BELOW. (THIS INFORMATION CAN BE OBTAINED EITHER FROM PRODUCT YIELD SUMMARY - PAGE III-4A OR THE OVERALL MATERIAL BALANCE - PAGE III-2A)

	LB/HR	MOL/HR	BPSD	MW	API	UOPK
UNSTABILIZED NAPHTHA	301375	2928.18	28650	102.9	64.77	>12.13 = 12.25

$$\frac{\text{LB/HR}}{\text{BPSD}} = \frac{301375}{28650} = 10.5192 \Rightarrow \text{API} = 64.77$$

VAPOR DENSITY CAN BE CALCULATED USING IDEAL GAS LAW.

$$\text{MOLES HYDROCARBON} = \frac{1039367}{102.9} = 10,100 \text{ MOL/HR.}$$

$$\text{MOLES STEAM} = \frac{31270}{18} = 1737 \text{ MOL/HR}$$

$$\text{VAPOR TOTAL} = 11837 \text{ MOL/HR.}$$

$$PV = NRT = \frac{M \cdot RT}{MW} \quad \text{DENSITY} = \frac{W}{V} = \frac{(MW)(P)}{(R)(T)}$$

$$\text{FOR MIXTURE } MW = \frac{1070637}{11837} = 90.43 \quad P = 10 + 14.7 = 24.7 \text{ PSIA}$$

$$T = 310 + 460 = 770^\circ \text{R.}$$

$$\text{DENSITY } d_v = \frac{(90.43)(24.7)}{(10.73)(770)} = \underline{\underline{0.270 \text{ LBS/FT}^3}}$$

TO CALCULATE LIQUID DENSITY AT FLOWING TEMPERATURE USE EITHER PD-885 OR PD-67 TO OBTAIN EXPANSION FACTOR. FOR PD-885 NEED TO CALCULATE PSEUDOCRITICAL TEMPERATURE (P_T). WE CALCULATED THE P_T FOR THE NAPHTHA PORTION OF THIS MATERIAL PREVIOUSLY (CRUDE COLUMN OVERHEAD - PAGE V-1A) AS 555°F TOTAL IS SLIGHTLY LIGHTER SAY $P_T = 580^\circ \text{F}$ (BASED ON ANAL. BY SP OF MIX AND CORRECTION D-2).

FROM PD-885 EXPANSION FACTOR = 0.834 ← USE

FROM PD-67 EXPANSION FACTOR = 0.83

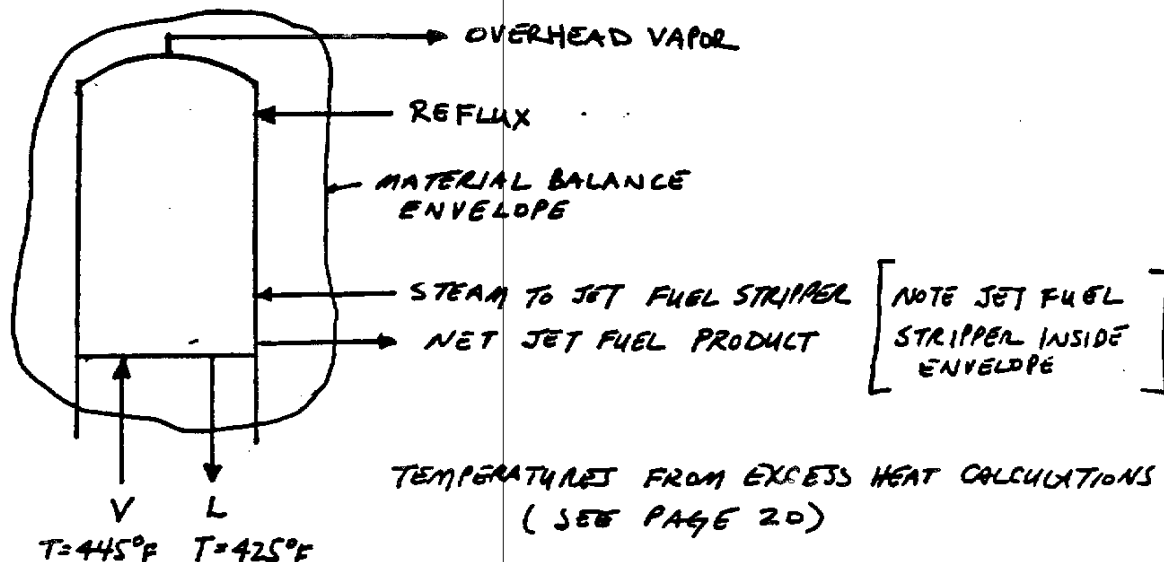
$$\text{DENSITY AT } 60^\circ \text{F} = (0.7209)(62.38) = 44.97 \text{ LB/FT}^3$$

$$\text{DENSITY AT } 300^\circ \text{F} = d_L = (44.97)(0.834) = \underline{\underline{37.5 \text{ LB/FT}^3}}$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER SIZE / TRAY LOADINGS

PAGE NO. 4
 DATE 10/15/91
 PROJECT _____
 BY EN MILLER

TRAYS BETWEEN JET FUEL DRAW AND DIESEL DRAW TRAY.
 IGNORE FOR THE MOMENT THE PUMP AROUND TRAYS IN THIS
 SECTION. THEY WILL BE CALCULATED SEPARATELY.



MATERIAL BALANCE

$$V + (\text{REFLUX}) + (\text{JF STR STEAM}) = L + (\text{OVERHEAD VAPOR}) + (\text{NET JET FUEL})$$

$$\text{INTERNAL REFLUX BELOW JET FUEL DRAW} = \underline{L = 391,975 \text{ LB/HR}} \quad (\text{SEE PAGE 28})$$

$$V + 432,438 + 7,660 = 391,975 + 765,083 + 196,035$$

$$\underline{V = 912,995 \text{ LB/HR}}$$

INTERNAL REFLUX HAS THE SAME PROPERTIES AS DRAW TRAY LIQUID
 MW = 183.3 API = 43.3 WOPK = 11.97 (SEE PAGE 10)

HYDROCARBON PORTION OF VAPOR INCLUDES NET PRODUCTS AT THAT POINT
 PLUS INTERNAL REFLUX FROM JET FUEL DRAW TRAY. SUMMARIZED

	LB/HR	MOL WT	MOL/HR
INTERNAL REFLUX L	391,975	183.3	2138.43
UNSTABILIZED NAPHTHA	301,375	102.9	2928.18
JET FUEL	196,035	187	1048.32
TOTAL	889,385	145.44	6114.93

STEAM PORTION OF VAPOR

	LB/HR	MOL WT	MOL/HR
STEAM TO TOWER.	15875	18	882
STEAM TO AGO STRIPPER	3495	18	194
STEAM TO DIESEL STRIPPER	4240	18	235.6
	23610	18	1311.6
HYDROCARBON (FROM PREVIOUS PAGE)	889385	145.44	6114.93
VAPOR TOTAL	912995	122.94	7426.53

$$P = 11.5 + 14.7 = 26.2 \text{ PSIA} \quad T = 445 + 460 = 905^{\circ}\text{R}$$

$$\text{DENSITY} = d_v = \frac{(122.94)(26.2)}{(10.73)(905)} = \underline{\underline{0.332 \text{ LB/FT}^3}}$$

CALCULATE FLOWING DENSITY OF LIQUID FROM JET FUEL TRAY

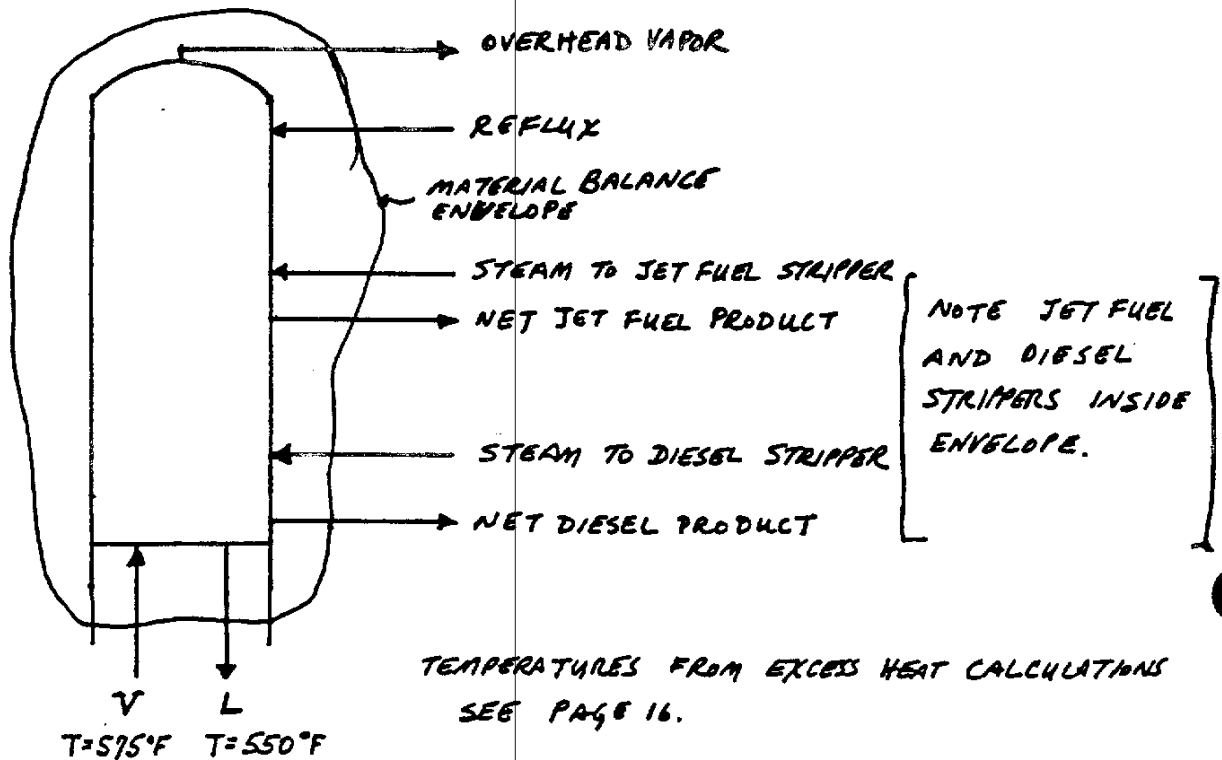
$$T_c \approx 765^{\circ}\text{F}$$

FROM CHART PD-885 @ 425°F EXPANSION FACTOR = 0.815

$$\text{DENSITY AT } 60^{\circ}\text{F} = (0.8095)(62.38) = 50.5 \text{ LB/FT}^3$$

$$\text{DENSITY @ } 425^{\circ}\text{F} = d_L = (50.5)(0.815) = \underline{\underline{41.16 \text{ LB/FT}^3}}$$

TRAYS BETWEEN DIESEL DRAW TRAY AND AGO DRAW TRAY.
IGNORE FOR THE MOMENT THE PUMP AROUND TRAYS IN THIS
SECTION. THEY WILL BE CALCULATED SEPARATELY.



MATERIAL BALANCE

$$V + (\text{REFLUX}) + (\text{STEAM TO JF STR}) + (\text{STEAM TO DIESEL STR}) = L + (\text{OVERHEAD VAPOR}) + (\text{NET JET FUEL}) + (\text{NET DIESEL})$$

$$\text{INTERNAL REFLUX BELOW DIESEL DRAW} = L = 670240 \text{ LB/HR (SEE PAGE 28)}$$

$$V + 432438 + 7660 + 4240 = 670240 + 765083 + 196035 + 113460$$

$$V = 1300480 \text{ LB/HR}$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER SIZE / TRAY LOADINGS

PAGE NO. 7
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 BY E. H. MILLER

INTERNAL REFLUX BELOW DRAW HAS THE SAME PROPERTIES AS THE DRAW TRAY LIQUID: MW = 246.8 API = 35.5 UOP K = 11.93 (SEE PAGE 8)

HYDROCARBON PORTION OF VAPOR INCLUDES NET PRODUCTS AT THAT POINT PLUS INTERNAL REFLUX FROM DIESEL DRAW TRAY.
 SUMMARIZED

	LB/HR.	MOL WT	MOL/HR.
INTERNAL REFLUX L	670242	246.8	2715.72
UNSTABILIZED NAPHTHA	301375	102.9	2928.18
JET FUEL	196035	187	1048.32
DIESEL	113460	253	448.46
TOTAL	1281110	179.41	7140.68
STEAM TO TOWER	15875	18	882
STEAM TO AGO STRIPPER.	3495	18	194
VAPOR TOTAL	1300480	158.27	8216.68

$$P = 13 + 14.7 = 27.7 \text{ PSIA} \quad T = 575 + 460 = 1035^\circ \text{R.}$$

$$\text{DENSITY} = d_v = \frac{(158.27)(27.7)}{(10.73)(1035)} = 0.395 \text{ LB/FT}^3$$

CALCULATE FLOWING DENSITY OF LIQUID FROM DIESEL DRAW TRAY.

$$p_{T_L} = 885^\circ \text{F} \text{ FROM CHART PD-185 @ } 550^\circ \text{F} \text{ EXPANSION FACTOR} = 0.78$$

$$\text{DENSITY AT } 60^\circ \text{F} = (0.8473)(62.38) = 52.85 \text{ LB/FT}^3$$

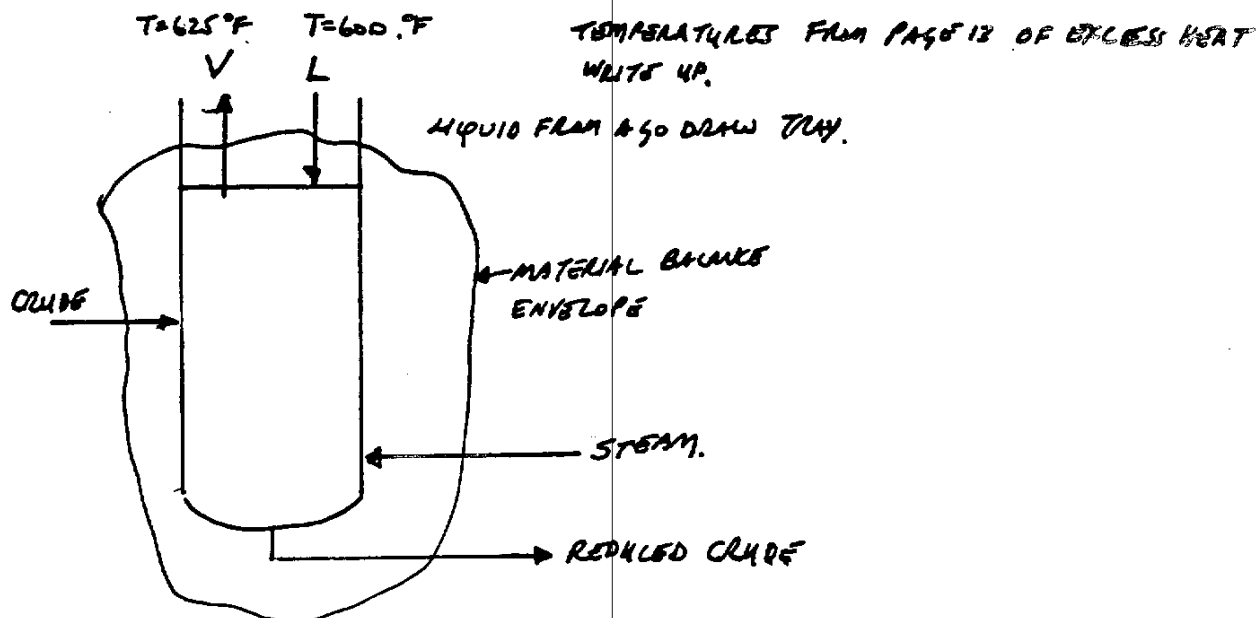
$$\text{DENSITY @ } 550^\circ \text{F} \quad d_L = (52.85)(0.78) = 41.23 \text{ LB/FT}^3$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER SIZE / TRAY LOADINGS

PAGE NO. 8
 DATE 10-15-91
 PROJECT EN MILLER

TRAYS BETWEEN A60 DRAW TRAY AND FLASH ZONE.

THIS TIME WE WILL DRAW OUR MATERIAL ENVELOPE FROM BOTTOM UP INSTEAD OF FROM TOP DOWN. THE ANSWER IS THE SAME.



MATERIAL BALANCE

$$V + (\text{REDUCED CRUDE}) = L + (\text{CRUDE}) + (\text{STEAM})$$

$$\text{INTERNAL REFLUX} = \underline{L} = \underline{575345 \text{ LB/HR}} \quad \text{SEE PAGE 29.}$$

$$V + 527895 = 575345 + 1235580 + 15875$$

$$\underline{V = 1298905 \text{ LB/HR}}$$

INTERNAL REFLUX HAS THE SAME PROPERTIES OF DRAW TRAY MATERIAL

$$MW = 303.9 \quad API = 33.9 \quad 401K = 11.87 \quad \text{SEE PAGE 5}$$

HYDROCARBON PORTION OF VAPOR RISING CONSISTS OF NET PRODUCTS LIFTED PLUS THE INTERNAL REFLUX BELOW A60 DRAW. SUMMARIZED

	LB/HR	MOL WT	MOL/HR
INTERNAL REFLUX L	575345	303.9	1893.2
UNSTABILIZED NAHT.	301375	102.9	2928.18
2GT FUEL	196035	187	1048.32
DIBSEL	113460	253	448.46
A60	96815	312	310.30
TOTAL	1283030	193.56	6628.46

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CLUDE UNIT DESIGN.
CLUDE TOWER SIZE / TRAY LOADINGS.

PAGE NO. 9
 DATE 10-15-91
 PROJECT CLUDE UNIT DESIGN.
 BY E. H. MILLER.

CALCULATE VAPOR DENSITY USING IDEAL GAS LAW. FIRST
 CALCULATE AVERAGE MOLECULAR WEIGHT OF STEAM / HYDROCARBON
 MIXTURE.

$$\text{STEAM} = \frac{15875}{18} = 882 \text{ MOL/HR}$$

$$\text{AV MOL WT} = \frac{1298905}{7510.46}$$

$$= 172.95$$

$$\text{HYDROCARBON} = \frac{1283030}{19356} = 662.46 \text{ MOL/HR}$$

$$\text{VAPOR TOTAL} = 7510.46 \text{ MOL/HR}$$

$$P = 14 + 14.7 = 28.7 \text{ PSIA} \quad T = 625 + 460 = 1085 \text{ }^{\circ}\text{R.}$$

$$\text{DENSITY } d_v = \frac{(172.95)(28.7)}{(10.73)(1085)} = \underline{\underline{0.426 \text{ LB/FT}^3}}$$

$$pT_c = 908 \text{ }^{\circ}\text{F} \quad \text{EXPANSION FACTOR} = 0.75$$

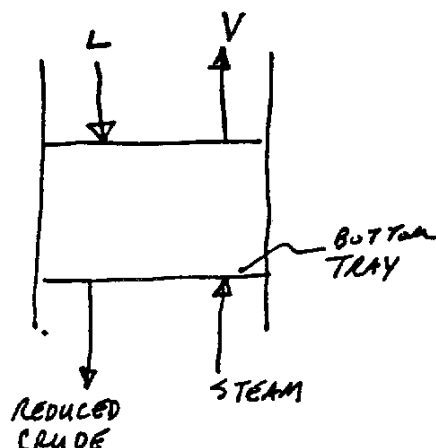
$$\text{Density of internal Reflux @ } 60^{\circ}\text{F} = (0.8555)(62.38) = 53.37 \text{ LB/FT.}$$

$$\text{DENSITY @ } 600^{\circ}\text{F} = d_L = (53.37)(0.751) = \underline{\underline{40.08 \text{ LB/FT}^3}}.$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT DESIGN
CRUDE TOWER SIZING / TRAY LOADINGS

PAGE NO. 10
 DATE 10-15-91
 PROJECT BY EN MILLER

DETERMINE LOADINGS IN THE BOTTOM SECTION OF CRUDE TOWER BELOW FLASH ZONE. THE BOTTOM TRAY LOADINGS ARE FOR THE VAPOR STEAM TO THE TOWER AND FOR THE ^{LIQUID} BOTTOMS FROM THE TOWER. THE TRAY LOADINGS FOR THIS SECTION OF THE TOWER ARE CALCULATED IN A FASHION SIMILAR TO THAT OF THE SIDELINE STRIPPERS.



THE HYDROCARBON PORTION OF THE VAPORS FROM THIS PORTION OF TOWER AT 10 VOL% OF LIQUID TO THE SECTION.

HYDROCARBON MATERIAL BALANCE.

$$L = V_{HC} + (\text{REDUCED CRUDE})$$

$$V_{HC} = 0.1 L$$

COMBINING

$$V_{HC} = \frac{(\text{REDUCED CRUDE}) (0.1)}{(0.9)} = \frac{(38100) (0.1)}{(0.9)} = 4233 \text{ BPSD}$$

FOR TRAY LOADING PURPOSES WE ASSUME THAT THE COMPOSITION OF THIS VAPOR IS THE SAME AS THE A90 PRODUCT. (PAGE III 7A)
 API = 29.5 MW = 312.

$$V_{HC} = (12.823)(4233) = 54280 \text{ LB/HR}$$

$$V_{HC} = \text{MOL/HR} = 54280 / 312 = 174 \text{ MOL/HR}$$

CUSTOMER ENGINEERING DESIGN SEMINAR

SUBJECT

CRUDE UNIT DESIGN.CRUDE TOWER SIZING / TRAY LOADINGSPAGE NO. 11DATE 10-15-91

PROJECT

BY E N MILLER

CALCULATE TOTAL VAPOR LOAD INCLUDING STEAM.

	LB/HR.	MO/HR.
V_{HC}	54280	174
STEAM.	<u>15875</u>	<u>882</u>
$V = \text{TOTAL VAPOR}$	70155	1056

$$\text{MIXTURE MOL WT} = 70155/1056 = .6643$$

$$P = 15 + 14.7 = 29.7 \text{ PSIA} \quad T = 715 + 460 = 1175 \text{ }^{\circ}\text{R.}$$

$$\text{DENSITY} = d_v = \frac{(66.43)(29.7)}{(10.73)(1175)} = \underline{\underline{0.157 \text{ LB/FT}^3}}$$

$$V = \underline{\underline{70155 \text{ LB/HR}}}$$

CALCULATE LIQUID LOAD.

	LB/HR	MW	MO/HR.	API
REDUCED CRUDE	527895	462.4	1141.65	17.5
STRIP OUT VAPOR	<u>54280</u>	<u>312</u>	<u>174</u>	<u>29.5</u>
$L = \text{LIQUID} = \text{TOTAL}$	582175	442.5	1315.65	18.62

$$\text{UOP } K = \left(\frac{527895}{582175} \right) (11.72) + \left(\frac{54280}{582175} \right) (41.87) = 11.73. \quad \text{CABP} \approx 903^{\circ}\text{F}$$

$$p T_c \approx 1165^{\circ}\text{F} \quad \text{EXPANSION FACTOR} = 0.764$$

$$\text{DENSITY AT } 60^{\circ}\text{F} = (0.9426)(62.34) = 58.8 \text{ LB/FT}^3$$

$$\text{DENSITY AT } 715^{\circ}\text{F} = d_L = (58.8)(0.764) = \underline{\underline{44.9 \text{ LB/FT}^3}}$$

$$L = \underline{\underline{582175 \text{ LB/HR.}}}$$

AT THE BOTTOM TRAY VAPOR LOAD IS STRIPPING STEAM.

$$V_s = 15875 \text{ LB/HR.}$$

ALLOW 1 PSI DP IN THIS SECTION $P = 16 + 14.7 = 20.7 \text{ PSIA}$.STEAM TO TOWER IS 50 PSIG (SATURATED). FROM STEAM TABLES $d_v = \underline{\underline{0.0220 \text{ LB/FT}^3}}$

$$\text{density of liquid @ } 675^{\circ}\text{F} = (0.9497)(62.38)(0.702) = \underline{\underline{46.33 \text{ LB/FT}^3}}$$

UOP17-1031A

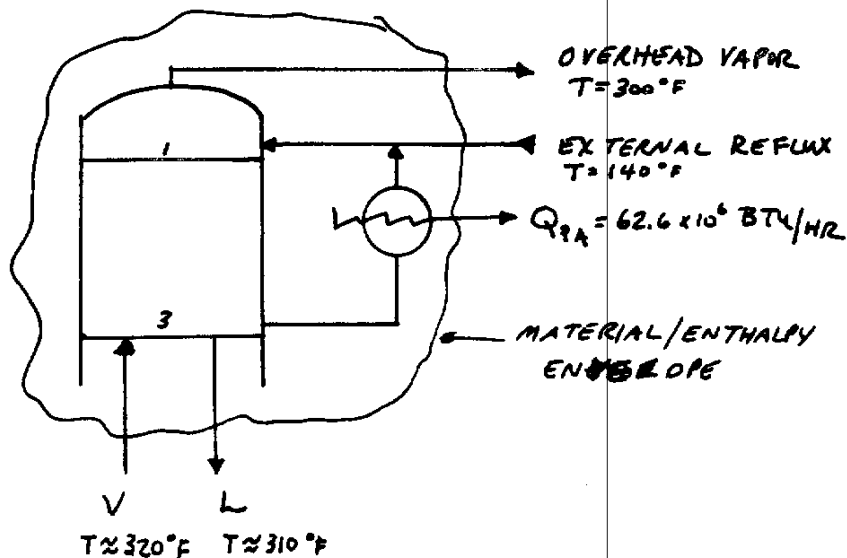
$$\text{REDUCED CRUDE } L = \underline{\underline{527895 \text{ LB/HR}}}$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT GRADE UNIT DESIGNGRADE TOWER STRIPING / TRAY LOADINGSPAGE NO. 12DATE 10-30-71

PROJECT

BY E. U. MILLER

LET'S REDO MATERIAL AND ENERGY BALANCE AT TOP OF COLUMN WITH PUMP AROUND IN PLACE.



MATERIAL BALANCE

$$V + (\text{EXTERNAL REFLUX}) = L + (\text{OVERHEAD VAPOR})$$

$$V = \text{STEAM} + \text{NET OVERHEAD PRODUCTS} + \text{INTERNAL REFLUX (L)}$$

$$(\text{OVERHEAD VAPOR}) = \text{STEAM} + \text{NET OVERHEAD PRODUCTS} + \text{EXTERNAL REFLUX.}$$

ENTHALPY BALANCE.

	IN (LB/HR)	OUT (LB/HR)	SPECIFIC ENTHALPY (BTU/LB)	IN (MM BTU/HR)	OUT (MM BTU/HR)
NET OVERHEAD PRODUCT	301375		272	81.974	79.262
STEAM	31270		1172.5	36664	36.361
EXTERNAL REFLUX (V)		146595	263		38.554
EXTERNAL REFLUX (L)	146595		44	6.450	
INTERNAL REFLUX (V)	L		272		
INTERNAL REFLUX (L)		L	148		
PUMP AROUND Q					62.6
TOTAL				125.088	216.777

$$\text{HEAT BROUGHT IN BY EXTERNAL REFLUX} = (216.777 - 125.088) = 91.689 \times 10^6 \text{ BTU/HR}$$

$$\text{INTERNAL REFLUX} = L = \frac{91.689 \times 10^6}{(272 - 148)} = \underline{\underline{739430 \text{ LB/HR.}}}$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT DESIGNPAGE NO. 13DATE 10-30-91

PROJECT

BY ENG MILLER

$$V = 739430 + 301375 + 31270 + 146595 - 146595$$

$$= \underline{\underline{1072075 \text{ LB/HR}}}$$

VAPOR DENSITY

$$\text{MOL OF HYDROCARBON} = \frac{1040805}{102.9} = 10115$$

$$\text{MOL OF STEAM} = \frac{31270}{18} = 1737$$

$$\text{TOTAL} = 11852 \text{ MOL/HR}$$

$$\text{AV. MOL WT} = \frac{1072075}{11852} = 90.46$$

$$\text{DENSITY OF VAPOR} = d_v = \frac{(90.46)(24.7)}{(10.73)(780)} = 0.267 \text{ LB/FT}^3$$

$$\text{DENSITY OF LIQUID AT } 60^\circ\text{F} = (0.7209)(62.38) = 44.97 \text{ LB/FT}^3$$

$$\text{EXPANSION FACTOR @ } 310^\circ\text{F} = 0.83$$

$$\text{DENSITY OF LIQUID AT } 310^\circ\text{F} = (0.83)(44.97) = 37.3 \text{ LB/FT}^3$$

LIQUID LOAD AT NAPHTHA PUMPAROUND TRAY.

$$= (\text{INTERNAL REFLUX}) + (\text{NAPHTHA PUMPAROUND RATE})$$

$$= 739430 + 934330 = \underline{\underline{1673760 \text{ LB/HR}}}$$

AT THE JET FUEL DRAW COLUMN LIQUID LOAD

$$\text{LIQUID LOAD} = \text{INTERNAL REFLUX} + \text{DRAW TRAY LIQUID}$$

$$= 391975 + 217225 = 609200 \text{ LB/HR}$$

LIQUID LOAD AT DIESEL PUMPAROUND TRAY

$$= (\text{INTERNAL REFLUX}) + \text{DIESEL PUMPAROUND RATE} + \text{DIESEL TO STRIPPER}$$

$$= 670240 + 441175 + 125740 = \underline{\underline{1237155 \text{ LB/HR}}}$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT DESIGNPAGE NO. 14DATE 10-30-91

PROJECT

BY E N MILLERCRUDE TOWER SIZING/TRAY LOADINGS

LIQUID LOAD AT A90 PUMP AROUND TRAY

$$\begin{aligned}
 &= \text{INTERNAL REFLUX} + \text{A90 PUMP AROUND RATE} + \text{A90 TO STRIPPER} \\
 &= 575345 + 222225 + 107260 \\
 &= \underline{\underline{904830 \text{ LB/HR}}}
 \end{aligned}$$

LOADS FOR A90 STRIPPER.

USE LIQUID TO STRIPPER FOR LIQUID LOAD = 107260 LB/HR.
 VAPOR FROM STRIPPER FOR VAPOR LOAD = 13940 LB/HR.

$$\text{VAPOR DENSITY} = d_v = \frac{(58.9)(14 + 14.7)}{(10.73)(595 + 460)} = 0.149 \text{ LB/FT}^3$$

$$\text{LIQUID DENSITY @ } 60^\circ = (12.8762)(62.38) = 54.657 \text{ LB/FT}^3$$

$$\text{EXPANSION FACTOR} = 0.773$$

$$\text{LIQUID DENSITY @ } 600 = (54.657)(0.773) = \underline{\underline{42.3 \text{ LB/FT}^3}}$$

LOADS FOR DIESEL STRIPPER.

USE LIQUID TO STRIPPER FOR LIQUID LOAD = 125740 LB/HR.
 VAPOR FROM STRIPPER FOR VAPOR LOAD = 16520 LB/HR

$$\text{VAPOR DENSITY} = d_v = \frac{(55.7)(13 + 14.7)}{(10.73)(545 + 460)} = 0.143 \text{ LB/FT}^3$$

$$\text{Liquid Density @ } 60^\circ\text{F} = (0.8473)(62.38) = 52.855 \text{ LB/FT}^3$$

$$\text{EXPANSION FACTOR} = 0.802$$

$$\text{LIQUID DENSITY @ } 550^\circ\text{F} = (0.802)(52.855) = \underline{\underline{42.4 \text{ LB/FT}^3}}$$

LOADS FOR JET FUEL STRIPPER.

$$\text{LIQUID} = 217225 \text{ LB/HR}$$

$$\text{VAPOR} = 28850$$

$$\text{density @ } 60 = (0.8095)(62.38) = 50.5$$

$$\text{EXPANSION FACTOR} = 0.815$$

$$\text{density @ } 425^\circ\text{F} = (50.5)(0.815) = 41.2.$$

$$d_v = \frac{(57.3)(11.5 + 14.7)}{(10.73)(420 + 460)} = 0.142$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CANOE UNIT DESIGNPAGE NO. 15DATE 10-30-91

PROJECT

BY E N MILLERCLUDE TOWER STRIPPER / TRAY LOADINGS.

SECTION TRAYS.	NAPHTHA P. A. 1-3	4-12	13-19	DIESEL P. A. 20-22	23-27	AGO P. A. 28-30
VAPOR:						
LB/HR.	1072075	1072075	912995	1300480	1300480	1298905
LB/FT ³	0.267	0.267	0.332	0.395	0.395	0.426
CFS	1115	1115	764	915	915	847
LIQUID:						
LB/HR	1673760	739430	391975	1237155	670242	904830
LB/FT ³	37.3	37.3	41.2	41.2	41.2	40.1
GPM	5595	2472	1186	3744	2028	2813
TEMP. °F	320°F	425	550	625	625	625
PRESSURE, PSIG	10	11.5	13	13	14	14

SECTION TRAYS.	31-36	37-39	40	KEROSENE STRIPPER ALL	DIESEL STRIPPER ALL	AGO STRIPPER ALL
VAPOR						
LB/HR.	1298905	70155	15875	28850	16520	13940
LB/FT ³	0.426	0.157	0.022	0.142	0.143	0.149
CFS	847	124	200	56.4	32.1	26.0
LIQUID						
LB/HR	575345	582175	527895	217225	125740	107260
LB/FT ³	40.1	44.9	46.3	41.2	42.4	42.3
GPM	1789	1617	1422	657	370	316
TEMP. °F	715		675	425	550	600
PRESSURE, PSIG	15		16	125	14	15

CRUDE TOWER SIZING / TRAY LOADINGS

PRELIMINARY SIZING OF CRUDE TOWER. USE VALVE TRAYS AND KOCH TRAY SIZING METHOD. DO FRACTIONATING TRAYS FIRST AND THEN LOOK AT PUMPAROUND TRAYS.

FOR FIRST ESTIMATE, WE COULD USE A "BARREL" FACTOR APPROACH.

$$D = \sqrt{\frac{\text{LIFT (BPSD)}}{130}} = \sqrt{\frac{(100,000 - 38100)}{130}}$$

$$= 21.8 \text{ FT} \Rightarrow 22 \text{ FT.}$$

TRY KOCH PRELIMINARY SIZING METHOD USING LOADINGS FOR TRAYS 4-12.

$$\text{CALCULATE } V_{\text{LOAD}} = (\text{ACFC}) \sqrt{\frac{d_v}{d_L - d_v}} = (1115) \sqrt{\frac{0.267}{37.3 - 0.267}}$$

$$= 94.7$$

FOAMING TENDENCY OF SYSTEM LOW, USE SYSTEM FACTOR = 1.0
USE 30" TRAY SPACING TRAY SPACE FACTOR = 1.142.

$$\text{CORRECTED } V_{\text{LOAD}} = \frac{94.7}{(1.142)(1.0)} = 82.9.$$

$$\text{CORRECTED LIQUID RATE} = \frac{\text{GPM}}{(\text{TRAY SPACING FACTOR})(\text{SYSTEM FACTOR})}$$

$$= \frac{2472}{(1.142)(1.0)} = 2165 \text{ GPM.}$$

CHART A IS OUT OF RANGE CONTINUE WITH DETAIL SIZING

$$\text{DOWNCOMER MINIMUM AREA} = \left(\frac{2472}{175}\right) \left(\frac{24}{30}\right) \left(\frac{1}{1}\right) = 11.3 \text{ FT}^2$$

THIS IS BASED ON 85% OF DOWNCOMER LIMIT. WE GENERALLY ARE SOMEWHAT MORE CONSERVATIVE, BECAUSE OF UNCERTAINTY IN PREDICTION OF DOWNCOMER FLOW LIMITS, AND LIMIT MINIMUM TO 75% OF CALCULATED DOWNCOMER LIMIT

$$\text{DOWNCOMER MINIMUM AREA} = (11.3) \left(\frac{85}{75}\right) = 12.8 \text{ FT}^2$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRAKE UNIT DESIGNPAGE NO. 17DATE 10-30-91PROJECT 11BY E N MILLERCRAKE TOWER SIZING / TRAY LOADINGS.

SINCE THIS IS A LARGE TOWER RATHER THAN USE THE KOCH CHARTS WE WILL USE THE CORD SEGMENT DATA IN GENERAL SECTION OF THE CHART BOOK.

FOR 22 FT DIAMETER TOWER $A_{TOWER} = \frac{\pi 22^2}{4} = 380.133 \text{ FT}^2$

FOR SINGLE PASS TRAY $A_D/A_T = 128/380.133 = 0.0337$

FROM CHARTS $L/D = 0.5252$ $H/D = 0.0745$

$L = (0.5252)(22) = 11.55 \text{ FT}$ $H = 19.668" \Rightarrow$

$\text{GPM/FT WIER} = \frac{2472}{11.55} = 213.9$ TOO HIGH

WE WILL END UP WITH EXCESSIVE PRESSURE DROP. UOP LIKES TO LIMIT GPM/FT WIER TO 120 MAX FOR NEW DESIGNS. THIS MEANS THAT FOR THIS COLUMN WE NEED A MINIMUM OF 2 PASS TRAYS.

IF WE KEEP THE DOWNCOMER AREA THE SAME FOR EACH OF THE SIDE DOWNCOMERS THEN $\text{GPM/FT WIER} = \frac{213.9}{2} = 106.9 \Rightarrow 107$

WHICH MEETS OUR CRITERIA.

IF WE SET SIDE DOWNCOMER "H" = 20 inches $H/D = 0.07576$.

$A_D = (0.034556)(380.133) = 13.136 \text{ FT}^2/\text{SIDE}$ $L/D = 0.5292$
 $A_D/A_T = 0.034556$

$A_D \text{ TOTAL} = (13.136)(2) = 26.272 \text{ FT}^2$ $L = (0.5292)(22) = 11.64 \text{ FT}.$

TO BALANCE DOWN COMER LOADS AREA OF CENTER DOWNCOMER SHOULD BE SAME AS SIDE DOWNCOMERS.

$\text{WIDTH} = \frac{26.272}{22} = 1.194 \text{ FT} \Rightarrow 14.33" \text{ MAKE } 15"$

$\text{BUBBLING AREA} = 380.133 - 26.272 - \left(\frac{15}{2}\right)(22) = 326.361 \text{ FT}^2.$

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGNPAGE NO. 18DATE 10-30-91

PROJECT

BY E. N. MILLERCRUDE TOWER SPRING / TRAY LOADINGS.

NOW CALCULATE PERCENT OF CALCULATED TRAY CAPACITY USING
EQUATIONS ON PAGE 9 OF KOCH MANUAL

$$C_{sf(met)} = \left(\frac{ACFS}{A_b} \right) \sqrt{\frac{d_v}{d_L - d_v}} = \frac{V_{LOAD}}{A_b \text{ (BUBBLING AREA)}}$$

$$= \frac{94.7}{326.31} = 0.290.$$

$$GPM/LW = \frac{(2472)}{(2)(11.64)} = 106.2.$$

$$C_{sf(met)} / (GPM/LW) = \left(\frac{0.290}{106.2} \right) = 0.0027.$$

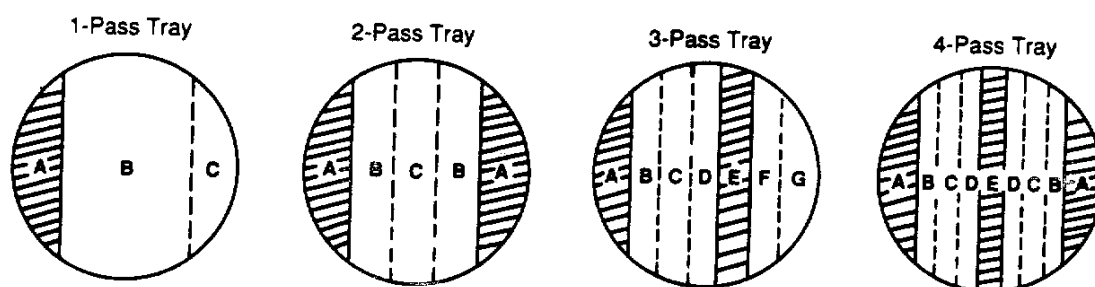
$$\% \text{ CAPACITY} = \frac{100}{(1.142)(1)(0.51)} \left[(0.290) + (0.001327)(106.2) \right]$$

$$= \underline{\underline{74 \%}}$$

$$\text{DOWN COMMER CAPACITY} = \left(\frac{2472}{26.272} \right) \left(\frac{85}{175} \right) \left(\frac{1}{1} \right) \left(\frac{24}{30} \right)$$

$$= \underline{\underline{36.6 \%}}$$

WE WILL NOW LOOK BRIEFLY AT THE PUMPAROUNDS. PROBABLY THE LIMITING SECTION WILL BE THE NAPHTHA PUMPAROUND. SINCE THE LIQUID RATE IS MORE THAN TWICE THAT OF THE SECTION WHICH WE COMPLETED AND SINCE DOWNCOMERS WERE WIDER LENGTH LIMITED, WE NEED FOUR PASS TRAYS IN THIS SECTION. FOR MULTIPASS TRAYS WE USE EQUAL FLOW PATH LENGTHS.



FLOW PATHS

A_{DA}

A_{DE}

A_{DC}

4

0.25

0.50

0.50

$$\text{DOWN COMER AREA MINIMUM} = \frac{(5595)}{(175)} \left(\frac{24}{30} \right) \left(\frac{85}{75} \right) = 28.99 = 29 \text{ FT}^2$$

$$\text{SIDE DOWN COMER} = \frac{(29)(.25)}{380.133} = 0.0190 = H/D \quad H/D = 0.0505$$

$$H = 13.33 \text{ inches} \Rightarrow 14"$$

$$\text{CENTER DOWN COMER} = \left(\frac{29}{22} \right) \left(\frac{12}{2} \right) = \frac{15.8}{2} \text{ inches} = \frac{16}{2} = 8 \text{ inches}$$

$$\text{REMAINING DISTANCE} = 132 - 14 - 8 = 110 \text{ in} \quad \text{IF DWN} = 10"$$

$$H_1/D = \frac{(14+52)}{(264)} = 0.235 \quad A_0/A_T = 0.1955 \quad H_2/D = \frac{(14+52+10)}{(264)} = 0.2879 \quad A_0/A_T = 0.238$$

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CANDE UNIT DESIGN.PAGE NO. 20DATE 10-30-91PROJECT BY E N MILLERCANDE TOWER SIZING / TRAY LOADINGS.

$$\begin{aligned} \text{AREA DOWN COMER} &= (0.2383 - 0.1955)(380.133) \\ &= 16.27 \quad \text{OK.} \end{aligned}$$

$$\begin{aligned} \text{AREA BUBBLING} &= 380.133 - (2)(7.76) - 2(16.27) - \left(\frac{8}{12}\right)(22) \\ &= 380.133 - 62.73 = 317.403 \text{ FT}^2. \end{aligned}$$

$$C_{SF(\text{net})} = \frac{(1115)}{(317.403)} \sqrt{\frac{0.267}{(37.3 - 0.267)}} = 0.2983$$

$$\begin{array}{ll} \text{WEIR LENGTH} & \text{SIDE DOWNCOMER } L/D = 0.4481 \quad L = 9.858 \text{ FT.} \\ & \text{CENTER DOWNCOMER} \quad L = 22 \text{ FT.} \end{array}$$

$$L_W (\text{TOTAL}) = (9.858 + 22)(2) = 63.716 \text{ FT.}$$

$$\text{GPM}/L_W = 5595 / 63.716 = 87.81 \text{ GPM/FT}$$

$$\begin{aligned} \% \text{ CAPACITY} &= \frac{100}{(1.142)(1)(0.51)} \left[0.2983 + (0.001327)(87.81) \right] \\ &= \underline{\underline{71.22 \%}} \end{aligned}$$

$$\begin{aligned} \text{DOWN COMER CAPACITY} &= \frac{5595}{2(7.76 + 16.27)} \left(\frac{85}{175} \right) \left(\frac{24}{30} \right) = \\ &= 45.24 \% \end{aligned}$$

$$\begin{aligned} \text{DOWN COMER CAPACITY} &= \frac{5595}{(16.27)(2)} \left(\frac{85}{175} \right) \left(\frac{24}{30} \right) \\ &= 66.81 \% \end{aligned}$$

MAKE CANDE COLUMN 22 FT IN DIAMETER
WITH 30" TRAY SPACING.

CRUDE TOWER SIZING / TRAY LOADINGS

LOOK AT BOTTOM SECTION BELOW FLASH ZONE. USE INTERNAL SWAGE.

$$\text{DOWN COMER AREA} = \frac{(1617)(85)(24)}{(175)(50)(30)} = 12.57 \text{ FT}^2$$

TRAY 14' INTERNAL DIAMETER $A_T = 153.938$.

$$A_D/A_T = \frac{12.57}{153.938} = 0.0816 \quad H/D = 0.1360 \quad L/D = 0.6856$$

FOR ONE PASS TRAYS $L = (0.6856)(14) = 9.598$

$$\text{GPM}/L = 1617/9.598 = 168.46. \quad \text{TOO HIGH}$$

FOR TWO PASS TRAYS $L_{\text{TOTAL}} = (9.598)(2) = 19.196$.

$$\text{GPM}/L_{\text{TOTAL}} = 1617/19.196 = 84.24 \text{ GPM}/\text{FT OK.}$$

$$H = (0.136)(14) = 22.848 \text{ inches.} \Rightarrow 23 \text{ inches.} \quad H/D = 0.1369$$

$$A_D/A_T = 0.0824 \quad A_D = 12.684 \text{ FT}^2.$$

$$\text{CENTER DOWN COMER} = \left(\frac{12.684}{14} \right) (12) = 10.87 \text{ inches} = 11 \text{ inches.}$$

$$A_{\text{BUBBLING}} = 153.938 - (2)(12.684) - \left(\frac{11}{12} \right) (14) = 115.736$$

$$C_{SF(\text{net})} = \frac{124}{115.736} \sqrt{\frac{0.157}{44.9 - 0.157}} = 0.0635$$

$$\text{GPM}/L_{\text{NW}} = 84.24.$$

$$\% \text{ CAPACITY} = \frac{100}{(1.142)(1)(0.51)} \left[0.0635 + (0.001327)(84.24) \right] = 29.84 \%$$

$$\text{DOWN COMER CAPACITY} = \frac{1617}{\left(\frac{11}{12} \right) (14)} \left(\frac{85}{175} \right) \left(\frac{24}{30} \right) = 49 \%$$

ON A PRELIMINARY BASIS WE WILL USE 14 FT FOR THIS SECTION OF TOWER

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGNPAGE NO. 22DATE 10-30-91

PROJECT

BY E N MILLERCRUDE TOWER SIZING / TRAY LOADINGSSIZE JET FUEL STRIPPER

$$\text{CALCULATE } V_{\text{LOAD}} = .56.4 \sqrt{\frac{0.142}{41.2 - 0.142}} = 3.32$$

USE SYSTEM FACTOR = 1.0 USE 24" TRAY SPACING.

FROM CHART A 5'0" DIAMETER SINGLE PASS TRAYS.
 WE TEND TO DESIGN STRIPPERS "LOOSE" SO LOOK AT
 6'0" DIAMETER SINGLE PASS TRAY.

$$\text{DOWN COMER AREA} = \left(\frac{657}{175}\right) \left(\frac{85}{75}\right) = 4.25 \text{ FT}^2$$

FROM CHART D/E CORD HEIGHT = 15" CORD LENGTH = 9.85 FT.

$$\text{GPM/L} = 657/4.85 = 135.5 \text{ HIGH.}$$

IF WE GO TO CORD HEIGHT OF 22" L = 5.51 GPM/L = 119.24

$$\text{DOWN COMER AREA} = 7.25 \text{ } \phi$$

$$\text{BUBBLING AREA} = 28.27 - (2)(7.25) = 13.77 \text{ FT}^2$$

$$C_{FS(WET)} = \frac{3.22}{13.77} = 0.234$$

FROM CHART C % CAPACITY = 77%.

LOOK AT TWO PASS TRAYS SET SIDE DOWNCOMER CORD HEIGHT AT 15"
 AND CENTER DOWNCOMER AT 9 INCHES (MINIMUM WIDTH IS 8")

$$\text{BUBBLING AREA} = (28.27) - (2)(4.25) - \frac{(9)(6)}{12} = 15.27.$$

$$C_{FSMET} = \frac{3.22}{15.27} = 0.211$$

$$\text{GPM/L} = \frac{657}{(2)(4.85)} = 67.7 \quad \% \text{ CAPACITY} = \underline{59}$$

USE 6'0" DIAMETER AND 2 PASS TRAYS WITH
 24 inch TRAY SPACING FOR JET FUEL STRIPPER.

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT CRUDE UNIT DESIGNPAGE NO. 23DATE 10-30-91

PROJECT

BY EN MILLERCRUDE TOWER SIZING / TRAY LOADINGSSIZE DIBSOL STRIPPER

$$\text{CALCULATE } V_{LATO} = 32.1 \sqrt{\frac{0.143}{42.4 - 0.143}} = 1.87$$

FROM CHART A DIAMETER = 4'0" FOR "LOOSE" SIZING
TRY 5'0" SINGLE PASS.

$$A_D \text{ MIN} = \left(\frac{370}{175} \right) \left(\frac{85}{50} \right) \approx 3.5 \text{ FT}^2.$$

CORD HEIGHT = 14.0 inches CORD LENGTH = 4.25 FEET

$$\text{GPM/FT} = 370 / 4.25 = 87.06 \text{ GPM/FT OK.}$$

$$\text{BUBBLING AREA} = 19.63 - (3.4)(2) = 12.83 \text{ FT}^2.$$

$$C_{FS(Net)} = \frac{1.87}{12.83} = 0.146. \quad \% \text{ CAPACITY} = \underline{\underline{51}}.$$

USE 5'0" DIAMETER WITH SINGLE PASS TRAYS.

SIZE AGO STRIPPER

SINCE THE LOADS FOR THESE TWO STRIPPERS
ARE SIMILAR, WE WILL USE THE SAME DIAMETER.

MAKE AGO STRIPPER 5'0" DIAMETER WITH SINGLE PASS TRAYS.

THE STUDENT CAN CONFIRM THIS FOR HIMSELF.



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HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR

CRUDE UNIT DESIGN

100000 BPSD - ARABIAN LIGHT CRUDE

CRUDE EXCHANGE AND HEATER SECTION

ENGINEERING DEPARTMENT

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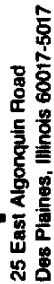
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PROJECT

BY EN MILLER

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STREAM	°API (LB/GAL)	K	MW	FLOWING SP. GR. (LB/CF)	°F	PSIG (MM HG ABS)	BPSD @ 60° F (10 ⁶ SCF/D)	FLOWING GPM (CFS)	MOLS PER HR.	LB/HR	BTU/ LB	10 ⁶ BTU/HR
CRUDE TO JET FUEL PRODUCT EXCH.	35.6	11.9	210.3		60		100000		5876.9	1235580	0	0
EXCHANGER DUTY											→	30.3
CRUDE TO NAPHTHA PA EXCHANGE					110		100000			1235580	245	30.3
EXCHANGER DUTY											→	62.6
CRUDE TO DIESEL PRODUCT EXCHANGE					210		100000			1235580	75	92.9
EXCHANGER DUTY											→	19.6
CRUDE TO DIESEL PA EXCHANGE					240		100000			1235580		112.5
EXCHANGER DUTY											→	13.5
CRUDE TO DESALTER					260		100000			1235580		126.0
DESALTER FEED WATER TO EXCHANGE					100		5000			72975		2.92
EXCHANGER DUTY											→	7.30
DESALTER FEED WATER TO DESALTER					200		5000			72975	140	10.22
DESALTER EFFLUENT WATER TO EXCHANGE					250					72975		13.87
EXCHANGER DUTY											→	7.30
DESALTER EFFLUENT WATER TO S.W. SUMMER					150					72975		6.57
DESALTERED CRUDE FROM DESALTER					250*		100000			1235580		119.9
* NOTE THERE IS DIFFERENCE IN ENTHALPY BALANCE ACROSS DESALTER.												



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100 000 BPSD - ARABIAN LIGHT CRUDE

CRUDE EXCHANGE AND HEATER SECTION

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100 000 BPSD - ARABIAN LIGHT CRUDE

CRUDE COLUMN SECTION

ENGINEERING DEPARTMENT

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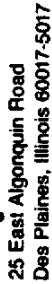
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DESALTED CRUDE TO FLASH ZONE	35.6	11.9	210.3		715	15	100,000		5876.9	1235380	454	560.8
TOTAL VAPOR.			154.1				(4409)		4447.3	746885	502	375.1
UNSTABILIZED NAPHTHA.	64.8	12.25	102.9						2928.2	301375	518	156.2
JET FUEL	42.8	11.97	187						1048.3	196035	500	98.0
DIESEL	35.0	11.93	253						446.5	113460	442	55.8
AGO	29.5	11.87	312						310.3	96815	480	46.5
OVER FLASH VAPOR.	26.5	11.87	350						112.0	39200	475	18.6
TOTAL LIQUID			474.6				35100		1029.6	444195	380	185.7
STEAM TO CRUDE COLUMN			18		298	50			881.9	15875	1152	18.3
REDUCED CRUDE TO CRUDE EXCH.	17.5	11.72	462.4		675		38100		1141.65	527895	363	191.6
AGO PUMP/ROUND TO CRUDE EXCHANGE	30.0	11.87	303.9		600		17383		731.2	222225	320	71.1
EXCHANGER DUTY												30.0
AGO PUMP/ROUND RETURN TO COLUMN	30.0	11.87	303.9		400		17383		731.2	222225	185	41.1
UNSTRIPPED AGO TO STRIPPER	30.0	11.87	303.9		600		8390		392.9	107260	320	34.3
STEAM TO AGO STRIPPER			18		298		50		194.2	3495	1152	4.0
VAPORS FROM AGO STRIPPER			58.9		595		(2.154)		236.8	13940	624	8.7
STEAM			18		595				194.2	3495	1354	4.6
HYDROCARBON.	34.5	11.87	245		595				426	10445	395	4.1
AGO PRODUCT TO CRUDE EXCHANGE	29.5	11.87	312		582		7550		310.3	96815	306	29.6



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UNSTRIPPED JET FUEL TO STRIPPER	43.3	11.97	183.3		425		18390		1185	217225		45.7
STEAM TO JET FUEL STRIPPER			18		298	50			4255	7666		8.8
VAPORS FROM JET FUEL STRIPPER			51.3		420		(5.1)		5622	28850		15.9
STEAM			18		420				4255	7660	1220	9.4
HYDROCARBON	47.8	11.97	155		420				136.7	21190	307	6.5
JET FUEL PRODUCT TO CRUDE EXCH.	42.8	11.97	187		408		16550		1048.3	196035	197	38.6
NAPHTHA PUMPAROUND TO CRUDE EXCH.					310					934330	148	138.3
EXCHANGER DUTY												62.6
NAPHTHA PUMPAROUND RETURN TO COLUMN					210					934330	81	95.7
COLUMN OVERHEAD VAPOR TO AIR COND.			78.7		300	10	(55.39)		6089.6	479240	321.8	154.2
STEAM			18						1737.1	31270	1162.8	36.4
UNSTABILIZED NAPHTHA	64.8	12.25	102.9						2928.2	301375	263	79.3
REFLUX	64.8	12.25	102.9						1424.3	146595	263	38.5
AIR CONDENSER DUTY												132.0
MATERIAL TO RECEIVER			78.7		140	5			6089.6	479240	46.3	22.2
REFLUX TO COLUMN	64.8	12.25	102.9		140				1424.3	146595	44	6.4
UNSTABILIZED NAPHTHA TO DEBUT.	64.8	12.25	102.9		140		28650		2928.2	301375	44	13.3
WATER TO S.W. STRIPPER			18		140				1937.1	31270	80	2.5



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100000 BPSD - ARABIAN LIGHT CRUDE

PRODUCT EXCHANGE AND COOLER CIRCUIT

ENGINEERING DEPARTMENT

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REDUCED CRUDE TO HOT CRUDE EXCH.	17.5	11.72	462.4		675		38100	1141.65		527895	363	191.6
EXCHANGER DUTY											→	93.3
REDUCED CRUDE TO COLD CRUDE EXCH.					412					527895	176.1	94.3
EXCHANGER DUTY											→	21.0
REDUCED CRUDE TO COOLER.					340					527895	138.9	73.3
COOLER DUTY											→	45.8
REDUCED CRUDE TO STORAGE					195					527895	52	27.5
AGO PRODUCT TO CRUDE EXCHANGE	29.5	11.87	312		582		7550		310.3	96815	306	29.6
EXCHANGER DUTY											→	14.6
AGO PRODUCT TO COOLER					350					96815	155	15.0
COOLER DUTY											→	12.2
AGO PRODUCT TO STORAGE	29.5	11.87	312		120	50				96815	29	2.8
DIESEL PRODUCT TO CRUDE EXCHANGE	35.0	11.93	253		530				444.5	113460	272	31.5
EXCHANGER DUTY											→	19.6
DIESEL PRODUCT TO COOLER					260					113460	104.9	11.9
COOLER DUTY											→	8.6
DIESEL PRODUCT TO STORAGE	35.0	11.93	253		120	50				113460	29	3.3



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PRODUCT EXCHANGE AND COOLER CIRCUIT

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