

CUSTOMER ENGINEERING DESIGN SEMINAR
SUBJECT CRUDE UNIT REVAMPPAGE NO. _____
DATE NOV 3, 1991
PROJECT _____
BY EN MILLER1 LINEAMIENTOS GENERALES
POR EL ANÁLISIS DEL PROBLEMA

ENGINEERING

DESIGN SEMINAR

CRUDE UNIT REVAMP

(CLASS NOTES)

NOV

1991

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT REVAMPPAGE NO. 1DATE NOV 3, 1991

PROJECT

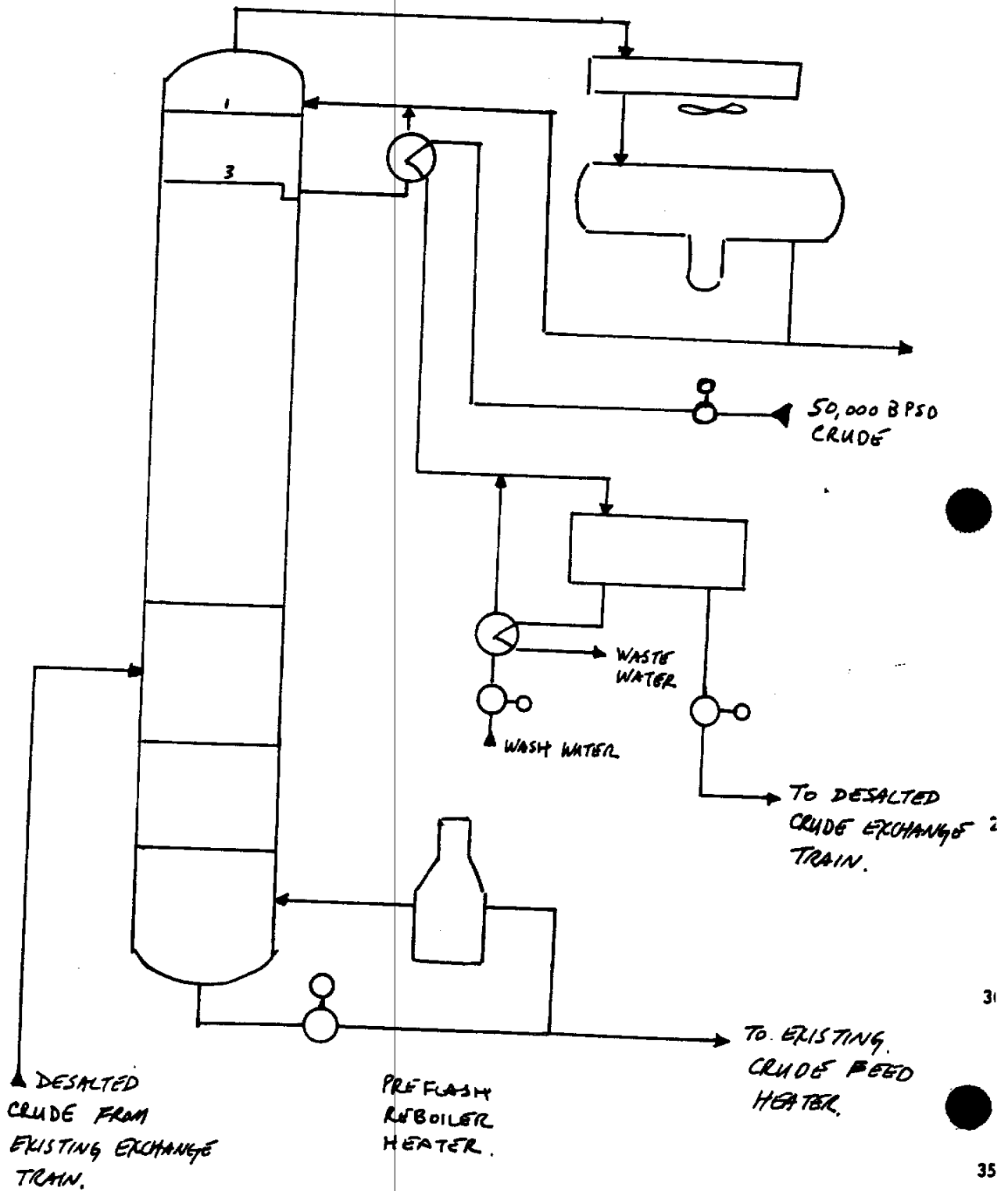
BY E N MILLER

WE ASSUME THE CRUDE UNIT WHOSE DESIGN WE OUTLINED IN THE PREVIOUS SESSIONS HAS BEEN BUILT AND OPERATING FOR A NUMBER OF YEARS. THERE IS A PROPOSAL TO EXPAND THE CRUDE UNIT CAPACITY FROM 100,000 TO 150,000 BPSD. OUR JOB IS TO DETERMINE IF THE PROPOSAL IS FEASIBLE AND TO MAKE SUGGESTIONS AS TO HOW THE REVAMP MIGHT BE DONE.

AFTER SOME INITIAL DISCUSSIONS, IT WAS EVERYONE'S OPINION THAT WE WOULD NEVER GET AN ADDITIONAL 50% THROUGH THE EXISTING CRUDE TOWER. SOME MEANS MUST BE FOUND TO MINIMIZE THE ADDITIONAL LOAD ON THE EXISTING TOWER. DURING OUR INITIAL DESIGN WORK WE FOUND THAT THE MOST HIGHLY LOADED PORTION OF THE TOWER WAS AT THE TOP. AT THIS POINT IT WAS SUGGESTED THAT A PREFLASH TOWER COULD BE ADDED TO THE CRUDE UNIT TAKING MOST OF THE NAPHTHA OVERHEAD AND THEN SENDING THE TOPPED CRUDE TO THE EXISTING CRUDE TOWER FOR RECOVERY OF PRODUCTS. THERE ARE TWO COMMON OPTIONS FOR USE OF PREFLASH TOWERS: (1) CRUDE TYPE OF TOWER WHERE ALL THE HEAT IS PUT IN THE FEED AND (2) REBOILED TOWER. WE WILL EVALUATE THE SECOND OPTION.

IT WAS ALSO DECIDED FOR THIS STUDY TO ADD A NEW DESALTER FOR THE INCREMENTAL 50000 BPSD CAPACITY SUPPLYING FEED PREHEAT FROM PREFLASH TOWER. AFTER THE DESALTER THE INCREMENTAL CAPACITY WILL BE INTEGRATED INTO THE EXISTING CRUDE PREHEAT TRAIN, REVAMPING AS REQUIRED. IT WAS ALSO DECIDED THAT THE PREFLASH FRACTIONATOR WOULD BE OPERATED AS A "DRY" FRACTIONATOR WITH NO STEAM STRIPPING.

A SKETCH OF THE PROPOSED PREFLASH TOWER SCHEME IS ATTACHED.



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BY EN MILLER

1. TYPICALLY FOR PREFLASH TOWERS, ABOUT 2 VOLUME % BASED ON CRUDE WILL NOT BE STRIPPED OUT. AFTER EVALUATION OF OPERATING DATA WE ESTIMATED THAT THE HEAVY NAPHTHA WOULD HAVE THE FOLLOWING PROPERTIES: BOILING RANGE 300-385, UOP K = 11.97 API = 50.3 MN = 145. SG = 0.770.

PREPARE SUMMARY OF NAPHTHA SPLIT BETWEEN PREFLASH TOWER AND CRUDE TOWER SIMILAR TO THE TABLE BELOW

	°API	UOP K	MOB WT	B PSD	LB/HR	MOL/HR
PF NAPHTHA (PREFLASH NAPHTHA)	61.33	12.14	108.6	34508	369072	3396.8
HVY NAPHTHA (CRUDE NAPHTHA)	50.3	11.97	145	3000	34018	234.6
NAPHTHA	60.4	12.13	111	37470	403090	3631.44

2. BASED ON THIS PRODUCT SPLIT PREPARE A YIELD STATEMENT SIMILAR TO THAT DONE FOR THE CRUDE DESIGN (III-4A). IN THIS CASE, THE YIELDS WILL BE IN TWO PARTS - PREFLASH OVERHEAD (PF NAPHTHA AND LIGHTER) AND FLASHED CRUDE (LIQUID FROM PREFLASH) TOWER. SUMMARY TABLE WILL LOOK SOMETHING LIKE THE FORM SHOWN BELOW.

	API	UOP K	MN	B PSD	LB/HR	MOL/HR	LV %
CRUDE							
ESTIMATED YIELDS							
C3							
...							
PF NAPHTHA							
FLASHED CRUDE							
HVY NAPHTHA							
...							

3. SINCE THE SHIFT IN NAPHTHA PROPERTIES ARE NOT LIKELY TO BE GREAT, WE ANTICIPATE BEING ABLE TO OPERATE THE PREFLASH TOWER OVERHEAD RECEIVER AT THE SAME CONDITIONS THAT WERE USED FOR THE CRUDE TOWER OVERHEAD RECEIVER IN THE CRUDE DESIGN. THESE CONDITIONS WERE A OPERATING PRESSURE OF 5 PSIA AND AN OPERATING TEMPERATURE OF 140°F. THE STUDENT IS TO CONFIRM THAT THE PREFLASH TOWER NET OVERHEAD IS ALL LIQUID AT THESE CONDITIONS. EQUILIBRIUM KTI DATA FOR THE PF NAPHTHA ARE SUMMARIZED IN THE FOLLOWING TABLE.

TEMPERATURE °F	KTI @ 14.696 PSIA	KTI @ 50 PSIA
320	49.6	54.0
300	39.9	43.6
250	21.4	23.7
200	10.1	11.3
150	4.02	4.6
140	3.27	3.71
130	2.63	3.0
120	2.10	2.40

4. THE CRUDE FLASH CURVES HAVE BEEN REDRAWN ON THE ATTACHED DISTILLATION CHART. ONE TECHNIQUE THAT HAS BEEN USED TO CREATE FLASH CURVES FOR "TOPPED" CRUDES IS A PROCEDURE SIMILAR TO THAT USED FOR VACUUM TOWER DESIGN. THE PREFLASH TOWER OVERHEAD CUT POINT TEMPERATURE ON THE WHOLE CRUDE 0 PSIA FLASH CURVE IS PROJECTED TO THE 0% DISTILLED POINT AND A LINE DRAWN FROM THIS POINT TO THE 100% DISTILLED POINT ON THE WHOLE CRUDE 0 PSIA FLASH CURVE. THIS LINE REPRESENTS THE FLASH CURVE FOR THE "TOPPED" CRUDE AT SOME REDUCED PRESSURE. THIS PRESSURE IS ESTIMATED BY MULTIPLYING MOLE FRACTION OF "TOPPED" CRUDE RELATIVE TO THE WHOLE CRUDE BY THE ABSOLUTE PRESSURE (14.7 PSIA).

$$\text{i.e., } \frac{(4658.31)}{(8941.44)} \times 14.7 = 7.66 \text{ PSIA.}$$

CUSTOMER ENGINEERING DESIGN SEMINAR

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BY E N MILLER

4. CONTINUED: USING THE 0% POINT ($\approx 400^{\circ}\text{F}$) ON THE FLASHED CRUDE 7.66 PSIA FLASH CURVE AS REFERENCE FOR COX CHART DRAW IN THE 0 PSIG AND THE 15 PSIG FLASHED CRUDE FLASH CURVES. SINCE WE HAD A HIGH FLASH ZONE TEMPERATURE FOR THE WHOLE CRUDE, WE WILL MAKE NO PROVISION FOR OVER FLASH TO KEEP THE TEMPERATURE OF THE MATERIAL TO FLASH ZONE DOWN. WHEN WE CORRELATED EFV CURVES FOR CURVATURE, WE FOUND ABOUT A 15°F DIFFERENCE BETWEEN THE LINEAR EFV CURVE AND THE ADJUSTED CURVE. WE WILL TAKE THE SAME DELTA CREDIT IN THE REVAMP. DETERMINE CRUDE TOWER FLASH ZONE TEMPERATURE. ON A CORRELATED BASIS YOU SHOULD GET ABOUT 735°F . EXPLAIN HOW THIS VALUE WAS OBTAINED.

5. IN ORDER TO COMPLETE HEAT AND MATERIAL BALANCES FOR THE PREFLASH TOWER WE NEED TO KNOW THE TEMPERATURE AND ENTHALPY OF THE FEED. TO CALCULATE THIS WE NEED TO EVALUATE CRUDE EXCHANGE NETWORK AND CRUDE TOWER FOR THE REVAMP OPERATION. ANOTHER APPROACH WOULD BE TO SET A TARGET FEED TEMPERATURE TO PREFLASH TOWER, COMPLETE PREFLASH TOWER CALCULATIONS AND THEN DO CRUDE TOWER AND EXCHANGE NETWORK CALCULATIONS TO DETERMINE IF TARGET PREFLASH TOWER FEED TEMPERATURE CAN BE MET. THEN ITERATE PROBLEM AS REQUIRED. WE COULD ALSO START WITH THE OBJECTIVE OF NOT REVAMPING THE EXISTING CRUDE HEATER AND WORKING BACKWARDS FROM THAT POINT TO DETERMINE IF THAT PART OF REVAMP OBJECTIVE CAN BE MET. LETS START WITH THIS OPTION.

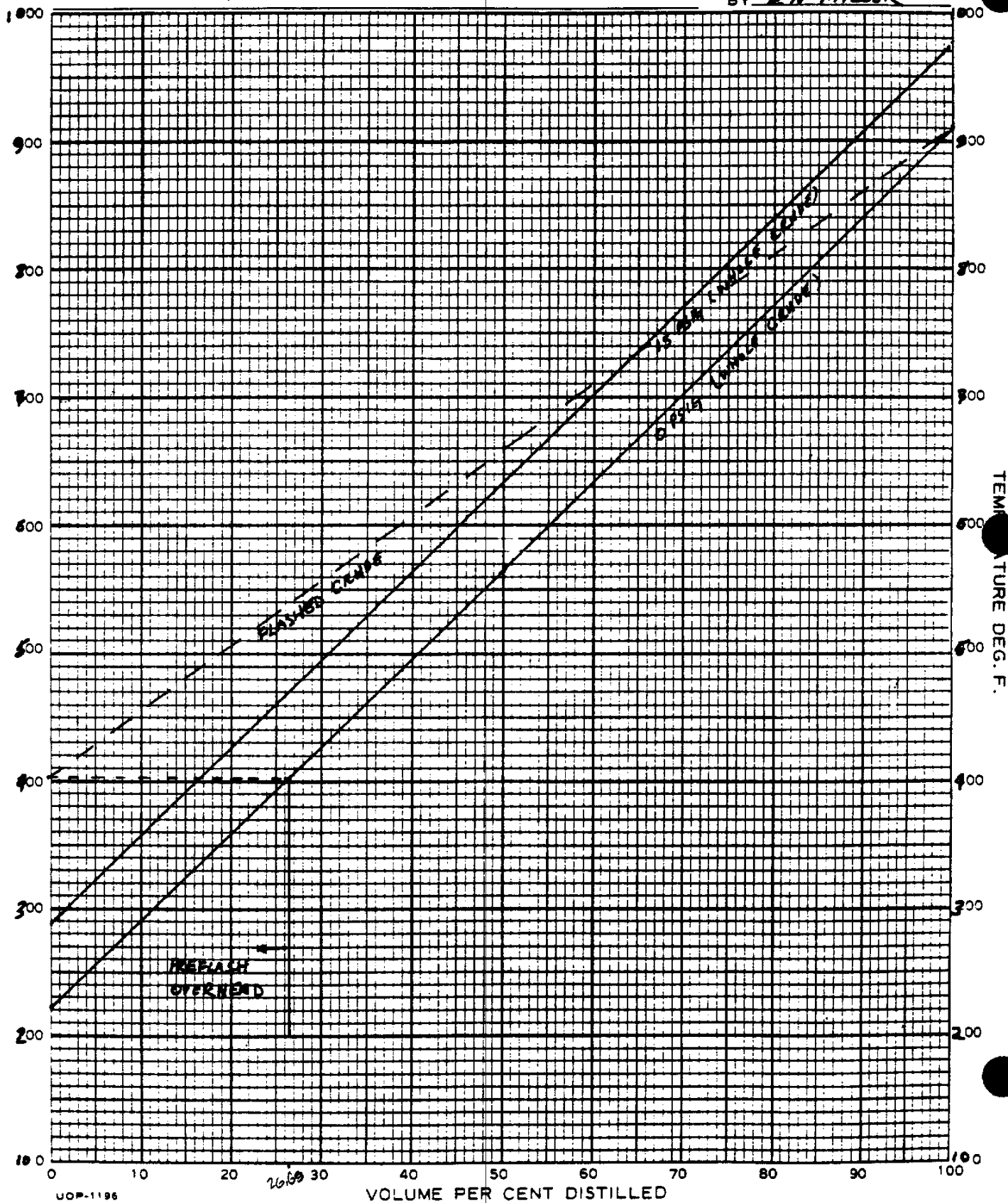
6. AS INDICATED IN ITEM 5, WE WOULD LIKE TO REVAMP THE UNIT WITHOUT ADDING ADDITIONAL CRUDE FEED HEATER. CAN THIS BE DONE? TESTS MADE WITH THE EXISTING CRUDE FEED HEATER INDICATED THAT IT COULD OPERATE A 110 % OF HEAT AND WEIGHT BALANCE DUTY OF 256.7 MMBA/HR WITHOUT MUCH DIFFICULTY. ON THIS BASIS CALCULATE REQUIRED PREFLASH TOWER BOTTOMS TEMPERATURE. USE THE NAPHTHA ENTHALPY CURVE OF NAPHTHA PORTION OF FEED.

UOP

DISTILLATION CHART

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT REVAMP
ARABIAN LIGHT FLASH CURVES

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CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT REVAMPPAGE NO. 7
DATE NOV 3, 1991
PROJECT
BY E N MILLER

7. FOR A REBOILED FLASH FRACTIONATOR, WE NORMALLY EXPECT ABOUT A 25°F RISE BETWEEN FEED AND BOTTOMS. AS DISCUSSED IN THE CRUDE DESIGN SECTION WE NEED A COLD REFLUX / DISTILLATE RATIO OF ABOUT 1.0 MINIMUM TO MEET FRACTIONATION REQUIREMENTS. SINCE THE OPERATING PRESSURE OF THE PREFLASH TOWER, IS THE SAME AS ORIGINAL CRUDE TOWER, WE EXPECT OVERHEAD VAPOR TEMPERATURE TO BE SIMILAR SO WE WILL USE 300°F FOR THIS DESIGN. WE CAN NOW DO A HEAT BALANCE FOR THE PREFLASH TOWER. NOTE WE CAN ONLY DETERMINE THE TOTAL HEAT INPUT REQUIREMENT FOR BOTH THE FEED AND THE REBOILER TOGETHER. HOW THIS HEAT REQUIREMENT IS SPLIT BETWEEN THE TWO SERVICES WILL BE CALCULATED LATER. ASSUME FOR NOW THAT ALL THE HEAT REMOVED FROM THIS TOWER IS WITH OVERHEAD CONDENSER. WE WILL ADDRESS PROBLEM OF DIVIDING THIS HEAT REMOVAL BETWEEN OVERHEAD CONDENSER AND TOP PUMPAROUND LATER. ON THE ABOVE BASIS DO THE HEAT BALANCE FOR THE PREFLASH TOWER.
8. IF THE FEED IS 25°F LOWER THAN BOTTOMS TEMPERATURE, BASED ON ITEM 7, CALCULATE REBOILER DUTY.
9. ORIGINAL REVAMP PROCESS SCHEME WAS BASED ON HEATING INCREMENTAL CRUDE CHALITY (50000 BPSD) TO DESALTER OPERATING TEMPERATURE OF 255-260°F WITH PREFLASH PUMP AROUND AT THE TOP. DETERMINE IF THIS IS POSSIBLE. ASSUME PUMP AROUND NAPHTHA COMES FROM THE TOWER AT 310°F AND RETURNS AT 210°F. FOR CONTROL EXTERNAL REFLUX SHOULD BE 10-20 %WW OF NET OVERHEAD PRODUCT.

10. AT THIS POINT, WE CAN CALCULATE THE HEAT AND MATERIAL BALANCE FOR THE CRUDE TOWER AT THE PROPOSED REVAMP OPERATING CONDITIONS. THE CALCULATIONAL PROCEDURE IS SIMILAR TO THAT USED FOR THE CRUDE UNIT DESIGN. WE WILL ASSUME THAT THE DRAW TRAY TEMPERATURES DO NOT CHANGE. DIRECTIONALLY THEY SHOULD INCREASE SLIGHTLY. FOR THE MOMENT WE WILL ASSUME THAT THE TOWER TOP TEMPERATURE IS 300 °F. IT WILL BE NECESSARY TO CONFIRM THIS ASSUMPTION LATER. ALSO REMEMBER THAT THE TOWER BOTTOMS TEMPERATURE IS ABOUT 40 °F LOWER THAN THE FLASH ZONE TEMPERATURE. OPERATING EXPERIENCE HAS INDICATED THAT LESS THAN DESIGN STEAM RATES ARE REQUIRED FOR THE SIDE CUT STRIPPERS. FOR THE REVAMP WE WILL USE THE SAME STEAM RATES FOR THE SIDECUT STRIPPERS AS WE DID FOR THE ORIGINAL DESIGN. THIS IS ABOUT 6.7 LB STEAM PER BARREL OF SIDECUT PRODUCT. FOR THE REVAMP WE WILL USE 8 LB STEAM / BARREL OF BOTTOMS FOR STEAM TO BOTTOM OF THE TOWER.

START BY DOING THE HEAT AND MATERIAL BALANCE CALCULATIONS FOR THE SIDE CUT STRIPPERS.

THEN DO THE EXCESS HEAT CALCULATIONS STARTING AT THE BOTTOM AND WORKING TO THE TOP OF THE TOWER. TARGET A MINIMUM OF 20% EXCESS HEAT AT THE DIESEL AND JET FUEL DRAWS. DETERMINE AVAILABLE PUMPAROUND HEAT REMOVAL DUTY ROUNDED DOWN TO THE NEAREST 5 MMBTU/HR. DO NOT CALCULATE PUMPAROUND RATES AT THIS TIME.

FOR THE MOMENT IGNORE THE NAPHTHA PUMPAROUND HEAT REMOVAL SECTION AND ASSUME THAT ALL EXCESS HEAT AT TOP IS REMOVED OVERHEAD.

THE ATTACHED HEAT AND WEIGHT SHEETS CAN BE COMPLETED AS DATA GENERATED TO MAKE IT EASIER TO KEEP TRACK OF INFORMATION.



25 East Algonquin Road
Des Plaines, Illinois 60017-5017

HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR.

CRUDE UNIT DESIGN - REVAMP

150 000 BPSD - ARABIAN LIGHT CRUDE

ENGINEERING DEPARTMENT

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DATE 10-20-91

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
DESALTED CRUDE TO FLASH ZONE												
TOTAL VAPOR.												
NAPHTHA												
JET FUEL												
DIESEL												
AGO												
OVER FLASH VAPOR.										0		
TOTAL LIQUID												
STEAM TO CRUDE COLUMN												
REDUCED CRUDE TO CRUDE EXCH.												
AGO PUMParound TO CRUDE EXCHANGE												
EXCHANGER DUTY												
AGO PUMParound RETURN TO COLUMN												
UNSTRIPPED AGO TO STRIPPER												
STEAM TO AGO STRIPPER												
VAPORS FROM AGO STRIPPER												
STEAM												
HYDROCARBON.												
AGO PRODUCT TO CRUDE EXCHANGE												

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HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR

CRUDE UNIT DESIGN - REVAMP

150,000 BPSD - ARABIAN LIGHT CRUDE

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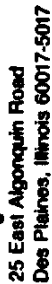
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CRUDE COLUMN SECTION

STREAM	° API (LB/GAL)	K	MM	FLOWING SP. GR. (LB/CF)	° F	PSIG (MM HG ABS)	BPSD @ 60°F (10 ³ SCF/D)	FLOWING GPM (CFS)	MOLES PER HR.	LB/HR	BTU/ LB	10 ⁶ BTU/HR
DIESEL PUMPROUND FROM COLUMN												
DIESEL PUMPROUND TO DEBIT EXCH												
EXCHANGER DUTY												
DIESEL PUMPROUND TO CAME EXCH												
EXCHANGER DUTY												
DIESEL PUMPROUND FR CAME EXCH												
DIESEL PUMPROUND TO DEBIT EXCH												
EXCHANGER DUTY												
DIESEL PUMPROUND FR DEBIT EXCH												
TO THE DIESEL PUMPROUND RETURN COLUMN												
UNSTRIPPED DIESEL TO STRIPPER												
STEAM TO DIESEL STRIPPER												
VAPORS FROM DIESEL STRIPPER												
STEAM												
HYDROCARBONS												
DIESEL PRODUCT TO CRUDE EXCHANGER												



ENGINEERING DESIGN SEMINAR

CRUDE UNIT DESIGN - REVAMP

150 000 BPSD - ARABIAN LIGHT CRUDE

CRUDE COLUMN SECTION

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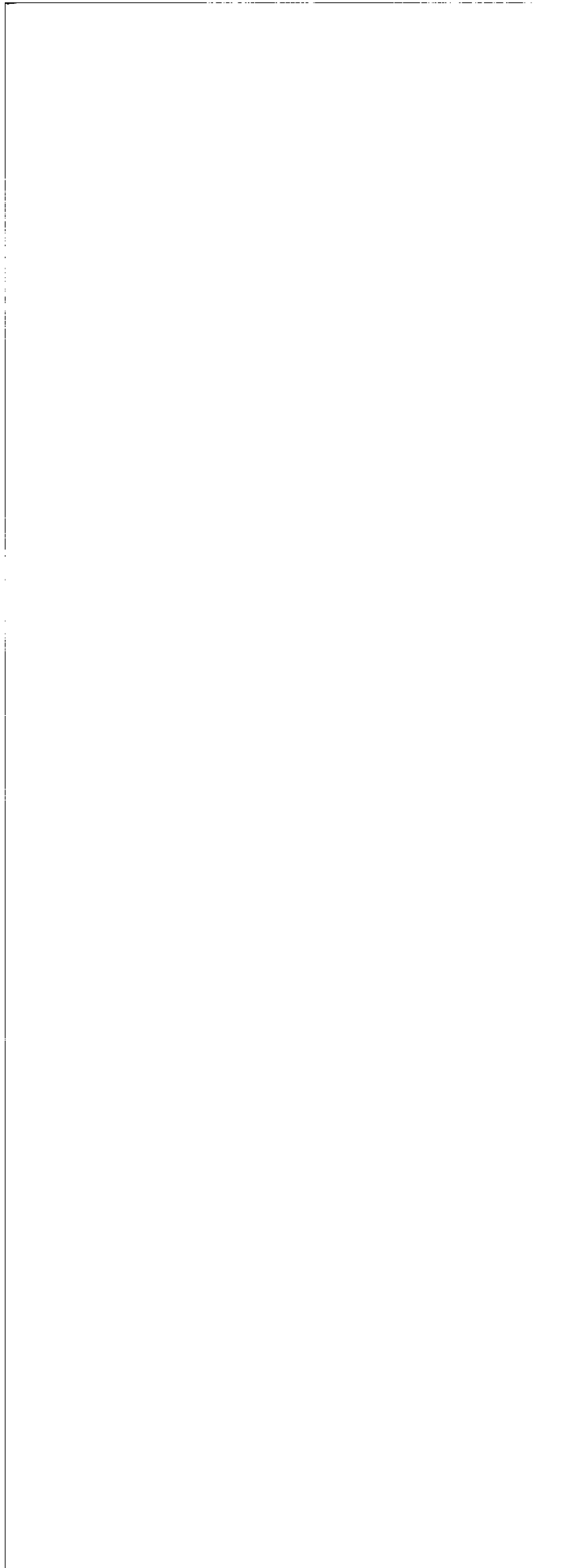
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BY **EN MILLER**

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11. WHEN MAKING EXCESS HEAT CALCULATIONS UP THE COLUMN YOU SHOULD HAVE FOUND THAT:

- (1) HEAT AVAILABLE FOR REMOVAL WITH A90 PUMPAROUND = 50 MMBTU/HR
- (2) HEAT AVAILABLE FOR REMOVAL WITH DIESEL PUMPAROUND = 65 MMBTU/HR
- (3) IGNORING NAPHTHA PUMPAROUND EXTERNAL REFLUX REQUIRE TO HEAT BALANCE COLUMN WAS 375130 LB/HR WHICH WAS EQUIVALENT TO 80.65 MMBTU/HR EXCESS HEAT AT TOP OF TOWER.

WE WILL NOW DETERMINE HOW MUCH OF THIS HEAT CAN BE REMOVED WITH NAPHTHA PUMP AROUND. FOR THE DEFINED NAPHTHA CUT EQUILIBRIUM KTI DATA ARE SUMMARIZED IN THE FOLLOWING TABLE.

TEMPERATURES.	14.7PSIA KTI.	50.8PSIA KTI
400 °F	27.04	30.37
350 °F	14.7	16.7
300 °F	7.1	8.2
250 °F	2.97	3.47
200 °F.	1.03	1.23
150 °F.	0.29	0.34
140 °F	0.21	0.26

IF NO NAPHTHA PUMPAROUND HEAT REMOVAL:
OVERHEAD VAPOR.

$$\text{STEAM} = 1913.5 \text{ MOL/HR.}$$

$$\text{NAPHTHA PRODUCT} = 234.97 \text{ MOL/HR.}$$

$$\text{REFLUX} = 2587.10 \text{ MOL/HR.}$$

WHEN OVERHEAD VAPOR AT 300 °F, IF CONSIDERED AS SINGLE COMPONENT
PP HC $\approx$ 7.1 PSIA

$$y_{HC} = \frac{7.1}{24.7} = 0.287 \quad y_{\text{STEAM}} = 1 - y_{HC} = 0.713$$

$$\text{TOTAL MOLES OVERHEAD VAPOR} = 1913.5 / 0.713 = 2683.7 \text{ MOL/HR.}$$

$$\text{MOLES OF HYDROCARBON} = 2683.7 - 1913.5 = 770.2 \text{ MOL/HR}$$

$$\text{MOLES OF REFLUX} = 770.2 - 234.97 = 535.2 \text{ MOL/HR}$$

$$\text{LB/HR EXTERNAL REFLUX} = 77605 \text{ LB/HR.}$$

11. CONTINUED.

$$\begin{aligned}\text{FRACTION OF TOTAL REFLUX AS EXTERNAL} &= \frac{77605}{375130} \times 100 \\ &= 20\%.\end{aligned}$$

THIS MEETS OUR 10-20% MINIMUM GUIDELINE.

$$\begin{aligned}\text{HEAT REMOVED BY EXTERNAL REFLUX} &= (77605)(258-43) \\ &= 16.685 \times 10^6 \text{ BTU/HR.}\end{aligned}$$

$$\begin{aligned}\text{HEAT AVAILABLE FOR NAPHTHA PUMPAROUND} \\ &= 80.653 - 16.685 = 63.96 \times 10^6 \text{ BTU/HR} \\ &\approx \underline{64 \times 10^6 \text{ BTU/HR}}\end{aligned}$$

THIS QUANTITY OF NAPHTHA PUMPAROUND HEAT CAN BE ADJUSTED ~~UP~~ UP BY REDUCING PUMPAROUND HEAT LOWER IN COLUMN. WE MAY WANT TO DO THIS LATER DEPENDING ON HOW WELL EXISTING HEAT EXCHANGE FITS THE REVAMP OPERATION. IT CAN ALSO BE ADJUSTED DOWN BY INCREASING EXTERNAL REFLUX RATE.

12. WE ARE NOW AT A POINT WHERE WE CAN BEGIN TO EVALUATE THE SUITABILITY OF THE EXISTING HEAT EXCHANGE NETWORK FOR THE REVAMP OPERATION. SINCE WE ADDED A NEW DESALTER FOR THE INCREMENTAL CRUDE CAPACITY THE CRUDE EXCHANGE TRAINS TO THE DESALTER ARE SPLIT. YOU SHOULD EVALUATE FIRST THE EXISTING CRUDE TRAIN EXCHANGE TO THE DESALTER. THERE ARE A NUMBER OF WAYS TO RATE EXISTING EXCHANGER FOR ALTERNATE OPERATIONS. A COMMON TECHNIQUE USED BY PROCESS ENGINEERS WHEN LIMITED DATA ARE AVAILABLE AND EXCHANGE CONDITIONS ARE SIMILAR IS TO CHECK "UA REQUIRED" FOR THE NEW SERVICE AND COMPARE IT WITH "UA AVAILABLE" FROM THE EXISTING OPERATION. "U" IS OVER-ALL HEAT TRANSFER COEFFICIENT AND "A" IS HEAT TRANSFER AREA. THE EQUATION

$$Q = UA \Delta T_{LM} \cdot F \text{ IS USED.}$$

F IS TEMA CORRECTION FACTOR WHICH CAN BE FOUND IN HEAT EXCHANGE SECTION OF DATA BOOK.

CONTINUED DESIGNS
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HEAT EXCHANGERS

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BASIC EQUATION: $Q = U A \Delta T$

WHERE Q IS BTU/HR TRANSFERRED

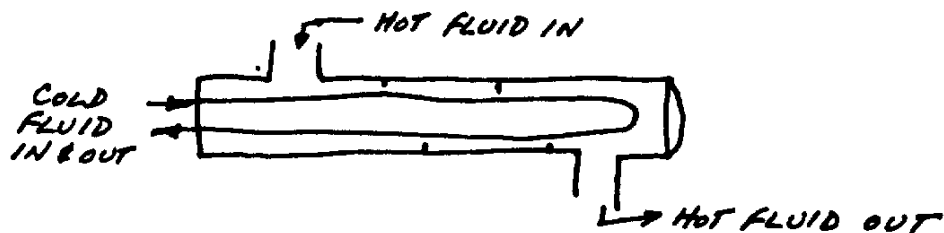
U IS THE HEAT TRANSFER COEFFICIENT, BTU/HR/FT²/°F

ΔT = CORRECTED LOG MEAN TEMP. DIFFERENCE, °F.

THE UNCORRECTED LOG MEAN TEMP DIFF. =

$$LMTD = \frac{\text{GREATER TEMP. DIFF.} - \text{LESSER TEMP. DIFF.}}{\log_e \frac{\text{GTD}}{\text{LTD}}}$$

THIS APPLIES ONLY TO TRUE COUNTER-CURRENT EXCHANGERS.
FOR MECHANICAL REASONS, MOST EXCHANGERS CONSIST OF A BUNDLE OF U-TUBES IN A SHELL:



IF THE HOT FLUID IS TO BE COOLED BELOW THE TEMP.
AT WHICH THE COLD FLUID LEAVES, THERE IS SAID TO BE A
TEMPERATURE CROSS AND THE CORRECTED LMTD ^{MAY BE} _A SUFFICIENTLY REDUCED
SO THAT THE EXCHANGER BECOMES UNREASONABLY EXPENSIVE. THE
USUAL SOLUTION IS TO INSTALL EXCHANGERS IN MULTIPLE SHELLS
IN SERIES, SO THAT THERE IS NO CROSS IN ANY ONE SHELL

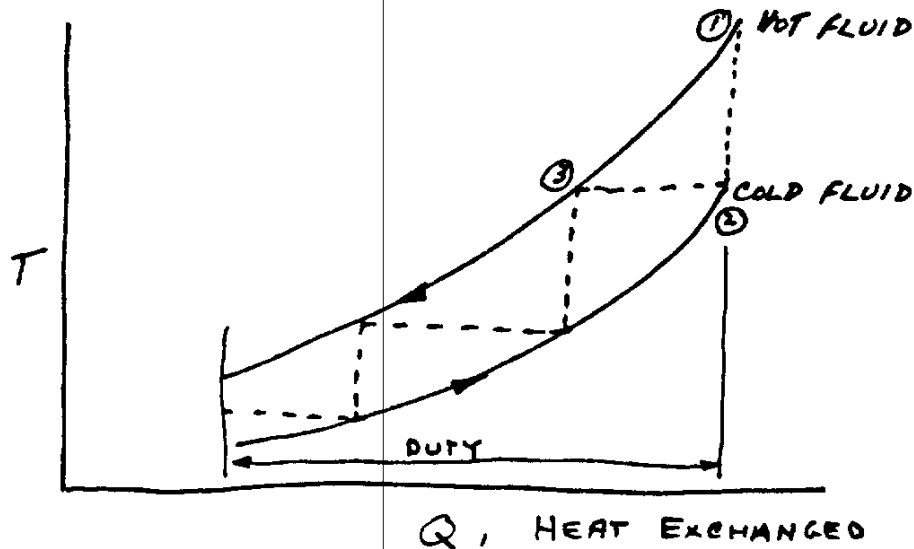
CONSIDER THE FOLLOWING DIAGRAM:

HEAT EXCHANGERS, CONT'D

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HERE WE HAVE PLOTTED THE TEMPS. OF THE HOT AND COLD FLUID AGAINST THE QUANTITY OF HEAT EXCHANGED. IF WE DROP A VERTICAL LINE FROM THE END OF THE HOT FLUID CURVE (POINT 1) TO THE COLD FLUID CURVE (POINT 2) AND THEN RUN A HORIZONTAL LINE FROM POINT 2 BACK TO THE HOT FLUID CURVE (POINT 3) THIS REPRESENTS THE DUTY IN A "NO-CROSS" EXCHANGER SHELL. IF WE CONTINUE THIS STEPPING OFF PROCESS WE SEE THAT THE TOTAL DUTY FOR THIS EXAMPLE WOULD REQUIRE 3 SHELLS IN SERIES.

IF WE DENOTE THE HOT FLUID IN AND OUT OF THE EXCHANGER BY THE SYMBOLS T_1 AND T_2 , AND CORRESPONDINGLY THE COLD FLUID IN AND OUT BY t_1 AND t_2 , A CORRECTION FACTOR "F" CAN BE DEFINED AS A FUNCTION OF THE HEAT CAPACITY RATIO "R" AND THE THERMAL EFFECTIVENESS "P". THESE LATTER ARE DEFINED AS

$$R = \frac{T_1 - T_2}{t_2 - t_1} \quad \left\{ \begin{array}{l} \text{SET 1} \\ \text{OF } R, P \end{array} \right. \quad \text{LET } R' = 1/R \quad \left\{ \begin{array}{l} \text{SET 2} \\ \text{OF } R, P \end{array} \right.$$

$$\text{AND } P = \frac{t_2 - t_1}{T_1 - t_1} \quad \left\{ \begin{array}{l} \text{SET 1} \\ \text{OF } R, P \end{array} \right. \quad P' = PR \quad \left\{ \begin{array}{l} \text{SET 2} \\ \text{OF } R, P \end{array} \right.$$

USE SET 2 IF SET 1 IS IN A DIFFICULT REGION.

THE MATHEMATICAL RELATIONSHIP BETWEEN F , P AND R IS SO COMPLICATED THAT GRAPHS ARE ALWAYS USED TO OBTAIN " F ". THESE GRAPHS WILL BE FOUND IN THE TEMA MANUAL AND IN MANY HANDBOOKS.

(FOR THE INTELLECTUALLY CURIOUS, THE EQUATIONS ARE SHOWN ON PAGE 4.)

EXAMPLE: SUPPOSE WE HAVE A HOT FLUID BEING COOLED FROM 500°F TO 325°F BY A COLDER FLUID WHICH IS HEATED FROM 200°F TO 450°F :

$$\begin{array}{r} 500 \rightarrow 325 \\ 450 \leftarrow 200 \\ \hline 50 \quad 125 \end{array}$$

$$\text{UNCORRECTED LMTD} = \frac{125-50}{\log_e\left(\frac{125}{50}\right)} = 81.9^{\circ}\text{F}$$

$$500 - 325 = 175 \quad R = \frac{175}{250} = 0.70$$

$$450 - 200 = 250$$

$$500 - 200 = 300 \quad P = \frac{250}{300} = 0.833$$

FROM CURVE FOR A SINGLE SHELL $F = 0.3$

" " " TWO " $F = 0.6$

" " " THREE " $F = 0.875$

A SINGLE SHELL WOULD BE COMPLETELY UNECONOMICAL. AN " F " FACTOR OF 0.8 IS USUALLY CONSIDERED TO BE THE LOWEST ACCEPTABLE. IN THIS CASE, WE WOULD USE 3 SHELLS, AND A CORRECTED LMTD OF

$$0.875 \times 81.9 = 71.7^{\circ}\text{F}.$$

NOTE: A "NO-CROSS" SHELL AS USED HERE MEANS ONE WHERE THE OUTLET TEMPERATURES (COLD SIDE & HOT SIDE) FROM ONE SHELL ARE THE SAME. THIS USUALLY GIVES A Q80 TEMA F FACTOR. THE TEMA CURVES FOR $F(R, P)$ ARE FOUND IN THE HEAT EXCHANGER HANDBOOKS.

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HEAT EXCHANGERS, CONT'D
MATHEMATICAL RELATIONSHIPS BETWEEN
F, R AND P

IF N IS THE NUMBER OF SHELL PASSES:

FOR THE CASE WHERE $R \neq 1$

$$\text{LET } X = \frac{1 - \left(\frac{1-RP}{1-P}\right)^{\frac{1}{N}}}{R - \left(\frac{1-RP}{1-P}\right)^{\frac{1}{N}}}$$

$$\text{THEN } F = \frac{\frac{\sqrt{R^2+1}}{R-1} \text{LN}_2 \left(\frac{1-X}{1-RX} \right)}{\text{LN}_2 \left(\frac{\frac{2}{X} - 1 - R + \sqrt{R^2+1}}{\frac{2}{X} - 1 - R - \sqrt{R^2+1}} \right)}$$

IF $R=1$

$$\text{LET } X = \frac{P}{N-NP+P}$$

$$\text{AND } F = \frac{\sqrt{2} \left(\frac{X}{1-X} \right)}{\text{LN}_2 \left(\frac{\frac{2}{X} - 2 + \sqrt{2}}{\frac{2}{X} - 2 - \sqrt{2}} \right)}$$

NOTE: LN_2 IS $\log_e$ COMMONLY CALLED "LN" TO DIFFERENTIATE IT WITH $\log_{10}$, COMMONLY CALLED "LOG".

12. Continued.

EFFECT OF DIFFERENCES IN FLOW BETWEEN REVAMP CONDITIONS AND ORIGINAL DESIGN IN EXCHANGERS ON HEAT TRANSFER COEFFICIENT CAN BE ESTIMATED AS FOLLOWS:

$$h_{\text{shell}} \propto G_{\text{shell}}^{0.6} \quad (\text{FOR SHELL SIDE})$$

$$h_{\text{tube}} \propto G_{\text{tube}}^{0.8} \quad (\text{FOR TUBE SIDE})$$

$$\text{Then } u'/u = \sqrt{(G'/G)_t^{0.8} \times (G'/G)_s^{0.6}}$$

SINCE WE HAVE NOT REALLY DECIDED WHICH FLUID SHOULD BE ON WHAT SIDE WE WILL USE THE FOLLOWING APPROXIMATION

$$u'/u \approx \sqrt{(G'/G)_1^{0.7} \times (G'/G)_2^{0.7}}$$

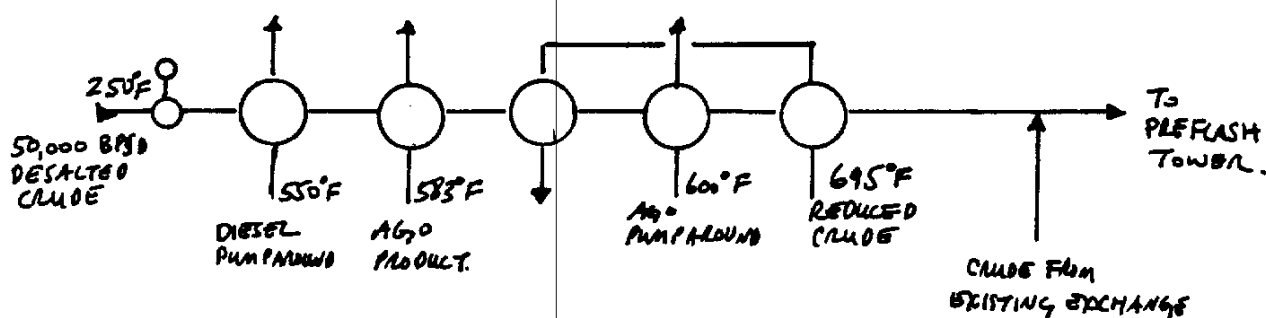
$$\text{OR } u'/u \approx [(G'/G)_1 \times (G'/G)_2]^{0.35}$$

IN THE REVAMP DESIGN THE CRUDE FLOW RATE REMAINS CONSTANT IN THE PART OF THE EXCHANGE TRAIN AND THE PRODUCT FLOW RATES HAVE ALL INCREASED SO THE EXISTING EXCHANGERS WILL BE ADEQUATE FOR THE REVAMP OPERATION. EVALUATE EXCHANGERS FOR MAXIMUM HEAT PICKUP IN THE REVAMP OPERATION WITH THE OBJECTIVE OF ELIMINATING THE NEED FOR THE DIESEL PUMP AROUND EXCHANGE IN THIS EXCHANGE TRAIN. WE EXPECT TO HEAT CRUDE TO 255-260°F WITHOUT THE DIESEL PUMP AROUND EXCHANGE IN SERVICE.

13. TO ANALYZE THE EXCHANGE SCHEME FOR THE DESIGNED CRUDE THERE ARE A NUMBER OF OPTIONS AVAILABLE. FOR THE PRESENT EVALUATION WE WILL ONLY LOOK AT ONE OF THE POSSIBILITIES. THE SCHEME WHICH IS SUGGESTED IS AIMED AT MINIMIZING CHANGES TO THE EXISTING FLOW SCHEME. IN GENERAL, IT IS DIFFICULT TO ADD EXCHANGE IN SERIES WITH THE EXISTING EXCHANGES

13. CONTINUED

BECAUSE OF PRESSURE DROP CONSTRAINTS. FOR THIS REASON WE WILL ADD ADDITIONAL EXCHANGES IN PARALLEL ADDED 50,000 BPSD OF CRUDE FROM THE NEW DESALTER. A SKETCH OF THE ADDED EXCHANGE TRAIN IS SHOWN BELOW



CALCULATE TEMPERATURES AND EXCHANGE SURFACE FOR EACH EXCHANGER IN THIS NEW EXCHANGE TRAIN WITH THE FOLLOWING ASSUMPTIONS.

1. ASSUME HEAT PICKED UP IN THE EXISTING EXCHANGER REMAINS UNCHANGED.
2. FOR PRODUCT EXCHANGES ASSUME $\frac{1}{3}$ OF THE REVAMP PRODUCT FLOW GOES THROUGH THE NEW EXCHANGER.
3. FOR THE PUMP AROUND STREAMS ASSUME A NEW PARALLEL BRANCH IS ADDED. DUTY FOR EACH SERVICE WILL BE AS FOLLOWS:

DIESEL PUMP AROUND	$Q = 18.5 \times 10^6 \text{ BTU/HR.}$
AGO PUMP AROUND	$Q = 20 \times 10^6 \text{ BTU/HR.}$

4. IN GENERAL, LIMIT COLD END APPROACH TEMPERATURE TO 50°F AT. AND/OR ONE SHELL OF EXCHANGE FOR ALL SERVICES EXCEPT REDUCED CRUDE EXCHANGES.
5. MAXIMUM ΔT FOR PUMP AROUND TOTAL SHOULD BE LIMITED TO 200°F MAXIMUM.

14. ONCE YOU HAVE COMPLETED CALCULATION OF THE NEW PARALLEL EXCHANGE TRAIN IN ITEM 13, CALCULATE THE TOTAL DESALTED CRUDE TEMPERATURE (AS A LIQUID) AND ENTHALPY. TEMPERATURE FROM EXISTING DESALTED CRUDE TRAIN IS 500°F AND YOU SHOULD HAVE CALCULATED A TEMPERATURE OF 520°F FROM THE NEW EXCHANGE TRAIN.

15. FROM ITEM 8 OF THE "ANSWERS", WE ESTIMATED A FEED TEMPERATURE OF 485°F. THIS WAS BASED ON USING ALL OF EXISTING CHARGE HEATER CAPACITY (110% OF ORIGINAL PROCESS DUTY). WE NOW FIND THAT WE CAN WITH THE ADDITION OF A PARALLEL EXCHANGE TRAIN HEAT THE FEED TO THE PREFLASH TOWER TO ABOUT 507°F WHICH IS MUCH HIGHER THAN OUR ORIGINAL TARGET. WE NOW NEED TO REVISIT THE PREFLASH TOWER HEAT BALANCE USING THE NEW FEED TEMPERATURE. A NUMBER OF THINGS HAPPEN AS WE WILL EXPLAIN IN THE FOLLOWING DISCUSSION.

a. WE STILL EXPECT ABOUT A 25°F RISE BETWEEN THE FEED TEMPERATURE AND THE COLUMN BOTTOMS WHEN THE PREFLASH TOWER IS REBOILED. ON THIS BASIS THE COLUMN BOTTOMS TEMPERATURE WILL BE ABOUT (507+25) 532°F.

b. COLUMN BOTTOMS, BY DEFINITION, IS A BUBBLE POINT LIQUID. BASED ON SPLIT BETWEEN OVERHEAD AND BOTTOMS IN THE ORIGINAL CALCULATIONS THE BOTTOMS TEMPERATURE WAS 510°F AT 15 PSIG. SINCE THE SPLIT IS THE SAME THE OPERATING PRESSURE AT BOTTOM OF THE TOWER (PREFLASH) MUST BE HIGHER TO GET A BOTTOMS TEMPERATURE OF 532°F. USING A COX CHART, WE ESTIMATE THIS PRESSURE TO BE 38 PSIG OR 23 PSIG.

c. ASSUMING A 5 PSI DIFFERENTIAL PRESSURE ACROSS THE TOWER, PRESSURE AT TOP BECOMES 18 PSIG. WITH A HIGHER PRESSURE THE TEMPERATURE AT THE TOP OF THE TOWER WILL BE HIGHER. AGAIN FROM THE COX CHART THE NEW OVERHEAD TEMPERATURE IS ABOUT 325°F.

CUSTOMER ENGINEERING DESIGN SEMINAR.
 SUBJECT CUMS UNIT REVAMP

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15. CONTINUED.

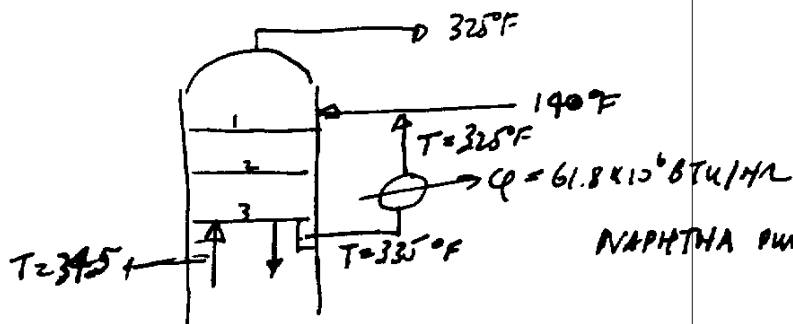
d. IF WE ASSUME ΔP BETWEEN TOP AND RECEIVER IS 6 PSI, THE RECEIVER WILL NOW OPERATE AT 12 PSIG INSTEAD OF THE 5 PSIG PREVIOUSLY ASSUMED. SINCE THE RECEIVER OPERATING CONDITIONS PRODUCED A SUBCOOLED LIQUID RUNNING THE SYSTEM AT A HIGHER PRESSURE AND THE SAME TEMPERATURE IS OK.

WE CAN NOW REWRITE HEAT AND MATERIAL BALANCES FOR THE NEW CONDITIONS ESTABLISHED ABOVE.

	IN LB/HR.	OUT LB/HR.	SPECIFIC ENTHALPY	IN MMBTU/HR.	OUT MMBTU/HR.
FEED	1853385			474.18	
REBOILER DUTY				Q_R	
NET OVERHEAD (V)		418000	275		114.95
REFLUX (L)	41800		44	18.39	
REFLUX (V)		418000	275	18.39	114.95
FLASHED CUMS (L)		1435,385	~ 285		<u>380.38</u>
				$Q_{LT} 492.57$	<u>620.28</u>

$$Q_{REBOILER} = 620.28 - 492.57 = 127.71 \text{ MMBTU/HR.}$$

WE ALSO NEED TO DO NAPHTHA PUMPAROUND CALCULATIONS



$$\text{NAPHTHA PUMPAROUND RATE} = \frac{61.8 \times 10^6}{(158 - 93)} = 950770 \text{ LB/HR.}$$

$$\text{EXTERNAL REFLUX RATE} = \frac{34.76 \times 10^6}{(275 - 44)} = 150,476 \text{ LB/HR.}$$

16. WE WILL NOW CALCULATE THE REBOILER CIRCULATION RATE.

FIRST DRAW THE 23514 EFV FLASH CURVE PARALLEL TO THE EXISTING 151514 CURVE ON PAGE 3 OF REVAMP NOTES. 0% POINT SHOULD BE 532°F. WE WILL SET THE REBOILER OUTLET TEMPERATURE AT 700°F. FROM THE EFV CURVE WE GET A PERCENT VAPORIZATION OF THE "TOLIED" CRUDE OF 33.5% NOW CALCULATE VAPOR COMPOSITION BASED ON 100 BBL'S OF FLASHED CRUDE.

	LV% TOTAL CRUDE	LV% FLASHED CRUDE	ACCUM LV% IN FLASHED CRUDE
HEAVY NAPHTHA.	2	2.723	2.723
JET FUEL	16.55	22.563	25.286
DIESEL	9.15	12.474	37.760
A90	7.55	10.293	

33.5% VAPORIZATION

ONLY A PORTION OF THE DIESEL IS VAPORIZED IN REBOILER SO PRORATE.

$$\text{DIESEL VAPOR} = 33.5 - 25.286 = 8.214 \text{ LV\%}$$

$$\text{DIESEL LIQUID} = 12.474 - 8.214 = 4.26 \text{ LV\%}$$

WE NOW NEED TO ESTIMATE PROPERTIES OF THE TWO PORTIONS OF THE DIESEL CUT. THIS CAN BE DONE FROM CRUDE ASSAY.

	API	UOP K	MW	LV% ON CRUDE	WT% ON CRUDE
DIESEL VAPOR	36.2	11.94	238	6.03	6.0
DIESEL LIQ	32.7	11.91	287	3.12	3.18
TOTAL DIESEL	35.0	11.93	253	9.15	9.18

$$\text{EST LAB OF DIESEL VAPOR} = 565^\circ\text{F. EST LAB P} = 582^\circ\text{F}$$

CALCULATE SPECIFIC ENTHALPY OF VAPOR BASED ON 100 BPD OF FLASHED CRUDE

	API	UOP K	MW	LB/HR.	MOLE/HR	BPD
HYV NAPH	50.3	11.97	145	30.92		2.723
JET FUEL	42.8	11.92	187	267.26		22.563
DIESEL VAPOR	36.2	11.94	238	101.12		8.214
TOTAL				399.30		

16. CONTINUED.

AT 700°F ENTHALPY OF VAPOR.

	LB/HR	BTU/LB	BTU/HR
HVY NAPHTHA	30.92	505	15614.6
JET FUEL	267.26	485	129621.1
DIESEL VAPOR	101.12	480	48537.6
	399.30		193773.7

$$\text{BTU/LB} = 485.28.$$

CALCULATE SPECIFIC ENTHALPY OF LIQUID BASED ON 100 BPD OF FLASHED CRUDE.

	API	UOP K	MW	LB/HR	MOLE/HR	BPD.
DIESEL LIQ.	32.7	11.91	287	39.23		3.12
AGO	29.5	11.87	312	96.82		7.55
RESID.	17.5	11.72	462.4	719.66		51.94
				855.71		

AT 700°F ENTHALPY OF LIQUID

	LB/HR	BTU/LB	BTU/HR
DIESEL LIQ.	39.23	400	15692
AGO	96.82	390	37453
RESID	719.66	380	273471
	855.71		327116

$$\text{BTU/LB} = 382.3$$

$$\text{SPECIFIC ENTHALPY OF MATERIAL FROM REBOILER} = \frac{193773.7 + 327116}{899.30 + 855.71}$$

$$= 415 \text{ BTU/LB}$$

SPECIFIC ENTHALPY OF LIQUID TO REBOILER = 265 BTU/LB.

$$\text{REBOILER CIRCULATION RATE} = \frac{117.71 \times 10^6}{(415 - 265)} = 784,733 \text{ LB/HR}$$



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HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SUMMARY

CANOE UNIT DESIGN - ROTAMP

150,000 GPD - ARABIAN LIGHT CANOE

EXISTING CANOE PROHEAT TRAIN.

ENGINEERING DEPARTMENT

PAGE NO. 1

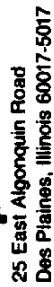
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PROJECT

BY E.A. 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
CANOE TO JET FUEL PRODUCT EXCH. EXCHANGER DUTY.												
CANOE TO NAPHTHA PA EXCHANGE EXCHANGER DUTY.												
CANOE TO DIESEL PRODUCT EXCHANGE EXCHANGER DUTY.												
CANOE TO DIESEL PA EXCHANGE EXCHANGER DUTY.												
CANOE TO DESALTER.												
DESALTER FEED WATER TO EXCHANGE EXCHANGER DUTY.												
DESALTER FEED WATER TO DESALTER.												
DESALTER EFFLUENT WATER TO EXCHANGE EXCHANGER DUTY.												
DESALTER EFFLUENT WATER TO SW STRIPPER												
DESALTER CANOE FROM DESALTER.												



ENGINEERING DESIGN SEMINAR.

CRUISE UNIT DESIGN - COVAMP

150,000 OPSO MODERN LIGHT CRUISE.

NEW CRUISE PHOTOGRAPHY TEAM.

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NEW CROPS PROTHEAT TEAM:

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ENGWISUNG POLITIK SEQUENZ-

ENGWISSUNG POLITIK SEMINAR

CALCUL UNIT DESIGN - REVAMP

50,000 BSO ARABIAN 4547 CMMR

NEW COME PACHOTAT TRAW.

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HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR

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CRUDE UNIT DESIGN - REFINERY

DATE NOV 1981

150,000 BPSD ARABIAN LIGHT CRUDE.

PROJECT

ENGINEERING DEPARTMENT

PRE FLASH COLUMN

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 PREFLASH COLUMN												
UNSTAB. NAPHTHA												
MVY NAPHTHA.												
FLASHED CRUDE.												
BOTTOMS FROM COLUMN												
BOTTOMS TO CRUDE COLUMN SECTION												
BOTTOMS TO REBOILER.												
REBOILER HEATER DUTY												
TOTAL FROM HEATER TO COLUMN												
VAPOR												
LIQUID												
NAPHTHA MINIMOUND TO EXCHANGER.												
EXCHANGER DUTY												
NAPHTHA MINIMOUND TO COLUMN												
VAPOR FROM COLUMN TOP.												
CONDENSED DUTY.												
MATERIAL TO REFLUXER.												
REFLUX												
NET UNSTAB. NAPHTHA TO FRACTIONATOR												

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CRUDE UNIT REVAMP.PAGE NO. 1DATE NOV 3, 1991

PROJECT

BY REN MILLER"ANSWERS"

1.

	API	UOP K	MW	BPSD	LB/HR	MOLE/HR
PF NAPHTHA.	61.3	12.15	108.65	34470	369020	3396.47
HVY NAPHTHA	50.3	11.97	145	3000	34070	234.97
NAPHTHA	60.4	12.13	111	37470	403090	3631.44

2.

API UOP K MW BPSD LB/HR MOLE/HR LV%

CRUDE

150,000

ESTIMATED YIELDS

C3	14.7	44.094	255	1890	42.84	0.17	
IC4	13.81	58.12	255	2095	36.05	0.17	
NLC4	13.50	58.12	1590	13560	233.23	1.06	
IC5	13.05	72.146	1095	9985	138.40	0.73	
NLC5	13.03	72.146	2205	26300	281.37	1.47	
CP	11.11	70.13	105	1150	16.40	0.07	
PF NAPHTHA	61.3	12.15	108.65	34470	369020	3396.47	22.98

FLASHED CRUDE

110 025 1435385 4658.31 73.35

HVY NAFT	50.3	11.97	145	3000	34070	234.97	2.0
JET FUEL	42.8	11.97	187	24825	294055	1572.49	16.55
DIESEL	35.0	11.93	253	13725	170190	672.69	9.15
A60	29.5	11.87	312	11325	145225	465.47	7.55
RESID	17.5	11.72	462.4	57150	791845	1712.47	38.10
Σ	35.6	11.92	210.3	150 000	1853385	8941.44	100.00

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CLUDE UNIT REVAMP.
ANSWERSPAGE NO. 2DATE NOV 3 1991

PROJECT

BY EN MILLER

3. CHECK PREFLASH TOWER OVERHEAD RECEIVER TO BE SURE PRODUCT IS ALL LIQUID AT PROPOSED OPERATING CONDITIONS. CALCULATE BUBBLE POINT OPERATING PRESSURE IS 5 PSIG OR 19.7 PSIA. IF WE USE K_T DATA AT 14.7 PSIA, IT SHOULD BE CLOSE ENOUGH.

COMPONENT.	f_i mol/hr	AT 14.7 PSIA K _T	$f_i \cdot K_i \cdot T$
C ₃	42.84	215.15	13361.786
iC ₄	36.05	102.92	3710.3
NC ₄	233.23	79.935	18643.2
iC ₅	138.40	36.899	5106.8
NC ₅	281.37	29.747	8369.9
CP	16.40	20.335	333.5
PF NAHTHA.	3396.47	3.27	1110.1
TOTAL	4144.76		50635.6

OK.

$$f_i K_i = 3444.6 < 4144.76 \quad \text{SINGLE PHASE.}$$

4. REFER TO ATTACHED CHART. CONTAINING FLASH CURVES.

$$\begin{aligned} \text{FOR TOPPED CRUDE WITH NO OVERFLASH LIFT} &= \frac{73.35 - 38.1}{73.35} \times 100 \\ &= 48\% \end{aligned}$$

TEMPERATURE FROM FLASHED CRUDE EFV CURVE AT 15 PSIG FOR 48% VAPOR IS 750°F.

ADJUST THIS VALUE BY 15°F TO ACCOUNT FOR FLASH CURVE CURVATURE.

$$\begin{aligned} \text{TEMPERATURE OF FEED TO CRUDE FLASH ZONE} &= 750 - 15 \\ &= 735^\circ\text{F} \end{aligned}$$

UOP

DISTILLATION CHART

CUSTOMER ENGINEERING DESIGN SEMMAR

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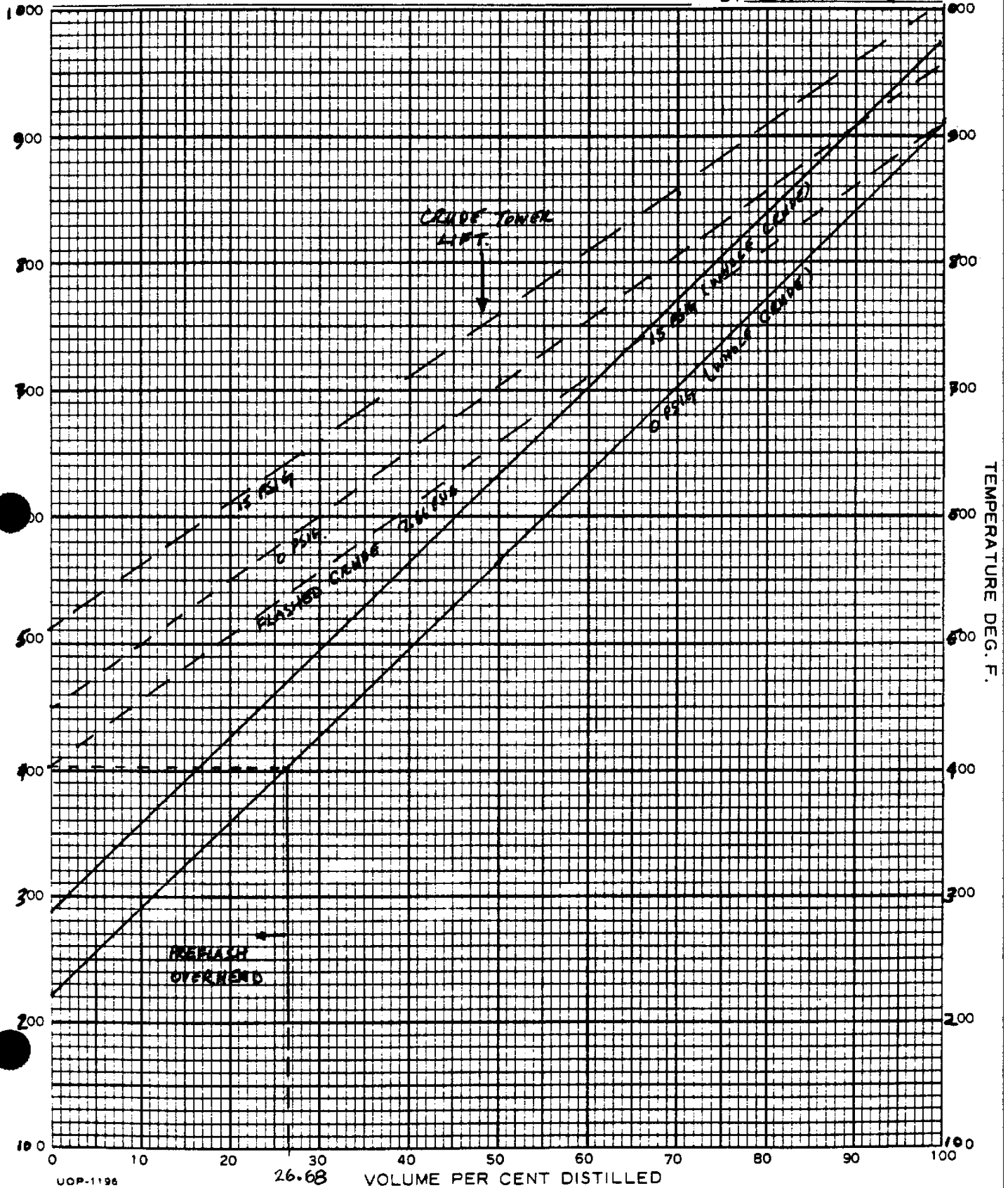
SUBJECT CRUDE UNIT REVAMP

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ARABIAN LIGHT FLASH CURVES

PROJECT

BY EN MILLER



6. CALCULATE ENTHALPY OF TOPPED CRUDE FEED TO CRUDE TOWER WITH NO OVERFLASH. AT 735°F

COMPONENT	LB/HR.	@ 735°F ENTHALPY. BTU/LB.	ENTHALPY. MM BTU/HR.
HVY NAPH (V)	30070	555	18.91
2ET FUEL (V)	294055	515	151.44
DIESEL (V)	170190	507	86.29
A40 (V)	145225	497	72.18
RESID (L)	791845	400	316.74
TOTAL	1435385	TOTAL	645.55

$$\text{HEATER DUTY} = 1.1 \times 250.7 = 275.77 \text{ MM BTU/HR.}$$

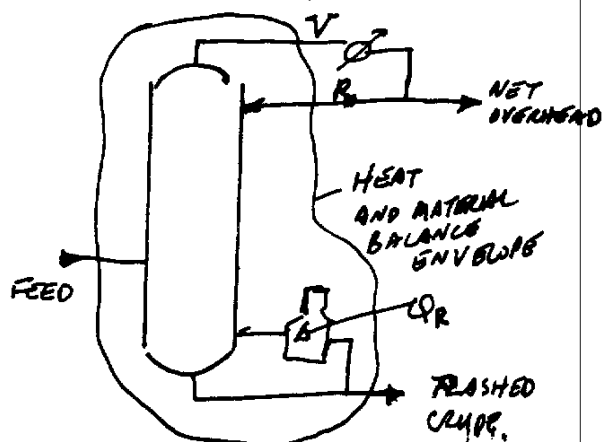
$$\begin{aligned} \text{ENTHALPY OF FEED TO HEATER AS A LIQUID} &= 645.55 - 275.77 \\ &= 369.78 \text{ MM BTU/HR.} \end{aligned}$$

$$\text{SPECIFIC ENTHALPY OF FEED TO HEATER} = \frac{369.78 \times 10^6}{1435385} = 257.6 \text{ BTU/LB.}$$

TEMPERATURE IS 510°F

ORIGINAL DESIGN WAS BASED ON A 500°F FEED TEMPERATURE TO CRUDE HEATER SO THIS IS NOT AN UNREASONABLE EXPECTATION

7. FROM YIELDS (SEE ITEM 2) NET OVERHEAD FROM PREFLASH TOWER IS 418 000 LB/HR. EXTERNAL REFLUX IS 418,000 LB/HR WITH SAME COMPOSITION



OVERHEAD TEMP = 300°F

BOTTOM TEMP = 570°F

REFLUX TEMP = 140°F

7 CONTINUED.

MATERIAL BALANCE

$$(FEED) + (REFLUX) = (OVERHEAD VAPOR) + (FLASHED CRUDE)$$

$$(OVERHEAD VAPOR) = (REFLUX) + (NET OVERHEAD)$$

HEAT BALANCE.

$$(FEED ENTHALPY) + (REFLUX ENTHALPY) + Q_R = (OVERHEAD VAPOR ENTHALPY) + (FLASHED CRUDE ENTHALPY)$$

	IN LB/HR	OUT LB/HR	SPECIFIC ENTHALPY BTU/LB	IN MM BTU/HR	OUT MM BTU/HR
FEED	1853385				
REFLUX (L)	418000				
REBOILER DUTY			44	18.39	
NET OVERHEAD (V)	418000	418000	263		109.93
REFLUX (V)	41	418000	263		109.98
FLASHED CRUDE (L)		1435385	257.6		369.78
TOTALS				$H_F + Q_R + 18.39$	588.65

$$Q_R + H_F = 588.65 - 18.39 = 570.26 \text{ MM BTU/HR.}$$

8. FEED TEMP ESTIMATED TO BE $510 - 25 = 485^\circ\text{F}$.

$$\begin{aligned} \text{AT THIS TEMPERATURE FEED ENTHALPY} &= (240)(1853385) \\ &= 444.81 \text{ MM BTU/HR} \end{aligned}$$

$$\text{REBOILER DUTY} = 570.26 - 444.81 = 125.45 \text{ MM BTU/HR.}$$

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT CRUDE UNIT REVAMP
ANSWERS

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9. CALCULATE DUTY REQUIRED TO HEAT 50,000 BPSD CRUDE FROM 60°F TO 258°F

$$\text{DUTY} = \overset{\text{LBS/HR}}{(617,795)} \overset{\text{BTU/LB}}{(100 - 0)} = 61.8 \text{ MM BTU/HR.}$$

$$\text{NAPHTHA PUMPAROUND RATE} = \frac{\overset{\text{BTU/HR}}{61.8 \times 10^6}}{(143 - 83) \overset{\text{BTU/LB}}{}} = 1030,000 \text{ LB/HR.}$$

TWO SHELLS IN SERIES REQUIRED AS A MINIMUM.

$$\text{TOTAL CONDENSER DUTY} = (109.935 - 18.39)(2) = 183.09 \text{ MM BTU/HR.}$$

MAKE HEAT BALANCE TO DETERMINE EXTERNAL REFLUX REQUIRED WITH NAPHTHA PUMPAROUND W PLACE.

$$H_F + Q_{\text{REB}} + \overset{\text{LIQ.}}{H_{\text{REFLUX}}} = H_{\text{FLASHED CRUDE}} + \overset{\text{VAPOR}}{H_{\text{REFLUX}}} + H_{\text{NET OVERHEAD}} + Q_{\text{PA.}}$$

$$\begin{aligned} \text{EXTERNAL REFLUX} &= \frac{(570.26 - 369.78 - 109.93 - 61.8) \times 10^6}{(263 - 44)} \\ &= \frac{28.75 \times 10^6}{219} = 131280 \text{ LB/HR} \end{aligned}$$

$$R/D = \frac{131280}{418000} = 31.4\% \quad \text{OK.}$$

10. DATA FOR DESALTED CRUDE TO FLASH ZONE CAN BE TAKEN DIRECTLY FROM CALCULATIONS MADE IN ITEM 6.

$$\text{REDUCED CRUDE TEMPERATURE} = 735 - 40 = 695^\circ\text{F.}$$

$$\text{STEAM TO CRUDE TOWER} = \frac{(8)(57150)}{24} = 19050 \text{ LB/HR}$$

$$\text{STEAM RATE} = 19050/18 = 1058.3 \text{ MOL/HR.}$$

$$\text{REDUCED CRUDE SPECIFIC ENTHALPY AT } 695^\circ\text{F} = 378 \text{ BTU/LB.}$$

CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT CANDE UNIT REVAMP.ANSWERSPAGE NO. 7DATE MAY 3, 1991

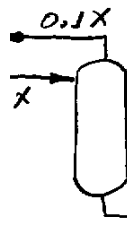
PROJECT

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10. CONTINUED.

AGO STRIPPER HEAT AND MATERIAL BALANCE.

AS BEFORE STRIPPED HYDROCARBON VAPOR 10% ON VOLUME BASIS OF FEED TO STRIPPER. SINCE PROPERTIES OF NET AGO PRODUCT HAVE NOT CHANGED AND STRIPPER PERCENTAGE THE SAME PROPERTIES OF THE STRIPPER HYDROCARBON VAPOR AND UNSTRIPPED AGO TO STRIPPER STAY THE SAME. DRAW TRAY TEMPERATURE IS 600°F.



$$\text{HYDROCARBON VOLUME BALANCE.} = 0.1X + 11325$$

$$(\text{UNSTRIPPED AGO}) = (\text{HYDROCARBON VAPOR}) + (\text{NET AGO PRODUCT})$$

$$(\text{HYDROCARBON VAPOR}) = (0.1)(\text{UNSTRIPPED AGO})$$

$$\Delta P = 34.5 = (29.5 + 5) \text{ PPM VAPOR NO STRIPPED.}$$

$$(\text{UNSTRIPPED AGO}) = (0.1)(\text{UNSTRIPPED AGO}) + 11325 \text{ BPSD}$$

$$\text{UNSTRIPPED AGO} = 11325 / 0.9 = 12583 \text{ BPSD.}$$

$$\text{HYDROCARBON VAPOR} = 12583 - 11325 = 1258 \text{ BPSD}$$

$$= (1258)(12.437) = 15645 \text{ LB/HR.}$$

$$\text{UNSTRIPPED AGO} = 15645 + 145225 = 160,870 \text{ LB/HR.}$$

$$\text{STEAM RATE IS SAME AS ORIGINAL DESIGN} = 3495 \text{ LB/HR.}$$

ENTHALPY BALANCE

$$H_{\text{UNSTRIPPED AGO}}^{\text{LIQ.}} + H_{\text{STEAM}}^{\text{VAPOR}} = H_{\text{VAPORS FROM STRIPPER}}^{\text{VAPOR}} + H_{\text{NET AGO PRODUCT}}^{\text{LIQUID}}$$

VAPORS FROM STRIPPER ARE AT 595°F

$$H_{\text{VAPORS FROM STRIPPER}}^{\text{VAPOR}} = H_{\text{STEAM}}^{\text{VAPOR}} + H_{\text{HYDROCARBON}}^{\text{VAPOR}}$$

SINCE TEMPERATURES AND COMPOSITIONS UNCHANGED, SPECIFIC ENTHALPY OF UNSTRIPPED AGO AND STRIPPER VAPOR ARE THE SAME AS ORIGINAL DESIGN.

$$H_{\text{NET AGO PRODUCT}}^{\text{LIQUID}} = [51.48 + 4.0 - (4.6 + 6.18)] \times 10^6 \text{ BTU/HR.}$$

$$= 44.7 \times 10^6 \text{ BTU/HR.}$$

$$\text{SPECIFIC ENTHALPY OF AGO PRODUCT} = \frac{44.7 \times 10^6}{145225} = 307.8 \text{ BTU/LB.}$$

$$\text{TEMPERATURE OF AGO FROM STRIPPER} = 583^\circ\text{F.}$$

CUSTOMER ENGINEERING DESIGN SEMINAL.SUBJECT CLUDE UNIT REVAMPPAGE NO. 8DATE NOV 3, 1991

PROJECT

BY EN MILLERANSWERS

10. CONTINUED.

DIESEL STRIPPER HEAT AND MATERIAL BALANCE. USE APPROACH OUTLINE ABOVE FOR AGO STRIPPER. NOTE NET PRODUCT RATES COME FROM YIELD STATEMENT DEVELOPED IN ITEM 2.

$$\text{UNSTRIPPED DIESEL} = \frac{13725}{0.9} = 15250 \text{ BPSD}$$

$$\begin{aligned} \text{STRIPPED HYDROCARBON} &= 15250 - 13725 = 1525 \text{ BPSD} \\ &= (1525)(12.038) = 18360 \text{ LB/HR.} \end{aligned}$$

$$\text{UNSTRIPPED DIESEL} = 18360 + 170190 = 188550 \text{ LB/HR.}$$

ENTHALPY BALANCE

$$\begin{aligned} H_{\text{NET DIESEL}}^{\text{LIQ}} &= [54.88 + 4.9 - (5.4 + 6.89)] \times 10^6 \text{ BTU/HR.} \\ &= 47.49 \times 10^6 \text{ BTU/HR.} \end{aligned}$$

$$\text{SPECIFIC ENTHALPY OF NET DIESEL} = \frac{47.49 \times 10^6}{170190} = 279 \text{ BTU/LB}$$

$$\text{TEMPERATURE OF NET DIESEL FROM STRIPPER} = 533^\circ\text{F}$$

JET FUEL STRIPPER.

$$\text{UNSTRIPPED JET FUEL} = \frac{24825}{0.9} = 27583 \text{ BPSD}$$

$$\begin{aligned} \text{STRIPPED HYDROCARBON} &= 27583 - 24825 = 2758 \text{ BPSD} \\ &= (2758)(11.515) = 31760 \text{ LB/HR} \end{aligned}$$

$$\text{UNSTRIPPED JET FUEL} = 31760 + 294055 = 325815 \text{ LB/HR}$$

ENTHALPY BALANCE

$$\begin{aligned} H_{\text{NET JET FUEL}}^{\text{LIQUID}} &= [68.55 + 8.8 - (9.4 + 9.75)] \times 10^6 \text{ BTU/HR.} \\ &= 58.2 \times 10^6 \text{ BTU/HR.} \end{aligned}$$

$$\text{SPECIFIC ENTHALPY OF JET FUEL PRODUCT} = \frac{58.2 \times 10^6}{294055} = 197.9 \text{ BTU/LB}$$

$$\text{TEMPERATURE OF JET FUEL PRODUCT} = 410^\circ\text{F.}$$

HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR.

CRUDE UNIT DESIGN - REVAMP

150 000 BPSD - ARABIAN LIGHT CRUDE

PAGE NO. 9

DATE 10-20-91

PROJECT

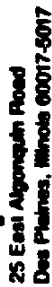
ENGINEERING DEPARTMENT

CRUDE COLUMN SECTION

BY EN MILLER

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STREAM	°API (LB/GAL)	K	MW	FLOWING SP. GR. (LB/CF)	°F	PSIG (MM HG ABS)	BPBD @ 60°F (10 ⁴ SCF/D)	FLOWING GPM (CF)	MOLES PER HR.	LBMM	BTU/ LB	10 ⁴ BTU/MM
DESALTED CRUDE TO FLASH ZONE					735	15				1435385		645.55
TOTAL VAPOR.										643540		328.81
NAPHTHA										39070	555	18.91
JET FUEL										294055	515	151.44
DIESEL										170190	507	86.29
AGO										145225	497	72.18
OVER FLASH VAPOR.										0		-
TOTAL LIQUID										791845	400	316.74
STEAM TO CRUDE COLUMN					298	50			6583	19050	1152	21.95
REDUCED CRUDE TO CRUDE EXCH.					695		57150			791845	378	299.32
AGO PUMParound TO CRUDE EXCHANGE												
EXCHANGER DUTY												
AGO PUMParound RETURN TO COLUMN												
UNSTRIPPED AGO TO STRIPPER	30.0	11.87	303.9		600		12583		529.33	160870	320	51.48
STEAM TO AGO STRIPPER			18		298	50			194.2	3495	1152	4.0
VAPORS FROM AGO STRIPPER			7417		595		(2347)		258.06	19140	533.2	10.78
STEAM			18		595				194.2	3495	1354	4.6
HYDROCARBON	34.5	11.87	245		595				6386	15645	395	6.18
AGO PRODUCT TO CRUDE EXCHANGE	29.5	11.87	312		583		11325		765.97	145225	308	44.7



ENGINEERING DESIGN SEMINAR

ENGINEERING DESIGN SEMINAR

CRYDE UNIT DESIGN - REVAMP

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PAGE NO. 10

DATE 10-20-91

PROJECT

BY **EV MILLER**

[illegible]

WOM-17-10294



25 East Algonquin Road
Des Plaines, Illinois 60017-5017

HEAT AND WEIGHT BALANCES

ENGINEERING DESIGN SEMINAR

CRUDE UNIT DESIGN - REVAMP

150,000 BPSD - ARABIAN LIGHT CRUDE

CRUDE COLUMN SECTION

ENGINEERING DEPARTMENT

PAGE NO. 16

DATE 10-20-91

PROJECT

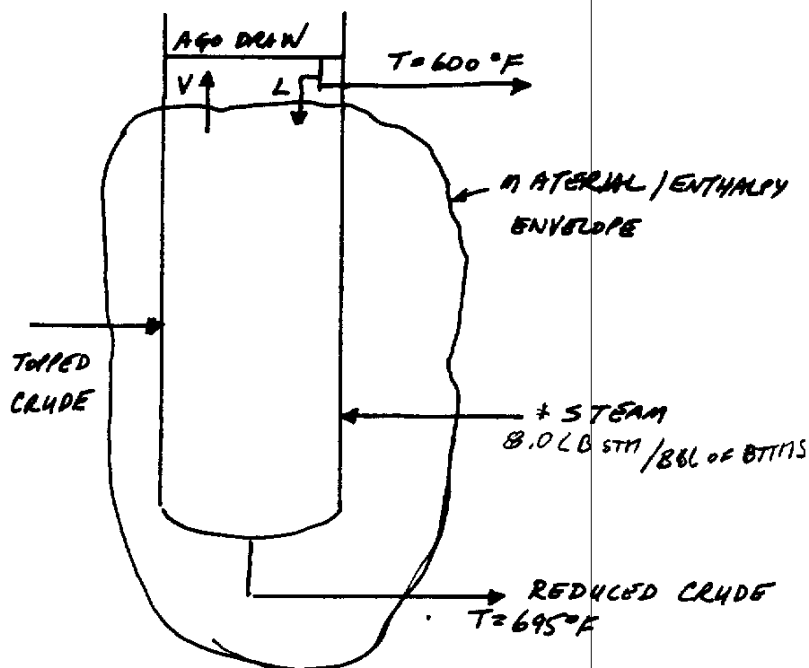
BY E N MILLER

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STREAM	°API (LB/GAL)	K	MM	FLOWING SP. GR. (LB/CF)	°F	PSIG (MM HG ABS)	BPSD @ 60°F (10 ⁴ SCF/D)	FLOWING GPM (CFS)	MOLES PER HR.	LB/MR	BTU/ LB	10 ⁴ BTU/MR
UNSTRIPPED JET FUEL TO STRIPPER	43.3	11.97	183.3		425					325815	2104	68.55
STEAM TO JET FUEL STRIPPER			18		298	50			425.5	7660	1152	8.8
VAPORS FROM JET FUEL STRIPPER			62.53		420				630.4	39420	485.8	19.15
STEAM			18		420				425.5	7660	1220	9.4
HYDROCARBON	47.8	11.97	155		420				204.9	31760	307	9.75
JET FUEL PRODUCT TO CRUDE EXCH.	42.8	11.97	187		410		24825		15725	294055	198	58.2
NAPHTHA PUMPAROUND TO CRUDE EXCH.												
EXCHANGER DUTY												
NAPHTHA PUMPAROUND RETURN TO COLUMN												
COLUMN OVERHEAD VAPOR TO AIR COND.												
STEAM												
NAPHTHA												
REFLUX												
AIR CONDENSER DUTY												
MATERIAL TO RECEIVER												
REFLUX TO COLUMN												
NAPHTHA TO DEBUT.												
WATER TO S.W. STRIPPER												

10. CONTINUED

NOW CALCULATE EXCESS HEAT FOR BOTTOM SECTION UP TO AGO DRAW AS DONE FOR THE CRUDE UNIT DESIGN.


 $L = \text{INTERNAL REFLUX}$
 $V = \text{TOTAL VAPOR RISING}$
 $= V_N + L_V$

WHERE

 $V_N = \text{NET RISING VAPORS}$
 $L_V = \text{INTERNAL REFLUX AS VAPOR.}$

ASSUME VAPORS TO AGO DRAW TRAY ARE AT 625°F.

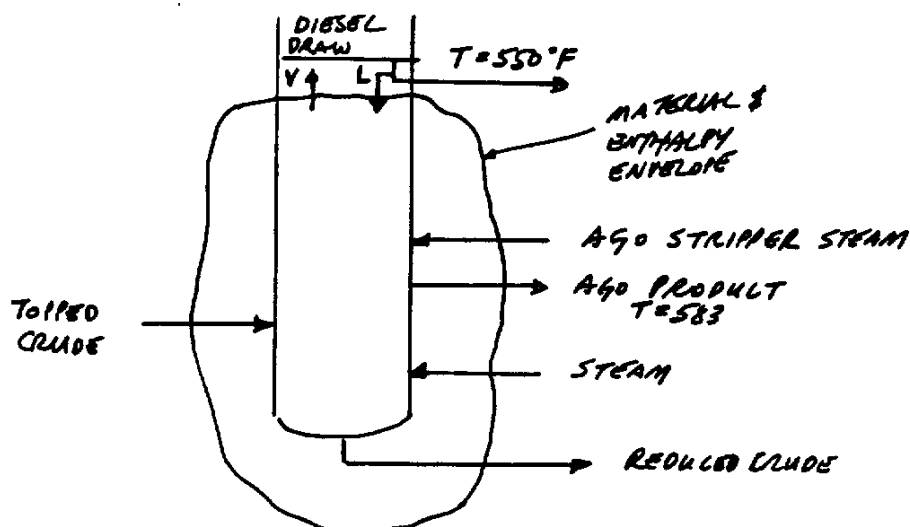
	MATERIAL BALANCE		ENTHALPY BTU/LB	ENTHALPY BALANCE	
	IN LB/HR	OUT LB/HR		IN MMBTU/HR	OUT MMBTU/HR
CRUDE (TOPPED)	1435385			645.55	
* STEAM	19050	19050	1319	21.95	25.13
INTERNAL REFLUX (L)	L			H_L	
INTERNAL REFLUX (V)		L_V			H_{LV}
NAPHTHA		34070	450		15.33
JET FUEL		294055	435		127.91
DIESEL.		170190	428		72.84
AGO.		145225	416		60.41
RESID		791845			299.32
				$H_L + 667.5$	$H_{LV} + 600.94$

HEAT AVAILABLE FOR INTERNAL REFLUX = $667.5 - 600.94 = 66.56$ MMBTU/HR

$$\text{EXCESS HEAT} = \left[\frac{66.56}{(25.13 + 15.33 + 127.91 + 72.84 + 60.41)} \right] \times 100 = \underline{\underline{22.07 \%}}$$

ANSWERS.

10. CONTINUED - CALCULATE EXCESS HEAT AT DIESEL DRAW TRAY

 $L = \text{INTERNAL REFLUX}$ $V = \text{TOTAL VAPOR RISING, WHOLE}$ $V_N = \text{NET RISING VAPORS}$ $L_V = \text{INTERNAL REFLUX AS VAPOR.}$ ASSUME VAPORS TO DIESEL DRAW TRAY ARE AT 575°F

	MATERIAL BALANCE			ENTHALPY BALANCE	
	IN LB/HR.	OUT LB/HR.	ENTHALPY. BTU/LB	IN MM BTU/HR.	OUT MM BTU/HR.
CRUDE (TOILED)	1435385			645.55	
STEAM.	19050	19050	1295	21.95	24.67
AG0 STR STEAM	3495	3495	1295	4.0	4.53
INTERNAL REFLUX (L)	L			H_L	
INTERNAL REFLUX (V)		L_V			H_{LV}
NAPHTHA		34070	417		14.21
JET FUEL		294055	400		117.62
DIESEL		170190	395		67.23
AG0 (L)		145225			44.7
RESID		791845			299.32
				$H_L + 671.5$	$H_{LV} + 572.28$

HEAT AVAILABLE FOR INTERNAL REFLUX = $671.5 - 572.28 = 99.2 \times 10^6 \text{ BTU/HR}$

$$\text{EXCESS HEAT} = \left[\frac{99.2 \times 10^6}{(24.67 + 4.53 + 14.21 + 117.62 + 67.23)} \right] \times 100 = 43.46\%$$

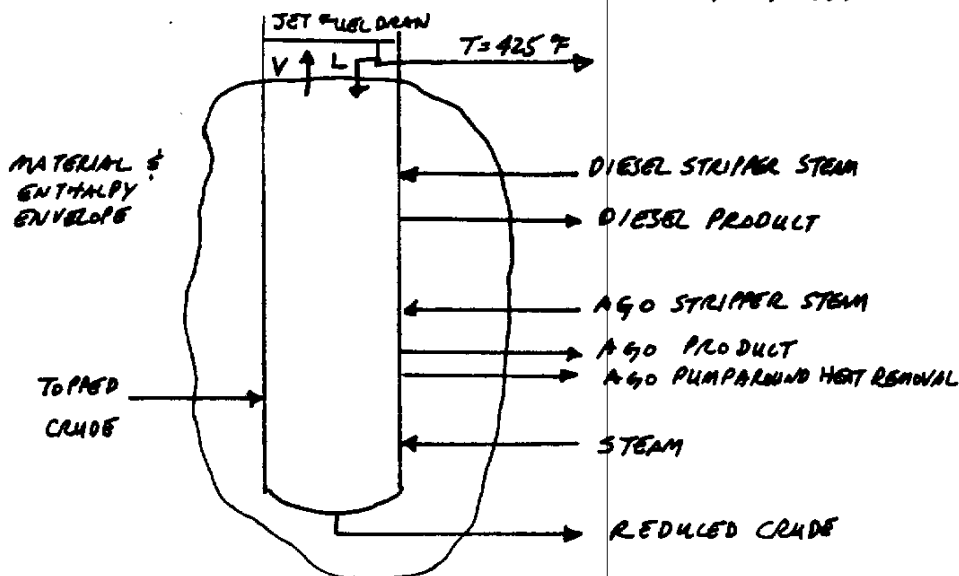
IF WE LIMIT EXCESS HEAT AT DRAW TO 20% CALCULATE HEAT THAT COULD BE REMOVED AT AG0 PUMP-AROUND = $99.2 - (0.2)(228.26)$

$$= 99.2 - 45.65 = 53.55$$

PUMP-AROUND HEAT AVAILABLE AT AG0 DRAW = $50 \times 10^6 \text{ BTU/HR.}$

ANSWERS

10. CONTINUED - CALCULATE EXCESS HEAT AT JET FUEL DRAW TRAY

 $L = \text{INTERNAL REFLUX}$ $V = \text{TOTAL VAPOR RISING WHERE}$ $V_N = \text{NET RISING VAPORS}$ $L_V = \text{INTERNAL REFLUX AS VAPOR}$

ASSUME VAPORS TO JET FUEL DRAW TRAY ARE AT 445°F

MATERIAL BALANCE

	IN LB/HR	OUT LB/HR	ENTHALPY BTU/LB
CRUDE (TOWED)	1435385		
STEAM	19050	19050	1232
AGO STR STEAM	3495	3495	1232
DIESEL STR STEAM	4240	4240	1232
INTERNAL REFLUX (L)	L		
INTERNAL REFLUX (V)		L_V	
NAPHTHA.		34070	340
JET FUEL		294055	322
DIESEL (L)		170190	
AGO (L)		145225	
RESID		791845	

AGO PUMP AROUND HEAT

ENTHALPY BALANCE

IN MMBTU/HR	OUT MMBTU/HR
645.55	
21.95	23.47
4.0	4.31
4.9	5.22
H_L	
	H_{LV}
	11.58
	94.69
	47.49
	44.70
	299.32
	<u>50.00</u>

$$H_L + 676.4 \quad H_{LV} + 580.78$$

$$\text{HEAT AVAILABLE FOR INTERNAL REFLUX} = 676.4 - 580.78 = 95.62 \times 10^6 \text{ BTU/HR}$$

$$\text{EXCESS HEAT} = \left[\frac{95.62 \times 10^6}{23.47 + 4.31 + 5.22 + 11.58 + 94.69} \right] \times 100 = 68.66\%$$

ANSWERS

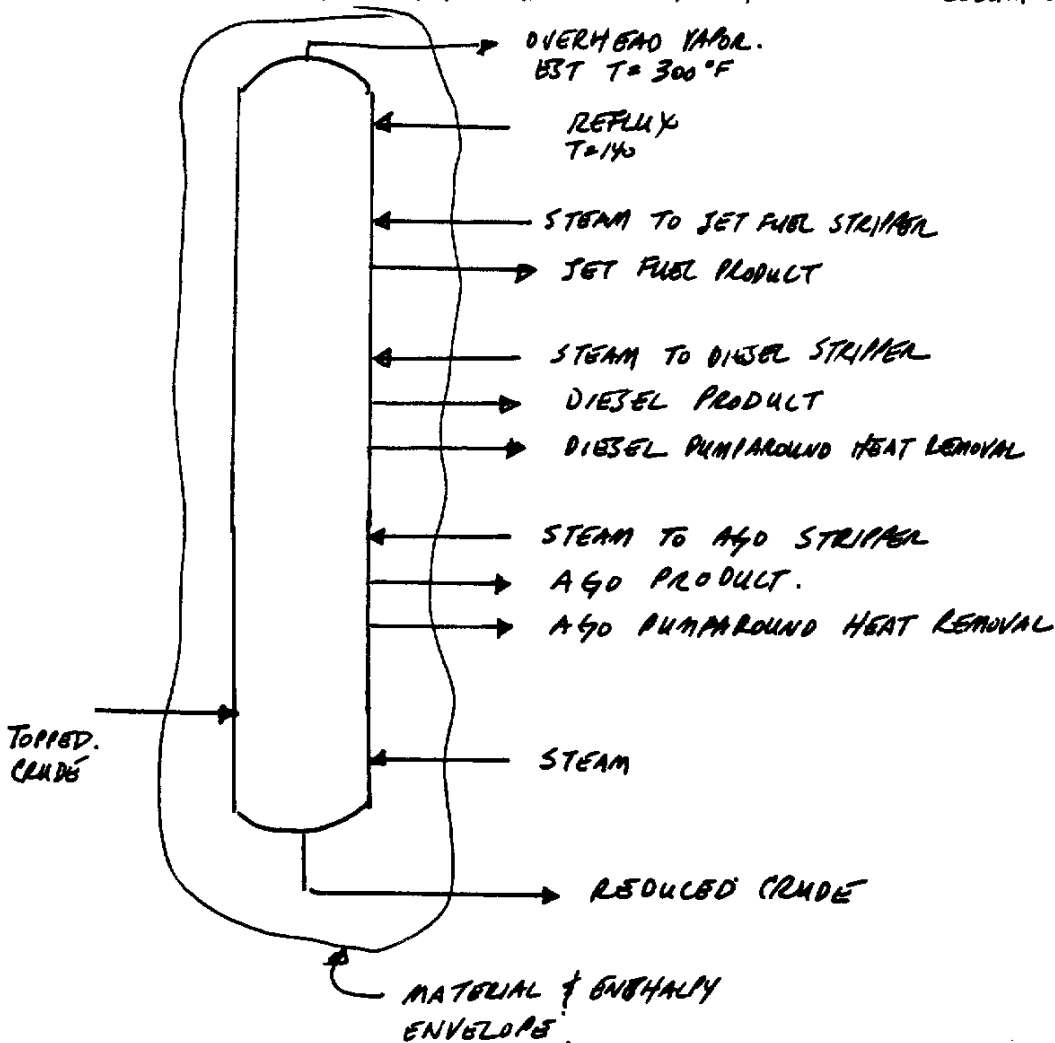
10. CONTINUED

IF WE LIMIT EXCESS HEAT AT JET FUEL DRAW TRAY TO 20%,
HEAT REMOVAL AVAILABLE AT DIESEL PUMPAROUND IS

$$= 95.62 - (0.2)(139.27) = \underline{67.77 \times 10^6 \text{ BTU/HR.}}$$

HEAT AVAILABLE AT DIESEL PUMPAROUND = 65 MM BTU/HR.

IGNORING NAPHTHA PUMPAROUND LOOK AT TOP OF COLUMN



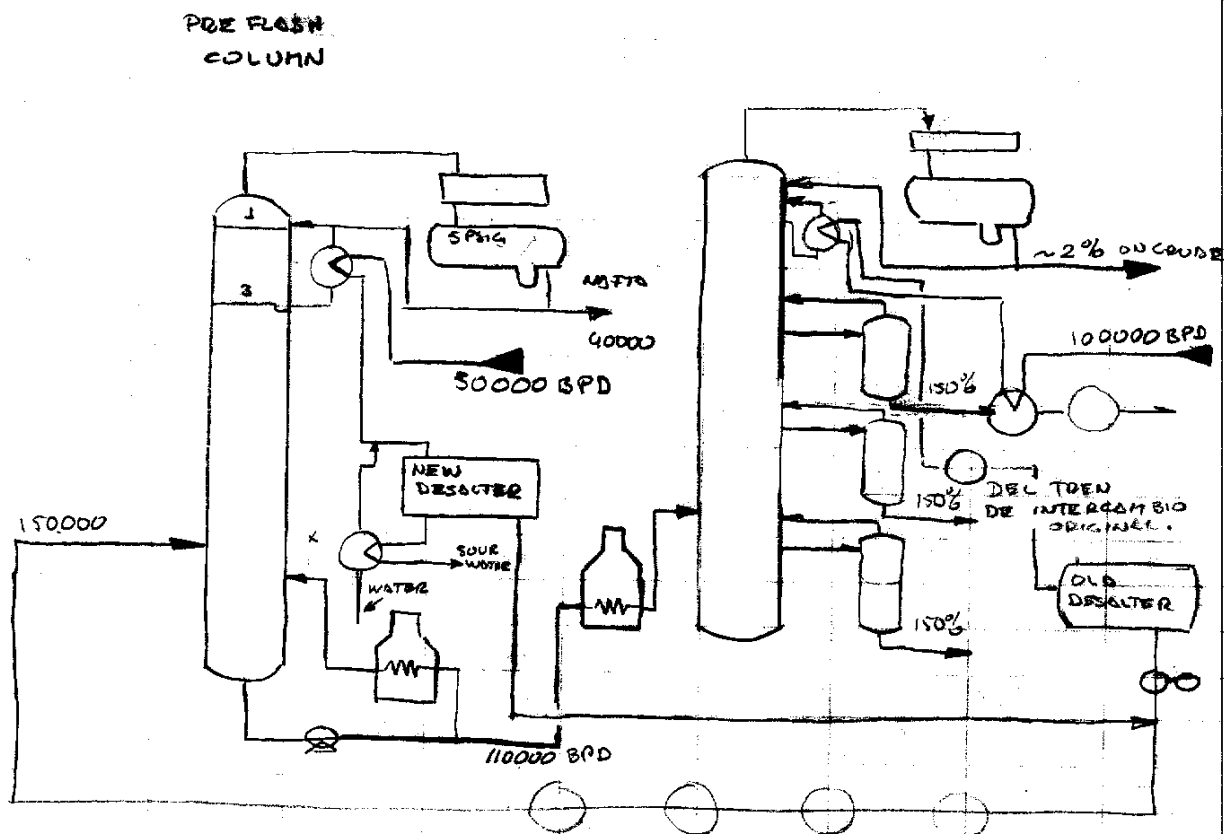
10. CONTINUED.

MATERIAL BALANCE			ENTHALPY BALANCE		
	IN	OUT		IN	OUT
	LB/HR.	LB/HR.	Btu/LB	MMBtu/HR	MMBtu/HR
CRUDE (TOPPED)	1435385			645.55	
STEAM.	19050	19050	1162.8	21.95	22.15
A90 STR STEAM	3495	3495	1162.8	4.0	4.06
DIESEL STR STEAM	4240	4240	1162.8	4.9	4.93
JET FUEL STR STEAM.	7660	7660	1162.8	8.8	8.907
REFLUX (L)	R _L			H _{RL}	
REFLUX (V)		R _V			H _{RV}
NAPHTHA.		34070	258		8.79
JET FUEL (L)		294055			58.20
DIESEL (L)		170190			47.49
A90 (L)		145225			44.70
RESID		791845			299.32
A90 P.A. HEAT.					50.0
DIESEL P.A. HEAT					65.0
				H _{RL} + 694.2	H _{RV} + 613.547

$$\text{EXCESS HEAT AT TOP OF TOWER} = (694.2 - 613.547) = 80.653 \text{ MMBtu/HR.}$$

$$\text{REFLUX REQUIRED TO REMOVE HEAT} = \frac{80.653 \times 10^4}{(258 - 43)} = 375,130 \text{ LB/HR}$$

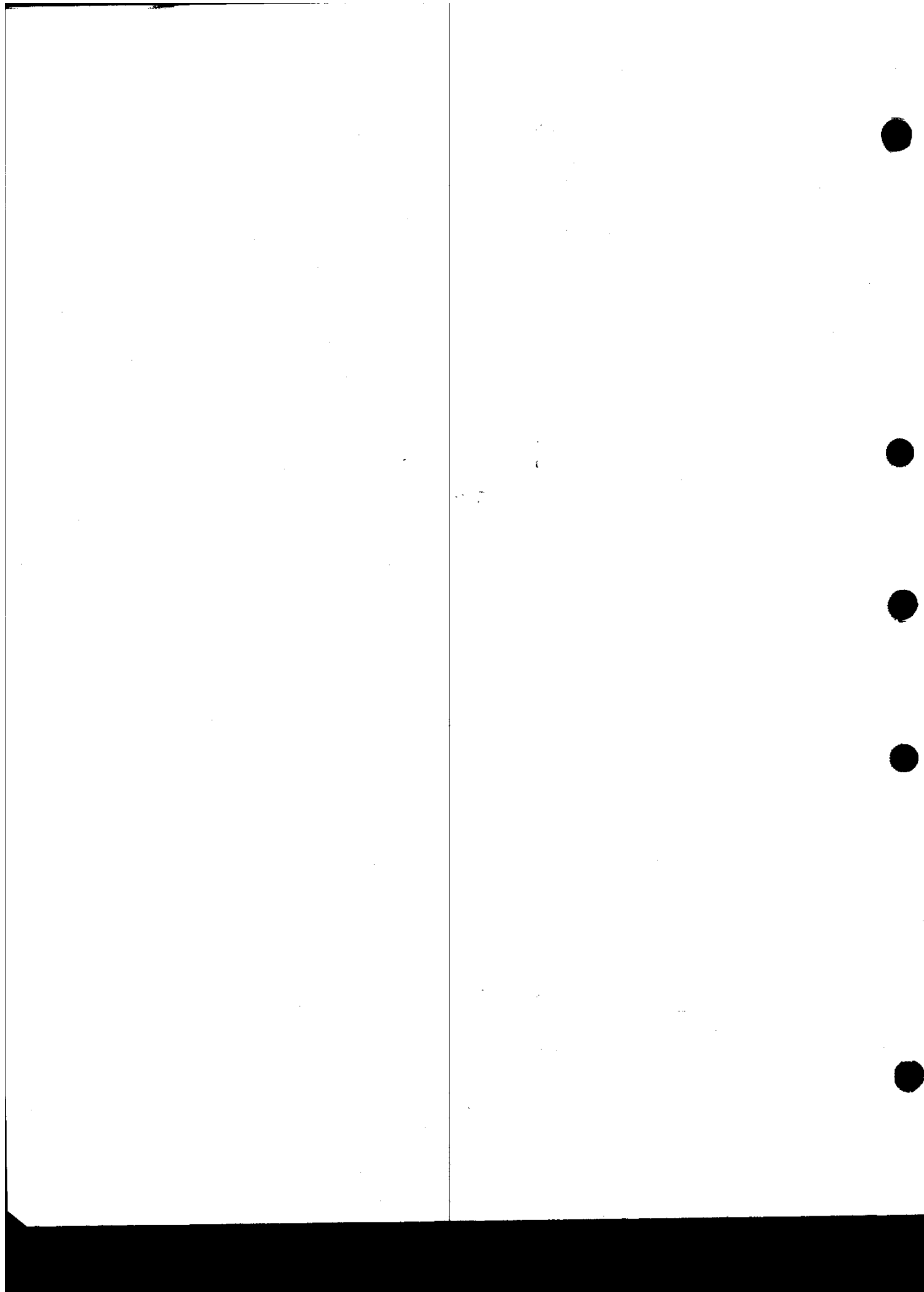
CRUDE UNIT REVAMP



$$K_1 W_1 + K_2 W_2 = K_3 W_3$$

$$K_1 = \frac{12.13 \times 403090 - 11.97 \times 34018}{369072} = 12.14$$

$$\Delta P_1 = \frac{60.4 \times 403090 - 50.3 \times 34018}{369072} = 61.33$$



NAPHTHA DE III 4A CRUDE UNIT DESIGN.

$$\text{NAPHTHA} = 2429.31 \text{ mol/hr} \times 1.5 = 3643.97$$

IF YOU RATED MOL/HR RATE BY 1.5 $\text{mol/hr} = \underline{3643.97}$

WE DIVIDED CALCULATED MASS FLOW BY MOL WT

$$\text{i.e. } \frac{403090}{111} = 3631.44 \text{ mol/hr.}$$

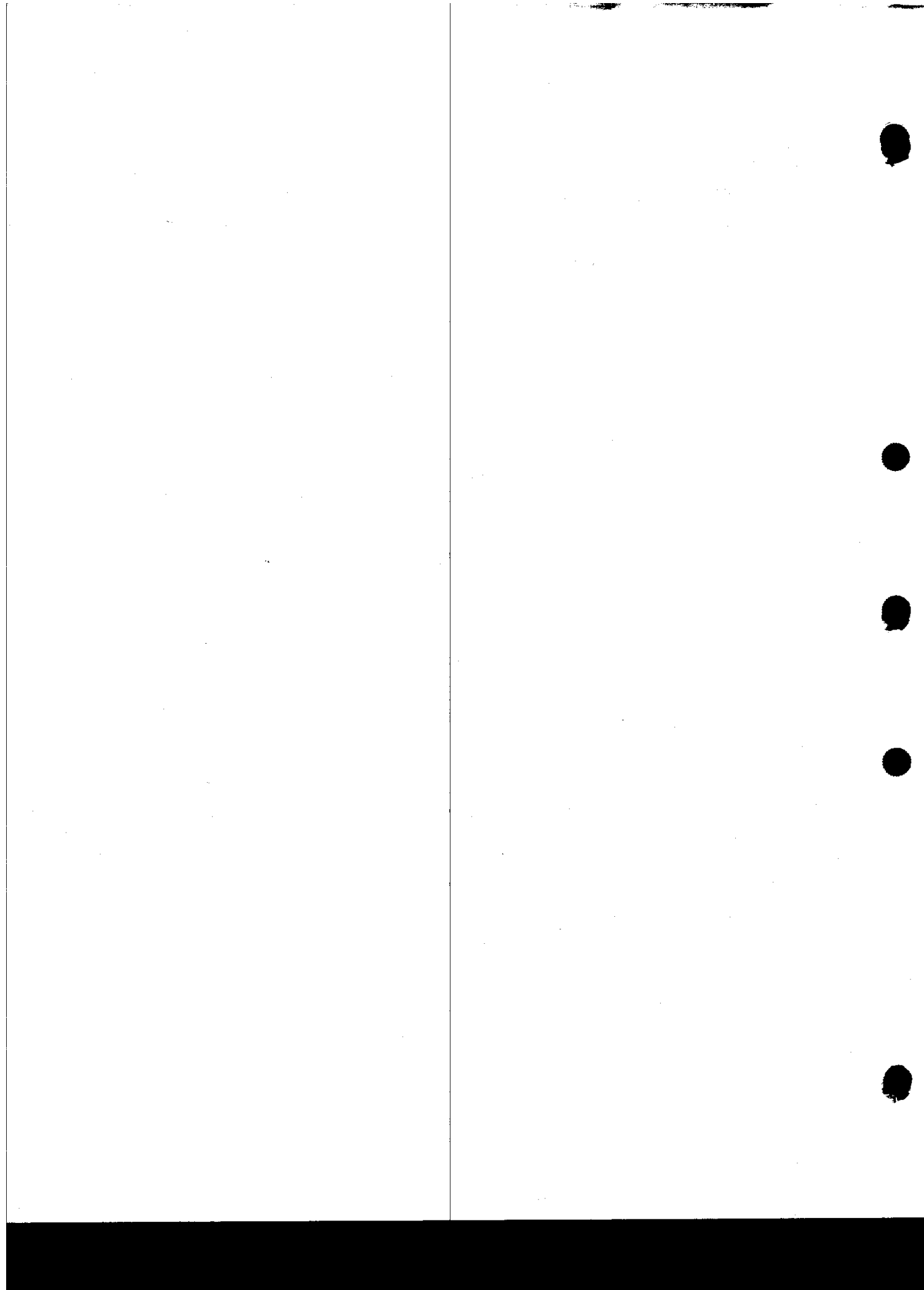
THERE IS SLIGHT DIFFERENCE DEPENDING ON HOW YOU CHOOSE TO DO CALCULATIONS HOWEVER THEY ARE FOR PRACTICAL PURPOSES THE SAME NUMBER

ESTIMATED YIELDS

	API	UOP R	MW	BPSD	LB/HR	MOL/HR	%LV
CRUDE	34.5	11.9	211.6	150000	1865550	8815.37	100%

ESTIM YIELDS

C3		14.7	44.094	255.0	1890.0	42.84	0.12
IC4		13.81	58.12	255.0	2092.5	36.05	0.12
NC4		13.50	58.12	1590.0	13560.0	233.24	1.06
IC5		13.05	72.146	1095.0	9982.5	138.36	0.73
NC5		13.03	72.146	2205.0	20302.5	281.43	1.47
CP		11.11	70.13	105.0	1147.5	16.40	0.07
PF NAPHTHA	61.33	12.14	108.6	34500	369072	3396.8	23.01
FLASHED CRUDE				110025	1435385	4658.51	73.35
HVY NAPHTHA	50.3	11.97	145	3000	34018	234.6	2.0
JET FUEL	42.8	11.97	187	24825	2940525	1572.48	16.55
DIESEL	35.0	11.93	253	13725	170190	672.64	9.15
A.G.O.	29.5	11.87	312	11325	1452225	465.45	7.55
RESID.	17.5	11.72	462.4	57150	791842.5	1712.48	38.10
	35.6	11.92	210.3	150000	1853370	8815.37	100.00



③

FOR OVA SYSTEM.

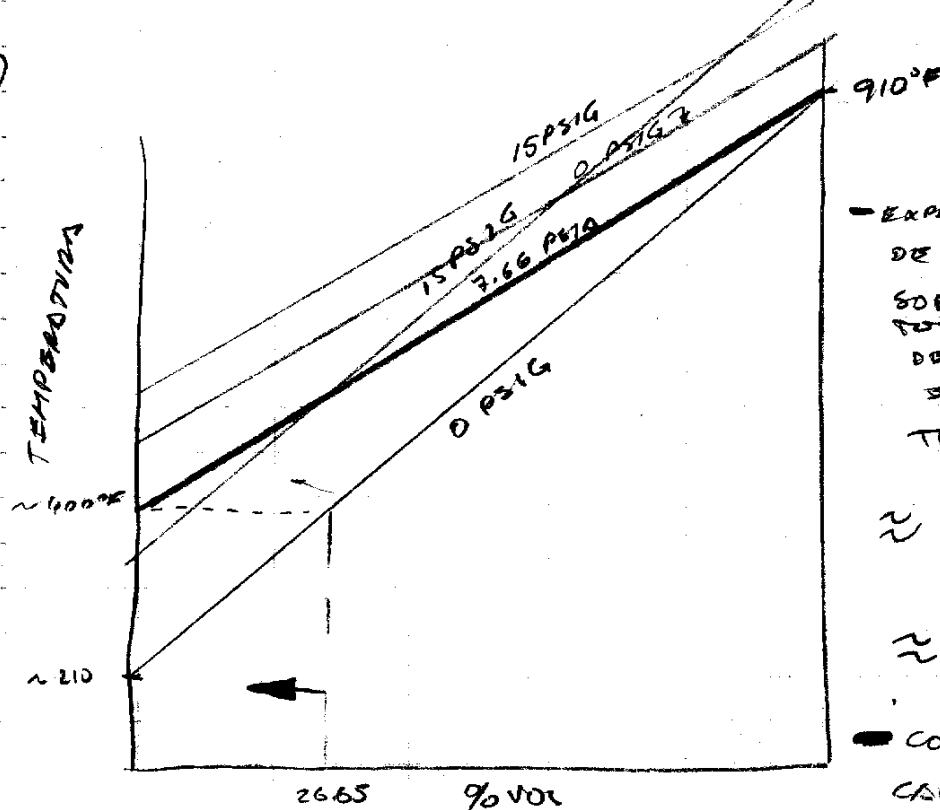
P = 19.7 PSIA

K_{TI} VAPOR @ 140°F = 3.33

CRUDE OVAH SYSTEM 140°F

	f _i	K _i	f _i K _i
C3	42.84	10.98	
IC4	36.08	5.27	
NC4	233.24	4.10	
IC5	138.36	1.9	
NC5	286.43	1.53	
CP	16.40	1.04	
PENAPTA	3396.8	0.169	

4195.12 >> 2901.4 Liquid. (SINGLE PHASE.)



REVISOR
CURVAS FLASH
PARA VACUUM
UNITS.

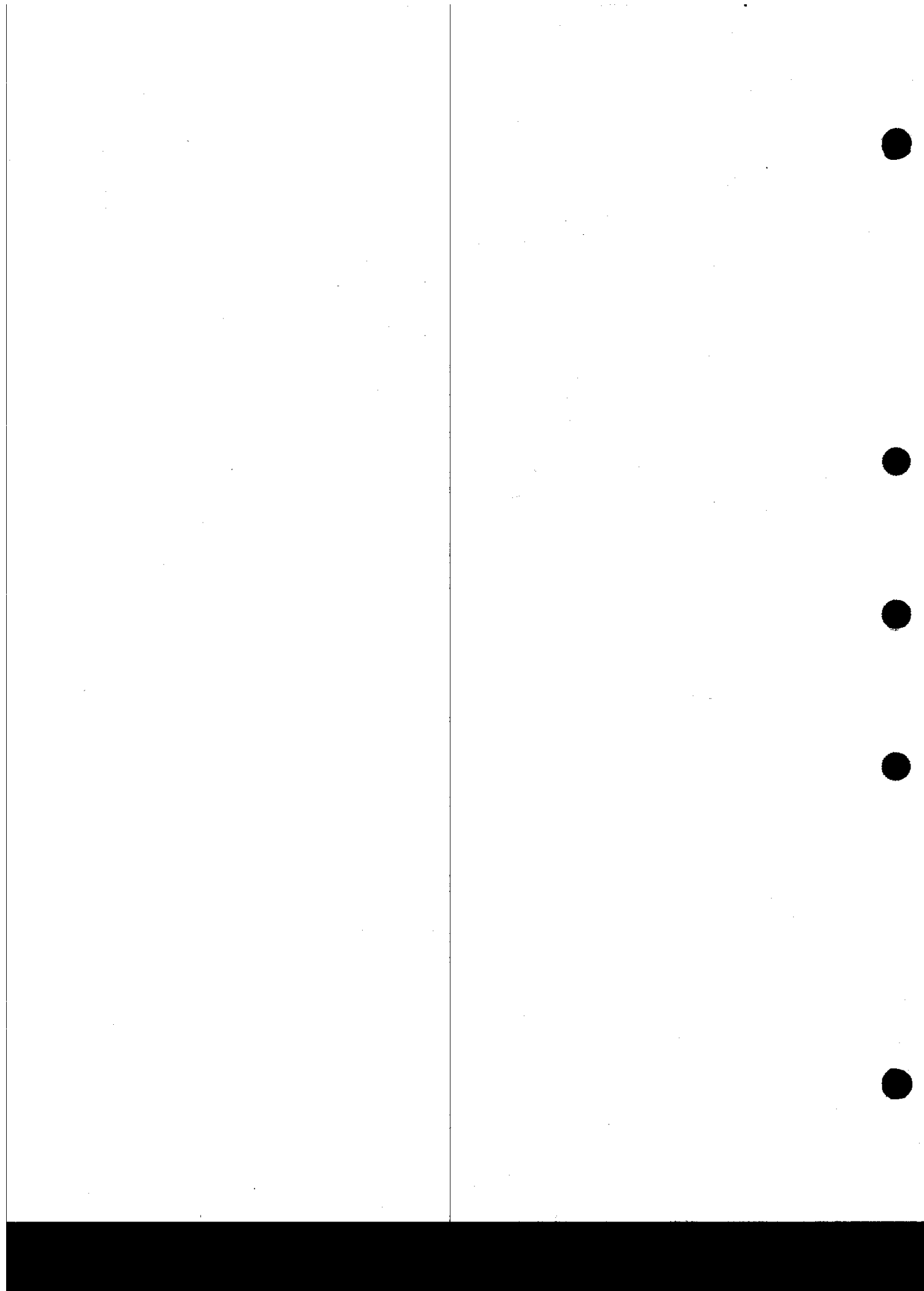
EXPANSION DE LA SECCION
DE LA CURVA DE LA FRACCION
SOBRE 2665 AL RANGO
TOTAL.
DETERMINAR PRESION
EQUIVALENTE OF
TOPPED CRUDE

$$\frac{\text{mol VAPOR}}{\text{mol TOTAL}} = 14.96$$

$$\approx \frac{4658}{8941} (14.96 \text{ PSIA})$$

$$\approx 7.66 \text{ PSIA}$$

CORRECCION MEDIANTE
CARTE DE COX
A 0 PSIG & 15 PSIG.



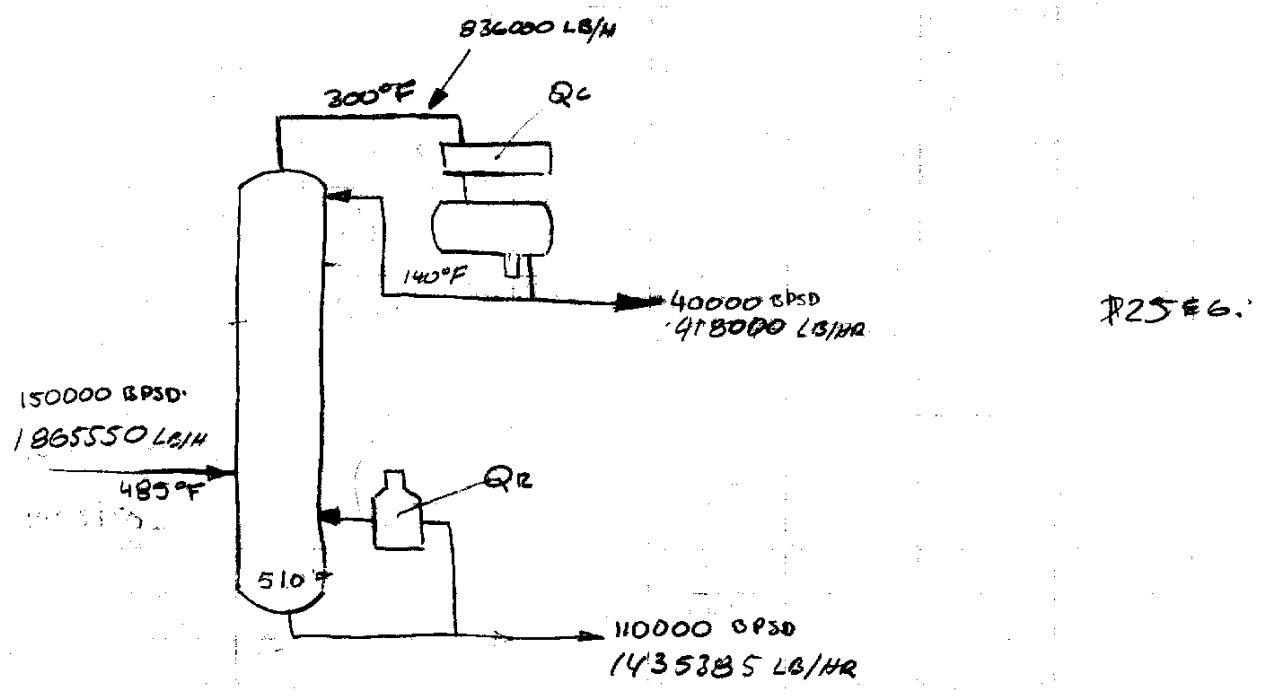
PARA TEMP CERRA A LO 2° COLUMNA. (TOPPED CRUDE FEED)
 SE UTILIZA EL PORCENTAJE DE CRUDO REDUCIDO + PARA
 DETERMINAR EL % DE LIFT

$$\% \text{ LIFT} = \frac{\text{FACC TOPPED CRUDE} - \text{FACC OF REDUCED CRUDE}}{73.35} \times 100 = 48\%$$

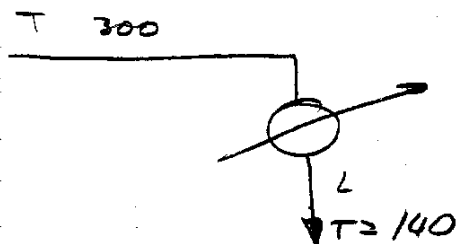
73.35

AT 48% OF LIFT TEMP $\approx 750^\circ\text{F}$
 CORRECTED FOR CURVATURE $750 - 15^\circ = \underline{\underline{735^\circ\text{F}}}$

HEAT AND MASS BALANCE FOR PRE FLASH

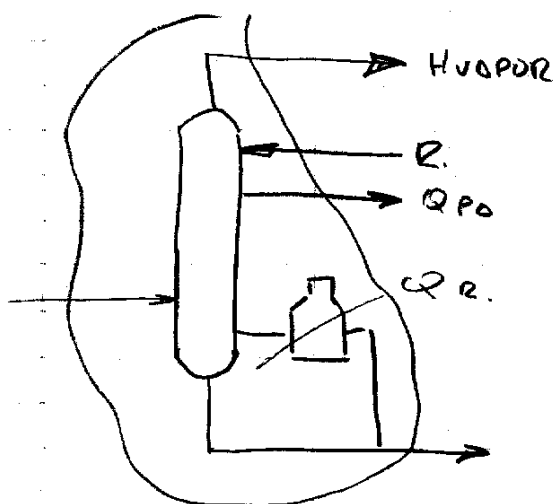






$$\begin{aligned} \text{VAPOR RATE} &= \text{NET} + \text{REFLUX} \\ &= 418000 + 418000 \\ L &= 832000 \text{ LB/HR.} \end{aligned}$$

$$\begin{aligned} 832000 \text{ LB/HR} \times 263 &= H_{\text{VAPOR}} = 109 \\ 832000 \text{ LB/HR} \times 44 &= H_{\text{LIQUID}} = 18.39 \end{aligned}$$



$$H_F + Q_R + H_R^{\text{LIQ}} = H_{\text{FLASH CRUDE}} + H_{\text{REFLUX}} + H_{\text{NET OVHD}} + Q_{PA}$$

$$H_{REF}^V - H_{REB} = H_F + Q_R - H_{FC} - H_{OVHD} - Q_{PA}$$

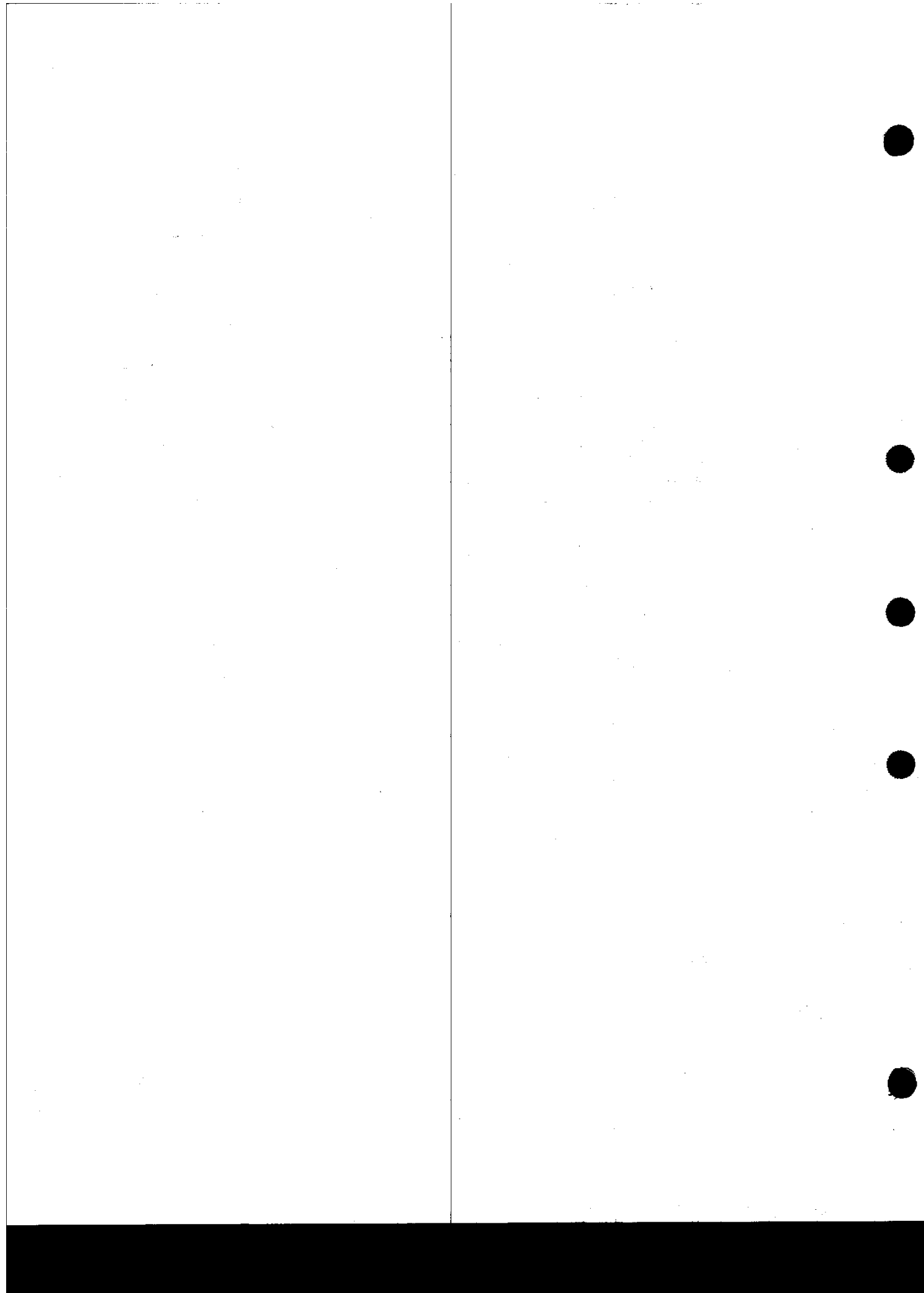
$$\begin{aligned} 444.81 + 125.45 - 369.78 - 109.93 - 61.8 \\ H_{REF}^V - H_{REB} = 28.75 \times 10^6 \end{aligned}$$

$$M_R \cdot H_R^V - M_P \cdot H_R^L$$

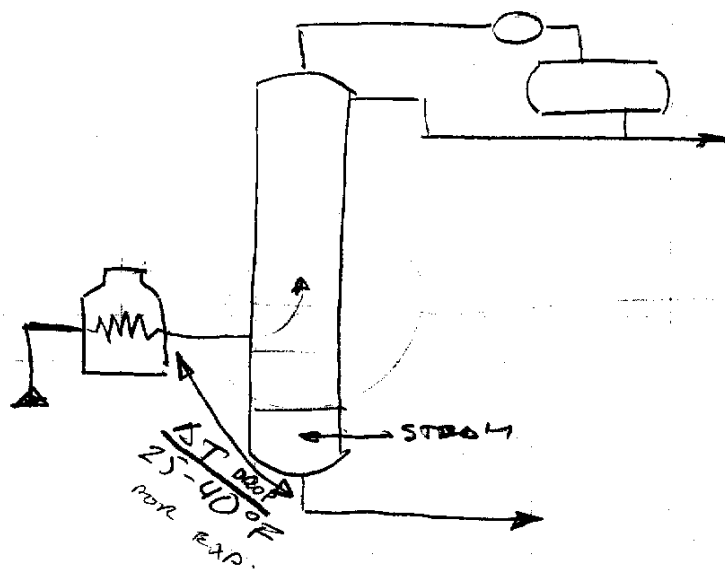
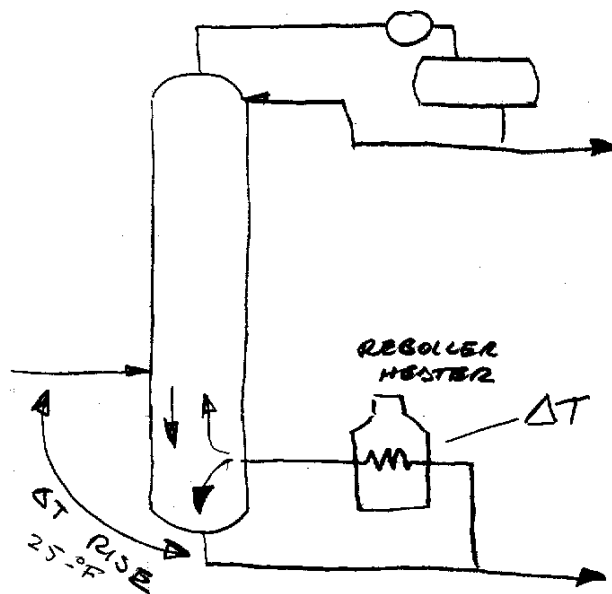
$$M_R [H_R^V - H_R^L] = 28.75 \times 10^6$$

$$M_R = \frac{28.75 \times 10^6}{263 - 44}$$

$$M_R = 131280 \text{ LB/HR.}$$

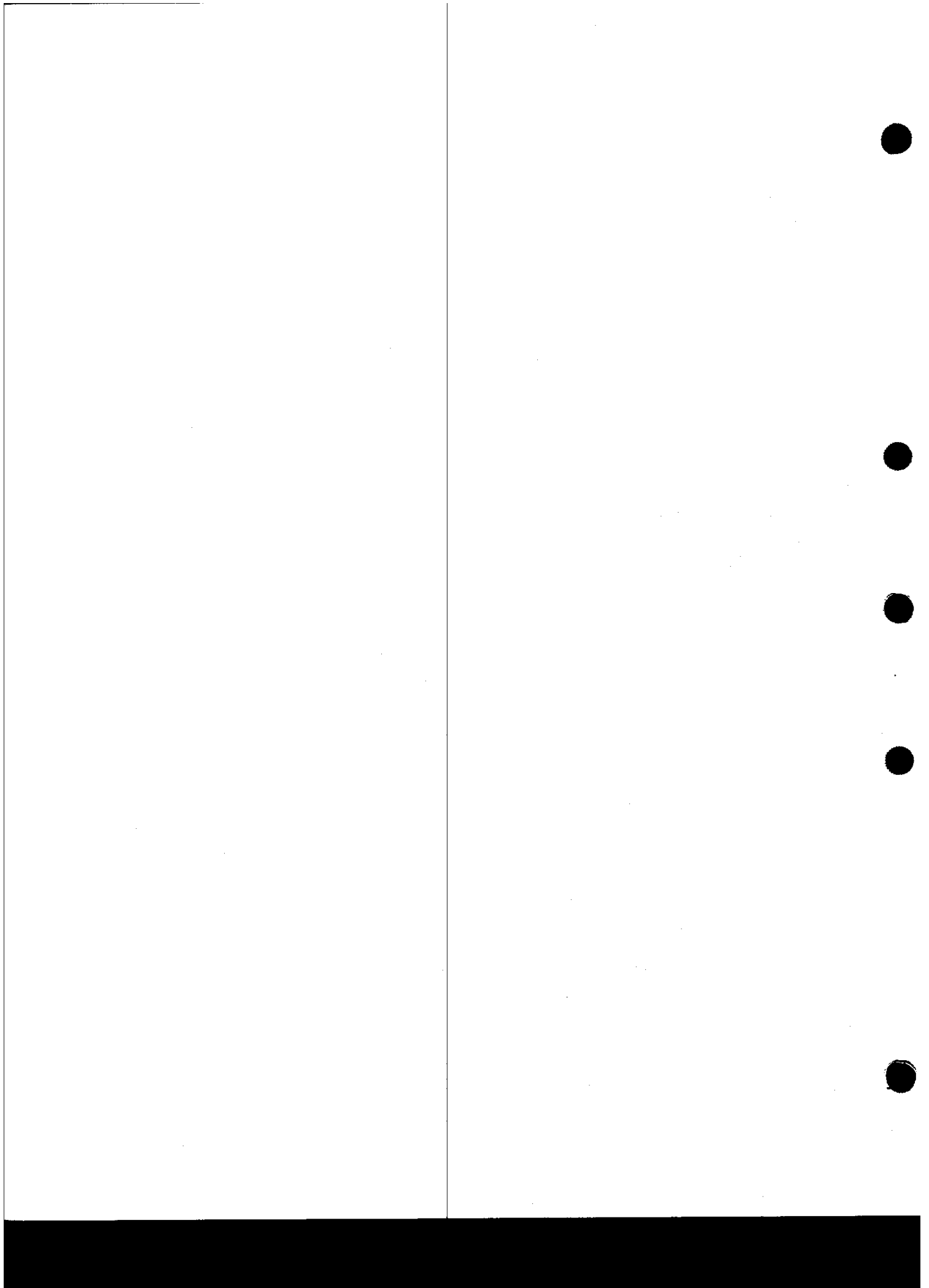


CRUDE UNIT REUAMP



ΔV_F
 ΔL_F

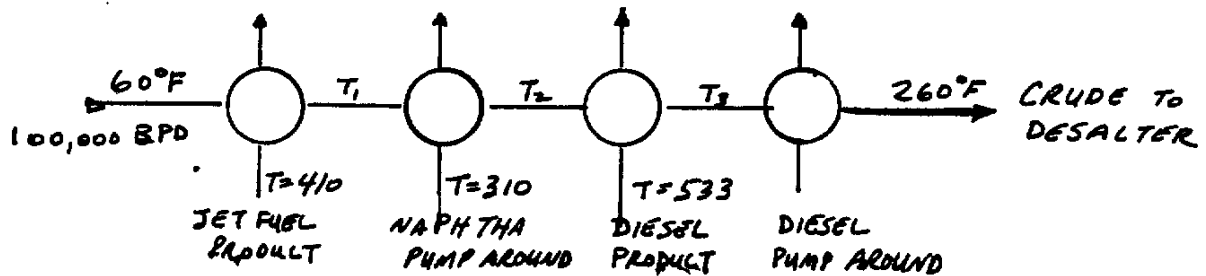
STEAM



"ANSWERS"

ITEM 12.

CRUDE TO DESALTER EXCHANGE TRAIN (REVAMP OPERATION)



FROM HEAT EXCHANGER SUMMARY (PAGE 24 OF HEAT EXCHANGE NETWORK WRITE UP)

$$\begin{aligned} \text{JET FUEL/CRUDE EXCHANGE } UA &= (3690)(55) = 0.20295 \times 10^6 \\ \text{NAPHTHA PUMP AROUND/CRUDE EXCHANGE } UA &= (14227)(55) = 0.7825 \times 10^6 \\ \text{DIESEL PRODUCT/CRUDE EXCHANGE } UA &= (3190)(50) = 0.1595 \times 10^6 \\ \text{DIESEL PUMP AROUND/CRUDE EXCHANGE } UA &= (2089)(50) = 0.1045 \times 10^6 \end{aligned}$$

SINCE CRUDE RATE IS UNCHANGED, WE CAN USE THE SAME CRUDE OUTLET TEMPERATURE FROM JET FUEL PRODUCT EXCHANGER.

$$T_1 = 110$$

$$Q = 30.3 \times 10^6 \text{ BTU/HR}$$

JET FUEL FROM EXCHANGER TEMPERATURE.

$$Q = \dot{M} H = H_{in} - H_{out} = 58.2 - H_{out} = 30.3$$

$$H_{out} = 58.2 - 30.3 = 27.9 \times 10^6 \text{ BTU/HR.}$$

$$\text{SPECIFIC ENTHALPY} = \frac{27.9 \times 10^6}{294055} = 94.9 \text{ BTU/LB.}$$

$$\text{TEMPERATURE} = 238^\circ\text{F}$$

"UA" REQUIRED FOR REVAMP OPERATION.

$$\frac{410 \rightarrow 238}{110 \leftarrow 60} \frac{172}{50} = 3.44 = R$$

$$\frac{300}{178} \frac{50}{350} = 0.14 = P$$

$$Q = UA \Delta T_{LM} \cdot F$$

$$\Delta T_{LM} = 233.7$$

$$F \approx 0.985$$

$$UA (\text{REQUIRED}) = \frac{30.3 \times 10^6}{(233.7)(0.985)} = 0.132 \times 10^6$$

$$UA (\text{AVAILABLE}) > UA (\text{REQUIRED}) \quad 0.203 \times 10^6 > 0.132 \times 10^6$$

EXCHANGER HAS EXCESS SURFACE FOR REVAMP.

"ANSWERS"

12. CONTINUED

WE DETERMINED THAT THE EXISTING JET FUEL/CRUDE EXCHANGER HAS MORE SURFACE THAN REQUIRED FOR THE REVAMP OPERATION AND WILL WORK. WE CAN EITHER CONTINUE WITH THE EXCHANGE EVALUATION OR ESTIMATE HOW MUCH HEAT CAN BE EXCHANGED WITH EXISTING EXCHANGER. THIS BECOMES A TRIAL AND ERROR CALCULATION.

ASSUME CRUDE FROM EXCHANGE TEMPERATURE IS 120°F

$$Q = [(39.5)(123550) - 0] = 36.45 \times 10^6 \text{ BTU/HR.}$$

WE CAN CALCULATE JET FUEL TEMPERATURE FROM EXCHANGER. ON A LMTD BASIS FROM PREVIOUS CALCULATION

$$\therefore \Delta T_2 / \Delta T_1 = Q_2 / Q_1 = \frac{36.45}{30.3} = 1.203.$$

$$\Delta T_2 = (1.203)(172) = 207 \quad T_{\text{OUT}} = 410 - 207 = 203^{\circ}\text{F}$$

$$\begin{array}{ccc} 410 \rightarrow 203 & 207 & \\ \frac{120 \leftarrow 60}{290} & \frac{60}{350} & = 3.45 = R \\ & & 0.17 = P. \end{array}$$

$$LMTD = 207.9$$

$$F = 0.955$$

$$U A_{\text{req}} = \frac{36.45 \times 10^6}{(207.9)(0.955)} = 0.184 \times 10^6 < 0.203 \times 10^6 \text{ AVAILABLE}$$

ASSUME CRUDE FROM EXCHANGE TEMPERATURE = 125

$$Q = \left(\frac{65}{60}\right)(36.45) = 39.49 \times 10^6 \text{ BTU/HR.}$$

$$\text{JET FUEL TEMPERATURE} = 410 - \left[\frac{(39.49)(36.45)}{(36.45)}\right] 207 = 410 - 224 = 186$$

$$\begin{array}{ccc} 410 \rightarrow 186 & 224 & \\ \frac{125 \leftarrow 60}{285} & \frac{65}{350} & = 3.75 = R \\ & & 0.186 = P \end{array}$$

$$LMTD = 194.8$$

$$F = 0.91$$

$$U A_{\text{req}} = \frac{39.49 \times 10^6}{(194.8)(0.91)} = 0.223 \times 10^6$$

12. CONTINUED

JET FUEL/CRUDE EXCHANGE. ADJUST U VALUE FOR INCREASED.
JET FUEL PRODUCT FLOW. I.P.

$$u'/u = \sqrt{[(1)(1.5)]^{0.7}} = (1.5)^{0.35} = 1.152.$$

$$UA'/UA = 1.152.$$

$$\text{FOR LAST ITERATION } UA'/UA = \frac{0.223 \times 10^4}{0.203 \times 10^4} = 1.098.$$

WE COULD MAKE ONE MORE ITERATION BUT CLOSE ENOUGH FOR OUR PURPOSES.

NAPHTHA PUMPAROUND / CRUDE EXCHANGE

FROM PREVIOUS CALCULATION (SEE ITEM 11) WE DETERMINED THAT 64×10^6 BTU/HR WAS AVAILABLE FROM NAPHTHA PUMPAROUND EXCHANGE. IN ORIGINAL DESIGN, NAPHTHA PUMPAROUND DUTY WAS 62.6×10^6 BTU/HR. FOR THE SAME ΔT FOR NAPHTHA PUMPAROUND, EVALUATE EXISTING EXCHANGER.

1. FIND CRUDE TEMPERATURE FROM EXCHANGER.

$$\text{ENTHALPY (BTU/LB)} = \frac{(64 + 39.49) \times 10^6}{1235580} = 83.76 \text{ BTU/LB}$$

FROM ENTHALPY/TEMP CHART TEMP = 227 °F

$$\begin{array}{rcl} 310 & \rightarrow & 210 \\ 227 & \leftarrow & 125 \\ \hline 83 & & 85 \end{array} \quad \begin{array}{l} \frac{200}{102} = 0.98 = R \\ \frac{185}{185} = 0.55 = P \end{array}$$

$$LMTD \approx 84$$

$$F \approx 0.66$$

SINCE ORIGINAL EXCHANGE, WAS ONE SHELL IN SERIES, THE HEAT TRANSFER IS GOING TO BE "DRIVING FORCE" LIMITED AND WE WILL NOT BE ABLE TO USE AVAILABLE PUMPAROUND HEAT WITH EXISTING EXCHANGER. TO ELIMINATE EXCHANGER TEMPERATURE CROSS, WE WILL NEED TO INCREASE THE NAPHTHA PUMPAROUND RATE. THIS REDUCES ΔT ON THE NAPHTHA PUMPAROUND SIDE FOR THE SAME HEAT TRANSFERRED.

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PROJECT

BY EN MILLER"ANSWERS"12. CONTINUED.

LET'S SET THE PUMP AROUND AT AT 80°F. I.E.
 SUPPLY TEMPERATURE = 310 °F, RETURN TEMPERATURE = 230 °F

$$\text{NAPHTHA PUMP AROUND RATE} = \frac{64 \times 10^6}{148 - 95} = 1207550 \text{ LB/HR}$$

$$\text{RATIO OF NEW RATE TO ORIGINAL} = 1207550 / 934330 = 1.292$$

$$\begin{array}{ccc} 310 \rightarrow 230 & 80 & \\ \frac{227}{83} \rightarrow \frac{125}{105} & \frac{102}{185} & = 0.784 = R \end{array}$$

$$\text{LMTD} = 13.57$$

$$F = 0.82$$

$$= 0.8341 \times 10^6$$

ADJUSTING UA FOR INCREASED NAPHTHA PUMP AROUND FLOW

$$u'/u = [(1)(1.292)]^{0.35} = 1.094$$

$$uA'/uA = 1.094$$

FOR CURRENT ITERATION

$$uA_{REQ} / uA_{ORIGINAL} = \frac{0.8341 \times 10^6}{0.7825 \times 10^6} = 1.066$$

SINCE "AVAILABLE" uA'/uA GREATER THAN uA'/uA "REQUIRED"
 EXISTING EXCHANGER WILL MEET PERFORMANCE EXPECTATION
 IN THE REVAMP OPERATION. I.E. $1.094 > 1.066$.

SINCE WE INCREASED NAPHTHA CIRCULATION BY ABOUT 30%,
 WE NEED TO CHECK EXISTING TRAYS IN THIS PART OF
 THE CRUDE TOWER TO BE SURE THAT THEY CAN HANDLE THE
 ADDITIONAL LIQUID LOAD. BECAUSE THE NET OVERHEAD FROM
 THIS TOWER FOR THE REVAMP OPERATION HAS BEEN GREATLY REDUCED
 WE DO NOT EXPECT ANY PROBLEMS. AT SOME POINT THIS
 ASSUMPTION NEEDS TO BE CONFIRMED.

"ANSWERS"

12. CONTINUED.

CRUDE / DIESEL PRODUCT EXCHANGE.

SINCE DIESEL PRODUCT RATE FOR THE REVAMP OPERATION IS 1.5 TIMES THE ORIGINAL DESIGN, WE SHOULD HAVE NO PROBLEMS IN HEATING THE CRUDE TO THE ORIGINAL TARGET TEMPERATURES OF 240°F IN THIS EXCHANGER. SO LETS SEE IF WE CAN HEAT THE CRUDE TO THE DESALTER OPERATING TEMPERATURE.

$$Q = [126.0 - (64 + 39.49)] \times 10^6 = 22.51 \times 10^6 \text{ BTU/HR.}$$

DETERMINE TEMPERATURE OF DIESEL PRODUCT FROM EXCHANGE

$$\text{ENTHALPY} = \frac{(47.49 - 22.51) \times 10^6}{170.190} = 146.78 \text{ BTU/LB.}$$

$$\text{TEMPERATURE} = 337^\circ\text{F}$$

$$\begin{array}{ccc} 533 \rightarrow 337 & \frac{196}{33} = 5.94 = R. \\ \frac{260}{273} \rightarrow \frac{227}{110} & \frac{33}{306} = 0.11 = P. \end{array}$$

$$\text{LMTD} = 179.3$$

$$F = 0.96.$$

$$U_{A_{REQ}} = \frac{22.51 \times 10^6}{(179.3)(0.96)} = 0.1308 \times 10^6$$

ADJUST UA FOR INCREASE DIESEL PRODUCT FLOW

$$U'/U = [(1)(1.5)]^{0.35} = 1.152.$$

$$U A' / U A = 1.152.$$

FOR CURRENT ITERATION

$$U_{A_{REQUIRED}} / U_{A_{ORIGINAL}} = \frac{0.1308 \times 10^6}{0.1045 \times 10^6} = 1.252$$

ON AN ADJUSTED BASIS $U_{A_{REQ}} / U_{A_{OILS}} > (U A' / U A) \text{ AVAILABLE}$

THEREFORE WE WOULD NOT EXPECT TO GET TO 260°F.

AT CRUDE TEMPERATURE OF 256°F FROM EXCHANGE.

$$Q = [(1235580)(100) - (64 + 39.49) \times 10^6] = 20.07 \times 10^6 \text{ BTU/HR.}$$

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12. CONTINUED

DETERMINE TEMPERATURE OF DIESEL PRODUCT FROM EXCHANGE

$$\text{ENTHALPY} = \frac{(47.49 - 20.07) \times 10^6}{170140} = 161.1 \text{ BTU/LB}$$

$$\text{TEMP} = 345^\circ\text{F}$$

$$\begin{array}{r} 533 \rightarrow 345 \\ \hline 256 \leftarrow 227 \\ \hline 277 \quad 118 \end{array} \quad \begin{array}{r} 188 \\ \hline 29 \\ \hline 306 \end{array} = 6.48 = R$$

$$\begin{array}{r} 29 \\ \hline 306 \end{array} = 0.095 = P$$

$$U_{A_{\text{REQ}}} = \frac{20.07 \times 10^6}{(186.3)(0.97)} = 0.11106 \times 10^6$$

$$\text{LMTD} = 186.3$$

$$F = 0.97$$

FOR CURRENT ITERATION $U_{A_{\text{REQ}}} / U_{A_{\text{OIL}}} = \frac{0.11106}{0.1045} = 1.063$

SINCE $(U_{A_{\text{REQ}}} / U_{A_{\text{OIL}}})_{\text{REVAMP}} < (U_{A'} / U_A)$

i.e. $1.063 < 1.152$

HENCE WE WOULD EXPECT TO BE ABLE OPERATE SOMEWHERE BETWEEN 256°F AND 260°F CRUDE TEMPERATURE FROM THE DIESEL PRODUCT EXCHANGE IN THE REVAMP OPERATION, SO FOR THE REVAMP DESIGN WE WILL PROVIDE A BYPASS AROUND THE DIESEL PUMP AROUND EXCHANGE.

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 SUBJECT CLAUDE UNIT REVAMP.
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13. RESULTED CLAUDE / DIESEL PUMP AROUND EXCHANGER (NEW)

CALCULATE CLAUDE TEMPERATURE FROM EXCHANGER (NEW)

$$\text{ENTHALPY} = \frac{[(617790)(97) + 18.5 \times 10^6]}{617790} = 126.95 \text{ BTU/LB}$$

$$\text{TEMPERATURE} = 304^\circ\text{F}$$

IF WE USE ONE SHELL, NO CROSS EXCHANGER THE MINIMUM PUMPAROUND RETURN TEMPERATURE IS 304°F .

CALCULATE PUMPAROUND RATE THROUGH THIS EXCHANGER.

$$\text{ADDED DIESEL PUMPAROUND RATE} = \frac{18.5 \times 10^6}{(291 - 124)} = 113500 \text{ LB/HR}$$

$$\text{OR TOTAL PUMPAROUND RATE} = 441175 + 113500 = 554675 \text{ LB/HR}$$

$$\text{OR } 1.26 \% \text{ OF ORIGINAL DESIGN RATE.}$$

CALCULATE EXCHANGER SURFACE REQUIREMENT.

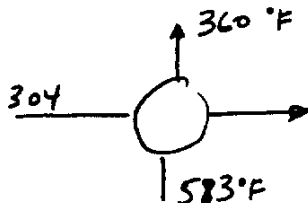
ASSUME $U \approx 50$.

$$\begin{array}{r} 550 \rightarrow 304 \\ 304 \leftarrow 250 \\ \hline 246 \quad 54 \end{array}$$

$$\text{LMTD} = 126.6 \quad F \approx 0.8$$

$$\begin{aligned} A &= \frac{18.5 \times 10^6}{(126.6)(0.8)(50)} \\ &= 3653 \text{ FT}^2 \end{aligned}$$

AGO PRODUCT EXCHANGER.



FOR 50°F COLD END APPROACH TEMPERATURE TEMPERATURE OF AGO PRODUCT FROM EXCHANGE $\approx 360^\circ\text{F}$

$$\begin{aligned} \text{CALCULATE DUTY : } &= \frac{44.7 \times 10^6 - (162)(145225)}{3} \\ &= 7.06 \times 10^6 \text{ BTU/HR.} \end{aligned}$$

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BY E. N. MILLER13. CONTINUED

CALCULATE CRUDE TEMPERATURE FROM EXCHANGE.

$$\text{ENTHALPY} = \frac{78.43 \times 10^6 + 7.06 \times 10^6}{617790} = 138.38 \text{ BTU/LB}$$

$$\text{TEMPERATURE} = 322.^\circ\text{F}$$

$$\begin{array}{r} 583 \rightarrow 360 \\ 322 \leftarrow 304 \\ \hline 261 \quad 56 \end{array} \quad \frac{223}{18} = 12.39 = R$$

$$\frac{18}{279} = 0.065 = P$$

$$\text{LMTD} = 133.2$$

$$F = 0.95$$

$$U \approx 50.$$

$$A = \frac{7.06 \times 10^6}{(133.2)(.95)(50)} = 1116 \text{ FT}^2$$

COLD REDUCED CRUDE EXCHANGE.

WE KNOW THE DUTY OF THE AGO PUMP AROUND. IF WE SET THE CRUDE FROM THE AGO PUMP AROUND AT 400°F WHICH WILL RESULT IN A ONE SHELL NO CROSS EXCHANGE FOR THE AGO PUMP AROUND WITH MAXIMUM ΔT ON AGO SIDE, WE CAN THEN CALCULATE DUTY OF COLD REDUCED CRUDE EXCHANGE.

$$114.29 \times 10^6$$

$$\begin{aligned} \text{TOTAL DUTY FOR BOTH SERVICES} &= (185)(617790) - 85.49 \times 10^6 \\ &= 28.8 \times 10^6 \text{ BTU/HR} \end{aligned}$$

SINCE AGO DUTY IS 20×10^6 BTU/HR, COLD REDUCED EXCHANGE DUTY IS 8.8×10^6 BTU/HR.

NOW CALCULATE CRUDE TEMPERATURE FROM EXCHANGE.

$$\text{ENTHALPY} = \frac{(85.49 \times 10^6 + 8.8 \times 10^6)}{617790} = 152.6 \text{ BTU/LB}$$

$$T = 348^\circ\text{F}$$

IF WE ASSUME 50°F COLD END APPROACH TEMPERATURE THEN REDUCED CRUDE TEMPERATURE FROM EXCHANGE IS $322 + 50 = 372^\circ\text{F}$.

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BY EW MILLER13. CONTINUED

CALCULATE TEMPERATURE OF REDUCED CRUDE TO COLD EXCHANGER.
REMEMBER ONLY $\frac{1}{3}$ OF TOTAL REDUCED CRUDE GOES TO THIS PART
OF THE EXCHANGE TRAIN.

$$\begin{aligned} \text{ENTHALPY OF RESID FROM EXCHANGE} &= \frac{(791845)(156)}{3} \\ &= 41.18 \times 10^6 \text{ BTU/HR.} \end{aligned}$$

$$\begin{aligned} \text{SPECIFIC ENTHALPY OF RESID TO EXCHANGE} &= \frac{(41.18 + 8.8) \times 10^6}{263948} \\ &= 189.36 \text{ BTU/LB.} \end{aligned}$$

TEMPERATURE OF RESID TO EXCHANGE = 428°F
FOR COLD REDUCED CRUDE / DESALTED CRUDE EXCHANGER.

$$\begin{array}{rcl} 428 & \rightarrow & 372 \\ \hline 348 & \leftarrow & 322 \\ \hline 80 & & 50 \end{array} \quad \frac{52}{26} = 2.15 = R$$

$$\frac{348}{80} \leftarrow \frac{322}{50} \quad \frac{26}{106} = 0.245 = P$$

$$\text{LMTD} = 63.83$$

$$F = 0.93$$

$$\begin{aligned} A &= \frac{8.8 \times 10^6}{(63.83)(0.93)(50)} \\ &= 2965 \text{ FT}^2. \end{aligned}$$

FOR A60 PUMPAROUND EXCHANGER.

$$600 \rightarrow 400$$

$$\frac{400}{200} \leftarrow \frac{348}{52}$$

$$\text{LMTD} = 109.87 \quad F = 0.8$$

$$\begin{aligned} Q &= 20 \times 10^6 \\ u &\approx 50 \end{aligned}$$

$$A = \frac{20 \times 10^6}{(109.87)(0.8)(50)} = 4551 \text{ FT}^2$$

HEAT AVAILABLE FROM REDUCED CRUDE FOR HOT REDUCED CRUDE/
CRUDE EXCHANGE IS

$$Q = \frac{299.32}{3} - 8.8 - 41.18 = 44.79 \times 10^6 \text{ BTU/HR}$$

TEMPERATURE OF DESALTED CRUDE FROM THIS EXCHANGER

$$\text{ENTHALPY} = \frac{(85.49 + 28.8 + 44.79) \times 10^6}{617790} = 285.6 \text{ BTU/LB.}$$

$$\text{TEMPERATURE} = 520^\circ\text{F.}$$

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13. Continued.

HOT REDUCED CRUDE / DESALTED CRUDE EXCHANGER.

$$\begin{array}{rcl} 695 & \rightarrow & 428 \\ \underline{520} & \leftarrow & \underline{400} \\ 175 & & 28 \end{array} \quad \begin{array}{l} \frac{267}{120} = 2.23 = R \\ \frac{120}{295} = 0.41 = P \end{array}$$

$$C_p = 49.79 \times 10^6 \text{ BTU/HR.}$$

$$U \approx 50.$$

$$LMTD = 80.2$$

$$F = 0.87$$

REQUIRES MINIMUM OF
THREE SHELLS IN SERIES

$$A_{\text{total}} = \frac{49.79 \times 10^6}{(80.2)(0.87)(50)}$$

$$= 14272 \text{ FT}^2$$

$$\text{AREA PER SHELL} = 14272 / 3 = 4757 \text{ FT}^2$$

14. CALCULATE ENTHALPY AND TEMPERATURE OF TOTAL CRUDE
TO PREFLASH TOWER.

FROM ORIGINAL DESIGN HT & WT BALANCE.

BTU/D	LB/HR.	TEMPERATURE	ENTHALPY MM BTU/HR
100,000	1235580	500	310.1
50,000	617790	520	164.08
TOTALS			
150,000	1853370	507°F	474.18
		=	=

Engineering Design Seminar
Crude Unit Design

M E R

November 1991



CUSTOMER ENGINEERING DESIGN SEMINAR
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WHAT IS MER. ?

- MINIMUM ENERGY REQUIREMENT
- MAXIMUM ENERGY RECOVERY

A PART OF 'PINCH TECHNOLOGY'
SOMETIMES KNOWN AS 'LINNHOFF' ANALYSIS

PINCH TECHNOLOGY IS PART OF AN
EMERGING ANALYSIS METHODOLOGY
KNOWN AS PROCESS DESIGN SYNTHESIS.

USED PRIMARILY FOR HEAT UTILIZATION
EVALUATIONS

ONLY PORTION OF PROCESS DESIGN SYNTHESIS
TECHNOLOGY WIDELY USED COMMERCIALY.

WHY USE LINNHOFF ?

- TARGETS FOR MINIMUM ENERGY REQUIREMENTS
- DESIGN OF NETWORKS FOR M.E.R.
- PROPOSE CHANGES IN PROCESS CONDITIONS
- CONSISTENT BASIS FOR PROCESS EVALUATIONS
- FUNDAMENTAL ANALYSIS OF COMPETITION
- EVALUATE ALTERNATE SCHEMES WITHOUT FULL H W B

HEAT
WEIGHT
BALANCES

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CUSTOMER ENGINEERING DESIGN SEMINAR
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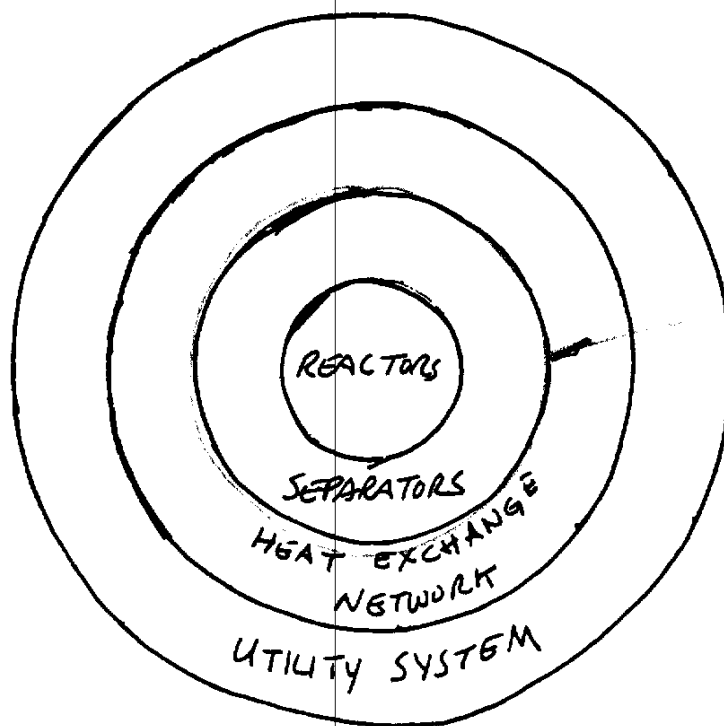
HOW TO USE MER?

- ^{CONSULTORIA} CONSULTANCY SERVICES.
- COMMERCIAL SOFTWARE
PROCESS SIMULATION PACKAGES.
(INCLUDED IN MOST)
DEDICATED SOFTWARE PROGRAMS.
- TECHNIQUE PUBLISHED IN OPEN LITERATURE
INTRODUCTION TO USE OF MER

CUSTOMER ENGINEERING DESIGN SECTOR
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ONION SKIN REPRESENTATION OF PROCESS DESIGN HIERARCHY



BE GENERAL
 HEAT &
 WEIGHT
 BALANCES

CORE - REACTORS $\begin{cases} \text{FEEDS} \\ \text{PRODUCTS} \\ \text{RECYCLES} \end{cases}$ FLOW RATE
SYSTEMS CONSTRAINTS

- SEPARATORS

AT THIS POINT BASIC PROCESS HEAT AND MASS BALANCE KNOWN

- 'STREAM DATA'

- HEAT RECOVERY SYSTEM $\begin{cases} \text{TARGETS} \\ \text{ECONOMICS} \end{cases}$

- UTILITIES $\begin{cases} \text{HOW MUCH} \\ \text{WHERE} \\ \text{LEVEL} \end{cases}$

TRADITIONAL APPROACH TO H.E.N.
(H.E.N. = HEAT EXCHANGE NETWORK)

- ① PROBLEM DEFINITION.
- ② DESIGN DERIVED FROM PREVIOUS EXPERIENCE

PINCH TECHNOLOGY PROPOSAL

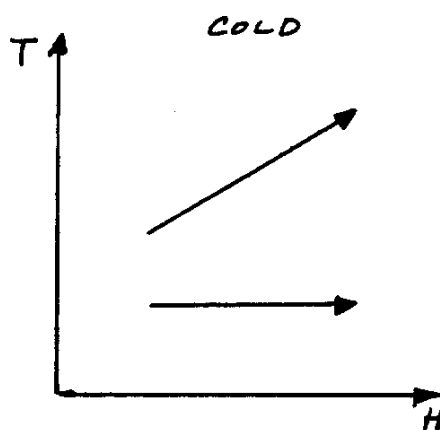
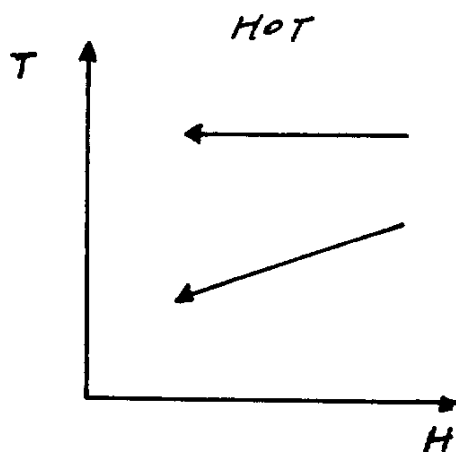
- ① PROBLEM DEFINITION
- ② TARGETS SET FOR MINIMUM.
ENERGY AND CAPITAL.
- ③ DESIGN DERIVED WHICH MEETS TARGETS.

STREAM DATA

"HOT" STREAMS REQUIRE COOLING.

"COLD" STREAMS REQUIRE HEATING.

T-H DIAGRAM.



DEFINITIONS.

H = STREAM ENTHALPY (RELATIVE).

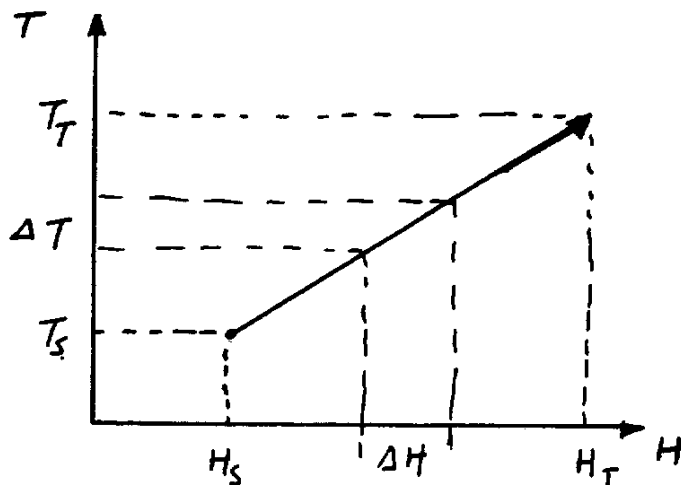
$$CP = (M)(C_p) = \frac{\Delta H}{\Delta T} = \frac{Q}{\Delta T}$$

CP = HEAT CAPACITY FLOW RATE

$$\left(\frac{\text{BTU}}{\text{HR} \cdot ^\circ\text{F}} \text{ OR } \frac{\text{MW}}{^\circ\text{C}} \right)$$

REMEMBER.

$$Q = UA \Delta T_{LM}$$

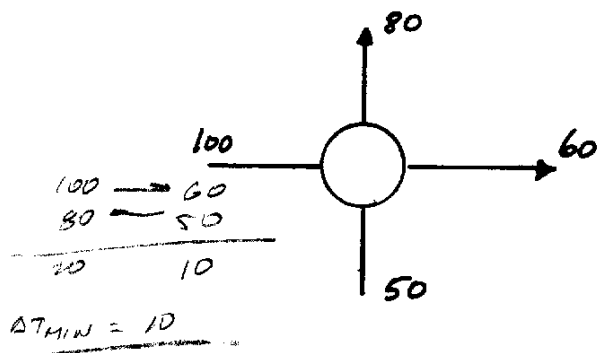


T_S = STREAM SOURCE TEMPERATURE T_S^H = HOT STREAM.
TEMP INITIAL SOURCE TEMPERATURE.

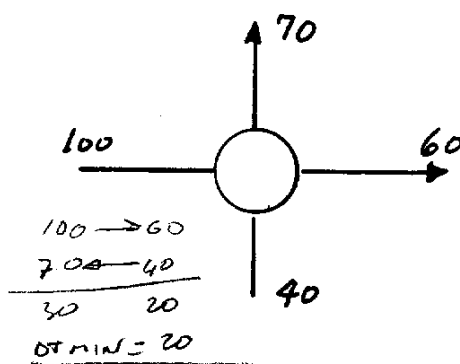
T_T = STREAM TARGET TEMPERATURE T_T^C = COLD STREAM.
TEMP FINAL TARGET TEMPERATURE.

ΔT_{MIN} = MINIMUM ALLOWABLE ΔT FOR EXCHANGE.
EXPERIENCE

EXAMPLES.



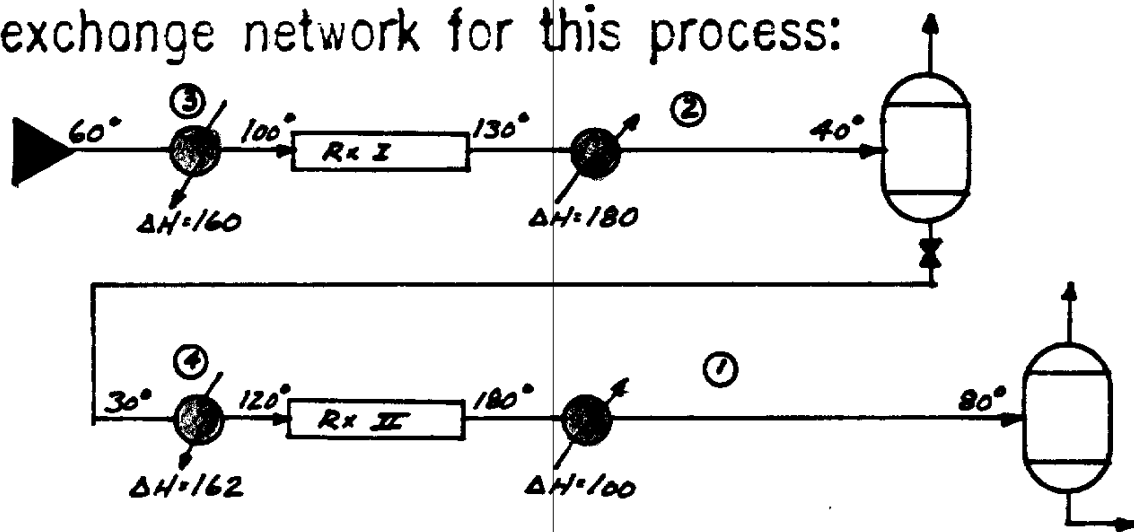
$\Delta T_{MIN} = ??$



$\Delta T_{MIN} = ??$

I. INTRODUCTORY EXAMPLE

Design an energy efficient heat exchange network for this process:



minimum approach $\Delta T = 10$

What is minimum heating duty?

What is minimum cooling duty?

How many services are required?

STREAM DEFINITIONS.

"HOT" STREAMS. REQUIRED ENTHALPY

$$\text{STREAM [1]} \quad T_{S1}^H = 180 \quad T_{T1}^H = 80 \quad \Delta H = 100 = Q$$

$$\text{HEAT CAPACITY} \quad CP_1 = \frac{100}{180 - 80} = 1.0$$

$$\text{STREAM [2]} \quad T_{S2}^H = 130 \quad T_{T2}^H = 40 \quad \Delta H = 180$$

$$CP_2 = \frac{180}{130 - 40} = 2.0$$

"COLD" STREAMS. REQUIRED COUNTERFLOW

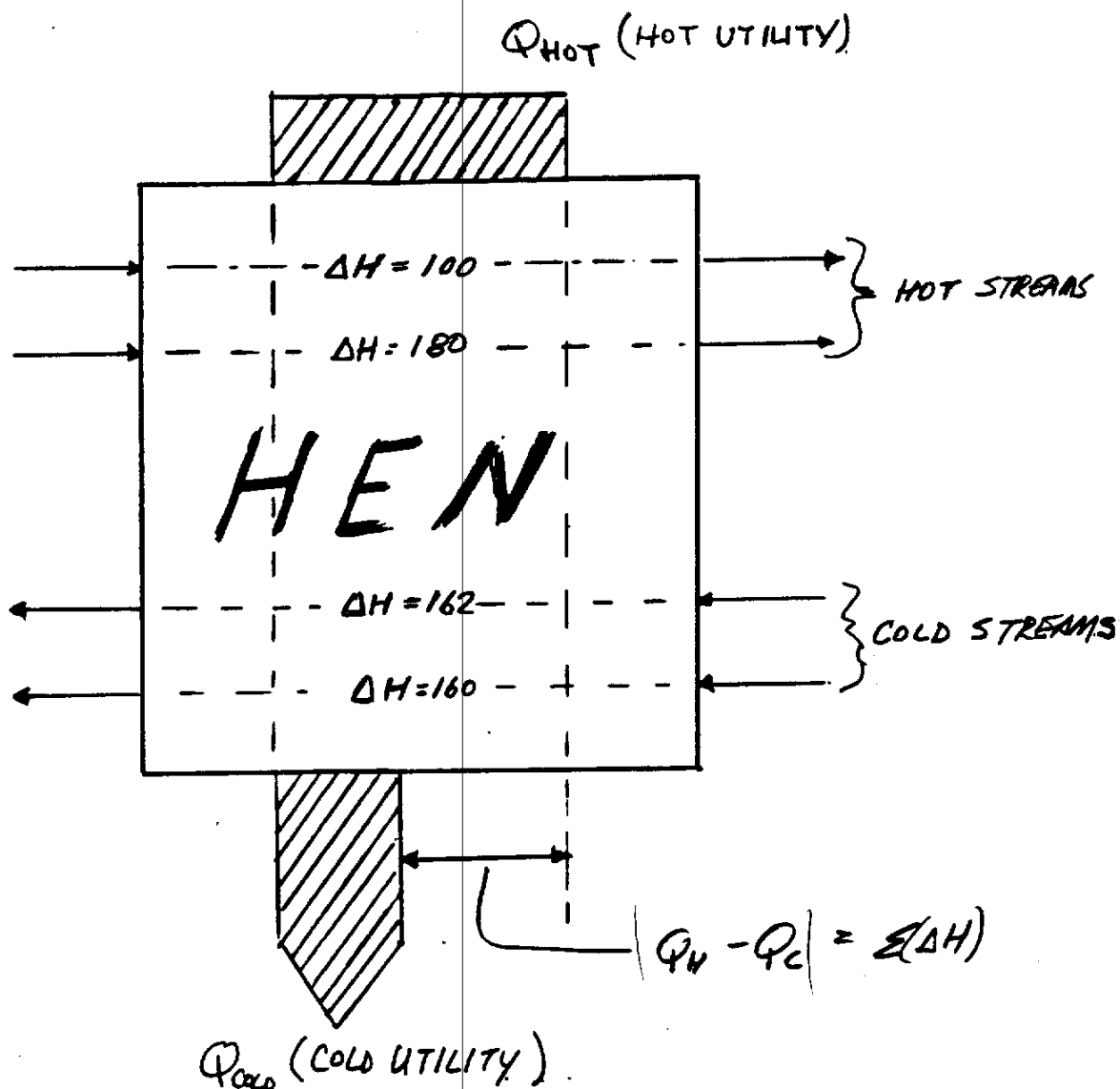
$$\text{STREAM [3]} \quad T_{S3}^C = 60 \quad T_{T3}^C = 100 \quad \Delta H = 160$$

$$CP_3 = \frac{160}{100 - 60} = 4.0$$

$$\text{STREAM [4]} \quad T_{S4}^C = 30 \quad T_{T4}^C = 120 \quad \Delta H = 162$$

$$CP_4 = \frac{162}{120 - 30} = 1.8$$

DEFINE $\Delta T_{MIN} = 10$ (APPROACH MIN)



CUSTOMER ENGINEERING DESIGN SEMINARSUBJECT MER ANALYSISCRUDE UNIT DESIGNSETTING ENERGY TARGETSPAGE NO. 11DATE NOV 1991

PROJECT

BY E N MILLERFIRST LAW CONSIDERATIONS
(CONSERVATION OF ENERGY)

ENTHALPY BALANCE (CLASS PROBLEM)

$$\Sigma \Delta H_c = 160 + 162 = 322 \quad Q_H \text{ REQUIRED FOR HEATING COLD STREAMS}$$

$$\Sigma \Delta H_H = 100 + 180 = 280 \quad Q_C \text{ REQUIRED FOR COOLING HOT STREAMS}$$

$$\text{EXCESS} = 42 \quad \text{NET HEAT REQUIRED FOR BALANCE}$$

$$Q_H - Q_C = \Sigma \Delta H = \text{CONSTANT}$$

NOTE

 Q_H INCREASES Q_C INCREASES Q_H DECREASES Q_C DECREASES

PODA SATISFACER
DE BALANCE
SE MONTARÁ
SIEMPRE
CONSTANTE

LOOK AT LIMITS

$$Q_H |_{\text{MIN}} = 42$$

$$Q_C |_{\text{MIN}} = 0$$

IS THIS SOLUTION FEASIBLE?

CON 2D LEV DE PERTURBACIONES.

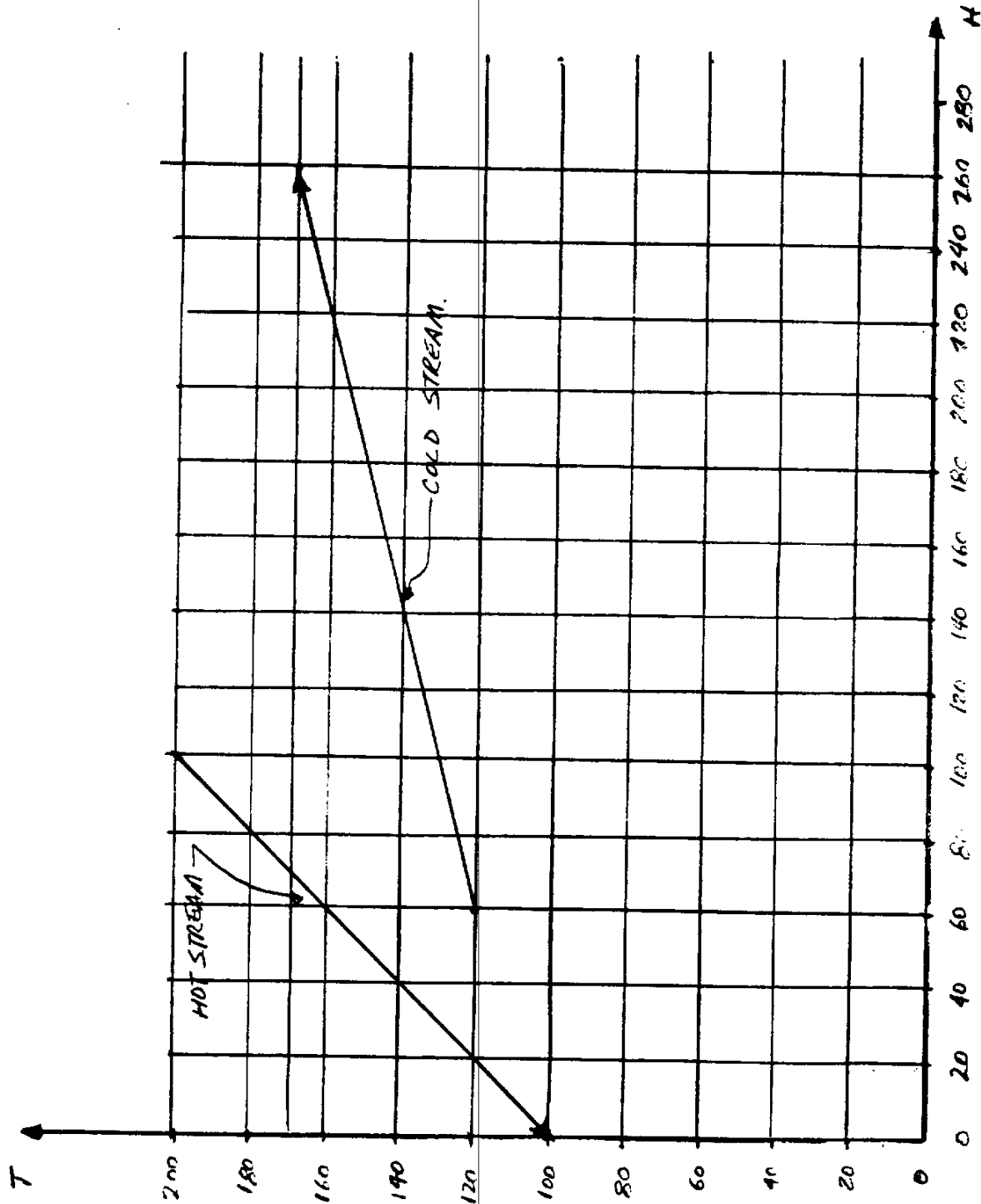
(FUERZA IMPULSORIA)

SE REQUIERE DE SUFICIENTE DIFERENCIA DE POTENCIAL

ΔT MIN EN ESTE COD.

CUSTOMER ENGINEERING DESIGN SEMINAR
 SUBJECT MER ANALYSIS
CRUDE UNIT DESIGN
TWO STREAM PROBLEM

PAGE NO. 12
 DATE NOV 1991
 PROJECT
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BASIC DATA

HOT STREAM

$$T_{\text{SUPPLY}}^H = 200 \quad T_{\text{TARGET}}^H = 100$$

$$Q = \Delta H = \text{ENTHALPY CHANGE} = 100$$

$$Q = m \dot{C}_p \Delta T = m \cdot \dot{C}_p (200 - 100) = 100$$

$$C_{P_{\text{HOT}}} = m \cdot C_p = 1.0$$

COLD STREAM

$$T_{\text{SUPPLY}}^C = 120 \quad T_{\text{TARGET}}^C = 170$$

$$Q = \Delta H = \text{ENTHALPY CHANGE} = 200$$

$$C_{P_{\text{COLD}}} = \frac{200}{170 - 120} = \underline{\underline{4.0}}$$

$$\Sigma \Delta H_c = 200 \quad \text{HEAT REQUIRED TO HEAT COLD STREAM}$$

$$\Sigma \Delta H_H = 100 \quad \text{COOLING REQUIRED TO COOL HOT STREAM}$$

$$\text{EXCESS} = \underline{100} \quad \text{NET HEATING REQUIRED}$$

$$\Rightarrow Q_H|_{\text{MIN}} = 100 \quad Q_C|_{\text{MIN}} = 0$$

POSSIBLE ??

 $\Delta H = 100$ $T = 200$ $T = 120$

Q = 100 CPAT
BTU/H

170 $T_{\text{S}} = 200 + 25$
115 $T_{\text{C}} = 145$

TRADITIONAL APPROACHHOT STREAM TEMP $200 \rightarrow 100$ $Q = 100$

HEAT COLD STREAM WITH HOT STREAM

$$Q = CP \cdot \Delta T$$

$$Q = 100$$

$$CP = 4.0$$

$$\Delta T = (T_T^C - T_S^C) = \frac{100}{4} = 25$$

$$T_T^C = 25 + 120 = 145$$

NOW LOOK AT OUR EXCHANGER TEMPERATURES

HOT	200	$\rightarrow$	100
COLD	145	$\leftarrow$	120
ΔT	55		-20

NOT FEASIBLE! WHY? 2ND LAW VIOLATION

WHAT IS THEORETICAL MINIMUM ENERGY REQUIREMENT? 2

$$\Delta H = Q = CP \Delta T$$

MER

i.e. MAXIMUM HEAT RECOVERY $\Rightarrow$ MER.

$$Q = 1 \times (200 - 120)$$

$$Q = 80. \Delta H_{MSA}$$

HOT	200	$\rightarrow$	120
COLD	140	$\leftarrow$	120
	60		0

$$Q = 80$$

$$\Delta T_c = \frac{80}{4} = 20$$

SUMMARIZE $Q_{HOT} - Q_{COLD} = 100 = \text{CONSTANT}$ HOT STREAM COOLING UTILITY MIN. $Q_{C|MIN} = (120 - 100)(1) = 20$ COLD STREAM HEATING UTILITY MIN $Q_{H|MIN} = 120$

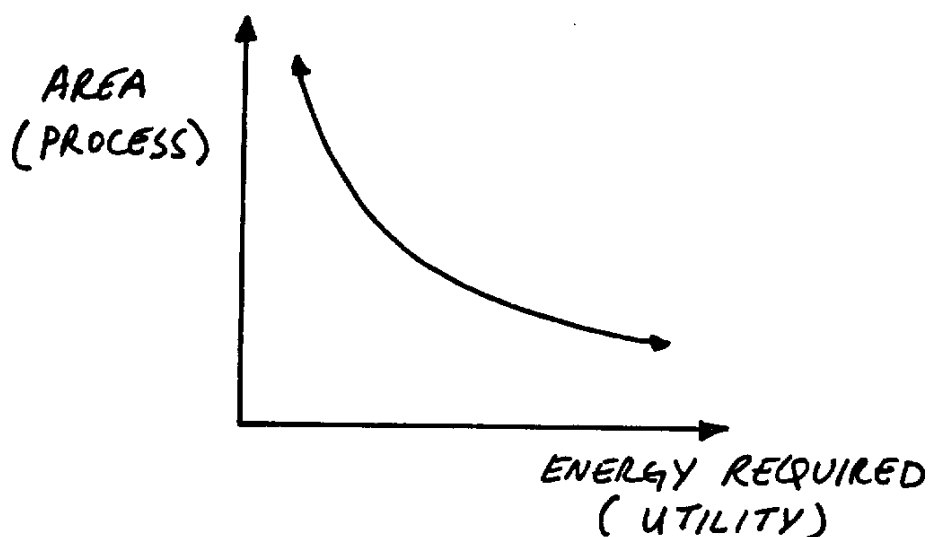
$$Q_{EXCHANGE|MAX} = 80$$

SOLUTION ECONOMICAL? WHY?

RECALL $Q = UA \Delta T_{LM}$

$$A = \frac{Q}{U \Delta T_{LM}}$$

$\Delta T_{LM} \rightarrow 0$ i.e. $\Delta T_{MIN} \rightarrow 0$ $A \rightarrow \infty$



REMEMBER $\Delta T_{MIN} \Rightarrow$ MINIMUM TEMPERATURE APPROACH

THIS IS OFTEN CHOSEN TO ACCOUNT FOR EFFECT OF CAPITAL COSTS.

SAY $\Delta T_{MIN} = 10$

i.e. $T_T^{HOT} = 130^\circ F$

$$T_S^{COLD} + \Delta T_{MIN} = T_T^{HOT}$$

$$\text{HEAT RECOVERED FROM HOT STREAM} = (1)(200 - 130) = 70$$

$$Q_{\text{COLD}} |_{\text{UTILITY}} = (130 - 100)(1) = 30$$

FIRST LAW

$$Q_{\text{HOT}} - Q_{\text{COLD}} = \underline{100} = \text{CONSTANT}$$

$$Q_{\text{HOT}} |_{\text{UTILITY}} = 130$$

TO SUMMARIZE
FOR

$$\Delta T_{\text{MIN}} = 10$$

$$Q_{\text{COLD}} |_{\text{MIN}} = 30$$

$$Q_{\text{HOT}} |_{\text{MIN}} = 130$$

$$Q_{\text{RECOVERY}} |_{\text{MAX}} = 70$$

LIKEWISE

$$\Delta T_{\text{MIN}} = 20$$

$$Q_{\text{COLD}} = 40$$

$$Q_{\text{HOT}} = 140$$

$$Q_{\text{RECOVERY}} = 60$$

$$\text{DIF} = 100$$

AT OTHER EXTREME

$$Q_{\text{COLD}} |_{\text{MAX}} = 100$$

$$Q_{\text{HOT}} |_{\text{MAX}} = 200$$

$$Q_{\text{RECOVERY}} = 0$$

SUMMARIZE

THERE IS A CORRELATION BETWEEN

ΔT_{MIN} AND $Q_{HOT}|_{MIN}$, $Q_{COLD}|_{MIN}$ REQ

AS ΔT_{MIN} VARIES

$Q_{HOT}|_{MIN}$ AND $Q_{COLD}|_{MIN}$ VARIES.

- MORE IN
- MORE OUT.

I.E.

$$Q_H|_{MIN} + X \text{ IMPLIES } Q_C|_{MIN} + X$$

WHAT DOES ONE DO IF THERE ARE MULTIPLE
HOT AND COLD STREAMS?

EXTEND CONCEPT

"COMPOSITE CURVES"

EXTEND TWO STREAM PROBLEM ANALYSIS

CONSTRUCT COMPOSITE CURVES

USE CLASS PROBLEM

FOR HOT COMPOSITE (SEE ATTACHED GRAPH)

STREAM 1 AND STREAM 2 USED.

$$\sum \Delta H_H = 100 + 180 = 280$$

REMEMBER

$$\Delta H = m \cdot C_p \cdot \Delta T = C_P \cdot \Delta T$$

FOR TEMPERATURE INTERVAL ~~180~~ TO 130

WE HAVE ONLY STREAM 1 $\Delta H_1 = (1)(180 - 130)$
 $= 50$

FOR TEMPERATURE INTERVAL 130 TO 80

STREAM 1 $\Delta H_1 = (1)(130 - 80) = 50$

STREAM 2 $\Delta H_2 = (2)(130 - 80) = 100$

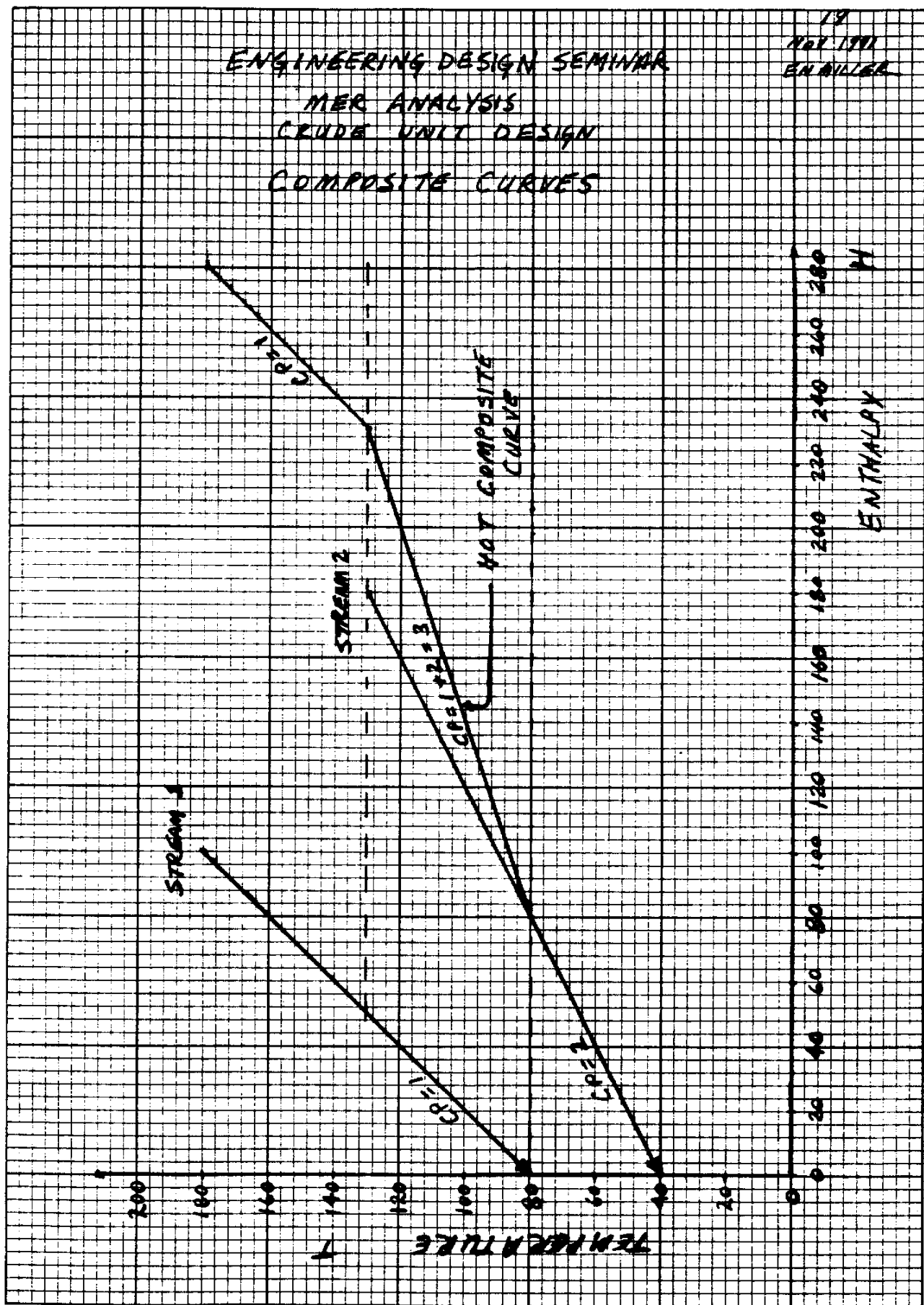
TOTAL $\Delta H = (1+2)(130-80) = 150$

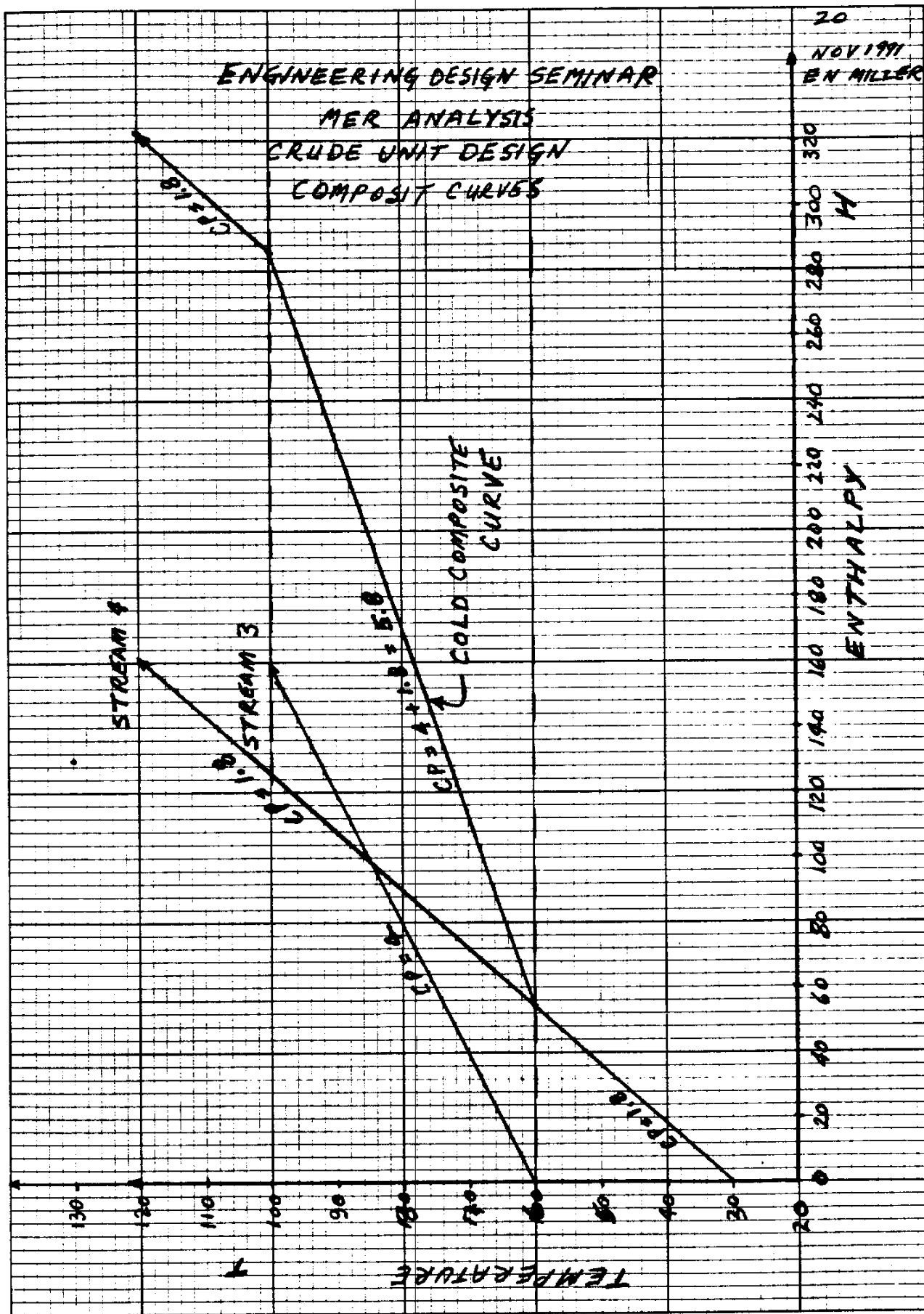
FOR TEMPERATURE INTERVAL 80 TO 40

WE HAVE ONLY STREAM 2 $\Delta H = (2)(80 - 40)$
 $= 80$

ENGINEERING DESIGN SEMINAR MER ANALYSIS CRUDE UNIT DESIGN COMPOSITE CURVES

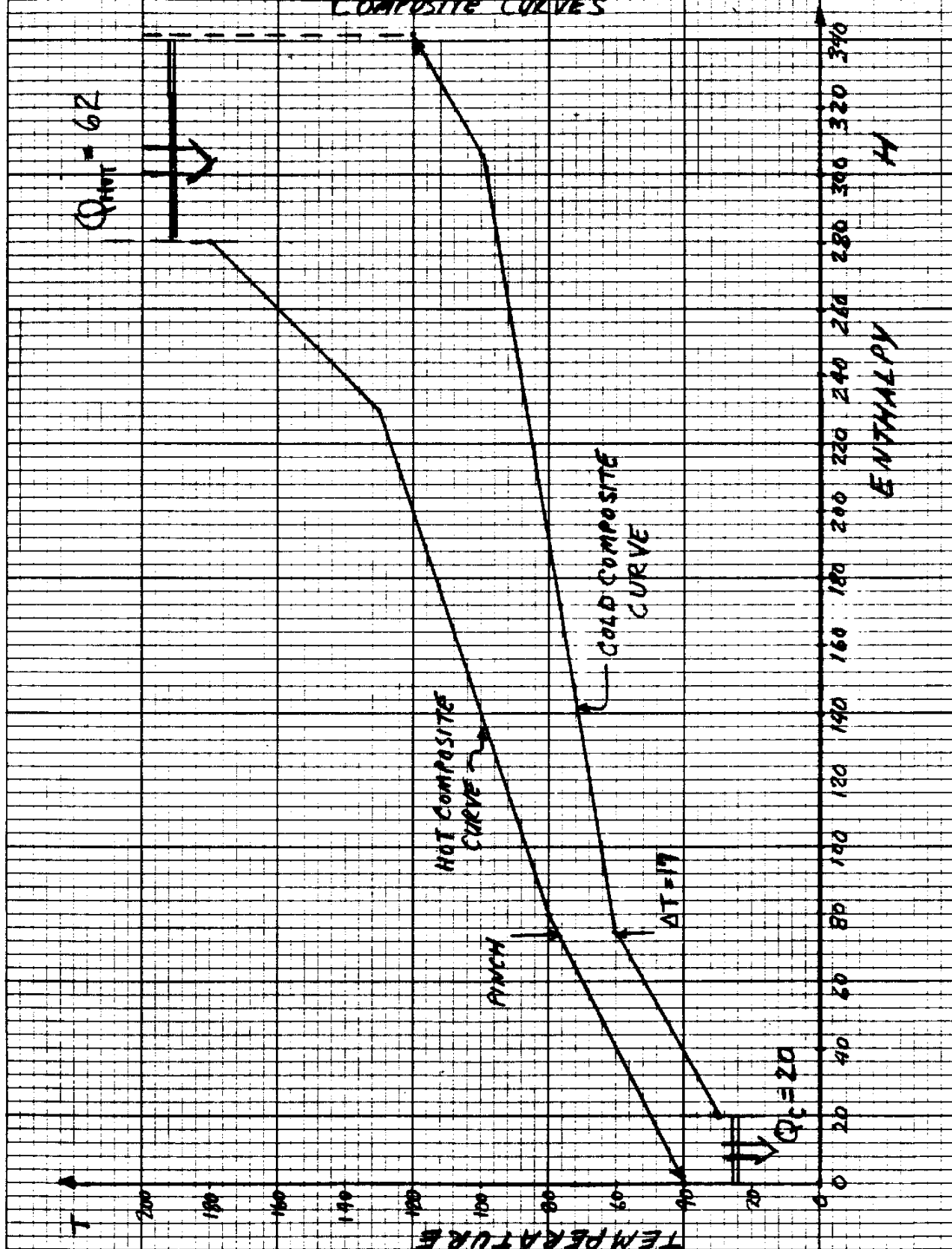
19
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ENGINEERING DESIGN SEMINAR HEAT ANALYSIS CRUDE UNIT DESIGN COMPOSITE CURVES

21
MAY 1991
KEN MILLER



WHAT HAVE WE DONE?

STARTING WITH

STREAM DATA

ΔT_{MIN}

CONSTRUCT

COMPOSITE CURVES FOR PROCESS

THIS ALLOWS

SETTING ENERGY TARGETS

MINIMUM HOT UTILITY TARGET

MINIMUM COLD UTILITY TARGET

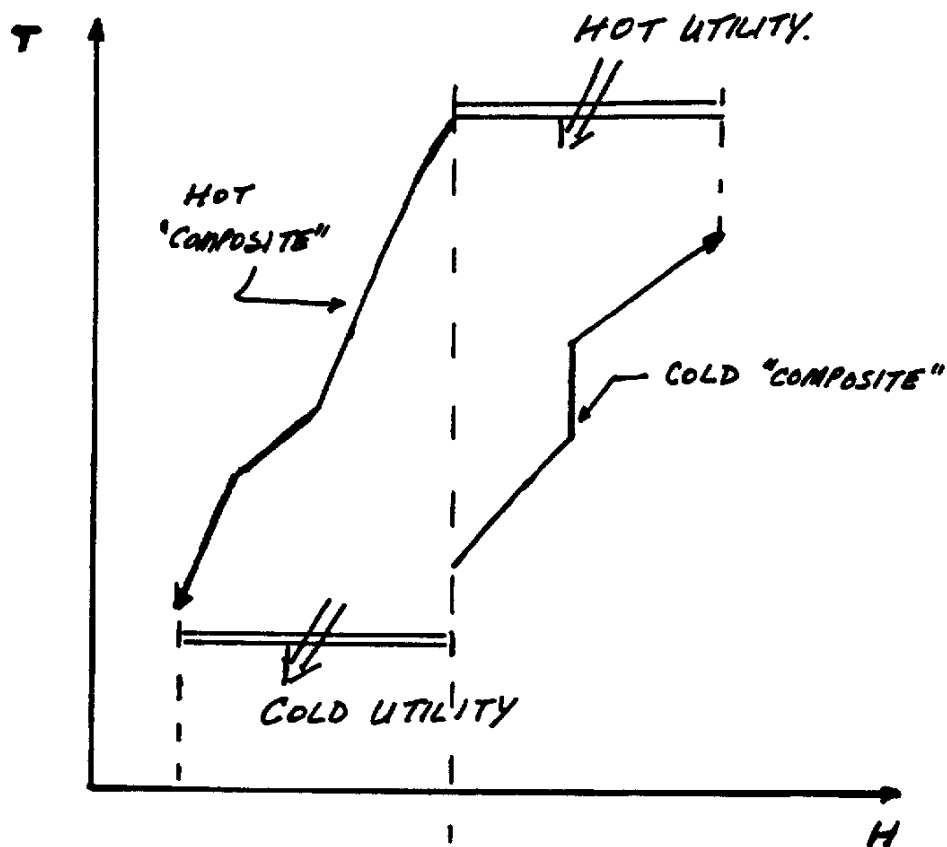
MAXIMUM HEAT RECOVERY TARGET

PRIOR TO H. E. N. DESIGN

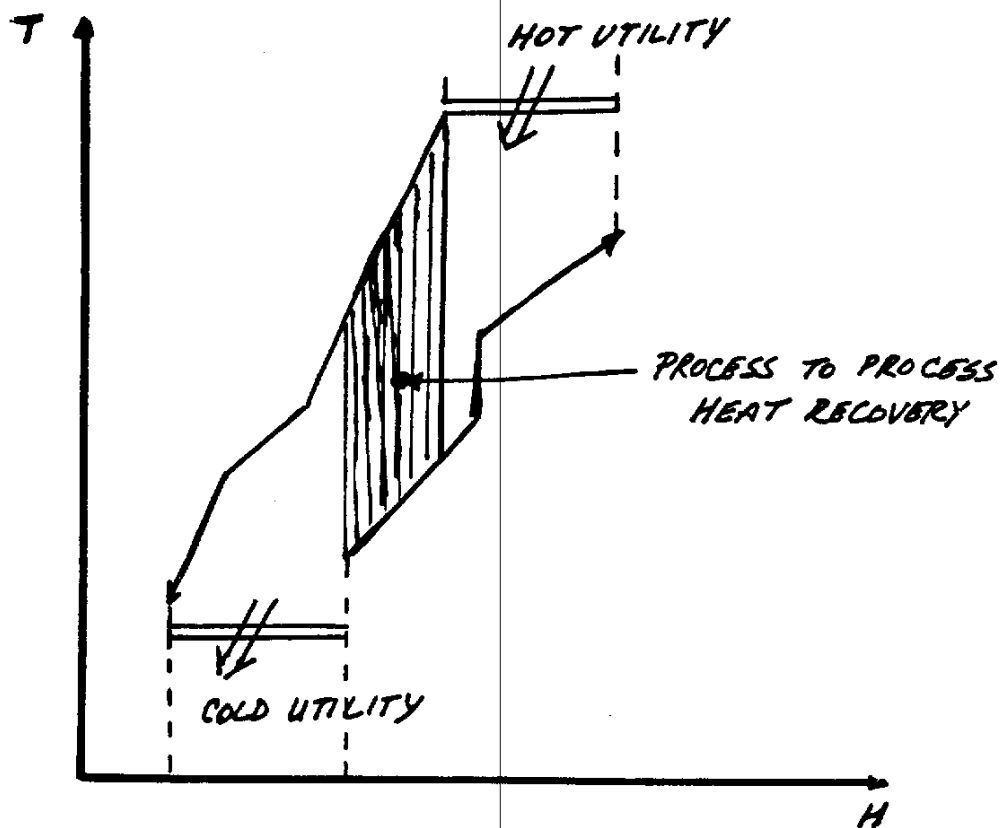
COMPOSITE CURVES ARE USEFUL FOR VISUALIZATION

PURPOSES BUT CUMBERSOME TO USE IN DESIGN.

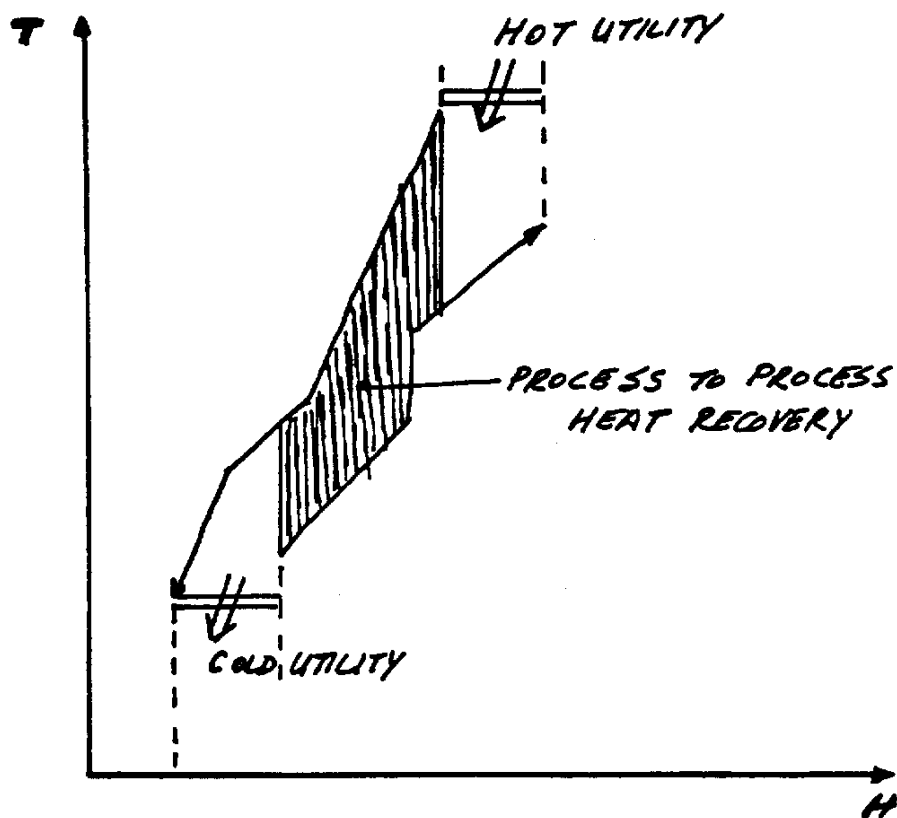
TOO BIG



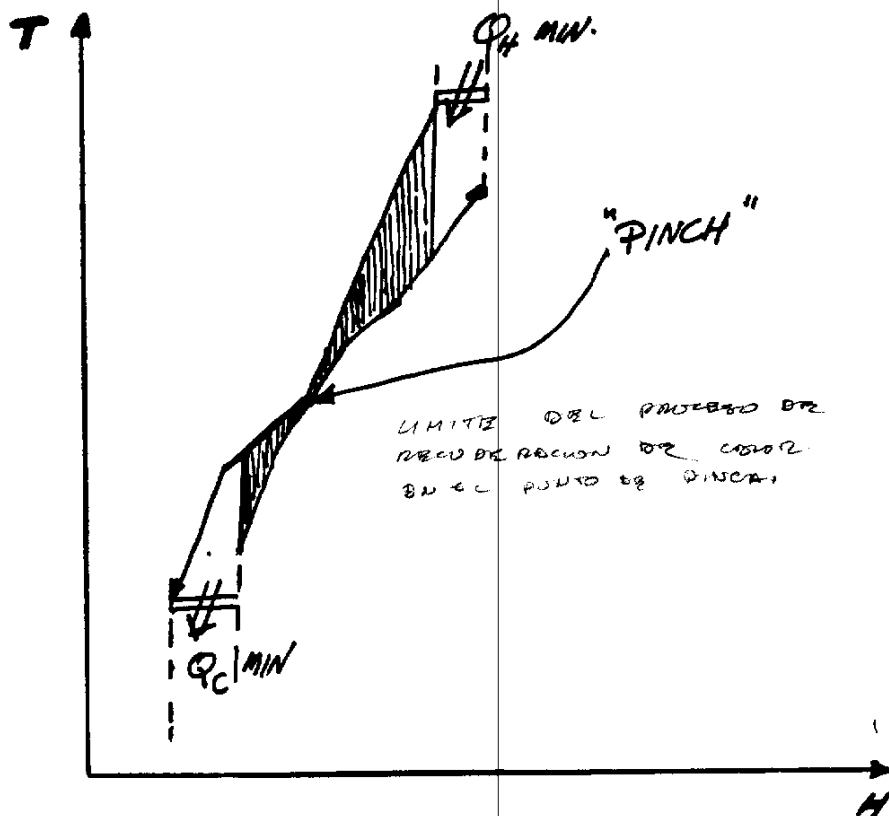
NO INTEGRATION.



PROCESS HEAT RECOVERY REDUCES
UTILITY REQUIREMENT



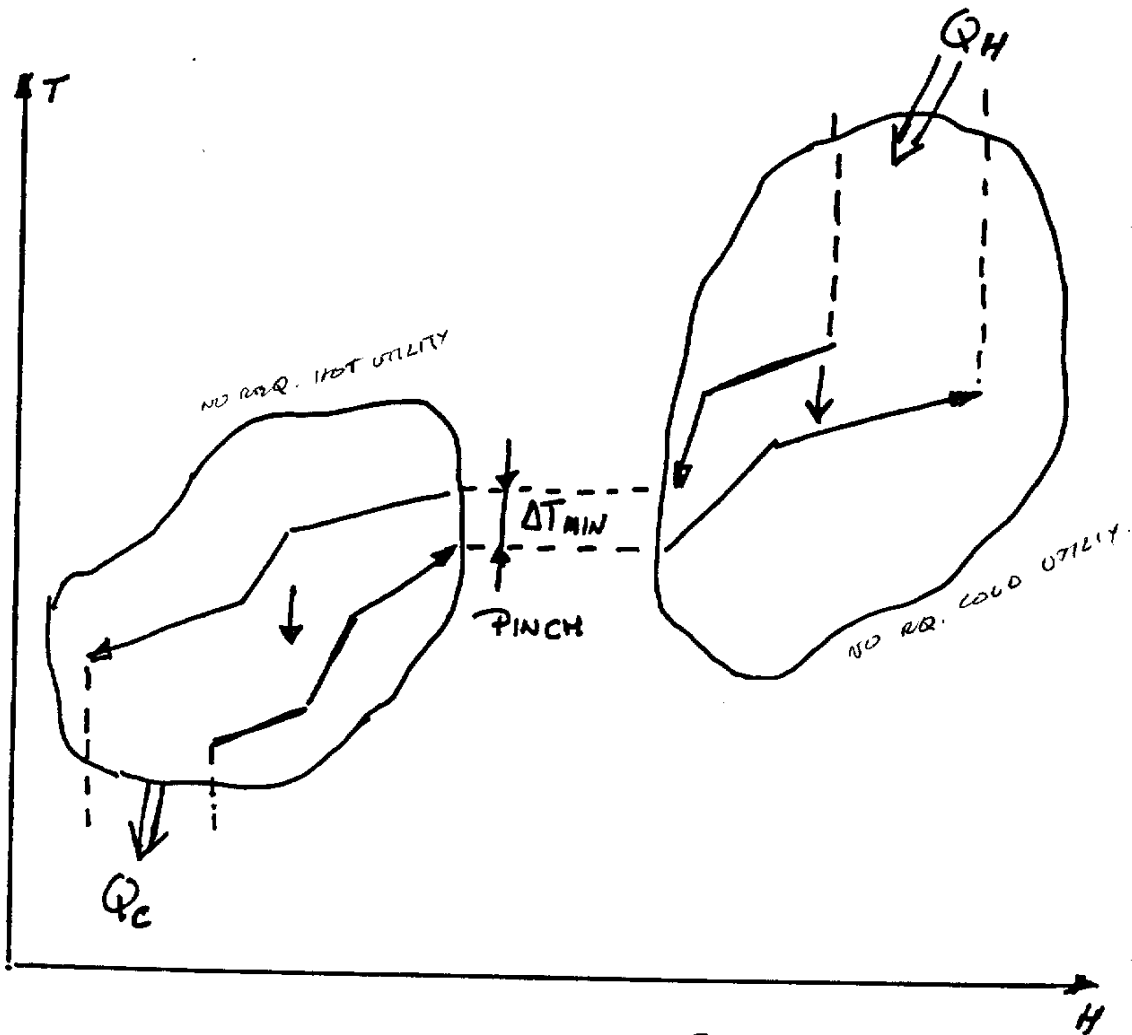
MORE
HEAT RECOVERY



THE "PINCH" LIMITS

PROCESS HEAT RECOVERY

AT MIN ES LA SITUACION ECONOMICA EN EL PUNTO
 DE PINCH (ESTRECHAMIENTO).



DIVIDE PROCESS AT PINCH.

CUSTOMER ENGINEERING DESIGN SEMINAR.SUBJECT MER ANALYSISCRUDE UNIT DESIGNCOMPOSITE CURVES.PAGE NO. 283vDATE NOV 1991

PROJECT

BY EN MILLERABOVE PINCH

HEAT REQUIREMENT OF COLD STREAMS ARE PROVIDED PARTIALLY BY HOT STREAMS AND PARTIALLY BY THE HOT UTILITY. THIS PART OF PROCESS IS A NET "HEAT SINK". IT IS IN ENTHALPY BALANCE WITH Q_H - HEAT FROM HOT UTILITY. NO COOLING IS REQUIRED.

BELOW PINCH

HEAT REQUIRED BY COLD STREAMS IS COMPLETELY PROVIDED BY THE HOT STREAMS. THE REMAINING HEAT IN THE HOT STREAMS IS REMOVED USING A COLD UTILITY. THIS PART OF THE PROCESS IS A NET "HEAT SOURCE". IT IS IN ENTHALPY BALANCE WITH Q_c - COOLING FROM EXTERNAL UTILITY. NO HEATING IS REQUIRED

IT IS POSSIBLE TO TRANSFER HEAT.

FROM ABOVE PINCH TO BELOW PINCH.

DRIVING FORCES POSITIVE.

$$\Delta T > \Delta T_{\min}.$$

IT IS NOT POSSIBLE TO TRANSFER HEAT.

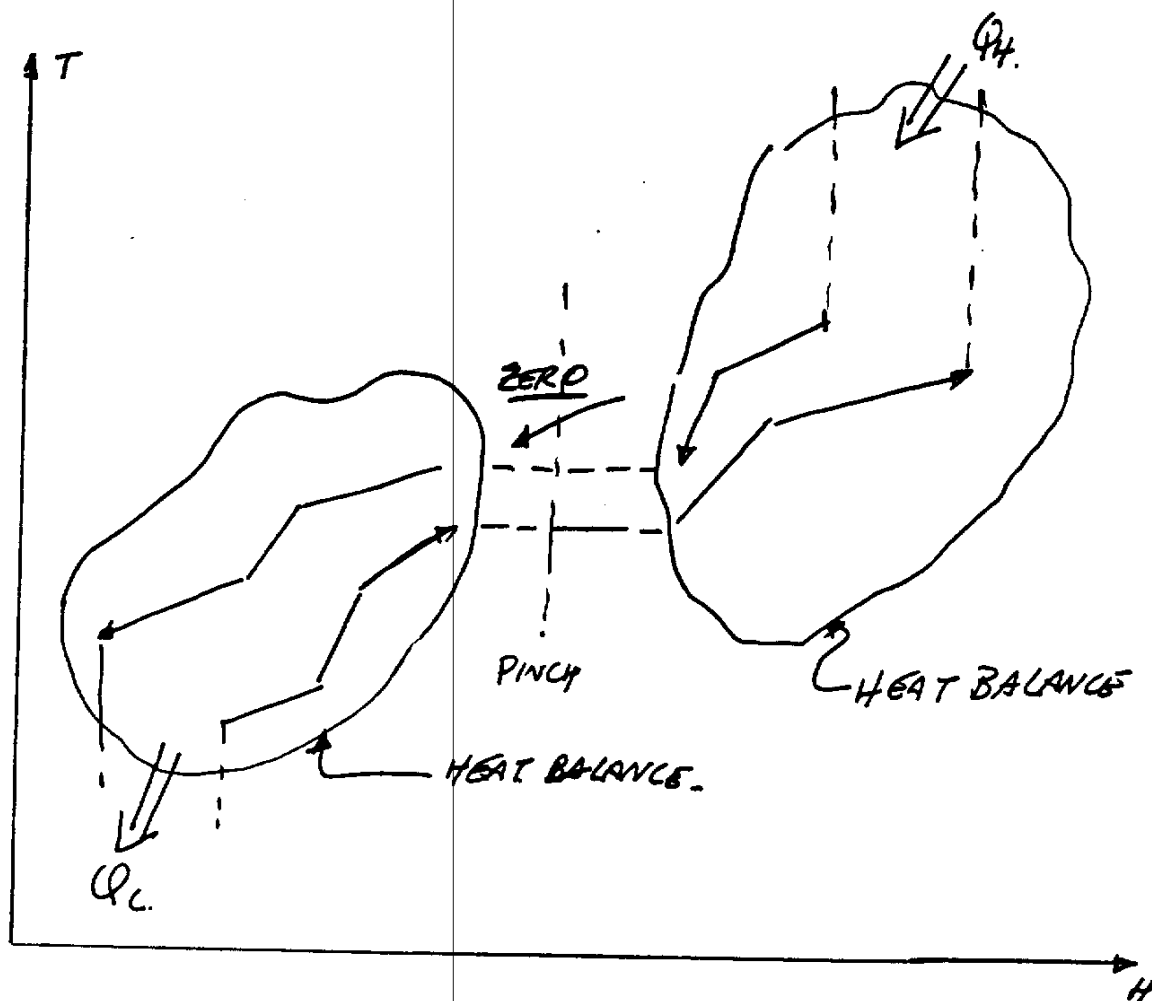
FROM BELOW PINCH TO ABOVE PINCH.

$$\Delta T < \Delta T_{\min}$$

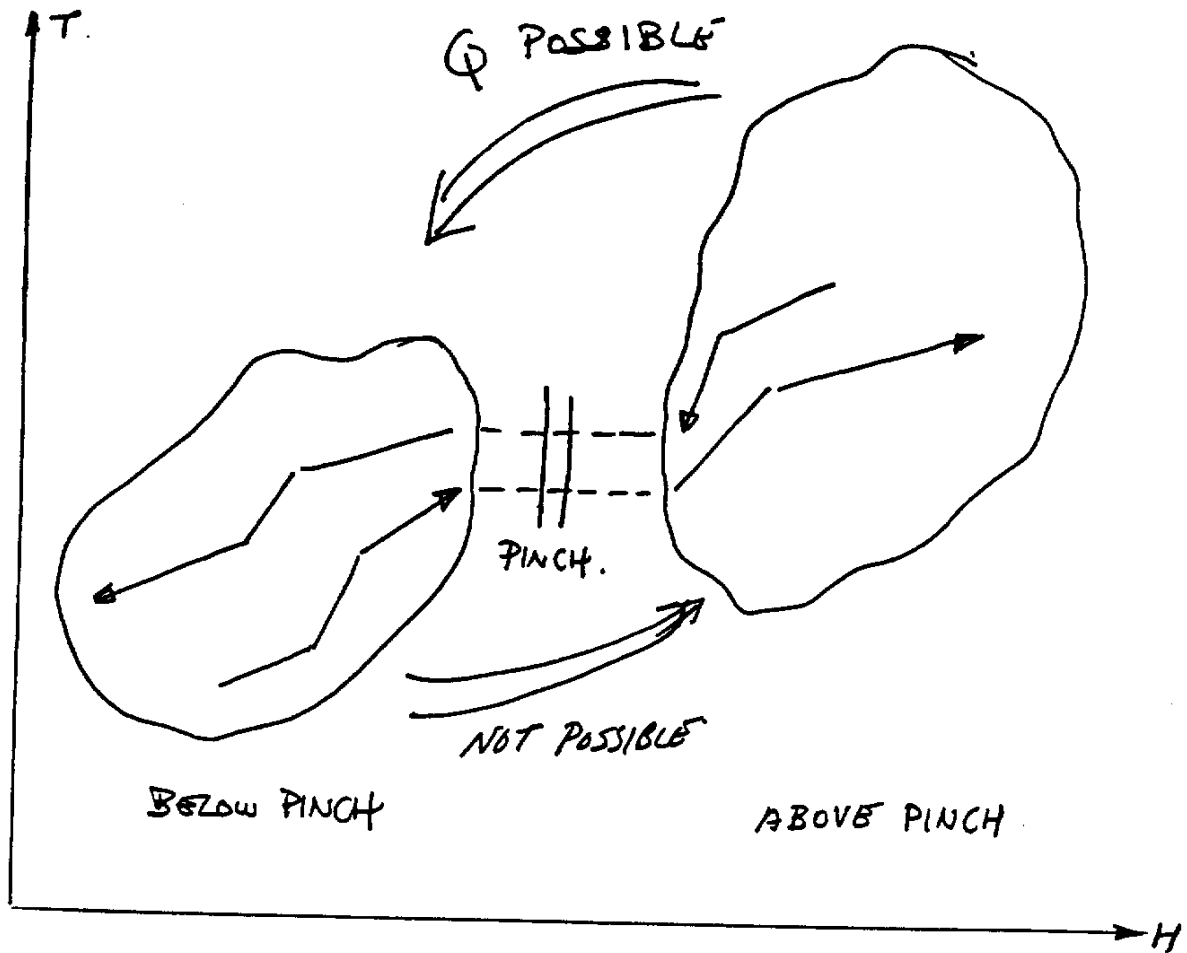
OR

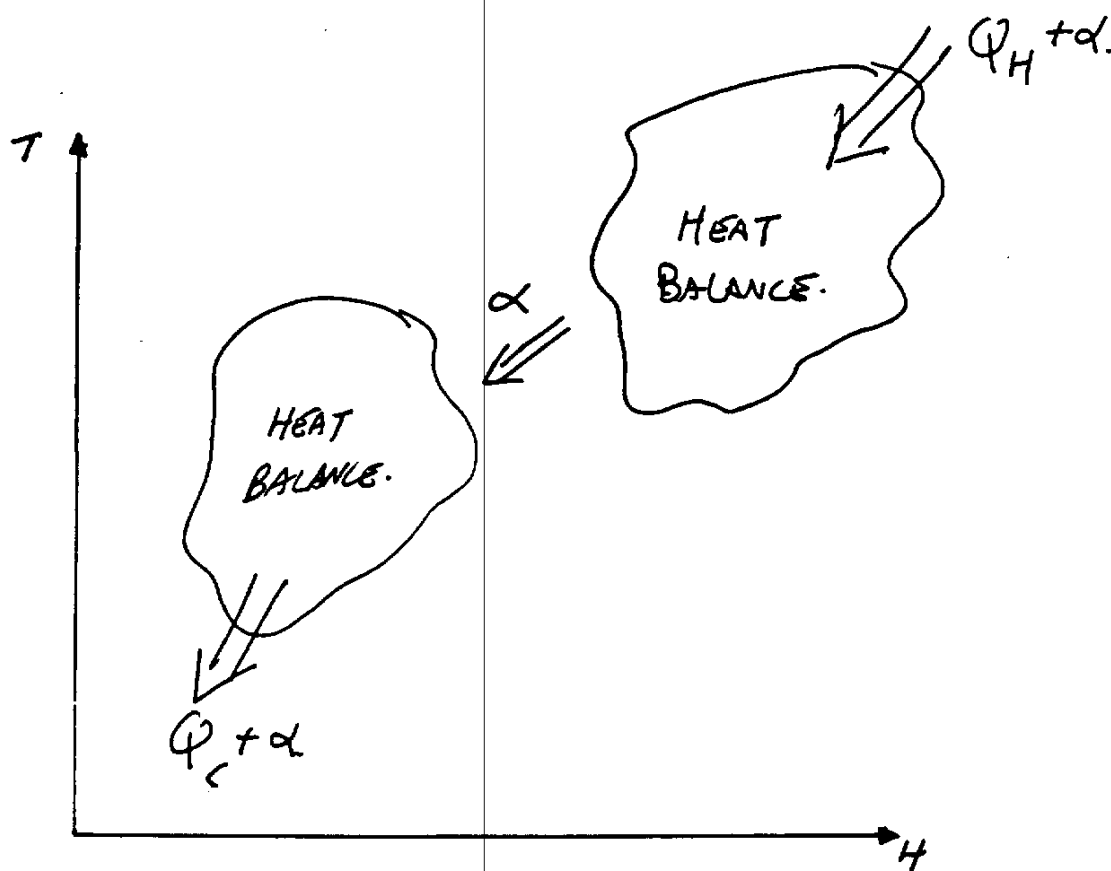
$$\Delta T < 0 \text{ (NEGATIVE).}$$

VIOLATION OF SECOND LAW.



WHEN DIVIDING PROCESS AT PINCH INTO TWO INDEPENDENT REGIONS WHICH ARE IN HEAT BALANCE IF THERE IS NO HEAT TRANSFER ACROSS PINCH.





IF WE TRANSFER α UNITS OF HEAT FROM ABOVE PINCH TO BELOW THE PINCH, TO REESTABLISH THE HEAT BALANCE ABOVE THE PINCH THE HOT UTILITY NEEDS TO BE INCREASED BY α UNITS TO $Q_h + \alpha$. BELOW THE PINCH WE ARE ADDING α UNITS TO 'HEAT SOURCE'. TO RETURN THE HEAT BALANCE THE COLD UTILITY NEEDS TO BE INCREASED BY α UNITS TO $Q_c + \alpha$.

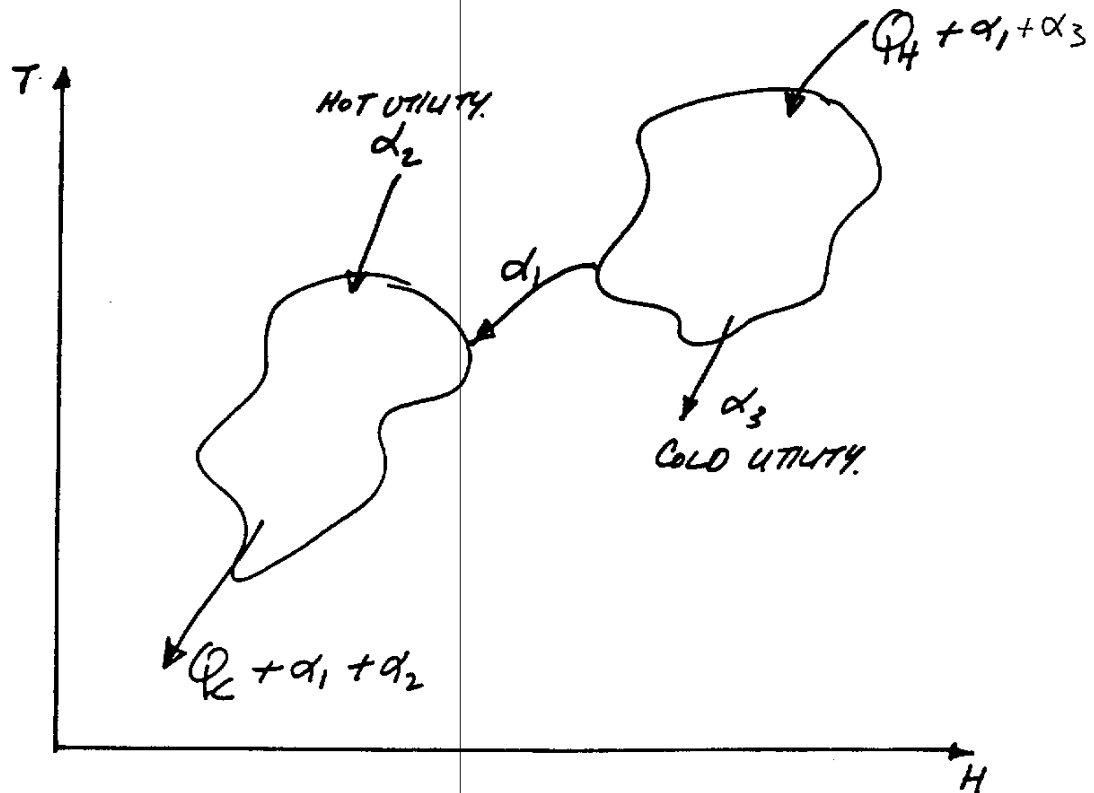
THE PINCH EQUATION.

$$A = T + XP.$$

ACTUAL UTILITY.

TARGET

CROSS PINCH.
HEAT TRANSFER.



TYPES OF CROSS PINCH HEAT FLOW.

1. HOT STREAMS ABOVE PINCH TO HEAT COLD STREAMS BELOW PINCH.
2. HOT UTILITY BEING USED TO HEAT COLD STREAMS BELOW PINCH.
3. COLD UTILITY BEING USED TO COOL HOT STREAMS ABOVE THE PINCH.

CONCLUSIONS:

SIGNIFICANCE OF PINCH.

- DON'T TRANSFER HEAT ACROSS.
- DON'T USE HOT UTILITIES BELOW.
- DON'T USE COLD UTILITIES ABOVE

AS MENTIONED PREVIOUSLY. WORKING WITH COMPOSITE CURVES UNWIELDY. LOOK AT ANALYTIC REPRESENTATION - CALLED PROBLEM TABLE ALGORITHM BY LINNHOFF

① ADJUST. TEMPERATURES FOR ΔT_{MIN}

$$i.e., T_i^{H'} = T_i^H - \frac{1}{2} \Delta T_{MIN}$$

$$T_i^{L'} = T_i^L + \frac{1}{2} \Delta T_{MIN}$$

COLD STREAM. SUPPLY $\frac{1}{2}$ TARGET TEMPERATURES INCREASED BY $\Delta T_{MIN}/2$

HOT STREAM SUPPLY $\frac{1}{2}$ TARGET TEMPERATURES DECREASED BY $\Delta T_{MIN}/2$.

FOR SAMPLE PROBLEM. $\Delta T_{MIN} = 10$.

REVIEW.

STREAM No.	TYPE.	T_s	T_f	ΔH	CP
STREAM ①	HOT	180	80	100	1
STREAM ②	HOT	130	40	180	2
STREAM ③	COLD	60	100	160	4
STREAM ④	COLD	30	120	162	1.8

$$\Delta T_{MIN} = 10$$

	T_s	T_T		T_s'	T_T'
1. HOT	180	80	$180 - 5 =$	175	$80 - 10 =$ 75
2. HOT	130	40	$170 - 10/2 =$	125	35
3. COLD	60	100	$60 + 10/2 =$	65	105
4. COLD	30	120	$30 + 10/2 =$	35	125

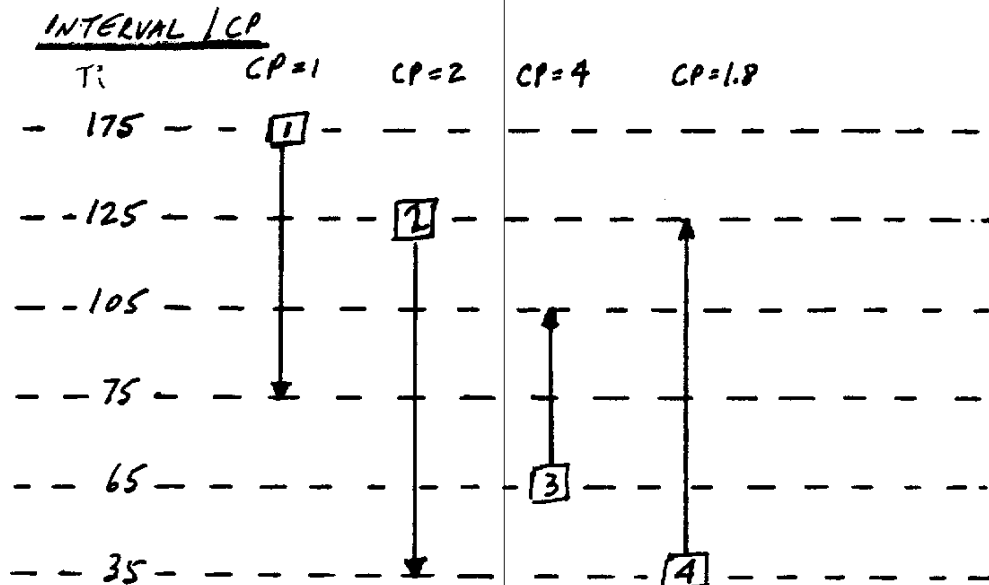
2. SET UP TEMPERATURE INTERVALS.

STREAM.	TYPE	$\frac{T_s'}{T_T'}$	REMOVE DUPLICATES	SET UP INTERVALS
1.	HOT	175 75	→ 175 ✓ → 75 ✓	--- 175 --- --- 125 ---
2.	HOT	125 35	→ 125 ✓ → 35 ✓	--- 105 ---
3.	COLD	65 105	→ 65 ✓ → 105 ✓	--- 75 --- --- 65 ---
4.	COLD	35 125	→ DUPLICATE → DUPLICATE	--- 35 ---

DESCENDING ORDER
OF MAGNITUDE:

CUSTOMER ENGINEERING DESIGN SEMINAR.
 SUBJECT MER ANALYSIS.
CLUDE UNIT DESIGN.
PROBLEM TABLE.

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 BY EN MILLER

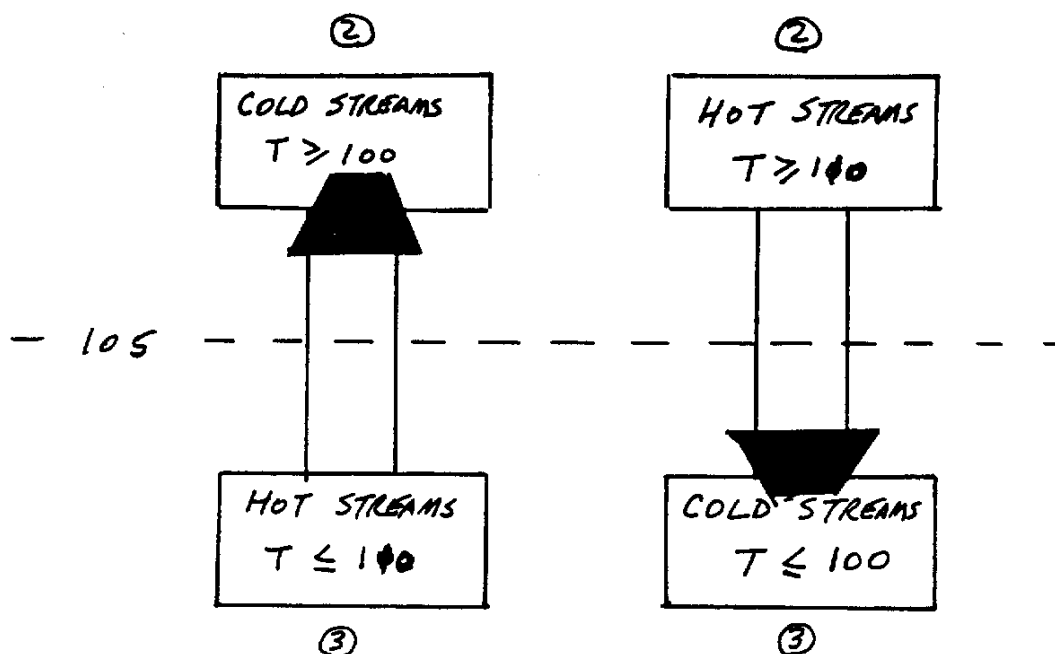


3. SET UP INTERVAL HEAT BALANCE

$$\Delta H_i = (T_i - T_{i+1}) (\sum CP_{\text{COLD}} - \sum CP_{\text{HOT}})$$

T_i INTERVAL	$T_i - T_{i+1}$	$\sum CP_{\text{COLD}} - \sum CP_{\text{HOT}}$	ΔH_i	SURPLUS OR DEFICIT
- 175 -				
①	50	0 - 1 = -1	-50	SURPLUS
- 125 -				
②	20	1.8 - (2+1) = -1.2	-24	SURPLUS
- 105 -				
③	30	(1.8+4) - (2+1) = +2.8	+84	DEFICIT
- 75 -				
④	10	(1.8+4) - 2 = +3.8	+38	DEFICIT
- 65 -				
⑤	30	1.8 - 2 = -0.2	-6	SURPLUS
- 35 -				

LOOK AT INTERVALS ② AND ③



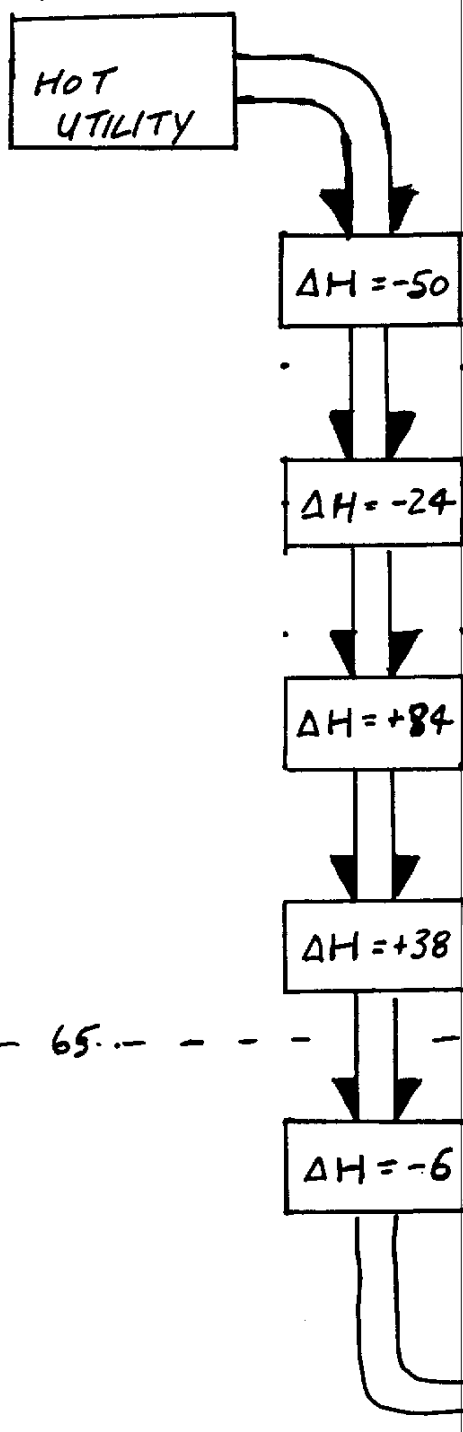
NOT POSSIBLE

ALWAYS POSSIBLE

4, IT IS POSSIBLE TO SET UP A HEAT
CASCADE.

HEAT CASCADE

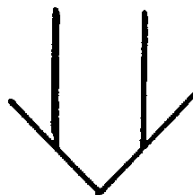
$$\Delta T_{MIN} = 10$$



TRIAL 1	TRIAL 2
0	<u>48</u> = $\phi_H _{MIN}$
50	98
74	122
10	
-10	38
<u>-48</u> - - - 0 - PINCH	
-42	6 = $\phi_c _{MIN}$

REVIEW PROBLEM TABLE ALGORITHM.

- ① ADJUST FOR ΔT_{MIN}
- ② SET UP TEMPERATURE INTERVALS
- ③ CALCULATE INTERNAL HEAT BALANCE
- ④ CASCADE FOR POSITIVE HEAT FLOW



PINCH, Q_H / MIN , Q_C / MIN
FOR

ΔT_{MIN} SPECIFIED

NOW WE CAN DRAW COMPOSITE CURVES WITHOUT TRIAL
& ERROR

FOR SAMPLE PROBLEM.

$$\Delta T_{MIN} = 10, \quad Q_H / MIN = 48, \quad Q_C / MIN = 6$$

PINCH HOT = 70°

PINCH COLD = 60°



ENGINEERING DEPARTMENT

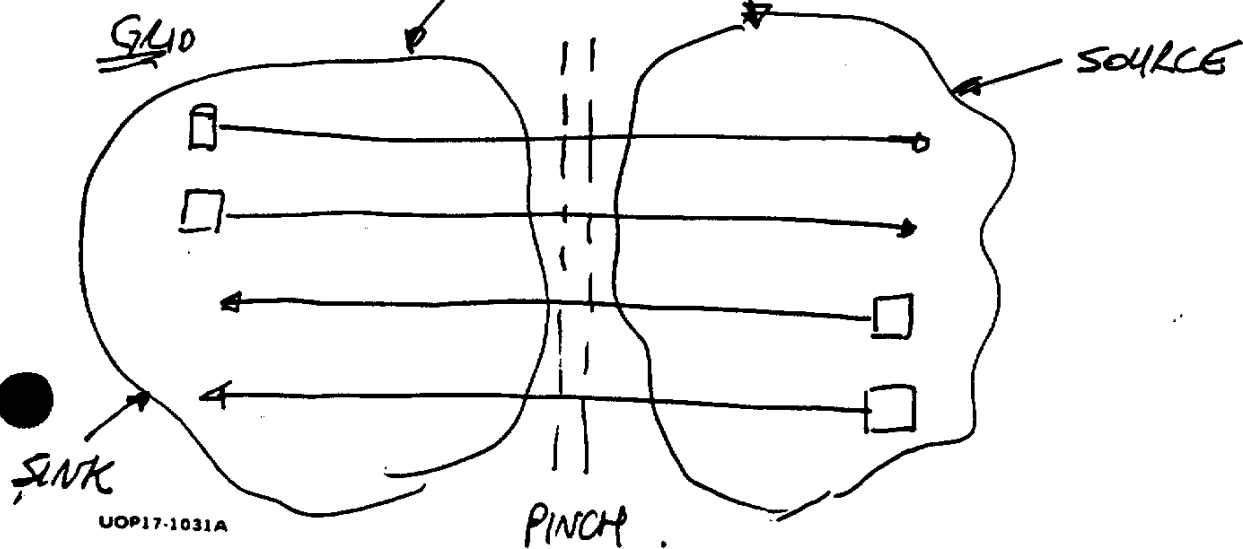
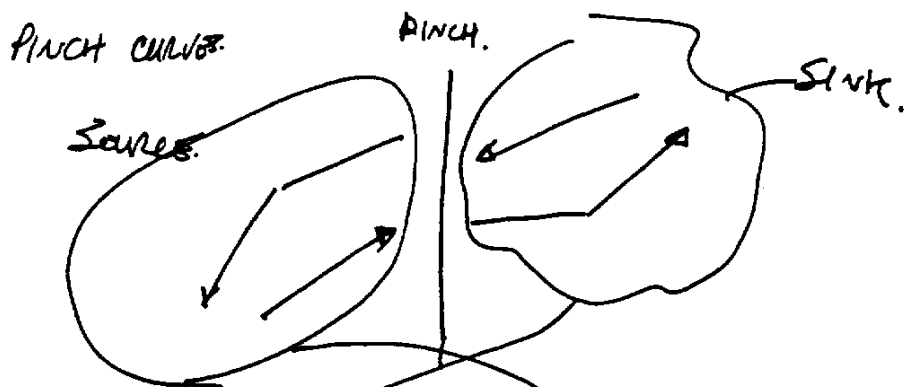
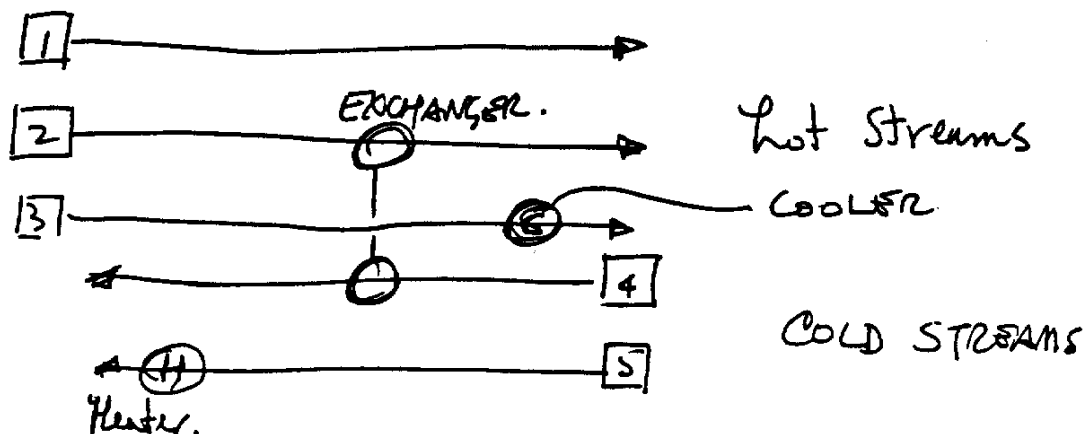
CUSTOMER ENGINEERING DESIGN SEMINAR
SUBJECT MER ANALYSIS
CLUDE UNIT DESIGN
HEN

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BY EN MILLER

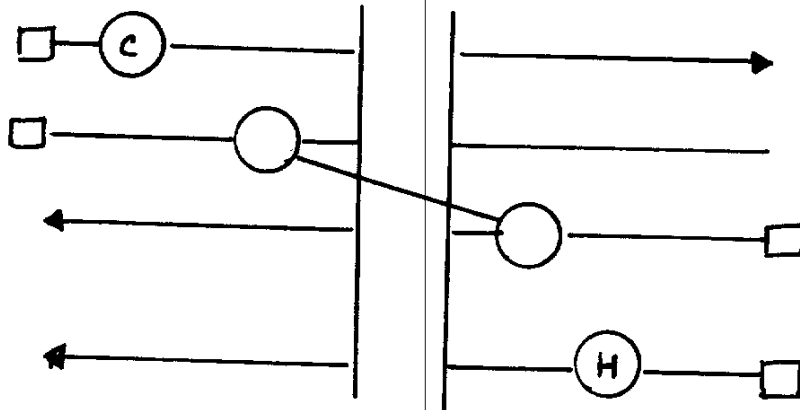
NOW TO

H. E. N.

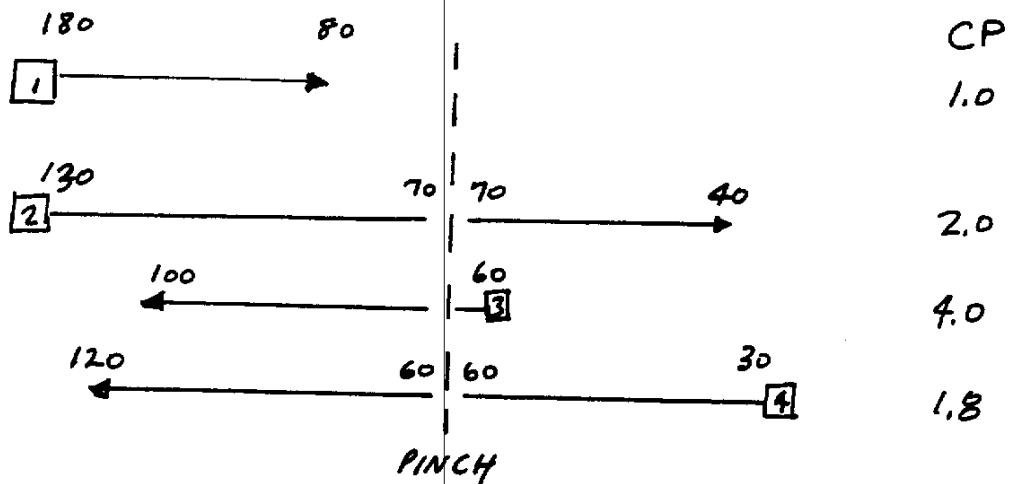
HEAT EXCHANGER NETWORK REPRESENTATION, USE OF GRID DIAGRAM.



VIOLATIONS OF PINCH ARE EASY TO SEE WITH GRID DIAGRAM



FOR SAMPLE PROBLEM DRAW NETWORK GRID

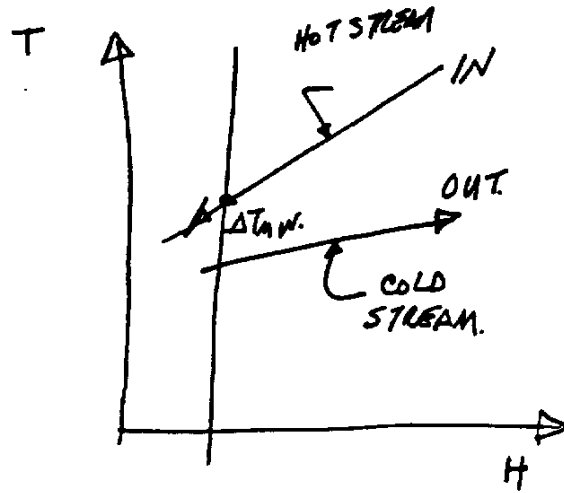


NOTE:

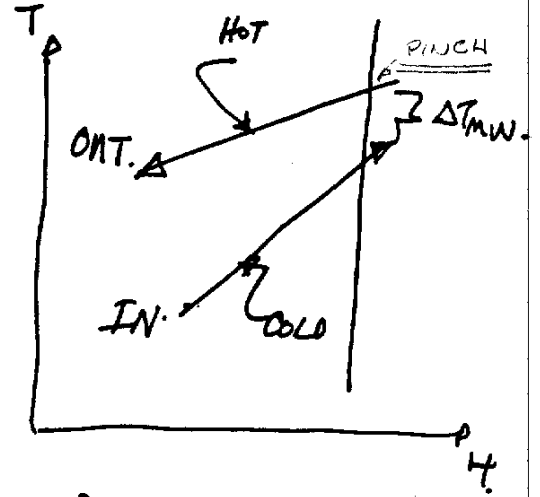
PROBLEM IS MOST CONSTRAINED AT PINCH.

ANY HOT STREAM CAN BE MATCHED WITH ANY COLD STREAM.

FEASIBLE MATCHES AT PINCH.



ABOVE
PINCH.



BELOW
PINCH.

$$C_{P_{OUT}} \geq C_{P_{IN}}$$

ie. ABOVE PINCH $C_{P_{COLD}} \geq C_{P_{HOT}}$

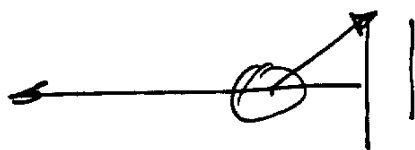
BELOW PINCH $C_{P_{HOT}} \geq C_{P_{COLD}}$

OBSERVATIONS.

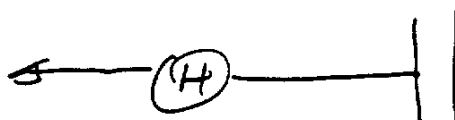
STREAMS OUT ABOVE PINCH - HEAT SINK
(COLD STREAMS)



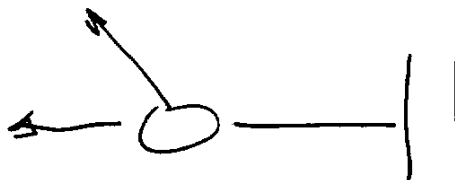
NOT POSSIBLE. STREAMS
REQUIRED HEATING.



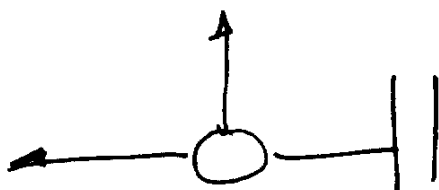
MATCH WITH HOT STREAM
BELOW PINCH. NOT
POSSIBLE. VIOLATES ΔT_{MIN}
ON SECOND LAW



POSSIBLE.



POSSIBLE. EXCHANGE
WITH HOT STREAM AWAY
FROM PINCH



POSSIBLE. EXCHANGE
WITH HOT STREAM AT
PINCH. NEED SMALLER
STREAM IN $C_{OUT} \geq C_{IN}$.

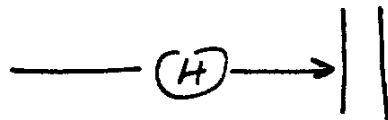
R.E.D

QSD $\Rightarrow$ EASY

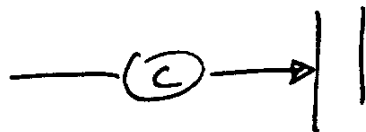
THREE OPTIONS.

OBSERVATIONS.

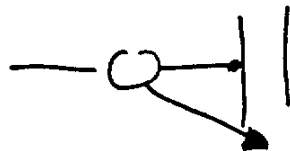
STREAMS IN ABOVE PINCH — HEAT SOURCE.
HOT STREAMS



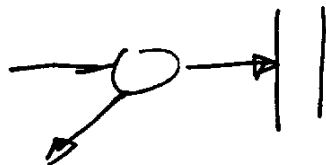
NOT POSSIBLE. STREAM
REQUIRES COOLING.



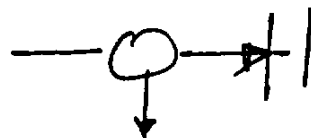
POSSIBLE. BUT DO NOT WANT
TO USE COOLER ABOVE PINCH.
INCREASES UTILITIES.



POSSIBLE. BUT DO NOT WANT
TO EXCHANGE HEAT ACROSS
PINCH. (MATCH OF COLD
STREAM BELOW PINCH).



NOT POSSIBLE. MATCH WITH
COLD STREAM ABOVE AND
AWAY FROM PINCH. VIOLATES
 ΔT_{MIN} OR 2ND LAW

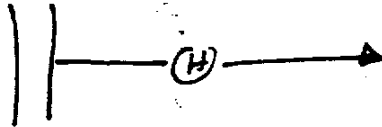


POSSIBLE. BUT NEED LARGER
COLD STREAM. MATCH AT
PINCH $CP_{OUT} \geq CP_{IN}$.

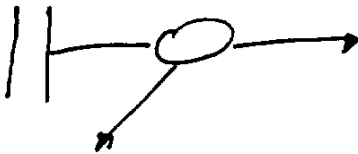
Q.E.D. $\Rightarrow$ DIFFICULT. ONE CHOICE

OBSERVATION.

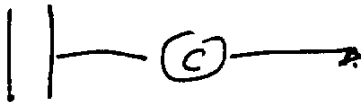
STREAMS OUT BELOW PINCH - HEAT SOURCE.
(HOT STREAMS)



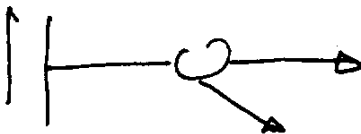
NOT POSSIBLE. HOT STREAM
REQUIRES COOLING.



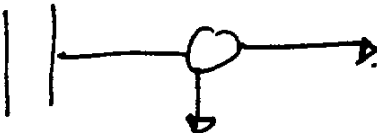
NOT POSSIBLE. MATCH WITH COLD
STREAM. ABOVE PINCH
VIOLATES ΔT_{MIN} OR 2ND LAW.



POSSIBLE



POSSIBLE. MATCH WITH COLD
STREAM. AWAY FROM PINCH

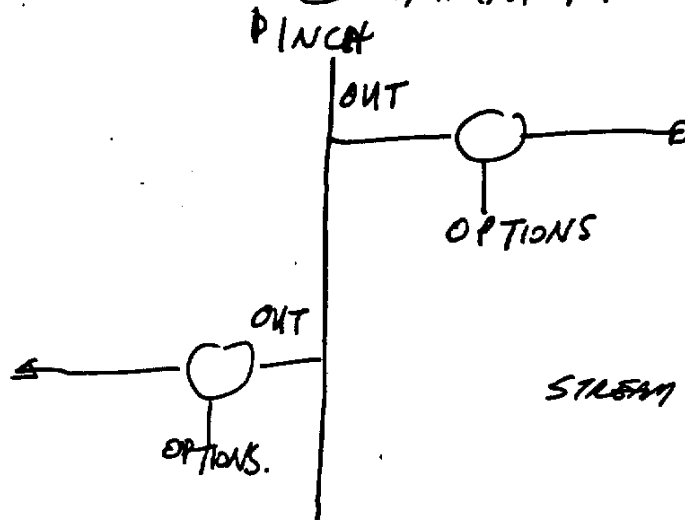


POSSIBLE. MATCH WITH COLD
STREAM AT PINCH. COLD
STREAM NEEDS SMALLER CP
 $CP_{OUT} > CP_{IN}$.

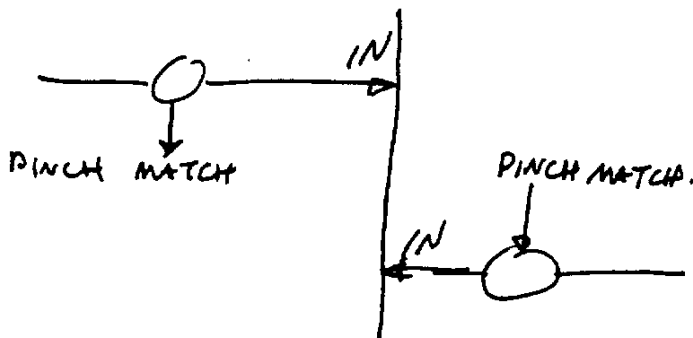
Q.E.D. $\Rightarrow$ EASY

THREE CHOICES.

SUMMARY.



STREAM MATCHES ARE EASY.



$$C_{POUT} \geq C_{PIN}$$

$$C_{COLD} \geq C_{HOT}$$

$$C_{POUT} \geq C_{PIN}$$

$$C_{HOT} \geq C_{COLD}$$

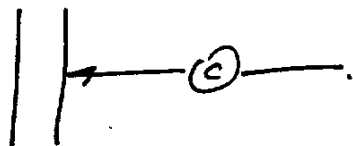
SO.

① START AT PINCH

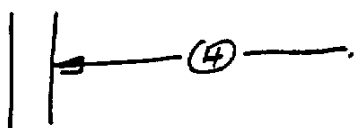
② START WITH STREAMS IN

OBSERVATION:

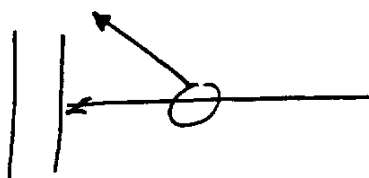
STREAMS IN BELOW PINCH - HEAT SINK.
(COLD STREAMS)



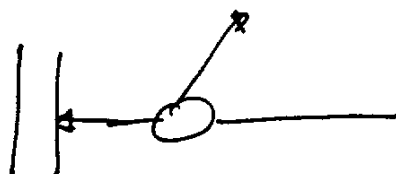
NOT POSSIBLE, COLD STREAM.
NEEDS HEATING.



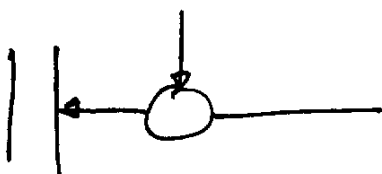
POSSIBLE, BUT INCREASES
UTILITY BELOW PINCH.



POSSIBLE, BUT INCREASES
UTILITY. CROSS PINCH EXCHANGE.



NOT POSSIBLE. MATCH WITH
HOT STREAM AWAY FROM
PINCH. VIOLATES ΔT_{MIN} OR
2ND LAW.



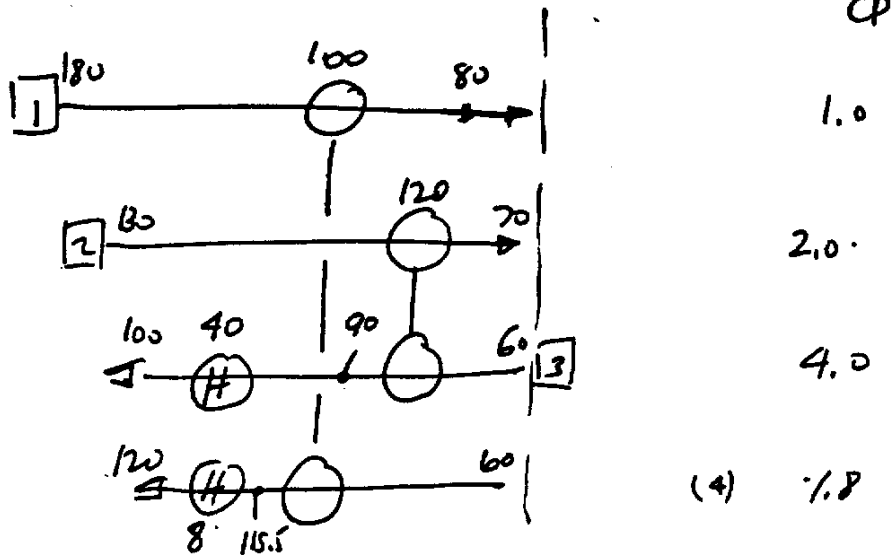
POSSIBLE BUT NEED A LARGER
HOT STREAM. BUT
 $CP_{HOT} \geq CP_{COLD}$.

Q.E.D. $\Rightarrow$ DIFFICULT. ONE MATCH.

SAMPLE PROBLEM.

DIVIDE PROBLEM INTO TWO PARTS

④ HOT END DESIGN.



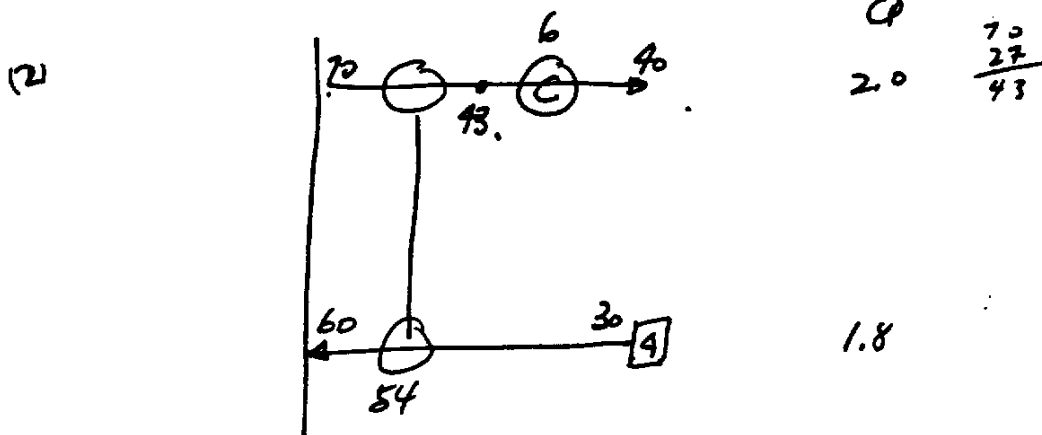
① MATCH (2) → (3) AT PINCH.

② MATCH (1) → (4)

ALL IN STREAMS MATCH.

IN STREAMS MATCH
CPOUT > CPIN.

③ COLD END DESIGN.

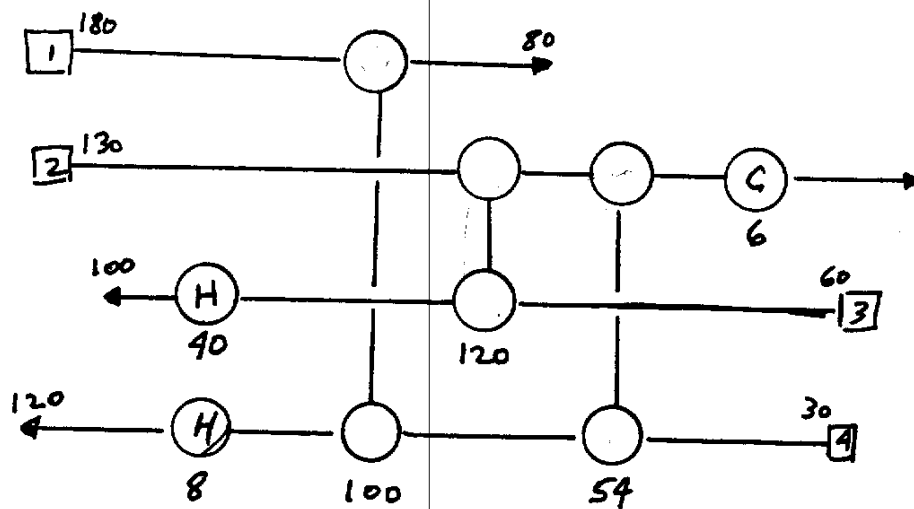
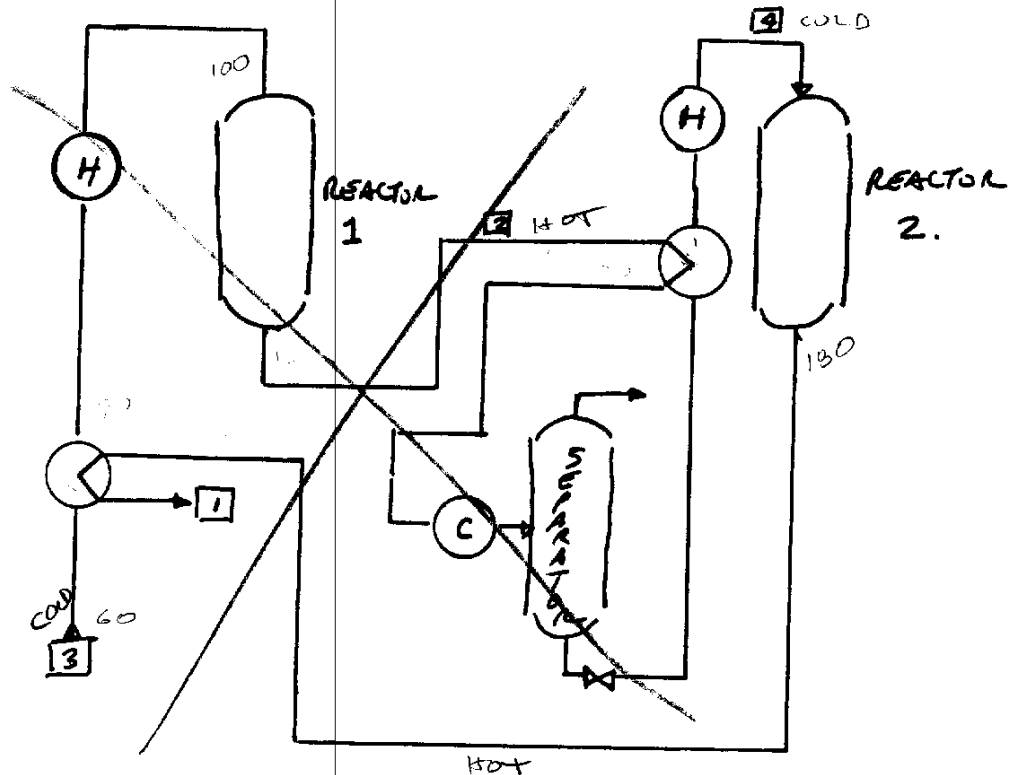


MATCH IN STREAMS CPOUT > CPIN.

CUSTOMER ENGINEERING DESIGN SEMINAR.
SUBJECT MER ANALYSIS
CRUDE UNIT DESIGN.
HEN FOR MER

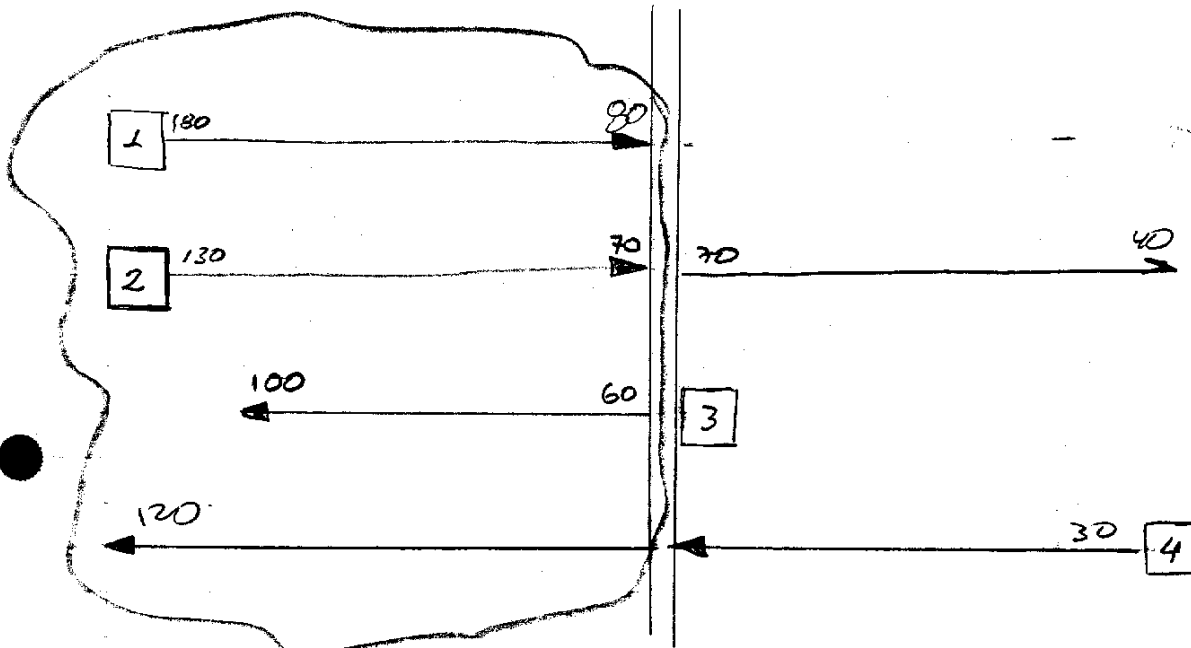
PAGE NO. 52
DATE NOV 1991
PROJECT _____
BY EN MILLER.

FINAL DESIGN



NUMBER OF UNITS = 6

1



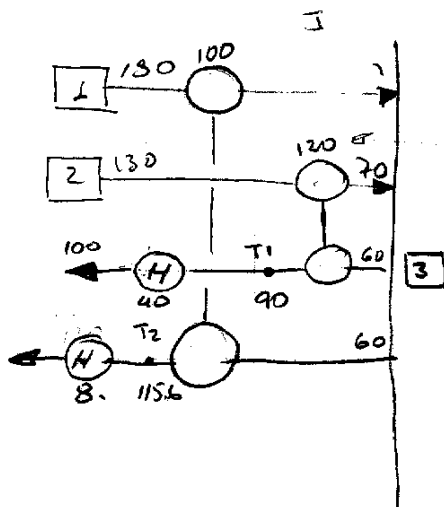
ABOVE
PINCH

$$CP = 2.0$$

$$CP = 2.0$$

$$CP = 4.0$$

$$CP = 1.3$$



①

$$CP_{OUT} \geq CP_{IN}$$

[2] WITH [4] ? NOT POSSIBLE

STREAM 2 & 3

$$\Delta H = (130 - 70)(2) = 120$$

$$\Delta H = (100 - 60)(4) = 160$$

$$\Delta T = 120/4 = 30$$

$$T_1 = 60 + 30$$

②

STREAM 1 & 4

$$\Delta H = (180 - 80) = 100$$

$$\Delta T = 100/1.8 = 55.6$$

$$T_2 = 80 + 55.6 = 135.6$$

③

STREAM 3

$$\Delta H_{TOTAL} = (100 - 60) \cdot 4.0 = 160$$

$$\Delta H_{EXCH} = 120$$

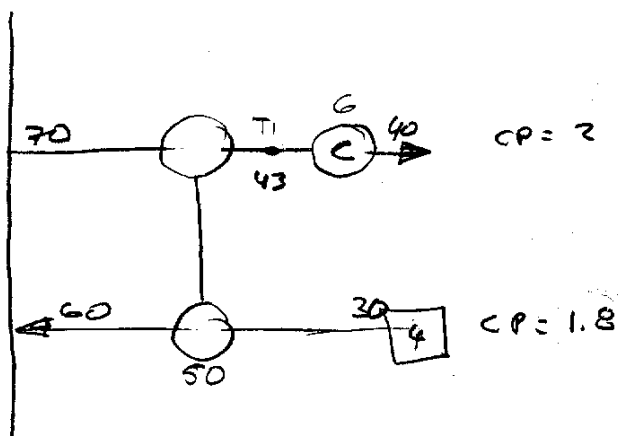
$$\Delta H_{HEATER} = 40$$

$$\Delta H_{TOTAL} = (120 - 60) \cdot 1.3$$

$$\Delta H_{TOTAL} = 108$$

$$\Delta H_{EXCH} = 100$$

$$\Delta H_{HEATER} = 8$$



$$\Delta H_2 = 1.8(60 - 30) = 54$$

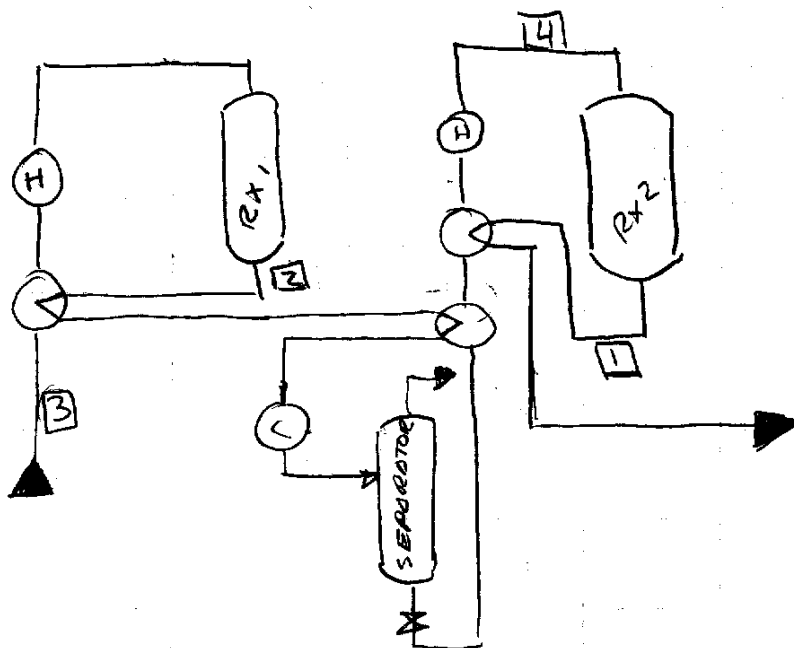
$$\Delta T = 54/2 = 27$$

$$T = 70 - 27 = 43$$

$$\Delta H_{TOT} = 2 \times (70 - 40) = 60$$

$$Q_{BAC} = 54$$

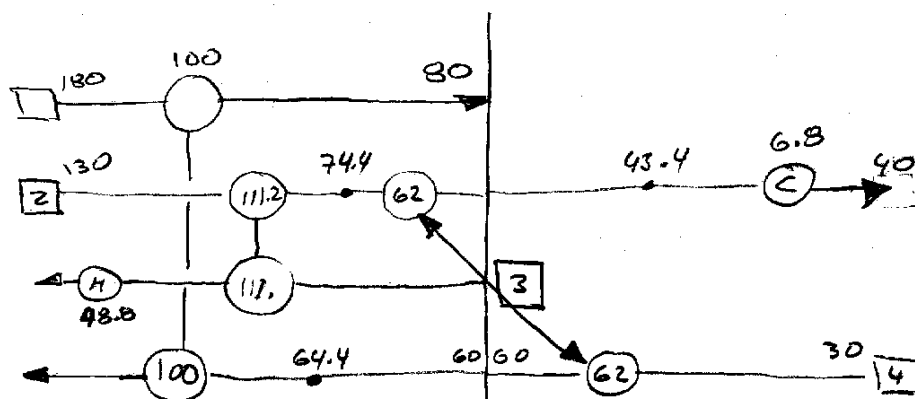
$$Q_{COOLBR} = 6$$



FINAL DESIGN

3

ELIMINANDO UN INTERCAMBIADOR (HEATER 3) STREAM



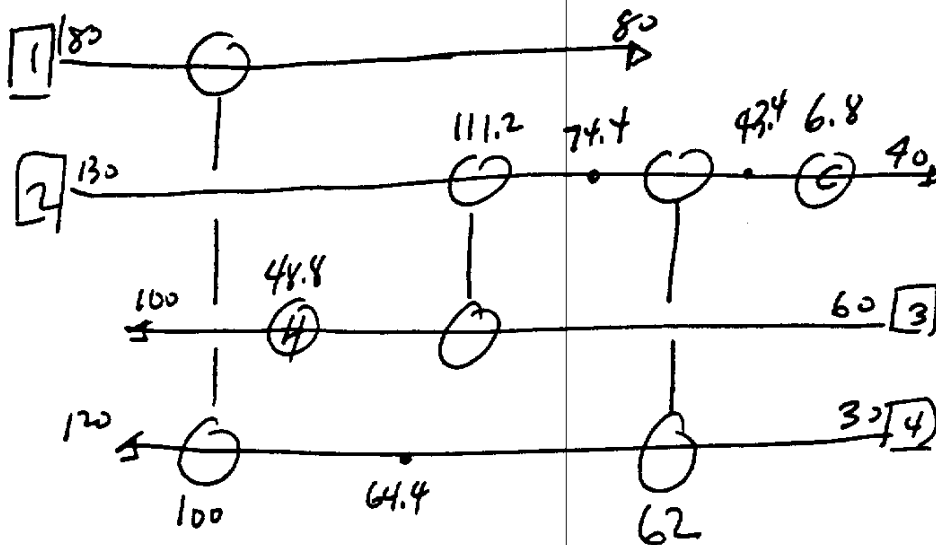
SE SACRIFICA LA EFICIENCIA DEL SISTEMA

DESIGN FOR MER.

- DIVIDE PROBLEM AT PINCH.
- START AT PINCH AND MOVE AWAY
- START WITH BIGGEST STREAMS IN.
- OBSERVE $CP_{OUT} > CP_{IN}$
- PLACE ALL PINCH MATCHES IN THIS WAY
- MAXIMIZE LOADS ON PINCH MATCHES.
- FILL IN THE REST.

SAMPLE PROBLEM. - REDUCE UNITS BY 1

- ① ELIMINATE HEATER ON STREAM ④
 BY CROSS EXCHANGE ② → ④.



NUMBER OF UNITS = 5

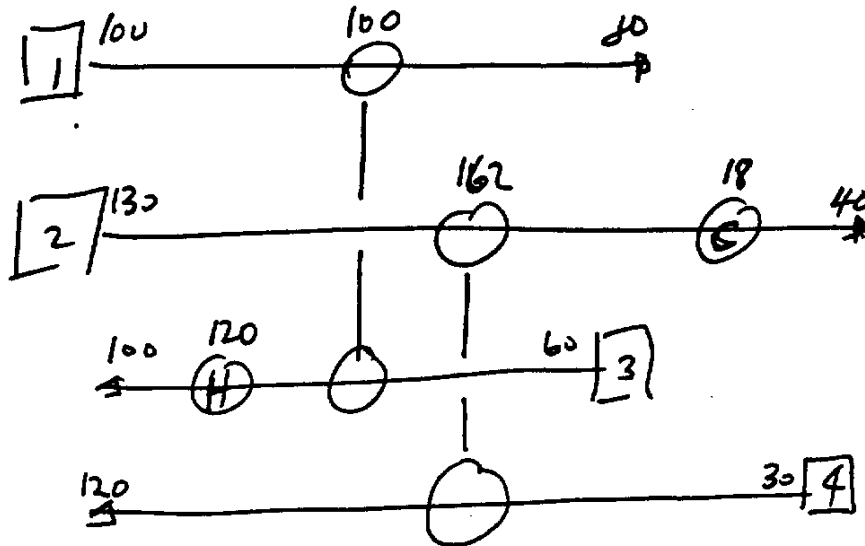
REMEMBER PINCH EQUATION

$$A = T + XP$$

$$48.8 = 48 + XP$$

$$XP = 0.8 \text{ UNIT CROSS PINCH EXCHANGE.}$$

MINIMUM UNITS



IN GENERAL

$$U_{\min} = N - 1$$

 N = NUMBER OF STREAMS (INCLUDING UTILITIES.

FROM GRAPH THEORY - EULER'S GENERAL NETWORK THEOREM.

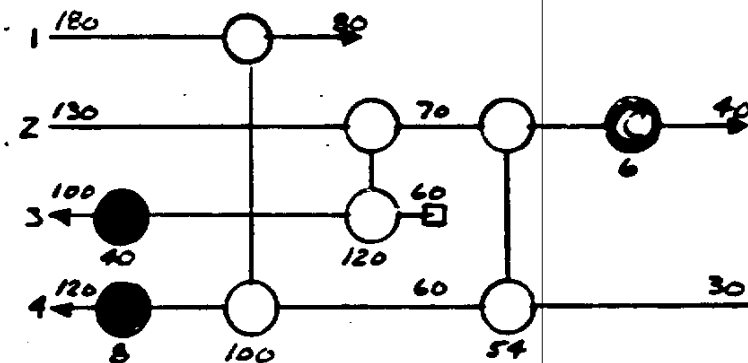
$$U = N + L - S$$

 U = NUMBER OF UNITS (INCLUDING HEATERS & COOLERS)

 N = NUMBER OF STREAMS (INCLUDING UTILITIES)

 S = NUMBER OF SEPARATE SYSTEMS.

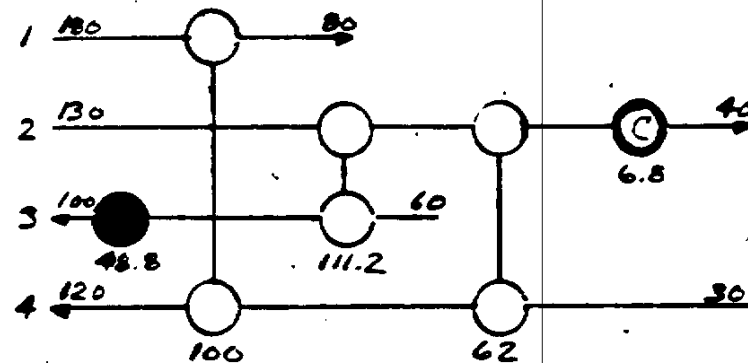
INTRODUCTORY EXAMPLE SOLUTIONS



$$Q_H = \frac{48}{6}$$

$$Q_C = \frac{48}{6}$$

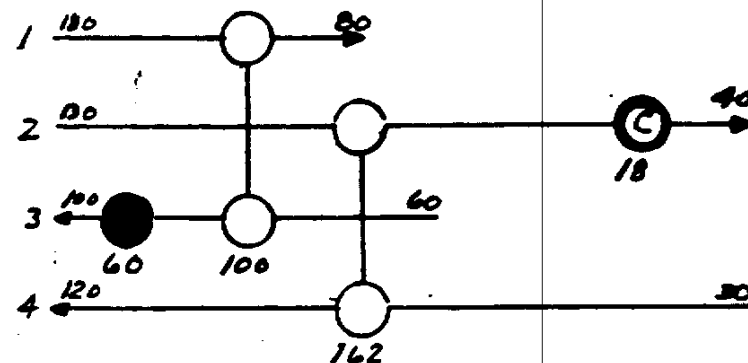
6 UNITS



$$Q_H = \frac{48.8}{6.8}$$

$$Q_C = \frac{48.8}{6.8}$$

5 UNITS

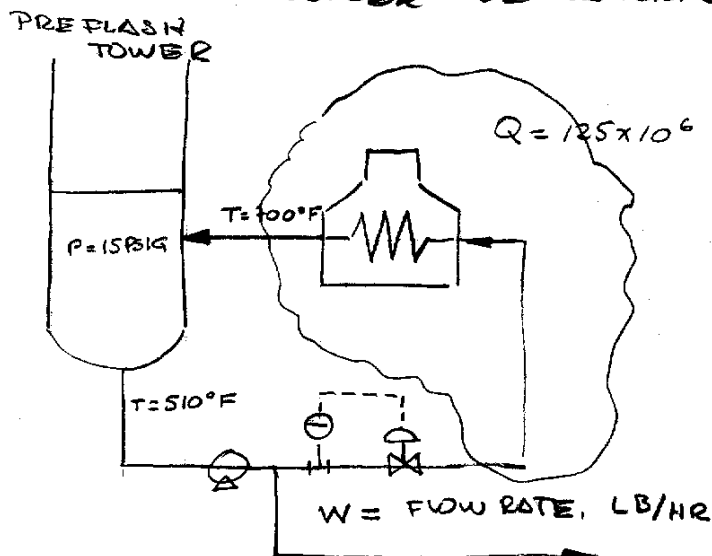


$$Q_H = \frac{60}{18}$$

$$Q_C = \frac{60}{18}$$

4 UNITS

DETERMINACION DEL FLUJO CIRCULANTE POR EL REBOILER DE LA TORRE PRE FLASH.



$$W_{IN} = W_{OUT} \quad \text{HOSS}$$

$$(W_{IN})(h_{LIQ}^{510}) + Q_R = W_{OUT}(h_{OUT}^{700})$$

$$W_{OUT} = V_{OUT} + L_{OUT}$$

$$V = \text{VAPOR}$$

$$L = \text{LIQUID}$$

$$(W_{IN})(h_{LIQ}^{510}) + Q_R = V_{OUT}(h_{VAP}^{700}) + (L_{OUT})(h_{LIQ}^{700})$$

DESCONOC. DESCONOCIDOS

$$W_{OUT} = V_{OUT} + L_{OUT}$$

LA FRACCION DE VAPOR Y LIQUIDO SE PUEDE DETERMINAR MEDIANTE LA CURVA FLASH DEL CAMPO TOPPED.

A 700°F EN LA CURVA FLASH SE TIENE QUE EL 38% ES VAPOR PERO NO SE CONOCE SU COMPOSICION.

PARA ESTO SE UTILIZA EL BREAK UP (POG #1 ANSWER)

	VOL RATE	% VOL	% VOL AUMENTADO	
HVY NAPT	3000 BPD	2.73	2.73	↑ VAPOR
JET FUEL	24825	22.56	25.29	
DIESEL	13725	12.47	37.76	
AGO	11325	10.29	48.05	↓ LIQUID
RESID	57150	51.94		
	110025			

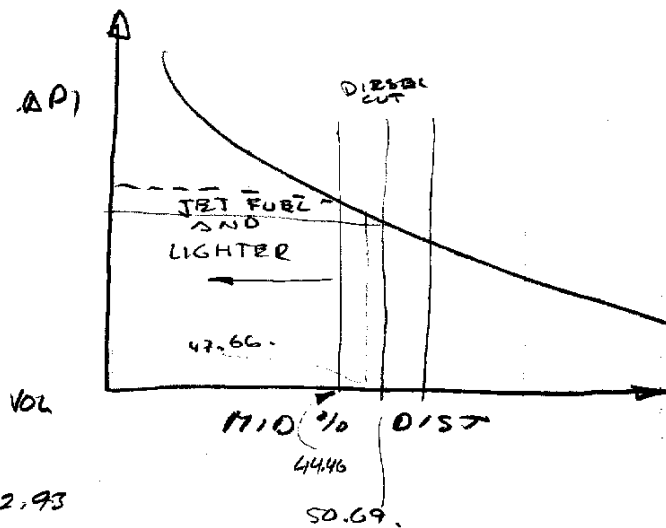
UTILIZANDO UNA TEMP DE 680°F PARA TENER UNA VAPORIZACION PORCENTUAL DEL DIESEL, SE TIENE QUE EL % VAPOR ES 34%

PARA DETERMINAR LA PROCCION DE DIESEL VAPORIZADA A LA
LA VAPORIZACION mediante la siguiente

	% VOL	% VOL A.C.
DIESEL VAP	8.71	34%
DIESEL LIQ	3.76	37.76.

	DBT	UOPK	HW	DB VOL FLU. NO CRUDO TOTAL	DBL % FLU. NO CRUDO	LB/AR PASADO	BPD FLU. CRU
DIE VAP				6.39	8.71	117700	9580
DIE LIQ				2.76	3.76	52490	4145
DIESEL TOT	3SD	11.93	253	9.15	12.47	170190	13725

PARA DETERMINAR LA API SE UTILIZA LA INFORMACION
DEL ASSAY DEL CRUDO (CURVA API VS MID % DIST



CS	2.93
NAP	25.98
JET	16.55
	44.46
DIESEL	6.39
	50.69

FRACC VAP.
RESP. CRUDO
TOTAL

MID.PT

$$\frac{6.39}{2} \rightarrow \frac{44.46}{47.66} \rightarrow 13.20$$

FRACC NO
VAP ORU
TOTAL.

DE LA CURVA PARA
50.69% $\rightarrow$ API = 36.5
UOPK = 11.94
HW =

DE FIG

$$DIBERG = 0.0871 \times 110025 = 9580 \text{ PSD.}$$

$$\Delta P = 36.5$$

$$DIBERG \text{ VAPOR} = \frac{\text{LB/HR}}{\text{BPD}} (9580)$$

$$= 12.289 \times 9580 = 117700 \text{ LB/HR}$$

FINALMENTE:

	API	UOPK	HW	0/10 VOL ON TOTAL CRUDE	% VOL PLASMA CRUDE	LB/HR	BPD ROSSO.
DIBERG VAP	36.5	11.94		6.39	8.71	117700	9580
DIBERG LIQ	31.5	11.91		2.76	3.76	52490	4145
TOTAL	35.0	11.93		9.15	12.47	170190	13725

PROCCION VAPORIZADO BASADO EN EL FLUJO TOTAL DE LOS FONDOS

	API	UOPK	CABP**	LB/HR	BTU/LB	BTU/HR
HV NAR	50.3	11.77	360	34070	490	
JET FUR	42.8	11.97	461	294055	505	
DIESEL	36.5	11.94	570	117700	500	
TOTAL				445825		224042075

$$HV = \text{BTU/LB} = 224042075 / 445825 = 502$$

** PARA CABP SE CONOCE API $\neq$ UOPK DEL GRÁFICO
CON CABP $\neq$ UPK $\neq$ TEMP SE OBTIENE ENTALPIA VAPOR
@ 680°F.

DE FORMA SIMILAR PARA LA PROCCION LIQUIDA.

	API	UOPK	CABP	LB/HR	BTU/LB	BTU/HR
DIESEL LIQ	31.5	11.91	660	52490	385	
AGO	29.5	11.87	690	145225	375	
RESID	17.5	11.72	940	79185	360	
				989550		359732225

$$h_{liq} = \frac{359732225}{989550} = 363$$

ENTALPIA DE LA MISTURA.

$$\begin{aligned}
 \text{BUT MIX} &= (H_V^{650}) \cdot \left(\frac{W_T \text{ PRODUCT}}{VAPOR} \right) + (h_l^{650}) \left(\frac{W_S \text{ PRODUCT}}{VAPOR} \right) \\
 &= 502 \times \frac{445025}{989550 + 445025} + 363 \frac{989550}{989550 + 445025} \\
 &= 406
 \end{aligned}$$

$$(W_{IN}) (h_{lg}^{510}) + Q_R = W_{OUT} (406)$$

$$W_{IN} \times 257.6 + 125 \times 10^6 = W_{IN} (406)$$

$$W_{IN} = \frac{125 \times 10^6}{406 - 257.6}$$

$$W = 342318.0 \text{ lb/hr.}$$