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DOE/CE/15303--T2

DE88 000377

FEASIBILITY OF APPLICATION OF THAYER METHOD TO CRUDE OIL DISTILLATION

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Submitted to:

TEXIM Associates, Inc.
15402 Wandering Trail, Friendswood, TX 77545
for:

Office of Energy-Related Inventions
U. S. Department of Energy

F201-87CE15303

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SUMMARY

The purpose of this report is to present the results of an evaluation of Victor R. Thayer's invention: "Process and Apparatus for reducing the Energy Required to Separate Liquids by Distillation", applied to crude oil distillation. In brief, Thayer claims that given a side draw-off pump-around on a distillation column for energy conservation, there would be a significant cost advantage by using on-tray heat exchange rather than external exchange.

In order to test the method calculations were based on a 200,000 bbl/stream day revamped crude unit, described in the literature. To implement energy conservation, an external pump-around of 54×10^6 Btu/h duty, producing 75,460 lb/h of 600 psig steam was installed, and chosen for analysis because tube side fouling would be minimal (treated boiler feed water). Fouling on the shell side would be the usual carbonaceous deposits encountered in crude service. Calculations were made to size two comparative cases.

Sizing the external case was straight forward. For the internal case, the column diameter was sized at 36ft requiring center-to-side downcomer flow. On-tray heat exchange required 4 parallel bundles: 5 rows x 51 2"od tubes/row, i.e. 128 U-tubes per tube sheet with a partitioned head for water flow in and steam out. The bundles would be manifolded together, and require access through the column wall for cleaning and removal. Also, 5 tube rows high would require additional column height to maintain liquid-vapor action, and a gradient for flow on the tray.

Preliminary cost estimates were made: external case, \$1,091,000 $\pm$ 15%; on-tray case, \$882,200 $\pm$ 20%; both values falling within the estimated uncertainty limits. Therefore, and coupled with much greater annual maintenance costs, a significant cost advantage of the on-tray system was not verified.

The primary conclusion of the evaluation was that, in our opinion, the Thayer method will not be used for retrofit of crude oil distillation units, and probably will not be used in the design of new crude units.

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1.00 INTRODUCTION

The objective of this report is to present the results of an evaluation of Victor R. Thayer's invention: "Process and Apparatus for Reducing the Energy Required to Separate Liquids by Distillation", (U. S. Patent #4,265,736, 5 May 1981), for the specific application of crude oil distillation. In brief, the essential claim of the invention is that, given a side-draw-off or pump-around on a distillation column, there is significant cost advantage to be gained if an on-tray heat exchange is used rather than an external exchange.

The energy conservation analysis by Bannon and Marple [1] indicates that it is possible to recover essentially all the heat required to bring the feed stream to its vaporization temperature, and that energy conservation is accomplished by heat exchange between the liquid feed and the vapor leaving the top of the tower in several stages rather than in one stage. Thayer proposes a method whereby this principle can be implemented; wherein, he claims on-tray heat exchange is favored economically over external heat exchange.

More specifically Thayer claims the following six (6, A-F incl) advantages for his method (see patent, loc. cit., columns 3-4):

- A. External heat exchangers are wholly or partially eliminated.
- B. Pumps, piping and flow-control instrumentation required to recirculate liquid through external heat exchangers are wholly or partially eliminated.
- C. The indirect transfer of heat is more rapid with heat exchange tubing installed on the trays as described because the liquid on the trays is violently agitated by the gas passing upward through the trays.
- D. The provision of heat exchange tubing on a large number of trays in the fractionating tower makes each tray so provided an individual condenser and thereby improves heat recovery because heat is exchanged in many more stages (internally) than can be economically provided when using external exchangers. This reduces the amount of makeup heat that needs to be supplied to the heater or vaporizer, such makeup heat being an important item of operating cost.

and also makes possible the use of a smaller heater, thereby saving capital cost.

E. The installation of heat exchange tubes on the trays of the fractionating tower eliminates the need for the external shells or vessels normally used to contain such tubes and thereby reduces the installed cost of such tubes.

F. All Trays in the tower work at substantially maximum efficiency for fractionating the feed mixture. In systems used in the prior art, whereby reflux liquid is withdrawn from the tower, cooled externally and then returned to the tower, the trays between which the reflux liquid flows between the point of return and the point of withdrawal do not work at maximum efficiency because the recycle of liquid partially counteracts the fractionation that otherwise occurs on each tray.

Thayer recognized that implementation of his method, for a given distillation column, would in the final analysis reduce to the cost involved; i.e. the capital, maintenance, and operating costs of the comparative cases, on-tray vs. external exchangers.

The authors' of this report had available three (3) items, two (2) prior reports by Howard Robb, NBS/OERI [2], and Professor David Kaufmann, PE [3]; and the evaluation request by Victor Thayer [4] to the U.S. Department of Commerce, including the patent, a draft of an unpublished paper submitted to the AIChE for consideration, and a narrative disclosure presenting suggested possible applications of the method.

Robb's report presented four (4) conclusions concerning Thayer's method: in brief, a) technical feasibility is uncertain; b) the potential for energy savings is significant; c) commercial feasibility is uncertain; and d) the next stage in the system development should be a parametric and design analysis to estimate operating characteristics and examine feasibility.

Kaufmann's report presented six (6) conclusions: in brief, a) the proposed method is technically feasible; b) the method has potential to reduce capital and operating costs for some types of distillation columns; c) the method has potential for saving energy by making thermal energy recovery more economical; d) some uncertainties in technical details of the process exist; e) major uncertainties are related to ease of operation,

maintainability, and achievable stream factor; f) implementation of the method may be more related to current world-wide excess capacity in refining and petrochemicals.

The evaluative comments of an unidentified chief engineer (called Cf-Engineer X hereafter) in an operating oil company, to whom Thayer submitted his patent and draft paper, are very very pertinent, obviously representing current operating experience, and are summarized here in brief: a) the maintenance difficulty (using on-tray exchangers) would increase dramatically; b) in-tower heat exchange is not desirable for several mechanical reasons-- the optimum crude column side draw-off location is probably 150 ft above grade, making maintenance expensive and difficult; c) heat exchangers use relatively thin tubes and leaks are common, testing and repair would be much more difficult and costly; d) hydraulic design of mass transfer trays is critical, an on-tray heat exchanger would severely affect liquid flows and add column height; e) external heat exchangers are carefully designed to achieve high efficiency, we doubt in-tower exchangers can be designed to achieve heat transfer effectiveness and comparable fouling even approaching that of external exchangers; and f) in these times flexibility of design is very important, as crude compositions and unit operations change-- external heat exchange would provide much more flexibility.

It should be mentioned that the authors of this report have operating plant experience. The senior author (HWPj) was employed by a major oil company as a senior technologist in one of their refineries, and had experience with crude units. The junior author (VEA) was employed by a chemical company and later with a major design and engineering construction company.

The sections of this report which follow describe the evaluation of the feasibility of application of the Thayer method to crude oil distillation.

2.00 CHOICE OF CRUDE COLUMN EXAMPLE

It was suggested by Thayer and others that a crude oil distillation column, with the numerous side draw-offs, might be a good example for a check of feasibility of the method. At the

outset it was necessary to find a published paper that represents modern crude oil distillation technology, so that scale, configuration, and energy conservation methods could be used as a point of departure. A paper by Waldsmith [5] from Chemical Engineering Progress, which discusses the modification of Shell's Norco Refinery Crude unit to process 200,000 bbl/stream day of North African (NA) crudes, satisfies the requirement. Pertinent features of the unit are as follows.

1) The unit was originally designed and built to process Louisiana Delta (LD) crude at the same feed rate of 200,000 bbl/sd. Note the scale of this unit (9.4×10^6 Tons/y) which is typical of modern refinery crude units, and 100-1000 times the scale of chemical plant units.

2) Modifications were required to debottleneck the unit to process the NA crude, provide maximum energy conservation, and retain necessary flexibility to process LD crude also (recall Cf-Engineer X's comment on flexibility).

3) The design of modifications were optimized to process a mixture of 125,000 NA crude plus 75,000 LD crude, bbl/sd, and also retain the capability of processing 200,000 bbl/sd LD crude. Design features were included to maximize energy conservation which could be economically justified in a reasonable length of time (comment: perhaps 2-3 years).

4) The modification work plan produced two preliminary designs: a) -revamp to process above stated crude mix at minimum capital investment, and 2) -revamp to recover maximum amount of heat (comment: perhaps up to 20-30% over original configuration) which could be economically justified. Cost estimates were prepared and a combination solution chosen.

5) In its original configuration the crude unit consists of three (3) columns: a) a lower column, taking the feed from a process heater, producing extra-heavy gas oil (XHGO) and bottoms to the vacuum column, and supplying overhead vapor to the second column; b) an upper (second) column, producing HGO, kerosene, naphtha and overhead light ends; and c) a vacuum column with a second process heater, producing pitch and pyrolysis and catalytic feedstocks.

The process flow of the revamped unit follows.

While insufficient information is provided in the Waldsmith paper to reproduce the internal column flows for the revamped case, a hypothetical set of flows and energy balance can be constructed to test application of the Thayer method. On-tray vs. external heat exchange, on tray 13 of the lower column, was the basis and approach taken for this report.

3.00 RESULTS OF CALCULATIONS

The detailed process engineering calculations are presented in Appendix A, A1. to A6. incl., covering: mass and energy balance, Trays 1 and 13; column sizing, Tray 1; shell and tube overall U, internal exchanger overall U; tube wall thickness; and mechanical configuration of on-tray tubes. Operating data, properties, and correlations, used to make the calculations were obtained from, Waldsmith [5], Oil & Gas Journal [6], Perry [7], Kern [8], and the API Technical Data Book [9].

Important results of the calculations in each appendix (A-series) follow. An important consideration was the degree of fouling in choosing an example to test the Thayer method. Experience has show that carbonaceous deposits are encountered in crude oil distillation, and must be removed during annual maintenance shut-downs, a fact that must be dealt with if implementation is to be achieved.

- 1) A1: Tray 13 with its pump-around producing 54×10^6 Btu/h of

steam was chosen as the location in the lower tower to test the method; primarily for the reason that treated boiler feed water, with its relatively low fouling factor (0.0005 cf. 0.0040 for crude oil), can be used to generate the steam, thereby reducing the frequency of tube-side cleaning. However, fouling on the shell-side must be dealt with. A diagram of the lower part of the column(p1/5) shows the flows involved and operating conditions.

2) A2: Having the flows involved a tower diameter of 36 ft was calculated, not uncommon for crude units, indicating that a split flow system of center-to-side downcomer bubble cap trays would be used, as is standard practice.

3) A3: For the external pump-around case, shell and tube heat exchanger calculations were made, to produce 75,460 lb/h of 600 psig steam from the 54×10^6 Btu/h, $U = 39.21$ Btu/h ft²deg, and requiring approx. 17,257 ft² of heat transfer surface.

4) A4: Similar calculations were made for the on-tray case, resulting in $U = 91.38$ and approx. 7,405 ft² of heat transfer surface for the same duty.

5) A5: 2"od x 0.20"wall thickness carbon steel tubes resulted for both internal and external cases.

6) A6: The mechanical configuration of the on-tray case required special attention. 20% of the column area was allocated to downcomers, disengaging, and distributing zones, leaving a 12.85 ft wide trapezoidal area (see A6 p2/6 for diagram) for each of 4 bundles of tubes: 5 rows x 51 tubes/row per tube sheet, i.e. 255 single tubes or 128 U-tubes per tube sheet. Note that the tube sheet is diagonal; also a partitioned head to direct flow in and out of the tubes would be required. Bundles would be in parallel and manifolded together through nozzles passing through the column wall. Bundles must be accessible through the wall of the column for cleaning and removal.

7) The very pertinent question arises: how will Tray 13 liquid-vapor action be modified as a result of the four bundles on the tray? It is obvious that with 5 rows of tubes normal flow would be substantially disturbed, and in fact additional column height would have to be provided to insure coverage of the tubes by the liquid-vapor froth and to produce a wier-to-downcomer

gradient necessary to maintain flow. If we were dealing with "clean" service, the possibility of replacing the bubble-cap tray with a perforated-plate tray could be considered, however because of inherent fouling such a substitution is not possible.

Based on the foregoing considerations preliminary cost estimates can be prepared for the two alternative cases, which follow.

4.00 COST ESTIMATES and COMPARISON

And now to the preliminary cost estimate calculations presented in Appendix A7. Unit cost data used were obtained from Guthrie [10, 1970], Peters and Timmerhaus [11, 1979], and recent economic indicators from Chemical Engineering [12, 1987].

For the external heat exchange system, as well as the internal on-tray system, all items of purchased equipment were estimated individually using the above references, and then escalated to current costs. Estimation of the external system was straight forward because of its conventional nature; however, since there is no existing experience for the internal system, only a rough visualization was possible, hence the cost estimate has greater uncertainty.

Tables 1 and 2 present a summary of the preliminary cost estimates for the two respective cases. It will be noted that our estimate indicates that purchased equipment is greater for the external heat exchange case than for the on-tray case. The total estimated costs are:

External Heat Exchange case = \$1,091,000 $\pm$ 15%

Internal On-Tray case = \$882,200 $\pm$ 20%

The assigned uncertainties of $\pm$ 15% and $\pm$ 20% respectively place the totals within the same range, i.e. the lower and upper limits being \$927,350 and \$1,058,600 respectively. In addition there will be a substantial increased maintenance cost for the on-tray case compared to the external case; no specific estimate of this difference was made.

**TABLE 1. Preliminary Cost Estimate for External
Heat Exchange - Pump-Around from Tray 13**

1) 10 Heat Exchangers, bundles + shells	\$ 339,800
2) Tray 13 (standard)	26,030
3) 2 Pumps + motors	22,504
4) Piping, valves, fittings + CV	17,170
5) Insulation	20,756
6) Instrumentation & Controls	24,360
7) Electrical Equipment + hookup	51,540
	<hr/>
purchased equipment	(502,160)
8) Equipment installation	167,400
9) Indirect Costs (29.4% of subtotal)	279,000
- engineering + supervision	
-construction expense	
-contractors fee	
	<hr/>
subtotal	\$ 948,560
10) Contingency (15%)	142,284
	<hr/>
Total (+15%)	\$ <u>1,091,000</u>

**TABLE 2. Preliminary Cost Estimate for On-Tray
Heat Exchange- on Tray 13**

1) 4 Heat Exchange bundles w/heads	\$ 170,680
2) New Design Tray 13	52,060
3) Manifold piping (internal & external), nozzles + CV	35,000
4) Increased tower height + access ways (for cleaning)	12,236
5) Service Platform w/hand rails and supports (at Tray 13 level)	28,472
6) Instrumentation & controls	24,360
7) Electrical eqwuipment + hookup	25,000
	<hr/>
purchased equipment	(347,808)
8) Equipment Installation	154,600
9) Indirect Costs (31.6% of subtotal)	231,870
-engineering + supervision	
-construction expense	
-contractors fee	
	<hr/>
subtotal	734,300
10) Contingency (20%)	147,900
	<hr/>
Total (+20%)	\$ <u>882,200</u>

5.00 CONCLUSIONS and DISCUSSION

Based on the analysis and synthesis carried out, it is our opinion that the following conclusions can be drawn.

1) The Thayer method will not be used for retrofit (revamp) of crude oil distillation units, and there is a high probability that it will not be used in the design of new crude units.

The preliminary cost comparison does not show a significant advantage in favor of the Thayer method, and coupled with the probable substantially increased yearly maintenance cost, it is highly doubtful that the method would be implemented for crude units.

2) For crude units, Thayer's six (6) claims (p1-2) do not produce a sufficient cost advantage to offset the complexity, increased cost of additional column height, and maintenance, for an on-tray unit.

3) Our analysis indicates that the evaluative comments by Cf-Engineer X, on application to crude units, are substantially correct on most points, and sufficient to greatly preclude application.

4) Application to crude units represent the most severe test for application of the Thayer method because of size and inherent fouling by carbonaceous deposits.

We do not close out possible application of the Thayer method to smaller scale "clean" service cases, where for an on-tray installation a bubble-cap tray can be replaced by a perforated tray, and with essentially non-fouling fluids (fouling factors <0.0002) application may become possible.

6.00 RECOMENDATION

It is recommended that this report be included with the other evaluative reports on the proposed Thayer technology, so that prospective process industry users may have the benefit of this analysis on crude units.

References Used

- [1] Bannon, R. P. and S. Marple, "Heat Recovery in Hydrocarbon Distillation", Chem. Eng. Progress, 74, p41-45, July 1978.
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- [3] Kaufmann, D., "Second Stage Evaluation- (same title as [2] above)", submitted to U. S. Dept of Commerce, 10 Aug 1983.
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- [5] Waldsmith, R.W., "Revamping Shell-Norco's Crude Unit", Chem. Eng. Progress, 76(9), p65-69, Sept 1980; (included as Appendix B.1).
- [6] Oil & Gas Journal, "Crude Oil Properties (USA)- Louisiana Delta", (1973); (included as Appendix B.2).
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- [10] Guthrie, K.M., Process Plant Estimating Evaluation and Control, Craftsman Book Company of America, Solana Beach, CA (1974).
- [11] Peters, M. S. and K. D. Timmerhaus, Plant Design and Economics for Chemical Engineers, third edition, McGraw-Hill Book Company, NY (1980).
- [12] Chemical Engineering, "Economic Indicators", 16 Mar 1987; (included as Appendix B.3).

APPENDIX A.

A1. Mass Balance, Trays 1 & 13

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Tray 13 lies immediately below the heavy gas oil draw-off; therefore, no liquid is assumed to flow to Tray 13 from Tray 14, i.e. $L_{14} = 0$. Vapor flow to Tray 13, V_{12} , is the vaporized diesel and lighter material used to determine the tower diameter below Tray 1. Heat exchange equipment on Tray 13 or the Tray 13 pump-around condenses diesel from the vapor stream to produce the liquid stream from Tray 13, L_{13} . Heat transferred, 54×10^6 Btu/h, condenses 0.634×10^6 lb/h of diesel.

The diagram below summarizes the calculations.

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A1.
2/5

TRAY 13

MATERIAL BALANCE:

VAPOR:
DIESEL AND
LIGHTER,

↑ VAPOR

VAPOR + GAS OIL

→ HEAVY GAS OIL

TRAY 13

 $T = 666^{\circ}\text{F}$

$$Q = (54)(10^6) \frac{\text{BTU}}{\text{HR}}$$

TO CONDENSE DIESEL

FEED

200,000 $\frac{\text{BBL}}{\text{DAY}}$

TO VACUUM TOWER

ASPHALT 20,600 $\frac{\text{BBL}}{\text{DAY}}$ + GAS OIL, 67,000.0 $\frac{\text{BBL}}{\text{DAY}}$
+ DIESEL (5,400)GAS OIL FLOWRATE IN FEED = 67,000 $\frac{\text{BBL}}{\text{DAY}}$ HEAT OF VAPORIZATION = 84 $\frac{\text{BTU}}{\text{LB}}$ (NELSON, 1949, FIG. 41, P. 142)DENSITY = 37 $\frac{\text{LB}_m}{\text{FT}^3}$, (A.P.I. 6A3.4)

DIESEL-

 $T = 666^{\circ}\text{F}$, $K = 11.84$, 34°API

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A1.
3/5

DIESEL CONDENSED ON TRAY 13

(54)(10 ⁶) BTU	LB _m	FT ³	7.4805 GAL	BBL	24 HR
Hour	84 BTU	37 LB _m	FT ³	426 GAL	DAY

$$= 642,857.1 \frac{\text{LB}_m}{\text{HR.}} = 17,374.52 \frac{\text{FT}^3}{\text{HR. (LIQUID)}} = 74,268,6156 \frac{\text{BBL}}{\text{DAY}}$$

$$= 74,268.62 \frac{\text{BBL}}{\text{DAY}} \text{ DIESEL RECTIFIED ON TRAY 13}$$

FEED TO VACUUM TOWER:

20,600 BBL/DAY ASPHALT

67,000 BBL/DAY GAS OIL -

+ 74,268.6 BBL/DAY DIESEL

 161,869. BBL/DAY
VAPOR BELOW TRAY 13, (COLUMN SIZING CALC'S)
(12 JUNE 87)

$$= 1,347,846.3 \frac{\text{LB}_m}{\text{HR}} = 8,304,4272 \frac{\text{LB}_m}{\text{HR.}}$$

$$= 6,907,865.6 \frac{\text{FT}^3}{\text{HR}}$$

VAPOR ABOVE TRAY 13,

$$1,347,846.3 \frac{\text{LB}_m}{\text{HR}} - 642,857.1 \frac{\text{LB}_m}{\text{HR}} = 704,989.2 \frac{\text{LB}_m}{\text{HR.}} \text{ (DIESEL CONDENSED)}$$

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A1.
4/5-

DIESEL CONDENSED: (REF. API, FIG. 2B2.1)

$$642,857.1 \frac{\text{LB}_m}{\text{HR}} \bigg| \frac{\text{LB}_m}{215 \text{ LB}_m} = 2935.42 \frac{\text{LB}_m}{\text{HR}}$$

VAPOR FLOW RATE ABOVE TRAY 13:

$$8304.4272 \frac{\text{LB-MOLE}}{\text{HOUR}} - 2935.42 \frac{\text{LB-MOLE}}{\text{HOUR}} =$$

$$= 5369.007 \frac{\text{LB-MOLE}}{\text{HR}} =$$

$$5369.007 \frac{\text{LB-MOLE}}{\text{HR}} \bigg| \frac{359 \text{ FT}^3}{\text{LB-MOLE}} \bigg| \frac{(460 + 666)^\circ\text{R}}{492^\circ\text{R}} =$$

$$= 4,411,250.4 \frac{\text{FT}^3}{\text{HR}} @ 666^\circ\text{F}, 1 \text{ ATM.}$$

30 JUNE 1987

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A1.
5/542.381 50 SHEETS 5 SQUARE
42.382 100 SHEETS 5 SQUARE
42.389 200 SHEETS 5 SQUARE
NATIONAL

MATERIAL BALANCE:

BOTTOM OF ATMOSPHERIC TOWER TO TRAY 13

E. Atthaus
(6-11-87)

A2. Column Sizing, Tray 1

1/5

THE CRUDE TOWER DIAMETER, 35.85 FT., WAS DETERMINED BY THE SOUDERS-BROWN CORRELATION GIVEN BY PERRY (4TH ED., EQ. 18-1) AND THEN ENLARGED BY A FACTOR OF 1.35 ($1/0.74$) TO ACCOUNT FOR AREA LOST TO DOWNCOMERS, SUPPORTS AND DISENGAGING AND DISTRIBUTING ZONES (PERRY, 4TH ED., TABLE 18-2).

THE TOWER DIAMETER WAS ESTIMATED BASED ON MAXIMUM VAPOR LOAD WHICH WOULD BE EXPERIENCED BELOW TRAY #1. ALL MATERIAL DIESEL AND LIGHTER ARE PRESUMED VAPORIZED. IDEAL GAS LAW CALCULATIONS DETERMINED VAPOR FLOWRATE AND DENSITY. LIQUID DENSITY IS THE DENSITY OF THE RESIDUUM.

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E. ALTHAUS

A2.
2/5

ESTIMATE CRUDE TOWER DIAMETER
BASED ON SOUDERS-BROWN CORRELATION,
PERIZY, 4TH ED, EQ. 18-1

BASE DIAMETER ON FLOWRATE AND
DENSITY AT BOTTOM OF TOWER.

PRELIMINARY MATERIAL BALANCE ON PAGE 1

DETERMINE MASS FLOWRATE OF C_4 & LIGHTER

$$\text{FLOWRATE} = \frac{2000 \text{ BBL (LW)}}{\text{DAY}}$$

$$\text{CRUDE DENSITY} = 30.6^\circ \text{ API}$$

$$= 0.8729 \text{ SP. GR } \frac{60}{60}, \text{ API TABLE}$$

$$= 7.269 \frac{\text{LB}}{\text{GAL}} = 54.38 \frac{\text{LB}}{\text{FT}^3} = 305.3 \frac{\text{LB}_m}{\text{BBL}}$$

$$\text{MOLECULAR WEIGHT BETWEEN } C_1 - C_4 = 37$$

MOLAR

MASS FLOWRATE =

$$\frac{2000 \text{ BBL}}{\text{DAY}} \left| \frac{305.3 \text{ lb}_m}{\text{BBL}} \right| \left| \frac{1 \text{ lb-mole}}{37 \text{ lb}} \right| \left| \frac{1 \text{ DAY}}{24 \text{ HR}} \right| = 687.61261 \frac{\text{LB}_m}{\text{HR}}$$

$$\text{MASS FLOWRATE} = 25441.6667 \frac{\text{lb}_m}{\text{HR}}$$

42.381 50 SHEETS 3 SQUARE
42.382 100 SHEETS 3 SQUARE
42.389 200 SHEETS 3 SQUARE

COMP- ONENT	TBP RANGE (°F)	VOL % %	FLOW RATE BBL DAY	GRAVITY O API ($\frac{15.6}{15.6}$)	CHAR. FACTOR	MW	MASS FLOW RATE ($\frac{LB_m}{Hr}$)	MOLAR FLOW RATE ($\frac{LB-mole}{Hr}$)
C ₄ $\frac{1}{2}$ LIGHT		1.0	2,000.0			37	25,444.667	687.61261
LIGHT NAPHTHA	57- 310	11.7	23,400.0	62.2 (255.14)	12.05	106	249,015.00	2349.1981
HEAVY NAPHTHA	310-375	5.8	11,600.0	46.9 (277.3)	11.64	137	134,028.333	978.3090
DIESEL	375-680	37.7	75,400.0	34.0 (299.0)		210	939,358.333	4,289.3075
W.R.GAS OIL	680- ASPHALT	33.5 →	87,600.0	22.0 18.8 (329.3)	11.84	11.84	1,347,846.333	8,304.4272
ASPHALT		10.3		9.6 (350.8)				

↑
REF. API, FIG ZB2.1

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LIQUID DENSITY = 18.8 ° API, RESIDUUM.
(GAS OIL + ASPHALT)

@ T = 680 °F

$\rho = 43.7 \text{ lbm/ft}^3$, API FIG. 6A3.4

LIQUID

GAS FLOW RATE =

$$\frac{8304.4272 \text{ lb-mole}}{\text{HOUR}} \times \frac{359 \text{ ft}^3}{\text{lb-mole}} \times \frac{(460 + 680)^\circ \text{R}}{492^\circ \text{R}} =$$

$$= 6907865.587 \frac{\text{ft}^3}{\text{HR}}$$

$$\text{GAS DENSITY} = \frac{(1,347,846.333 \text{ LBm}) (\cancel{\text{HR}})}{(\cancel{\text{HR}}) (6,907,865.587) (\text{ft}^3)}$$

$$= 0.1951176258 \frac{\text{LBm}}{\text{ft}^3}, 680^\circ \text{F}, 1 \text{ ATM}$$

GO TO SOUDERS-BROWN CORRELATION,
PERRY, 4TH ED., P. 18-6

FOR 24" TRAY SPACING, 1" LIQUID SEAL

$$K_V = 0.172$$

$$U_{\text{max}} = K_V \sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \quad \text{PERRY, EQ 18-1, 4TH ED}$$

$$U_{\text{MAX}} = 0.172 \sqrt{\frac{43.7 \frac{\text{lbm}}{\text{ft}^3} - 0.195117 \frac{\text{LBm}}{\text{ft}^3}}{0.195117 \frac{\text{LBm}}{\text{ft}^3}}}$$

$$U_{\text{MAX}} = 2.56832 \frac{\text{FT}}{\text{SEC.}}$$

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CROSS SECTIONAL AREA =

$$\frac{6,907,865.6 (\text{FT}^3)}{(\text{HOUR})} \times \frac{(\text{SEC})}{2,56832 (\text{FT})} \times \frac{1 \text{ HOUR}}{3600 \text{ SEC}}$$

$$= 747.12313 \text{ FT}^2$$

ACTIVE AREA = (0.74) (TOTAL AREA), TABLE 18-2
PERRY, 4TH ED.

TRAY CROSS SECTIONAL AREA =

$$= \frac{747.123 \text{ FT}^2}{0.74} = 1009.6258 \text{ FT}^2$$

COLUMN DIAMETER =

$$\frac{\pi}{4} D^2 = A; D^2 = \frac{4A}{\pi}; D = \sqrt{\frac{4A}{\pi}}$$

$$D = \sqrt{\frac{(4.0)(1009.6258 \text{ FT}^2)}{\pi}}$$

$$D = 35.8538 \text{ FT.}$$

A3. SHELL & TUBE, OVERALL U

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SHELL AND TUBE HEAT EXCHANGER
CALCULATIONS ARE BASED ON THIS
MECHANICAL LAYOUT PRESUMING THAT
ONLY ONE EXCHANGER WILL BE REQUIRED.

ONE TUBESIDE PASS,

SHELL INSIDE DIAMETER = 48 INCHES

TUBE OUTSIDE DIAMETER = 2.0 INCHES

TUBE WALL THICKNESS = 0.2 INCHES

TUBE PITCH = 3.0 INCHES, TRIANGULAR

BAFFLE SPACING = 2.0 FEET.

NUMBER OF TUBES PER SHELL = 168 (GRAPHICAL ESTIMATE)

TUBE LENGTH = 20 FEET.

THIS HEAT EXCHANGER USES 666 °F OIL
FROM TRAY 13 TO GENERATE 600 PSIG
SATURATED STEAM (499 °F). A 30 °F
APPROACH TEMPERATURE WAS CHOSEN BETWEEN
COOLED OIL AND BOILER FEED WATER.
THIS APPROACH AND THE STEAM DUTY
MENTIONED BY WALDSMITH SPECIFY OIL
AND STEAM FLOW RATES.

A DIAGRAM SUMMARIZING EXCHANGER CONFIGURATION IS GIVEN BELOW.

$$Q = 54 \times 10^6 \text{ BTU/HR}$$

$$\Delta T_{LM} = 79.8^\circ\text{F}$$

THE SHELL SIDE HEAT TRANSFER COEFFICIENT WAS BASED ON AN EQUATION FROM KERN, P. 137. THE CALCULATED VALUE WAS, $h_o = 61.630 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$

THE TUBE SIDE HEAT TRANSFER COEFFICIENT WAS DETERMINED USING FIG. 24 FROM KERN (P. 834). ACCORDING TO KERN, THIS FACTOR CAN BE MULTIPLIED BY 1.25 FOR BOILING

SERVICE (KERN, P. 471). THE RESULTING
INSIDE COEFFICIENT IS, $h_i = 289.94 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$

THE OVERALL HEAT TRANSFER COEFFICIENT
WAS DETERMINED USING THESE COEFFICIENTS
AND FOULING RESISTANCES GIVEN BY KERN
(TABLE 12, P. 845). THE OVERALL COEFFICIENT

IS $U_o = 39.212 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$

THIS RESULTS IN A REQUIRED SURFACE AREA,
 $A_o = 17,257. \text{ FT}^2$. TEN (ACTUALLY 9.8)
OF THE PRESUMED VESSELS WOULD BE REQUIRED.
FLOW RATES WOULD HAVE TO BE ADJUSTED TO
DEVELOP MORE ACCURATE SHELL AND
TUBE SIDE COEFFICIENTS FOR THIS CASE.

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4/15PRESUME A 30°F APPROACHTHEN, $T_{hR} = 529^{\circ}\text{F}$

Q

$$\Delta T_{LM} = \frac{\Delta T_2 - \Delta T_1}{\ln(\Delta T_2 / \Delta T_1)}$$

$$\Delta T_2 = 666^{\circ}\text{F} - 499^{\circ}\text{F} = 167^{\circ}\text{F}$$

$$\Delta T_2 = 529^{\circ}\text{F} - 499^{\circ}\text{F} = 30^{\circ}\text{F}$$

$$\Delta T_{LM} = \frac{167^{\circ}\text{F} - 30^{\circ}\text{F}}{\ln(167^{\circ}\text{F} / 30^{\circ}\text{F})} ; \Delta T_{LM} = 79.7997^{\circ}\text{F}$$

$$Q = 54 \text{ mm BTU / HR}$$

$$Q = \dot{m} C_p \Delta T_H ; \dot{m} = \frac{Q}{C_p \Delta T_H}$$

$$\dot{m} = \frac{(54)(10^6)(\text{BTU})(\text{LB}_m)(^{\circ}\text{F})}{(\text{Hour})(0.745)(\text{BTU})(666 - 529)(^{\circ}\text{F})}$$

$$\dot{m} = 529,074.61 \frac{\text{LB}_m}{\text{HR}}$$

$$\rho = 7.68489 \frac{\text{LB}_m}{\text{GAL}} \quad (22^{\circ}\text{API OIL AT } 60^{\circ}\text{F})$$

$$\dot{m} = 1639.1926 \text{ BBL / HR}$$

$$\dot{m} = 39,340.6 \text{ BBL / DAY}$$

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DETERMINE SHELL SIDE HEAT TRANSFER COEFFICIENT.

OIL BEING COOLED FROM 666°F TO 529°F

EQUATION FROM KERN, P. 137.

$$\frac{h_o D_e}{k} = 0.36 \left(\frac{D_e G_s}{\mu} \right) \left(\frac{c \mu}{k} \right)^{1/3} \left(\frac{\mu}{\mu_w} \right)^{1/4}$$

DETERMINE SHELL SIDE CROSSFLOW AREA

$$Q_s = \frac{ID \times C' B}{P_T \times 144} (ft^2)$$

SHELL ID = 4 FEET, = 48 INCHES, ARBITRARY

CLEARANCE, $C' = 1$ INCH

BAFFLE SPACING = 2.0 FEET, $B = 24$ INCHES

PITCH = 3.0 INCHES

TUBE O.D. = 2.0 INCHES.

$$Q_s = \frac{(48)(INCHES)(1)(INCH)(24)(INCHES)}{(3.0)(INCHES)(144)(IN^2/FT^2)}$$

$$Q_s = 2.66667 \text{ FT}^2$$

DETERMINE MASS VELOCITY, HERN, Eq. 7.2.

$$G_s = \frac{W}{A_s} = \frac{529,074.61 \text{ LB}_m}{\text{HR}} \bigg| \frac{1}{2.66667 \text{ FT}^2}$$

$$G_s = 198,402.9785 \frac{\text{LB}_m}{\text{FT}^2 \text{ HR}}$$

DETERMINE EQUIVALENT DIAMETER FOR A TUBE,

USE TRIANGULAR PITCH. HERN, Eq. 7.5.

$$d_e = \frac{(4) \left(\left(\frac{1}{2} P_T \right) (0.86 P_T) - \left(\frac{1}{2} \pi d_o^2 / 4 \right) \right)}{\frac{1}{2} \pi d_o} \quad [\text{INCHES}]$$

$$P_T = 3.0 \text{ INCHES}$$

$$d_o = 2.0 \text{ INCHES}$$

$$d_e = \frac{(4) \left[\left(\frac{1}{2} \right) (3.0 \text{ INCHES}) (0.86) (3.0 \text{ INCHES}) - \left(\frac{1}{2} \pi \right) (2.0 \text{ IN})^2 / 4 \right]}{\frac{1}{2} \pi (2.0 \text{ INCHES})}$$

$$d_e = \frac{(4) [3.87 - 1.5708] \text{ IN}^2}{\left(\frac{1}{2} \right) (\pi) (2.0) (\text{INCHES})}$$

$$d_e = 2.92743 \text{ INCHES}$$

$$D_e = 0.243953 \text{ FT,}$$

CHECK SHELL SIDE REYNOLDS NUMBER

$$Re = \frac{D_e G}{\mu}$$

$$\begin{aligned} \mu &= 160 \text{ CENTISTOKES @ } 100^\circ\text{F} \\ &= 9.8 \text{ CENTISTOKES @ } 210^\circ\text{F} \end{aligned} \quad \left. \vphantom{\begin{aligned} \mu &= 160 \text{ CENTISTOKES @ } 100^\circ\text{F} \\ &= 9.8 \text{ CENTISTOKES @ } 210^\circ\text{F} \end{aligned}} \right] \text{API FIG 11A4.1}$$

$$\text{BULK TEMP} = \frac{666^\circ\text{F} + 529^\circ\text{F}}{2} = 597.5^\circ\text{F}$$

$$\text{WALL TEMP} = 515^\circ\text{F}$$

$$\begin{aligned} \mu_{600^\circ\text{F}} &= 0.735 \text{ CENTISTOKES} \\ \mu_{515^\circ\text{F}} &= 1.100 \text{ CENTISTOKES} \end{aligned} \quad \left. \vphantom{\begin{aligned} \mu_{600^\circ\text{F}} &= 0.735 \text{ CENTISTOKES} \\ \mu_{515^\circ\text{F}} &= 1.100 \text{ CENTISTOKES} \end{aligned}} \right] \text{A.P.I., 11A4.2}$$

CHECK DENSITY OF OIL AT BULK & WALL TEMP

$$\rho = 7.68489 \frac{\text{LB}_m}{\text{GAL}} \quad @ \quad 60^\circ\text{F}$$

USE API CHART GA3.3

$$\rho = 7.68489 \frac{\text{LB}_m}{\text{GAL}} \mid \frac{0.119826 \text{ (gm)}}{\text{(cm)}} \frac{\text{(GAL)}}{\text{(LE)}}$$

$$\rho = 0.92084 \frac{\text{gm}}{\text{ml}} \quad @ \quad 60^\circ\text{F}$$

$$\rho = 0.731 \text{ g/ml} \quad @ \quad 600^\circ\text{F}$$

$$\rho = 0.763 \text{ g/ml} \quad @ \quad 515^\circ\text{F}$$

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$$\mu = \frac{0.735 \text{ CENTISTOKES}}{(1.00)} \times \frac{0.01 \text{ STROKE}}{1 \text{ CENTISTROKE}} \times \frac{1 \text{ CM}^2}{\text{STROKE}} \times \frac{(0.763)}{0.73} \times \frac{\text{g}}{\text{ML}} \rightarrow$$

$$\rightarrow \frac{1 \text{ ML}}{1 \text{ CM}^3} \times \frac{1 \text{ lbm}}{454 \text{ gm}} \times \frac{2.54 \text{ CM}}{1 \text{ INCH}} \times \frac{12 \text{ INCH}}{1 \text{ FT}} \times \frac{3600 \text{ SEC}}{1 \text{ HOUR}}$$

$$\mu = 1.298573 \frac{\text{lbm}}{\text{FT-HR}} \quad ; \quad \mu_{515^\circ\text{F}} = 1.844107 \frac{\text{lbm}}{\text{FT-HR}}$$

$$Re = \frac{D_e G_s}{\mu} = \frac{(0.243953 \text{ FT}) (198402.9785 \text{ (LBm) (FT) (HR)})}{(1.298573 \text{ (LBm)})}$$

$$Re = 37,272,45713 \quad \text{SHELL SIDE}$$

$$\text{SPECIFIC HEAT OF OIL, } C = 0.745 \frac{\text{BTU}}{(\text{LBm})(^\circ\text{F})} \quad \text{API, FIG 7 E2.2}$$

$$\text{THERMAL CONDUCTIVITY, } K = 0.050 \frac{\text{BTU}}{(\text{FT})(\text{HR})(^\circ\text{F})} \quad \text{API, FIG 12 A 3.1}$$

$$\frac{h_o D_e}{K} = 0.36 \left(\frac{D_e G_s}{\mu} \right)^{0.55} \left(\frac{C \mu}{K} \right)^{1/3} \left(\frac{\mu}{\mu_w} \right)^{0.14} \quad \text{KERN, p.137}$$

$$\frac{h_o D_e}{K} = (0.36) \left(\frac{D_e G_s}{\mu} \right)^{0.55} \left(\frac{(0.745 \text{ BTU}) (1.298573 \text{ (LBm) (FT) (HR) (}^\circ\text{F)})}{(\text{LBm})(^\circ\text{F}) (\text{FT})(\text{HR}) (0.050 \text{ (BTU)})} \right)^{1/3} \rightarrow$$

$$\rightarrow \left(\frac{1.298573 \text{ (lbm) (FT) (HR)}}{1.844107 \text{ (FT) (HR) (LBm)}} \right)^{0.14}$$

$$\frac{h_o D_e}{K} = (0.36) (37,272,45713)^{0.55} (19.348693)^{1/3} (0.7041744)^{0.14}$$

$$\frac{h_o D_e}{K} = 300.69418$$

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$$\frac{h_o D_e}{k} = 300.69418$$

$$h_o = \frac{(300.69418)(k)}{D_e} = \frac{(300.69418)(0.050)(\text{BTU})}{(\text{FT})(\text{HR})(\text{OF})(0.243953)(\text{FT})}$$

$$h_o = 61.629532 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(\text{OF})}$$

ESTIMATE NUMBER OF EXCHANGER TUBES

7	3
6	8
5	11
4	12
3	13
2	14
1	5
0	6
1	15
2	14
3	13
4	12
5	11
6	8
7	3

168 TUBES

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10/15DETERMINE INSIDE HEAT TRANSFERCOEFFICIENT, USE FIG. 24, KERN, P. 834

$$\text{BOILING COEFFICIENT} = (1.25) \text{ (FIG. 24)} \\ \text{(KERN, P. 471.)}$$

DETERMINE TUBE SIDE REYNOLD'S NUMBER

$$Re = \frac{DG}{\mu} =$$

$$D = \frac{2'' - (2)(0.2'')}{12 \text{ IN}} = 0.13333 \text{ FT.}$$

$$G = \frac{(54)(10^6) \text{ BTU}}{\text{HR}} \left| \frac{\text{LBW}}{715,6142 \text{ BTU}} \right| \left| \frac{4}{168 \text{ TUBES}} \right| \left| \frac{1}{\pi (0.13333 \text{ FT})^2} \right|$$

$$G = (3.2168)(10^4) \frac{\text{LBW}}{\text{FT}^2 \cdot \text{HR}} = 32169.045 \frac{\text{LBW}}{\text{FT}^2 \cdot \text{HR}}$$

$$\mu \Rightarrow \frac{1}{\mu} = 2.1482 \left[(t - 8.435) + \sqrt{8078.4 + (t - 8.435)} \right]^{-1.20}$$

PERRY, 4TH ED., TABLE 3-267, $t = ^\circ\text{C}$, $\mu = \text{POISES}$.

$$t = 499 ^\circ\text{F} = 259.62778 ^\circ\text{C}$$

$$\frac{1}{\mu} = 592.7278 \quad ; \quad \mu = (1.007325)(10^{-3}) \text{ POISES}$$

$$\mu = 0.1007 \text{ centipoises}$$

$$\mu = 0.243773 \frac{\text{lbw}}{\text{FT} \cdot \text{HR.}}$$

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$$Re = \frac{DG}{\mu} = \frac{(0.13333 \text{ FT}) (32169.045 \text{ LB}_m) (\text{FT}) (\text{HR})}{(\text{FT}^2) (\text{HR}) (0.243773) (\text{LB}_m)}$$

$$Re = 17595.10$$

$$L/D = \frac{20 \text{ FT.}}{0.13333 \text{ FT}} = 150$$

$$j_H = 65, \text{ KERN, FIG. 24, p. 834.}$$

$$j_H = \frac{h_L D}{k} \left(\frac{c_p \mu}{k} \right)^{-1/3} \left(\frac{\mu}{\mu_w} \right)^{-0.14}$$

$$k = 0.376 \frac{\text{BTU}}{(\text{FT})(\text{HR})(^\circ\text{F})} \quad @ \quad 420^\circ\text{F}$$

$$k = 0.675 \frac{\text{BTU}}{(\text{FT})(\text{HR})(^\circ\text{F})} \quad @ \quad 620^\circ\text{F}$$

PERRY
4TH ED.
p. 3-204

BY INTERPOLATION :

$$k = 0.494105 \frac{\text{BTU}}{(\text{FT})(\text{HR})(^\circ\text{F})} \quad @ \quad T = 495^\circ\text{F}$$

$$C = 1.8 \frac{\text{BTU}}{\text{lb}_m ^\circ\text{F}}, \text{ KEENAN \& KEYES, 1969, P. 121,}$$

$$D = 0.13333 \text{ FT.}$$

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$$j_H = \frac{(h_i)(0.113333)(FT)(FT)(HR)(^{\circ}F)}{(0.4941)(BTU)} \frac{(1.8)(BTU)(0.243773)(lbm)}{(lbm)(^{\circ}F)(FT)(HR)(0.4941)(BTU)}$$

$$\rightarrow (1)^{-0.14}$$

$$j_H = \frac{h_i (0.280233)(FT^2)(HR)(^{\circ}F)}{BTU}$$

$$h_i = \frac{65 \quad BTU}{0.280233 \quad FT^2 \quad HR \quad ^{\circ}F}$$

$$h_i = 231.94989 \frac{BTU}{FT^2 \quad HR \quad ^{\circ}F}$$

Times 1.25 for BOILING

$$h_i = 289.9373684 \frac{BTU}{(FT^2)(HR)(^{\circ}F)}$$

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DETERMINE OVERALL HEAT TRANSFER COEFFICIENT
BASED ON OUTSIDE AREA.

$$U_o = \frac{1}{\frac{D_o}{D_i h_{di}} + \frac{D_o}{D_i h_i} + \frac{X_w D_o}{k_m \bar{D}_L} + \frac{1}{h_o} + \frac{1}{h_{do}}}$$

$$\frac{1}{h_{di}} = 0.001, \text{ KERN, P. 845 TABLE 12}$$

BOILER FEED WATER

$$D_o = 2'' = 0.1666667 \text{ FT}$$

$$D_i = 1.6'' = 0.133333 \text{ FT}$$

$$\frac{D_o}{D_i h_{di}} = \frac{(0.166667)(\text{FT})(0.001)(\text{FT}^2)(\text{HR})(^{\circ}\text{F})}{(0.133333)(\text{FT})(\text{BTU})} = (1.25)(10^{-3})$$

$$\frac{D_o}{D_i h_{di}} = (1.25)(10^{-3}) \frac{(\text{FT}^2)(\text{HR})(^{\circ}\text{F})}{(\text{BTU})} \leftarrow \textcircled{1}$$

$$\frac{D_o}{D_i h_i} = (1.25) \frac{1}{289.9374(\text{BTU})} \frac{(\text{FT}^2)(\text{HR})(^{\circ}\text{F})}{(\text{BTU})}$$

$$\frac{D_o}{D_i h_i} = (4.311275938)(10^{-3}) \leftarrow \textcircled{2}$$

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$$\frac{X_w D_o}{K_m \bar{D}_L} =$$

$$X_w = 0.2 \text{ IN} = 0.016666 \text{ FT}$$

$$K_m = \frac{26 \text{ (BTU)}}{(\text{FT})(\text{HR})(^\circ\text{R})}, \text{ KERN, P. 799, TABLE 3,}$$

$$D_o = 2.0 \text{ INCHES} = 0.1666667 \text{ FT.}$$

$$\bar{D}_L = \frac{D_o - D_i}{\ln(D_o/D_i)} = 0.1493807 \text{ FT.}$$

$$\frac{X_w D_o}{K_m \bar{D}_L} = \frac{(0.0166667)(\text{FT})(0.1666667)(\text{FT})(\text{FT})(\text{HR})(^\circ\text{F})}{26(\text{BTU})(0.1493807)(\text{FT})}$$

$$\frac{X_w D_o}{K_m \bar{D}_L} = (7.152036883)(10^{-4}) \frac{\text{FT}^2 \text{ HR } ^\circ\text{F}}{\text{BTU}}$$

$$\frac{1}{h_o} = \frac{1}{61.629532 \text{ BTU}} \frac{\text{FT HR } ^\circ\text{F}}{\text{BTU}}$$

$$\frac{1}{h_o} = 0.0162259872 \leftarrow (4)$$

$$\frac{1}{h_{d_o}} = 0.003 \frac{\text{FT}^2 \text{ HR } ^\circ\text{F}}{\text{BTU}} \leftarrow (5)$$

KERN, P. 846, TABLE 12,
CRACKING UNIT, GAS OIL FEED, 5000[°]F

$$U_o = \frac{1}{(1) + (2) + (3) + (4) + (5) + \dots}$$

$$U_o = \frac{1}{(1.25)(10^{-3}) + (4.3113)(10^{-3}) + (7.15204)(10^{-4}) + 0.01622 + 0.003}$$

$$U_o = \frac{1}{0.025503} = 39.211893 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^{\circ}\text{F})}$$

$$Q = U_o A \Delta T_{Lm} ; A = Q / (U_o \Delta T_{Lm})$$

$$Q = 54 \text{ MM BTU / HR}$$

$$\Delta T_{Lm} = 79.7997^{\circ}\text{F}$$

$$A = \frac{(54)(10^6)(\text{BTU})(\text{FT}^2)(\text{HR})(^{\circ}\text{F})}{(\text{HR})(39.211893)(\text{BTU})(79.7997)(^{\circ}\text{F})}$$

$$A = 17,257.373 \text{ FT}^2 \text{ REQUIRED}$$

A4. INTERNAL EXCHANGER, OVERALL U

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KAUFMAN (SECTION 3.2.1, PP. 6-7) PROPOSES AN EXTERNAL HEAT TRANSFER COEFFICIENT FOR THAYER-TYPE SERVICE,

$$h_o = \frac{600 \text{ BTU}}{(\text{FT}^2)(\text{HR})(\text{OF})}, \text{ WHEN SUBSTITUTED}$$

INTO THE EQUATION FOR OVERALL HEAT TRANSFER COEFFICIENT THE

$$\text{RESULT IS, } U_o = \frac{91.381 \text{ BTU}}{(\text{FT}^2)(\text{HR})(\text{OF})}.$$

THE RESULTING TUBE AREA IS 7405.2 FT² OR 14,143 FEET OF 2-INCH O.D. TUBE.

18 MAY 1987

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TRAY 13 PUMPAROUND EXCHANGER CALCULATIONS
BASED ON INFORMATION GLEANED FROM
WALDSMITH'S ARTICLE. (CEP, 9/80, pp 65-69.)

SHELL, NORCO, DU-S, GENERATES AN ADDITIONAL
54 MM BTU/HR, 660 PSIG STEAM.

ADDITIONAL 26,500 FT² SURFACE ADDED
FOR 660 PSIG STEAM

STEAM PRESSURE = 660 PSIG = 674.7 psia

$t_{\text{STEAM}} = 498.94^{\circ}\text{F}$

$\Delta H_{\text{VAP}} = 715.6142 \text{ BTU/LB}_m$

STEAM FLOW RATE (SATURATED STEAM)

$$= \frac{(54)(10^6) \text{ BTU}}{\text{HOUR}} \bigg| \frac{\text{LB}_m}{715.6142 \text{ BTU}}$$

$$= 75,459.7 \frac{\text{LB}_m \text{ STEAM}}{\text{HOUR}}$$

THIS WOULD BE THE STEAM FLOW RATE
IF ALL 660 PSIG STEAM IS GENERATED
AT TRAY 13

15 MAY 1987

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PRELIMINARY MATERIAL BALANCE ON SHELL
DU-5 ATMOSPHERIC CRUDE COLUMN,
NORCO, LA., BASED ON WALDSMITH, CHEMICAL
ENGINEERING PROGRESS, 9/80, pp. 65-69,
AND CRUDE OIL DATA, OIL AND GAS
JOURNAL,

CRUDE OIL FEED RATE = 200,000 $\frac{\text{BBL}}{\text{DAY}}$

COMPONENT	TBP RANGE (°F)	VOL %	FLOW RATE BBL/DAY
C ₄ & LIGHT		1.0	2,000.0
LIGHT NAPHTHA	57-310	11.7	23,400.0
HEAVY NAPHTHA	310-375	5.8	11,600.0
DIESEL	375-680	37.7	75,400.0
WIDE RANGE GAS OIL	680-ASPHALT	33.5	67,000.0
ASPHALT		10.3	20,600.0
			RESIDUUM, 680°F ⁺ 87,600.

ALL RESIDUUM, (680°F⁺) FLOWS TO VACUUM
DISTILLATION = 87,600 BBL/DAY

FLOWRATE OF MATERIAL UP COLUMN =
= 112,400 BBL/DAY = 4683.33 BBL/HR.

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SHELL AND TUBE EXCHANGER DESIGN

DETERMINE OIL FLOW RATE TO
PRODUCE $(54)(10^6)$ BTU/HR

$$\dot{m} = (54)(10^6) \frac{\text{BTU}}{\text{HR}} \left| \frac{\text{LB}_m}{80 \text{ BTU}} \right.$$

$$\dot{m} = 675,000 \frac{\text{LB}_m}{\text{HR}}$$

$$\dot{m} = 50,191.2 \frac{\text{BBL}}{\text{DAY}} = 2091.3 \frac{\text{BBL}}{\text{HR}}$$

ΔH_{VAP} OIL
NELSON, 1949,
p. 142

27 MAY 1987

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FROM TBP CURVE IT APPEARS THAT
2% OF CRUDE HAS A BOILING
POINT BETWEEN 680°F AND 666°F ,
(666°F MENTIONED BY WALDSMITH AS
TEMP USED FOR GENERATION OF 660PSIG
STEAM)

OIL FLOWRATE THROUGH TRAY 13 PUMP AROUND EXCHANGER.

EXCHANGER WILL BE SIZED TO CONDENSE
THIS 2% OF TOTAL FLOW.

$$\frac{0.02 \times 200,000 \text{ BBL/DAY}}{24 \text{ HOURS}} = 166,667 \frac{\text{BBL}}{\text{HR.}}$$

WIDE-RANGE GAS OIL = 22°A.P.I.

CHARACTERIZATION FACTOR = 11.84
(FROM O.E.G.J.)

MOLECULAR WEIGHT (A.P.I., FIG. 2B2.1) = 415

HEAT OF VAPORIZATION = 80 BTU/LB

(NELSON, 1949, FIG 41, P. 142)

$$\text{SP. GR. } \frac{60^{\circ}}{60^{\circ}} = \frac{141.5}{^{\circ}\text{API} + 131.5} = \frac{141.5}{22^{\circ}\text{API} + 131.5} = 0.921824$$

DENSITY OF H_2O @ 60°F = $0.9989705 \frac{\text{gm}}{\text{mL}}$

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$$\rho = \frac{0.921824 \text{ SP.GR} \times 0.9989205 \text{ gm}}{8.3452 \text{ (LB) (LITER)} \times \frac{1 \text{ kg}}{1000 \text{ g}}} \times \frac{1 \text{ kg}}{2.2046 \text{ lb}} \times \frac{1 \text{ L}}{0.26417 \text{ GAL}}$$

$$\rho = 7.68489 \frac{\text{LB}}{\text{GAL}}$$

$$\dot{M} = 166.667 \frac{\text{BBL}}{\text{HR}} \times \frac{42 \text{ GAL}}{1 \text{ BBL}} \times \frac{7.68489 \text{ LB}}{\text{GAL}}$$

$$\dot{M}_{\text{OIL}} = 53,794.31 \frac{\text{LB}}{\text{HR}}$$

$$\Delta H = \left(53,794.31 \frac{\text{LB}}{\text{HR}} \right) \left(80 \frac{\text{BTU}}{\text{LB}_m} \right) = 4,303,544 \frac{\text{BTU}}{\text{HR}}$$

DUTY REQUIRED TO CONDENSE WIDE
RANGE GAS OIL IN TRAY 13
PUMPAROUND

STEAM FLOW RATE BASED ON TRAY 13
 DUTY

$$\dot{M}_{\text{STEAM}} = 4,303,544 \frac{\text{BTU}}{\text{HR}} \times \frac{1 \text{ lb}_m \text{ STEAM}}{715.6142 \text{ BTU}}$$

$$\dot{M}_{\text{STEAM}} = 6013.777 \frac{\text{lb}_m \text{ STEAM}}{\text{HR}}$$

18 JUNE 1987

E. ALTHAUS

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DETERMINATION OF OVERALL HEAT TRANSFER
COEFFICIENT, U_o , FOR A THAYER-TYPE
EXCHANGER ON TRAY 13.

FOR A SHELL AND TUBE EXCHANGER,

$$h_o = 61.629532 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$$

GIVING $U_o = 39.211893 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$

FROM KAUFMAN, SECTION 3.2.1, (PP. 6-7)

Try $h_o = 600 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$

$$\frac{1}{h_o} = \frac{1}{600} = (1.666667)(10^{-3}) \frac{(\text{FT}^2)(\text{HR})(^\circ\text{F})}{\text{B.T.U.}}$$

TERM 4, PAGE 21 OF CALCULATIONS.

$$U_o = \frac{1}{(1.25)(10^{-3}) + (4.3113)(10^{-3}) + (7.15204)(10^{-4}) + (1.66667)(10^{-3}) + 0.003}$$

$$U_o = \frac{1}{0.0109431}$$

$$U_o = 91.381397 \frac{\text{BTU}}{(\text{FT}^2)(\text{HR})(^\circ\text{F})}$$

$$Q = U_o A \Delta T_{Lm} ; \quad A = Q / (U_o \Delta T_{Lm})$$

$$A = \frac{(54)(106)(\text{BTU})(\text{FT}^2)(\text{HR})(\text{OF})}{(\text{HR})(91.381397)(\text{BTU})(79.7997)(\text{OF})}$$

$$A = 7405.1644 \text{ FT}^2$$

USING 2" O.D. TUBES.

$$\text{CIRCUMFERENCE} = 0.5235988 \text{ FT.}$$

14,142.8223 FT OF 2" TUBE REQUIRED

USING 30-FT. LONG TUBES,
471.43 TUBES REQUIRED.

FOR 472 TUBES LAID HORIZANTALLY ON
3-INCH CENTERS,

$$\frac{472 \text{ TUBES} \times 3 \text{ INCHES}}{\text{TUBE}} \times \frac{1 \text{ FT}}{12 \text{ INCHES}} = 118 \text{ FEET}$$

THEREFORE, THE THAYER EXCHANGER
WOULD REQUIRE FOUR OR FIVE
LAYERS OF TUBES ON TRAY 13.

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DETERMINE HEIGHT OF THAYER TUBE
BUNDLE ON TRAY 13.

$$(3'')^2 = x^2 + x^2; \quad 9 \text{ in}^2 = 2x^2;$$

$$x^2 = 4.5 \text{ in}^2; \quad x = 2.12''$$

FOR FOUR ROWS OF TUBES,
BUNDLE THICKNESS = 8.4 INCHES

FOR FIVE ROWS OF TUBES,
BUNDLE THICKNESS = 10.5 INCHES.

A5. Tube Wall Thickness

TWO-INCH OUTSIDE-DIAMETER TUBES HAVING 0.2-INCH TUBE WALL THICKNESS WERE CHOSEN FOR BOTH INTERNAL AND EXTERNAL HEAT EXCHANGER DESIGN. THIS DECISION WAS BASED ON WALL THICKNESS CALCULATIONS AT 685 PSIG. AND CARBON STEEL PIPE STRESS CORRESPONDING TO A WALL THICKNESS OF 850 °F. ACTUAL OPERATING CONDITIONS WILL BE 499 °F AND 660 PSIG.

PRESSURE, DESIGN = 700 PSIA

CARBON STEEL TUBES,

MAXIMUM STRESS = 7900 PSI,

@ MAX TUBEWALL TEMP = 850 °F

(RASE, REACTOR DESIGN, I, P. 440)

PIPE THICKNESS EQUATION, (RASE, PLANT DESIGN, P. 395)

$$t = \frac{PD}{2S} + C$$

P = INTERNAL PRESSURE, PSIG, = 685 PSIG

S = ALLOWABLE WORKING STRESS, PSI

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$$S = 7500 \text{ PSI}$$

$$D = \text{O.D. INCHES, TRY } 2" \text{ \& } 3"$$

$$t = \text{THICKNESS OF PIPE, INCHES}$$

$$C = \text{CORROSION ALLOWANCE} = 0.10 \text{ INCH}$$

FOR 2" TUBE

$$t = \frac{(685)(\text{PSIG})(2)(\text{INCHES})}{(2.0)(7500)(\text{PSI})} + 0.10 \text{ IN}$$

$$t = 0.18671 \text{ INCHES.}$$

FOR 3" TUBE:

$$t = \frac{(685)(\text{PSIG})(3)(\text{INCHES})}{(2.0)(7500)(\text{PSI})} + 0.10 \text{ IN}$$

$$t = 0.2301 \text{ INCHES}$$

USE TUBE THICKNESS = 0.2"

A6. MECHANICAL CONFIGURATION, ON-TRAY TUBES

E. Althaus
(7-2-87)

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TWENTY PER CENT OF THE AREA OF TRAY 13 WAS ALOCATED TO DOWNCOMERS, DISENGAGING ZONES, AND DISTRIBUTING ZONES. THIS LEAVES AN AREA 12.86 FEET WIDE ON EITHER SIDE OF TRAY 13 FOR TUBE BUNDLES. U-TUBES WILL BE USED WITH LENGTHS RANGING FROM 11.5 TO 17.5 FEET FROM TUBE SHEET TO VESSEL CENTERLINE. FIFTY-ONE TUBES WILL FIT ON EACH ROW. FIVE ROWS WILL BE REQUIRED IN EACH TUBESHEET. 255 SINGLE TUBES OR 128 U-TUBES WILL BE PLACED ON EACH TUBESHEET.

THE PROPOSED TRAY-13 LAYOUT HAS BEEN SKETCHED ON THE FOLLOWING PAGE

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TRAY 13 LAYOUT

42.381 30 SHEETS 3 SQUARE
42.382 100 SHEETS 3 SQUARE
42.383 200 SHEETS 3 SQUARE
NATIONAL

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TRAY 13 LAYOUT,

TRAY 13 DIAMETER = 35.85 FT.

10% OF AREA TO BE USED FOR DOWNCOMERS

10% OF AREA USED FOR DISTRIBUTING AND
DISENGAGING.6% OF AREA USED FOR SUPPORTS
(PERRY, 4TH ED., P. 18-7)

TOTAL DOWNCOMER AREA

$$(0.10) \left(\frac{\pi}{4} \right) (35.85 \text{ FT})^2 = 100.94 \text{ FT}^2$$

FROM TRAY 14 TO CENTER OF TRAY 13,
LENGTH OF CENTRAL DOWNCOMER = 35.85 FT.

$$\text{WIDTH: } \left(\frac{1}{2} \right) (100.94 \text{ FT}^2) \left(\frac{1}{35.85 \text{ FT}} \right) = 1.41 \text{ FT.}$$

OR 0.705 EACH SIDE OF CENTER LINE.

FROM EDGES OF TRAY 13 TO TRAY 12
2 DOWNCOMERS, AREA EACH = 25.235 FT²

MENSURATIONAL FORMULAS, PERRY, P. 2-6

$$A (\text{SEGMENT}) = \frac{1}{2} r^2 (\theta - \sin \theta)$$

$$\theta - \sin \theta = \frac{A}{\frac{1}{2} r^2} = \frac{25.235 \text{ FT}^2}{\left(\frac{1}{2} \right) (17.925 \text{ FT}^2)} = 1.41$$

$$\theta - \sin \theta = 0.157077$$

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Θ (RAD)	$\Theta - \sin \Theta$
1	0.15853
0.9	0.11667
0.99	0.15397
0.999	0.15807
0.995	0.15624
0.996	0.156696
0.997	0.15715

USE $\Theta = 0.997$ RAD.LENGTH OF CORD, $L = 2r \sin\left(\frac{\Theta}{2}\right)$

$$L = (2)(17.925 \text{ FT}) \sin\left(\frac{0.997 \text{ RAD}}{2}\right)$$

$$L = 17.14 \text{ FT}$$

DISTANCE FROM CENTERLINE TO DISCHARGE WEIR

$$d = \frac{1}{2} \sqrt{4r^2 - L^2}$$

$$d = \frac{1}{2} \sqrt{(4)(17.925 \text{ FT.})^2 - (17.14 \text{ FT})^2}$$

$$d = 15.7435 \text{ FT.}$$

WIDTH OF EXIT DOWNCOMER =

$$17.925 \text{ FT} - 15.7435 \text{ FT} = 2.181 \text{ FT.}$$

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2 DISTRIBUTING AREAS, 2 DISENGAGING AREAS,

$$\text{TOTAL} = 100.94 \text{ FT}, \quad \text{AREA EACH} = \frac{\text{TOTAL}}{4}$$

$$\text{AREA EACH} = 25.235 \text{ FT}^2$$

WIDTH OF EACH DISTRIBUTING AREA.

$$= \frac{25.235 \text{ FT}^2}{35.85 \text{ FT}} = 0.7039 \text{ FT}$$

WIDTH OF EACH DISENGAGING AREA:

$$\frac{25.235 \text{ FT}^2}{17.14 \text{ FT}} = 1.472 \text{ FT.}$$

WIDTH OF EACH TUBE BUNDLE

17.925 FT - TOWER RADIUS

- 0.705 FT - $\frac{1}{2}$ CENTRAL DOWNCOMER

- 2.181 FT. EXIT DOWNCOMER

- 0.7039 FT. DISTRIBUTING AREA

- 1.472 FT DISENGAGING AREA

12.86 FT.

TUBES WILL RANGE IN LENGTH FROM 11.5-17.5 FT.

AVERAGE TUBE LENGTH = 14.5 FEET

TOTAL NUMBER OF TUBES REQUIRED =

$$\frac{14143 \text{ FT}}{14.5 \text{ FT}} = 975.38 = \underline{\underline{976 \text{ TUBES}}}$$

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WIDTH OF EACH TUBE BANK = 12.86 FEET.

CENTER TO CENTER DISTANCE FOR TUBES = 3 INCHES

NO. OF TUBES PER TUBE ROW/TUBE SHEET

$$= \frac{12.86 \text{ FT}}{3 \text{ INCHES}} \left| \frac{12 \text{ INCHES}}{\text{FT}} \right| = 51.44 = 51 \frac{\text{TUBES}}{\text{ROW}} \frac{\text{PER}}{\text{SIDE}}$$

204 TUBES PER /ROW, ALL TUBE SHEETS.

NO. OF ROWS OF TUBES =

$$\frac{976 \text{ TUBES}}{204 \text{ TUBES/ROW}} = 4.78 = 5 \text{ ROWS}$$

EACH TUBE SHEET WILL HAVE 255 - 2" TUBES
ON 5 ROWS OR 128 U-TUBES.

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Cost Estimate Items	
External Heat Exchangers	On-Tray Heat Exchangers
1) Purchased <ul style="list-style-type: none"> - bundles + shells - tray 13 - pumps (2) + motors (2) - piping & valves, insulation - control valves 2) Instrumentation & Controls 3) Electrical equipment 4) Equipment installation 5) Indirect costs <ul style="list-style-type: none"> - engineering & supervision - construction expense - contractor's fee 6) Contingency (15%)	1) Purchased Equipment <ul style="list-style-type: none"> - bundles - modified tray 13 - manifold piping + nozzles - platforms - removal equip. - control valves 2) Increased tower height 3) Instruments & controls 4) Electrical equipment 5) Equipment installation 6) Indirect costs <ul style="list-style-type: none"> - engineering & supervision - construction expense - contractor's fee 7) Contingency (20%)
	Increased maintenance cost ?

Heat Exchanger Cost(External Shell & Tube units, 10 units, 17,257 ft²)

$$\text{MS Escalation Index, } I = \left(\frac{818.7}{301} \right) = 2.72$$

(1970)

$$\text{Shell (\$/unit)} = 3700 (1.00) (1.25) 2.72 = \$12,580$$

$$\text{Bundle (\$/unit)} = 8820 [0.79 + 0.052 + 0.05]^{(0.892)} (1.00) (2.72) = \$21,400$$

$$(\$/10\text{-units}) = 10 (12,580 + 21,400) = \$339,800$$

[Internal (on-tray) units, 4 units, 7405 ft²]

$$(\$/\text{unit}) = (1.75) (15,000) (0.892) (1.00) (2.72) (0.67) = \$42,670$$

$$(\$/4\text{units}) = 4 (42,670) = \$170,680$$

Tray 13:

$$(\$/\text{tray}) = (760 \text{ ft}^2) (25/\text{ft}^2) (1.00) \left(\frac{818.7}{599.4} \right)^{1.37} = \$96,030$$

(1979)

New Tray 13 (modified, more structural support, higher entrance and exit weirs, segmental baffles)

$$(\$/\text{Tray}) = (760) (25) (2.00) 1.37 = \$52,060$$

OIL FROM TOWER

660°F

HEAVY GAS OIL, $K=11.84$, 34°API

$$\rho = 37 \text{ Lbm/ft}^3$$

$$MW = 219 \text{ Lbm/LB-MOLE}$$

$$\dot{m} = (0.529)(10^6) \text{ Lbm/HR}$$

$$529,075 \text{ \# / HR}$$

$$\mu = 1.2986 \frac{\text{lbm}}{\text{FT-HR}} \quad (600^\circ\text{F})$$

OIL RETURNED TO TOWER

$T = 529^\circ\text{F}$

$$\dot{m} = 529,075 \text{ Lbm/HR.}$$

$$\mu = 1.8441 \frac{\text{Lbm}}{\text{FT-HR}}$$

BOILER FEED WATER

600 PSIG SAT'D LIQ., 499°F
614.7 PSIA

$$\dot{m} = 75,460 \text{ Lbm/HR, LIQUID}$$

$$\mu = 0.1007 \text{ cp} = 0.24783 \frac{\text{Lbm}}{\text{FT-HR}}$$

STEAM

600 PSIG. SAT'D STEAM., 499°F
614.7 PSIA

$$\dot{m} = 75,460 \text{ Lbm/HR, VAPOR}$$

$$\mu = 0.0204 \text{ cp, KERN, P. 825}$$

DETERMINING ECONOMIC DIAMETERS OF OIL LINES.

PERRY 4TH ED., NOMOGRAPH, FIG. 5-31

OIL FROM TOWER TO MANIFOLD

$$\rho = 37 \text{ LB}_m / \text{FT}^3$$

$$\text{FLOW RATE} = \frac{529,075 \cancel{\text{ LB}_m} / \text{HOUR}}{\cancel{\text{HOUR}} / 60 \text{ MIN}} \bigg/ 37 \text{ LB}_m / \text{FT}^3$$

$$\text{FLOWRATE} = 238.3221 \text{ FT}^3 / \text{MIN} = 1783 \text{ G.P.M.}$$

$$\text{ECONOMIC DIAMETER} = 10'' \text{ pipe } (10.25'' \frac{\text{I.D.}}{\text{ON CHART}})$$

10'' SCH. 40 OK, ID = 10.02''

FOR FLOW MANIFOLD

$$\text{FLOW} = \left(\frac{1}{2}\right) (238.3221 \text{ FT}^3 / \text{MIN}) = 119.161 \text{ FT}^3 / \text{MIN}$$

$$\text{ECONOMIC DIAMETER} = 7\frac{1}{4}'' \text{ I.D.}$$

$$\text{USE } 8'' \text{ SCH. 40 PIPE, ID} = 7.981''$$

FOR INDIVIDUAL HEAT EXCHANGER LINES.

$$\text{FLOW} = \left(\frac{1}{10}\right) (238.3221 \text{ FT}^3 / \text{MIN}) = 23.832 \text{ FT}^3 / \text{MIN}$$

$$\text{ECONOMIC DIAMETER} = 3.25''$$

$$\text{USE } 3'' \text{ SCH. 40 PIPE, ID} = 3.068''$$

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PRESSURE DROP, PERRY, 4TH ED. FIG. 5-26.

$$\frac{(\mu \text{ (CP)})^{0.16}}{\rho \text{ (lbm/FT}^3\text{)} (1 \text{ ATM})} = \frac{\left(1.2986 \frac{\text{lbm}}{\text{FT-HR}} \left| \frac{(\text{FT-HR})(\text{CP})}{2.42 \text{ lbm}} \right|^{0.16}}{37 \text{ lbm/FT}^3} = 0.0245$$

$\Delta P = 0.006 \text{ PSI/FT}$, 10" pipe

$\Delta P = 0.02 \text{ PSI/FT}$, 3" pipe

OIL PIPE PRESSURE DROP CALCULATIONS,
EXCLUDING EXCHANGER INTERNAL DP

PIPING COMPONENT	DIA (IN. NOM)	ΔP (PSI/FT)	LENGTH (FT)	ΔP (PSI)
VESSEL NOZZLE	10"	0.006	31'	0.186
✓ GATE	10"		7'	0.042
✓ ELBOWS	4-10"		18 each	0.432
✓ TEE'S	2-10"		12' ea -	0.144
✓ CONTROL V.	10"		330'	1.98
✓ TEE	10"		47' (HARD)	0.282
CONTRACT.	10" / 3"	0.02	10'	0.2
GATE	3"	0.02	2'	0.04
ENLARGE	3"	0.006	31'	0.186
CONTRACT	→ 3"	0.02	5'	0.10
✓ GATE	3"	0.02	2'	0.04
ENLARGE	3" → 10"	0.02	10'	0.2

- 3.066 PSI
511 equiv FT

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PIPING COMP.	DIA (INCHES, NOM.)	ΔP (PSI/FT)	LENGTH (FT)	ΔP (PSI)
TEE	10"	0.006	47	0.282
ELBOWS	4-10"		18' each	0.432
GATE	10"		7'	0.042
ENLARGE	10" →		31'	0.186
PIPE TO GROUND	10"		100'	0.60
PIPE TO VESSEL	10"	N	100'	0.60
EXCHANGER PIPE	3"	0.02	20'	0.4
				6.374
GATE (PUMP ISOLATION)	10", 2 ea	0.006	7'	0.084
				6.458

OIL LINE MATERIALS

✓ 10" GATE VALVES	—	6
✓ 10" ELBOWS		12
✓ 10" CONTROL VALVE		1
✓ 10" TEE'S		4
✓ 3" GATE VALVES		20
✓ 10" PIPE		250'
✓ 3" PIPE		300'

Pumps & Motors (piping $\Delta P = 6.5 + 10 = 16.5 \text{ psi}$, 1783 gpm)

$$\$(2, \text{pumps}) = (2) 2200 (1.38) (1.00) 2.72 = \underline{\$16,520}$$

$$\text{BHP} = \frac{\text{gpm}(\Delta P)}{1713 \text{ E}} = \frac{1783(16.5)}{1713(0.60)} = 28.6 \text{ say } 30 \text{ HP}$$

$$\$(2, \text{motors}) = 2(1100)(1.00)2.72 = \underline{\$5984}$$

Piping & Fittings Cost (incl d CV)

$$\underline{10" \text{ pipe}} = 250(7.0)(2.72) = \$4760$$

$$\underline{3" \text{ pipe}} = 300(1.50)(2.72) = \$1224$$

$$\underline{10" \text{ gvs}} = 6(550)(1.37) = \$4521$$

$$\underline{3" \text{ gvs}} = 20(110)(1.37) = \$3014$$

$$\underline{10" \text{ CV}} = (1)(1.50)(800)(1.37) = \$1644$$

$$\underline{10" \text{ Ells}} = (12)(75)(1.37) = \$1233$$

$$\underline{10" \text{ Tees}} = (4)(140)(1.37) = \$767$$

$$\Sigma_1 = \underline{\$17,170}$$

Insulation

$$\underline{10" \text{ pipe}} = 250(14)(1.37) = \$4795$$

$$\underline{3" \text{ pipe}} = 300(5.50)(1.37) = \$2261$$

$$\underline{14 \text{ Exchrs}} = 10(1000)(1.37) = \$13,700$$

$$\Sigma_2 = \underline{\$20,756}$$

On Tray Configuration

Increased Tower height (1.00 ft) + access ways.

$$\text{\$} = 2.50 (\text{\$}15.91 \text{ ft}^2) (113.1 \text{ ft}^2) 2.72 = \underline{\text{\$}12,236}$$

Service Platform w/hand rail and supports

$$\text{\$} = (1.75) (113.1 \text{ ft}) (\text{\$}1051 \text{ ft}) (1.37) = \underline{\text{\$}28,472}$$

ATTACHMENT 2

**FEASIBILITY OF APPLICATION OF THAYER
METHOD OF DISTILLATION IN 'CLEAN' SERVICE**

ATTACHMENT 2