TO: H. S. Dewey, Process Engineer

FROM: F. J. Newman, Power Plant Superintendent

In accordance with the instructions of Mr. L. P. McGillicudy, the following information on utilities is forwarded:

Steam

There are available 13,000 lb./hr. of 150 lb./sq. in. gauge steam @ 45¢/1,000 lb. and 1,500 lb./hr. of 15 lb./sq. in. gauge steam @ 15¢/1,000 lb. Steam requirements above those available will necessitate the erection of additional steam-producing facilities at the rate listed in Table 2. Steam which comes in direct contact with propane or tallow cannot be returned to the boiler and must be exhausted to the atmosphere or fed to the sewer. Steam lost in this manner will be taken as a loss of 10¢/1,000 lb.

Cooling Water

Cooling water is available at a pressure such that pumps will not be required for circulating the water through process equipment. The cost of cooling water is 3.0¢/1,000 gal. for any amount required. The water is available at 90°F. and cannot be returned above 105°F.

City water @ 8¢/1,000 gal. should be used in direct contact with propane or tallow and after being used must be fed to the sewer.

Electrical Power

Electrical power is available at the rate of 1¢/kw.-hr. in any quantity required.

TABLE 2

COST OF STEAM IF ADDITIONAL FACILITIES ARE REQUIRED

(These figures include investment and operating costs for the new installation.)

Steam, lb./hr.	Cost/1,000 lb. 15 lb./sq.in.gauge	Cost/1,000 lb. 150 lb./sq.in.gauge
1,000	$0.78	$1.05
5,000	0.68	0.90
10,000	0.58	0.78
15,000	0.52	0.70
20,000	0.48	0.65
25,000	0.42	0.60

TO: H. S. Dewey, Process Engineer

FROM: W. L. Curtis, Accountant

At the request of Mr. L. P. McGillicudy I am sending you information regarding the economics of the proposed tallow-decolorization plant.

Fixed charges on per annum basis

Depreciation based on a 10-yr. pay-off will be 10% of constructed cost.

Miscellaneous taxes and insurance will be 1 1/2% of constructed cost.

Interest will be 4 1/2% of constructed cost.

Overhead will be 6% of constructed cost.

Labor, including pay-roll burden, has been estimated to be $95,000/yr.

Maintenance has been estimated to be $35,000/yr.

Revenues

Color bodies are estimated to be worth 1¢/lb.

SOLUTION

Robert P. Bannon, University of Illinois

PLANT MEMORANDUM

TO: L. P. McGillicudy, Chief Engineer

FROM: H. S. Dewey, Process Engineer

RE: PROPANE EXTRACTION PLANT FOR DECOLORIZATION OF TALLOW

Decolorized tallow can be produced by the propane extraction process for 0.305 cent/lb. The present caustic process costs 0.499 cent/lb. At your request a propane extraction plant has been designed to decolorize 8,600 lb./hr. of animal tallow producing 8,170 lb./hr. of decolorized tallow (Lovibond color of three) and 430 lb./hr. of color bodies. The costs of the proposed plant are as follows:

Total constructed cost	$139,110
Net annual cost	185,450
Annual fixed charges	30,600
Annual operating cost	186,850
Annual by-product revenue	32,000

A complete report on the design of the proposed plant is attached hereto.

The construction of this plant is strongly recommended.

INTRODUCTION

A liquid-liquid extraction process for the decolorization of animal tallow has been developed in the laboratory. The solvent for the new process is liquid propane under pressure. In order to determine whether the propane extraction method is more economical than the present caustic method (0.499 cent/lb.), a plant capable of treating 8,600 lb./hr. of raw tallow was designed and the cost estimated. The designed output is 8,170 lb./hr. of decolorized tallow (Lovibond color of three) and 430 lb./hr. of color bodies.

DESCRIPTION OF THE PROCESS

Raw tallow stored at 125°F. and atmospheric pressure (Figure 1) is pumped (P1) through exchanger E9, in which it is heated to 165°F., to the top baffle of a continuous countercurrent propane extraction tower (T1). The extraction tower is operated isothermally at 165°F. and 450 lb./sq.in.abs. tower-top pressure. Extract consisting of propane and refined tallow is removed as overhead, and raffinate composed of color bodies and propane comprise the bottoms.

Overhead Stream. The overhead stream is heated (E1) to 175°F. with 150-lb. steam. Slightly less than half the solvent in the overhead is vaporized here. This vapor is separated from the liquid in flash drum D1 (450 lb./sq.in.abs.) and is condensed in E2 with tallow and boiling propane from D1 or in E3 with cooling water. From the condenser the propane flows back to the storage tank. The liquid from D1 is throttled to 350 lb./sq.in.abs. and passed through exchanger E2, where most of the remaining solvent is vaporized. This mixture of liquid and vapor is separated in drum D2 (350 lb./sq.in.abs. and 152°F.). Vapor from D2 is condensed with cooling water in E5 at 350 lb./sq.in., flows through holding tank H2, and is pumped (P3) back to the propane storage tank. The liquid from drum D2 is throttled to 230 lb./sq.in.abs. and heated (E4) to 252°F. with 150-lb. steam. The vapor formed in E4 is separated from the tallow in flash drum D3. The vapor from D3 is condensed with cooling water (E6), flows through holding tank H1, and is pumped back to the propane storage tank. The refined-tallow stream from D3 is throttled down to 17 lb./sq.in.abs. and into steam stripper S1 where 150-lb. steam strips out the remaining solvent. The tallow from S1 flows by gravity and stripper pressure through exchanger E9, where it heats up raw tallow feed; through exchanger E10, where it is further cooled to its storage temperature with cooling water; and then to the refined-tallow storage tank. The mixture of steam and propane vapor from stripper S1 is separated in a jet condenser J1. The propane vapor from J1 is compressed (C1) to 230 lb./sq.in.abs., condensed (E6) with cooling water, and pumped (P4) to the storage tank.

Bottoms Stream. The bottoms stream of propane and color bodies from the extraction tower (T1) is throttled to 230 lb./sq.in.abs. and heated (E8) with 150-lb. steam to 237°F. The vapor formed is separated in flash drum D4. This vapor is also condensed in E6 and pumped (P4) back to the propane storage tank. The liquid from D4 is throttled down to 17 lb./sq.in.abs. and into steam stripper S2, where 150-lb. steam strips out the remaining solvent. The color bodies from S2 flow by gravity and stripper pressure through a cooler (E7) to storage. The mixture of steam and vapor from stripper S2 joins the stream from stripper S1 and goes through the jet condenser as described.

EQUIPMENT LIST

I. Extraction Tower T1
   Diameter: 8.5 ft.
   Height: 50.5 ft.
   No. of baffles: 26
   Operating temperature and pressure: 165°F. and 450 lb./sq.in.abs.
   Cost: $30,704.

II. Flash Drums

	D1	D2	D3	D4
Diameter, ft.	4.5	4.0	1.5	0.5
Height, ft.	21.0	9.0	11.5	6.0
Operating temp., °F.	175	152	252	237
Operating press., lb./sq.in.abs.	450	350	230	230
Cost, $	5,623	1,580	697	128

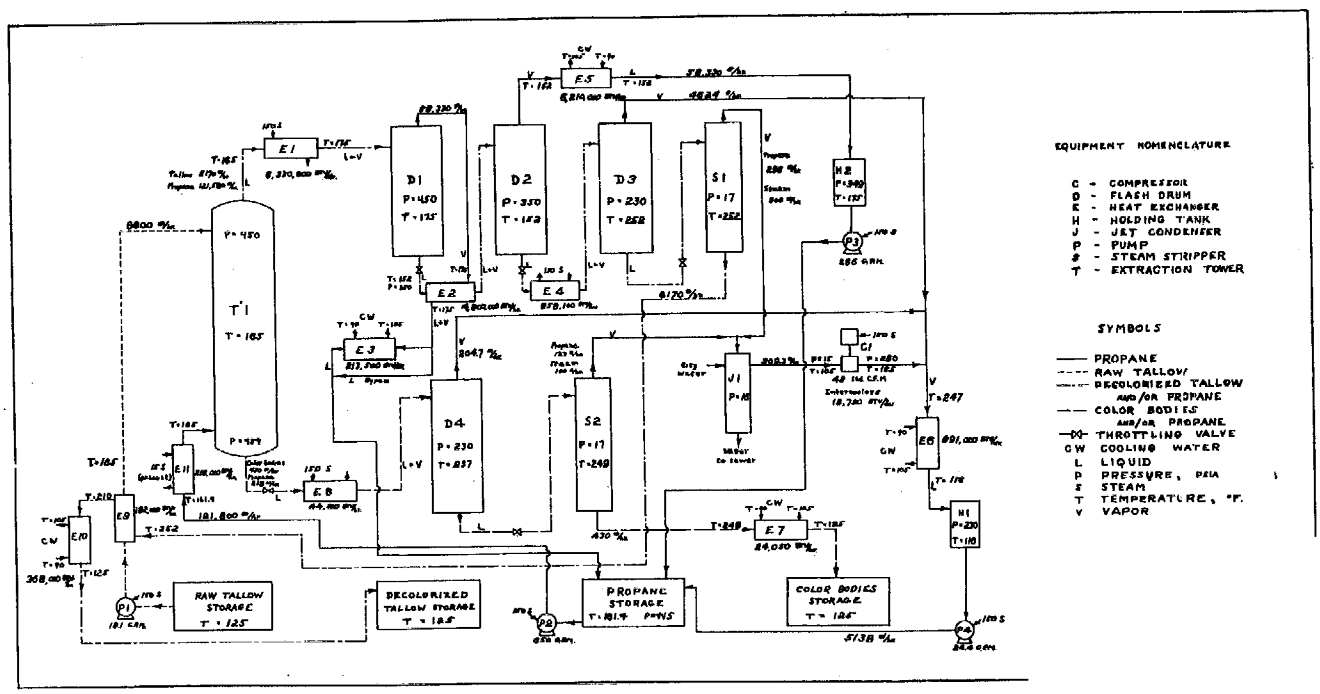


Fig. 1. Flow sheet for proposed tallow-decolorization plant.

III. Steam Strippers – Packed with 1-in. Raschig Rings

	S1	S2
Diameter, ft.	1.0	1.0
Height, ft.	25	25
Operating temp., °F.	252	249
Operating press., lb./sq.in.abs.	17	17
Cost, $	397.50	397.50

IV. Jet Condenser J1

Diameter: 0.5 ft.
Height: 17 ft.
Operating pressure: 16 lb./sq.in.abs.
Cost: $222

V. Heat Exchangers: Shell and Tube

Exchanger duty	B.t.u./hr.	ΔT, °F.	U, B.t.u./(hr.)(ft./°F.)	Area, ft.	Max. operating pressure, lb./sq.in.abs.	Cost, $
E1	6,330,000	Heating 196 Vapor. 191	202 204	162	450	2,485
E2	4,800,000	23	91.2	2,290	450	11,700
E3	213,500	77.5	171.5	16	450	350
E4	858,100	158.5	220	16	230	256
E5	6,210,000	54.5	171.5	665	350	3,840
E6	991,000	desup. 68 cond. 17.5	14.3 171.5	563	230	3,740
E7	24,050	73	42.9	7.7	15	109
E8	44,500	165	217	1.24	230	28
E9	182,000	86	25	85	450	1,820
E10	368,000	63.6	42.9	135	15	1,335
E11	219,000	87.5	107	23.4	450	512

VI. Compressor C1 - Steam-driven Reciprocating Type, Three Stages with Intercoolers

Capacity: 42.0 std.cu.ft./min.
Power delivered: 10.7 hp.
Operating range: 15 to 230 lb./sq.in.abs.
Cost: $1,070

VII. Pumps - Steam-driven Reciprocating Type

	P1	P2	P3	P4
Capacity, gal./min.	19.1	650	286	24.4
Pressure head, lb./sq.in.abs.	450	19	100	115
Power delivered, hp.	5.05	5.09	16.2	2.81
Cost, $	580	585	1,555	534

VIII. Holding Tanks

	H1	H2
Height, ft.	6.0	9.5
Diameter, ft.	2.0	5.0
Operating pressure, lb./sq.in.abs.	230	349
Cost, $	370	1,900

IX. Propane Storage Tank

Length: 25 ft.
Diameter: 9 ft.
Operating pressure: 445 lb./sq.in.abs.
Cost: Already paid out

Judging Committee comment: This item cannot be neglected; only tallow storage has been paid out.

COST SUMMARY

Total constructed cost		$139,110	
Annual investment cost		$ 30,600	
Annual operating costs:			
Utilities	$53,000		
Propane	3,850		
Labor	95,000		
Maintenance	35,000		
Total		$186,850	
Total annual cost		$217,450	
Annual revenue:			
Value of color bodies		$ 32,000	
Net annual cost		$185,450	
Net cost per pound of tallow produced			$0.00305
Net annual saving of extraction process over caustic process			$118,050

DISCUSSION OF THE PROCESS

The process design must necessarily start with the extraction tower. From the equilibrium curves it is seen that, as the temperature increases, the concentration differences between phases in equilibrium increases, and therefore fewer equilibrium stages would be needed for a given separation at 165° than at 160°F. or lower. For this reason the tower was designed for isothermal operation at 165°F. The tower-top temperature was necessarily fixed at 165°F. owing to miscibility considerations.

In a solvent-recovery system a large quantity of heat is necessary to vaporize the propane. In order to make operating costs as low as possible, it is necessary to recover as much heat from the propane as possible. The propane-recovery process is essentially an evaporation process with several complications. According to the equilibrium constant nomograph in reference 2, the lower the pressure, the higher is the temperature necessary for a given liquid-vapor equilibrium. This would seem to indicate that the multiple-effect evaporator principle could not be used; however, these equilibrium constants do not hold for solutions containing a substance of high and one of low molecular weight when the mole fraction of the high-molecular-weight substance is negligible (5). It is possible, therefore, to realize the heat economy of multiple-effect evaporation in the first two flash drums of the overhead propane-recovery system.

Slightly less than one half the propane in the overhead stream was vaporized at 450 lb./sq.in. abs. and 175°F. with 150 lb. steam, and this vapor was used as a heat source to vaporize most of the remaining solvent at 350 lb./sq.in.abs. and 152°F. (One-hundred-and-fifty-pound steam was used in E1 because it was cheaper than 15-pound steam for the quantity needed.) A little more propane was vaporized in E1 than was necessary to use in E2. This was done to permit better and much easier control of the process. It is a little more expensive to do this, but it greatly reduces the difficulty of control inherent in this type of design. An additional exchanger (E3) was provided to condense this excess vapor. The pressure for the second flash drum (D2) was set at 350 lb./sq.in. abs., which was considered to be the best pressure for a balance between a ΔT on the exchanger (E2) preceding the drum and the amount of propane that could be separated in the drum. Since the mole fraction of tallow in the liquid in drum D2 is appreciable, an equilibrium constant was used to determine the propane concentration in the liquid. A third flash drum was necessary to reduce the solvent concentration to the required value for entering stripper S1. This third drum was operated at as low a pressure as possible that would permit condensation of the vapor with cooling water. This low pressure resulted in the temperature in the drum being as low as possible, and steam was thereby saved. The bottoms stream was small and contained so little solvent that only one flash drum at 230 lb./sq.in.abs. was needed.

Holding or surge tanks were considered to be necessary between propane condensers and pumps. Without tanks at these locations surging would tend to decrease the effective area of the condenser because of flooding.

The vapor streams from the two strippers were combined so that only one jet condenser was needed. Practically all water introduced into the propane stream in the jet condenser will be removed in the intercoolers of compressor C1. However, over a considerable time there may be a build-up of water in the propane system. Since propane and water are slightly miscible, this water could not be separated out in holding tank H1.

The color bodies leaving stripper S2 were cooled to 125°F. before being sent to storage. This may or may not be necessary depending on subsequent processes and the holding time in storage.

APPENDIX

Discussion of Calculations

Only the calculations for the final design are included in this report. Several other designs, from which the final design evolved, were partially calculated.

Each piece of equipment is calculated separately and assumptions made are stated for the first of each type of equipment. General assumptions made are

1. Heat losses are negligible.
2. No tallow or color bodies are vaporized in any equipment or lost in any other way.

Thermodynamic data for propane were obtained from references 1 and 3 and vapor-liquid equilibrium constants from reference 2. Thermodynamic data for steam were obtained from steam tables.

DESIGN CALCULATIONS

I. Extraction tower

Material balances (S.F.B.)

Over-all balance: Feed = raffinate + extract

$$8{,}600 = R + 8{,}170$$

$$R = 430 \text{ lb./hr.}$$

Tallow balance $FX_F = RX_R + EY_E$

$$(8{,}600)(0.860) = 430X_R + (8{,}170 \times 0.894)$$

$$X_R = 0.214$$

Mole balance $N_F = N_R + N_E$

$$\frac{8{,}600}{750} = N_R + \frac{8{,}170}{740}$$

$$N_R = 0.4265 \text{ lb. mole/hr.}$$

$$\text{M.W. bottoms} = \frac{430}{0.4265} = 1{,}008$$

From Ponchon - Savarit diagram (Figure 2)

$$\frac{S_P}{R} = 283$$

$S_P = 283 \cdot 430 = 121{,}800$ lb./hr.

Solvent out in bottoms =
$430 \times 0.5057 = 218$ lb./hr.

Solvent out in overhead =
$121{,}800 - 218 = 121{,}580$ lb./hr.

Number of theoretical stages = 3.88

Number of baffles =

$$\frac{\text{No. of theoretical stages}}{\text{baffle efficiency}} = \frac{3.88}{0.15} = 25.9$$

$\therefore$ 26 baffles are needed.

Since tower is isothermal, no heating coils are necessary and the feed tray is the top tray.

Height = (No. of baffles - 1) 1.5
+ height of settling sections
= (26 - 1) 1.5 + 8 + 5 = 50.5 ft.

Since 450 lb./sq.in.abs. is the minimum allowable pressure, this will be the tower top pressure.

Tower bottom pressure = 450 + liquid head

$$= 450 + \text{height} \times \text{average density} \times \frac{1}{144}$$

Average density =

$$\frac{8{,}600\ 0.906 - 3\times10^{-4}(165-100)\ 62.4 + 121{,}800\times24.4}{8{,}600 + 121{,}800}$$

= 26.4 lb./cu.ft.

Tower bottom pressure

$$= 450 + 50.5 \times 26.4 \times \frac{1}{144} = 459 \text{ lb./sq.in.abs.}$$

Diameter

Volume of propane phase at top
= volume of propane + volume of extract
$= 121{,}580 \times 0.041$

$$= \frac{8{,}170}{62.4\left[0.905 - 3\times10^{-4}(165 - 100)\right]}$$

= 5,138 cu.ft./hr.

Volume of propane phase at propane entrance = $121{,}800 \times 0.041$ = 5,000 cu. ft. /hr.

Therefore, maximum superficial velocity will be at top of tower.

Maximum allowable superficial velocity = 90 ft./hr.

$$\text{Tower diameter} = 2\sqrt{\frac{\frac{5{,}138}{90}}{\pi}} = 8.5 \text{ ft.}$$

Cost: It is assumed that interpolation of cost data is permissible for in-between diameters.

Vessel cost	= 50.5 × 152 × 4.00	= $30,704
Baffle cost	= 26 × 180	= $ 4,680
Total tower cost		= $35,384

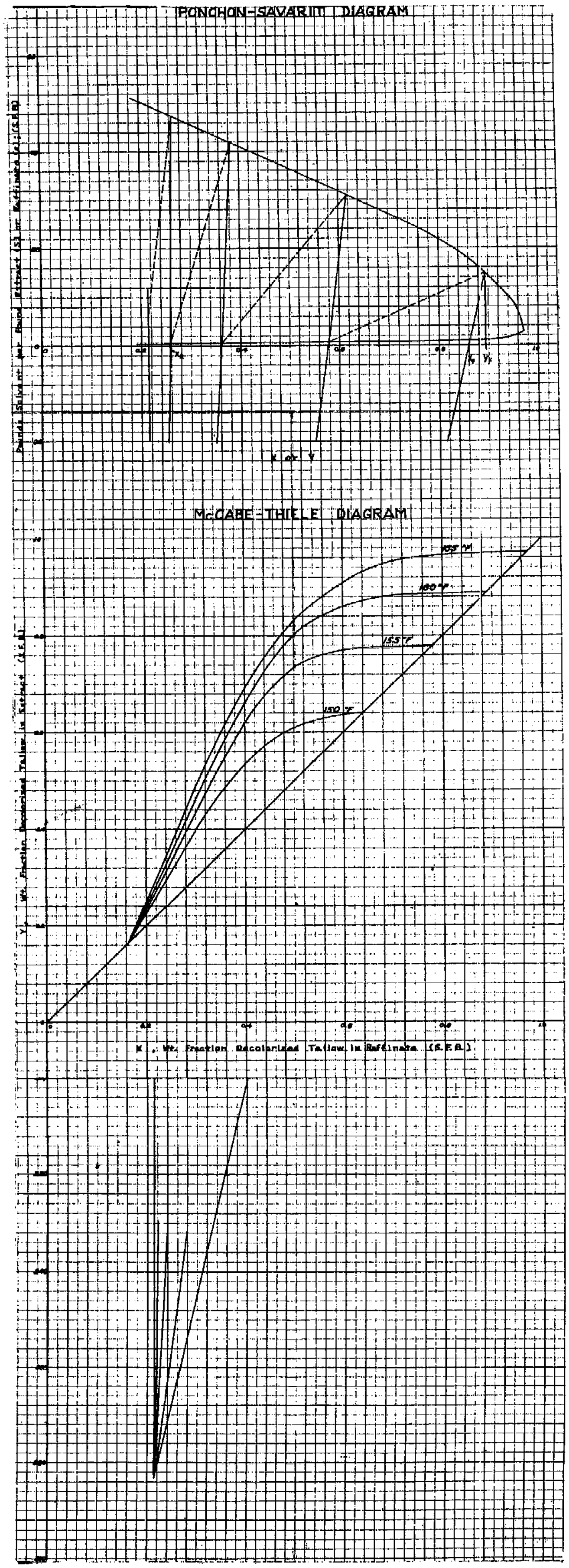


Fig. 2.

II. Overhead Flash Drums

Overhead stream

Tallow = 8,170 lb./hr. = 11.04 lb. moles/hr.
Propane = 121,580 lb./hr. = 2,760 lb. moles/hr.

It is desired to balance the amount of propane vaporized in exchanger E1 and E2 so that the vapor from flash drum D1 will be a little more than just sufficient to meet the heat requirements of E2.

Flash drum D2

Pressure = 350 lb./sq.in.abs.
Temperature = 152°F.
K = 1.10 from equilibrium constant nomograph (2); as the mole fraction of tallow is appreciable this K is assumed to hold.
Mole fraction propane in vapor = 1

Mole fraction propane in liquid $= \frac{1}{K} = 0.91$

Moles in exit liquid stream $= \frac{11.04}{1.0 - 0.91}$

= 123 lb. moles/hr.

Liquid propane leaving D2
= (123 - 11.04)44 = 4,920 lb./hr.
Propane vapor separated in D1 and D2
= 121,580 - 4,920 = 116,660 lb./hr.
x = propane vaporized in E1, lb./hr.
y = propane vaporized in E2, lb./hr.

Exchanger E2 duty
= y(408.4 - 322.5) + 4,920(301.8 - 322.5)
+ 8,170 × 0.53 (152 - 175)
= 85.9y - 201,600 B.t.u./hr.

Heat available from propane vapor from D1
= x(408.5 - 322.5) = 86.0x

Therefore:

$$86.0x = 85.9y - 201{,}600$$
$$x + y = 116{,}660$$
$$x = 57{,}160 \text{ lb./hr.}$$
$$y = 59{,}500 \text{ lb./hr.}$$

In order to provide an extra margin of heat in exchanger E2 for better control of the process, x and y will be made equal.

$$x = y = \frac{1}{2} \cdot 116{,}660 = 58{,}330 \text{ lb./hr.}$$

1. Flash drum D1

Pressure = 450 lb./sq.in.abs.
Temperature = 175°F., the boiling point of pure propane at 450 lb./sq.in.abs.

It is assumed that the presence of the tallow does not affect this temperature appreciably. Equilibrium constants do not hold when the mole fraction of the tallow is negligible, as is the case here. The solution is treated as if it were pure propane.

Exit streams:
Propane vapor = x = 58,330 lb./hr.
Propane liquid = 121,580 - 58,330
= 63,250 lb./hr.
Tallow = 8,170 lb./hr.

Diameter:
Vapor rate = 58,330 lb./hr. = 16.2 lb./sec.
Sp. vol. of vapor = 0.194 cu.ft./lb.
= 16.2 · 0.194 = 3.14 cu.ft./sec.

$$\rho G = \frac{1}{\text{sp.vol.}} = 5.15 \text{ lb./cu.ft.}$$

$$\rho \text{tallow} = [0.905 - 3\times10^{-4}(175 - 100)]\, 62.4 = 55.0 \text{ lb./cu.ft.}$$

$$\rho \text{propane liquid} = \frac{1}{0.0425} = 23.5 \text{ lb./cu.ft.}$$

$$\rho L_{avg.} = \frac{55.0\times 8{,}170 + 23.5\times 63{,}250}{8{,}170 + 63{,}250} = 27.1 \text{ lb./cu.ft.}$$

Judging Committee comment: Specific volumes are additive not densities.

$$D^2 = \frac{8 \times \text{cu.ft./sec.}}{0.227\pi\sqrt{\frac{\rho_L - \rho_G}{\rho_G}}} = \frac{8 \times 3.14}{0.227\,\pi\sqrt{\frac{27.1 - 5.15}{5.15}}}$$

D = 4.13 ft. to nearest $\frac{1}{2}$ ft. = 4.5 ft.

Height:

$$h_L = \frac{(8{,}170 + 63{,}250)1/12}{27.1 \times \pi \left(\frac{4.5}{2}\right)^2} = 13.8 \text{ ft.}$$

$$h = 2 + D \times 1.17 + h_L = 2 + 4.5 \times 1.17 + 13.8 = 21.0 \text{ ft.}$$

Cost:

Vessel cost = 21 × 64 × 4.00	= $5,375
Mist separator cost = 4 × 62 =	248
Total cost	= $5,623

2. Flash drum D2

Diameter:
Vapor rate = 58,330 lb./hr. = 16.2 lb./sec.
Sp. vol. of vapor = 0.270 cu.ft./lb.
= 0.270 × 16.2 = 4.38 cu.ft./sec.

$$\rho_G = \frac{1}{\text{sp.vol.}} = 3.70 \text{ lb./cu.ft.}$$

$$\rho_{\text{tallow}} = [0.905 - 3\times10^{-4}(152 - 100)]\, 62.4 = 55.4 \text{ lb./cu.ft.}$$

$$\rho_{\text{propane liquid}} = 26.0 \text{ lb./cu.ft.}$$

$$\rho_{L_{avg.}} = \frac{55.4 \times 8{,}170 + 26.0 \times 4{,}920}{8{,}170 + 4{,}920} = 44.3 \text{ lb./cu.ft.}$$

Judging Committee comment: Specific volumes are additive not densities.

$$D^2 = \frac{8 \cdot 4.38}{0.227\pi \sqrt{\frac{44.3 - 3.70}{3.70}}}$$

D = 3.85 ft.   to nearest $\frac{1}{2}$ ft. = 4.0 ft.

Height:

$$h_L = \frac{8{,}170 + 4{,}920}{44.3 \cdot \pi \left(\frac{4}{2}\right)^2} = 1.96 \text{ ft.}$$

h = 2 + 4.0 · 1.17 + 1.96 = 9.0 ft.

Cost:
Vessel cost = 9.0 × 56 × 2.70 = $1,360
Mist separator cost = 4 × 55 = $ 220
Total cost = $1,580

3. Flash drum D3

Pressure = 230 lb./sq.in.abs.
Entering stream:
Tallow = 8,170 lb./hr. = 11.04 lb. moles/hr.
Propane = 4,920 lb./hr. = 112 lb. moles/hr.
Exit streams:
Tallow = 8,170 lb./hr.

$$\text{Propane liquid} = 0.035 \times \frac{8{,}170}{1.0 - 0.035}$$

= 296 lb./hr.

Propane vapor = 4,920 - 296 = 4,624 lb./hr.

Mole fraction propane in exit liquid

$$= \frac{\frac{296}{44}}{\frac{296}{44} + 11.04} = 0.379$$

$$K = \frac{y^1}{x^1} = \frac{1}{0.379} = 2.64$$

∴ Temperature = 252°F. [equil. constant nomograph (2)]

Diameter:
Vapor rate = 4,624 lb./hr. = 1.285 lb./sec.
Sp. vol. vapor = 0.62 cu.ft./lb.
= 0.62 × 1.285 = 0.796 cu.ft./sec.

$$\rho_G = \frac{1}{\text{sp.vol.}} = 1.61 \text{ lb./cu.ft.}$$

$$\rho_L = \left[0.905 - 3\times10^{-4}(252 - 100)\right] 62.4$$
= 53.6 lb./cu.ft.

(Above propane critical temperature, density of solution is same as density of tallow.)

$$D^2 = \frac{8 \cdot 0.796}{0.227\pi \sqrt{\frac{53.6 - 1.61}{1.61}}}$$

D = 1.25 ft.   to nearest $\frac{1}{2}$ ft. = 1.5 ft.

Height:

$$h_L = \frac{(8{,}170 + 296)1/12}{53.6 \times \pi \left(\frac{1.5}{2}\right)^2} = 7.45 \text{ ft.}$$

h = 2 + (1.5 × 1.17) + 7.45 = 11.2 ft.
to nearest $\frac{1}{2}$ ft. = 11.5 ft.

Cost:
Vessel cost = 11.5 × 22 × 2.20 = $557
Mist separator = 4 × 35 = $140
Total cost = $697

III. Overhead Steam Stripper S1
Pressure = 17 lb./sq.in.
Entering stream at 252°F.
Tallow = 8,170 lb./hr.
Propane = 296 lb./hr.

It is assumed that the temperature change on throttling from the flash-drum pressure to the stripper pressure is negligible because of the small percentage of propane in the stream.

Heat balance:

The propane is above its critical temperature (206°F.) and therefore has no heat of vaporization.

Approximate heat capacity for propane in this range is 0.5 B.t.u./(lb.)(°F.)

Heat necessary to raise temperature of tallow stream 1°F. = 8,170 × 0.53 + 296 × 0.5 = 4,477 B.t.u./hr.

Heat given up by steam in cooling to incoming tallow stream temperature
= 300(1,195.6 - 1,169.3) = 7,900 B.t.u./hr.

The steam could not raise the temperature of the tallow as much as two degrees. Therefore it is assumed that the tallow stream leaves at the same temperature it entered and that the heat given off by the steam is dissipated in heat effects owing to stripping and in losses.

Exit streams at 252°F. and 16 lb./sq.in.abs.
Liquid: tallow = 8,170 lb./hr.
negligible amount of propane
Vapor: steam = 300 lb./hr.
propane = 296 lb./hr.

Size (given): 25 ft. high and 1 ft. in diameter packed with 15 ft. of 1-in. Raschig rings

Cost:
Vessel cost = 25 × 15 × 0.90 = $337.50

$$\text{Volume of packing} = \pi \left(\frac{1}{2}\right)^2 \times 15$$
= 11.8 cu.ft.
round off to 12 cu.ft.

Cost of packing = 12 × 5 = $ 60.00
Total stripper cost = $397.50

IV. Bottoms flash drum D4

Pressure = 230 lb./sq.in.abs.

Entering stream:

Color bodies = 430 lb./hr.

Propane = 218 lb./hr.

Exit streams:

Color bodies = 430 lb./hr.

$$\text{Propane liquid} = 0.03 \times \frac{430}{1.0 - 0.03}$$

= 13.3 lb./hr.

Propane vapor = 218 - 13.3 = 204.7 lb./hr.

Mole fraction propane in exit liquid

$$= \frac{\frac{13.3}{44}}{\left(\frac{13.3}{44}\right) + \left(\frac{430}{1{,}008}\right)} = 0.415$$

$$K = \frac{y^1}{x^1} = \frac{1}{0.415} = 2.41$$

Temperature = 237°F. [equilibrium constant nomograph (2)]

Diameter:

Vapor rate = 204.7 lb./hr. = 0.057 lb./sec.

Sp. vol. of vapor = 0.62 cu.ft./lb.

= 0.62 × 0.057 = 0.0354 cu.ft./sec.

$$\rho_G = \frac{1}{\text{sp.vol.}} = 1.61 \text{ lb./cu.ft.}$$

$$\rho_L = \left[0.980 - 3\times10^{-4}(237 - 100)\right] 62.4$$

= 58.4 lb./cu.ft.

$$D^2 = \frac{8 \times 0.0354}{0.227\pi \sqrt{\frac{58.4 - 1.61}{1.61}}}$$

D = 0.26 ft. to nearest $\frac{1}{2}$ ft. = 0.5 ft.

Height:

$$h_L = \frac{(430 + 13.3)1/12}{58.4\pi \left(\frac{0.5}{2}\right)^2} = 3.22 \text{ ft.}$$

h = 2 + 0.5 · 1.17 + 3.22 = 6 ft.

Cost:

Vessel cost = 6 × 8 × 2.20 =	$106
Mist separator = 4 × 20 =	$ 80
Total cost =	$186

V. Bottoms Steam Stripper S2

Pressure = 17 lb./sq.in.abs.

Entering stream at 237°F.

Color bodies = 430 lb./hr.

Propane = 13.3 lb./hr.

Heat balance: Assumptions made for overhead stripper cannot be made here because the bottoms stream is much smaller than the overhead stream.

Heat given up by steam = heat absorbed by bottoms stream

100 (ΔH steam) = 430 × 0.50 (t-237) + 13.3 × 0.5 (t-237)

By trial and error, t = 249°F.

Exit streams at 16 lb./sq.in.abs. and 249°F.

Liquid: color bodies = 430 lb./hr.
negligible amount of propane

Vapor: steam = 100 lb./hr.
propane = 13.3 lb./hr.

Size and cost same as for overhead stripper.

VI. Jet Condenser J1

Vapor streams from both strippers are joined and fed to a common jet condenser.

Balance to determine temperature of entering vapor:

300 (ΔH steam from t to 252°F.)
+ 296 × 0.5 (252-t)
- 100 (ΔH steam from 249 to t)
- 13.3 × 0.5 (t-249) = 0

t = 251°F.

Duty = 400(1,168.1 - 184.4) + 400(216 - 120) + 309.3 × 0.5(251-105) = 453,700 B.t.u./hr.

$$\text{City water consumption} = \frac{457{,}700}{1(120-90)8.33}$$

= 1,820 gal./hr.

Diameter:

Maximum vapor velocity will occur at vapor entrance.

Sp. vol. steam = 26.34 cu.ft./lb.

$$\text{Sp. vol. propane} = \frac{NRT}{P} = \frac{\frac{1}{44} \times 10.73 \times 709}{16}$$

= 10.8 cu.ft./lb.

$$\text{Average sp. vol.} = \frac{26.31 \times 400 + 10.8 \times 309.3}{400 + 309.3}$$

= 19.6 cu.ft./lb.

$$\rho_G = \frac{1}{19.6} = 0.051 \text{ lb./cu.ft.}$$

ρ_L = 62.4 lb./cu.ft.

$$\text{Max. vapor velocity} = \frac{0.136\sqrt{\rho_L - \rho_G}}{\rho_G}$$

$$= \frac{0.136\sqrt{62.4 - 0.051}}{0.051} = 21.1 \text{ ft./sec.}$$

$$\text{Cross-sectional area} = \frac{19.6 \times \frac{709.3}{3{,}600}}{21.1}$$

= 0.183 sq.ft.

$$\text{Diameter} = 2\sqrt{\frac{0.183}{\pi}} = 0.5 \text{ ft.}$$

Height = 17 ft. (given)

Cost = 17 × 8 × 0.90 + 5 × 20 = $222

VII. Heat Exchangers

Shell-and-tube countercurrent flow. The thermal resistance of the tubes and the inside-outside area factor are neglected in all the exchangers.

E1:

Hot side: 150-lb. steam
$t = 366^oF.$

Cold side: overhead stream from extraction tower
$t_{in} = 165^oF.$
$t_{out} = 175^oF.$

For the quantity needed, 150-lb. steam was cheaper than 15-lb. steam.

Heating duty = 8,170 × 0.53(175-165) + 121,580(322.5-312) = 1,320,000 B.t.u./hr.

Vaporizing duty = 58,330(408.5-322.5) = 5,010,000 B.t.u./hr.

Total duty = 6,320,000 B.t.u./hr.

$$\text{Steam consumption} = \frac{6,320,000}{857} = 7,370 \text{ lb./hr.}$$

Exchanger must be divided into two parts to calculate area.

Heating:

ln mean $\Delta T = 196^oF.$

$h_{tallow+propane} = 252$ B.t.u./(hr.)(sq.ft./$^oF.$) [Figure 3]

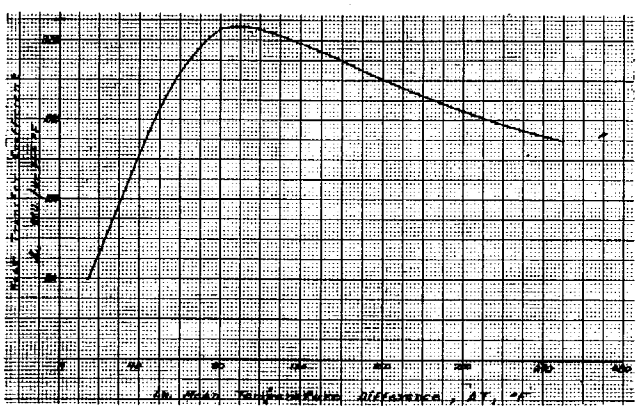


Fig. 3. Effect of ΔT on heat transfer coefficient for tallow plus boiling propane.

$$U = \frac{1}{\frac{1}{1,000}+\frac{1}{252}} = 202 \text{ B.t.u./(hr.)(sq.ft./}^oF.)$$

$$\text{Area} = \frac{1,320,000}{202 \times 196} = 33.3 \text{ sq.ft.}$$

Vaporizing at $175^oF.$

$\Delta T = 191^oF.$

$h_{tallow+propane} = 256$ B.t.u./(hr.)(sq.ft./$^oF.$)

$$U = \frac{1}{\frac{1}{1,000}+\frac{1}{256}} = 204 \text{ B.t.u./(hr.)(sq.ft./}^oF.)$$

$$\text{Area} = \frac{5,010,000}{204 \times 191} = 129 \text{ sq.ft.}$$

Total area = 162 sq.ft.

Cost = 162 × 10.50 × 1.46 = $2,485

It is assumed that cost data for exchangers cannot be interpolated.

E2:

Hot side: propane vapor from flash drum D1 $t = 175^oF.$

Cold side: tallow and liquid propane from flash drum D1 $t = 152^oF.$

Duty = 85.9 × 58,330 - 201,600 = 4,800,000 B.t.u./hr.

$$\text{Vapor condensed} = \frac{4,800,000}{86.0} = 55,850 \text{ lb./hr.}$$

It is assumed that on throttling before the stream goes through the exchanger the temperature drops to $152^oF.$, the boiling point of propane at 350 lb./sq.in.abs.

$\Delta T = 175 - 152 = 23^oF.$

$$U = \frac{1}{\frac{1}{400}+\frac{1}{118}} = 91.2 \text{ B.t.u./(hr.)(sq.ft./}^oF.)$$

$$\text{Area} = \frac{4,800,000}{91.2 \cdot 23} = 2,290 \text{ sq.ft.}$$

Cost = 2,290 × 3.50 × 1.46 = $11,700

E3:

This exchanger is used to condense any vapor not condensed in E2. The system is designed to give a slight excess of vapor over what is needed for heating in E2 in order to permit better control of the process.

Hot side: propane vapor from D1
$t = 175^oF.$

Cold side: cooling water
$t_{in} = 90^oF.$
$t_{out} = 105^oF.$

Amount of propane condensed = 58,330 - 55,850 = 2,480 lb./hr.

Duty = 2,480 (86.0) = 213,500 B.t.u./hr.

$$\text{Cooling-water consumption} = \frac{213,500}{15 \times 8.33} = 1,710 \text{ gal./hr.}$$

lm $\Delta T = 77.5^oF.$

$$U = \frac{1}{\frac{1}{300}+\frac{1}{400}} = 171.5 \text{ B.t.u./(hr.)(sq.ft./}^oF.)$$

$$\text{Area} = \frac{213,500}{77.5 \times 171.5} = 16.0 \text{ sq.ft.}$$

Cost = 16.0 × 15.00 × 1.46 = $350

E4:

Hot side: 150-lb. steam

$t = 366^{o}F.$

Cold side: tallow and boiling propane from drum D2

$t_{in} = 152^{o}F.$
$t_{out} = 252^{o}F.$

Duty = (4,920-296)(475-302) + 296 ×0.5(252-152) + 8,170 × 0.53(252-152) = 858,100 B.t.u./hr.

$$\text{Steam consumption} = \frac{858,100}{.857} = 1,000 \text{ lb./hr.}$$

It is assumed that the boiling point of the tallow-propane mixture changes with concentration of propane.

ln mean $\Delta T = 158.5^{o}F.$

$$U = \frac{1}{\frac{1}{1,000} + \frac{1}{282}} = 220 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

$$\text{Area} = \frac{858,100}{220 \times 158.5} = 16.0 \text{ sq.ft.}$$

Cost = 16.0 × 15.00 × 1.07 = \$256

E5:

Hot side: propane vapor from flash drum D2

$t_{condensing} = 152^{o}F.$

Cold side: cooling water

$t_{in} = 90^{o}F.$
$t_{out} = 105^{o}F.$

Duty = 58,330 (106.6) = 6,210,000 B.t.u./hr.

$$\text{Cooling-water consumption} = \frac{6,210,000}{15 \times 8.33}$$

= 49,700 gal./hr.

ln mean $\Delta T = 54.5^{o}F.$
U = 171.5 B.t.u./(hr.)(sq.ft./^{o}F.)

$$\text{Area} = \frac{6,210,000}{54.5 \times 171.5} = 665 \text{ sq.ft.}$$

Cost = 665 × 4.90 × 1.18 = \$3,840

E6:

Hot side: propane vapor from

flash drum D3	4,624 lb./hr. at 252^{o}F.
flash drum D4	204.7 lb./hr. at 237^{o}F.
compressor C1	309.3 lb./hr. at 165^{o}F.

Cold side: cooling water

$t_{in} = 90^{o}F.$
$t_{out} = 105^{o}F.$

Average vapor temperature in

$$= \frac{4,624 \times 252 + 204.7 \times 237 + 309.3 \times 665}{4,624 + 204.7 + 309.3}$$

$= 247^{o}F.$

Heat capacity assumed constant over range involved.

Condensing temperature at 230 lb./sq.in. abs. = 116^{o}F.
Desuperheating duty = 5,138 × 0.5(247-116) = 336,000 B.t.u./hr.
Condensing duty = 5,138(127.8) = 655,000 B.t.u./hr.
Total duty = 336,000 + 655,000 = 991,000 B.t.u./hr.

$$\text{Cooling-water consumption} = \frac{991,000}{(105-90)8.33}$$

= 7,940 gal./hr.

Desuperheating:

$$\text{ln mean } \Delta T = \frac{247-105-(116-90)}{\ln \frac{247-105}{116-90}} = 68^{o}F.$$

(Judging Committee comment: Adding liquid propane to superheated vapor until temperature of resulting vapor is such that tube wall is at or below condensing temperature makes it possible to increase heat transfer coefficient and thus lower appreciably surface area.)

$$U = \frac{1}{\frac{1}{15} + \frac{1}{300}} = 14.3 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

$$\text{Area} = \frac{Q}{U \cdot \Delta T} = \frac{336,000}{14.3 \cdot 68} = 345 \text{ sq.ft.}$$

Condensing:

ln mean $\Delta T = 17.5^{o}F.$

$$U = \frac{1}{\frac{1}{400} + \frac{1}{300}} = 171.5 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

$$\text{Area} = \frac{655,000}{17.5 \times 171.5} = 218 \text{ sq.ft.}$$

Total area = 345 + 218 = 563 sq.ft.
Cost = 563 x 6.15 × 1.07 = \$3,740

E7:

Hot side: color bodies from stripper S2

$t_{in} = 237^{o}F.$
$t_{out} = 125^{o}F.$

Cold side: cooling water

$t_{in} = 90^{o}F.$
$t_{out} = 105^{o}F.$

The storage temperature of color bodies is set at 125^{o}F.

Duty = 430 × 0.50(237-125) = 24,050 B.t.u./hr.

$$\text{Cooling-water consumption} = \frac{24,050}{15 \times 8.33}$$

= 193 gal./hr.

ln mean $\Delta T = 73^{o}F.$

$$U = \frac{1}{\frac{1}{300} + \frac{1}{50}} = 42.9 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

The h for color bodies is assumed to be the same for tallow.

$$\text{Area} = \frac{24{,}050}{42.9 \times 73} = 7.7 \text{ sq.ft.}$$

$$\text{Cost} = 7.7 \times 15.00 \times 0.94 = \$109$$

E8:

Hot side: 150 lb./sq.in. gauge steam
$t = 366^{o}F.$
Cold side: color bodies and propane from extraction tower
$t_{in} = 165^{o}F.$
$t_{out} = 237^{o}F.$

It is assumed that the boiling point of the propane solution of the color bodies changes with concentration. The temperature drop on throttling is neglected in the calculation of ΔT.

$$\text{Duty} = 430 \times 0.50(237-165) + 204.7(465-312) + 13.3 \times 0.5(237-165) = 44{,}500 \text{ B.t.u./hr.}$$

$$\text{Steam consumption} = \frac{44{,}500}{857} = 52 \text{ lb./hr.}$$

ln mean $\Delta T = 165^{o}F.$

$$U = \frac{1}{\frac{1}{1{,}000} + \frac{1}{277}} = 217 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

$$\text{Area} = \frac{44{,}500}{217 \times 165} = 1.24 \text{ sq.ft.}$$

$$\text{Cost} = 1.3 \times 15.00 \times 1.46 = \$28$$

E9:

Hot side: tallow from stripper S1
$t_{in} = 252^{o}F.$
Cold side: raw tallow from storage
$t_{in} = 125^{o}F.$
$t_{out} = 165^{o}F.$

$$\text{Duty} = 8{,}600 \times 0.529(165-125) = 182{,}000 \text{ B.t.u./hr.}$$

$$t \text{ out of hot stream} = 252 - \frac{182{,}000}{8{,}170 \times 0.53} = 210^{o}F.$$

ln mean $\Delta T = 86^{o}F.$

$$U = \frac{1}{\frac{1}{50} + \frac{1}{50}} = 25 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

$$\text{Area} = \frac{182{,}000}{25 \times 86} = 85 \text{ sq.ft.}$$

$$\text{Cost} = 85 \times 15.00 \times 1.46 = \$1{,}820$$

E10:

Hot side: tallow from stripper S1 after passing through E9
$t_{in} = 210^{o}F.$
$t_{out} = 125^{o}F.$
Cold side: cooling water
$t_{in} = 90^{o}F.$
$t_{out} = 105^{o}F.$

$$\text{Duty} = 8{,}170 \times 0.53(210-125) = 368{,}000 \text{ B.t.u./hr.}$$

$$\text{Cooling-water consumption} = \frac{368{,}000}{15 \times 8.33} = 2{,}940 \text{ gal./hr.}$$

$$\text{ln mean } \Delta T = \frac{210-105-(125-90)}{\ln \frac{210-105}{125-90}} = 63.6^{o}F.$$

$$U = \frac{1}{\frac{1}{50} + \frac{1}{300}} = 42.9 \text{ B.t.u./(hr.)(sq.ft./}^{o}\text{F.)}$$

$$\text{Area} = \frac{368{,}000}{42.9 \times 63.6} = 135 \text{ sq.ft.}$$

$$\text{Cost} = 135 \times 10.50 \times 0.94 = \$1{,}335$$

E11:

Hot side: exhaust steam, 15 lb./sq.in. gauge
$t = 250^{o}F.$
Cold side: propane from storage tank
$t_{out} = 165^{o}F.$

The propane storage temperature will be the average temperature of the streams flowing into the storage tank. Heat losses from the storage tank will be negligible because the throughput is large in relation to the volume of the tank.

Propane-storage temperature

$$= \frac{58{,}330 \times 175 + 5{,}138 \times 116 + 58{,}330 \times 152}{58{,}330 + 5{,}138 + 58{,}330} = 161.4^{o}F.$$

Constant heat capacity assumed over range involved.

$$\text{Duty} = 121{,}800 \times 0.5(165-161.4) = 219{,}000 \text{ B.t.u./hr.}$$

Fifteen-pound exhaust steam will be used for heating.

Exhaust steam is 89.5% quality.

$$\text{Exhaust-steam consumption} = \frac{219{,}000}{945 \times 0.895} = 259 \text{ lb./hr.}$$

This amount is available from P1.

$$\text{ln mean } \Delta T = \frac{250-161.4-(250-165)}{\ln \frac{250-161.4}{250-165}} = 86.8^{o}F.$$

$$U = \frac{1}{\frac{1}{1,000} + \frac{1}{120}} = 107 \text{ B.t.u./(hr.)(sq.ft./}^{\circ}\text{F.)}$$

$$\text{Area} = \frac{219,000}{107 \times 86.8^{\circ}\text{F.}} = 23.4 \text{ sq.ft.}$$

$$\text{Cost} = 23.4 \times 15.00 \times 1.46 = \$512$$

VIII. Compressor C1

At the rate of 309.3 lb./hr. propane vapor enters at 105°F. and 15 lb./sq.in.abs. and leaves at 165°F. and 230 lb./sq.in.abs. Compressor is three stage with intercoolers.

The calculations are made by the method described in reference 4 (Mollier diagram for propane used from reference 1).

Stage	Pressure, lb./sq.in.abs.	$(H_2\text{-}H_1)_S$, B.t.u./lb.	Temperature after compression, °F.
1	15 to 38	22	155
2	36 to 93	22	160
3	91 to 230	22	165

$$\text{Total } (H_2\text{-}H_1)_S = 66 \text{ B.t.u./lb.}$$

$$(H_2\text{-}H_1) \text{ actual} = \frac{66}{0.75} = 88 \text{ B.t.u./lb.}$$

$$\text{Compressor rating} = 88 \times \frac{309.3}{60} \times \frac{778}{33,000}$$

$$= 10.7 \text{ hp.}$$

$$\text{Capacity} = \frac{309.3}{60 \cdot 44} \times 359 = 42.0 \text{ std. cu.ft./min.}$$

Heat removed in intercoolers

$$= 309.3 \left[0.4(105\text{-}100) + 2 \frac{22}{0.75}\right]$$

$$= 18,750 \text{ B.t.u./hr.}$$

$$\text{Cooling-water consumption} = \frac{18,750}{1(105\text{-}90)8.33}$$

$$= 150 \text{ gal./hr.}$$

Steam drive:

Power per pound of steam

$$= 0.60(1,195.6\text{-}106.5)0.000393$$

$$= 0.0308 \text{ hp.-hr./lb.}$$

$$\text{150-lb. steam consumption} = \frac{10.7}{0.0308}$$

$$= 347 \text{ lb./hr.}$$

$$\text{Compressor cost} = 10.7 \times 100 = \$1,070$$

IX. Pumps

In the calculation of the pumps, friction and velocity heads are necessarily neglected. It is assumed that the pump cost is based on horsepower delivered to liquid.

(Judging Committee comment: Horsepower should be figured on installed basis.)

P1 - raw tallow to extraction tower, 8,600 lb./hr.

P_{in} = 14.7 lb./sq.in.abs.

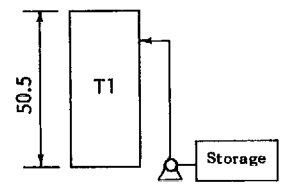
P_{out} = 450 lb./sq.in.abs.

Power delivered to liquid

$$= q\left[x_2 - x_1 + v(P_2 - P_1)\right]$$

$$= \frac{8,600}{3,600}\left[50.5 + \frac{(450\text{-}14.7)144}{\left[0.906 - 3\times10^{-4}(125\text{-}100)\right] 62.4}\right]\frac{1}{550}$$

$$= 5.05 \text{ hp.}$$



$$\text{Input to pump} = \frac{5.05}{0.60} = 8.41 \text{ hp.}$$

$$\text{150-lb. steam consumption} = \frac{8.41}{0.0308}$$

$$= 273 \text{ lb./hr.}$$

$$\text{Capacity} = \frac{8,600}{60} \cdot \frac{1}{8.33 \times 0.90} = 19.1 \text{ gal./min.}$$

$$\text{Pump cost} = 5.05 \times 115 = \$580$$

P2 - propane to extraction tower, 121,800 lb./hr.

Difference in elevation is neglected.

P_{in} = 445 lb./sq.in.abs.

P_{out} = 459 + 5 = 464 lb./sq.in.abs.

Tower pressure drop = 5 lb./sq.in.

Power to liquid = $q \times v(P_2\text{-}P_1)$

$$= \frac{121,800}{3,600}\, 0.041(464\text{-}445)\frac{144}{550}$$

$$= 5.09 \text{ hp.}$$

$$\text{Input to pump} = \frac{5.09}{0.60} = 8.48 \text{ hp.}$$

$$\text{150-lb. steam consumption} = \frac{8.48}{0.0308}$$

$$= 275 \text{ lb./hr.}$$

$$\text{Capacity} = \frac{121,800}{60 \times 3.2 \text{ lb./gal.}} = 650 \text{ gal./min.}$$

$$\text{Cost} = 5.09 \times 115 = \$585$$

P3 - propane from holding tank H2, 58,330 lb./hr.

Difference in elevation is assumed negligible.

P_{in} = 349 lb./sq.in.abs.
P_{out} = 449 lb./sq.in.abs.

Power to liquid

$$= \frac{58,330}{3,600} \times 0.0385(449-349)144 \times \frac{1}{550}$$

= 16.2 hp.

$$\text{Input to pump} = \frac{16.2}{0.60} = 27.0 \text{ hp.}$$

$$\text{150-lb. steam consumption} = \frac{27.0}{0.0308}$$

= 876 lb./hr.

$$\text{Capacity} = \frac{58,330}{60} \times \frac{1}{3.4 \text{ lb./gal.}} = 286 \text{ gal./min.}$$

Cost = 16.2 × 96 = $1,555

P4 - propane from holding tank H1 to storage, 5,138 lb./hr.

Differences in elevation are assumed negligible.

P_{in} = 230 lb./sq.in.abs.
P_{out} = 445 lb./sq.in.abs.

Power to liquid

$$= \frac{5,138}{3,600} \times 0.035(445-230)\frac{144}{550} = 2.81 \text{ hp.}$$

$$\text{Input to pump} = \frac{2.81}{0.60} = 4.69 \text{ hp.}$$

$$\text{150-lb. steam consumption} = \frac{4.69}{0.0308}$$

= 152 lb./hr.

$$\text{Capacity} = \frac{5,138}{60} \times \frac{1}{3.5} = 24.4 \text{ gal./min.}$$

Cost = 2.81 × 190 = $534

X. Holding Tanks - designed for 5 min. holding time.

$$\text{H1 - flow} = 5,138 \quad \frac{1}{12} = 428 \text{ lb./5 min.}$$

Sp. vol. propane = 0.035 cu.ft./lb.

Volume of holding tank = 428 × 0.035
= 18.3 cu.ft.

Diameter arbitrarily fixed at 2 ft.

$$\text{Height} = \frac{18.3}{\left(\frac{2}{2}\right)^2} = 5.83 \text{ ft.}$$

round off to 6.0 ft.

Cost = 6 × 28 × 2.20 = $370
Operating pressure = 230 lb./sq.in.abs.

$$\text{H2 - flow} = 58,330 \times \frac{1}{12} = 4,855 \text{ lb./5 min.}$$

Sp. vol. = 0.0385 cu.ft./lb.
Volume of tank = 4,855 × 0.0385 = 187 cu.ft.
Diameter arbitrarily fixed at 5 ft.

$$\text{Height} = \frac{187}{\pi\left(\frac{5}{2}\right)^2} = 9.5 \text{ ft.}$$

Operating pressure = 349 lb./sq.in.abs.
Cost = 9.5 × 74 × 2.70 = $1,900

XI. Propane Inventory

$$\text{Volume of extraction tower} = 50.5 \times \pi\left(\frac{8.5}{2}\right)^2$$

= 2,870 sq.ft.

Propane in extraction tower

$$= \frac{121,800}{121,800+8,600} \; 2,870 \times 24.4 = 65,400 \text{ lb.}$$

The flash drums and holding tanks are assumed to run normally at one half their holding capacity.

Propane in D1 =	63,250 × 1/12 × 1/2 =	2,640 lb.
D2 =	4,920 × 1/12 × 1/2 =	205 lb.
D3 =	296 × 1/12 × 1/2 =	12.5 lb.
D4 =	13.3 × 1/12 × 1/2 =	0.5 lb.
H1 =	1/2 × 428 =	214 lb.
H2 =	1/2 × 4,855 =	2,428 lb.

$$\text{Storage tank} = 25\pi\left(\frac{9}{2}\right)^2 \times \frac{1}{2} \times 24.4 = 19,400 \text{ lb.}$$

The amount of propane in the piping and heat exchangers cannot be calculated and is therefore neglected.

Total propane in system = 88,300 lb.
Density of liquid at 60°F. = 4.24 lb./gal.

$$\text{Original propane cost} = \frac{88,300}{4.24} \times 0.06 = \$1,250$$

XII. Utilities

150-lb. steam consumption

Steam strippers	S1 =	300 lb./hr.
	S2 =	100 lb./hr.
Compressor	C1 =	347 lb./hr.
Heat exchangers	E1 =	7,370 lb./hr.
	E4 =	1,000 lb./hr.
	E8 =	52 lb./hr.
Pumps	P1 =	273 lb./hr.
	P2 =	275 lb./hr.
	P3 =	876 lb./hr.
	P4 =	152 lb./hr.
Instrument compressor	=	25 lb./hr.
Total 150-lb. steam consumption	=	10,440 lb./hr.

Cost per year
= 365 × 0.85 × 24 × 10,440 × 0.00045
= $35,000

Additional cost for stripping steam
= 365 × 0.85 × 24 × 400 × 0.00010 = $298

15-lb. steam consumption

Heat Exchanger E11 = 259 lb./hr.
Building and pump house heating
= 200 lb./hr.
Steam for storage facilities = 400 lb./hr.
(cost of this item is to be neglected)

Pump and compressor exhaust steam is to be used for these purposes.

Consumption of 15-lb. steam from boiler
= 0 lb./hr.
Total cost of steam per year = \$35,300

Cooling-water consumption:

Heat exchangers	E3	=	1,710 gal./hr.
	E5	=	49,700 gal./hr.
	E6	=	7,940 gal./hr.
	E7	=	193 gal./hr.
	E10	=	2,940 gal./hr.
Compressor C1 inter-coolers		=	150 gal./hr.
Total cooling-water consumption		=	62,633 gal./hr.

Annual cooling-water cost
$= 365 \times 0.85 \times 24 \times 62{,}633 \times 0.00003$
= \$14,000

City-water consumption in jet condenser
= 1,820 gal./hr.
Annual city-water cost
$= 365 \times 0.85 \times 24 \times 1{,}820 \times 0.00008$
= \$1,083
Total water cost per year = \$15,100

Jet condenser	J1 =	\$ 222
Heat exchangers	E1 =	2,485
	E2 =	11,700
	E3 =	350
	E4 =	256
	E5 =	3,840
	E6 =	3,740
	E7 =	109
	E8 =	28
	E9 =	1,820
	E10 =	1,335
	E11 =	512
Compressor	C1 =	1,070
Pumps	P1 =	580
	P2 =	585
	P3 =	1,555
	P4 =	534
Holding tanks	H1 =	370
	H2 =	1,900
Propane	=	1,250

Total basic-process-equipment cost
= \$78,650
Auxiliary-equipment-material cost
= 78,650 0.305 = \$24,000
Freight, erection labor, taxes, etc.
= (78,650 + 24,000)0.21 = \$21,550
Electrical Power
Power for lighting, etc. = 35 kw.
Cost per year
$= 365 \times 0.85 \times 24 \times 35 \times 0.01 = \$2{,}600$
Total annual utilities cost
= 35,300 + 15,100 + 2,600 = \$53,000
Propane
Propane loss = 121,800 0.0003
= 36.54 lb./hr.
Cost to replace propane lost

$$= \frac{36.54}{4.24}\ 0.06 = \$0.517/\text{hr.}$$

Annual propane cost
$= 365 \times 0.85 \times 24 \cdot 0.517 = \$3{,}850$

XIII. Total Costs

Basic-process-equipment-material cost

Extraction tower	T1 =	\$35,384
Flash drums	D1 =	5,623
	D2 =	1,580
	D3 =	697
	D4 =	186
Steam strippers	S1 =	397.50
	S2 =	397.50

Total direct cost
= 78,650 + 24,000 + 21,550 = \$124,200
Total indirect cost
$= 124{,}200 \times 0.12 = \$14{,}910$
Constructed cost
= 124,200 + 14,910 = \$139,110
Total annual fixed charges
$= 139{,}110 \times 0.22 = \$30{,}600$

XIV. Cost Summary

Total costs per year

Fixed charges	=	\$30,600	
Utilities	=	53,000	
Propane	=	3,850	
Labor	=	95,000	
Maintenance	=	35,000	
Total annual cost			\$217,450

Annual revenues
Color bodies
$= 365 \times 0.85 \times 24 \times 430 \times 0.01 = \$\ 32{,}000$
Net annual cost \$185,450
Tallow produced per year
$= 365 \times 0.85 \times 24 \times 8{,}170 = 60{,}800{,}000$ lb.
Cost per lb. of tallow produced

$$= \frac{185{,}450}{60{,}800{,}000} = \$0.00305$$

Annual cost of caustic process
= 60,800,000 0.00499 = \$303,500
Annual saving of extraction over caustic process = 303,500 - 185,450 = \$118,050

$$\%\ \text{saving} = \frac{0.499 - 0.305}{0.499}\ 100 = 39\%$$

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