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LOW TEMPERATURE HEAT RECOVERY FOR A
SYNTHETIC NATURAL GAS PLANT

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LOW TEMPERATURE HEAT RECOVERY FOR A SYNTHETIC NATURAL GAS PLANT

By

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ABSTRACT

LOW TEMPERATURE HEAT RECOVERY FOR A SYNTHETIC NATURAL GAS PLANT

By

Craig K. Kendziorski

A gasification plant rejects nearly a billion Btu of low temperature (less than 350°F) waste heat each hour from various coolers and condensers.

The purpose of this thesis was to determine and then evaluate possible heat recovery schemes.

One scheme evalusted was an organic working fluid rankine cycle. This process uses waste heat from a condensing vapor stream to produce valuable shaft work or electricity. Payback periods of over five years are required.

The production of 50 psig steam from waste heat (for use outside of the plant) was also evaluated. Only 17,000 lb/hr of this low pressure steam can be produced however.

Preheating the feed to one of the plant's distillation columns with the hot bottoms product from the same column was also evaluated. This scheme would save a quarter of a million dollars in fuel each year with a seven month payback period.

ACKNOWLEDGMENTS

The author wishes to express his sincerest appreciation to Dr. Bruce W. Wilkinson, Department of Chemical Engineering, for his patience and guidance throughout this project.

The valuable assistance and financial support from the people of Consumers Power Company is also gratefully acknowledged.

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INTRODUCTION

The rejection of large quantities of low temperature waste heat to the environment is a common occurrence in almost all industrial operations. With the cost of energy (especially oil) increasing at an alarming rate, industry is forced into taking a hard look at the potential of salvaging some of this energy.

Consumers Power Company, owners of this nation's first synthetic natural gas (SNG) plant located in Marysville, Michigan is faced with this problem. This plant gasifies lighter petroleum fractions into SNG by the "Catalytic Rich Gas" process developed in England. Production began in late 1973 and was brought up to full capacity of 200 million cubic feet per day in 1974 with the addition of a second gasification train.

This plant suffers from very large heat losses with most of the heat being rejected at low (less than 300°F) temperatures. This appears to be more a problem inherent in the process itself than in an inefficient design. With the hopes of recovering some of this waste heat, Consumers Power Company sponsored this project.

This problem was approached in three steps:

1. Determination of the waste heat rejection points. This involved determining the quantity and temperature of heat rejected from 35 different points in the process.

2. Evaluation of these points to determine which have potential for heat recovery and what technique might best be used for each one.
3. Conceptual design of the necessary equipment to accomplish the recovery and determine its economic potential.

Appendices A and B are essentially independent design of two heat recovery techniques given detailed evaluation.

Payback periods for these designs were based on an assumed stream factor of 90%. During the extent of this project (9/78 - 8/79), plant operation has been on an intermittent schedule. Justifiable payback periods would be substantially longer if this operating schedule were continued.

On September 5, 1979 Consumers Power Company announced it would mothball the Marysville SNG Plant for an indefinite period. No capital expenditures can therefore be justified until the facility is to be restored to service.

DISCUSSION

The Marysville SNG plant is a complex chemical process as depicted in Figure 1, the Overall Process Flow Diagram. This plant has the flexibility to operate with the following feedstocks:

LPG	(Liquid Petroleum Gas)
Condensate	(Petroleum Off Products)
Center Cut	(Natural Gas Liquids)

Any desirable combination of the three main feedstocks are displaced or pumped from the feedstock caverns to the feed settling drums and desalters.

The center cut feedstock is fractionated in the debutanizer where the overhead cut (C_3 - C_4) is joined by the LPG feedstock and sent to the LPG Desulfurization Section. The bottom cut from the debutanizer is joined by the condensate feedstock and is fractionated in the splitter. The bottom of the splitter is normally hydrodesulfurized for use as plant fuel but may be returned via pipeline to the nearby Buckeye Pipeline Company. The overhead of the splitter is sent to the Feed (Naptha) Desulfurization Section.

Intermediate storage facilities are available for the desulfurized LPG and Naptha before being sent to the two gasification trains.

The feeds to the Gasification Sections are vaporized and then mixed with process steam. This mixture is first passed thru

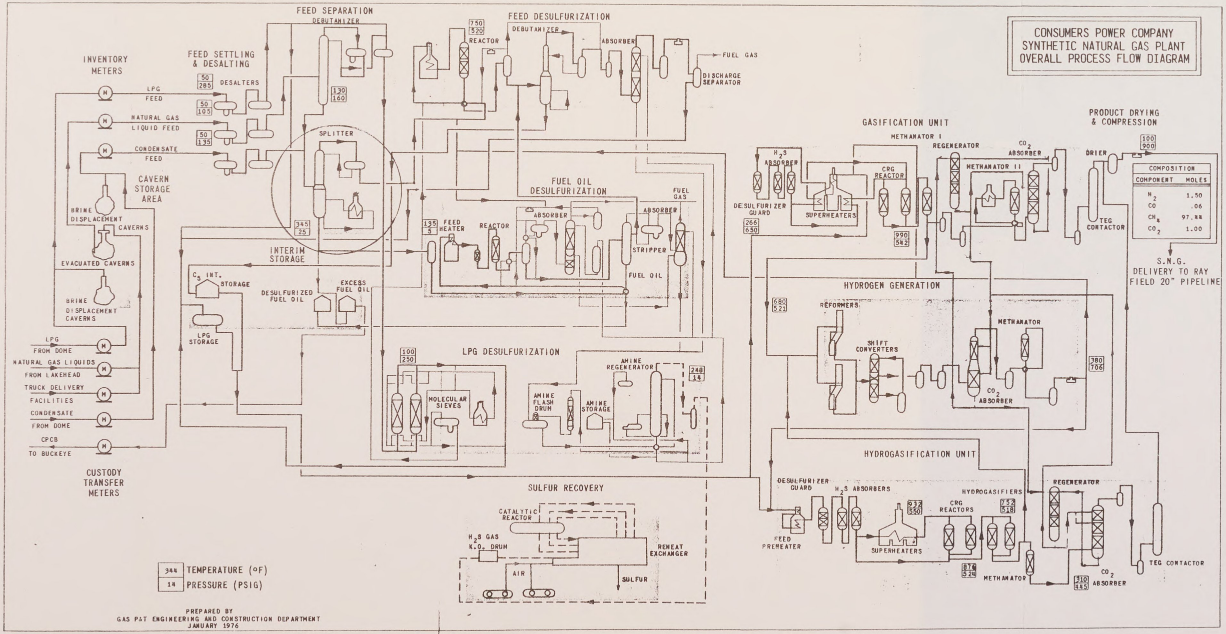
CONSUMERS POWER COMPANY
SYNTHETIC NATURAL GAS PLANT
OVERALL PROCESS FLOW DIAGRAM
INVENTORY METERS
LPG FEED
NATURAL GAS LIQUID FEED
CONDENSATE FEED
CAVERN STORAGE AREA
BRINE DISPLACEMENT
CAVERNS
EVACUATED CAVERNS
BRINE DISPLACEMENT CAVERNS
LPG FROM DOME
NATURAL GAS LIQUIDS FROM LAKEHEAD
TRUCK DELIVERY FACILITIES
CONDENSATE FROM DOME
CPCB TO BUCKEYE
CUSTODY TRANSFER METERS
FEED SETTLING & DESALTING
DESALTERS
50 285
50 105
50 135
FEED SEPARATION
DEBUTANIZER
130 160
SPLITTER
345 25
INTERIM STORAGE
C_5 INT. STORAGE
LPG STORAGE
DESULFURIZED FUEL OIL
EXCESS FUEL OIL
FEED DESULFURIZATION
750 520
REACTOR
DEBUTANIZER
ABSORBER
FUEL GAS
DISCHARGE SEPARATOR
FUEL OIL DESULFURIZATION
135 5
FEED HEATER
REACTOR
ABSORBER
ABSORBER
FUEL GAS
STRIPPER
FUEL OIL
LPG DESULFURIZATION
100 250
MOLECULAR SIEVES
AMINE REGENERATOR
248 14
AMINE FLASH DRUM
AMINE STORAGE
SULFUR RECOVERY
CATALYTIC REACTOR
H_2S GAS K.O. DRUM
AIR
REHEAT EXCHANGER
SULFUR
GASIFICATION UNIT
METHANATOR I
H_2S ABSORBER
DESULFURIZER GUARD
266 650
SUPERHEATERS
CRG REACTOR
990 542
REGENERATOR
CO_2 ABSORBER
METHANATOR II
PRODUCT DRYING & COMPRESSION
100 900
DRIER
TEG CONTACTOR
COMPOSITION
COMPONENT MOLES
H_2 1.50
CO .06
CH_4 97.44
CO_2 1.00
S.N.G. DELIVERY TO RAY FIELD 20" PIPELINE
HYDROGEN GENERATION
680 521
REFORMERS
SHIFT CONVERTERS
METHANATOR
CO_2 ABSORBER
380 706
HYDROGASIFICATION UNIT
DESULFURIZER GUARD
H_2S ABSORBERS
FEED PREHEATER
932 550
SUPERHEATERS
CRG REACTORS
HYDROGASIFIERS
752 518
874 524
METHANATOR
REGENERATOR
310 445
CO_2 ABSORBER
TEG CONTACTOR
344 TEMPERATURE (°F)
14 PRESSURE (PSIG)
PREPARED BY
GAS P&T ENGINEERING AND CONSTRUCTION DEPARTMENT
JANUARY 1976

FIGURE 1

Catalytic Rich Gas reactors and then Methanators.

This gas now contains mostly methane (SNG) and CO_2. It is cooled and after removal of excess CO_2 by absorption, the remaining SNG is fed to a glycol drier. The gas is now compressed and admitted to Consumers' pipeline system for distribution.

In addition to the main process flow as described above, the plant has an Amine System, Hot Pottasium Carbonate System, Hydrogen Plant and Steam Plant which serve as necessary auxiliaries to the process.

Due to natural gas availability and petroleum costs, the Marysville plant did not operate at full capacity during the time of this project. For this reason, no actual process operating data were taken directly from the plant. It would also be difficult to directly obtain meaningful data due to testing equipment required and the variations in feedstocks. However, actual plant data were available for some plant streams in the process from previous years.

The best source of plant operating conditions was a computer simulation of the process supplied by Consumers Power Company. This was used as a source of operating conditions for most of the work. The computer simulation was used with a more detailed set of process flow diagrams than the one pictured in Figure 1. The computer simulation gave temperatures, pressures, compositions, flow rates and some other information corresponding to almost every stream for a detailed process flow diagram. The computer simulation assumed plant operation at full capacity of 200 million cubic feet per day of SNG using 859.8 gpm of centercut and 811.4 gpm of condensate as feedstock.

Since this plant is relatively new (1973), it appears to be efficiently designed with respect to energy. Discharges of liquids and vapors are normally recovered and used as fuel. Many process streams are already heated by the cooling of other process streams. Medium and low pressure steam is generated from hot process streams at several locations. Boiler feed water is preheated by waste process heat and the flue gas from some of the fired heaters. Boiler combustion air is also preheated by boiler flue gas. This plant also uses a steam cogeneration system whereby boilers produce 600,000 lb/hr of superheated 1220 psig steam, which is used to drive the large SNG compressors as well as some other equipment. The exhaust steam from these turbines is further used as a source of process heat. With this abundance of low pressure steam which must be condensed, there is little incentive to replace it with some other source of heat.

Using the information from the computer simulation, energy balances were derived around every major piece of equipment where heat was rejected. The flue gas from the 10 fired heaters were not included since exact temperatures were unknown. The total amount of energy rejected by 35 major pieces of equipment was calculated. This information is tabulated in Table I.

It might be noted that the total calculated heat rejection rate is 750 million Btu/hr. At $2/million Btu this energy would be worth almost $12,000,000/year. Obviously it would be nice to recover some of this "free" energy.

TABLE 1. Waste Heat Rejection Points
(in order of descending temperature)

Equipment Description	C.P.Co. Equipment Number	Temp. In (°F)	Temp. Out (°F)	Calculated Energy (Million Btu/hr)	Comment
1) Splitter Bottoms Product Cooler	EC-1202	525	135	32.2	Liq. Fuel Oil
2) Hydrogen Cooler	EA-4010	350	100	2.5	Gas, some water condensation
3) Jet Cooler	EA-3007X	350	-	0.18	Condensing water
4) Gas Cooler	EC-4003	338	266	13.7	Gas stream, some water condensing
5) C_5^+ Product Cooler	EC-1303	328	150	36.7	Petroleum Liquid
6) Effluent Condenser	EA-1404	325	110	2.0	Gas, some water condensing
7) Methanator II Cooler	EC-2010	322	300	11.2*	CH_4 Gas
8) Stripper OVHD Condenser	EA-1405	310	110	2.78	Condensation
9) Splitter OVHD Condenser	EC-1201	292	135	97.7	Condensation, C_6
10) Methanator I Cooler	EC-2007	288	253	32.6*	CH_4 Gas
11) Hydrogasifier Cooler	EC-2507	280	246	36.0*	Gas, some water condensation
12) LPG Gas Cooler	EA-1312	262	100	11.5*	Gas and condensation
13) Product Cooler	EA-1406	260	120	5.7	Petroleum Liquid
14) Reactor Effluent Condenser	EC-1301	255	130	34.0	Some water condensation
15) Carbonate Cooler	EC-2011	250	158	50.0	Pottasium Carbonate Solution
16) Carbonate Cooler	EC-2511	250	158	50.0	Pottasium Carbonate Solution
17) Solution Cooler	EC-4002	250	158	20.3	Pottasium Carbonate Solution
18) BFW Cooler	EA-1407	250	125	0.63*	Liquid Water
19) Regenerator OVHD Condenser	EC-1501	248	130	5.8*	Condensation DEA
20) Acid Gas Cooler	EC-2513	230	150	99.2	Condensation water
21) Stripper Reflux Condenser	EA-7001	230	190	2.6	Condensation water
22) Acid Gas Cooler	EC-2013	200	153	97.0	Condensation water
23) Cooler	EA-3006X	200	-	0.78	TEG Liquid
24) SNG Final Cooler	EA-3007	183	100	17.7	CH_4 Gas
25) Lean Amine Cooler	EA-1501	180	125	3.5	DEA Liquid
26) Stripper Cooler	EA-7003	176	120	3.0	Liquid water
27) CRG Effluent Cooler	EC-2514	153	110	5.9	CH_4 Gas
28) CRG Effluent Cooler	EC-2015	153	110	5.7	CH_4 Gas
29) Trim Cooler	EA-1306	150	100	5.4*	Liquid
30) Cooler	EA-3004X	140	110	1.0	Liquid water
31) Debutanizer OVHD Condenser	EA-1201	137	100	42.6	Petroleum Condensation
32) Debutanizer OVHD Condenser	EA-1304	136	100	17.0*	Condensation
33) Reactor Trim Cooler	EA-1302	130	100	7.2	Liquid
34) Air Cooler**	EC-2012	950	-	11.5*	Start-up only
35) Desorber Condenser**	EA-1505	-	-	32.0*	Used only part time

Total Waste Heat = 750 million Btu/hr

Value of Waste Heat ($2/million Btu) = $11,800,000/year
(assuming .9 stream factor)

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*Design value.

**Used only for startups or regeneration of catalyst.

Table 1 contains the key process streams that were used in the evaluation of the heat recovery techniques and locations. Several criteria were used in the selection process. First, the temperature at which the heat is available is crucial. The greater the temperature, the more recovery potential a heat source has. For this reason, the heat sources are listed in order of descending temperature. Second is the amount of energy that is available. Even though they may be at high temperatures, some sources are insignificantly small to be considered. Finally, the short comment giving some idea of the fluid and phases present is important. This gives a qualitative idea of the heat transfer coefficient and other fluid properties. The availability of space for new equipment at the heat source is also very important. If there appeared to be potential for some heat recovery scheme, then the location of the heat source and the equipment around it was determined.

Using these criteria, the SNG plant was studied to determine potential heat recovery techniques. Numerous proposals were made at this initial stage but only three were deemed worthy of further evaluation. All three proposals evaluated centered around the splitter column. This column is circumscribed in Figure 1. Unfortunately, the overhead stream condenser and the bottom product cooler are not pictured.

A brief description of the three heat recovery schemes as well as the reasons behind their choice is given in the next three subsections. Details of the proposed schemes and the calculations made are given in the Appendices.

It should be noted that all designs assumed an on stream efficiency of 90%. Payback periods are, of course, strongly dependent on this assumption.

Splitter Column Feed Preheater

As mentioned earlier, the heating of one process stream with another is not a viable alternative if by product steam is replaced as a heating medium. Since the reboiler for the splitter distillation column uses a fired heater rather than steam (because of the high temperatures involved), preheating of the feed to this column would result in direct fuel savings for the reboiler.

The bottoms product from this column contains large amounts of sensible heat at high temperatures. This heat (some 32 million Btu/hr) is normally removed in the splitter bottoms air coolers (see Table 1). The use of this heat to preheat the column feed is proposed. Location is no problem as both streams serve the same column.

A heat exchanger was designed to recover some of the waste heat of the splitter bottoms product to preheat the feed to the same column. Because of the complexity of this design it is completely described and discussed in Appendix A.

Rankine Cycle

One of the newer schemes in heat recovery technology is the organic working fluid rankine cycle. This cycle is similar to the cycle used by power plants but employs an organic as the working fluid instead of water.

Figure 2 shows a schematic of a simple rankine cycle. The temperature of the waste heat source is normally much less than the

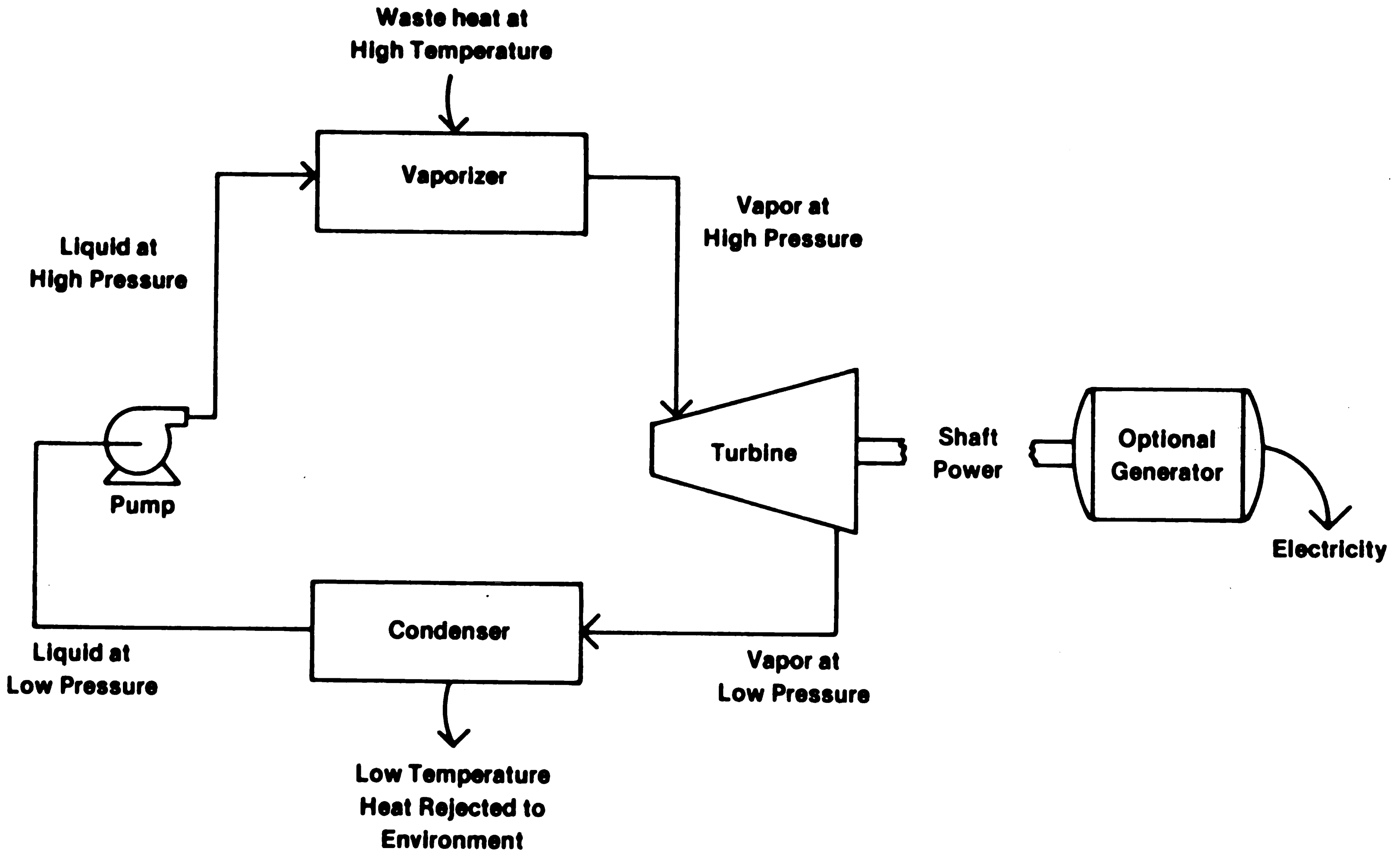

Figure 2 Simple Rankine Cycle

temperature of a steam boiler. This keeps the efficiency of rankine cycles low (10-20%) but the heat source is free and the shaft work or electricity produced is a valuable commodity.

None of the first eight heat sources listed in Table 1 were selected for the rankine cycle design because of one of the following reasons:

a) Not enough energy available (less than 5 million Btu/hr).

b) Poor heat transfer coefficients for gaseous streams.

c) Heat source number one already used in preheater design with good results.

The splitter overhead stream, (heat source number nine) was chosen for the rankine cycle design for the following reasons:

a) Large quantities of heat available.

b) Most of this heat (59.3 million Btu/hr is available at a reasonable temperature (292°F) from the condensation of the splitter overhead. The remaining heat duty is sensible heat of the condensate which is not used.

c) Good heat transfer coefficients can be expected from this condensing stream.

d) Opportunity to use the turbine shaft power directly to drive the 600 HP HDS pump located beneath the column.

The estimated temperature of the condensing overhead stream was later lowered to 260°F to more accurately resemble actual plant operation and the multicomponent nature of this stream. The complete rankine cycle design is presented in Appendix B.

Steam Generation Capabilities

The Marysville plant has problems with excess low pressure steam so it might seem contradictory to produce even more low pressure steam. If a nearby market were found for this steam however, it could become a valuable product. Possible markets could be heating for nearby homes and businesses or for use in a nearby industry. Sufficient quantities would be needed to make this economical.

Steam of at least 50 psig would probably be needed for distribution to market. With a minimum steam generation approach temperature of 35°F, this would require a waste heat source of at least 338°F. Only the top three sources listed in Table 1 seem reasonable and only the heat presently rejected by the splitter bottoms product cooler is of any significant quantity.

Cooling of this stream to 338°F would recover 15.85 million Btu/hr. This would generate a maximum of 17,400 lb/hr of 50 psig steam worth $274,000/year at $2/1000 lb. Assuming an overall heat transfer coefficient of 150 Btu/(hr)(ft^2)(°F), the required heat exchange area would be 1030 ft^2 with a purchased cost of $14,300(11) (January 1979). Using an installation factor of 4.74(2) the installed cost would be $67,900. Although this does not include the cost of water treatment, piping, and pumpimg, the payout period could be as low as 0.25 years with such a scheme. In order to justify further exploration of this option, a customer would have to be identified whose steam demand matched the steam availability from this source. It might be noted that the steam generating option is in conflict with the feed preheat option presented earlier.

CONCLUSIONS AND RECOMMENDATIONS

It has been estimated that in excess of 750 million Btu/hr of waste heat is being rejected from the plant via some 35 process streams. This heat represents fuel or wasted feedstock worth almost 12 million dollars each year. This heat is generally available at temperatures less than 350°F. With the cogeneration system presently used, the opportunity for simple heat interchange between process streams is greatly reduced.

Steam Generation

The production of low pressure steam for use outside of the plant is not extremely attractive despite the short 0.25 year payback period. Only the splitter bottoms stream is of high enough temperature for the production of low pressure steam and only 17,400 lb/hr of 50 psig steam (worth $274,000/year @ $2/1000 lb) can be produced from it. This small quantity of steam probably would not justify the cost and problems envolved in delivering to an outside market.

Rankine Cycle

An 1,800 kilowatt rankine cycle using Fluorinol-70 as a working fluid was designed to operate using the heat from the condensing splitter overhead stream. Estimated installed cost of this process is $2,100,000 and the power generated would have a value of $425,000/

year based on an electrical value of 3¢/Kw-hr. Estimated operating costs are $207,000/year and most of this ($171,000/year) is fixed cost. The overhead condensing air coolers presently in operation must also have substantial operating costs, which could be credited to the rankine cycle (exact value unknown). Payback period including the $207,000/year operating cost is 9.3 years or without operating cost 4.8 years. Actual payback period should be somewhere between the two depending on the actual operating costs of the air coolers.

The advantages and disadvantages of this rankine cycle are listed below.

Advantages

1. Produces $425,000/year of shaft of power or electricity.
2. 600 HP steam turbine driven HDS Pump located beneath this column may use some of this power directly (electrical back-up motor is also available).

Disadvantages

1. Added complexity.
2. Substantial payback period.
3. Questionable ability to function with varying splitter overhead conditions.
4. Substantial amounts of cooling water needed (3540 gpm). This may require enlargement of cooling water facilities.
5. Large equipment size.

The rankine cycle concept might have more potential if sized to power the 600 HP HDS pump only. This would eliminate the generator and the need to tie into the electrical grid. Electrical back-up motor is already available and could be used during any rankine cycle down time. The splitter overhead stream could still be used, but the splitter bottoms product stream would probably give even better results (higher cycle temperatures and efficiencies would be possible using the bottoms stream in this smaller rankine cycle).

It should be noted that the bottoms stream was used successfully in the Preheater Design.

Feed Preheater

The best results were obtained in the feed preheater design for the splitter column . A brief summary of these results is shown in Table 2.

TABLE 2. Feed Preheater Summary

Amount of Heat Transferred	16.58 million Btu/hr
Estimated Installed Cost of Heat Exchanger	\$115,700(2,10)
Fuel Savings in Reboiler @ \$2/million Btu	\$207,000/year
Payback Period	0.56 year

The feed preheater has a short payback period of 0.56 years and excellent fuel savings of \$207,000/year. As indicated in Appendix A, preheating the feed would have no effect on column top and bottom products. However, there would be substantial changes in internal flow rates in certain regions of the column. It was assumed that the column would need no modifications to handle these changes. The overhead air condenser duty was increased by 3.48 million Btu/hr (7.1%) due to changes in vapor rate. If the displaced bottoms product air coolers could be used, sufficient overhead condenser capacity could be achieved.

This preheater was designed to have a minimum payback period. The amount of heat recovered may be increased beyond this design with an ever increasing capital cost. Figure 20, Appendix A shows how payback period changes with splitter feed temperature.

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APPENDICES

APPENDIX A

FEED PREHEATER DESIGN FOR SPLITTER COLUMN

Appendix A

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1. SUMMARY

The feed preheater for the splitter column was placed after the debutanizer bottoms and condensate feed stream were joined (see Figure 3 in Process Description Section following). The advantages of having the preheater function on the combined streams are:

a) Larger flow rates with less temperature rise per heat input.

b) Ability to achieve substantial preheat with any combination of feedstocks.

The feed preheater recovers 16.58 million Btu/hr from the bottoms product stream. Of this energy 13.1 million Btu/hr is truly recovered as reboiler savings. The other 3.48 million Btu/hr shows up as increased overhead condenser duty (external reflux increases by 7.1%). Additional heat transfer capability from the released bottoms product air coolers should be more than is needed by the overhead condensers. It is hoped that some of the excess bottoms product air coolers can be adapted and used as overhead condensers. If this is not possible, then more overhead condensers will have to be purchased or the preheat exchanger could simply be bypassed on extremely hot days when the condensers can no longer keep up with demand.

Preheating the feed will have no effect on the composition of the product streams from the column. Vapor and liquid rates were substantially changed in some regions of the column, however. A more detailed study would be needed to determine the effects of these changes under all operating conditions. It was assumed for this preliminary design that no internal changes would be needed in the column.

The installed cost of the feed preheat exchanger is $115,700 (January 1979) and the energy (fuel) savings in the reboiler is $207,000/year ($2.00/million Btu). Estimated payback period is seven months.

2. BASIS OF DESIGN

This design is based on plant operation at full capacity (200 million cubic feet of SNG per day) using 859.8 gpm of center-cut and 811.4 gpm of condensate as feedstock. A .9 stream factor was also used. Compositions and flow rates of the feed to the column, the bottoms product, and the overhead product were taken from the Computer Simulation (2/10/73), supplied by Consumers Power Company.

Temperature of the feed without preheating was chosen as 200°F. The actual plant data shown in Table 3 agrees very well with this feed temperature. Bottoms product temperature was taken as 473°F. This temperature was chosen before the plant data was obtained but agreement is excellent.

The distillation column would operate with somewhat different internal and external vapor-liquid rates if the preheater were installed (extent of these changes is discussed in the Distillation Program Section). It was assumed in this design that these changes would not affect the column's operation or require any internal modifications to the column. The same feed plate would be used. It was also assumed that space would be available for the placement of the preheater near the splitter column. Use of a feed preheater would increase the heat duty demands on the overhead condenser. This was not taken into account by this design but it was hoped that some of

TABLE 3. Splitter Column Plant Data from the Marysville SNG Plant

Date	Condensate Flow (gpm)	Centercut Flow (gpm)	OVHD Vapor Temp. (°F)	Bottoms Temp. (°F)	Feed Temp. (°F)
9/24/75	990	576	266	514	193
10/24/75[a]	750	925	216	396	196
11/24/75	1060	510	271	511	198
12/24/75	970	300	256	514	191
1/31/76	710-1000	480	256	516	141-187
2/21/76	778	580	268	523	196
3/23/76[b]	670	400	240	424	184
10/25/76	380	220	238	430	191
11/25/76	1046	345	232	448	189
12/24/76	933	490	259	519	201
1/24/77	876	540	248	494	192
2/24/77	806	445	239	455	189
3/16/77[c]	764	510	250	484	193

Data taken on 24th of each month if plant running normally and data available.

[a] Data not consistent for that day.

[b] Troubles in Desalters.

[c] Plant shut down before 24th.

the excess bottoms product air coolers could be modified and used with the present air cooled condensers. No salvage value was taken for extra air coolers released from service. Energy savings in the reboiler were valued at $2.0/million Btu.

The average heat of vaporization of the feed stream was assumed to be 13,000 Btu/lb-mole. This value is not completely constant, but was assumed so for this preliminary design.

3. PROCESS DESCRIPTION

Figure 3 shows a schematic of the splitter column with feed preheater.

The debutanizer bottoms stream (136,520 lb/hr) is expanded from 165 psig to 33 psig. This cools this stream from over 300°F to somewhere near 200°F (no exact information given). This stream then joins with 297,130 lb/hr of condensate feed from the desalters. The resultant feed (433,650 lb/hr, 200°F and 33 psig) enters the shell side of the proposed feed preheater. On the tube side flows 126,990 lb/hr of bottoms product from the splitter column. A total of 16.58 million Btu/hr of heat is recovered by the feed as the bottoms stream's temperature is lowered from 473°F to 272.1°F. The resultant fuel oil is then sent to air coolers where its temperature is reduced to 135°F. The feed stream is heated from 200°F to 235°F and partially vaporized by the preheater. The feed leaves the preheater as a two phase mixture of 950 lb-moles/hr vapor and 3316.3 lb-moles/hr of liquid. The feed undergoes a pressure drop of 10 psi in the preheater and is now ready to enter onto plate 16 of the splitter column.

Not all of the 16.58 million Btu/hr recovered by the preheater comes out as savings in the reboiler. The reboiler duty is decreased by 13.1 million Btu/hr and the other 3.48 million Btu/hr shows up as increased condenser duty.

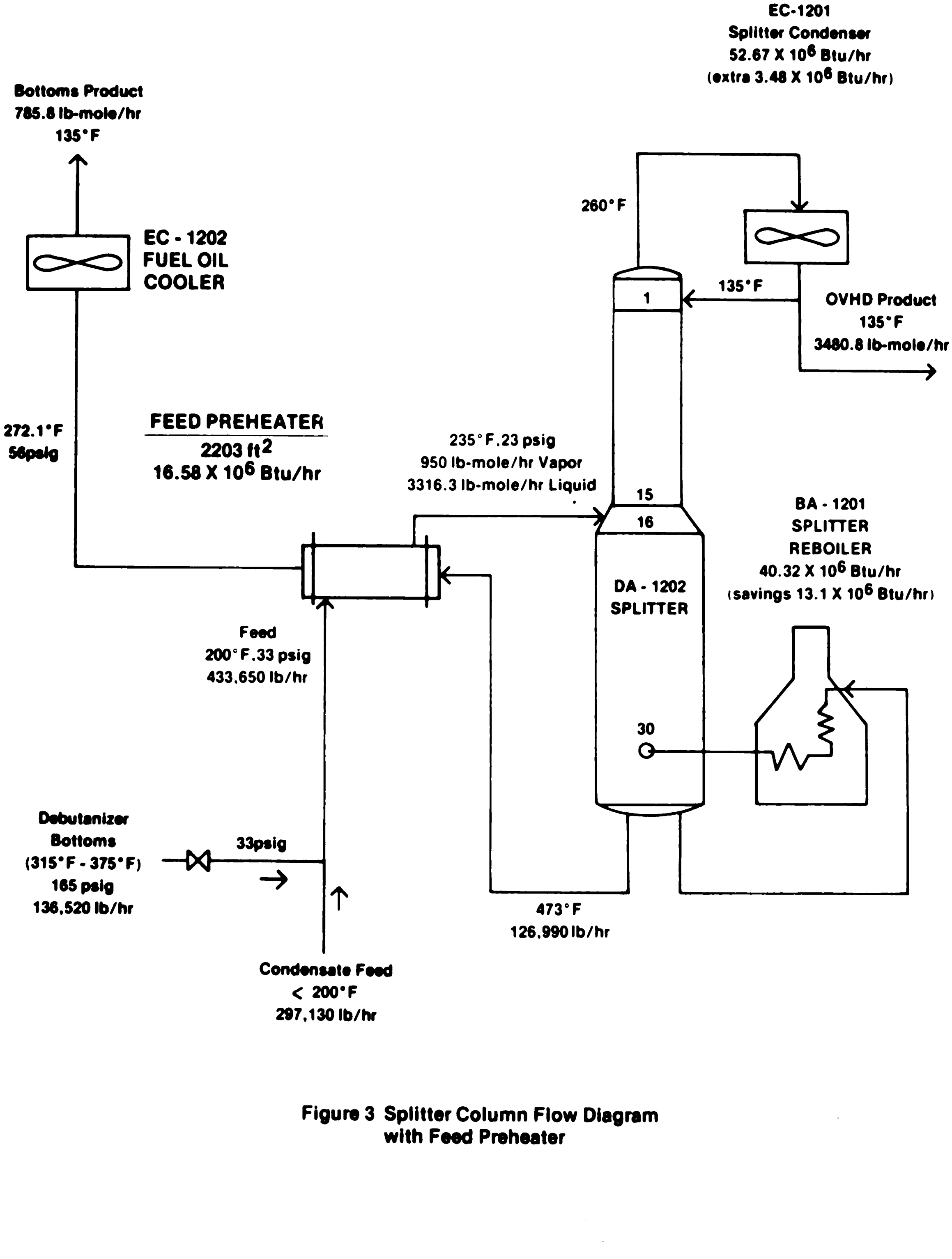

Figure 3 Splitter Column Flow Diagram with Feed Preheater

4. PREHEATER DESCRIPTION

Only one major piece of equipment is added to the existing process. This is a horizontal shell and tube feed preheat exchanger. See Tables 4 and 5 for the preliminary specifications.

The feed to the column is placed on the shell side of the preheater because of its greater flow rate. Less fouling is also expected from this stream because of its cooler temperatures and lower molecular weight than the bottoms product. To ease cleaning, a one inch square pitch tube arrangement is used.

The bottom product is placed on the tube side. Four tube passes were used to increase the velocity to 2.84 ft/sec. The much higher temperatures and molecular weight of this fluid may lead to greater fouling. The tube side flow would ease cleaning when necessary.

The inside heat transfer coefficient was calculated to be 238.8 Btu/(hr)(ft^2)(°F) by the Dittus Boelter equation. The shell side calculations are divided into two parts. Heating of the subcooled liquid feed takes place in the first 2.8 feet of the preheater. Here 4.23 million Btu/hr of heat is transferred as the feed is heated from 200°F to 215°F. In the next 11.6 feet vaporization of 950 lb-moles/hr of the original 4266.3 takes place. The temperature rises from 215°F to 235°F as 12.35 million Btu/hr are transferred. For the subcooled section the outside heat transfer coefficient (h_o) was

TABLE 4. Feed Preheater Description

Heat Duty	16.58 million Btu/hr
Temperature Driving Force	125.5°F
Overall Heat Transfer Coefficients (see Calculations Section)	
Vaporization Section	55.3 Btu/(hr)(ft^2)(°F)
Subcooled Section	79.6 Btu/(hr)(ft^2)(°F)
Surface Area	2203 ft^2
Material of Construction	
Shell	Carbon Steel
Tubes	Carbon Steel
Heat Exchanger Cost (Jan. 1979)	$24,400(10)
Installed Cost (4.74(2))	$115,700
Energy Savings ($2/million Btu)	$207,000/year
Payback Period*	0.56 yr

---

*Ignores extra condensor requirements of 3.48 million Btu/hr (increase of 7.1%). Displaced Bottoms Product coolers should be usable (16.58 million Btu/hr capacity).

TABLE 5. Feed Preheater Specifications

Area	2203 ft^2
Tube Length	14.4 ft
Number of Tubes	780
BWG	12
OD	.75"
ID	.532"
Pitch	1" square
Shell ID	35"
Number of Passes	
Tube Side	4
Shell Side	1
Baffle Spacing (25% cut)	16"

Tube Side	Shell Side
Fluid: Splitter Bottoms	Fluid: Feed
Flow Rate: 126,990 lb/hr	Flow Rate: 433,650 lb/hr
Inlet Temp.: 473°F	Inlet Temp.: 200°F
Outlet Temp.: 272.1°F	Outlet Temp.: 235°F
Pressure Out: ≈ 56 psig	Vapor: 900 lb-moles/hr
Pressure Drop: = 0.67 psi	Liquid: 3366.3 lb-moles/hr
	Pressure Out: ~ 23 psig
	Pressure Drop: 10.0 psi

found to be 300.2 Btu/(hr)(ft^2)(°F) by the following equation from Kern.$^{(5)}$

$$\frac{h_o D}{k} = 0.36\ (Re_s)^{.55}\ (Pr)^{1/3}\ .$$

For the boiling section h_o was found to be 113.1 Btu/(hr)(ft^2)(°F) by the following equation from Peters and Timmerhaus$^{(11)}$ for film boiling outside horizontal tubes.

$$h_o = 0.62 \left[\frac{k_v^3\ \rho_v(\rho_L - \rho_v) g\ \lambda_c}{D_o\ \mu_v\ \Delta t_f} \right]^{1/4} .$$

An overall fouling coefficient of 300 Btu/(hr)(ft^2)(°F) was used as recommended by Peters and Timmerhaus.$^{(11)}$ The overall heat transfer coefficients (U_o) are then 79.6 Btu/(hr)(ft^2)(°F) for the subcooled section and 55.3 Btu/(hr)(ft^2)(°F) for the boiling section.

Because of the phase change and the multiple tube passes, calculation of the true log mean temperature difference $(\Delta T\ \mathit{lm})$ is difficult. For this reason a very conservative $\Delta T\mathit{lm}$ was calculated using 215°F as the inlet temperature of the feed rather than the actual 200°F. This $\Delta T\mathit{lm}$ was used for both sections.

A total heat transfer area of 2,203 ft^2 is needed to recover the 16.58 million Btu/hr from the bottoms product. A tube length of 14.4 feet in a shell of approximately 3 feet in diameter is used. These dimensions should fit easily into any available space near the column. Carbon steel tubes and shell should be adequate for these fluids. Cost of this heat exchanger is \$24,400$^{(10)}$ (January 1979).

The pressure drop on the tube side was estimated to be 0.67 psi by the following equation from Kern.$^{(5)}$

$$\Delta P_t = \frac{1}{2}\,\frac{f\,G^2\,L\,N}{5.22\times 10^{10}\,D\,S} \quad .$$

The pressure drop on the shell side was estimated to be 10 psi by the following equation from Kern.$^{(5)}$

$$\Delta P_s = \frac{f\,G_s^2\,D_s\,(N+1)}{5.22\times 10^{10}\,D_e\,S} \quad .$$

The amount of pressure drop on the shell side is irrelevant since this stream is available at higher pressures and can be throttled down to any desirable pressure.

5. ECONOMIC ANALYSIS

The cost of the feed preheat exchanger (the only major piece of equipment needed) is \$24,400(10) (January 1979). Using a factor of 4.74(2) to take into account all other factors entering into the installation of this heat exchanger an installed cost of \$115,700 is obtained. The energy recovered is worth \$207,000/year. This assumes a 0.9 stream factor and an energy value for the reboiler of \$2.00/ million Btu. Payback period is only .56 year.

Not included are additional maintenance and other costs. Further, the cost to convert some of the present bottoms product air coolers for use as air cooled condensers is not included. Cost of any modifications to the distillation column if needed were also not included.

TABLE 6. Feed Preheater Economic Breakdown

Preheater Cost (January 1979)	\$24,400(10)
Installed Cost (4.74(2) factor)	\$115,700
Energy Saved	13.1 million Btu/hr
Value of Energy (\$2.00/million Btu)	\$207,000/year
Payback Period	0.56 year

6. PARAMETRIC STUDIES AND OPTIMIZATION

The purpose of this study was to determine the optimum feed temperature to the splitter column. Since this temperature corresponds to a certain amount of preheat, the purpose was essentially to find the optimum amount of preheat. The emphasis was to keep the payback period as low as possible while recovering as much energy as possible.

To simplify the procedure, a counter current exchanger with an overall heat transfer coefficient of 50 Btu/(hr)(ft^2)(°F) was assumed. Heat exchanger costs were found in Peters and Timmerhaus (11, p. 568) All other assumptions and values are as stated in the Basis of Design Section.

Using Figure 5 (Vapor Content vs Temperature diagram found in Section 7), the vapor content of the feed at the desired temperature was found. The necessary amount of preheat was then calculated as was the preheater size and cost. The vapor content of the feed was used in a multicomponent distillation design program (Section 7) to determine the energy savings in the reboiler as well as the extra condenser duty required. The value of the energy savings and the payback period could then be determined. The results of the optimization study are tabulated in Table 7 and shown in Figure 4, in which a plot of Payback Period vs Feed Temperature is shown.

Their exists a broad feed temperature range (215-240°F) over which the payback period is essentially constant. The temperature

TABLE 7. Optimization Results Varying Splitter Feed Temperature

New Feed Temp. (°F)	Preheat (10^6 Btu/hr)	Bottoms Product Outlet Temp. (°F)	ΔT_m Subcool Section (°F)	Boiling Section (°F)	Preheater Area (ft^2)	Cost (Jan. 1979) ($)	Installed Cost (10^3 $)	Reboiler Savings (10^6 Btu/hr)	Energy Value ($/yr)	Extra Condenser Duty (10^6 Btu/hr)	Payback Period (yrs)
215	4.23	422	239	-	350	6,150	29	3.38	63,000	0.85	0.46
220	7.80	379	196	233	740	9,500	45	6.21	115,000	1.59	0.39
225	11.02	340	157	210	1190	11,800	62	8.76	162,000	2.26	0.38
230	13.98	304	121	187	1740	17,700	84	11.08	205,000	2.90	0.41
235	16.58	272	89.0	165	2450	23,700	112	13.1	243,000	3.48	0.46
240	18.55	248	65.0	146	3260	28,400	135	14.63	271,000	3.92	.50
245	22.10	205	17.6	110	8071	68,600	325	17.34	322,000	4.76	1.01

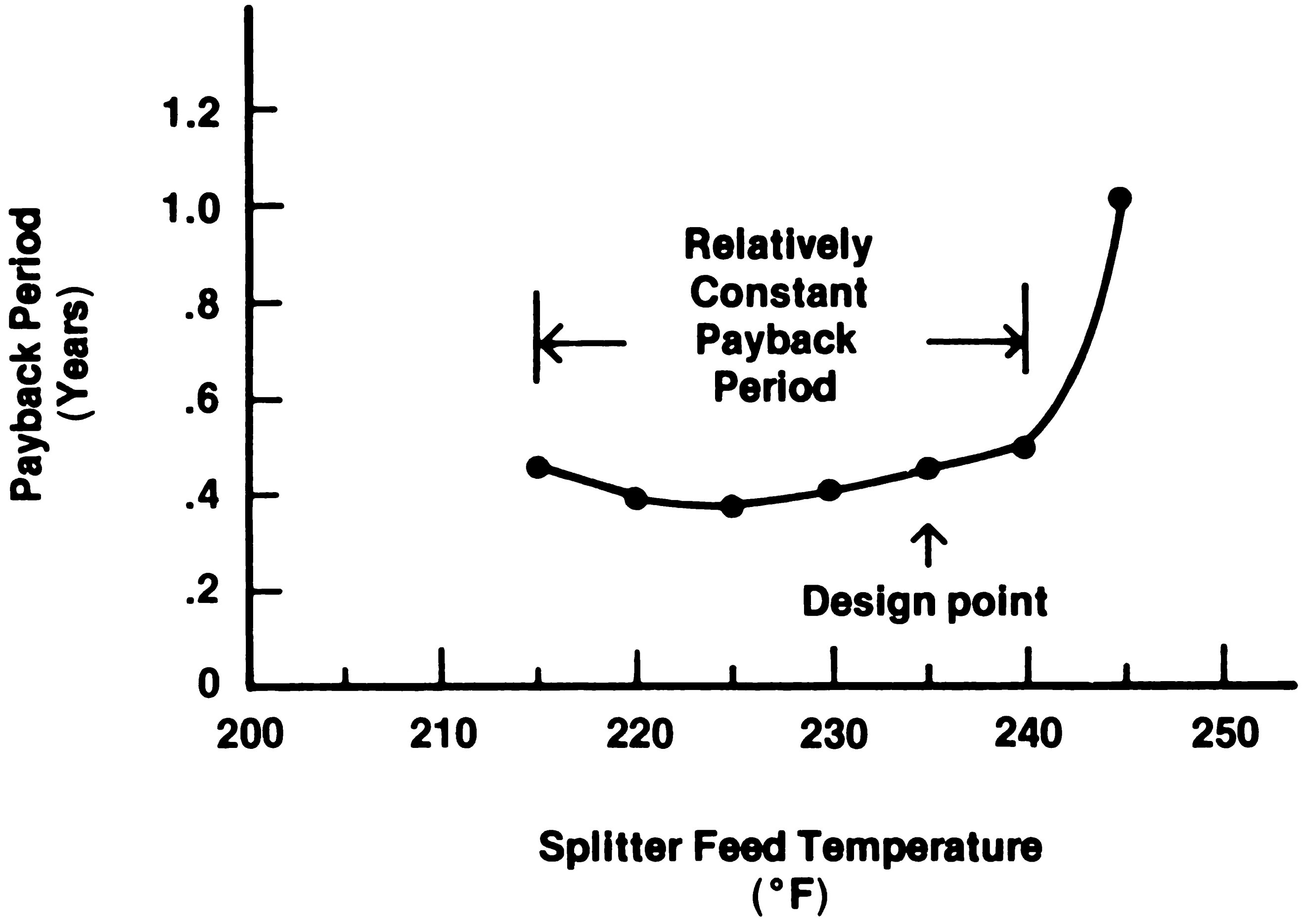

Figure 4 - Payback Period VS. Splitter Feed Temperature

of 235°F was chosen for the design. It has a short payback period of 0.48 years and an energy recovery of 13.1 million Btu/hr from 16.58 million Btu/hr of preheat.

7. COMPUTER PROGRAMS

Vapor-Liquid Equilibria Program

Of major importance in this design is the amount of vapor contained in the feed to the distillation column. To determine this a very simple computer program was written to solve the numerous independent equations obtained.

The pressure of the splitter distillation column is approximately 23 psig. At this pressure K values were obtained for each of the feed components at various temperatures. Using these values in the program, the vapor-liquid compositions and amounts were obtained at various assumed temperatures. Of particular interest is the quantity of vapor in the feed at each temperature. This information was used to plot Figure 5.

From this plot, the amount of vapor contained in the feed at any temperature could be determined. This curve was extrapolated back to zero vapor content to find the initial boiling temperature.

Distillation Program

The purpose of this program was to determine the effect of preheat on the splitter column's operation. Of major interest is the preheat's effect on reboiler heat duty.

The relative volatilities of the components change slightly from the top to the bottom of the column. To compensate for this,

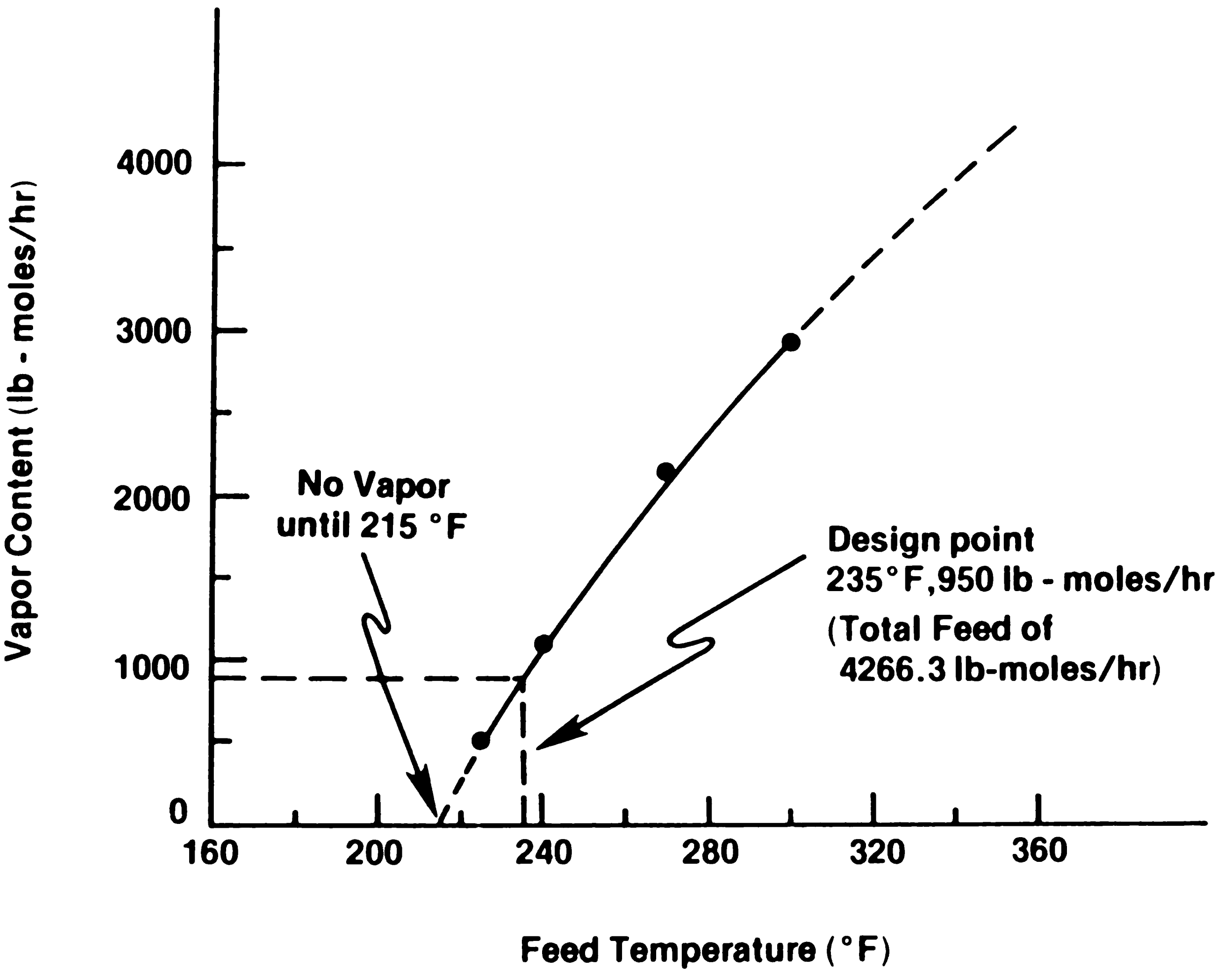

Figure 5 - Vapor Content VS. Feed Temperature

```
C       VAPOR-LIQUID EQUILIBRIA CALCULATIONS

        PROGRAM CRAIG (INPUT,OUTPUT)
        DIMENSION FXF(17),X(17),Y(27),XMOL(17),YMOL(17)
        REAL K(17),L,LO,LG(5)
C       VAPOR-LIQUID EQ.   FOR SPLITTER FEED
        READ 100,(FXF(I),I=1,17)
        F=4266.3
C       LG=  INITIAL GUESSES OF LIQUID,  BASED ON PREVIOUS RUNS
        LG(1)=1305.  $ LG(2)=2120.  $ LG(3)=3164. $ LG(4)=3712.0
        DO 111 LL=1,4
        L=LG(LL)
        READ 100,T
        READ 100,(K(I),I=1,17)
        PRINT 50
        V=F-L
1       SUMX=0.0
        DO 5 I=1,17
        X(I)=FXF(I)/(L+K(I)*V)
5       SUMX=SUMX+X(I)
        IF(ABS(SUMX-1.0).LE..00001) GO TO 10
        L=L+.05
        V=F-L
        IF (L.GT.4700.) GO TO 10
        GO TO 1
10      SUMY=0.0
        DO 20 I=1,17
        XMOL(I)=X(I)*L
        Y(I)=X(I)*K(I)
        SUMY=SUMY+Y(I)
        YMOL(I)=Y(I)*V
20      CONTINUE
        PRINT 300,T,V,L,SUMX,SUMY
        PRINT 350
        DO 22 I=1,17
22      PRINT 400,I,K(I),X(I),XMOL(I),Y(I),YMOL(I)
111     CONTINUE
50      FORMAT("1",///,35X,*VAPOR-LIQUID EQUILIBRIUM CALCULATIONS FOR SPLI
       1TTER FEED*,//////)
100     FORMAT(7(F10.5))
300     FORMAT(10X,*TEMPERATURE =*,F6.2,10X,*VAPOR =*,F9.2,10X,*LIQUID =*,
       1F9.2,10X,*SUMX =*,F7.5,8X,*SUMY=*,F7.5,//////)
350     FORMAT(6X,*COMPONENT*,11X,*K VALUE*,12X,*X VALUE*,10X,*LIQUID MCLE
       1S*,11X,*Y VALUE*,10X,*VAPOR MOLES*,//)
400     FORMAT(10X,I2,5(10X,F10.5),/)
        STOP
        END
 86.3     1202.8     566.5      182.       46.4       344.6      312.
 138.3    263.7      69.6       221.       32.7       319.4      209.1
 205.6    43.8       22.5
  300.
 9.8      4.9        2.4        2.01       2.12       1.3        1.55
  1.0     .75        .39        .464       .45        .27        .127
  .0348   .0054      .0007
  270.
  8.2     3.8        1.75       1.4        1.47       .9         1.1
  .774    .5         .26        .31        .28        .174       .07
  .0174   .0023      .0005
  240.
  6.8     2.9        1.3        1.01       1.01       .61        .69
  .463    .34        .16        .19        .16        .0967      .0387
  .0093   .001       .0003
  225.
   5.8    2.5        1.1        .90        .81        .50        .58
   .348   .27        .12        .145       .12        .077        .029
   .0062  .00062     .00015

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the relative volatilities at the average temperature of the upper half of the column (313°F) were used for the plates above the feed and the relative volatilities at the average temperature of the lower half of the column (420°F) were used for the plates below the feed.

The heat of vaporization is also not constant throughout the column. It ranges from 11,350 Btu/lb-mole at the top to 16,200 Btu/lb-mole at the bottom. To model this, the heat of vaporization was assumed to change in a linear fashion from 13,000 Btu/lb-mole at the feed to its known value at either end of the column. (The vapor and liquid rates are more closely modeled when the heat of vaporization is not assumed to be constant.)

Compositions and flow rates of the feed, overhead product and bottoms product streams were obtained from the computer simulation supplied by Consumers Power Company. Only three of the 17 components of the feed are found in both the top and bottom products in significant quantities. Two of these, (the C_8 aromatics and the C_8 napthenes) were grouped together because of their quantities and the closeness of their volatilities. This left only two key components to be considered.

Table 8 shows the preheat's effect on each component and the total product flow rate in the column's bottom stream. Component 14 undergoes the greatest change (increasing by 0.65 lb-mole/hr). The total flow rate increased by 0.61 lb-mole/hr (less than 1/10 of one percent). This change is small and the effects are beneficial as the separation is slightly increased. The overhead product showed similar changes and was not tabulated.

TABLE 8. Effects of Preheat on Bottoms Product

Feed Temperature = 235°F
(Obtained from Distillation Program)

Bottoms Product

Component	With no Preheat (lb-moles/hr)	With 16.58 million Btu/hr Preheat (lb-moles/hr)
1. Butane	.00	00
2. Pentane	.00	00
3. C_6(P)	.00	.00
4. C_6(N)	.00	.00
5. Benzene	.00	.00
6. C_7(P)	.09	.09
7. C_7(N)	.00	.00
8. Toluene	.00	.00
9. C_8(P)	.90	.91
10. C_9(IP)	.77	.74
11. Nothing	-	-
12. C_8(A) and C_8(N)	60.80	60.80
13. 300/350	242.00	242.00
14. 350/400	208.43	209.08
15. 400/500	205.60	205.60
16. 500/600	43.80	43.80
17. 600	22.50	22.50
Total Bottoms	784.9	785.5

Product Flow Rate (lb-moles/hr)

Bottoms Product Flow Rate from C.P. Co. Computer Simulation = 783.8 lb-moles/hr

Table 9 summarizes the important results from the distillation program. Reboiler duty is decreased by 13.1 million Btu/hr as 79% of the preheat energy added shows up as energy savings. The other 3.48 million Btu/hr goes into increased condenser duty. The effects of the preheat on the internal flow rates are also shown. They are substantial in some regions of the column. Both vapor rates and liquid rates decrease by approximately 20% in the lower half of the column. Vapor rate increases only slightly in the upper half, but liquid rate increases by nearly 40%.

TABLE 9. Effects of Preheat on Splitter Column

	No Preheat (Feed Temp. 200°F)	16.58 million Btu/hr of Preheat (Feed Temp. 235°F)	% Change from no Preheat
Reboiler Duty (million Btu/hr)	53.42	40.32	-24.5
Reboiler Savings = <u>13.1 million Btu/hr</u>			
Condenser Duty (million Btu/hr)	49.19	52.67	+ 7.1
Extra Condenser Duty = <u>3.48 million Btu/hr</u>			
Vapor Rate (lb-moles/hr)			
Top of Column (Plate 1)	4334	4640	+ 7.1
Feed Plate (Plate 16)	3816	4051	+ 6.2
Bottom of Column (Plate 30)	3298	2489	-24.5
Liquid Rate (lb-moles/hr)			
Top of Column (Plate 1)	811	1115	+37.5
Feed Plate (Plate 16)	4823	3833	-20.5
Bottom of Column (Plate 30)	4025	3288	-19.8
Bottoms Product (lb-moles/hr)	784.9	785.5	+ .08
Overhead Product (lb-moles/hr)	3481.4	3480.8	- .02

```
C     DISTILATION  CALCULATIONS
      PROGRAM CRAIG (INPUT,OUTPUT)
      DIMENSION Y(17,40),X(17,40),RV(17),FXF(17),BXB(17),DXD(17),DEN(17)
      DIMENSION V(31),H(31)
      REAL LB,LA,LIQ(31)
C
C     RV=RELATIVE VOLATILITY
C     FXF=FEED IN LB-MOLES/HR.
C     BXB(I)=LB-MOLES OF I IN BOTTOMS   DXD(I)=LB-MOLES OF I IN TOP
C     Y=VAPOR COMPOSITION  FIRST INDEX IS COMPONENT, SECOND IS PLATE NO.
C     X=LIQUID COMPOSITION FIRST INDEX IS COMPONENT, SECOND IS PLATE NO.
C     VA=VAPOR RATE ABOVE   VB=VAPOR RATE BELOW
C     LA=LIQUID RATE ABOVE   LB=LIQUID RATE BELOW
C     HVAP=HEAT OF VAPORIZATION AT FEED PLATE IN  BTU/LB-MOLE
C     SUBC=AMOUNT OF SUBCOOLING OF FEED IN BTU/HR/LB-MOLE
C     NPA=NUMBER OF PLATES ABOVE FEED  NPB=NUMBER OF PLATES BELOW FEED
      DO 999 JKL=1,8
      READ 975,TEMP,SUBC,VAPOR
      PRINT 200
C
      FXF(1)=86.3  $  FXF(2)=1202.8  $  FXF(3)=566.5
      FXF(4)=182.0  $  FXF(5)=46.4  $  FXF(6)=344.6
      FXF(7)=312.0  $  FXF(8)=138.3  $  FXF(9)=263.7
      FXF(10)=32.7 $ FXF(11)=0. $ FXF(12)=290.6
      FXF(13)=319.4  $  FXF(14)=209.1  $       FXF(15)=205.6
      FXF(16)=43.8  $  FXF(17)=22.5
      FEED=4266.3
      HVAP=13000.
      VAMVB=VAPOR-FEED*(SUBC/HVAP)
      NPA=15 $ NPB=15
      H(15)=13000.
      DIFFB=(16200.-13000.)/(NPB-1.)
      DIFFT=(13000-11350.)/(NPA)
      VA=5000.
      DV=500.
      VB=VA-VAMVB-100.
C
C     DETERMINATION OF HVAP,PLATE TO PLATE, ASSUMED LINEAR
      DO 888 JJ=1,14
      KK=15-JJ
      JK=JJ+15
      H(KK)=H(KK+1)+DIFFB
888   H(JK)=H(JK-1)-DIFFT
      H(30)=H(15)-DIFFT*15.
      RO=0.0
      BXB(12)=60.8 $ BXB(11)=0. $ DXD(13)=77.4 $ DXD(11)=0.
      DO 3 L=1,10
3     BXB(L)=1.E-90
      DO 8 L=14,17
8     DXD(L)=1.E-90
      JF=31-NPA
      PRINT 400,TEMP,NPA,NPB
2     VB=VB+DV
1     DO 9 L=1,12
9     DXD(L)=FXF(L)-BXB(L)
      DO 12 L=13,17
12    BXB(L)=FXF(L)-DXD(L)
      D=0.0  $  B=0.0
      DO 13 I=1,17
      D=D+DXD(I)
13    B=B+BXB(I)
      VA=VB+VAMVB
      LB=VB+B
      LA=VA-D
      DO 15 I=1,17
      X(I,1)=BXB(I)/B
      Y(I,30)=DXD(I)/D
15    X(I,31)=Y(I,30)
C
      V(15)=VA
      DO 99 JJ=1,15
      JK=JJ+15
99    V(JK)=V(JK-1)*H(JK-1)/H(JK)
      V(14)=(V(15)-VAMVB)*H(15)/H(14)
      DO 88 JJ=1,13
      KK=14-JJ
88    V(KK)=V(KK+1)*H(KK+1)/H(KK)
      DO 77 JJ=1,14
```

```
77      LIQ(JJ+1)=V(JJ)+B
        DO 79 JJ=16,30
79      LIQ(JJ)=V(JJ-1)-D
C
C       BOTTOM TO FEED
C       RELATIVE VOLATILITY FOR BOTTOM OF COLUMN
        RV(1)=428. $ RV(2)=226. $ RV(3)=143. $ RV(4)=209. $ RV(5)=219.
        RV(6)=85.7 $ RV(7)=148. $ RV(8)=114. $ RV(9)=72.  $ RV(10)=62.
        RV(11)=1. $ RV(12)=54.8 $ RV(13)=39.8 $ RV(14)=22.6 $ RV(15)=9.05
        RV(16)=2.21 $ RV(17)=1.
        DO 20 J=1,NPB
        DENOM=0.0
        DO 10 I=1,17
        DEN(I)=RV(I)*X(I,J)
10      DENOM=DENOM+DEN(I)
        DO 16 I=1,17
        Y(I,J)=DEN(I)/DENOM
16      X(I,J+1)=(V(J)*Y(I,J)+BXB(I))/LIQ(J+1)
20      CONTINUE
C
C       TOP TO FEED
C       RELATIVE VOLATILITY FOR TOP OF COLUMN
        RV(1)=3529. $ RV(2)=1647. $ RV(3)=882. $ RV(4)=823. $ RV(5)=823.
        RV(6)=471. $ RV(7)=588. $ RV(8)=412. $ RV(9)=382. $ RV(10)=382.
        RV(11)=1. $ RV(12)=174. $ RV(13)=121. $ RV(14)=52.9 $ RV(15)=17.4
        RV(16)=2.74 $ RV(17)=1.0
        DO 22 JA=1,NPA
        J=31-JA
        DENOM=0.0
        DO 35 I=1,17
        DEN(I)=Y(I,J)/RV(I)
35      DENOM=DENOM+DEN(I)
        DO 40 I=1,17
        X(I,J)=DEN(I)/DENOM
40      Y(I,J-1)=(DXD(I)+LIQ(J)*X(I,J))/V(J-1)
22      CONTINUE
C
        DO 55 I=1,10
55      BXB(I)=BXB(I)*X(I,JF)/X(I,NPB)
        DO 60 I=14,17
60      DXD(I)=DXD(I)*X(I,NPB)/X(I,JF)
        DO 65 I=1,10
        IF(ABS(X(I,JF)/X(I,NPB)-1.0).GT..0001) GO TO 1
65      CONTINUE
        DO 63 I=14,17
        IF (ABS(X(I,NPB)/X(I,JF)-1.).GT..0001) GO TO 1
63      CONTINUE
        R=X(13,JF)/X(12,JF)/X(13,NPB)*X(12,NPB)
        IF (ABS(R-1.).LE..0005) GO TO 90
        DV=DV*(R-1.)/(RO-R)
        RO=R
C
        GO TO 2
90      PRINT 300,VA,VB,LA,LB,R
        PRINT 350
        PRINT 375,(NJ,NJ=1,17)
        DO 80 J=1,NPB
80      PRINT 100,(X(I,J),I=1,17)
        DO 95 J=JF,31
95      PRINT 100,(X(I,J),I=1,17)
        PRINT 850,B,D
        PRINT 900
        DO 920 I=1,17
920     PRINT 950,I,X(I,1),BXB(I),I,Y(I,30),DXD(I)
        PRINT 500
        DO 550 L=1,30
550     PRINT 600,L,H(L),V(L),LIQ(L)
        PRINT 700
        NPT=NPA+NPB
        FOV=VAPOR/FEED
        BOILD=H(1)*V(1)/(10.**6)
        CONDD=V(30)*H(30)/(10.**6)
        PRINT 800,NPT,TEMP,FOV,V(30),V(1),BOILD,CONDD
999     CONTINUE
100     FORMAT(" ",17(E8.3))
200     FORMAT("1",///,30X,* MULTICOMPONENT DISTILLATION BY CRAIG KENDZIOR
       1SKI*,///)
300     FORMAT(" ",//,5X,*VA=*,F9.3,5X,*VB=*,F9.3,5X,*LA=*,F9.3,5X,*LB=*,F
       19.3,5X,*R=*,F9.3,/)
350     FORMAT(" ",/////,52X,*PLATE TO PLATE COMPOSITIONS*,/,52X,*FROM BOT
       1TOM OF COLUMN TO TOP*,//)
```

```
375    FORMAT(4X,17(I2,6X),/)
400    FORMAT(" ",//////,20X,*FEED TEMPERATURE =*,F8.3,20X,*NPA=*,I2,20X,
      1*NPB=*,I2,/)
500    FORMAT("1",////,20X,*PLATE*,20X,*HVAP*,20X,*VAPOR*,20X,*LIQUID*,//
      1)
600    FORMAT(21X,I2,18X,F10.3,14X,F10.3,16X,F10.3,/)
700    FORMAT("1",//////,62X,*SUMMARY*,////)
800    FORMAT(20X,*TOTAL NUMBER OF PLATES = *,I2,//,20X,*FEED TEMPERATURE
      1 =*,F10.3,//,20X,*FRACTION OF VAPOR IN FEED =*,F5.3,//,20X,*VAPOR
      1RATE AT TOP =*,F10.3,//,20X,*BOIL-UP =*,F10.3,//,20X,*REBOILER DUT
      1Y (MMBTU/HR) =*,F10.3,//,20X,*CONDENSOR DUTY (MMBTU/HR) =*,F10.3,/
      1,"1",)
850    FORMAT("1",//,13X,*BOTTOMS PRODUCT  (LBMOLES/HR) =*,F10.3,15X,*OVE
      1RHEAD PRODUCT (LBMOLES/HR) =*,F10.3,///)
900    FORMAT(10X,*COMPONENT*,5X,*MOLAR CONC.*,5X,*AMOUNT*,20X,*COMPONENT
      1*,5X,*MOLAR CONC.*,5X,*AMOUNT*,//)
950    FORMAT(14X,I2,7X,E8.3,4X,F10.3,25X,I2,9X,E8.3,4X,F10.3,/)
975    FORMAT(3(F10.3))
       STOP
       END
   200.    991.       0.0
   215.    0.0        0.0
   220.    0.0        275.
   225.    0.0        522.4
   230.    0.0        750.
   235.      0.0      950.
   240.    0.0        1102.
   245.    0.0        1375.


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8. DESIGN CALCULATIONS

Physical Properties

<u>Splitter Bottoms Product</u> (373°F)

ρ_L = 41.28 lb/ft^2 *

μ_L = .254 Cp*

C_p = .65 Btu/(lb)(°F)$^{(5)}$

K_L = .071 Btu/(hr)(ft)(°F)$^{(5)}$

<u>Splitter Column Feed</u> (225°F)

ρ_L = 40.94 lb/ft^3 *

ρ_V = .43 lb/ft^3 *

μ_L = .223 Cp*

μ_V = .009 Cp*

k_L = .076 Btu/(hr)(ft)(°F)$^{(5)}$

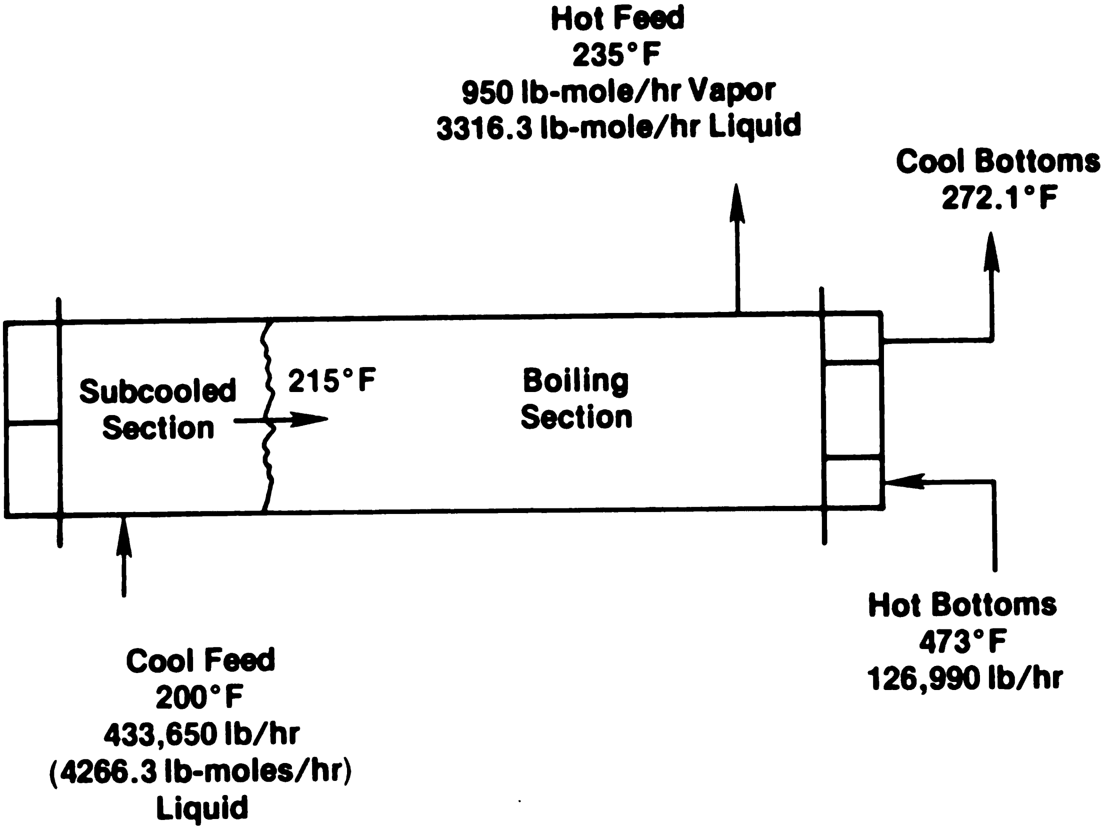
k_V = .0125 Btu/(hr)(ft)(°F)$^{(5)}$

λ_c = 13,000 Btu/lb-mole

ΔH_{vap} = 13,000 Btu/lb-mole

Note: Sources as referenced in Bibliography unless otherwise noted.

*From Computer Simulation.



Q Total = 16.58×10^6 Btu/hr

Q Subcool = 4.23×10^6 Btu/hr

Q Boiling = 12.35×10^6 Btu/hr

Figure 6 - Feed Preheat Exchanger

The feed enters the shell side of the preheat exchanger as a subcooled liquid at 200°F and exits a two phase mixture at 235°F. Based on Figure 5, (Vapor Content vs Feed Temperature) the exit conditions of the original 4266.3 lb-moles/hr (433,650 lb/hr) are:

950 lb-moles/hr	Vapor
3316.3 lb-moles/hr	Liquid

The amount of heat required to heat the subcooled feed from 200°F to its boiling point of 215°F is:

$$Q_{\text{subcooled}} = 433{,}650 \times .65\ (215-200)$$
$$= \underline{4.23 \times 10^6 \text{ Btu/hr}}$$

The amount of heat required for the partial vaporization is:

$$Q_{\text{boiling}} = 950 \times 13{,}000$$
$$= \underline{12.35 \times 10^6 \text{ Btu/hr}}$$

The total heat that must be transferred in the feed preheat exchanger is therefore:

$$Q_{\text{total}} = \underline{16.58 \times 10^6 \text{ Btu/hr}}$$

Inside Heat Transfer Coefficient (h_i)

Using the information from Table 5 (Feed Preheater Specifications) the tube cross-section and then the bulk velocity of the bottoms product are determined.

$$\text{Area} = (780/4)(.2223/144) = 0.301 \text{ ft}^2$$

$$U_b = 126{,}990/(41.28 \times .301)$$
$$= 10{,}220 \text{ ft/hr} = 2.84 \text{ ft/sec}$$

The Reynolds number indicates turbulent flow.

$$Re = \frac{D \times U_b \times \rho}{\mu}$$

$$= \frac{(.532/12)(10,220)(41.28)}{(.254 \times 2.42)} = 30,430 \; .$$

The Prandtl number is:

$$Pr = \frac{C_p \; \mu}{k} = \frac{(.65)(.254 \times 2.42)}{.071}$$

$$= 5.627.$$

Using the Dittus Boelter Equation:

$$\frac{h_i \; D}{k} = .023 \; (Re)^{.8} \; (Pr)^{.3}$$

h_i was found to be 238.8 Btu/(hr)(ft^2)(°F).

Outside Heat Transfer Coefficient (h_o)

This calculation is divided into two sections. In the first section of the heat exchanger the subcooled liquid feed is heated from 200°F to 215°F. In the rest of the heat exchanger partial vaporization of the feed takes place.

Subcooled Section

The shell side crossflow area (A_s) is first determined (see Kern$^{(5, \text{ p. } 138)}$),

$$A_s = \frac{ID \times C'B}{P_T \times 144} = \frac{(35)(.1875)(16)}{(1.0)(144)}$$

$$= .729 \text{ ft}^2$$

and then the mass velocity.

$$G_s = 433{,}650/.729 = 594{,}720 \text{ lb/(hr)(ft}^2).$$

The Equivalent Diameter (De) is found to be .0792 ft from Kern(5, p. 838). The shell side Reynolds number is then found.

$$Re_s = \frac{De\ G_s}{\mu} = \frac{(.0792)(594{,}720)}{(.223 \times 2.42)}$$

$$= 87{,}280\ .$$

The Prandtl number is:

$$Pr = \frac{C_p\ \mu}{k} = \frac{(.65)(.223 \times 2.42)}{.076} = 4.62\ .$$

This equation from Kern(5, p. 137) for liquid flow outside of tubes was used to determine h_o for this section.

$$\frac{h_o\ De}{k} = 0.36\ (Re_s)^{.55}\ (Pr)^{1/3}$$

$$h_o \text{ (subcooled)} = \underline{300.2 \text{ Btu/(hr)(ft}^2)(°\text{F)}}$$

Boiling Section

The average temperature-difference driving force across the film (Δt_f) is 121.3°F. This includes the correction for the inside heat transfer coefficient of 238.8 Btu(hr)(ft^2)(°F).

$$\Delta t_f = \text{Ave. Tube Wall Temperature-Ave. Boiling Temperature}$$

The outside heat transfer coefficient (h_o) was calculated using this equation from Peters and Timmerhaus(11, p. 546) for boiling liquids outside horizontal tubes.

$$h_o = 0.62\left[\frac{k_v^3\ \rho_v(\rho_L - \rho_v)\ g\ \lambda_c}{D_o\ \mu_v\ \Delta t_f}\right]^{1/4}$$

h_o (Boiling) = <u>113.4 Btu/(hr)(ft^2)(°F)</u>

<u>Overall Heat Transfer Coefficients</u> (U_o)

An overall fouling coefficient (h_f) of 300 Btu/(hr)(ft^2)(°F) was used as recommended by Peters and Timmerhaus(11, p. 550).

The corresponding U_o's were found as follows:

$$U_o = \frac{1}{\frac{1.0}{h_o} + \frac{1.0}{h_f} + \frac{.75}{(.532 \times h_i)}}$$

U_o(subcooled) = <u>79.6 Btu/(hr)(ft^2)(°F)</u>

U_o(Boiling) = <u>55.3 Btu/(hr)(ft^2)(°F)</u>

<u>Heat Transfer Area and Cost</u>

It is difficult to calculate the exact temperature driving force because of the multiple tube passes and the two distinct sections. A very conservative log mean temperature difference ($\Delta T_{\ell m}$) was used for both sections (see Preheater Description, Section 4). A 0.99 $\Delta T_{\ell m}$ correction factor was used to correct for the multiple tube passes (see Kern(5, p. 828).

$$\Delta T_{\ell m} = 0.99\ \frac{(473\text{-}235) - (272.1 - 215)}{\ell n\ \frac{(473\text{-}235)}{(272.1\text{-}215)}}$$

$$= 125.5\ °\mathrm{F}.$$

The required areas were then calculated for each section.

$$Q = (U_o)(A_o)(\Delta\ T_m)$$

A (Subcooled) = 423 ft^2

A (Boiling) = 1,780 ft^2

Total Required Heat Exchange Area = <u>2,203 ft^2</u>

Corresponding tube length for the 780, 3/4" tubes is 14.4 ft.

Estimated Cost of this heat exchanger is $9,700(10) (1958). Using the appropriate economic indicators for fabricated equipment(14) the January, 1974 cost is $24,400.

Tube Side Pressure Drop

The tube side Reynolds number was previously determined to be 30,430. The corresponding friction factor (f) is .0002 ft^2/in^2 (see Kern(5, p. 836)). The specific gravity of this fluid is:

$$S = 41.28/62.4 = 0.662\ .$$

The mass velocity is:

$$G = W/A_t = 126{,}990/.301 = 421{,}900\ lb/(hr)(ft^2).$$

This equation from Kern(5) was then used to calculate the pressure drop:

$$P_t = \frac{1}{2}\,\frac{f\ G^2\ L\ N}{5.22 \times 10^{10}\ D\ S}$$

$$= \frac{1}{2}\,\frac{(.002)(421{,}900)^2(14.4)(4)}{5.22 \times 10^{10}(.532/12)(.662)}\ .$$

Pressure Drop Tube Side = <u>0.67 psi</u>.

Shell Side Pressure Drop

This calculation must also be done for the two sections. The length of the exchanger where boiling takes place is:

$$14.4 \times (1780/2203) = 11.6 \text{ ft}$$

In the first 2.8 ft. heating of the subcooled liquid must take place.

Subcooled Section

The shell side Reynolds number was previously determined to be 87,280. This corresponds to a shell side friction factor of .0014 ft^2/in^2 (see Kern(5, p. 839)).

From before

$G_s = 594{,}720 \text{ lb/(hr)(ft}^2)$

$D_s = 35/12 = 2.92 \text{ ft}$

$De = .0792 \text{ ft}$

$S = 40.94/62.4 = 0.656$

The number of passes was then determined.

$$(N+1) = (2.8/16)12 = 2.1$$

This equation from Kern(5) was used to calculate the pressure drop:

$$\Delta P_s = \frac{f\, G_s^2\, D_s\, (N+1)}{5.22 \times 10^{10}\, D_e\, S}$$

ΔP_s (Subcooled Section) = $\underline{1.12 \text{ psi}}$.

Boiling Section

As before

G_s = 594,720 lb/(hr)(ft^2)

f = .0014 ft^2/in^2

D_s = 2.92 ft

D_e = .0792 ft

$(N+1) = (11.6/16)12 = 8.7.$

The average specific gravity is used here.

$S_{inlet} = 0.656$

$S_{outlet} = .43/62.4 = 0.03$

$S_{mean} = 0.343.$

Using the same equation as used in the subcooled section:

ΔP_s (Boiling Section) = $\underline{8.88 \text{ psi}}$.

The total pressure drop is simply the summation of these two.

Pressure Drop Shell Side = $\underline{10.0 \text{ psi}}$.

APPENDIX B

RANKINE CYCLE DESIGN USING SPLITTER OVERHEAD STREAM

Appendix B

Table of Contents

1. SUMMARY

The rankine cycle generates 1,800 Kw of electrical power from the 59.3 million Btu/hr received from the splitter overhead stream at 260°F. Efficiency of the cycle is 11%. Fluorinol-70 was chosen as the working fluid.

Installed cost of the equipment is $2,040,000. The electricity generated is valued at $425,300/year @ 3¢/Kw-hr. Estimated payback period is 9.32 years if operating costs are included but only 4.8 years if they are not. Actual payback period will be somewhere between the two depending on the operating costs of the air coolers presently used.

There are several disadvantages to this rankine cycle design. They are listed below:

1. Added Complexity
2. Substantial Payback Periods
3. Ability to Function Under Varying Conditions
4. Large Equipment Size
5. Large Amounts of Cooling Water Used (3540 gpm) Might Require Increased Cooling Tower Capacity

The rankine cycle can be designed to produce shaft power or to generate electricity if a generator is included. This design assumed a generator would be used. The rankine cycle is more effective if the shaft power is used directly since equipment costs are reduced by the generator cost and the cycle is more efficient without generator and motor inefficiencies. Since heat exchange

equipment makes up 60% of the capital cost of the rankine cycle and with its modular design, unit capital cost per Kw is independent of size (within certain limits). If desired then, a smaller rankine cycle design could be just as effective if sized to meet a particular need. Located beneath the splitter column where this cycle is designed is a 600 HP (450 Kw) HDS Pump. It is estimated that a 450 Kw rankine cycle (without generator) could be designed to power the HDS Pump only, with no effect on payback period.

2. BASIS OF DESIGN

This design is based on the availability of 59.3 million Btu/hr of waste heat from the splitter overhead stream at a temperature of 260°F. Only the heat available from the condensation of this stream is used and it is calculated with the plant operating at full capacity. The overhead stream is a multicomponent mixture of light hydrocarbons and will not condense at a constant temperature. Depending on feed composition and column operation, the splitter overhead temperature may range from 184-290°F. There exists very limited actual plant data, but it indicates the overhead temperature is normally greater than 250°F. In this design it is assumed that an average temperature of 260°F over this stream's condensing range can be obtained. In practice this might be difficult for some feedstocks, and require minor changes in column operation (such as less complete separation, or slight increase in column pressure).

A stream factor of .9 was used. Cooling water, and not air cooling is necessary to achieve the low temperatures required in the condenser for reasonable cycle efficiency. A design temperature for cooling water of 70°F was used. It was assumed that cooling water could be made available at a cost of 2¢/1000 gal. No allowances were made for enlargement of cooling water facilities.

Fluorinol-70 was chosen as the working fluid for the rankine cycle (see Fluid Selection). Cost of the Fluorinol-70 was neglected.

Turbine efficiency was taken as 80%, and a pump efficiency of 75% was used. All heat exchangers were designed to utilize 20 foot lengths.

A simple, single installation factor of 4.74 was used as recommended by Chilton.(2) Electricity was valued at 3¢/Kw-hr(6,12) for evaluation purposes, although the effects of other values were considered.

3. PROCESS DESCRIPTION

The flow diagram for the rankine cycle is schematically depicted in Figure 7. An enthalpy-pressure diagram for the Fluorinol-70 working fluid is shown in Figure 9 (Design Calculations Section). Numbered streams in this discussion refer to similar numbers on those figures.

Stream 1 consists of 227,000 lb/hr of subcooled liquid Fluorinol-70 at 48 psia and 120°F. It is divided into two equal streams and sent to identical vaporizers. Here a total of 17,186 ft^2 of heat transfer area is available to transfer the 59.3 million Btu/hr from the condensation of the splitter overhead stream at 260°F. The overhead stream will then be sent to the original air coolers to be further cooled to the desired temperature. The Fluorinol-70 leaves the vaporizers as a saturated vapor at 40 psia and 220°F. It is then recombined and sent to the turbine.

Stream 2 enters the turbine at 40 psia and is expanded to 4 psia and 120°F. A small amount of condensation (3.8%) takes place in the turbine. Turbine efficiency is 80% and the turbine will deliver 1856 Kw of shaft power while the 227,000 lb/hr of Fluorinol-70 undergoes an enthalpy drop of 27.94 Btu/lb. The turbine shaft power would best be used directly to power nearby equipment or it may be connected to a 97% efficient generator to produce electricity.

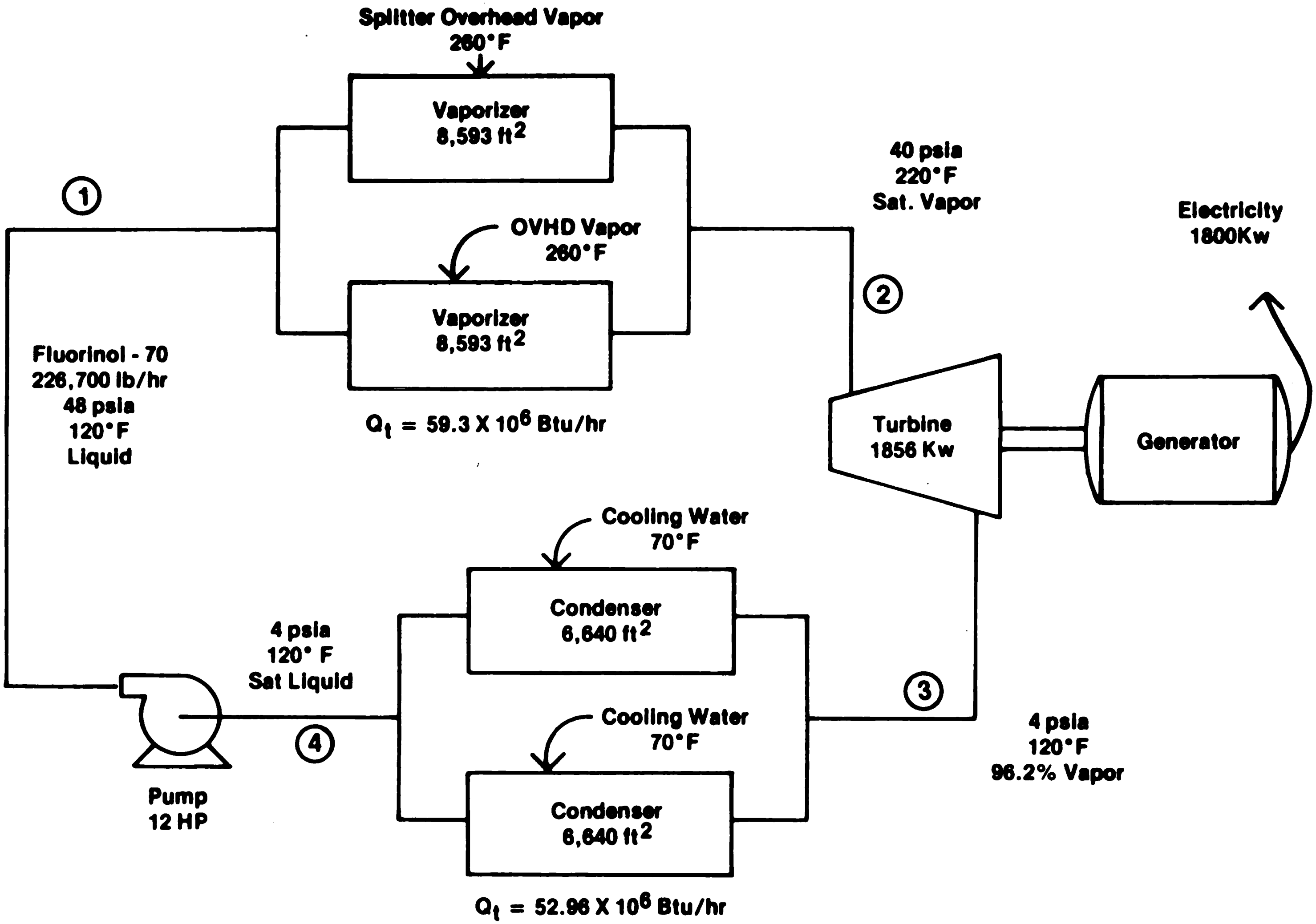


Figure 7 - Rankine Cycle Flow Diagram, (Fluorinol - 70)

The turbine exhaust (stream 3) is then sent to two identical condensers. A total of 13,280 ft^2 of heat transfer area is available to remove the 52.96 million Btu/hr of heat required for complete condensation. About 3540 gpm of cooling water at 70°F with a 30°F temperature rise is needed to accomplish this.

The Fluorinol-70 (now a saturated liquid at 4 psia and 120°F) is recombined into one stream and sent to a single 12 HP pump. Here the pressure is raised to 48 psia, and the cycle is ready to repeat itself.

4. EQUIPMENT DESCRIPTION

Condensers:

The two identical condensers combine to remove the 52.96 million Btu/hr required for complete condensation of the turbine exhaust vapors. 3540 gal/min of cooling water flowing through four passes on the tube side of the condenser is needed. Water velocity in the tubes is 3.3 ft/sec. The inside heat transfer coefficient (h_i) was calculated by the Dittus Boelter equation:

$$\frac{h_i D}{k} = .023\ (Re)^{.8}\ (Pr)^{.3}\ .$$

Using this equation h_i was calculated to be 807 Btu/(hr)(ft^2)(°F).

Fluorinol-70 is condensing on the shell side of the tubes at 120°F. Condensation actually takes place over the temperature range of 118°F - 122°F, but the average temperature was deemed acceptable for preliminary design purposes. Calculation of the outside heat transfer coefficient (h_o) is more difficult and uncertain than the calculation of h_i. This equation from Kern(5) for condensation on horizontal tubes was used.

$$h_o = 1.51\ \left(\frac{\mu}{k^3\ \rho^2 g}\right)^{-1/3}\ \left(\frac{4\ G''}{\mu}\right)^{-1/3}\ .$$

From this equation h_o was calculated to be 226.1 Btu/(hr)(ft^2)(°F).

This is much less than the 700-750 Btu/(hr)(ft^2)(°F) recommended by Halocarbons Company.$^{(4)}$

Inside and outside fouling coefficients of 1000 Btu/(hr)(ft^2)(°F) were used as recommended in Perrys.$^{(10)}$ The overall heat transfer coefficient (U_o) was calculated to be 123.3 Btu/(hr)(ft^2)(°F). This gave a required heat transfer area of 6,640 ft^2 for each condenser. Each condenser has 1268 tubes of one inch diameter on a 1¼" square pitch. Square pitch was chosen for cleaning and maintenance purposes. Carbon steel tubes and shells should be adequate for these fluids and temperatures.

Pressure drops were rather minimal, only 1.25 psi on the tube side and .1 psi on the shell side. Cost of both condensers is \$123,000$^{(11)}$ (January 1979). For more detailed specifications see Tables 10 and 11.

Vaporizers:

The two identical vaporizers are used in parallel to recover a total of 59.3 million Btu/hr available from the splitter overhead stream. Condensation of the overhead stream takes place on the tube side of the horizontal exchangers. Heating of the liquid Fluorinol-70 followed by its vaporization takes place on the shell side.

Calculation of h_i was done using the same equation as used in the calculation of h_o for the condenser, except with a change in the calculation of the loading per linear foot (G") as done by Kern$^{(5, p. 266)}$. Using that equation h_i was calculated to be 471.2 Btu/(hr)(ft^2)(°F).

TABLE 10. Condenser Description

Number of Heat Exchangers	2
Total Heat Duty	52.96 million Btu/hr
$\Delta T_{\ell n}$	32.74 °F
U_o	123.3 Btu/(hr)(ft^2)(°F)
Total Surface Area	13,280 ft^2
Material of Construction	
Shell	Carbon Steel
Tubes	Carbon Steel
Cost of Both Exchangers	123,000(11) (January 1979)

TABLE 11. Condenser Specifications (per unit)

Area	6640 ft^2
Tube length	20 ft
Number of Tubes	1268
BWG	14
OD	1"
ID	.834"
Pitch	1¼" square
Shell ID	54"
Number of Passes	
Tube Side	4
Shell Side	1
Baffle Spacing (25% cut)	36"

Tube Side	Shell Side
Fluid: Cooling Water	Fluid: Fluorinol-70 (vapor in)
Flow Rate: 1,770 gpm	Flow Rate: 113,350 lb/hr
Inlet Temp.: 70 °F	Inlet Temp.: 120 °F
Outlet Temp.: 100 °F	Outlet Temp.: 120 °F
Pressure Out: -	Pressure Out: 3.7 psi
Pressure Drop: 1.25 psi	Pressure Drop: .1 psi
	Condensation

Boiling of the Fluorinol-70 takes place over a range of 2.6°F. Again the average of 220°F was used. Flow rate in pounds per hour of the Fluorinol-70 is only half that of the overhead stream in the tubes. To make h_o as large as possible, baffle spacing was kept very small, only 6" throughout the exchanger. This is less than the minimum recommended spacing of 10 inches,(5,10) but it is necessary and not impractical with the given flow rates. For the liquid section h_o was calculated to be 122.5 Btu/(hr)(ft^2)(°F) by this equation from Kern.(5)

$$h_o = J_h \frac{k}{D_e} (Pr)^{1/3} .$$

This equation can also be used for the section where vaporization takes place. Here h_o was found to be 129.4 Btu/(hr)(ft^2)(°F).

Fouling coefficients of 1000 Btu/(hr)(ft^2)(°F) were again used. This gave a U_o of 77.5 Btu/(hr)(ft^2)(°F) in the liquid section, and 80.2 Btu/(hr)(ft^2)(°F) in the boiling section. Total required heat transfer area is 17,185 ft^2, of which 89% is for the actual vaporization of the Fluorinol-70. Pressure drop on the tube side is a negligible .08 psi. Due to the small baffle spacing pressure drop on the shell side is 7.4 psi. Total cost of both vaporizers is $142,000(11) (January 1979). See Tables 12 and 13 for more specific details.

Turbine:

See Table 14 for turbine description. In the turbine 1856 Kw (2488 HP) of power are generated by the expansion of 226,700 lb/hr

TABLE 12. Vaporizer Descriptions

Number of Heat Exchangers	2
Total Heat Duty	59.3 million Btu/hr
ΔT_{lm}	
Subcooled Section	79.8°F
Boiling Section	40°F
U_o	
Subcooled Section	77.5 Btu/(hr)(ft^2)(°F)
Boiling Section	80.2 Btu/(hr)(ft^2)(°F)
Total Surface Area	17,185 ft^2
Material of Construction	
Shell	Carbon Steel
Tube	Carbon Steel
Cost of Both Exchangers	$142,000(11) (January 1979)

TABLE 13. Vaporizer Specifications (per unit)

Area	8,593 ft^2
Tube length	20 ft
Number of Tubes	1641
BWG	14
OD	1"
ID	.834"
Pitch	1¼" square
Shell ID	60"
Number of Passes	
Tube Side	1
Shell Side	1
Baffle Spacing (25% cut)	6"

Tube Side	Shell Side
Fluid: Splitter OVHD (Vapor In)	Fluid: Fluorinol-70 (Liquid In)
Flow Rate: 230,000 lb/hr	Flow Rate: 113,400 lb/hr
Inlet Temp.: 260°F	Inlet Temp.: 120°F
Outlet Temp.: 260°F	Outlet Temp.: 220°F
Pressure Out: 20 psig	Pressure Out: 39.7 psia
Pressure Drop: .08 psi	Pressure Drop: 7.4 psi
Condensation	Boiling

TABLE 14. Turbine Description

Fluid	Fluorinol-70 Vapor
Flow Rate	226,700 lb/hr
Turbine Efficiency	80%
Enthalpy Change Across Turbine	27.94 Btu/lb
Pressure Change	35.3 psi
Turbine Power	1856 Kw 2488 HP
Value of Shaft Power (3¢/Kw-hr)	$436,900/year
Turbine Cost (January 1979)	$104,000(9)

of Fluorinol-70 from 40 psia to 4 psia. Enthalpy change of the working fluid is 27.94 Btu/lb. At 3¢/Kw-hr the value of the turbine shaft power is $436,000/year.

About 3.8% of the Fluorinol-70 will condense in the turbine. For this reason a lower turbine efficiency of 80% was assumed in the design. The axial flow turbine is recommended as the best type for this application because of its slower rotational speeds and ability to maintain high part load efficiency.(9)

Turbine cost was obtained by sizing up a very similar turbine used for the expansion of Fluorinol-85 in a rankine cycle by Thermoelectron Corporation in their detailed report (Reference 9, page 3-125). Selling price for their 6 stage, 928 Kw axial turbine of 83% efficiency was $51,000 (November 1974). Using this scale up equation with the normal .6 power factor;

$$\text{Turbine Cost (January 1979)} = 51{,}000\ \left(\frac{\text{Power (Kw)}}{928}\right)^{.6}\left(\frac{251.3}{186.1}\right)$$

the turbine cost was found to be $104,000 (January 1979). This price should be relatively conservative considering the higher efficiency and many stages in their turbine.

Generator:

The turbine power of 1856 Kw could either be used directly or converted to electric power. If there is no use for the turbine shaft power in the immediate vicinity of the splitter column then a generator will be needed.

With an efficiency of 97% electrical power from the generator is 1800 Kw. At 3¢/Kw-hr this electricity has a value of $425,700/year.

Cost of the generator is $59,800 and was estimated as an electric motor(10) of 1856 Kw. See Table 15 for more details.

TABLE 15. Generator Description

Efficiency	97%
Power Output	1800 Kw
Generator (3¢/Kw-hr)	$59,800(10) (January 1979)
Electricity Value	$425,700/year

Pump and Accessories:

Pumping requirements for the Fluorinol-70 are minimal, and only a 12 HP pump is needed. Pump is best kept separate from turbine, and powered by an electric motor. Total cost of pump and motor is only $1,600 (January 1979).

TABLE 16. Fluorinol-70 Pump Description

Fluorinol-70 Flow Rate	226,700 lb/hr (337 gpm)
Pressure in	4 psia
Pressure out	50 psia
Total Head	80 ft of Fluorinol-70
Pump Efficiency	75%
Pump Power	12 HP
Pump Cost	$1,300(11) (January 1979)
Motor Cost	$300(10) (January 1979)

No allowances were made for accessory equipment such as surge tanks or storage tanks. Their costs should be small and were not deemed essential for this preliminary design.

5. ECONOMIC ANALYSIS

Total cost for the major pieces of equipment (see Table 17) is $430,400. Most of this cost is for the large heat exchange equipment. Using a factor of 4.74(2) to take into account engineering, construction, contingency, and all other factors entering into the design and construction of a fluid process, gives an installed cost of $2,040,000. Value of the 1800 Kw of electricity generated at 3¢/Kw-hr is $425,800/year (see Table 18).

Variable costs total $35,700/year and most of this is cooling water costs. No cost was included in this design for any enlargement in cooling water facilities, if needed. Fixed costs are $171,300/year and include maintenance, materials, overhead, taxes and insurance. These large fixed costs equal over 40% of the value of the electricity produced. This cost has a very detrimental effect on the payback period.

Payback period including all direct costs is 9.23 years. Payback period ignoring operating costs is only 4.8 years. Essentially all of these operating costs must in some degree be charged against the air coolers presently in operation. The true payback period must lie somewhere between 4.8 and 9.23 years, depending on the present costs associated with the operation of the air coolers. The cost to operate the air coolers is probably not as large as

TABLE 17. Rankine Cycle Equipment Cost (January 1979)

Condensers	\$123,000(11)
Vaporizers	\$142,000(11)
Turbine	\$104,000(4)
Pump	\$ 1,300(11)
Pump motor	\$ 300(10)
Generator	\$ 59,800(10)
Total	\$430,400
Without Generator	\$370,600
Installed Cost (4.74(2))	\$2,040,000
Without Generator	\$1,757,000

TABLE 18. Rankine Cycle Economic Breakdown

Stream Factor = .9

Fixed Investment = $2,040,000

	Unit Price	Amount	Value ($/year)
Generated Power	(3¢/Kw hr)	1800 Kw	425,800

Operating Costs

Variable			Cost ($/year)
Pumping	(3¢/Kw-hr)	8.9 Kw	2,200
Water	(2¢/1000 gal)	3540 gpm	33,500
		Subtotal Variable Cost	35,700
Fixed			
Maintenance Labor (2% Fixed)			40,800
Supervisory (30% Maintenance Labor)			12,200
Maintenance Materials (2% Fixed)			40,800
Taxes and Insurance (2% Fixed)			40,800
Operating Supplies (10% Labor)			4,100
Overhead (80% Labor)			32,700
		Subtotal Fixed Cost	171,300

Total Operating Costs (35,700 + 171,300)	$207,000/year
Yearly Savings (425,800 - 207,000)	$218,800/year
Payback Period (2,040,000/218,800) (with Operating Costs)	9.32 year
Payback Period (2,040,000/425,800) (ignoring Operating Costs)	4.80 year

those for an expensive and much more complicated rankine cycle so the actual payback period is probably closer to 9.23 years. No salvage value was considered for the air coolers not needed with the rankine cycle.

Net yearly revenue from this process is $218,800 (including operating costs). Obviously the value of the produced electricity is of extreme importance. Figure 8 shows the effect of electrical value on payback period. Anything less than the 3¢/Kw-hr this design is based on quickly makes the payback period completely unreasonable.

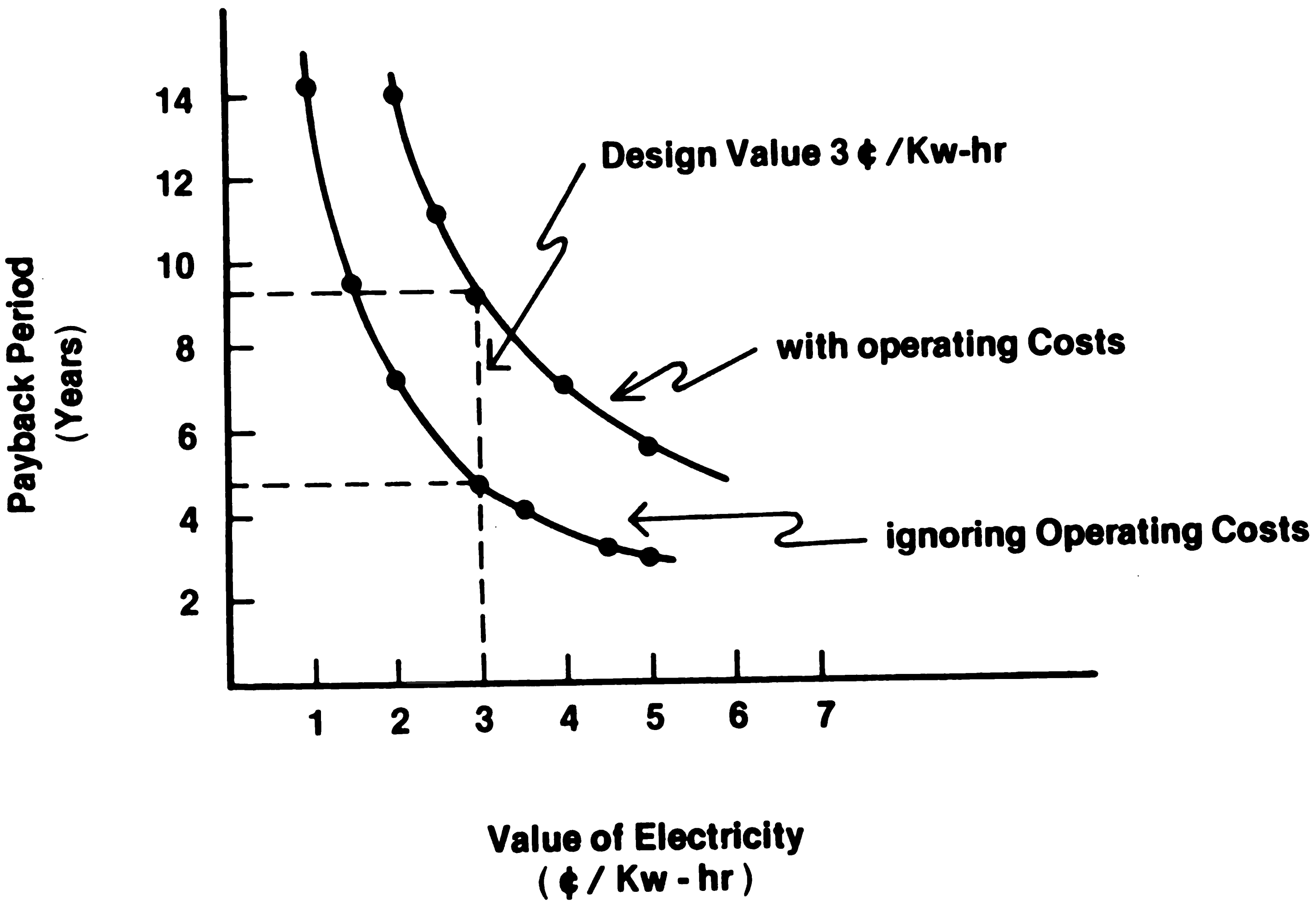


Figure 8 - Pay Period VS. Value of Electricity

6. FLUID SELECTION

Selection of the organic working fluid is the key step in the design of a rankine cycle process. Table 19 lists the major criteria to be considered.

TABLE 19. Fluid Selection Criteria

1) Toxicity
2) Flammability
3) Fluid Stability
4) Corrosiveness
5) Thermodynamics of Fluid
6) Heat Transfer Coefficients
7) Working Pressures
8) Molecular Weight
9) Availability and Accuracy of Thermodynamic Properties
10) Availability and Cost of Fluid

A mixture of trifluoroethanol and water, with a brand name of Fluorinol was chosen. Fluorinols are not extremely toxic (a major problem with many organics). Trifluoroethanol is not classified as toxic via the dermal or inhalation pathways,(7) nor is it a primary skin irritant.(4) Animal studies have shown trifluoroethanol to cause severe eye damage similar to many other organics, such as isopropyl alcohol and toluene.(4)

Fluorinols have been found to be very stable at temperatures much higher than those attained in this process.(4) It's compatability with most metals including carbon steel is good, with no significant corrosion problems, or effects on the Fluorinol fluid.(4)

Most organics, including Fluorinols, have lower values of thermal conductivity than water so heat transfer coefficients are normally not as good as those for water. Fluorinols have the advantage of lower working pressures. In this design, the maximum pressure is only 40 psia. This gives less expensive equipment design and added safety, as compared to fluids requiring much higher pressures.

Molecular weight is very important in simplifying turbine design. A turbine used with a high molecular weight fluid is smaller in size, needs less stages for the same enthalpy drop and is much cheaper than a turbine used with a lower molecular weight fluid.(13) Trifluoroethanol has a molecular weight of 100.

Fluorinols have good thermodynamic properties. Data on fluorinols is available and considered to be quite accurate. Since Fluorinols are a mixture of two different fluids, the right proportions can be used to meet the specific needs of the given cycle. One disadvantage of a mixture is that the dew point and bubble point are not exactly the same. This difference is only a few degrees, and the linear average was used.

The specific Fluorinol working fluid was chosen to meet the following turbine conditions:

1. Turbine inlet temperature of 220°F and exhaust temperature of 120°F (these temperatures determined in Optimization Section).
2. Turbine efficiency of 85%.
3. Turbine inlet from vaporizer available as saturated vapor.
4. Turbine exhaust should be a slightly superheated vapor (2-3 Btu/lb of superheat). This avoids any possible problems from liquid in the turbine and still gives good cycle efficiency.

Enthalpy-pressure charts supplied by Halocarbons Products Corporation(4) were used in the fluid selection. Fluorinol-70 best met these turbine conditions an estimated 2.5 Btu/lb of superheat in the turbine exhaust.

Later in the actual design, tables were used instead of the enthalpy-pressure charts. (The tables were reported to be "slightly" more accurate.) Actually, there are major discrepancies in some regions. It was then found that the turbine exhaust would contain some liquid, rather than being superheated vapor. Turbine efficiency was lowered to 80% to take this into account. Condensation of 3.8% of the vapor takes place in the turbine. This is small and was assumed to be allowable. Design could be redone with another Fluorinol like Fluorinol-85, and thus eliminating condensation with no major changes in the results. Table 20 contains a more complete description of Fluorinol-70.

TABLE 20. Fluorinol-70 Composition and Properties

Composition

70 mole percent trifluoroethanol

30 mole percent water

92.835 weight percent trifluoroethanol

7.165 weight percent water

Average molecular weight 77.443

Freezing point -40°F

Maximum recommended operating temperature 625°F

7. PARAMETRIC STUDIES AND OPTIMIZATION

The purpose of the optimization procedure was to find reasonable approach temperatures for the condenser and vaporizer. Emphasis was to keep the payback period low, but still produce as much power as possible.

To simplify the procedure, all capital costs except those of the three major pieces of equipment; condenser, vaporizer, and turbine, were ignored. Pumping, water costs, maintenance and operating costs also were ignored. Heat exchange area was valued at $\$10/ft^2$. Turbine cost was determined by scale up of a similar type turbine (see Turbine, Equipment Description Section). Turbine efficiency was taken as 85%. The enthalpy-pressure charts were used rather than the tables to simplify the procedure. In the actual design the tables were used and some major differences were found, as noted in the Fluid Selection Section. The use of the enthalpy-pressure charts has little effect on the outcome of the optimization, however. Approximate overall heat transfer coefficients of 60 Btu/(hr)(ft^2)(°F) for the vaporizer, and 130 Btu/(hr)(ft^2)(°F) for the condenser were used. These values agreed quite well with the actual values of 80.2 Btu/(hr)(ft^2)(°F) for the vaporizer, and 123.3 Btu/(hr)(ft^2)(°F) for the condenser. Pressure drop was assumed negligible, but its effects on the overall heat transfer coefficients should also be optimized for a final design.

As seen in Table 21 the condenser temperature was first set at 110°F and the vaporizer outlet temperature was varied. A vaporizer outlet temperature of 210°F was found to give the shortest payback period of 4.90 years. Next the vaporizer outlet temperature was set at 210°F and the condenser temperature was varied. A condenser temperature of 120°F was found to have the shortest payback period of 4.68 years. Increasing vaporizer temperature to 220°F was found to have no effect on the payback period so a vaporizer outlet temperature of 220°F and a condenser temperature of 120°F were chosen for the actual design.

TABLE 21. Optimization Results

		Vaporizer		Condenser			Turbine					
Vaporizer Temp. (°F)	Condenser Temp. (°F)	ΔT (°F)	Area (ft^2)	ΔT (°F)	Q (10^6 Btu/hr)	Area (ft^2)	Power (Kw)	Value at 3¢/Kw-hr ($1000)	Turbine Cost ($1000)	Total Equipment Cost ($1000)	Installed Cost ($1000)	Payback Period (years)
240	110	20	49,400	21.6	51.0	18,200	2455	580	124	799	3,788	6.54
220	110	40	24,700	21.6	51.7	18,400	2216	524	116	555	2,581	4.92
200	110	60	16,500	21.6	52.8	18,800	1857	439	104	458	2,169	4.94
190	110	70	14,100	21.6	53.5	19,100	1704	403	99	431	2,042	5.07
210	110	50	19,800	21.6	52.4	18,700	2019	478	110	494	2,342	4.90
210	120	50	19,800	32.7	53.1	12,500	1824	431	103	426	2,018	4.68
210	130	50	19,800	43.3	53.8	9,600	1620	383	96	390	1,846	4.82
220	120	40	24,700	32.7	52.4	12,300	2060	487	111	481	2,278	4.68

Note: Condenser and Vaporizer Costs 10 times area ($10/$ft^2$).

8. DESIGN CALCULATIONS

The energy available from the condensation of the splitter overhead stream was first calculated. This stream is a multicomponent mixture and the calculation is shown in Table 22. The total energy available is 59.3 million Btu/hr at the design conditions.

Thermodynamics

The information for Fluorinol-70 was obtained from Table 23 and can be seen in Figure 9. The purpose of this calculation is to determine the turbine exhaust conditions (enthalpy and quality).

$H_2 = 299.1 \qquad\qquad S_2 = .4614$

$$S_3(\text{ideal}) = S_2 = (X)S_3(\text{Vapor}) + (1 - X)S_3(\text{Liquid})$$

$$.4614 = (X)(.4896) + (1 - X)(.0702)$$

$$X_{\text{ideal}} = 93.3\%\ \text{vapor}$$

$$\begin{aligned} H_3(\text{ideal}) &= (X)H_3(\text{Vapor}) + (1 - X)H_3(\text{Liquid}) \\ &= (.933)(280.5) + (1 - .933)(37.55) \\ &= 264.2\ \text{Btu/lb} \end{aligned}$$

$$\begin{aligned} H_{\text{Turbine}} &= (\text{Turbine Efficiency})(\text{Ideal Enthalpy Change}) \\ &= (.8)(299.1 - 264.2) \\ &= \underline{27.94\ \text{Btu/lb}} \end{aligned}$$

TABLE 22. Determination of Energy Available in Condensation of Splitter Overhead Vapor at 260°F

ΔH (Condensation) = (Flow Rate)(M.W.)(ΔH/lb)

	Component	Flow Rate (lb-moles/hr)	Molecular Weight (lb/lb-mole)	Specific ΔH of Condensation (Btu/lb)	ΔH of Condensation (10^6 Btu/hr)
1	Butane	129.1	58	115	.86
2	Pentane	1804.2	72	120	15.59
3	C_6(P)	849.8	86	125	9.14
4	C_6(N)	273.0	84.2	125	2.87
5	Benzene	69.6	78.1	155	.84
6	C_7(P)	56.9	100	130	6.72
7	C_7(N)	468.0	98.2	130	5.97
8	Toluene	207.5	92.2	150	2.87
9	C_8(P)	395.6	114	130	5.86
10	C_8(N)	99.8	112.2	145	1.62
11	C_8(A)	245.0	106.2	145	3.77
12	C_9(IP)	49.0	128	145	.91
13	300/350	116.1	135.5	145	2.28
Total		5223.6 lb-mole/hr (456,00 lb/hr)		59.3 × 10^6 Btu/hr	

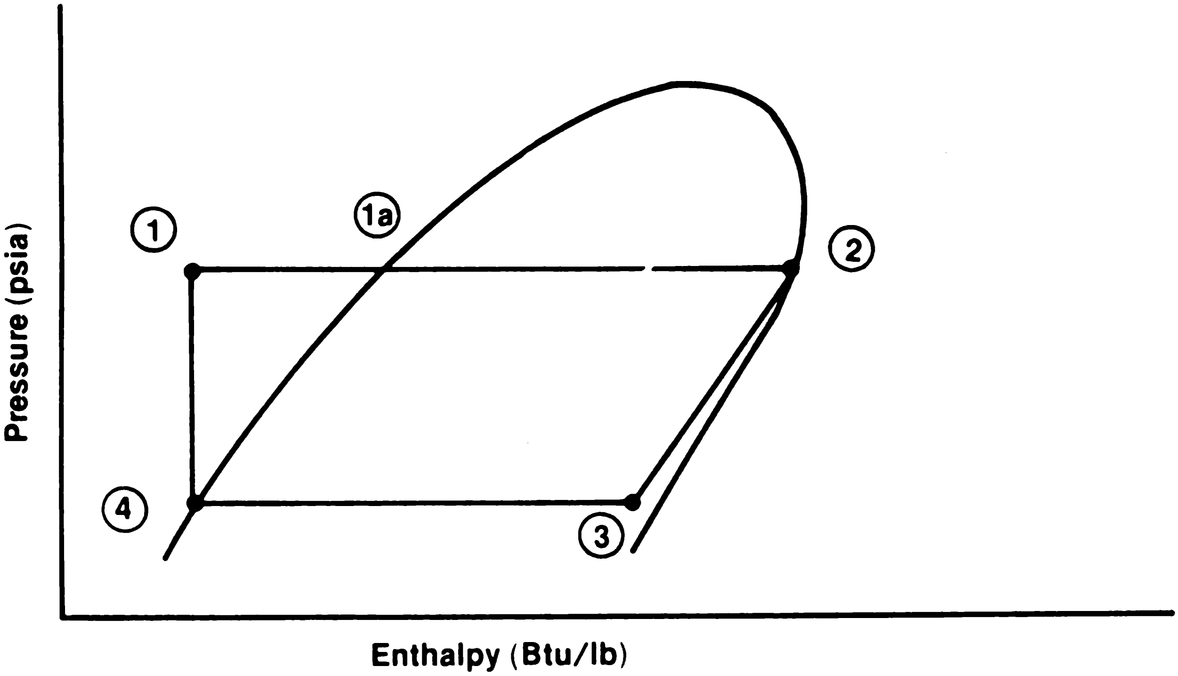
Average M.W. 88.06

TABLE 23. Fluorinol-70

Saturated Liquid and Saturated Vapor Properties

Pressure	Temp. °F		Sat. Liq.	Enthalpy BTU/lb		Entropy BTU/lb °R		Volume ft^3/lb	
psia	Liquid	Vapor	Spec. Heat	Liquid	Vapor	Liquid	Vapor	Liquid	Vapor
1	73.1	78.0	.4384	16.81	271.27	.0328	.5085	.0114	75.64
4	118.4	122.8	.4883	37.83	280.62	.0707	.4893	.0119	20.20
7	139.5	143.5	.5070	48.35	284.79	.0885	.4820	.0121	11.87
10	154.0	157.8	.5151	55.77	287.60	.1008	.4775	.0123	8.427
14.7	170.6	174.1	.5310	64.49	290.71	.1148	.4728	.0125	5.832
20	184.8	188.0	.5407	72.11	293.29	.1267	.4691	.0127	4.335
30	204.6	207.5	.5536	82.96	296.76	.1433	.4644	.0130	2.930
40	219.7	222.3	.5630	91.39	299.28	.1558	.4612	.0132	2.214
50	232.0	234.4	.5705	98.37	301.25	.1659	.4587	.0134	1.778
60	242.5	244.6	.5768	104.36	302.86	.1745	.4567	.0136	1.487
70	252.1	253.5	.5826	109.92	304.22	.1823	.4551	.0138	1.277
80	259.8	261.7	.5873	114.47	305.41	.1886	.4536	.0140	1.117
90	267.3	268.9	.5920	118.86	306.44	.1946	.4524	.0141	.9944
100	274.1	275.6	.5963	122.89	307.36	.2001	.4513	.0143	.8942
120	286.2	287.6	.6043	130.18	308.94	.2099	.4493	.0146	.7428
140	296.9	298.1	.6118	136.70	310.22	.2185	.4476	.0149	.6343
160	306.5	307.6	.6189	142.61	311.31	.2262	.4462	.0152	.5519
180	315.3	316.2	.6259	148.03	312.23	.2332	.4449	.0154	.4877
200	323.3	324.1	.6326	153.05	313.03	.2395	.4437	.0157	.4361
250	340.9	341.4	.6493	164.34	314.54	.2536	.4411	.0163	.3436
300	356.0	356.4	.6663	174.32	315.49	.2658	.4389	.0169	.2808
350	369.4	369.6	.6846	183.31	316.26	.2765	.4368	.0176	.2357
400	381.3	381.6	.7041	191.62	316.63	.2863	.4348	.0182	.2014
450	392.4	392.5	.7266	199.50	316.72	.2954	.4329	.0189	.1746
500	402.5	402.6	.7507	206.96	316.58	.3039	.4309	.0196	.1529
550	411.9	411.9	.7791	214.12	316.21	.3119	.4290	.0204	.1351
600	420.6	420.7	.8113	221.08	315.63	.3196	.4270	.0213	.1201
650	428.9	428.9	.8535	227.91	314.69	.3271	.4247	.0222	.1071
700	436.8	436.7	.9038	234.36	313.30	.3347	.4221	.0233	.0953
750	444.3	444.1	.9664	241.81	311.75	.3421	.4195	.0245	.0852
800	451.4	451.1	1.051	248.94	309.64	.3497	.4163	.0260	.0760
850	458.0		1.219	256.17		.3574		.0276	
900	464.7		1.474	265.18		.3669		.0303	
957.7	471.6		1.740	276.27		.3785		.0343	

Reproduced from "Fluorinol Data", Halocarbons Product Corporation
Reference 4.



	Pressure (psia)	Temperature (°F)	Enthalpy (Btu/lb)	Entropy [Btu/(lb)(°F)]
1	39.3	120	37.55	-
1a	39.3	220	90.8	.0132
2	39.3	220	299.1	.4614
3	4.0	120	271.2	.4737
4	4.0	120	37.55	.0119

Figure 9 - Enthalpy - Pressure Diagram

$H_3(\text{actual}) = 299.1 - 27.94 = 271.16$

$H_3 = (X)\ H_3(\text{Vapor}) + (1 - X)\ H_3(\text{Liquid})$

$271.16 = (X)\ (280.5) + (1 - X)(37.55)$

$X = 96.2\%$ Vapor

Turbine exhaust contains 3.8% liquid.

The required flow rate (W) of Fluorinol-70 is then easily determined.

$W = 59.3 \times 10^6/(299.1 - 37.55) = 226{,}700$ lb/hr

The Carnot efficiency of a cycle operating between 220°F and 120°F is 14.7%. The efficiency of this cycle is 10.7%.

Vaporizer Calculations

The two identical vaporizers are described in Tables 12 and 13 (Equipment Description Section). The heat duty and flow rates are evenly divided and these values are seen on Figure 10.

Physical Data

Splitter Overhead (260°F)

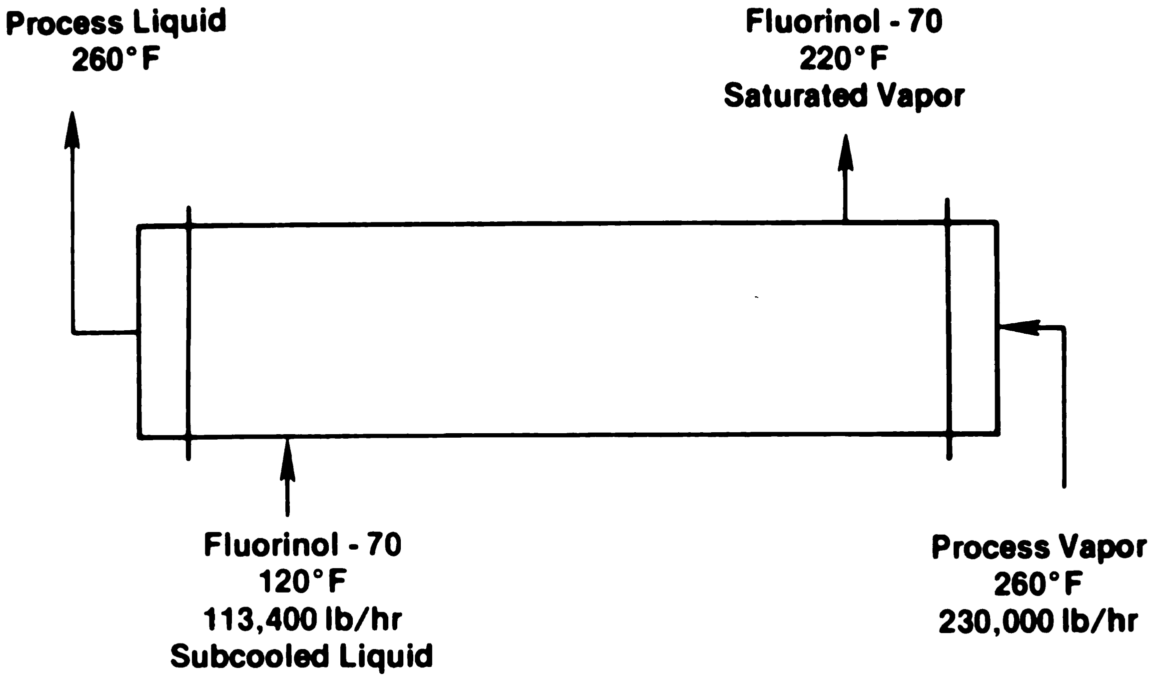
$\rho_L = 41.32$ lb/ft^3 *

$\mu_L = .14$ cp*

$\mu_V = .009$ cp$^{(8)}$

$\rho_V = .43$ lb/ft^3 *

$k_L = .0746$ Btu/(hr)(ft)(°F)



Q / exchanger = 26.65 X 10^6 Btu/hr

Figure 10 - Fluorinol - 70 Vaporizer

Fluorinol-70 (220°F)

ρ_L = 75.76 lb/ft^3 (4)

ρ_V = .442 lb/ft^3 (4)

μ_L = .5 cp$^{(4)}$

μ_V = .013 cp$^{(4)}$

k_L = .072 Btu/(hr)(ft)(°F)$^{(4)}$

k_V = .0134 Btu/(hr)(ft)(°F) (mixing law using water and polar organic)

λ_c = 208.3 Btu/lb$^{(4)}$

(170°F)

μ_L = .65 cp$^{(4)}$

k_L = .075 Btu/(hr)(ft)(°F)$^{(4)}$

Cp = .54 Btu/(lb)(°F) (derived from Tables,$^{(4)}$, $C_p = \frac{dH}{dT}$).

Note: As referenced in Bibliography unless otherwise noted.

*Computer Simulation.

Determination of h_i

The linear loading is calculated as in Kern$^{(5,\ p.\ 265)}$. This equation is for loading on a single horizontal tube which should approximate the loading inside of horizontal tubes.

$$G'' = \frac{W}{L\ N_t} = \frac{230{,}000}{(20)(1610)} = 7.14 \text{ lb/(hr)(ft)}.$$

This equation from Kern$^{(5,\ p.\ 265)}$ was used to calculate h_i:

$$h_i = 1.51 \left[\frac{\mu^2}{k^3\ \rho^2\ g}\right]^{-1/3} \left[\frac{4\ G''}{\mu}\right]^{-1/3}$$

$$h_i = 1.51 \left[\frac{(.14 \times 2.42)^2}{(.0746)^3(41.32)^2(4.15 \times 10^8)}\right]^{-1/3} \left[\frac{4\ (7.14)}{(.14 \times 2.42)}\right]^{-1/3}$$

$$h_i = 471.2 \text{ Btu/(hr)(ft}^2\text{)(°F)}$$

Determination of h_o

Two h_o s must be calculated, one for the subcooled section and one for the section where vaporization takes place.

Subcooled Section

The mass velocity is calculated as in Kern$^{(5,\ p.\ 138)}$.

$$G_s = \frac{W}{A_s} = \frac{113{,}400}{(60)(.25)(6)/(144 \times 1.25)}$$

$$= 226{,}800\ \mathrm{lb/(hr)(ft^2)}.$$

The shell side Reynolds number may now be calculated. The equivalent diameter is found in Kern$^{(5,\ p.\ 838)}$.

$$Re_s = \frac{G_s D_e}{\mu_L} = \frac{(226{,}800)(.99/12)}{(.65 \times 2.42)}$$

$$= 11{,}900.$$

With this Reynolds number the dimensionless heat transfer factor (J_h) is 60 (see graph in Kern$^{(5,\ p.\ 838)}$). This equation from Kern$^{(5)}$ is then used to calculate h_o.

$$h_o = J_h\ k/D_e\ \left(\frac{C_p\ \mu}{k}\right)^{1/3}$$

h_o(subcooled section) = $\underline{122.5\ \mathrm{Btu/(hr)(ft^2)(°F)}}$.

Boiling Section

This calculation parallels the subcooled section, except that all physical constants are at 220°F instead of 170°F. As before:

$$G_s = 226{,}800\ \mathrm{lb/(hr)(ft^2)}$$

$$Re = \frac{(226{,}800)(.99/12)}{(.5 \times 2.42)} = 15{,}500 \ .$$

The corresponding J_h is 71 and

$$h_o = 71 \left[\frac{.072}{(.99/12)}\right]\left[\frac{(.54)(.5 \times 2.42)}{.072}\right]^{1/3}$$

$$h_o(\text{Boiling Section}) = \underline{129.4 \text{ Btu/(hr)(ft}^2\text{)(°F)}}.$$

Required Areas

Outside and inside fouling coefficients of 1000 Btu/(hr)(ft^2)(°F) were selected as recommended in Perry.(10) The overall heat transfer coefficients for both sections were then calculated.

$$U_o = \frac{1}{\frac{1}{h_o} + \frac{1}{h_{of}} + \frac{1}{(.832)h_i} + \frac{1}{(.832)h_{if}}}$$

$$U_o(\text{Subcooled Section}) = \underline{77.5 \text{ Btu/(hr)(ft}^2\text{)(°F)}}$$

$$U_o(\text{Boiling Section}) = \underline{80.2 \text{ Btu/(hr)(ft}^2\text{)(°F)}}.$$

Area of Subcooled Section

The temperature driving force in this section is:

$$\Delta T_{\ell n} = \frac{(260 - 220) - (260 - 120)}{\ell n \left[\frac{260-220}{260-120}\right]} = 79.8°\text{F} \ .$$

The amount of heat transfer in this section by both vaporizers is:

$$Q = 226{,}700\ (90.8 - 37.55) =$$

$$Q(\text{Subcooled}) = 12.06 \times 10^6 \text{ Btu/hr} \ .$$

The required area is:

$$Q(\text{Subcooled}) = U_o\ A_o\ \Delta T_{\ell n}$$

$$A_o(\text{Subcooled}) = \underline{1{,}950 \text{ ft}^2}.$$

Area of Boiling Section

The temperature driving force in this section is a constant 40°F. The rest of the heat (42.24×10^6 Btu/hr) is transferred in this section (total vaporization takes place).

$$Q(\text{Boiling}) = U_o A_o \Delta T$$

$$A_o(\text{Boiling}) = \underline{14{,}731 \text{ ft}^2}.$$

The total required heat transfer area is 16,681 ft^2. The area available in the two heat exchangers is $\underline{17{,}185 \text{ ft}^2}$. This extra 3% was ignored.

Shell Side Pressure Drop

This calculation is also divided into two sections. In the first 2.3 ft. heating of the subcooled liquid occurs and in the next 17.7 ft. complete vaporization occurs.

Subcooled Section

As before:

$$G_s = 226{,}800$$

$$Re_s = 11{,}900.$$

With this Reynolds number the friction factor is .002 ft^2/in^2 (from Kern(5, p. 839)). The number of passes is:

$$(N+1) = 2.3/.5 = 4.6\ .$$

The specific gravity of this fluid is 1.214. The pressure drop was calculated with this equation from Kern.(5)

$$\Delta P_s(\text{subcooled section}) = \frac{f\ G_s^2\ D_s\ (N+1)}{5.22 \times 10^{10}\ D_e\ S} = .5 \text{ psi.}$$

Boiling Section

This calculation parallels the subcooled section.

$G_s = 226{,}000$

$Re_s = 15{,}500 \Rightarrow f = .002 \text{ ft}^2/\text{in}^2$

$(N+1) = 17.7/.5 = 35.4.$

The average specific gravity is used.

$S_{inlet} = 1.214$

$S_{outlet} = .0071$

$S_{ave} = .6105$

$\Delta P_s(\text{Boiling Section}) = 6.9$ psi.

Total Shell Side Pressure Drop = 7.4 psi.

Tube Side Pressure Drop

The total tube cross-sectional area is 6.23 ft^2/exchanger. The mass velocity in the tubes is now calculated.

$$G = W/A_t = 230{,}000/6.23 = 36{,}900 \text{ lb/(hr)(ft}^2\text{)} .$$

The tube side Reynolds number can now be calculated.

$$Re = \frac{D\ G_t}{\mu} = \frac{(.834/12)(36{,}900)}{(.009 \times 2.42)} = 117{,}800.$$

The tube side friction factor is .00015 ft^2/in^2 (Kern(5, p. 836)). The specific gravity of the process vapor is:

$$S = .43/62.4 = .0067.$$

This equation from Kern$^{(5)}$ for tube side pressure drop was used:

$$\Delta P_t = \frac{1}{2}\,\frac{f\ G^2\ L\ N}{5.22 \times 10^{10}\ D\ S} \; .$$

Pressure Drop Tube Side = .08 psi.

Condenser Calculations

The two identical condensers are described in Tables 10 and 11 (Equipment Description Section). The heat duty and flow rates are evenly divided between the two condensers.

Physical Data

Water (85°F)

ρ_L = 62.2 lb/ft^3 *

ρ_L = .8 cp*

k_L = .363 Btu/(hr)(ft)(°F)$^{(1)}$

C_p = 1.0 Btu/(lb)(°F)$^{(1)}$

Fluorinol-70 (120°F)

ρ_L = 84 lb/ft^3 $^{(4)}$

μ_L = .90 cp$^{(4)}$

μ_V = .013 cp$^{(4)}$

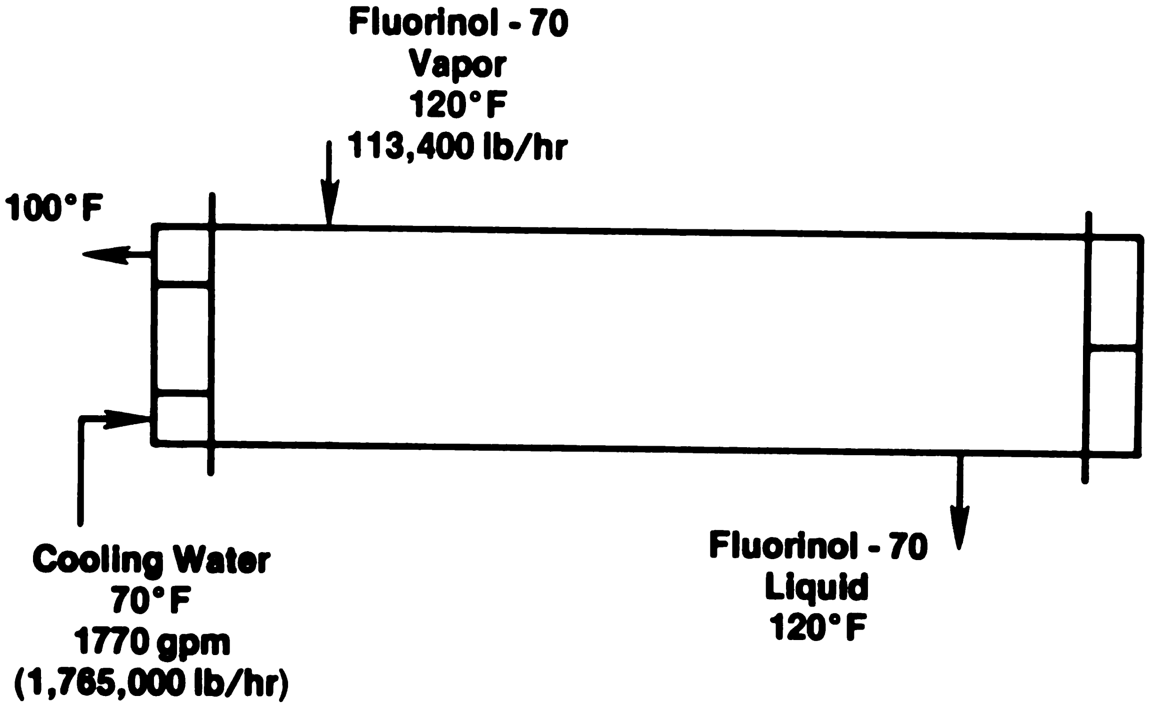
k_L = .0785 Btu/(hr)(ft)(°F)$^{(4)}$

*Steam Tables.

The total amount of heat transferred by both condensers is:

$$Q = 226{,}700\ (271.2 - 37.55) = 52.96 \times 10^6 \text{ Btu/hr} \; .$$

The total water flow rate is:



Q / exchanger = 26.48 X 10^6 Btu/hr

Figure 11 - Fluorinol - 70 Condenser

$$W = \frac{52.96 \quad 10^6}{(1)(100-70)} = 1.765 \times 10^6 \text{ lb/hr}$$

$$= 28{,}380 \text{ ft}^3\text{/hr}$$

$$= 3540 \text{ gpm.}$$

Determination of h_i

The tube side velocity is:

$$U_b = \frac{28{,}380}{(.834/24)^2(1268)(2/4)} = 11{,}800 \text{ ft/hr} = 3.3 \text{ ft/sec.}$$

The Reynolds number is in the turbulent region.

$$Re = \frac{D\, U_b\, \rho}{\mu} = \frac{(.834/12)(11{,}800)(62.2)}{(.8 \times 2.42)} = 26{,}350 \; .$$

The Prandlt number is:

$$Pr = \frac{C_p\, \mu}{k} = \frac{(1)(.8 \times 2.42)}{.363} = 5.33 \; .$$

Using the Dittus Boelter equation for heating of liquids, h_i was calculated.

$$\frac{h_i\, D}{k} = .023\, (Re)^{.8}\, (Pr)^{.4}$$

$$h_i = \underline{807 \text{ Btu/(hr)(ft}^2\text{)(°F)}}$$

Determination of h_o

The linear loading (G") is calculated as in Kern(5, p. 266)

$$G'' = \frac{W}{L\, N_t^{2/3}} = \frac{113{,}400}{(20)(1268)^{2/3}} = 48.4 \text{ lb/(hr)(ft)} \; .$$

This equation from Kern(5, p. 265) was used to calculate h_o

$$h_o = 1.51 \left[\frac{\mu^2}{k^3 \rho^2 g} \right]^{-1/3} \left[\frac{4\ G''}{\mu} \right]^{-1/3}$$

$$h_o = \underline{226.1 \text{ Btu/(hr)(ft}^2\text{)(°F)}}.$$

Required Area

Outside and inside fouling coefficients of 1000 Btu/(hr)(ft^2)(°F) were selected as recommended in Perry.(10) U_o was then calculated for the condenser.

$$U_o = \frac{1}{\frac{1}{226.1} + \frac{1}{1000} + \frac{1}{(.832)(807)} + \frac{1}{(.832)(1000)}}$$

$$= \underline{123.3 \text{ Btu/(hr)(ft}^2\text{)(°F)}}.$$

The temperature driving force in the condenser is 32.74°F. The correction for multiple tube passes is 1.0 at these temperatures. The total required area is:

$$52.96 \times 10^6 = 123.3\ A_o\ 32.74$$

$$\text{Required Area} = 13{,}119 \text{ ft}^2 .$$

The available area is 13,280 ft. This extra 1.2% was ignored.

Pressure Drop Shell Side

The mass velocity is:

$$G_s = W/A_s = \frac{113{,}400}{(54)(.25)(54)/(144 \times 1.25)} = 2800 \text{ lb/(hr)(ft}^2\text{)} .$$

The shell side Reynolds number is:

$$Re = \frac{(.99/12)(2800)}{(.013 \times 2.42)} = 7{,}340 .$$

This gives a friction factor of .00205 ft^2/in^2. The number of passes are:

$$(N+1) = (20)(12)/(54) = 4.44.$$

The specific gravity (S) of the Fluorinol-70 vapor is .00077. The shell side pressure drop is calculated using the same equation found in the Vaporizer Shell Side Calculation.

$$\Delta P_s = \frac{(.00205)(2800)^2(54/12)(4.44)}{5.22 \times 10^{10}\ (.99/12)(.00077)}$$

$$= \underline{.1\ \text{psi}}.$$

Pressure Drop Tube Side

With the already determined Reynolds number of 26,350, the corresponding friction factor is .00021 ft^2/in^2 (Kern(5, p. 836)). The specific gravity of water at 85°F is .997. The mass velocity is:

$$G = W/A_t = \frac{(1,765,000/2)}{(1268/4)(\pi)(.834/24)^2} = 734,000\ \text{lb/(hr)(ft}^2)\ .$$

This equation from Kern(5) for tube side pressure drop was used.

$$\Delta P_t = \frac{1}{2}\,\frac{f\ G^2\ L\ N}{5.22 \times 10^{10}\ D\ S} = \frac{1}{2}\,\frac{(.00021)(734,000)^2(20)(4)}{5.22 \times 10^{10}(.834/12)(.997)}$$

$$= \underline{1.25\ \text{psi}}.$$

Pump

The work required for the 75% efficient pump pictured in Figure 12 is:

$$W_s = \left[\frac{(50-4) \times 144}{84}\right]\left(\frac{1}{.75}\right)(226{,}700) = 2.36 \times 10^7 \text{ ft-lb/hr}$$

$$= \underline{12 \text{ HP}}$$

$$= \underline{8.88 \text{ Kw}} \; .$$

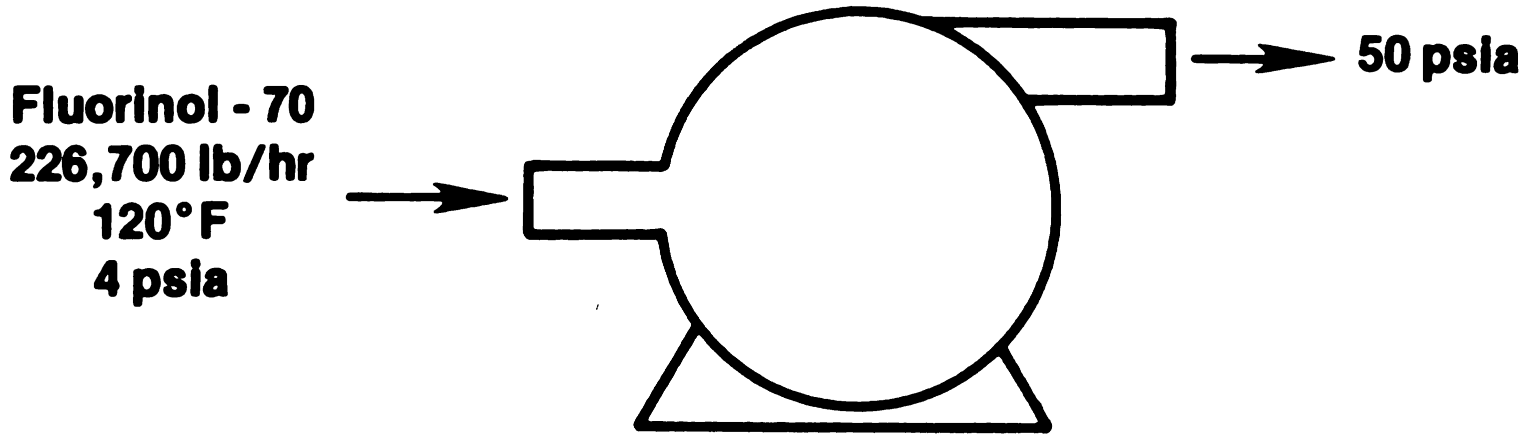


Figure 12 - Fluorinol - 70 Pump

Turbine and Generator

The enthalpy change across the 80% efficient turbine was already calculated to be 27.94 Btu/lb. The turbine power is:

$$\text{Power} = W \quad H_T = 6.334 \times 10^6 \text{ Btu/hr}$$

$$= \underline{1856 \text{ Kw}}$$

$$= \underline{2488 \text{ HP}}$$

Power output from the 97% efficient Generator is 1800 Kw.

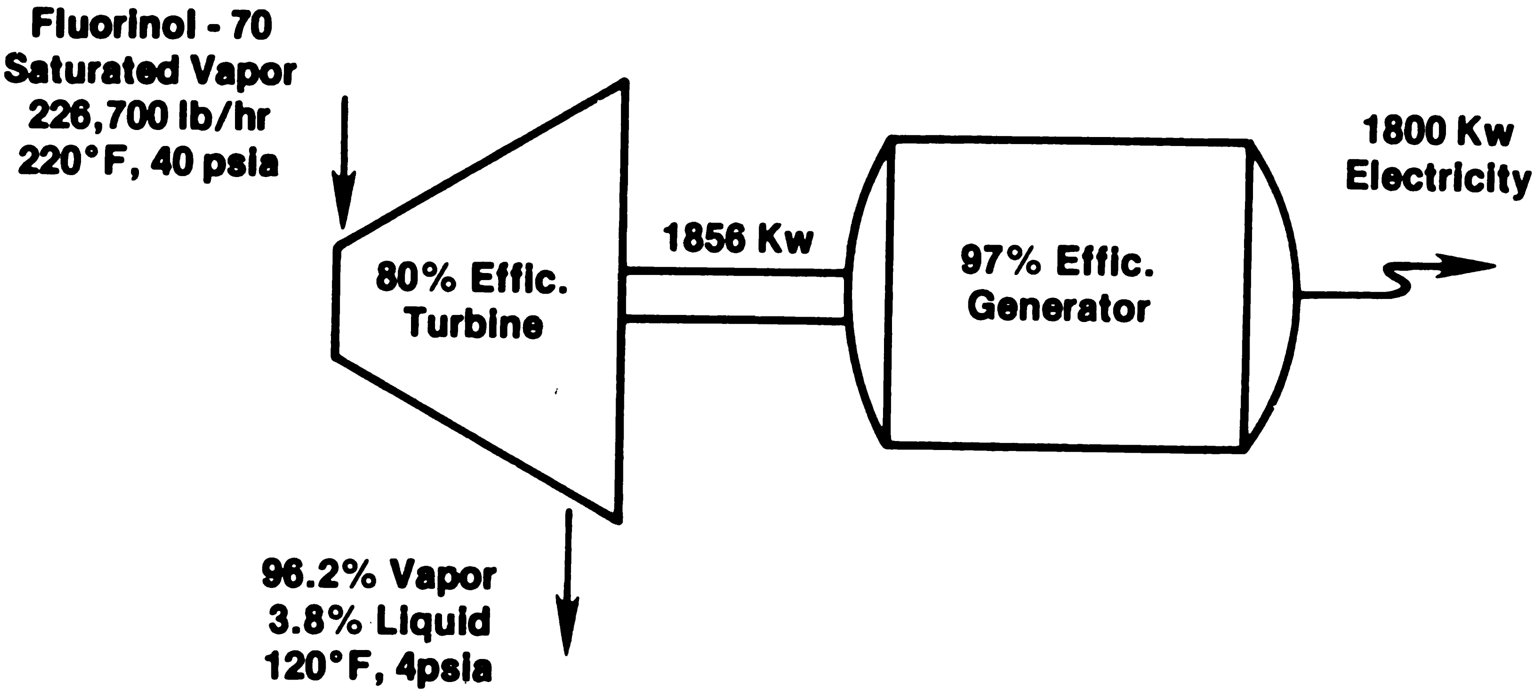


Figure 13 - Turbine and Generator