Solution
1996

Production of Non-Alcoholic Beer for Gulf Brewing Company

Process Design Report

B.A. Designer
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1996 Hops & Brew Way
Gulfcoast, Texas 70011

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Report

PRO-D-ZINE, INC.
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Gulfcoast, Texas 70011
March 9, 1996

TO: Big Boss

FROM: B.A. Designer

CC: Anita Brewsky, C.O. Owner

SUBJECT: Non-alcoholic beer for Gulf Brewing Company

Preliminary design work for the production of non-alcoholic beer at the Gulf Brewery has been completed. Your regular beer, Prairie Premium, is the feed to the process so that normal brewing operations are not disturbed.

In a very gentle manner, ethyl alcohol will be removed from the regular beer. Great care has been taken to preserve authentic beer taste in the non-alcoholic brew. First, several key volatile taste components are removed. Ethanol is then distilled at a low temperature from the volatile-less beer and the product is recombined with the volatiles. Alternative means of separation such as reverse osmosis and dialysis were considered, but since they are largely untried and unfamiliar to Pro-D-Zine, low pressure distillation was selected. Keeping quality in mind, sources of oxygen are strictly excluded. Additionally, the temperature of the beer never rises above 125.8 °F and the beer is never exposed to heating medium temperatures above 150 °F. Since beer is normally pasteurized at 140 °F without damage, these temperatures are in no way problematic, and scorching is avoided.

Once a suitable process was selected and refined, the process was optimized with respect to plant capacity using the six-tenths rule for relating plant capacity to total capital investment. Operating costs were assumed to vary with capacity in a linear manner. The results of this analysis showed that the optimum capacity is 614,000 bbl/year, yielding a 20% rate of return and having a net present worth of $2.1 million at 12%. The total capital investment required was $6.2 million and yearly operating costs were $5.6 million. Surprisingly, though, when a rigorous design at this capacity was completed, the rate of return turned out to be 98.7% and the net present worth $21 million. The total capital investment called for is $4.7 million and yearly operating costs are $2.4 million. Clearly the six-tenths rule and the assumption of linear operating costs were not accurate, but fortunately they were inaccurate in a positive direction.

Included in the report which follows are both the "optimum" design and a design at a lower capacity of 239,000 bbl/year which yielded a 7.4% rate of return. Pro-D-Zine realizes that the return on neither capacity is within the 12-20% range set by the managers, but endorses the larger capacity nevertheless. Further study of medium-range capacities could be undertaken, but the larger capacity process promises to be very profitable. It is equivalent to between one-fourth and one-half the non-alcoholic beer capacity of industry leader Anheuser-Busch and would give Gulf Brewing Company a significant market share.

Attachments: Complete process design report and calculation appendix

Introduction

Four weeks ago, C.O. Owner of Pro-D-Zine assigned this designer the tasks of defining an economical project for Gulf Brewing Company to produce and sell non-alcoholic beer and determining the economic potential of Gulf Brewing's entry into the non-alcoholic beer market. Since gaining a controlling interest in Gulf Brewing, Big Boss has had a strong desire to produce non-alcoholic beer, feeling that such a project would me quite economical. The brewery managers, however, have tried to convince Big Boss that this project would be, on the contrary, uneconomical, and would require too great an initial capital investment. Furthermore, the managers are adamantly opposed to any disturbances to the normal brewing process.

The purpose of this work is to complete a rigorous and economic preliminary design for the production of non-alcoholic beer in order to settle the dispute between the brewery managers and Big Boss over the nature of the economics. The Chief Executive Officer of Gulf Brewing Company, Anita Brewsky, will support an economical project.

The scope of this work is limited to removing the alcohol from a regular beer feed stream and storing the product (in this manner, normal brewing operations are not affected). Any bottling or other finishing of the non-alcoholic beer is left up to the brewers.

The specific economic goal of the work is to maximize the net present worth of the project while keeping the internal rate of return between 12 and 20%. Plant capacity will play a large role in the economics; as capacity increases, so do the IRR (to a point) and the net present worth (indefinitely). The amount of netback on the non-alcoholic beer can also be varied to make the economics more favorable.

Although the economics are a large part of the driving force behind this project, they are second in consideration to product quality. Taste is key to the acceptance of a non-alcoholic beer and great care must be taken to avoid damaging vital taste components, either by heat or by oxygenation.

The overall goal of this work, then, is to provide a preliminary design for a non-alcoholic beer facility which produces a high-quality, pleasant-tasting product without disturbing the normal brewing process. The facility should be of optimum capacity and net present worth should be maximized subject to an internal rate of return constraint of 12 to 20%.

This report and attached appendix document several different process designs completed, including the optimum. Results of various analyses and conclusions and recommendations drawn from the results are presented. Details included are project premises, process flow diagrams and stream attributes, a process description, discussion of special safety and environmental concerns, utility and operating costs summaries, an equipment summary, and all relevant economic and optimization calculations and results. The calculation appendix follows the formal report.

Summary and Discussion

As explained in the introduction, this work is in response to the instruction of C.O. Owner of Pro-D-Zine on behalf of Big Boss regarding Gulf Brewing Company. Big Boss believes that a non-alcoholic beer can be produced and sold at profit, but the brewery managers disagree. The goal of this work is to provide an economically attractive preliminary design for a non-alcoholic beer facility. This facility should be of optimum capacity and its net present worth should be maximized subject to an internal rate of return range of 12 to 20%. Just as importantly, the process should not degrade any key beer taste components either by heat or oxygenation and should not disturb the normal brewing process.

When the design work was begun, a review of the literature concerning beer and alcohol removal from beer was conducted. It quickly became apparent that the economic attractiveness of this operation lay not only in the sale of the non-alcoholic brew, but also in the excise tax savings accrued. By law, "non-alcoholic" beverages are those with an ethanol content of less than 0.5% by volume and are not subject to liquor excise taxes.

The literature review also made clear that several types of processes for the removal of alcohol from beer abound. Of these processes, several were eliminated immediately as they involve disturbing the normal brewing process so that alcohol is never produced in the first place. Such a process would violate the brewery managers' request. Three principal processes remained:

1. Reverse osmosis,
2. Dialysis, and
3. Distillation.

The first two processes are fairly new and have the disadvantage that at low ethanol concentrations such as the 0.5 vol% legal limit, a large amount of filtration is necessary and the beer may be inadvertently stripped of key flavor compounds. One advantage is that these processes avoid the high temperatures often associated with distillation, which may scorch the beer.

The first two processes were determined to be too untried and too unfamiliar to Pro-D-Zine. Distillation, an area in which Pro-D-Zine has much expertise, was selected as the best alternative. In order to avoid the problems associated with high temperatures, an extremely low column overhead pressure of 0.5 psia was selected. This pressure is only slightly higher than the minimum practical overhead pressure of 0.2 psia cited by Perry and Green. It is estimated that with specially designed trays, the bottoms pressure of the distillation column required will not exceed 2.0 psia. Due to the distillation operation being carried out under vacuum, excessive temperatures are avoided. The overhead temperature is 52.8 °F and the bottoms temperature, which is the maximum beer temperature in the process, is 125.8 °F. Since normal beer pasteurization temperature is 140 °F, these temperatures are not considered problematic. Refrigeration, rather than cooling water, must be used to condense the column overhead, but the product quality outweighs the increased operating costs.

The other safeguard against scorching of the beer that is employed in the process is the use of recirculating hot water heat exchangers whenever heating of the beer is required. Heating is necessary both to preheat the feed beer and to reboil the bottoms stream of the distillation column. Normal 100 psig steam heating was considered and has the advantage of being more economical, but subjecting the beer to the 338 °F temperature of saturated steam at this pressure was judged a poor idea. Exposure to hot surfaces at this temperature would certainly scorch the beer, degrading key taste components. Therefore, in both cases where heating is required, a system of recirculating hot water which is reheated by steam in a separate exchanger is used. In this manner, the beer is never exposed to heating medium temperatures greater than 150 °F.

Several other taste considerations were taken into account in designing the process. First, the volatile taste components and carbon dioxide are removed from the preheated feed in a flash drum and the volatiles are condensed. This liquid is then recombined with the de-alcoholized product from the distillation column, thus retaining the key flavor compounds. After this stream is cooled, a small amount of regular beer is blended back into it to impart additional flavor. A sufficient amount of regular beer is added so that the product approaches, but does not exceed, the 0.5 vol% legal limit. The final step is to mix in CO_2 to make up for the amount lost in the flash and the distillation column.

Once the process was conceived, a base case flowsheet was drawn, material and energy balances were completed, equipment was sized and costed, and the economic analysis was performed. As expected, the first case was far from optimum, having a net present worth of -$9.5 million at 12%. When this case was examined further, it was determined that the costs of distillation, both capital and operating, were excessive. A new case, termed "Optimization I," with a more efficient column was completed and found to be more economic. Below is a comparison of the two cases.

Table 5.1 Economic Comparison of Initial Cases

	Total Capital Investment	Yearly Operating Costs	Net Present Worth at 12%	Internal Rate of Return
Revised Base Case	$8.03 million	$3.50 million	- $9.51 million	N/A
Optimization I	$3.53 million	$2.23 million	- $676,000	0.07

Clearly, the economics of Optimization I were better but still unacceptable. One of the goals of this work is to maximize the net present worth of the project while keeping the internal rate of return between 12 and 20%. Therefore, a new series of optimizations was undertaken. It may have been possible to optimize with respect to column overhead pressure since the pressure used was 0.3 psi above the minimum. (As pressure decreases, relative volatility increases, thus reducing both the number of stages and the vapor rate required for a given separation. Also, at a lower pressure, the beer would be subjected to even lower temperatures.) However, because the pressure range for optimization would have been small, this optimization was judged to be trivial and another optimization variable was chosen. That variable was plant capacity.

As capacity increases, so do the rate of return (initially) and the net present worth (indefinitely). These trends can be seen in Figure 5.1. However, capacity requirements quickly become ridiculous, exceeding the total non-alcoholic beer production of industry leader Anheuser-Busch. The upper limit on internal rate of return helps to keep the project realistic. Also, had it been necessary, the netback on the non-alcoholic brew could have been adjusted to make the economics more favorable at a lower capacity.

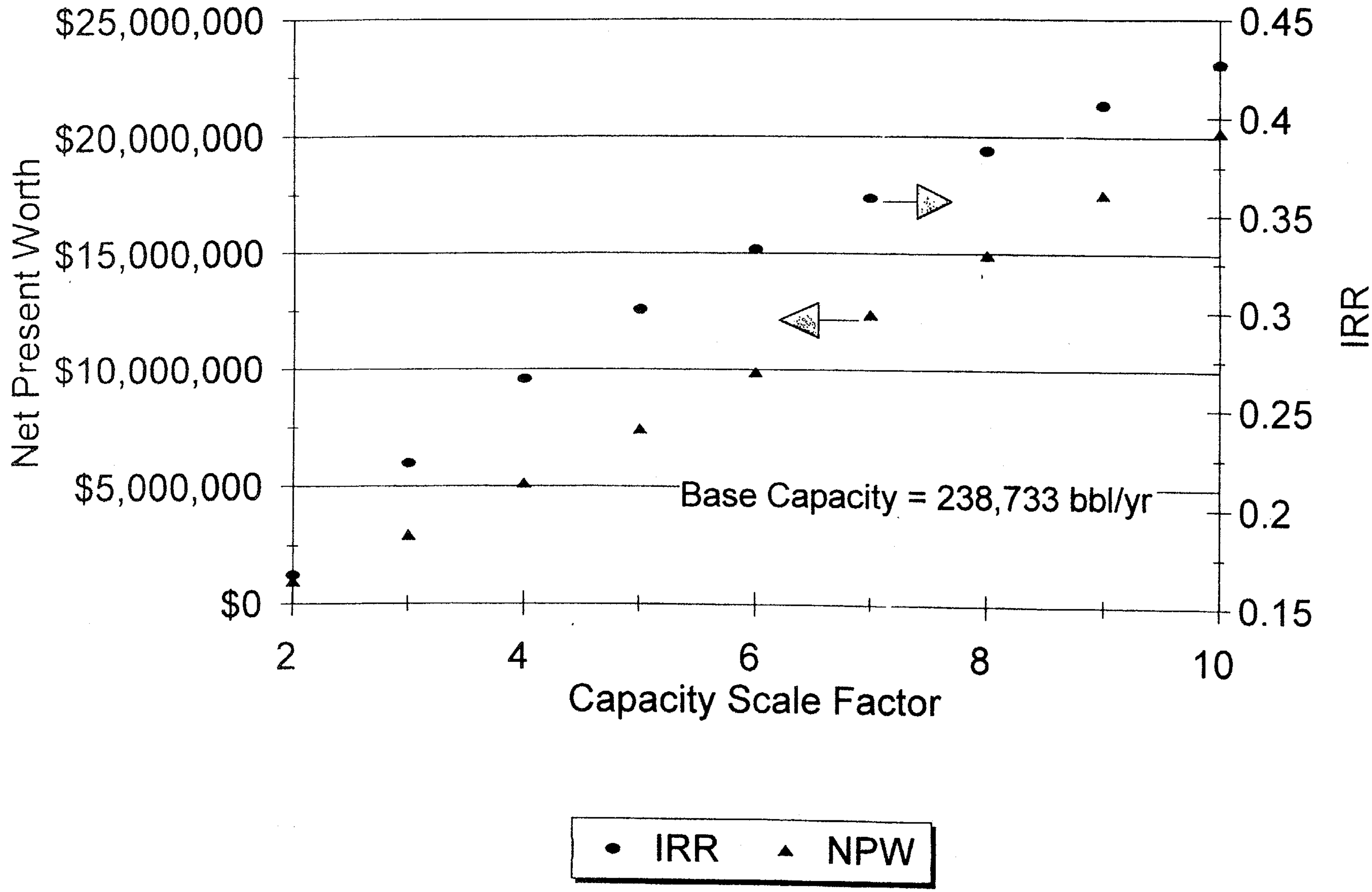
Due to the four week time constraint on the project, the capacity optimization was performed using the six-tenths rule to relate total capital investment to capacity rather than by rigorous design calculations. Operating costs were assumed to vary in a linear manner with capacity. The result of this optimization process was that a capacity of 614,315 bbl/year, or 2.57 times the 238,733 bbl/year base capacity, was optimum. This case had a net present worth of \$2.1 million and an internal rate of return of 20%. The details of these calculations can be found in Table 17.2 of the Economic Methods and Analysis section of this report, or approximate values can be read from Figure 5.1.

After the optimum capacity was determined, rigorous design calculations were completed at this new value. Curiously, when the economic analysis was performed, the net present worth was \$21.1 million and the internal rate of return 98.7%, both much higher than expected. The manufacturing cost was \$5.75/bbl. The six-tenths rule, which, according to Peters and Timmerhaus, "should only be used in the absence of other information," proved to be inaccurate. Also, it was later discovered that the assumption of linear variation of operating costs was inaccurate. The cost of operating labor, which constitutes a large portion of the yearly operating costs for this project (about 34%), does not vary with capacity as long as the number of operators remains the same. This fact was neglected. Fortunately, both inaccuracies were in a positive direction. Due to time constraints, no additional optimizations were undertaken.

One additional calculation was performed to determine whether the 614,315 bbl/year figure is reasonable. Using data from the 1994 Anheuser-Busch Annual Report, it was determined that this capacity is anywhere from one-fourth to one-half that of Anheuser-Busch's yearly non-alcoholic beer capacity (for details of this calculation, see p. 236 of the calculation appendix). Thus, the figure is quite reasonable.

Gulf Brewing now has two options. If either the managers or executives are concerned about the uncertainty surrounding the optimum capacity, further study of the range between 238,733 and 614,315 bbl/year may be warranted. The optimum lies between these values. However, a plant with the larger capacity has been shown to be quite reasonable and promises to be extremely lucrative even though its internal rate of return is outside the permitted range. Therefore, it is the recommendation of Pro-D-Zine that Gulf Brewing proceed immediately with building such a plant with a target startup date of January 1, 1999. The complete details of the preliminary process design are presented in this report.

Fig. 5.1 IRR and NPW vs. Capacity



Conclusions

The principal conclusion to be drawn from the work that has been done is that it is indeed possible to produce and sell a non-alcoholic brew at profit. The salient economic details of the best design are:

Table 6.1 Economic Conclusions

Optimum Capacity	614,315 bbl/year
Net Present Worth	$21.1 million
Internal Rate of Return	98.7 %
Total Capital Investment	$4.7 million
Yearly Operating Cost	$2.4 million
Manufacturing Cost	$5.75/bbl

Full details of the economic analysis can be found in Table 17.3 of the Economic Methods and Analysis section of this report.

Again, it must be emphasized that the six-tenths rule analysis predicted that this capacity would have a net present worth of $2.1 million and an internal rate of return of 20%. The rigorous design economics turned out to be much more attractive due to inaccuracies discussed in the Innovation and Optimization section of this report. Due to the four week time constraint, no further capacity optimizations were performed. The optimum capacity lies somewhere between the base capacity of 238,733 bbl/year and 614,315 bbl/year. This larger capacity, however, has been shown to be quite reasonable in comparison to the estimated Anheuser-Busch non-alcoholic beer production of 1.3 million to 2.2 million bbl/year.

The instructions of C.O. Owner, in addition to designing an economically attractive process, were to "be conservative in producing a quality product" and to avoid degrading any taste components, either by heat or by oxygenation. These objectives have also been met. The temperature of the beer never rises above 125.8 °F and the beer is never exposed to heating medium temperatures greater than 150 °F. Sources of oxygen are strictly excluded. The key volatile taste components are initially removed and then re-added to the beer. In addition, sufficient regular beer is blended with the product to enhance the taste while keeping the alcohol content below the legal limit of 0.5 vol%. By this combination of methods, the authentic taste of the beer, which is so key to the acceptance of a non-alcoholic brew, is preserved.

Recommendations

The primary recommendation of Pro-D-Zine is that Gulf Brewing Company proceed immediately with building a 614,315 bbl/year non-alcoholic beer facility with a target startup date of January 1, 1999 based on the optimum design presented in this report. Pro-D-Zine is aware that the process economics are so favorable that they are outside the internal rate of return range set by the brewery managers. However, it is precisely because the economics are so favorable that this design is recommended. The capacity has been shown to be quite reasonable and the product quality high.

If the brewery managers or executives remain uncomfortable with this capacity and its high rate of return, Pro-D-Zine recommends that further study of medium range capacities be undertaken. Since a quality design has already been developed, such a study would be time-consuming but not difficult. However, it must be emphasized again that Pro-D-Zine believes the optimum design presented in this report to be the best, most logical, and most economical choice for Gulf Brewing.

Project Premises

1. Overall Schedule and Battery Limits

- The plant will startup at full capacity on January 1, 1999.
- The project life will be at least ten years
- The battery limits are the feed line from cold regular beer storage and the product storage tanks.
- It is assumed that any additional beer finishing processes will be carried out by the brewers.
- Bottling the beer is outside the scope of this project.

2. Feed and Product Specifications

A. Ethanol. Specifications on any ethanol byproduct that could be sold are:

Table 8.1 Ethanol specifications.

	EtOH/H_2O azeotrope	Anhydrous EtOH	Anhydrous Motor Fuel EtOH	Hydrous Motor Fuel EtOH
Volume % EtOH	96	100	99.5	95

B. Final beer product. The final beer product cannot contain more than the following amounts of the following components:

Table 8.2 Product specifications.

1 ppm O_2
0.02 ppm diacetyls
0.1 ppm iron
0.2 ppm copper
2 ppm aluminum
0.1 ppb thiols
6 ppb *o*-dichlorophenols

Based on material balance calculations, the final non-alcoholic beer product should have the following composition:

Table 8.3 Non-alcoholic beer composition*.

98.5 wt% water
0.3 wt% ethanol**
0.55 wt% carbon dioxide
0.23 wt% pyruvic acid
0.19 wt% ethyl acetate
0.23 wt% non-characterizable volatiles

* This composition is based on an assumption about the composition of the feed beer that will be explained in the next section.

** By law, any beer labeled "non-alcoholic" must contain less than 0.5 vol% ethanol.

C. Feed beer. Based on information about the composition of beer taken from several literature sources, the composition of the feed to the process is assumed to be adequately represented by:

Table 8.4 Feed beer composition.

94.65 wt% water
4.2 wt% ethanol
0.53 wt% carbon dioxide
0.21 wt% pyruvic acid
0.21 wt% ethyl acetate
0.2 wt% non-characterizable volatiles

It is impossible to deal with every component of regular beer, as there are literally hundreds of components. However, the above mixture has been judged to be representative. The water, ethanol, and carbon dioxide are straightforward. Pyruvic acid and ethyl acetate were chosen to represent all organic acids and esters, respectively, because they are the most prevalent, according to Engan. The 0.2 wt% "non-characterizable volatiles" are a small yet important part of the mixture. They represent the "other" compounds frequently referred to in the beer literature. These compounds are the major taste-imparting compounds which are often missing in non-alcoholic beer. For purposes of this problem, the volatiles were assumed to be more volatile than water and ethanol but just as condensable. That is, they go with the vapor out of the flash drum but are condensed immediately thereafter and sent to the recombiner (see the Process Flow Diagram). In this manner, the vital taste components of the beer are retained.

For a more thorough explanation of how this composition was determined, see pp. 21-23 and 109-110 of the calculation appendix.

3. Costs of Raw Material and Utilities. The costs of steam, cooling water, and refrigeration were obtained from Peters and Timmerhaus. The cost of CO_2 was found in the *Kirk-Othmer Encyclopedia of Chemical Technology*. The cost of electricity was estimated from data given in a previous course. All costs were inflated to 1999 dollars and were as follows:

Table 8.5 Raw Material and Utilities Costs

Steam	\$2.185/1000 lb
Cooling Water	\$0.117/1000 gal
Refrigeration	\$0.0081/10^3 Btu removed
Electricity	\$0.0494/kWhr
CO_2	\$0.0416/lb

4. Product Selling Prices

A. Ethanol. If the alcohol byproduct were sold (the design was completed assuming no sale), the selling prices would be:

190 proof	\$2.73/gal
Fuel grade	\$1.14/gal

B. Non-alcoholic beer. As per the suggestion of C.O. Owner, netback on the NA beer product was assumed to be the same as that of Prairie Premium, \$5.00/bbl.

5. Economic Premises

- Discounted cash flow analyses were performed using the methods espoused by Peters and Timmerhaus and White, Agee, and Case.

- January 1, 1999 costs were used throughout. Costs were not inflated year-by-year.

- The plant will startup at full capacity on January 1, 1999.

•When necessary to inflate cost data, ratios of the Chemical Engineering Index were used to inflate first to 1994 dollars. 1994 costs were then inflated to 1999 dollars using an inflation rate of 2.5% per year.

•Incremental income tax rates used were 34% for federal taxes and 6% for state taxes

•Excise taxes were considered to be $0.388/gal for capacities less than 60,000 bbl/year, $0.448/gal otherwise

•Depreciation was assumed to apply to all of the Direct Plant Costs but not to Indirect Costs, Contractor's Fees, Contingency, or Working Capital. Seven-year Modified Accelerated Cost Recovery System (MACRS) factors were used. The factors are as follows:

Table 8.6 MACRS Factors

Applicable Year	Factor
1	0.1429
2	0.2449
3	0.1749
4	0.1249
5	0.0893
6	0.0892
7	0.0893
8	0.0446
Total	1.00

•The Gulf Brewery is in a sold-out position and additional capacity must be added by 1999.

•Additional economic assumptions are detailed in the Economic Methods and Analysis section of this report.

6. **Environmental Requirements.** There are no unusual environmental requirements associated with this project. The Safety and Environmental Considerations section of this report discusses two minor concerns.

7. **Processing Limitations.** There are no specific processing limitations for this project.

8. Existing Brewery Operation

- The normal brewing process for the regular beer will not be disturbed.

- 100 psig saturated steam is assumed to be available on site.

- 90 °F cooling water with a maximum return temperature of 120 °F is assumed to be available on site.

- Since the existing brewery must keep its products cooled to a temperature of 34 °F, ammonia refrigeration at 24 °F (minimum 10 °F approach temperature) is assumed to be available on site.

9. Extraordinary Costs. The following extraordinary advertising costs are assumed to occur in the first four years of the project:

Table 8.7 Advertising costs.

Year	Cost
1	$350,000
2	$250,000
3	$150,000
4	$90,000

10. Labor Cost. The process is assumed to require one additional operator position with four operators per shift position at a 1999 inflated labor cost of $21.46 per hour (labor cost from Peters and Timmerhaus).

11. Product Quality Considerations. Taste is key to acceptance of a non-alcoholic beer. *The process must not degrade the beer in any way, either by heat or by oxygenation.* The following measures to preserve taste were put into place:

- The beer is never heated directly by steam, either in the feed preheater or in the column reboiler. The high temperature would certainly scorch the beer. Instead, a recirculating hot water system is used (see the Process Flow Diagram). The hot water approaches the beer temperature within 10 °F and then is reheated by steam. This process creates more entropy than direct steam heating, but is worthwhile because of the preserved product quality.

•Distillation of the ethanol in the beer occurs at a low pressure, 0.5 psia. At this pressure, beer temperatures never rise above 125.8 °F, which is less than the 140 °F pasteurization temperature suggested by Hardwick.

•Before the beer enters the alcohol removal tower, it is subjected to a flash at slightly less than atmospheric pressure (13.2 psia). This step removes some of the key volatile flavor components, which are then recombined with the de-alcoholized beer from the distillation column.

•A small amount of the feed beer stream is diverted and mixed in-line with the final product. The amount is kept small enough that the final product does not exceed the 0.5 vol% legal ethanol limit for "non-alcoholic" beer. The amount is sufficient, however, to impart some authentic beer taste to the de-alcoholized brew.

•Due to the food-grade nature of this application, all equipment with which the beer comes in contact will be constructed of stainless steel.

12. Calculations

•The ChemCad III process simulator was used for all physical property calculations.

•The process simulator was also used for rigorous multi-stage distillation calculations (the Simultaneous Corrective Distillation Simulation (SCDS) model was used), enthalpy calculations, and flash calculations. Distillation calculations were confirmed by Fenske-Underwood-Gilliland or McCabe-Thiele methods.

•The binary ethanol/water data at low pressures given by Hirata, Nagahama, and Ohe were assumed to be applicable to this project. The data were entered into the simulator and regressed to find parameters for the TK-Modified-Wilson activity coefficient equation.

•The enthalpy model used was the latent-heat model described on the next page, which is reproduced from the ChemCad user manual. Use of this model is the reason for the negative enthalpies shown on the Stream Attributes sheets.

11.6.6 LATENT-HEAT MODEL → Used in all enthalpy calculations

This model is mostly suited to chemical systems. The enthalpies are calculated by the following equations:

$$H_L = \Delta H_{f(L)}^{298.15} + \int_{298.15}^{T} CpLd\,T$$

$$HV = H_{L(bubble)} + H_{V(at\ bubble)} + \int_{T_{bubble}}^{T} Cpv\,dT$$

where

H_L	=	the liquid enthalpy
$\Delta H_{f(L)}^{298.15}$	=	the heat of formation of liquid at 298.15 K
Cpl	=	the liquid heat capacity
HV	=	the vapor enthalpy
$H_{L\,(bubble)}$	=	the liquid enthalpy at bubble point and system pressure
$H_{V\,(at\ bubble)}$	=	the heat of vaporization at bubble point
Cpv	=	the heat capacity of vapor

The coefficients *a,...,e* for most chemical components are available in the database. If the coefficients are not available, this enthalpy option should not be used.

The heat of vaporization at temperature T is calculated by the DIPPR equation if the coefficients are available:

$$Hv = A\,(1 - T_r) ** \left(B + CT_r + DT_r^2\right)$$

where

Hv is in J/Kmol
Tr is the reduced temperature

If the DIPPR coefficients are not available, the heat of vaporization at temperature T is calculated by the Watson equation:

$$Hv(T_2) = Hv(T_1) * \left[\frac{1 - Tr_2}{1 - Tr_1}\right]^{0.38}$$

where

Tr1 and Tr2 are the reduced temperature at T1 and T2 respectively.

The heat of vaporization at component normal boiling point is stored in the database. If it is not available it is estimated by the following equation:

$$Hv(Tb) = 1.093 * R * Tc/1.8 * (Tbr * (ln\,(Pc/14.696) - 1)/(0.93 - Tbr))$$

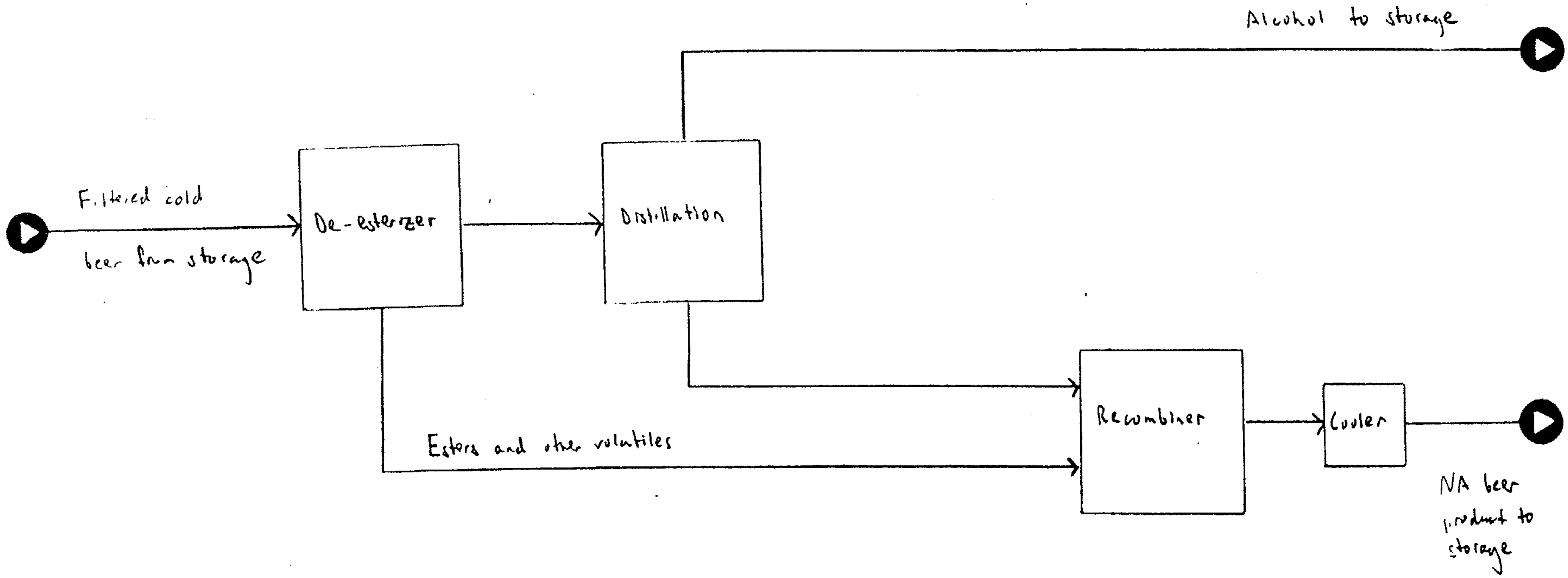
where

Tb is the component normal boiling point.
Tc is the critical temperature
Pc is the critical pressure
Tbr = Tb/Tc

The Soave-Redlich-Kwong entropy is used for this option.

Note: The standard reference state for all enthalpy calculations is the liquid heat of formation at 298.15°K.

Block Flow Diagram

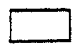


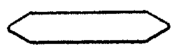
Explanation of Flowsheet Symbols

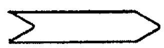
All flowsheets included in this report follow PRO-D-ZINE standards, and the details of process equipment should be obvious. However, the flow symbols may vary from other standards and so an explanation is included here.

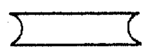

Stream Number


Pressure, psia

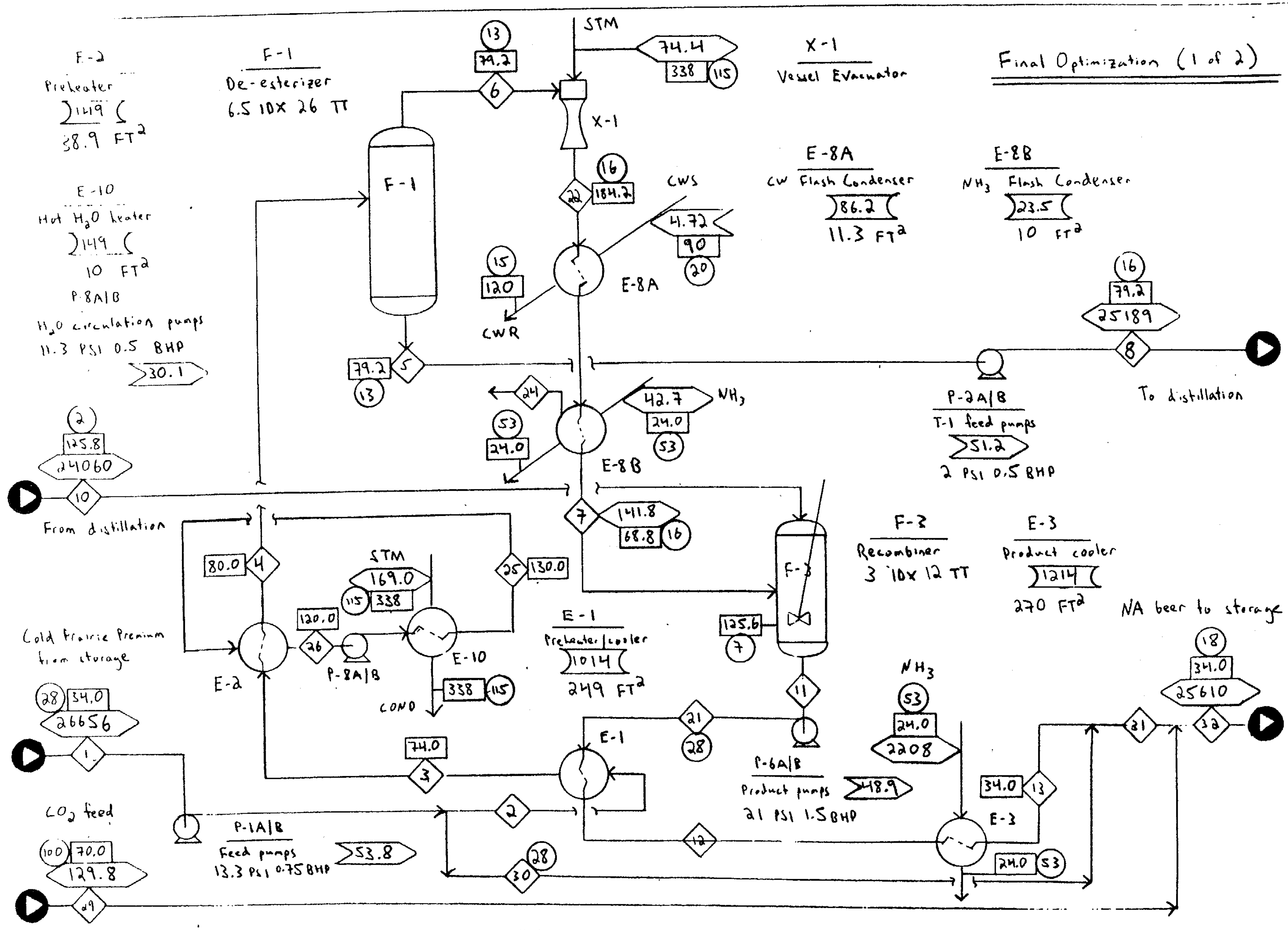

Temperature, °F

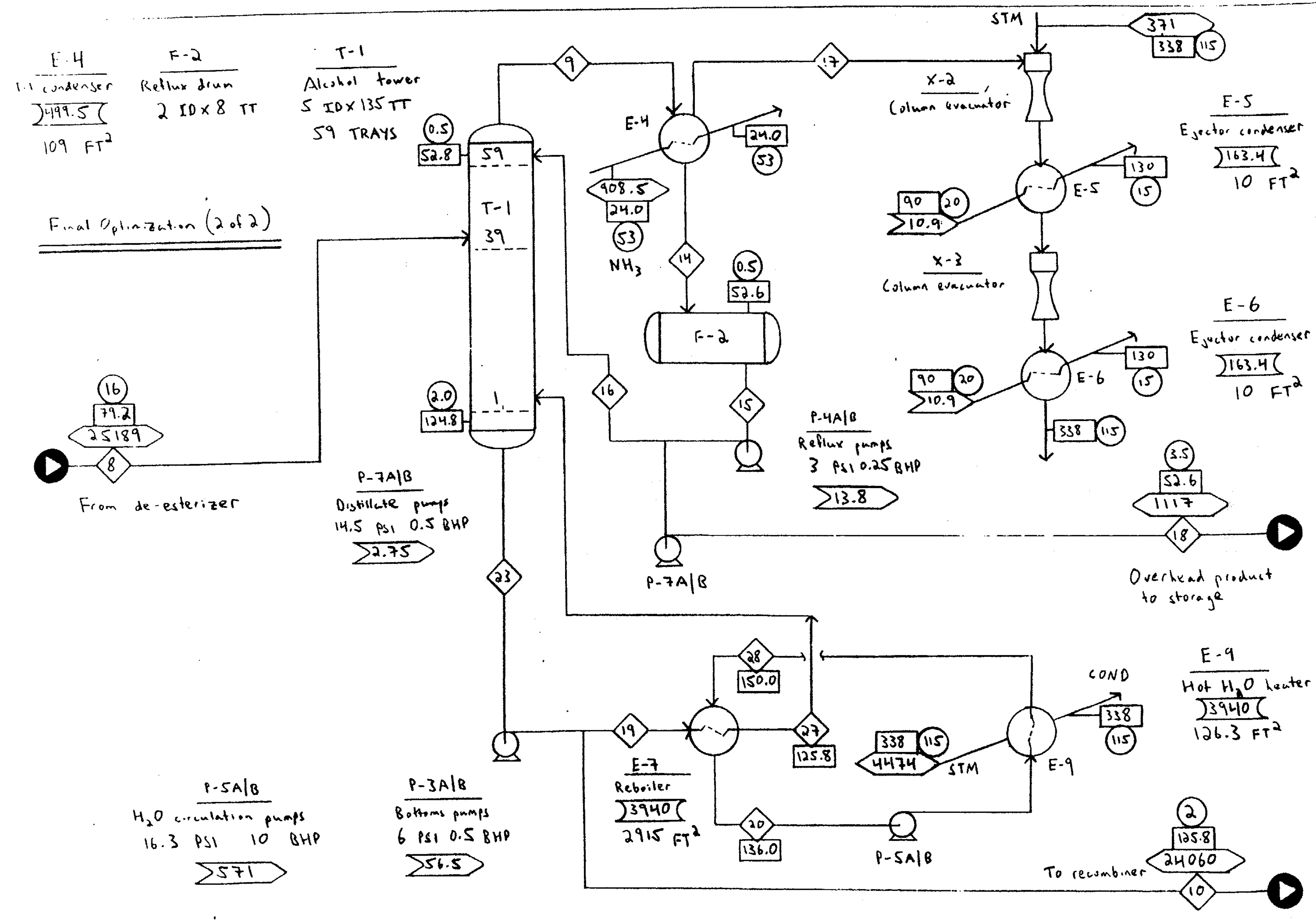

Mass Flow, lb/hr


Volumetric Flow, gpm


Duty, 10^3 Btu/hr

Final Optimization (1 of 2)





E-4
condenser
499.5
109 FT²
F-2
Reflux drum
2 ID x 8 TT
T-1
Alcohol tower
5 ID x 135 TT
59 TRAYS
Final Optimization (2 of 2)
STM
371
338
115
9
17
X-2
Column evacuator
E-5
Ejector condenser
163.4
10 FT²
0.5
52.8
59
T-1
39
1
2.0
124.8
E-4
24.0
53
908.5
24.0
53
NH₃
130
15
90
20
10.9
14
0.5
52.6
F-2
X-3
Column evacuator
E-6
Ejector condenser
163.4
10 FT²
16
79.2
25189
8
From de-esterizer
16
15
P-4A|B
Reflux pumps
3 PSI 0.25 BHP
13.8
130
15
90
20
10.9
338
115
3.5
52.6
1117
18
Overhead product to storage
P-7A|B
Distillate pumps
14.5 PSI 0.5 BHP
2.75
23
P-7A|B
28
150.0
19
27
125.8
E-7
Reboiler
3940
2915 FT²
20
136.0
P-5A|B
COND
338
115
338
115
4474
STM
E-9
E-9
Hot H₂O heater
3940
126.3 FT²
2
125.8
24060
10
To recombiner
P-5A|B
H₂O circulation pumps
16.3 PSI 10 BHP
571
P-3A|B
Bottoms pumps
6 PSI 0.5 BHP
56.5

Stream Attributes

Final Optimization (1 of 3)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Title	Feed to Process	Feed to Separation	Partially preheated feed	Preheated Feed	Flash Liquid	Flash Vapor	E-8 liquid effluent	T-1 Feed	T-1 Overhead	T-1 Bottoms to recombiner	Recombiner Outlet	Partially cooled recombined
Water	25230	24019	24019	24019	24017	2.14	75.1	24017	326.5	23952	24027	24027
Ethanol	1120	1066.3	1066.3	1066.3	1055.4	10.78	9.34	1055.4	5207.8	13.9	23.2	23.2
Carbon Dioxide	141.0	134.3	134.3	134.3	12.25	122.0	4.53	12.25	12.25	—	4.53	4.53
Pyruvic Acid	55.8	53.3	53.3	53.3	53.3	—	—	53.3	—	53.3	53.3	53.3
Ethyl Acetate	55.8	53.3	53.3	53.3	50.9	2.40	2.08	50.9	50.8	40.9	42.9	42.9
Air	—	—	—	—	—	—	0.0003	—	—	—	0.0003	0.0003
Volatiles	53.3	50.7	50.7	50.7	—	50.7	50.7	—	—	—	50.7	50.7
Total	26656	25377	25377	25377	25189	187.8	141.8	25189	5597.4	24060	24201	24201
Volumetric	53.8	51.2	51.5	51.5	51.2	31.4	0.31	50.9	24374	48.6	49.0	48.6
Temperature	34.0	24.0	74.0	80.0	79.2	79.2	68.8	79.2	52.8	125.8	125.6	84.3
Pressure	28.0	28.0	23.0	18.0	13.2	13.2	15.7	15.2	0.50	2.0	7.0	23.0
Phase	L	L	L	L	L	V	L	L	V	L	L	L
Molecular Weight	18.61	18.61	18.61	18.61	18.55	43.54	20.24	18.55	42.41	18.08	18.09	18.09
Density	61.785	61.785	61.455	61.362	61.507	0.106	59.641	61.507	0.004	61.607	61.589	62.146
Enthalpy	-176081	-167634	-163445	-167636	-167127	-696.5	-871.5	-166782	-13609	-162459	-163367	-164374

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Stream Attributes

Final Optimization (2 of 3)

Stream Number	13	14	15	16	17	18	19	20	21	22	23	24
Stream Title	Preliminary Product	T-1 Condensate	Reflux Drum Effluent	Pumped Reflux	T-1 Air + CO_2	Overhead Product	T-1 bottoms to reboiler	Reboiler H_2O	Pumped recombiner outlet	X-1 Discharge	T-1 total bottoms	E-8 vapor effluent
Water	24027	326.5	326.5	261.2	—	65.3	3820.6	281409	24027	76.4	27772	1.34
Ethanol	23.2	5208	5208	4166	—	1041.6	26.73	—	23.2	10.78	40.6	1.44
Carbon Dioxide	4.53	—	—	—	12.25	—	—	—	4.53	122.0	—	117.4
Pyruvic Acid	53.3	—	—	—	—	—	0.51	—	53.3	—	53.8	—
Ethyl Acetate	42.9	50.8	50.8	40.7	—	10.16	18.27	—	42.9	2.21	59.2	0.152
Air	0.0003	—	—	—	26.7	—	—	—	0.0003	17.3	—	17.3
Volatiles	50.7	—	—	—	—	—	—	—	50.7	50.7	—	—
Total	24201	5585	5585	4468	39.0	1117	3866	281409	24201	279.4	27926	137.6
Volumetric	48.4	13.8	13.8	11.0	220	2.75	7.82	570.7	48.9	63.4	56.5	20.2
Temperature	34.0	52.6	52.6	52.6	52.8	52.6	124.8	136.0	125.6	1841.2	124.8	68.8
Pressure	18.0	0.5	0.5	3.5	0.5	3.5	8.0	14.7	28.0	15.7	2.0	15.7
Phase	L	L	L	L	V	L	L	L	L	V	L	V
Molecular Weight	18.09	42.39	42.39	42.39	32.44	42.39	18.09	18.016	18.085	32.09	18.09	40.83
Density	62.422	50.487	50.487	50.487	0.003	50.487	61.60	61.434	61.589	0.073	61.60	0.114
Enthalpy	-165585	-15875	-15875	-12700	-45.82	-3175	-26093	-1,902,404	-163367	-1331	-188472	-462.2

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes

Final Optimization (3 of 3)

Stream Number	25	26	27	28	29	30	31	32				
Stream Title	Preheater H_2O	Preheater H_2O	Vapor to T-1	Reboiler H_2O	CO_2 Feed	Bypass reg. beer	NAB w/o CO_2	Final NAB Product				
Water	14908	14908	3820.6	281409	–	1210.3	25237	25237				
Ethanol	–	–	26.7	–	–	53.8	76.9	76.9				
Carbon Dioxide	–	–	–	–	129.8	6.76	11.30	141.0				
Pyruvic Acid	–	–	0.51	–	–	2.68	55.8	55.8				
Ethyl Acetate	–	–	18.3	–	–	2.68	45.5	45.5				
Air	–	–	–	–	–	–	0.0003	0.0003				
Volatiles	–	–	–	–	–	2.55	53.3	53.3				
Total	14908	14908	3866	281409	129.8	1278.8	25480	25610				
Volumetric	30.1	30.1	11126	573.3	2.70	2.57	50.9	51.2				
Temperature	130.0	120.0	125.8	150.0	70.0	34.0	34.0	34.0				
Pressure	26.0	16.0	2.0	26.0	100	28.0	18.0	18.0				
Phase	L	L	V	L	V	L	L	L				
Molecular Weight	18.016	18.016	18.09	18.016	44.01	18.61	18.11	18.16				
Density	61.697	61.536	0.006	61.177	0.804	61.785	62.389	62.361				
Enthalpy	-100872	-101021	-22058	-1,898,475	-499.9	-8447.4	-174092	-174606				

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Process Description

Cold, regular beer from storage, Stream 1, is pumped by the P-1A/B feed pump. The pump discharge is split into two streams, Stream 2 and Stream 30. The latter bypasses the process and is mixed with de-alcoholized beer, Stream 31, to form the final product, Stream 32.

The regular beer in Stream 2 exchanges heat with the Recombiner outlet (tube side) in E-1, cooling the product and preheating the feed to the separation. The outlet of E-1 is Stream 3, which is further preheated in E-4 by recirculating hot water (shell side), and leaves as Stream 4.

Because product quality, particularly scorching, is such an important concern in this process, recirculating hot water at 130 °F is used for heating rather than the available 100 psig steam at 338 °F, which would certainly damage the product. The water, Stream 26, is pumped by P-8A/B to the tube side of E-10 where it is heated by steam. The hot water leaves as Stream 25 and enters the shell side of E-2. The higher entropy and energy costs of this process are balanced by improved product quality.

The fully preheated feed, Stream 4, is flashed into the De-esterizer vessel, F-1, which is maintained at a pressure of 13.2 psia by the X-1 ejector. Following the suggestion of Pabst (p. R-24), F-1 is ten times as large as required in order to ensure that foaming and liquid carryover into the vapor stream do not occur and defeat the purpose of the process.

Stream 6, the flash vapor, contains small amounts of water, ethanol, and ethyl acetate, but consists mainly of carbon dioxide and non-characterizable volatiles. This stream is entrained by ejector motive steam and passes through the vessel evacuator, X-1, from which it leaves as Stream 22 at a pressure of 15.7 psia.

Stream 22 is cooled initially by cooling water (shell side) in E-8A and then further cooled by ammonia refrigeration (shell side) in E-8B. This cooling condenses all of the water vapor (including the motive steam), all of the volatiles, most of the ethanol and ethyl acetate, and some of the carbon dioxide. The non-condensed material, primarily CO_2 and leaked air, is allowed to vent to the atmosphere as Stream 24. The condensate, Stream 7, flows to the agitated Recombiner vessel, F-3.

The liquid from the flash, Stream 5, is pumped by the T-1 feed pumps, P-2A/B. Stream 8, the pump discharge, enters the 59-tray, 5 ft diameter, 135 ft tall alcohol removal tower, T-1, on tray 39 (numbering up the column). T-1 is maintained at 0.5 psia by steam-jet ejectors X-2 and X-3. Of the ethanol in the feed, 98.7 % is recovered in the distillate. In order to make up for the addition of water to Stream 7 by the motive steam for X-1, some of the water in Stream 8 is allowed to go up the column (this approach also makes the separation easier). The column overhead, Stream 9, flows to the shell side of the condenser, E-4, where all components except CO_2 and leaked air are condensed by ammonia refrigeration. The CO_2 and air, Stream 17, are entrained by motive steam and pass through the column evacuators, the intercondenser, and the aftercondenser, X-2, X-3, E-5, and E-6, respectively. The ejector exhaust is returned to wastewater cleanup.

The condensed overhead, Stream 14, passes through the reflux drum, F-2, exiting as Stream 15. This stream is pumped by P-4A/B and splits into Streams 16 and 18. Stream 16 is reflux to the column. Stream 18, the distillate product, is pumped by P-7A/B to storage tank V-2. The reason for two different sets of pumps is to ensure that by the time Stream 16 re-enters the column, line losses have dropped its pressure back down to 0.5 psia. Stream 18, which must travel all the way to storage, requires a higher pressure.

The column bottoms, Stream 23, is pumped by P-3A/B and splits into Stream 19 and Stream 10. Stream 19 is reboiled on the tube side of E-7 by another recirculating hot water circuit and re-enters the column as Stream 27. The hot water, Streams 28 and 20, is pumped by P-5A/B and is reheated by steam (shell side) in E-9.

Stream 10, the bottoms product, flows to F-3, the agitated recombiner, where it is mixed with Stream 7, the condensed flash vapor and steam. The condensed steam in Stream 7 makes up for the water lost up the column.

Stream 11, the recombiner outlet, is pumped by P-6A/B, exiting as Stream 21. This stream is partially cooled by incoming feed (shell side) in E-1, leaving as Stream 12. This stream is further cooled to the product storage temperature of 34 °F by ammonia refrigeration (shell side) in E-3.

Stream 13, the effluent from E-3, is mixed for taste purposes with a small amount of regular beer bypass, Stream 30, to form Stream 31. Makeup CO_2, Stream 29, is then added to this stream to form Stream 32, the final product which goes to the storage tank V-1.

For additional details about specific pieces of equipment, see Table 15.1, the Equipment Information Summary.

Safety and Environmental Considerations

Two process streams merit special environmental consideration: Stream 18, the distillate product, and Stream 24, the vent gas from the flash vapor condenser.

Stream 18 is a liquid with the following composition:

65.3 lb/h	(5.8 wt%)	water
1041.6 lb/h	(93.3 wt%)	ethanol
10.16 lb/h	(0.9 wt%)	ethyl acetate

The design and economic calculations have been completed assuming that it is impossible to sell this byproduct. As noted in the Economic Methods and Analysis section of this report, the specifications on ethanol, even fuel grade, are too stringent and this overhead stream cannot be sold. A storage tank sufficient for one week's storage of this material at 90% capacity has been sized and costed (see the Equipment Information Summary), but clearly some means of disposal is necessary.

First, simply disposing of the byproduct into a nearby river or sewage system was explored. However, this option was abandoned because the average Biological Oxygen Demand (BOD) of this stream was found to be too large. According to one reference (*Kirk-Othmer* 1992), sources with an average BOD level of 600 mg/L or higher are generally assessed extra charges by municipalities. The BOD level of this stream was found to be in excess of 1.5 million mg/L (for details of this calculation, see the Environmental Considerations section of the calculation appendix). Obviously, diverting this stream to the local sewage system is not a valid option.

Due to time constraints, this particular environmental problem was not explored any further. It is recommended that other options such as burning this stream or using an on-site water treatment facility be studied. Additionally, it may be possible to find a buyer for the byproduct whose specifications are less stringent.

The other stream under consideration, Stream 24, consists of the non-condensed material from the flash vapor and has the following composition:

1.34 lb/h	(0.82 wt%)	water
1.44 lb/h	(0.88 wt%)	ethanol
117.4 lb/h	(85.3 wt%)	CO_2
0.152 lb/h	(0.10 wt%)	ethyl acetate
17.8 lb/h	(12.9 wt%)	air

The design calls for this material to be vented to the atmosphere, as none of the components are considered environmentally problematic. However, with increasing concern about the effect that carbon dioxide may have on global warming and ever-changing environmental laws, it is recommended that this stream be examined from a legal standpoint. It may be necessary to install a small scrubber or other suitable environmental protection device.

Table 13.1 Utility Summary

A. Electricity at $0.0494/kWhr

User	Yearly Cost
P-1A/B	$232.20
P-2A/B	$154.80
P-3A/B	$154.80
P-4A/B	$77.40
P-5A/B	$3096.00
P-6A/B	$464.30
P-7A/B	$154.80
P-8A/B	$154.80
Total	**$4489.10**

B. 100 psig Steam at $2.815/1000 lb

User	Amount (lb/h)	Yearly Cost
X-1	74.4	$1759.15
X-2 and X-3	371.0	$8872.09
E-9	4474.0	$105,785.29
E-10	169.0	$3995.91
Total	**5088.4**	**$120,312.44**

Table 13.1 (cont'd)

C. Cooling Water at $0.117/1000 gal

User	Amount (gpm)	Yearly Cost
E-5	10.9	$644.31
E-6	10.9	$644.31
E-8A	4.72	$279.00
Total	**26.52**	**$1567.63**

D. Ammonia Refrigeration at $0.0081/10³ Btu

User	Amount (10³ Btu/hr)	Yearly Cost
E-3	1214	$82,473.09
E-4	499.5	$33933.53
E-8B	23.5	$1596.47
Total	**1737**	**$118,003.10**

Total Utilities Cost = $244,372/yr

Table 14.1 Operating Cost Summary

Note: This table has been constructed according to Table 27, pages 210-211 of Peters and Timmerhaus. For a more complete table of costs, see Sections II and III of the economic analysis.

I. Manufacturing Costs

A. Direct Production Costs

1. Raw Materials	$45,395.47
2. Operating Labor ($21.00/hr per shift position)	$821,741.76
3. Direct Supervisory and Clerical Labor (17.5% of OL)	$143,804.81
4. Total Utilities	$244,372.27
5. Maintenance and Repairs (3% of Fixed Capital Investment)	$105,396.70
6. Operating Supplies (15% of Maintenance and Repairs)	$15,809.51
7. Laboratory charges (15% of Operating Labor)	$123,261.26

B. Fixed Charges

1. Local taxes and insurance (3% of Fixed Capital Investment)	$105,396.70

C. Plant Overhead Costs

1. 60% of Operating Labor, Supervision, and Maintenance	$642,595.96
Total =	**$2,247,774**

II. General Expenses

A. Administrative Costs

1. 15% of Operating Labor, Supervision, and Maintenance	$160,641.49

III. Total Product Cost

A. Manufacturing Costs + General Expenses	**$2,408,416**

IV. Gross Earnings Cost

A. (Total Income - Total Product Cost) x 40%	**$265,154.63**

Table 15.1 Equipment Information Summary

A. Pumps

Number	Description	Body Type	Suct. P (psia)	Disch. P (psia)	GPM	Head (ft)	Motor BHP
P-1A/B	Feed	1.5 x 1.5	14.7	28.0	53.8	31.0	0.75
P-2A/B	T-1 Feed	2 x 2	13.2	15.2	51.2	4.67	0.5
P-3/AB	Bottoms	2 x 2	2.0	8.0	56.5	13.9	0.5
P-4A/B	Reflux	1.25 x 1.25	0.5	3.5	13.8	8.6	0.25
P-5A/B	Hot water	4 x 3	14.7	31.0	571	38.1	10
P-6A/B	Product	1.5 x 1.5	7.0	28.0	48.9	48.5	1.5
P-7A/B	Distillate	1.5 x 1.5	2.5	17.0	2.75	41.8	0.5
P-8A/B	Hot water	1.5 x 1.5	14.7	26.0	30.1	26.4	0.5

B. Drums

Number	Description	Orientation	ID (ft)	L (ft)	Capacity (gal)	P (psia)	T (°F)
F-1	De-esterizer	Vertical	6.5	26.0	6454	13.2	79.2
F-2	Reflux	Horizontal	2.0	8.0	187.7	0.5	52.6
F-3	Recombiner	Vertical	3.0	12.0	634.3	7.0	125.6

C. Towers

Number	Description	Trays	Type	Feed to tray	ID (ft)	H (ft)
T-1	Alcohol removal	59	Sieve	39	5.0	135

D. Ejectors

Number	Description	Suction P (psia)	Discharge P (psia)	Stages	Steam/stage (lb/h)
X-1	F-1 Evacuator	13.2	15.7	1	74.4
X-2 and X-3	T-1 Evacuators	0.5	15.7	2	185.5

Table 15.1 (cont'd)

E. Heat Exchangers

Number	Description	Shell Side	Tube Side	Type	Duty (10^3 Btu/hr)	Area (ft^2)
E-1	Preheater	Feed	Product	F.T.S.*	1014	249
E-2	Preheater	Hot water	Feed	F.T.S.	149	38.9
E-3	Cooler	Ammonia	Product	F.T.S.	1214	270
E-4	Condenser	Overhead	Ammonia	F.T.S.	499.5	109
E-5	X-2 Condenser	Steam	CW	F.T.S.	163.4	10.0
E-6	X-2 Condenser	Steam	CW	F.T.S.	163.4	10.0
E-7	Reboiler	Hot water	Bottoms	F.T.S.	3940	2915
E-8A	CW Flash cond.	CW	Fsh. vapor	F.T.S.	86.2	11.3
E-8B	NH_3 Flash cond.	Ammonia	Fsh. vapor	F.T.S.	23.5	10.0
E-9	Hot water heater	Steam	H_2O circ.	F.T.S.	3940	126.3
E-10	Hot water heater	Steam	H_2O circ.	F.T.S.	149	10.0

*F.T.S. = Fixed-Tube-Sheet

F. Storage Tanks

Number	Description	Type	Capacity (gal)	Diameter (ft)	Height (ft)
V-1	NA beer storage	Cone-roof	563,281	57.5	29.0
V-2	Distillate storage	Cone-roof	31,277	22.0	11.0

Material of Construction: All of the above equipment is constructed of 304 stainless steel except P-5A/B, P-8A/B, X-2, X-3, E-5, E-6, E-9, and E-10. These pieces of equipment only see recirculating hot water and never process fluid.

Table 16.1 Total Capital Estimate

This table contains itemized total plant capital cost estimaies.

A. Direct Costs

Purchased equipment-delivered	$10,651,49.06
Purchased equipment-installation	$189,755.55
Instrumentation and Controls	$89,847.62
Piping	$329,441.29
Electrical	$54,906.88
Buildings	$106,514.91
Yard Improvements	$106,514.91
Service Facilities	$745,604.34
Total	**$2,687,735**

B. Indirect Costs

Engineering and Supervision	$351,499.19
Construction	$436,711.12
Total	**$788,210.31**

Total Direct and Indirect Plant Costs = $3,475,965

C. Fees and Contingency

Contractor's fee	$173,798.24
Contingency	$347,596.49

Table 16.1 (cont'd)

Fixed Capital Investment = $3,997,359

Working Capital (15% of TCI) = $705,416

<u>Total Capital Investment = $4,702,776</u>

Table 16.2 Individual Capital Estimate

The next three pages contain itemized capital cost estimates for individual pieces of equipment.

Table 16.2

Equipment	Number	Purchase Cost	Inflation Factor	Delivered Cost	Material	Pressure	Temperature	Total Del. Cost	Installation	Controls	Piping	Electrical
Pump	P1-A/B	3731	1	3917.55	1	1	1	3917.55	1841.25	705.16	2585.58	430.93
Pump	P2-A/B	4172	1	4380.60	1	1	1	4380.60	2058.88	788.51	2891.20	481.87
Pump	P3-A/B	4381	1	4600.05	1	1	1	4600.05	2162.02	828.01	3036.03	506.01
Pump	P4-A/B	2893	1	3037.65	1	1.1	1	3341.42	1570.47	601.45	2205.33	367.56
Pump	P5-A/B	4965	1	5213.25	1	1	1	5213.25	2450.23	938.39	3440.75	573.46
Pump	P6-A/B	5371	1	5639.55	1	1.1	1	6203.51	2915.65	1116.63	4094.31	682.39
Pump	P7-A/B	5115	1	5370.75	1	1.1	1	5907.83	2776.68	1063.41	3899.16	649.86
Pump	P8-A/B	3786	1	3975.30	1	1	1	3975.30	1868.39	715.55	2623.70	437.28
Drivers	1 to 8-A/B	4239	1	4450.95	1	1	1	4450.95	2091.95	801.17	2937.63	489.60
Drum	F-1	13160	1	13818.00	1	1	1	13818.00	6494.46	2487.24	9119.88	1519.98
Drum	F-2	3552	1	3729.60	1	1.2	1	4475.52	2103.49	805.59	2953.84	492.31
Drum	F-3	11879	1	12472.95	1	1.1	1	13720.25	6448.52	2469.64	9055.36	1509.23
*Tower	T-1	471663	1	471663.00	1	1.2	1	565995.60	0.00	0.00	0.00	0.00
*Purchase cost correlation includes installation and auxiliaries												
**Ejector	X-1	9073	1	9526.65	1	1	1	9526.65	0.00	1714.80	6287.59	1047.93
Ejector system	X-2, X-3, E-5, E-6	81761	1	85849.05	1	1	1	85849.05	0.00	15452.83	56660.37	9443.40
**Cost of ejectors includes installation												
Exchanger	E-1	6988	1	7337.40	1	1	1.05	7704.27	3621.01	1386.77	5084.82	847.47
Exchanger	E-2	7204	1	7564.20	1	1	1.05	7942.41	3732.93	1429.63	5241.99	873.67
Exchanger	E-3	7453	1	7825.65	1	1	1	7825.65	3678.06	1408.62	5164.93	860.82
Exchanger	E-4	4658	1	4890.90	1	1.2	1.05	6162.53	2896.39	1109.26	4067.27	677.88
Exchanger	E-5	Included in X-2 and X-3 ejector system										
Exchanger	E-6	Included in X-2 and X-3 ejector system										
Exchanger	E-7	38432	1	40353.60	1	1.1	1.05	46608.41	21905.95	8389.51	30761.55	5126.92
Exchanger	E-8A	2562	1	2690.10	1	1	1.05	2824.61	1327.56	508.43	1864.24	310.71
Exchanger	E-8B	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Exchanger	E-9	5124	1	5380.20	1	1	1.05	5649.21	2655.13	1016.86	3728.48	621.41
Exchanger	E-10	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Storage tank	V-1	139752	1	146739.60	1.39	1	1	203968.04	95864.98	36714.25	134618.91	22436.48
Storage tank	V-2	24457	1	25679.85	1.39	1	1	35694.99	16776.65	6425.10	23558.69	3926.45
							Totals	1065149.06	189775.55	89847.62	329441.29	54906.88

Table 16.2 (cont'd)

Building	Yard	Service	Total Direct Cost	Engineering & Supervision	Construction	Total Indirect
391 76	391.76	2742.29	13006.27	1292.79	1606.20	2898.99
438 06	438.06	3066.42	14543.59	1445.60	1796.05	3241.64
460 01	460.01	3220.04	15272.17	1518.02	1886.02	3404.04
334 14	334.14	2338.99	11093.50	1102.67	1369.98	2472.65
521 33	521.33	3649.28	17307.99	1720.37	2137.43	3857.81
620.35	620.35	4342.45	20595.64	2047.16	2543.44	4590.59
590.78	590.78	4135.48	19613.98	1949.58	2422.21	4371.79
397.53	397.53	2782.71	13198.00	1311.85	1629.87	2941.72
445 10	445.10	3115.66	14777.15	1468.81	1824.89	3293.70
1381.80	1381.80	9672.60	45875.76	4559.94	5665.38	10225.32
447 55	447.55	3132.66	14858.73	1476.92	1834.96	3311.88
1372.02	1372.02	9604.17	45551.21	4527.68	5625.30	10152.98
56599.56	56599.56	396196.92	1075391.64	186778.55	232058.20	418836.74
952 67	952.67	6668.66	27150.95	3143.79	3905.93	7049.72
8584.91	8584.91	60094.34	244669.79	28330.19	35198.11	63528.30
770.43	770.43	5392.99	25578.18	2542.41	3158.75	5701.16
794.24	794.24	5559.69	26368.80	2621.00	3256.39	5877.38
782 57	782.57	5477.96	25981.16	2582.46	3208.52	5790.98
616 25	616.25	4313.77	20459.61	2033.64	2526.64	4560.28
4660.84	4660.84	32625.89	154739.91	15380.77	19109.45	34490.22
282.46	282.46	1977.22	9377.69	932.12	1158.09	2090.21
269.67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
564.92	564.92	3954 45	18755.38	1864.24	2316.18	4180.42
269.67	269.67	1887 70	8953.09	889.92	1105.65	1995.57
20396.80	20396.80	142777.63	677173.91	67309.45	83626.90	150936.35
3569.50	3569.50	24986.49	118507.37	11779.35	14634.95	26414.29
106514.91	106514.91	745604.34	2687754.56	351499.19	436711.12	788210.31

Table 16.2 (cont'd)

Direct + Indirect	Contractor	Contingency	Fixed Capital Inv.	Working Capital	Total Capital Inv.
15905.25	795.26	1590.53	18291.04	3227.83	21518.87
17785.24	889.26	1778.52	20453.02	3609.36	24062.38
18676.20	933.81	1867.62	21477.63	3790.17	25267.80
13566.14	678.31	1356.61	15601.07	2753.13	18354.20
21165.80	1058.29	2116.58	24340.66	4295.41	28636.08
25186.23	1259.31	2518.62	28964.16	5111.32	34075.49
23985.77	1199.29	2398.58	27583.63	4867.70	32451.34
16139.72	806.99	1613.97	18560.68	3275.41	21836.09
18070.86	903.54	1807.09	20781.49	3667.32	24448.81
56101.08	2805.05	5610.11	64516.24	11385.22	75901.46
18170.61	908.53	1817.06	20896.20	3687.57	24583.77
55704.19	2785.21	5570.42	64059.82	11304.67	75364.50
1494228.38	74711.42	149422.84	1718362.64	303240.47	2021603.11
34200.67	1710.03	3420.07	39330.77	6940.72	46271.50
308198.09	15409.90	30819.81	354427.80	62546.08	416973.89
31279.34	1563.97	3127.93	35971.24	6347.87	42319.10
32246.18	1612.31	3224.62	37083.11	6544.08	43627.19
31772.14	1588.61	3177.21	36537.96	6447.88	42985.84
25019.89	1250.99	2501.99	28772.87	5077.57	33850.44
189230.14	9461.51	18923.01	217614.66	38402.59	256017.24
11467.90	573.39	1146.79	13188.08	2327.31	15515.39
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
22935.79	1146.79	2293.58	26376.16	4654.62	31030.78
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
828110.26	41405.51	82811.03	952326.80	168057.67	1120384.47
144921.67	7246.08	14492.17	166659.92	29410.57	196070.49
3475964.86	173798.24	347596.49	3997359.59	705416.40	4702775.99

Grand Total Capital Investment: $4,702,776

Economic Methods and Analysis

The economic analysis for this project was done using the discounted cash flow methods espoused by Peters and Timmerhaus and White, Agee, and Case. Depreciation, taxes, and other premises were as stated in section 8 of this report. Several additional assumptions were made, however, and warrant discussion here.

Assumptions

1. Item 2 on page 4 of the problem statement says that Prairie Premium has a Before Tax Net Income of $5.00/bbl. Item 8, page 5, then states that the non-alcoholic beer will have the same "netback" as Prairie Premium. "Netback" is a term with which PRO-D-ZINE is unfamiliar, but these statements have been interpreted to mean that the Before Tax Net Income of the NA beer is also $5.00/bbl.

2. The opportunity cost of selling NA beer is $5.00/bbl sold. Thus, before taxes are considered, the net profit from this operation is zero.

3. Excise taxes were estimated from data taken from the Anheuser-Busch 1994 Annual Report and were considered to be $0.388/gal for a capacity of less than 60,000 bbl per year and $0.448/gal otherwise. The money saved in excise taxes by selling excise-exempt NA beer, rather than Prairie Premium, is the source of profit in this operation.

4. Regarding item 3, page 4, of the problem statement, since the cost of brewing the feed to the process is the same as the cost of brewing *any* of the regular beers, it is also the same as the cost of brewing the Prairie Premium. Since a net income figure ($5.00/bbl) was provided for the Prairie Premium, there was no distinct "feed cost" for this design problem. The feed cost is included in the $5.00/bbl figure.

5. A 12% minimum attractive rate of return was used in Net Present Worth calculations. Item 5, page 5, of the problem statement says that the acceptable range of IRR is from 12 to 20%. Therefore, 12% must be the *minimum* attractive rate of return.

6. All costs were inflated first to 1994 dollars using ratios of the Chemical Engineering Index and then to 1999 dollars using the 2.5% per year inflation rate specified in the problem statement.

7. The economic analysis was conducted as though no sales of the mainly ethanol byproduct were possible. The specifications on ethanol, even fuel grade, are too stringent and producing such a byproduct was judged uneconomical.

Further details of the economic methods used can be found in the problem statement, in the calculation appendix, and in the economic calculation example at the end of this section. It should be noted, however, that the economic example was begun before the actual process design. Therefore, it should be read through in its entirety, keeping in mind that several assumptions and premises changed as the work progressed.

Results

The economic analysis of the base case, before any optimization whatsoever was performed, was too poor (- $9.5 million present worth) to report here and can be found in the calculation appendix. Once the distillation column was optimized, the economics improved somewhat, but were still poor, yielding only a 7.4% rate of return. The cash flow table for this optimization is shown in Table 17.1.

In order to make the process profitable, the capacity was optimized using the six-tenths rule, as detailed in the Innovation and Optimization section of this report. A graphical representation of the results of this optimization can be seen in Figure 17.1. The optimization analysis showed that at a capacity of 614,315 bbl/yr, the process would yield an internal rate of return of exactly 20%, the upper limit given in the problem statement. This economic analysis is shown in Table 17.2.

Curiously, when the final, rigorous economic calculations were performed at the above capacity, the Net Present Worth and the Internal Rate of Return were both much higher than expected, as described earlier in this report. For reasons discussed in the next section, the six-tenths rule and the scale-up of operating costs were not accurate for this process. Fortunately, the process made more money than expected, not less. The detailed results of the final economic analysis are shown in Table 17.3.

Additionally, one cash flow analysis was conducted in which the selling price of the NA beer was $3.00/bbl higher than that of Prairie Premium. The results can be seen in Table 17.4.

Table 17.1

Optimization I Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	2040826.70	568339.07	391374.87	529507.17				
2000	1					2232901.44	350000	238733	5.00
2001	2					2232901.44	250000	238733	5.00
2002	3					2232901.44	150000	238733	5.00
2003	4					2232901.44	90000	238733	5.00
2004	5					2232901.44		238733	5.00
2005	6					2232901.44		238733	5.00
2006	7					2232901.44		238733	5.00
2007	8					2232901.44		238733	5.00
2008	9					2232901.44		238733	5.00
2009	10					2232901.44		238733	5.00

Table 17.1 (cont'd)

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-3530047.81
1193665	1193665	4492000.128	0	0	0.1429	291634.14	3020627.89	1208251.15	700847.53
1193665	1193665	4492000.128	0	0	0.2449	499798.46	3992201.67	1596880.67	412218.02
1193665	1193665	4492000.128	0	0	0.1749	356940.59	4135059.54	1654023.82	455074.87
1193665	1193665	4492000.128	0	0	0.1249	254899.25	4237100.87	1694840.35	474258.34
1193665	1193665	4492000.128	0	0	0.0893	182245.82	4309754.30	1723901.72	535196.97
1193665	1193665	4492000.128	0	0	0.0892	182041.74	4309958.39	1723983.35	535115.34
1193665	1193665	4492000.128	0	0	0.0893	182245.82	4309754.30	1723901.72	535196.97
1193665	1193665	4492000.128	0	0	0.0446	91020.87	4400979.26	1760391.70	498706.99
1193665	1193665	4492000.128	0	0			4492000.13	1796800.05	462298.64
1193665	1193665	4492000.128	0	0			4492000.13	1796800.05	462298.64
								NPW @ 12%	-6.1650E+05
								IRR	0.074
								NPW @ IRR	5.6E-11

Net Present Worth at 12 %
- ($616,497)

Internal Rate of Return =
0.074

Table 17.2

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
2.573	$6,223,972	$5,292,494	5600052

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	3763349.43	838818.85	690325.24
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Table 17.7 (cont'd)

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
933969.45						
	5600051.80	350000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80	250000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80	150000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80	90000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967

Table 17.2 (cont'd)

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($6,226,463)
11558945.32	0	0	0.1429	537782.63	9841424.58	3936569.83	$1,672,324
11558945.32	0	0	0.2449	921644.28	10637301.04	4254920.42	$1,453,973
11558945.32	0	0	0.1749	658209.82	10900735.50	4360294.20	$1,448,599
11558945.32	0	0	0.1249	470042.34	11088902.97	4435561.19	$1,433,332
11558945.32	0	0	0.0893	336067.10	11222878.21	4489151.29	$1,469,742
11558945.32	0	0	0.0892	335690.77	11223254.55	4489301.82	$1,469,592
11558945.32	0	0	0.0893	336067.10	11222878.21	4489151.29	$1,469,742
11558945.32	0	0	0.0446	167845.38	11391099.93	4556439.97	$1,402,454
11558945.32	0	0			11558945.32	4623578.13	$1,335,315
11558945.32	0	0			11558945.32	4623578.13	$1,335,315
						NPW @ 12%	$2,089,015
						IRR	0.200
						NPW @ IRR	6.8E-11

Net Present Worth at 12 %
$2,089,016

Internal Rate of Return =
0.200

Fig. 17.1 IRR and NPW vs. Capacity

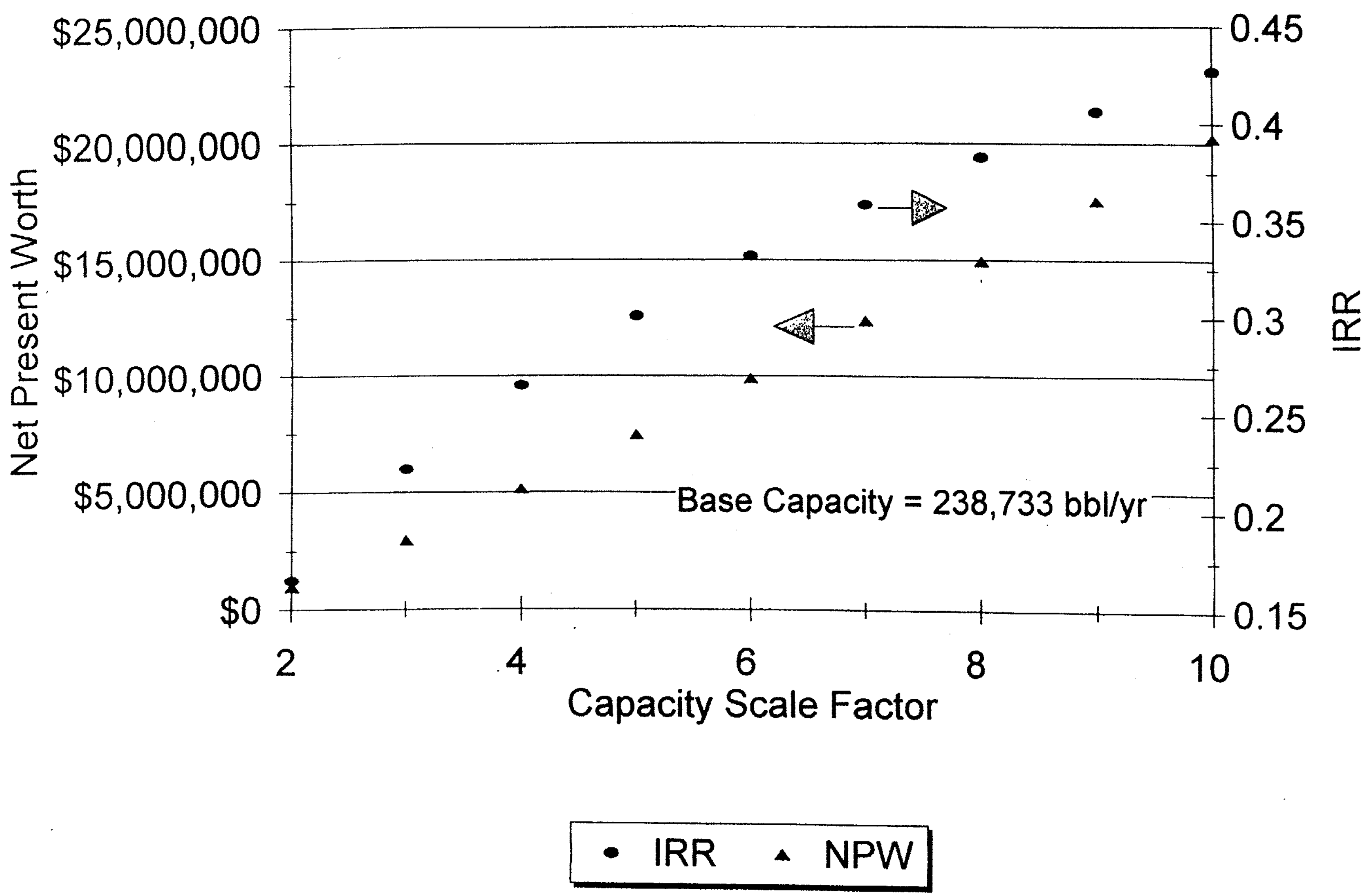


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Table 17.3

Final Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section I: Capital Investment

Operating Pressure (psia)	Correction
0 08	1.3
0 2	1.2
0 7	1.1
8 to 100	1.0
700	1.1
3000	1.2
6000	1.3

Operating Temperature (C)	Correction
-80	1.3
0	1.0
100	1.05
600	1.1
5000	1.2
10000	1.4

Material	Correction
Carbon steel	1.0
Bronze	1.05
Carbon/molybdenum	1.065
Aluminum	1.075
Cast steel	1.11
Stainless steel	1.28 to 1.5
Worthite alloy	1.41
Hastelloy C alloy	1.54
Monel alloy	1.65
Nickel/inconel alloy	1.71
Titanium	2.0

Note. Correcting factors may have already been taken into account in earlier estimates. For example, inflation factors are shown as 1.0, since inflation of 1.1646 (see Economic Calculations, p. 9) is already included

Table 17.3 (cont'd)

Equipment	Number	Purchase Cost	Inflation Factor	Delivered Cost	Material	Pressure	Temperature	Total Del. Cost	Installation	Controls	Piping	Electrical
Pump	P1-A/B	3731	1	3917.55	1	1	1	3917.55	1841.25	705.16	2585.58	430.93
Pump	P2-A/B	4172	1	4380.60	1	1	1	4380.60	2058.88	788.51	2891.20	481.87
Pump	P3-A/B	4381	1	4600.05	1	1	1	4600.05	2162.02	828.01	3036.03	506.01
Pump	P4-A/B	2893	1	3037.65	1	1.1	1	3341.42	1570.47	601.45	2205.33	367.56
Pump	P5-A/B	4965	1	5213.25	1	1	1	5213.25	2450.23	938.39	3440.75	573.46
Pump	P6-A/B	5371	1	5639.55	1	1.1	1	6203.51	2915.65	1116.63	4094.31	682.39
Pump	P7-A/B	5115	1	5370.75	1	1.1	1	5907.83	2776.68	1063.41	3899.16	649.86
Pump	P8-A/B	3786	1	3975.30	1	1	1	3975.30	1868.39	715.55	2623.70	437.28
Drivers	1 to 8-A/B	4239	1	4450.95	1	1	1	4450.95	2091.95	801.17	2937.63	489.60
Drum	F-1	13160	1	13818.00	1	1	1	13818.00	6494.46	2487.24	9119.88	1519.98
Drum	F-2	3552	1	3729.60	1	1.2	1	4475.52	2103.49	805.59	2953.84	492.31
Drum	F-3	11879	1	12472.95	1	1.1	1	13720.25	6448.52	2469.64	9055.36	1509.23
*Tower	T-1	471663	1	471663.00	1	1.2	1	565995.60	0.00	0.00	0.00	0.00
*Purchase cost correlation includes installation and auxiliaries												
**Ejector	X-1	9073	1	9526.65	1	1	1	9526.65	0.00	1714.80	6287.59	1047.93
Ejector system	X-2, X-3, E-5, E-6	81761	1	85849.05	1	1	1	85849.05	0.00	15452.83	56660.37	9443.40
**Cost of ejectors includes installation												
Exchanger	E-1	6988	1	7337.40	1	1	1.05	7704.27	3621.01	1386.77	5084.82	847.47
Exchanger	E-2	7204	1	7564.20	1	1	1.05	7942.41	3732.93	1429.63	5241.99	873.67
Exchanger	E-3	7453	1	7825.65	1	1	1	7825.65	3678.06	1408.62	5164.93	860.82
Exchanger	E-4	4658	1	4890.90	1	1.2	1.05	6162.53	2896.39	1109.26	4067.27	677.88
Exchanger	E-5	Included in X-2 and X-3 ejector system										
Exchanger	E-6	Included in X-2 and X-3 ejector system										
Exchanger	E-7	38432	1	40353.60	1	1.1	1.05	46608.41	21905.95	8389.51	30761.55	5126.92
Exchanger	E-8A	2562	1	2690.10	1	1	1.05	2824.61	1327.56	508.43	1864.24	310.71
Exchanger	E-8B	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Exchanger	E-9	5124	1	5380.20	1	1	1.05	5649.21	2655.13	1016.86	3728.48	621.41
Exchanger	E-10	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Storage tank	V-1	139752	1	146739.60	1.39	1	1	203968.04	95864.98	36714.25	134618.91	22436.48
Storage tank	V-2	24457	1	25679.85	1.39	1	1	35694.99	16776.65	6425.10	23558.69	3926.45
							Totals	1065149.06	189775.55	89847.62	329441.29	54906.88

Table 17.3 (cont'd)

Building	Yard	Service	Total Direct Cost	Engineering & Supervision	Construction	Total Indirect
391.76	391.76	2742.29	13006.27	1292.79	1606.20	2898.99
438.06	438.06	3066.42	14543.59	1445.60	1796.05	3241.64
460.01	460.01	3220.04	15272.17	1518.02	1886.02	3404.04
334.14	334.14	2338.99	11093.50	1102.67	1369.98	2472.65
521.33	521.33	3649.28	17307.99	1720.37	2137.43	3857.81
620.35	620.35	4342.45	20595.64	2047.16	2543.44	4590.59
590.78	590.78	4135.48	19613.98	1949.58	2422.21	4371.79
397.53	397.53	2782.71	13198.00	1311.85	1629.87	2941.72
445.10	445.10	3115.66	14777.15	1468.81	1824.89	3293.70
1381.80	1381.80	9672.60	45875.76	4559.94	5665.38	10225.32
447.55	447.55	3132.86	14858.73	1476.92	1834.96	3311.88
1372.02	1372.02	9604.17	45551.21	4527.68	5625.30	10152.98
56599.56	56599.56	396196.92	1075391.64	186778.55	232058.20	418836.74
952.67	952.67	6668.66	27150.95	3143.79	3905.93	7049.72
8584.91	8584.91	60094.34	244669.79	28330.19	35198.11	63528.30
770.43	770.43	5392.99	25578.18	2542.41	3158.75	5701.16
794.24	794.24	5559.69	26368.80	2621.00	3256.39	5877.38
782.57	782.57	5477.96	25981.16	2582.46	3208.52	5790.98
616.25	616.25	4313.77	20459.61	2033.64	2526.64	4560.28
4660.84	4660.84	32625.89	154739.91	15380.77	19109.45	34490.22
282.46	282.46	1977.22	9377.69	932.12	1158.09	2090.21
269.67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
564.92	564.92	3954.45	18755.38	1864.24	2316.18	4180.42
269.67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
20396.80	20396.80	142777.63	677173.91	67309.45	83626.90	150936.35
3569.50	3569.50	24986.49	118507.37	11779.35	14634.95	26414.29
106514.91	106514.91	745604.34	2687754.56	351499.19	436711.12	788210.31

Table 17.3 (cont'd)

Direct + Indirect	Contractor	Contingency	Fixed Capital Inv.	Working Capital	Total Capital Inv.
15905.25	795.26	1590.53	18291.04	3227.83	21518.87
17785.24	889.26	1778.52	20453.02	3609.36	24062.38
18676.20	933.81	1867.62	21477.63	3790.17	25267.80
13566.14	673.31	1356.61	15601.07	2753.13	18354.20
21165.80	1058.29	2116.58	24340.66	4295.41	28636.08
25186.23	1259.31	2518.62	28964.16	5111.32	34075.49
23985.77	1199.29	2398.58	27583.63	4867.70	32451.34
16139.72	806.99	1613.97	18560.68	3275.41	21836.09
18070.86	903.54	1807.09	20781.49	3667.32	24448.81
56101.08	2805.05	5610.11	64516.24	11385.22	75901.46
18170.61	908.53	1817.06	20896.20	3687.57	24583.77
55704.19	2785.21	5570.42	64059.82	11304.67	75364.50
1494228.38	74711.42	149422.84	1718362.64	303240.47	2021603.11
34200.67	1710.03	3420.07	39330.77	6940.72	46271.50
308198.09	15409.90	30819.81	354427.80	62546.08	416973.89
31279.34	1563.97	3127.93	35971.24	6347.87	42319.10
32246.18	1612.31	3224.62	37083.11	6544.08	43627.19
31772.14	1588.61	3177.21	36537.96	6447.88	42985.84
25019.89	1250.99	2501.99	28772.87	5077.57	33850.44
189230.14	9461.51	18923.01	217614.66	38402.59	256017.24
11467.90	573.39	1146.79	13188.08	2327.31	15515.39
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
22935.79	1146.79	2293.58	26376.16	4654.62	31030.78
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
828110.26	41405.51	82811.03	952326.80	168057.67	1120384.47
144921.67	7246.08	14492.17	166659.92	29410.57	196070.49
3475964.86	173798.24	347596.49	3997359.59	705416.40	4702775.99

Grand Total Capital Investment:
$4,702,776

Table 17.3 (cont'd)

Final Optimization Economic Calculations

All cost figures in dollars per year unless otherwise stated

Section II: Operating Costs

Operator Positions	Inflation Factor	Operating Labor	Direct Supervision & Clerical Labor	Maintenance & Repairs	Operating Supplies	Laboratory Charges	Local taxes & Insurance	Plant Overhead	Administrative Expenses
1	1.1646	821741.76	143804.81	105396.70	15809.51	123261.26	105396.70	642565.96	160641.49

Steam use (lb/h)	Steam cost
74.4	1759.15
169	3995.91
371	8772.09
4474	105785.29
Total	120312.44

Process Water (gpm)	Process Water Cost
0.00	0.00

Cooling Water (gpm)	Cooling Water Cost
4.72	279.00
10.9	644.31
10.9	644.31
Total	1567.63

CO_2 (lb/h)	CO_2 Cost
129.80	45395.47

	Electricity Costs
	232.20
	154.80
	154.80
	77.40
	3096.00
	464.30
	154.80
	154.80
Total	4489.10

Refrigeration (1000 Btu/hr)	Cost at $2.00/288KBtu	Inflation factor	1999 Cost
23.5	1370.83	1.1646	1596.47
499.5	29137.50	1.1646	33933.53
1214	70816.67	1.1646	82473.09
		Total	118003.10

Total Operating Cost
2.4084E+06 $/yr

Table 17.3 (cont'd)

Final Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	2687754.56	788210.31	521394.73	705416.40				
2000	1					2450505.78	350000	614260	5.00
2001	2					2450505.78	250000	614260	5.00
2002	3					2450505.78	150000	614260	5.00
2003	4					2450505.78	90000	614260	5.00
2004	5					2450505.78		614260	5.00
2005	6					2450505.78		614260	5.00
2006	7					2450505.78		614260	5.00
2007	8					2450505.78		614260	5.00
2008	9					2450505.78		614260	5.00
2009	10					2450505.78		614260	5.00

Table 17.3 (cont'd)

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-4702775.99
3071300	3071300	11557916.16	0	0	0.1429	384080.13	9946998.23	3978799.29	4778611.09
3071300	3071300	11557916.16	0	0	0.2449	658231.09	10899685.07	4359874.03	4497536.35
3071300	3071300	11557916.16	0	0	0.1749	470088.27	11087827.89	4435131.16	4522279.22
3071300	3071300	11557916.16	0	0	0.1249	335700.54	11222215.62	4488886.25	4528524.13
3071300	3071300	11557916.16	0	0	0.0893	240016.48	11317899.68	4527159.87	4580250.51
3071300	3071300	11557916.16	0	0	0.0892	239747.71	11318168.45	4527267.38	4580143.00
3071300	3071300	11557916.16	0	0	0.0893	240016.48	11317899.68	4527159.87	4580250.51
3071300	3071300	11557916.16	0	0	0.0446	119873.85	11438042.31	4575216.92	4532193.46
3071300	3071300	11557916.16	0	0			11557916.16	4623166.46	4484243.92
3071300	3071300	11557916.16	0	0			11557916.16	4623166.46	4484243.92
								NPW @ 12%	2.1129E+07
								IRR	0.987
								NPW @ IRR	6.4E-11

Net Present Worth at 12 %
$21,128,697

Internal Rate of Return =
0.987

Table 17.4 Unequal netback

Final Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	2687754.56	788210.31	521394.73	705416.40				
2000	1					2450505.78	350000	614260	8.00
2001	2					2450505.78	250000	614260	8.00
2002	3					2450505.78	150000	614260	8.00
2003	4					2450505.78	90000	614260	8.00
2004	5					2450505.78		614260	8.00
2005	6					2450505.78		614260	8.00
2006	7					2450505.78		614260	8.00
2007	8					2450505.78		614260	8.00
2008	9					2450505.78		614260	8.00
2009	10					2450505.78		614260	8.00

Table 17.4 (cont'd)

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-4702775.99
4914080	3071300	11557916.16	0	0	0.1429	384080.13	11789778.23	4715911.29	5884279.09
4914080	3071300	11557916.16	0	0	0.2449	658231.09	12742465.07	5096986.03	5603204.35
4914080	3071300	11557916.16	0	0	0.1749	470088.27	12930607.89	5172243.16	5627947.22
4914080	3071300	11557916.16	0	0	0.1249	335700.54	13064995.62	5225998.25	5634192.13
4914080	3071300	11557916.16	0	0	0.0893	240016.48	13160679.68	5264271.87	5685918.51
4914080	3071300	11557916.16	0	0	0.0892	239747.71	13160948.45	5264379.38	5685811.00
4914080	3071300	11557916.16	0	0	0.0893	240016.48	13160679.68	5264271.87	5685918.51
4914080	3071300	11557916.16	0	0	0.0446	119873.85	13280822.31	5312328.92	5637861.46
4914080	3071300	11557916.16	0	0			13400696.16	5360278.46	5589911.92
4914080	3071300	11557916.16	0	0			13400696.16	5360278.46	5589911.92
								NPW @ 12%	2.7376E+07
								IRR	1.226
								NPW @ IRR	1.9E-10

Net Present Worth at 12 %
$27,375,968

Internal Rate of Return =
1.226

Economic Calculation Example

Preliminary Economics

Purpose: Before I tackle the problem completely, I'd like to pull off a little bit that I can handle and get out of the way. Also, I'd like to get some idea of what I'll need to calculate and what I'll need to do. Therefore, I'll set up the economics first.

The actual economics will be done on a spreadsheet, but I'll define the bases here.

Assumptions / Calculation Basis

There are a few parts of the problem statement that seem to require some interpretation. I'll do my best. However, these assumptions may change the outcome.

1. Referring to items 2 and 8, pages 4 and 5 of problem statement, the Before Tax Net Income of PP is \$5.00/bbl. Since NAB
(it says "profit")
has the same "netback", I'll assume that the Before Tax Net
(whatever this means)
Income of NAB is also \$5.00/bbl.

2. Since the profit from PP is to be considered an opportunity cost, it follows that the combination of profit from NAB and cost of PP is:

$$\$5.00\ bbl\ (NAB) - \$5.00\ (PP) = 0$$

This can be adjusted by \$3-\$5/bbl if necessary.

** Note: this zero figure may seem like a problem. However, consider that there are no excise taxes to be paid on NAB.

3. Regarding item 3, page 4 of problem statement, since the cost of brewing the feed to the process is the same as the cost of brewing any of the regular beers, it is also the same as the cost of brewing the PP. Since we have a net income (\$5/bbl) for the PP, there is no separate "feed cost" for this design problem. It is included in the \$5.00/bbl figure. This assumption will be important.

Capital Investment

After the process design is completed, there will be several pieces of equipment comprising the total capital investment. Following the techniques of Peters and Timmerhaus each piece of equipment will have a line something like this example:

Equip	Purchase cost	Inflation from 1990 to 1999 $	Delivered cost
Pump	$2000	$2000(1+0.025)^9 = 2498$	$2498(1+0.05) = 2623$

given in problem statement

P+T uses January 1, 1990 figures (Preface, page xiii)

Freight = 5% of purchase price

Material correction	P correction	T correction	Purchased equip. delivered
$2623(1.39) = 3646$	$3646(1.0)$	$3646(1.0)$	$3646

(P+T give 1.28-1.5 values for stainless steel

... NEXT, upon summing all of the purchased equipment delivered costs, I will add in installment costs as follows (Fluid-processing plant)

Total delivered equipment	Installation Factor	Controls Factor	Piping Factor	Electrical Factor	Building Factor
$3646	0.47	0.18	0.66	0.11	0.10

Yard Factor	Service factor	Total direct plant cost
0.10	0.70	$3646(0.47+0.18+0.66+0.11+0.10+0.10+0.70+1)$

$$= 3646(3.32) \boxed{= 8459}$$

Oops. Mistake. This should be 12,105 but I won't change it now so as not to throw off the rest of the example..

** Note: I have intentionally left out the cost of land. It is assumed that the land is already available and owned by Gulf Brewery

Now put in the indirect costs:

Delivered cost	Engineering + supervision	Construction expenses
2623	$2623(0.33) = 866$	$2623(0.41) = 1075$

Another mistake, Should be 3646

Total indirect costs = \$1941

Sum the indirect and direct costs:

$$8459 + 1941 = \$10400$$

Add in contingency and contractor fees:

Direct + Indirect	Contractor	Contingency	Fixed capital investment
\$10400	$0.05(10400) = 520$	$0.10(10400) = 1040$	\$11960

* Working capital is next. As per Table 26 of P+T, it can be from 10-20% of the total capital investment. I'll compromise and use 15%. I need to solve for it:

$$\text{Fixed capital} + \text{working capital} = \text{total capital}$$

$$\text{Fixed capital} + 0.15(\text{total capital}) = \text{total capital}$$

$$\Rightarrow \text{total capital} = \frac{\text{Fixed capital}}{0.85}$$

It follows that

$$\text{working capital} = 0.15\left(\frac{\text{fixed capital}}{0.85}\right)$$

So for this example:

Fixed capital	Working capital	Total capital investment
\$11960	$0.15\left(\frac{11960}{0.85}\right) = 2111$	$11960 + 2111$ $\boxed{= \$14071 \text{ for this example}}$

$\Longrightarrow$ Once the design is in place, the real calculations will be performed in exactly this manner

Operating Costs

Another major component of the design economics is the operating costs. Once again, I'll go through an example of how the calculations will be performed. This is a continuation of the previous example.

* Stream factor: I'll assume that the plant runs on a continuous basis 95.8% of the time. This percentage corresponds to 8400 operating hours per year

Operating Labor (OL): as given in problem statement (item 6, page 5), the process will require 1 operator position. On page 8, it is stated that there are 4 operators per shift position. Table 23 of P+T gives the labor rate for operators at 21.00 $/h. Thus, the cost of labor is:

$$OL = (1\ \text{position})(4\ \text{operators}/\text{position})\left(8400\ \frac{h}{yr}\right)\left(21\ \frac{\$}{h}\right)(1.025)^{10}$$

convert to 1999 dollars using problem statement supplied 2.5% per year
Table is in 1989 dollars

$$= \underline{\underline{\$\,903228\ /yr}}$$ (no small sum)

... Now for some other operating costs (per percentages given in problem statement)

Direct Supervision and Clerical Labor	Maintenance + Repairs	Operating Supplies	Laboratory Charges
0.175 (903228)	0.03 (11960)	0.15 (359)	0.15 (903228)
= 158065	= 359	= 54	= 135485
17.5% of OL	3% of fixed capital investment	15% of M + R	15% of OL

⟶ I don't anticipate any patents or royalty charges. However, I do anticipate:

Local Taxes + Insurance	Plant Overhead	Administrative Expenses
0.03 (11960) = 359	0.60 (903228 + 158065 + 359) = 637051	0.15 (903228 + 158065 + 359) = 159263
3% of fixed capital investment	60% of OL, Supervision, and Maintenance	15% of OL, Supervision, and Maintenance

... Utility costs are the final piece of operating costs. I anticipate there being a need for 100 psig steam, process water (possibly), and cooling water.

Costs of these utilities are given by P+T. For sake of example, I'll assume 300 lb/h of steam, 500 gal/h process water, 500 gal/h cooling water. In this case, the utility costs will be:

page 200 of P+T, 1989 costs

For steam: $\left(300\ \frac{lb}{h}\right)\left(\frac{2.40\ \$}{1000\ lb}\right)\left(8400\ \frac{h}{yr}\right)\left(1.025\right)^{10} = 7742\ \$/yr$

P+T value; inflation from 1989

For process H_2O: $\left(500\ \frac{gal}{h}\right)\left(\frac{0.80\ \$}{1000\ gal}\right)\left(8400\ \frac{h}{yr}\right)\left(1.025\right)^{10} = 4301\ \$/yr$

For cooling H_2O: $\left(500\ \frac{gal}{h}\right)\left(\frac{0.10\ \$}{1000\ gal}\right)\left(8400\ \frac{h}{yr}\right)\left(1.025\right)^{10} = 538\ \$/yr$

Total utility $= 12581\ \$/yr$

The total operating costs come out to:

903328
158065
359
54
135485
359
637051
159263
12581

$\boxed{\$\ 2,006,545\ /yr}$

This concludes the example of how operating costs will be calculated.

Taxes and Depreciation (34% Federal, 6% State)

Taxes and depreciation are the next logical step in the economic analysis. Of course, depreciation can be deducted from taxable income. However, there are additional deductions listed by White, Agee, and Case (*Principles of Engineering Economic Analysis*, 3rd ed., John Wiley and Sons, New York, 1989). p.288 These items include:

salaries
wages
repairs
rent
taxes (other than income taxes)

Now, tax codes change quite often, so it's a little bit of a guessing game, but I will assume that for this problem I can deduct:

Operating Labor
Direct Supervision and Clerical Labor
Maintenance + Repairs

what I believe are conservative estimates $\left\{ \begin{array}{l} \text{50\% of local taxes and insurance (deduct the taxes, not insurance)} \\ \text{50\% of administrative expenses (for executive salaries)} \end{array} \right.$

And of course depreciation. Per problem statement, I'll use 7 yrs MACRS

Year of Project	MACRS Factor
1	0.1429
2	0.2449
3	0.1749
4	0.1249
5	0.0893
6	0.0892
7	0.0893
8	0.0446

(Table 4 of P+T, p. 287)

My interpretation of the depreciation rules (see also p. 262 of White, Agee, and Case) is that it applies to all of the <u>Direct Plant Costs</u>, <u>but not to the Indirect Costs, the Contractor's fee, the Contingency, or the Working Capital.</u>

⟶ <u>Excise taxes</u>

There are, of course, no excise taxes on NA beer, but since these economics are being done on the basis of producing NAB instead of Prairie Premium, I can claim as income <u>the excise taxes that I don't have to pay by producing NAB rather than PP.</u>

This idea is key to the whole project

The best immediately available source of information on excise taxes is page 45 of the Anheuser - Busch 1994 Annual Report (page R-1). It gives values of total federal and state excise taxes paid and it also gives gross profits for 1994, 1993, and 1992.

	1994	1993	1992
Federal + state (1) excise taxes	1679.7 mm\$	1679.8 mm\$	1668.6 mm\$
Gross Profits (2)	4269.4 mm\$	4085.6 mm\$	4084.6 mm\$
(1) ÷ (2)	0.3934	0.4112	0.4085

The computed values in the third row vary a little bit, but they're fairly consistent. The state excise taxes depend on what state(s) have the most beer sales. For purposes of this problem, I'll use the mean of the 3 values:

$$\boxed{\frac{\text{Excise taxes}}{\text{Gross Profits}} = 0.4044}$$

This is only an estimate, and may be revised at a later date, but for now it will play an important role in the economic calculations.

Advertising

As per the problem statement, there are to be extraordinary advertising costs during the first four years of the project:

Year 1	\$350,000
Year 2	\$250,000
Year 3	\$150,000
Year 4	\$90,000

Sales of byproducts

I do not know yet what the state of the byproducts will be, but I anticipate there being some money to be made in selling the ethanol that is removed from the beer. The going price according to *Chemical Marketing Reporter* is:

190 pf. \$2.56 - \$2.90/gal ⇒ Ave = \$2.73/gal

fuel grade \$1.08 - \$1.20/gal ⇒ Ave = \$1.14/gal

... Putting it All Together

Of course, when the different cases are calculated, the economics will be done on a spreadsheet, but here is an example of how they will be calculated.

A Beginning of Year	B End of year	C Direct Plant Costs	D Indirect Plant Costs	E Fees and Contingency
1999	0	8459	1941	1040 + 520 = 1560
2000	1	—	—	—

F Working Capital	G Total Operating Costs	H Advertising	I NAB Sold (bbl)
2111	—	—	—
—	2,006,545	350,000	4×10^6

J Profit from NAB	K Opportunity Cost of NAB	L Excise taxes saved	M Sale of byproducts (gal)
—	—	—	—
$5(4\times10^6) = 20\times10^6$	$5(4\times10^6) = 20\times10^6$	$0.41044\,(20\times10^6) = 8.088\times10^6$	400,000

** Assuming equal netback

N Profit from byproducts	O MACRS Factor	$P = O \times C$ Depreciation	$Q = J - K + L + N - P$ Taxable Income
—	—	—	—
$1.14\,(400{,}000) = 456000$	0.1429	$0.1429(8459) = 1209$	$20\times10^6 - 20\times10^6 + 8.088\times10^6 + 456000 - 1209 = 8{,}542{,}791$

... Putting it All Together [cont'd]

$R = 0.40(Q)$ Taxes	$S = J - K + L + N - R - H - G - F - E - D - C$ Net Cash Flow
—	-14071
$0.40(8542791)$ $= 3,417,116$	$20\times10^6 - 20\times10^6 + 8.088\times10^6 + - 3,417,116 - 350000 - 2,006,545$ $= \underline{\underline{2,314,339}}$

And so on. This was just an example, but in the real calculations, the analysis will be continued for a project life of 10 years. Then the IRR* can be determined. Also, the net present worth can be determined at different interest rates. Based on this analysis, the optimum design can be determined. It will also be possible to determine whether or not it is necessary to claim a higher netback for NAB than for PP.

This concludes the preliminary economic calculation example.

* Note: as per problem statement, the IRR may vary continuously between 12% and 20%

Additions:

Since completing the economic example, it has come to my attention that a better way to convert 1990 dollars to 1999 dollars is to use the Chemical Engineering Index (CEI) to convert to 1994 dollars (the latest information I can find), then use the assumed 2.5%/yr inflation rate.

CEI value in 1989 = 355
CEI value in 1990 = 357.5
CEI value in 1994 = 368

Therefore, the inflation factor from 1990 to 1999 will be:

$$I.F._{1990\rightarrow 1999} = \underbrace{\left(\frac{368}{357.5}\right)}_{1990 \text{ to } 1994}\underbrace{\left(1.025\right)^5}_{1994 \text{ to } 1999} = \boxed{1.1646}$$

Price of CO_2

I don't know yet, but the design may involve the purchase of CO_2. The only data on its price that I could find was from 1985, during which year 4 million tons worth \$260 million were produced.
Converting to a 1999 cost:

$$\left(\frac{260 \times 10^6 \ \$}{4 \times 10^6 \text{ tons}}\right)\left(\frac{368}{325}\right)\left(1.025\right)^5 = \$83.27/\text{ton}$$

CEI in 1985
1985 → 1994

1994 → 1999

Cost of refrigeration

Since the brewing company already has to keep its beer cold ($\sim 34°F$), I have proceeded with my base case assuming that NH_3 refrigeration is available on site. Therefore, there is no capital investment involved with the refrigeration system. However, there is an operating cost given by Peters and Timmerhaus (p. 815) as:

Refrigeration (ammonia) to 34°F – \$2.00/ton-day
(288,000 Btu removed)

I will assume that this cost is a 1990 cost like most of the others in this reference. The appropriate inflation factor is 1.1646
Thus, the cost will be calculated in this manner:

$$\left(\text{Refrigeration duty } \frac{10^3 \text{ Btu}}{\text{hr}}\right) \cdot \frac{\$2.00}{288 \times 10^3 \text{ Btu}} \cdot 8400 \frac{\text{hr}}{\text{yr}} \cdot 1.1646$$

Economic Revision

As I stated before, the economic calculations could change when I have more information. I now have more information regarding excise taxes. Originally, I estimated excise taxes as a percentage of gross profits (preliminary economics, page 7). However, I now have information that says that excise taxes are to be assessed on a per gallon basis (Kirk-Othmer Encyclopedia of Chemical Technology, 1992). The rates are:

Federal: \$0.23 / gal up to 60000 bbl/yr.

\$0.29 / gal over 60000 bbl/yr.

State: varies

In order to get an estimate of the combined effect of federal and state excise taxes, I'll again use data from the Anheuser-Busch Annual Report.

	1994	1993
Beer sold	88.5 million bbl · 0.985	87.3 million bbl · 0.985
Excise taxes paid	\$1.6797 billion	\$1.6798 billion
Dollars per gallon excise taxes	$T = \dfrac{\$1.6797 \times 10^9 \cdot 0.985}{88.5 \times 10^6\ \text{bbl} \times 42\ \text{gal/bbl}}$ $= \$0.445/\text{gal}$	$T = \dfrac{\$1.6798\ \text{billion} \cdot 0.985}{87.3 \times 10^6\ \text{bbl} \times 42\ \text{gal/bbl}}$ $= \$0.451/\text{gal}$

Since some of the A-B beer sold was non-alcoholic (nominally 1.5%), I have multiplied the total beer sales by 98.5%.
The average excise tax I'll use will be:

$$\text{Ave} = (0.445 + 0.451)/2$$

$$\boxed{\begin{array}{ll} = \$0.448/\text{gal} & (> 60000\ \text{bbl/yr}) \\ \$0.388/\text{gal} & (< 60000\ \text{bbl/yr}) \rightarrow 6¢\ \text{less} \end{array}}$$

Kirk - Othmer 1992

ages is somewhat different in the provinces, especially as far as the retail sales are concerned (24). The per capita consumption was 82 L in 1987 and output increased from 20.8 × 10^6 hL in 1975 to 23.8 × 10^6 hL in 1988.

The United States. Although there are American breweries more than 100 years old, the brewing industry in the United States is young. With their technical and scientific expertise, American brewers are among the leaders in the brewing industry.

The United States' definition of beer is as follows. Beer shall mean beer, ale, porter, stout, and other similar fermented beverages of any name of description containing >0.5% by volume of alcohol, brewed or produced from malt, wholly or in part, or from any substitute thereof. A cereal beverage shall mean a malt beverage, either fermented or unfermented that contains, when ready for consumption, <0.5% by volume of alcohol. These latter beverages are wholly tax-free, but the former must pay federal and state taxes without regard to beer strength, but related to the size of the brewery as far as the federal tax is concerned, ie, the tax is $0.23 per U.S. gallon ($0.06/L) up to a yearly output of 60,000 barrels (2.52 × 10^6 gallons = 9.5 × 10^6 L) but $0.29 ($0.076/L) per U.S. gallon over 60,000 barrels.

Beers produced in the United States are nearly all of continental European types, although there are some ales of British style produced as well. The largest market is for the pilsner type with an original gravity of 11–12°P. They are light in taste, pale in color, rather bland, and with no pronounced hop aroma. In total production the United States brewing industry is the world's largest having an output of over 230 × 10^6 hL in 1988. The per capita consumption of about 90 L ranks below many other countries. The United States also have the world's largest breweries and five of them accounted for 95% of the beer market in 1988. Packaged beer accounted for 82% of beer sold in 1985.

Environmental Problems

Breweries must consider pollution of their effluent and also the availability of good quality incoming water; the price is going up rapidly. In the 1970s the beer:water ratio used was 1:10 or more, but today, in well managed breweries, it is 1:5–6, which is nearly impossible to improve further. The BOD level of effluent from breweries varies between 1000–2000 mg/L. Normal household sewage water has a BOD level of 600 mg/L; average BOD level exceeding 600 mg/L is charged extra. Thus every company has an economic interest in producing effluents at as low a BOD level as possible. Table 13 shows the composition of the effluent in various steps of the brewing process, and likewise the most critical points. To minimize the polluting substances in the effluent, and consequently the payment, several things can be done. Traditionally the by-products of the brewery, ie, spent grains, surplus yeast, and carbon dioxide, have been separated and used reasonably for other purposes, but this could be done with more care. By careful handling of all relevant methods, ie, beer transfers, cleaning, cleaning solutions, disinfectants, spent kieselguhr, and sludge, the BOD level can be reduced to below 1000 mg/L, and if the total water consumption has been reduced to 6 L per liter beer produced, this is probably the lower limit by traditional methods.

Innovation and Optimization

Before any optimization was attempted, a base case was completed. The economic analysis of this case showed that optimization was definitely necessary--the Net Present Worth was - $9.5 million! One reason the base case was so uneconomical was the distillation column that was used. When the base case was first being calculated, a Fenske-Underwood-Gilliland calculation was performed, the appropriate numbers were entered into the process simulator, and the result was accepted simply because it worked. While this was the correct decision at the time due to the necessity of completing a base case, the column used certainly did not perform well enough for a final design. It had excessive capital costs of $4 million and yearly operating costs of nearly $1 million.

Because the material balance of the column was fairly rigidly fixed, the main optimization variable was the reflux ratio. The original reflux ratio was 40, and it clearly needed to be lowered. By the phrase "fairly rigidly fixed," it is meant that the revelation to let some of the water go up the column to make up for the motive steam added to the flash vapor did not come until after the base case. Letting some of the water go with the overhead makes the separation easier and allows use of a lower reflux ratio. Thus, the reflux ratio was varied until the correct amount of water went up the column. The conclusion from the simulator was that a reflux ratio of 4.0 was optimum. This conclusion was verified by a McCabe-Thiele calculation. This new reflux ratio resulted in significant decreases in column capital costs due to the lower diameter required to accommodate the vapor. Also, significant savings in refrigeration and steam costs occurred. Below is a comparison of the "Revised Base Case" and the "Optimization I" columns. Full details of this optimization can be found in the Optimization I section of the calculation appendix.

Table 18.1 Comparison of Distillation Alternatives

	Capital Cost	Yearly Operating Costs
Revised Base Case	$4.04 million	$0.97 million
Optimization I	$1.28 million	$54,299

The other optimization variable for this process was capacity. As capacity increases, the Internal Rate of Return increases steadily before eventually leveling off. The Net Present Worth increases indefinitely. Both trends are shown in Figure 18.1. When the design work was begun, one capacity (238,733 bbl/year) was arbitrarily chosen as the starting point. The economic analysis showed, however, that, even after Optimization I, this capacity was inadequate, yielding only a 7.0% rate of return. Clearly, other capacities needed to be explored.

In the interest of time, rigorous calculations at different capacities were not performed. Instead, the six-tenths rule given by Peters and Timmerhaus for relating Total Capital Investment to plant capacity was used. The rule is as follows:

$$\textit{Cost of equipment a} = \textit{cost of equipment b} * \left(\frac{\textit{capacity of equipment a}}{\textit{capacity of equipment b}}\right)^{0.6}$$

Operating costs were assumed to vary in a linear manner with capacity. Using these relationships, cash flow tables for different multiples of the original capacity were constructed. The important details of the calculations are shown in Table 18.2 and Figure 18.1. For complete details of the results, see the Capacity Optimization section of the calculation appendix.

The optimum capacity (subject to the 20% upper limit on IRR) was determined by this method to be 2.573 times the base value, or 614,315 bbl/year. The economic analysis at this capacity was presented in Table 17.2. Once this value was determined, rigorous sizing and costing calculations were completed and a more formal economic analysis was undertaken. Surprisingly, both the Net Present Worth and the Internal Rate of Return were much higher than estimated by the six-tenths rule--the IRR was 98.7%! This rigorous analysis was presented in Table 17.3.

The reason that the rigorous economics are so different from the predicted economics is that the six-tenths rule is only an estimate and "should only be used in the absence of other information," according to Peters and Timmerhaus. Clearly it did not apply well to this process. Additionally, an error was made regarding operating costs. The fact that operating labor costs, which account for about one-third of the total operating costs, do not increase with capacity was neglected.

Thus, the true optimum capacity is unknown. It is greater than the Optimization I capacity of 238,733 bbl/year and less than 614,315 bbl/year. Strictly speaking, the optimum capacity is infinite--what is desirable to find in this case is the capacity corresponding to a 20% IRR.

Due to the time constraint on this problem, the exact optimum capacity remains unknown. However, a check was made to determine whether the 614,315 bbl/year figure was reasonable. Using data from the Anheuser-Busch 1994 Annual Report and the suggestion of Marchbanks that sales of non-alcoholic beers account for 1.5-2.5% of total beer sales, this capacity was calculated to be from one-fourth to one-half that of Anheuser-Busch. (For details of this calculation, see the Capacity Optimization section of the calculation appendix.) Thus, the "optimum" capacity, while violating the IRR constraint, is quite reasonable.

Note: In a normal design problem, there would be other distillation optimization variables such as pressure. In this case, though, the product quality is so important that it is desired to keep column pressure as low as possible so that temperatures will also be low and the beer will not be scorched. According to Perry and Green, the minimum practical overhead condenser pressure is 10 torr, or 0.2 psia. A slightly higher pressure of 0.5 psia was used. Thus, some limited pressure optimization was possible but was not deemed an efficient use of time.

Table 18.2 Summary of Capacity Optimization Results

Results were generated using six-tenths rule and linear operating costs

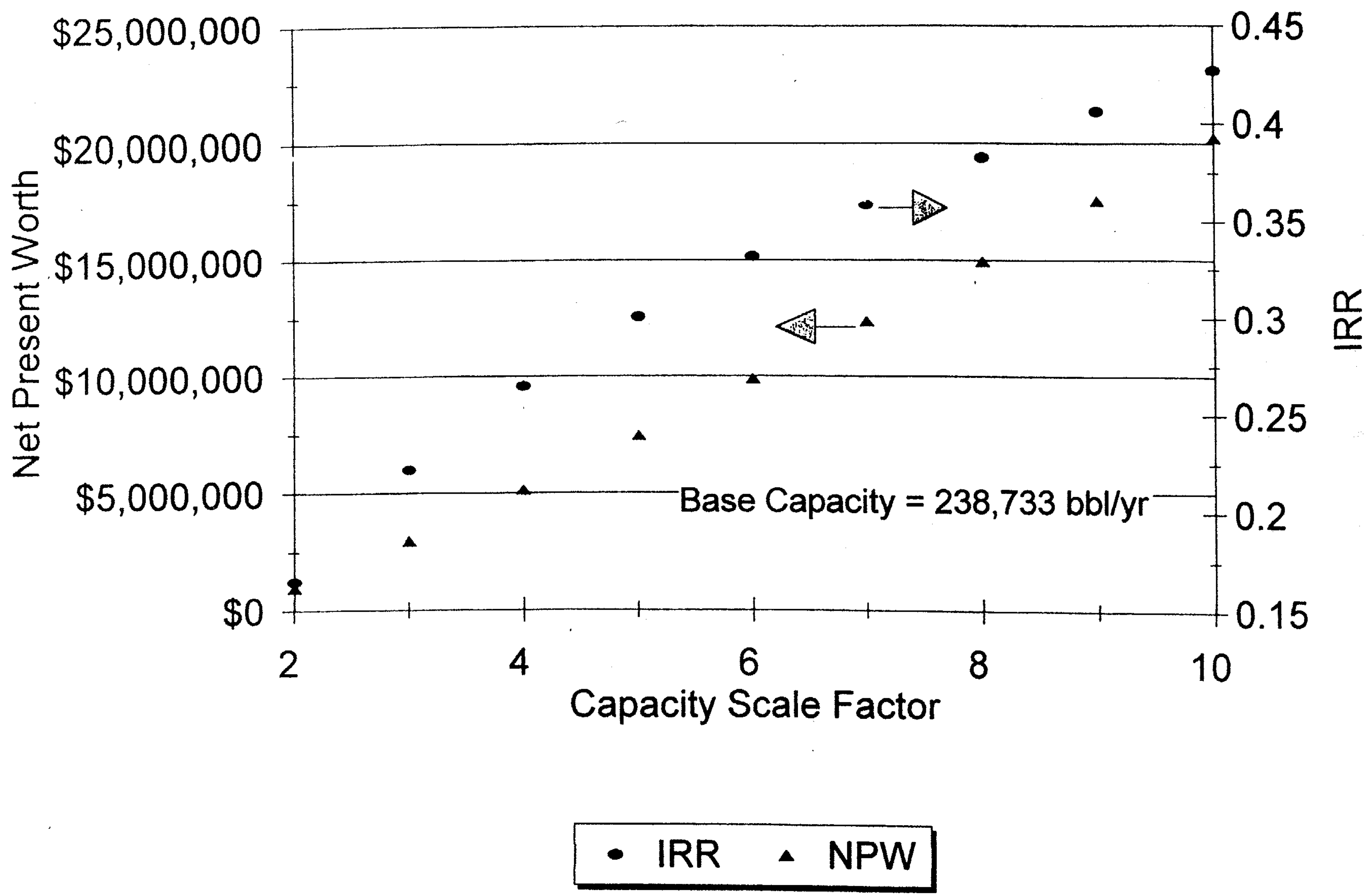
Base Capacity = 238,733 bbl/yr

Base Total Capital Investment = $3,530,048

Base Operating Costs = $2,052,869/yr

Scale Factor	IRR	NPW @ 12%
2	0.165	$990,293
3	0.222	$2,963,229
4	0.265	$5,147,449
5	0.301	$7,470,869
6	0.332	$9,895,477
7	0.359	$12,398,204
8	0.383	$14,963,752
9	0.406	$17,581,340
10	0.427	$20,243,015

Fig. 18.1 IRR and NPW vs. Capacity



Literature Cited in Report

Engan, S., 1981. "Beer Composition: Volatile Substances," Chapter 3 from *Food Science and Technology, A Series of Monographs,* Vol. 2. Academic Press.

Hardwick, W.A., 1983. "Beer," Chapter 3, *Biotechnology*, Verlag Chemie, Vol. 5.

Hirata, M., Nagahama, K., Ohe, S., 1975. *Computer Aided Data Book of Vapor-Liquid Equilibria.* El Sevier Scientific.

Hodgins, K.R., "De-alcoholization of Beer."

Kirk-Othmer Encyclopedia of Chemical Technology, 1992. John Wiley and Sons.

Marchbanks, C., 1986. "Non- and Low-alcohol Beers: How They are Made." *Brewing and Distillers Int.*

Perry, H., Green, D., 1994. *Perry's Chemical Engineers' Handbook.* McGraw-Hill, 6th ed.

Peters, M.S., Timmerhaus, K.D., 1990. *Plant Design and Economics for Chemical Engineers.* McGraw-Hill, 4th ed.

Selections from the 1994 Annual Report of Anheuser-Busch.

White, J.A., Agee, M.H., Case, K.E., 1989. *Principles of Engineering Economic Analysis.* John Wiley and Sons, 3rd ed.

Solution
1996

Calculation Appendix

This appendix includes all pertinent hand calculations, process simulator results, and other work done in completing the design.

Table of Contents

Table of Contents (cont'd)

Database

Database Development

Purpose: A design is only as good as the data on which it is founded. Some systems are extremely sensitive to the data while others are not. One never knows how sensitive the design is until one explores the data.

This problem is slightly different than other design problems in that there are no data provided in the problem statement. Thus, it is up to the designer to find data. The problem is further complicated by the fact that beer is a mixture of many compounds.

VLE Data: Among beer's components, two dominate: water and ethanol. Obviously, some sort of separation is required. Distillation is the most obvious choice, and therefore I have compiled some VLE data at different pressures for the EtOH/H_2O system. The data are taken from:

Hirata and Ohe, Computer Aided Data Book of Vapor-Liquid Equilibria, El Sevier Scientific, Tokyo 1975.

Simulator development: In order to facilitate ease of calculations, the VLE data were inputted to the process simulator ChemCad, version 3. They were then regressed* to find parameters for the TK-Wilson equation. Previous design experience and personal preference indicate that this model is the best one to use for distillation calculations.

VLE Data for other components: Beer consists of many other components, including inorganic salts, CO_2, organic acids, and esters. It is not known yet which of the other components are the most important. However, some initial data was gathered nonetheless. Acetic acid/water VLE Data was compiled.

Physical properties: In addition to good VLE data, good physical property values are obviously important to a good design. I plan on obtaining physical properties from the ChemCad process simulator, but since simulator property libraries vary in accuracy, it is prudent to have experimental values for comparison. Therefore, I have compiled some information on the physical properties of ethanol and CO_2. Additional information will be taken from Perry's Chemical Engineer's Handbook.

Special note on water: the physical properties of water are well known and will be taken from Perry's or the steam tables of Smith and Van Ness, Introduction to Chemical Engineering Thermodynamics.

* The quantity $\sum \left(\frac{x_{calc}}{x_{exp}} - 1 \right)$ was minimized

Contents of Database

The following is a list of the contents to date of my database for this problem. Additional information will probably be added as time goes by.

1. Ethanol | Water VLE data (3 pages)
2. ChemCad simulations of EtOH | H_2O VLE (7 pages)
3. Ethyl acetate | water VLE data (1 page)
4. Physical properties of ethanol (1 page)
5. Physical properties of CO_2 (1 page)
6. ChemCad VLE simulation of H_2O | pyruvic acid (1 page)

[530] ETHANOL(1) - WATER(2)

DATA FROM KIRSCHBAUM.E..F.GERSTNER.Z.VDI BEIH.VERFAHRENSTECHNIK NO.1, P.10(1939)

X_1	Y_1	Temperature [°C]	Pressure [mmHg]
0.0500	0.2710	32.00	50.00
0.1000	0.4310	28.90	50.00
0.2000	0.5590	25.75	50.00
0.3000	0.6060	24.40	50.00
0.4000	0.6420	23.70	50.00
0.5000	0.6810	23.10	50.00
0.6000	0.7250	22.65	50.00
0.7000	0.7760	22.20	50.00
0.8000	0.8390	21.90	50.00
0.9000	0.9082	21.65	50.00
0.9500	0.9529	21.60	50.00

Antoine constants

Components	A	B	C
1	8.04494	1554.300	222.650
2	8.10765	1750.286	235.000

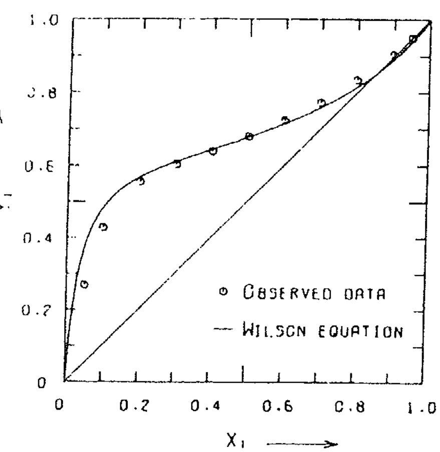


FIG. X - Y CURVE

Wilson parameters

$\Lambda_{12} = 0.18482$
$\Lambda_{21} = 0.71159$

Error* on Wilson equation

Y_1 : 0.0242
T [°C] : 0.49

* $\dfrac{\sum |Y_{1CALC} - Y_{1OBS}|}{\text{DATA POINTS}}$

$\dfrac{\sum |T_{CALC} - T_{OBS}|}{\text{DATA POINTS}}$

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[531] ETHANOL(1) - WATER(2)

DATA FROM KIRSCHBAUM.E..F.GERSTNER.Z.VDI BEIH.VERFAHRENSTECHNIK NO.1, P.10(1933).

X_1	Y_1	Temperature [°C]	Pressure [mmHg]
0.0500	0.2840	44.90	100.00
0.1000	0.4390	41.55	100.00
0.2000	0.5600	38.25	100.00
0.3000	0.6040	37.10	100.00
0.4000	0.6410	36.45	100.00
0.5000	0.6770	35.90	100.00
0.6000	0.7210	35.45	100.00
0.7000	0.7730	35.00	100.00
0.8000	0.8360	34.60	100.00
0.9000	0.9078	34.35	100.00
0.9500	0.9512	34.25	100.00

Antoine constants

Components	A	B	C
1	8.04494	1554.300	222.650
2	8.10765	1750.286	235.000

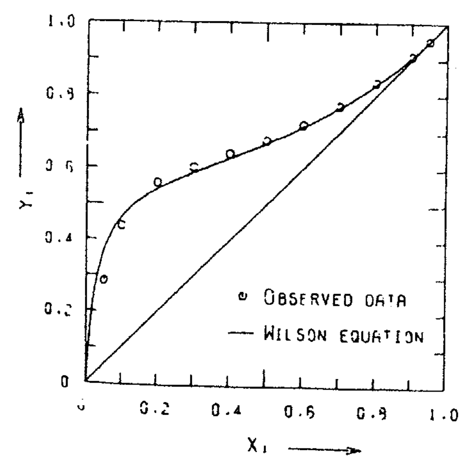

FIG. X - Y CURVE

Wilson parameters

$\Lambda_{12} = 0.15348$
$\Lambda_{21} = 0.89934$

Error* on Wilson equation

Y_1 : 0.0159
T [°C] : 0.34

* $\dfrac{\sum |Y_{1CALC} - Y_{1OBS}|}{\text{DATA POINTS}}$

$\dfrac{\sum |T_{CALC} - T_{OBS}|}{\text{DATA POINTS}}$

M.HIRATA AND S.OHE

ETHYL ALCOHOL(1)-WATER(2)

DATA FROM BEEBE A.H., COULTER K.E., LINDSAY R.A., BAKER E.M.: IND ENG CHEM.34.1501(1942)

X_1	Y_1	T [°C]	P [mmHg]	X_1	Y_1	T [°C]	P [mmHg]
0.0160	0.1460	62.00	190.00	0.9416	0.8502	49.10	190.00
0.0170	0.2755	60.00	190.00	0.8735	0.8790	49.00	190.00
0.0550	0.3656	57.20	190.00	0.8970	0.8990	49.80	190.00
[illegible]	0.4125	55.30	190.00	0.9485	0.9466	49.50	190.00
[illegible]	0.5015	52.20	190.00	0.9600	0.9590	47.60	190.00
[illegible]	0.5455	53.00	190.00	0.9719	0.9700	50.30	190.00
0.2345	0.5650	52.40	190.00	0.9812	0.9798	48.60	190.00
0.3535	0.6045	50.10	190.00				
0.4705	0.6445	49.80	190.00				
0.4970	0.6540	48.90	190.00				
[illegible]	0.6925	50.50	190.00				
0.6525	0.7260	48.50	190.00				
0.7000	0.7550	49.90	190.00				
0.7200	0.7685	48.70	190.00				
0.7895	0.9152	50.20	190.00				

ANTOINE CONSTANTS

COMPONENTS	A	B	C
1	8.04494	1554.300	222.650
2	8.10765	1750.285	235.000

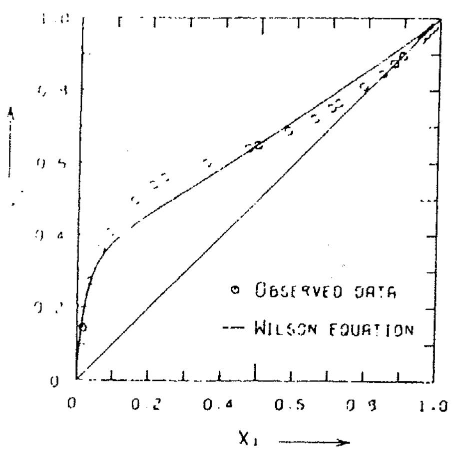

FIG X - Y CURVE

WILSON PARAMETERS

$\Lambda_{12} = 0.08103$
$\Lambda_{21} = 1.47540$

ERROR* ON WILSON EQUATION

Y_1 : 0.0307
T [°C] : 1.33

$$* \quad \frac{\sum |Y_{1CALC} - Y_{1OBS}|}{\text{DATA POINTS}} \qquad \frac{\sum |T_{CALC} - T_{OBS}|}{\text{DATA POINTS}}$$

ETHANOL(1) - WATER(2)

DATA FROM KIRSCHBAUM,E., GERSTNER: Z.VDI BEIH.VERFAHRENSTECHNIK.NO.1. P.10(1939)

X_1	Y_1	TEMPERATURE [°C]	PRESSURE [mmHg]
0.0500	0.3070	63.90	250.00
0.1000	0.4490	59.95	250.00
0.2000	0.5540	56.70	250.00
0.3000	0.5990	55.15	250.00
0.4000	0.6340	54.40	250.00
0.5000	0.6790	54.00	250.00
0.6000	0.7150	53.65	250.00
0.7000	0.7670	53.35	250.00
0.8000	0.8310	53.10	250.00
0.9000	0.9017	53.00	250.00
0.9500	0.9419	53.00	250.00

ANTOINE CONSTANTS

COMPONENTS	A	B	C
1	8.04494	1554.300	222.650
2	8.10765	1750.286	235.000

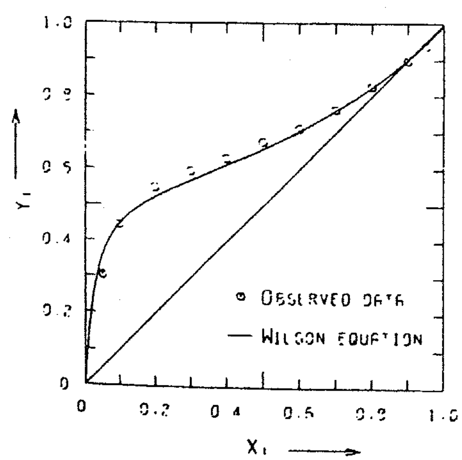

FIG X - Y CURVE

WILSON PARAMETERS

$\Lambda_{12} = 0.13044$
$\Lambda_{21} = 0.96906$

ERROR* ON WILSON EQUATION

Y_1 : 0.0155
T [°C] : 0.54

$$* \quad \frac{\sum |Y_{1CALC} - Y_{1OBS}|}{\text{DATA POINTS}} \qquad \frac{\sum |T_{CALC} - T_{OBS}|}{\text{DATA POINTS}}$$

ETHYL ALCOHOL(1)-WATER(2)

DATA FROM BEEBE A.H.,COULTER K.E.,LINDSAY R.A.,BAKER E.M.:IND.ENG.CHEM.34.1501(1942)

X_1	Y_1	T [°C]	P [mmHg]	X_1	Y_1	T [°C]	P [mmHg]
0.0160	0.1470	78.10	380.00	0.8420	0.8488	62.70	380.00
0.0315	0.2505	76.00	380.00	0.8749	0.8768	62.50	380.00
0.0600	0.3765	72.40	380.00	0.8967	0.8973	63.60	380.00
0.0855	0.4300	69.30	380.00	0.9485	0.9440	63.50	380.00
0.1465	0.5005	67.70	380.00	0.9727	0.9692	63.00	380.00
0.2060	0.5415	67.50	380.00				
0.2360	0.5600	67.10	380.00				
0.3495	0.5345	65.30	380.00				
0.4675	0.6410	64.70	380.00				
0.4875	0.6425	64.30	380.00				
0.5930	0.6890	64.40	380.00				
0.6525	0.7250	64.20	380.00				
0.7000	0.7495	63.80	380.00				
0.7175	0.7680	63.20	380.00				
0.7890	0.8111	63.90	380.00				

ANTOINE CONSTANTS

COMPONENTS	A	B	C
1	8.04494	1554.300	222.650
2	7.96681	1668.210	228.000

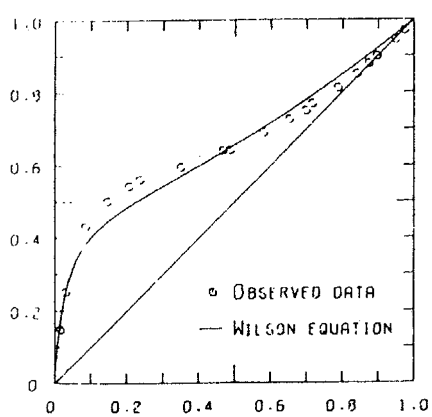

FIG   X - Y CURVE

WILSON PARAMETERS

Λ_{12} = 0.11585
Λ_{21} = 1.26068

ERROR* ON WILSON EQUATION

Y_1 : 0.0264
T [°C] : 0.84

$$* \quad \frac{\Sigma |Y_{1CALC} - Y_{1OBS}|}{\text{DATA POINTS}} \qquad \frac{\Sigma |T_{CALC} - T_{OBS}|}{\text{DATA POINTS}}$$

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ETHANOL(1) - WATER(2)

DATA FROM KIRSCHBAUM,E.,F.GERSTNER:Z.VDI BEIH.VERFAHRENSTECHNIK.NO.1, P.10(1939)

X_1	Y_1	TEMPERATURE [°C]	PRESSURE [mmHg]
0.0500	0.3180	79.60	500.00
0.1000	0.4530	75.70	500.00
0.2000	0.5430	72.55	500.00
0.3000	0.5850	71.05	500.00
0.4000	0.6210	70.25	500.00
0.5000	0.6590	69.60	500.00
0.6000	0.7060	69.05	500.00
0.7000	0.7600	68.55	500.00
0.8000	0.8240	68.15	500.00
0.9000	0.9005	68.00	500.00
0.9500	0.9462	68.00	500.00

ANTOINE CONSTANTS

COMPONENTS	A	B	C
1	8.04494	1554.300	222.650
2	7.96681	1668.210	228.000

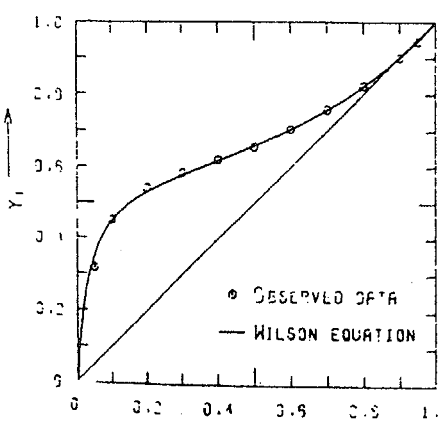


FIG   X - Y CURVE

WILSON PARAMETERS

Λ_{12} = 0.16240
Λ_{21} = 0.88292

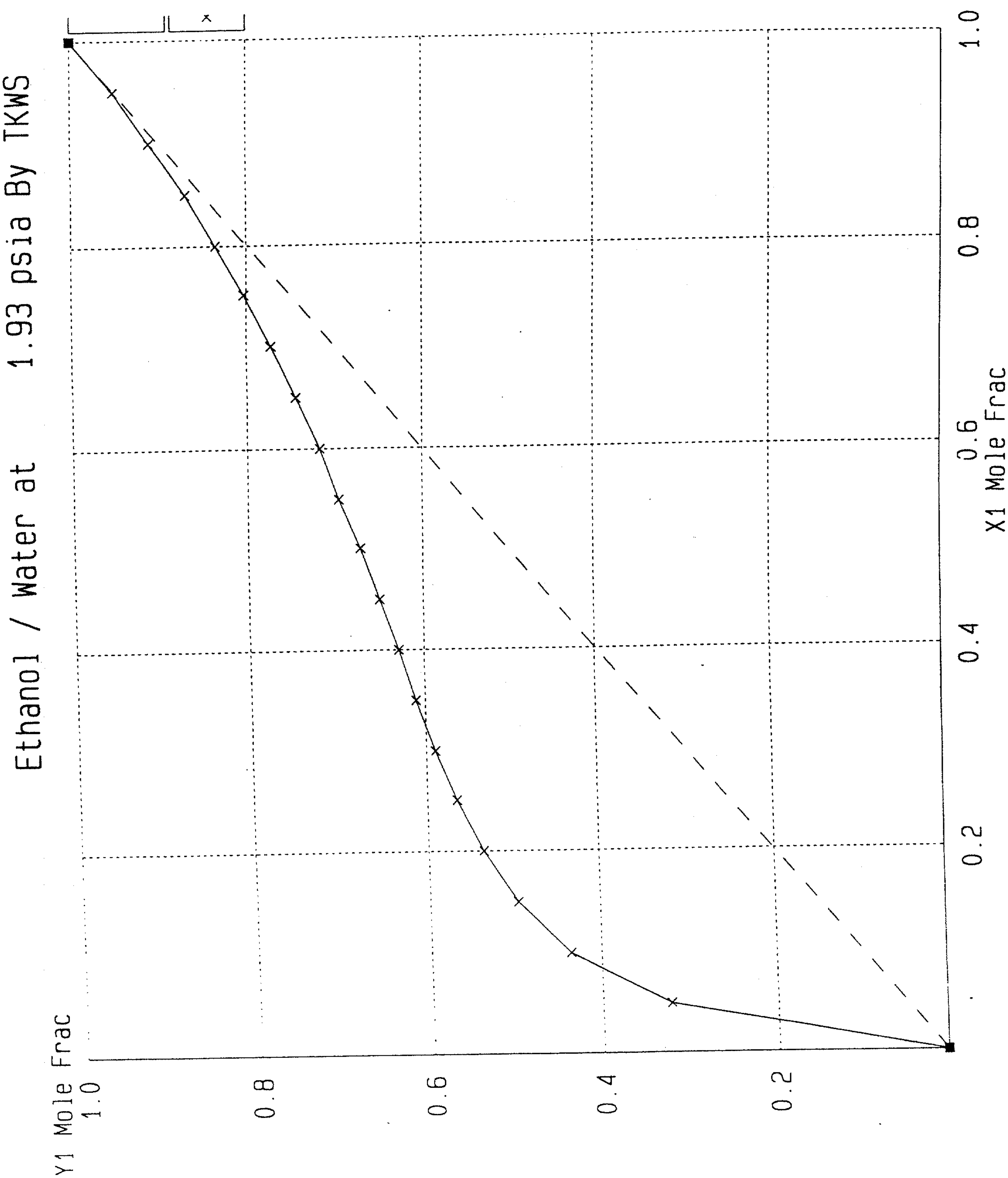
ERROR* ON WILSON EQUATION

Y_1 : 0.0056
T [°C] : 0.18

$$* \quad \frac{\Sigma |Y_{1CALC} - Y_{1OBS}|}{\text{DATA POINTS}} \qquad \frac{\Sigma |T_{CALC} - T_{OBS}|}{\text{DATA POINTS}}$$

M.HIRATA AND S.OHE

Database figure 1: EtOH/H_2O VLE simulation from ChemCad



Database Table 1: VLE Data for EtOH/H_2O regressed in ChemCad

Regression has converged in 60 iterations. (TK Wilson)

System: Ethanol / Water

Experimental Data			Deviation	
T Deg F	X1	Y1	D_T	D_Y1
89.60	0.0500	0.2710	-0.68	0.0571
84.02	0.1000	0.4310	-0.02	0.0138
78.35	0.2000	0.5590	1.30	-0.0151
75.92	0.3000	0.6060	1.59	-0.0097
74.66	0.4000	0.6420	1.39	-0.0039
73.58	0.5000	0.6810	1.30	-0.0016
72.77	0.6000	0.7250	1.11	-0.0005
71.96	0.7000	0.7760	1.08	0.0002
71.42	0.8000	0.8390	0.96	-0.0018
70.97	0.9000	0.9082	0.97	0.0022
70.88	0.9500	0.9529	0.94	-0.0001
112.82	0.0500	0.2840	-0.43	0.0365
106.79	0.1000	0.4390	0.37	-0.0035
100.85	0.2000	0.5600	1.67	-0.0260
98.78	0.3000	0.6040	1.45	-0.0170
97.61	0.4000	0.6410	1.03	-0.0109
96.62	0.5000	0.6770	0.73	-0.0040
95.81	0.6000	0.7210	0.45	-0.0010
95.00	0.7000	0.7730	0.34	0.0005
94.28	0.8000	0.8360	0.35	0.0001
93.83	0.9000	0.9078	0.32	0.0026
93.65	0.9500	0.9512	0.36	0.0018
143.60	0.0160	0.1460	0.08	0.0004
140.00	0.0370	0.2755	-1.40	-0.0120
134.96	0.0650	0.3650	-0.75	-0.0083
131.54	0.0900	0.4125	0.04	-0.0028
125.96	0.1580	0.5015	1.41	-0.0087
127.40	0.2090	0.5455	-1.87	-0.0155
126.32	0.2385	0.5650	-1.61	-0.0178
122.18	0.3535	0.6045	0.18	-0.0023
121.64	0.4705	0.6445	-1.05	0.0091
120.02	0.4970	0.6540	0.22	0.0117
122.90	0.5805	0.6925	-3.68	0.0134
119.30	0.6525	0.7260	-0.84	0.0179
121.64	0.7000	0.7550	-3.63	0.0162
119.66	0.7200	0.7685	-1.82	0.0147
122.36	0.7895	0.8152	-5.06	0.0128
120.38	0.8416	0.8502	-3.40	0.0148
120.20	0.8735	0.8790	-3.38	0.0103
121.64	0.8970	0.8990	-4.92	0.0090
121.10	0.9485	0.9466	-4.55	0.0053
117.68	0.9600	0.9580	-1.16	0.0043
122.54	0.9719	0.9700	-6.04	0.0032
119.48	0.9812	0.9798	-2.99	0.0021
Mean Deviation			1.57	0.0096
Max. Deviation			6.04	0.0571

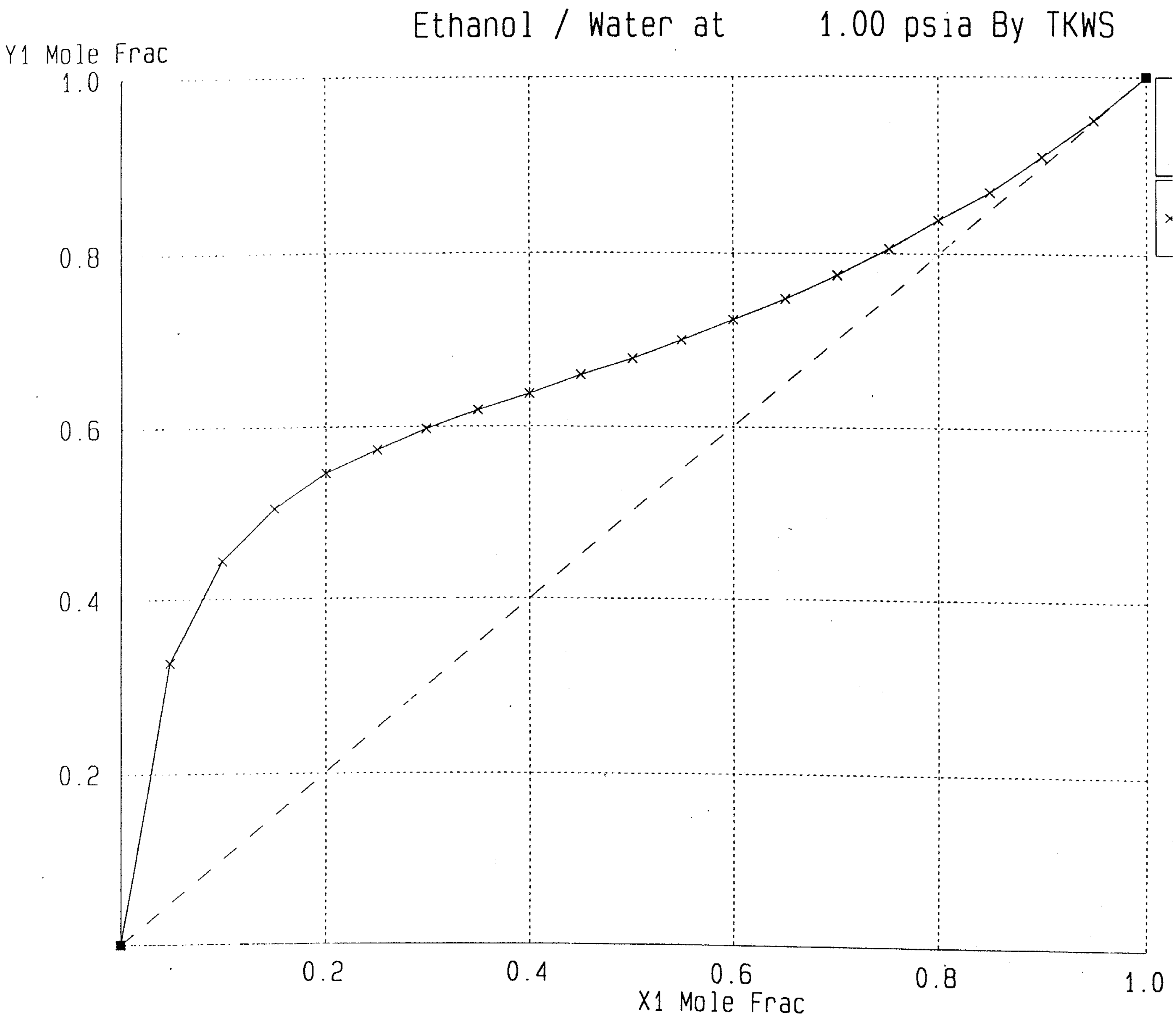
Regressed Parameters:
A12 = -3.1536e+002
A21 = 6.6232e+002

Database Table 2: Sample EtOH/H_2O output at 1.93 psia

XY data for Ethanol / Water

Wilson parameters: -315.361 662.320

T Deg F	P psia	Mole Fractions X1	Y1	Gamma1	Gamma2
124.71111	1.93420	0.00000	0.00000	5.16135	1.00000
112.36021	1.93420	0.05000	0.32143	3.79961	1.00771
107.14384	1.93420	0.10000	0.43599	2.97889	1.02777
104.31554	1.93420	0.15000	0.49572	2.44585	1.05717
102.52129	1.93420	0.20000	0.53420	2.08065	1.09414
101.23851	1.93420	0.25000	0.56301	1.82012	1.13759
100.23198	1.93420	0.30000	0.58713	1.62835	1.18681
99.38713	1.93420	0.35000	0.60905	1.48371	1.24137
98.64458	1.93420	0.40000	0.63017	1.37257	1.30098
97.97337	1.93420	0.45000	0.65133	1.28594	1.36550
97.35741	1.93420	0.50000	0.67310	1.21775	1.43484
96.78880	1.93420	0.55000	0.69589	1.16375	1.50900
96.26476	1.93420	0.60000	0.72003	1.12091	1.58800
95.78520	1.93420	0.65000	0.74584	1.08699	1.67191
95.35191	1.93420	0.70000	0.77359	1.06036	1.76081
94.96800	1.93420	0.75000	0.80358	1.03976	1.85479
94.63750	1.93420	0.80000	0.83614	1.02423	1.95395
94.36516	1.93420	0.85000	0.87161	1.01302	2.05841
94.15630	1.93420	0.90000	0.91042	1.00555	2.16824
94.01714	1.93420	0.95000	0.95303	1.00133	2.28352
93.95457	1.93420	1.00000	1.00000	1.00000	2.40430



Ethanol / Water at 1.00 psia By TKWS
Y1 Mole Frac
1.0
0.8
0.6
0.4
0.2
0.2
0.4
0.6
0.8
1.0
X1 Mole Frac

Database Table 3

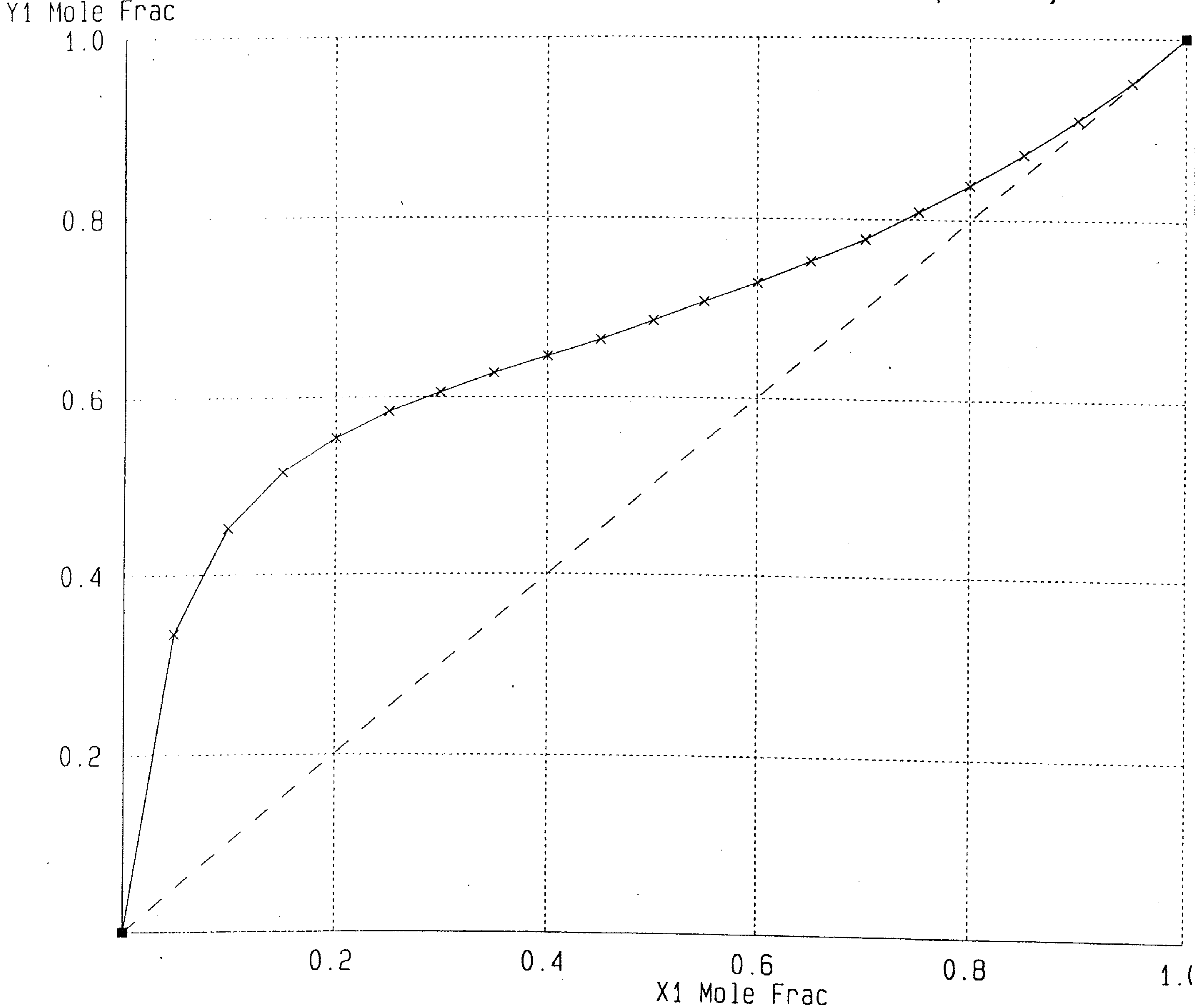
XY data for Ethanol / Water

Wilson parameters: -315.361 662.320

T Deg F	P psia	Mole Fractions X1	Y1	Gamma1	Gamma2
101.57835	1.00000	0.00000	0.00000	5.18967	1.00000
89.96910	1.00000	0.05000	0.32870	3.83855	1.00758
85.05522	1.00000	0.10000	0.44489	3.01801	1.02736
82.38829	1.00000	0.15000	0.50514	2.48150	1.05651
80.70067	1.00000	0.20000	0.54368	2.11183	1.09334
79.50084	1.00000	0.25000	0.57227	1.84680	1.13684
78.56554	1.00000	0.30000	0.59597	1.65089	1.18637
77.78502	1.00000	0.35000	0.61733	1.50257	1.24155
77.10197	1.00000	0.40000	0.63778	1.38819	1.30216
76.48589	1.00000	0.45000	0.65819	1.29874	1.36813
75.92076	1.00000	0.50000	0.67914	1.22810	1.43945
75.39879	1.00000	0.55000	0.70106	1.17198	1.51619
74.91685	1.00000	0.60000	0.72433	1.12731	1.59845
74.47484	1.00000	0.65000	0.74925	1.09184	1.68638
74.07475	1.00000	0.70000	0.77613	1.06390	1.78018
73.71971	1.00000	0.75000	0.80531	1.04220	1.88003
73.41399	1.00000	0.80000	0.83714	1.02579	1.98616
73.16270	1.00000	0.85000	0.87201	1.01390	2.09880
72.97191	1.00000	0.90000	0.91040	1.00594	2.21815
72.84850	1.00000	0.95000	0.95285	1.00143	2.34442
72.80034	1.00000	1.00000	1.00000	1.00000	2.47780

Databook figure 3 EtOH | H_2O VLE from ChemCAD

Ethanol / Water at 0.50 psia By TKWS



Database Table 4

XY data for Ethanol / Water

Wilson parameters: -315.361 662.320

T Deg F	P psia	Mole Fractions X1	Y1	Gamma1	Gamma2
79.41301	0.50000	0.00000	0.00000	5.20761	1.00000
68.50969	0.50000	0.05000	0.33578	3.87083	1.00743
63.87804	0.50000	0.10000	0.45365	3.05315	1.02691
61.35881	0.50000	0.15000	0.51450	2.51488	1.05575
59.76714	0.50000	0.20000	0.55319	2.14178	1.09236
58.64079	0.50000	0.25000	0.58163	1.87293	1.13580
57.76830	0.50000	0.30000	0.60499	1.67328	1.18552
57.04472	0.50000	0.35000	0.62585	1.52152	1.24119
56.41441	0.50000	0.40000	0.64568	1.40403	1.30268
55.84758	0.50000	0.45000	0.66536	1.31183	1.36998
55.32811	0.50000	0.50000	0.68552	1.23877	1.44316
54.84801	0.50000	0.55000	0.70660	1.18053	1.52238
54.40404	0.50000	0.60000	0.72898	1.13401	1.60786
53.99596	0.50000	0.65000	0.75300	1.09695	1.69986
53.62560	0.50000	0.70000	0.77899	1.06764	1.79867
53.29619	0.50000	0.75000	0.80731	1.04480	1.90466
53.01213	0.50000	0.80000	0.83836	1.02746	2.01817
52.77904	0.50000	0.85000	0.87257	1.01484	2.13960
52.60332	0.50000	0.90000	0.91048	1.00636	2.26935
52.49266	0.50000	0.95000	0.95270	1.00154	2.40781
52.45598	0.50000	1.00000	1.00000	1.00000	2.55536

Ethyl acetate | H_2O VLE Data (from Computer-Aided Data Book of Vapor-Liquid Equilibrium) 1/1

511 ETHYL ACETATE(1) - WATER(2)*

DATA FROM M.KOMASTU,M.HIRATA:KOUKA. VOL.72, P.1419(1969)

X_1	Y_1	TEMPERATURE [°C]	PRESSURE [mmHg]
0.0050	0.3520	40.60	150.00
0.0080	0.6160	36.30	150.00
0.0110	0.7590	32.40	150.00
0.2260	0.8700	32.00	150.00
0.8660	0.8660	31.90	150.00
0.8700	0.8700	31.90	150.00
0.8710	0.8710	32.00	150.00
0.8900	0.8900	32.80	150.00

ANTOINE CONSTANTS

COMPONENTS	A	B	C
1	7.09808	1238.710	217.000
2	8.10765	1750.286	235.000

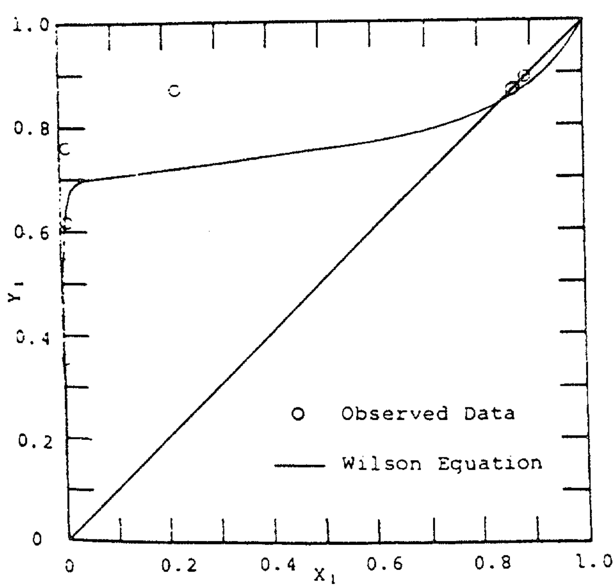


FIG X - Y CURVE

Wilson Paramerters

$\Lambda_{12} = 0.01397$

$\Lambda_{21} = 0.41167$

Error* on Wilson Equation

Y_1 : 0.0664

T[°C] : 3.14

$* \dfrac{\Sigma|Y_1calc - Y_1obs|}{\text{Data Points}}$

$\dfrac{\Sigma|Tcalc - Tobs|}{\text{Data Points}}$

M.HIRATA AND S.OHE

Physical Properties of Ethanol (- Ullmann's Encyclopedia of Industrial Chemistry) 4/1

i [64-17-5] or ethyl alcohol)H), M_r 46.7, is also referred to as rit, spirit of wine, grain alcohol, abso- l, and ethyl hydrate. Depending on its :nt, preparation, and final use, several oducts exist on the market. The 99 % ten referred to as absolute alcohol) is sively for tinctures and pharmaceu- rations, as a solvent and preservative, eptic, and in perfume. Ethanol is an functional component of alcoholic which are produced by fermentation ible carbohydrates. The fermentation may constitute (after processing and everage, e.g., in the case of beer or ie alcohol can be concentrated from to produce high-alcohol-containing he alcohol is used for purpo es other beverage, it is denatured by the addi- ostances such as methanol, pyridine, yde, or sublimate. The denatured al- ien used by industry and commerce, as a solvent, as a raw material for ring chemicals, or as a fuel.

ally synthesized ethanol is usually m petroleum sources by the hydration : (see Chap. 3) and has found wide as industrial alcohol. Various routes of ethanol are depicted in

an be considered to be one of the oducts. Fermented beer was con- b, nd wine was produced as early as : distillation process probably orginated in the turies. At this time, the "spiritual" effect of recognized: thus, the name *spiritus* was given drinks. The first wine distillates were used for ther than beverage purposes.

Until the 17th century, alcoholic fermentation was considered to be a spoilage process whereby the yeast produced was eliminated. The nature of fermentation was initially clarified in the 19th century with the discovery of the microscope, which showed that yeast cells were living organisms. However, recognition of the fact that these living organisms are responsible for the fermentation process took about 150 years.

In the 19th century, two theories were developed to explain the mechanism of fermentation the "mechanistic" and the "vital" processes. LOUIS PASTEUR (1822–1895) promoted the vital theory, which stated that living organisms were responsible for the conversion of sugar to alcohol. The mechanistic theory was supported by JUSTUS FREIHERR VON LIEBIG (1803–1873) and by FRIEDRICH WÖHLER (1800–1882). A convincing proof of the mechanistic mechanism, by which physicochemical processes lead to chemical conversion of sugar to ethanol, came from EDWARD BUCHNER (1860–1917), who demonstrated that alcoholic fermentation is related not to the living cell but to a substance in the fermentation broth, which was later identified as an enzyme. As is now known, enzymes are ultimately responsible for the complex conversion of carbohydrates to ethanol.

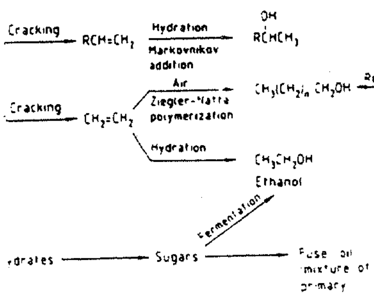


Figure 1. Industrial sources of ethanol

1. Physical Properties

Ethanol in its pure form (absolute alcohol) is a colorless liquid. It is miscible in all proportions with water and also with ether, acetone, benzene, and some other organic solvents. Anhydrous alcohol is hygroscopic: at a water uptake of 0.3–0.4 %, a certain stability does occur. Various physical properties of anhydrous ethanol are as follows:

bp	78 39 °C
fp	−114 15 °C
n^{20}	1 36048
d_4^{15}	0.79356
d_4^{20}	0 78942
d_{15}^{15}	0 79425
d_{20}^{20}	0.79044

Surface tension at 20 °C	22.03 mN m
C_p (16–21 °C)	2.415 J g^{-1} K^{-1}
Heat of fusion	4.64 kJ mol
Heat of evaporation	
At 70 °C	855.66 kJ/kg
At 80 °C	900.83 kJ/kg
At 100 °C	799.05 kJ/kg
Heat of combustion (at constant volume)	1370.82 kJ/mol
Thermal conductivity at 20 °C	18 μW m^{-1} K^{-1}
Dynamic viscosity η	1.19 mPa s
Volumetric expansion coefficient	1.1 × 10^{-3} K^{-1} *
Heat of mixing 30 wt % ethanol and 70 wt % water at 17.33 °C	39.32 J/g
Flash point (in a closed vessel)	13 °C
Autoignition temperature	425 °C
Explosion limit (amount of ethanol in a mixture with air)	
Lower, 3.5 vol %	67 g/m^3
Upper, 15 vol %	290 g/m^3
Maximum explosion pressure	736 kN m^2
Specific conductivity	135 × 10^{-11} Ω^{-1} cm^{-1}
Dilution number at 20 °C (diethyl ether = 1)	8 2
Diffusion coefficient for vapors at 20 °C and 101 3 kPa	0.12 cm^{-1}
Heating value	
Upper	29 895 kJ/kg
Lower	29 964 kJ/kg

* In practice, the volume increase for 1000 L is taken as : L K

Tables 1 and 2 show the freezing and flash points of ethanol–water mixtures. Table 3 shows the vapor–liquid equilibria of ethanol–water mixtures; the azeotropic mixture contains 95.57 wt % ethanol and 4.43 wt % water. Therefore, the highest concentration of ethanol, obtained by distillation from an ethanol–water mixture, is 95.57 wt%. Azeotropic distillation, with the help of a tertiary solvent (e.g., benzene), must be introduced to produce absolute (anhydrous) ethanol.

When ethanol is mixed with water, the volume contracts slightly, as shown in Table 4. For example, when 52 volumes of absolute ethanol and 48 volumes of water are mixed, 96.3 volumes of diluted ethanol result.

Table 1. Freezing points of ethanol–water mixtures

Ethanol, vol %	fp, °C	Ethanol, vol %	fp, °C
50	−36 9	30	−15.3
45	−28.1	25	−11 3
40	−24	20	−7 6
38	−22.3	15	−5 1
[illegible]	−16 8	10	−3

Table 2. Flash points of ethanol–water mixtures

Ethanol, wt %	Flash point, °C	Ethanol, wt %	Flash point, °C
100	13	50	24.5
94.5	16	40	26.5
80	19.5	30	30
70	21.5	10	46
60	22.6	5.5	56

Table 3. Vapor–liquid equilibria and boiling points of the ethanol–water system at 101.3 kPa

Ethanol in liquid, wt %	Ethanol in vapor, wt %	bp, °C
0	0.00	100.00
1	6.5	98.90
3	20.5	96.75
5	38.0	94.95
10	52.0	91 45
15	59.5	88.95
20	64.8	87.15
25	68 6	85.75
30	71 4	84.65
35	73.3	83.75
40	74.7	83.10
45	75.9	82.45
50	77.1	81.90
55	78.2	81 45
60	79 4	81.00
65	80 7	80.60
70	82.2	80.20
75	83.9	79.80
80	85.9	79.35
85	88.3	78.95
90	91.3	78 50
95	95.04	78.15
97	96.86	78.20
99	98.93	78.25
100	100.00	78.30

Table 4. Volume contraction for 1 mol of ethanol–water mixture at 15 °C

$x^{*}_{C_2H_5OH}$	d_4^{15}	ΔV^{**}
1 00000	0.79354	0.0000
0.97483	0.79666	−0.1063
0.92674	0.80270	−0.2903
0.88143	0.80851	−0.4429
0.81817	0.81688	−0.6250
0.74155	0.82756	−0.8028
0.65639	0.84032	−0.9495
0.56708	0.85510	−1.0540
0 47734	0.87186	−1.1123
0.35078	0.89980	−1 1116
0.22077	0.93482	−0.9695
0.15550	0.95355	−0.7864
0.07909	0.97311	−0 4030
0.05987	0.97790	−0.2869
0 03726	0 98435	−0.1586
0.01605	0 99196	−0 0582
0.00792	0.99542	−0.0267
0 00000	0.99913	0.0000

* Mole fraction of ethanol. ** Change in volume.

Physical Properties of CO_2 (from Kirk-Othmer Encyclopedia of Chemical Technology)

TABLE 1 Physical Constants of Carbon Dioxide

Molecular weight	44.01
Density of the gas at 60°F and 14.696 lb/in²abs	0.1166 lb/ft³
Critical temperature	88°F (31.0°C)
Critical pressure	1073 lb/in²abs (73.01 atm)
Specific heat:	
Gas at 60°F and 14.696 lb/in²abs; constant pressure	0.201 Btu/lb · °F
Gas at 60°F; constant volume	0.1546 Btu/lb · °F
Ratio of specific heat at constant pressure to specific heat at constant volume, 60°F	1.30
Specific gravity (air = 1.0)	1.528
Triple point (solid, liquid, and gas co-exist)	75.1 lb/in²abs and −70°F (5.11 atm and −56.6°C)
Atmospheric sublimation point	−109°F (78.5°C)
Thermal conductivity:	
at −58°F (−50°C)	0.0064 Btu/hr · ft² · °F/ft
at 32°F (0°C)	0.0084 Btu/hr · ft² · °F/ft
at 212°F (100°C)	0.0128 Btu/hr · ft² · °F/ft
Viscosity, gas:	
at 0°F (−17.8°C)	0.013 cP
at 100°F (37.8°C)	0.0155 cP
at 200°F (93.3°C)	0.018 cP
Latent heat of vaporization:	
at triple point (−70°F; −56.6°C)	149.7 Btu/lb
at 0°F (−17.8°C)	120.1 Btu/lb
at 32°F (0°C)	100.9 Btu/lb

Physical Constants

The physical constants of carbon dioxide are listed in Table 1.

Occurrence in Nature

Carbon dioxide is present in the atmosphere at an average concentration of 300 ppm, vol. (0.03 vol.%). CO_2 is formed by the respiration of all forms of animal life and consumed by plant life in the formation of carbohydrates by photosynthesis. Carbon dioxide occurs in the gases formed by the combustion of wood, coal, oil, and gas. CO_2 is a constituent of the products of the decay of organic matter, and occurs, along with methane (and, frequently, hydrogen sulfide), in "marsh gas" and "sewer gas." CO_2 also occurs, along with methane, in coal seams. Much natural gas contains CO_2. Some petroleum contains dissolved CO_2. Water from certain springs and artesian wells is saturated or supersaturated with CO_2. Some deep gas wells have produced almost pure carbon dioxide which, in some cases, has contained as little as 5 ppm of impurities.

XY data for Water / Pyruvic Acid (from ChemCad, SRK)

alue model: SRK

T Deg F	P psia	Mole Fractions X1	Y1
323.76987	13.22600	0.00000	0.00000
270.16946	13.22600	0.10000	0.68503
240.72307	13.22600	0.20000	0.85565
223.48613	13.22600	0.30000	0.91582
212.99431	13.22600	0.40000	0.94263
206.87187	13.22600	0.50000	0.95590
203.99876	13.22600	0.60000	0.96198
203.76707	13.22600	0.70000	0.96242
205.57579	13.22600	0.80000	0.95478
207.72447	13.22600	0.90000	0.92607
212.16057	13.22600	1.00000	1.00000

Thermodynamic Consistency

A thermodynamic consistency check of the binary EtOH | H_2O data inputted into Chemcad was conducted using the method of Graehling. The next page shows the data used to make the plot on $\ln \gamma_1/\gamma_2$ vs. x_1.

After plotting, the data was fit to the following 5^{th} order polynomial using the program Tablecurve.

$$\ln \gamma_1/\gamma_2 = 1.4531 - 4.2454\,x_1 + 1.5506\,x_1^2 + 1.1198\,x_1^3 - 0.7091\,x_1^4 - 0.1386\,x_1^5$$

Solving on the computer for x_1 when $\ln \gamma_1/\gamma_2 = 0$, I get

$$x_1 = 0.421 \qquad \text{when} \qquad \ln \gamma_1/\gamma_2 = 0$$

I'll let A = area above x-axis
B = area below x-axis

$$A = \int_0^{0.421} \ln \gamma_1/\gamma_2 \, dx$$

$$B = \int_{1.0}^{0.421} \ln \gamma_1/\gamma_2 \, dx$$

The indefinite integral is evaluated first:

$$\int \ln \gamma_1/\gamma_2 \, dx = 1.4531\,x - \frac{4.2454}{2}x^2 + \frac{1.5506\,x^3}{3} + \frac{1.1198}{4}x^4 - \frac{0.7091\,x^5}{5} - \frac{0.1386\,x^6}{6}$$

Evaluating the areas,

$$A = 0.2809$$

$$B = 0.3186$$

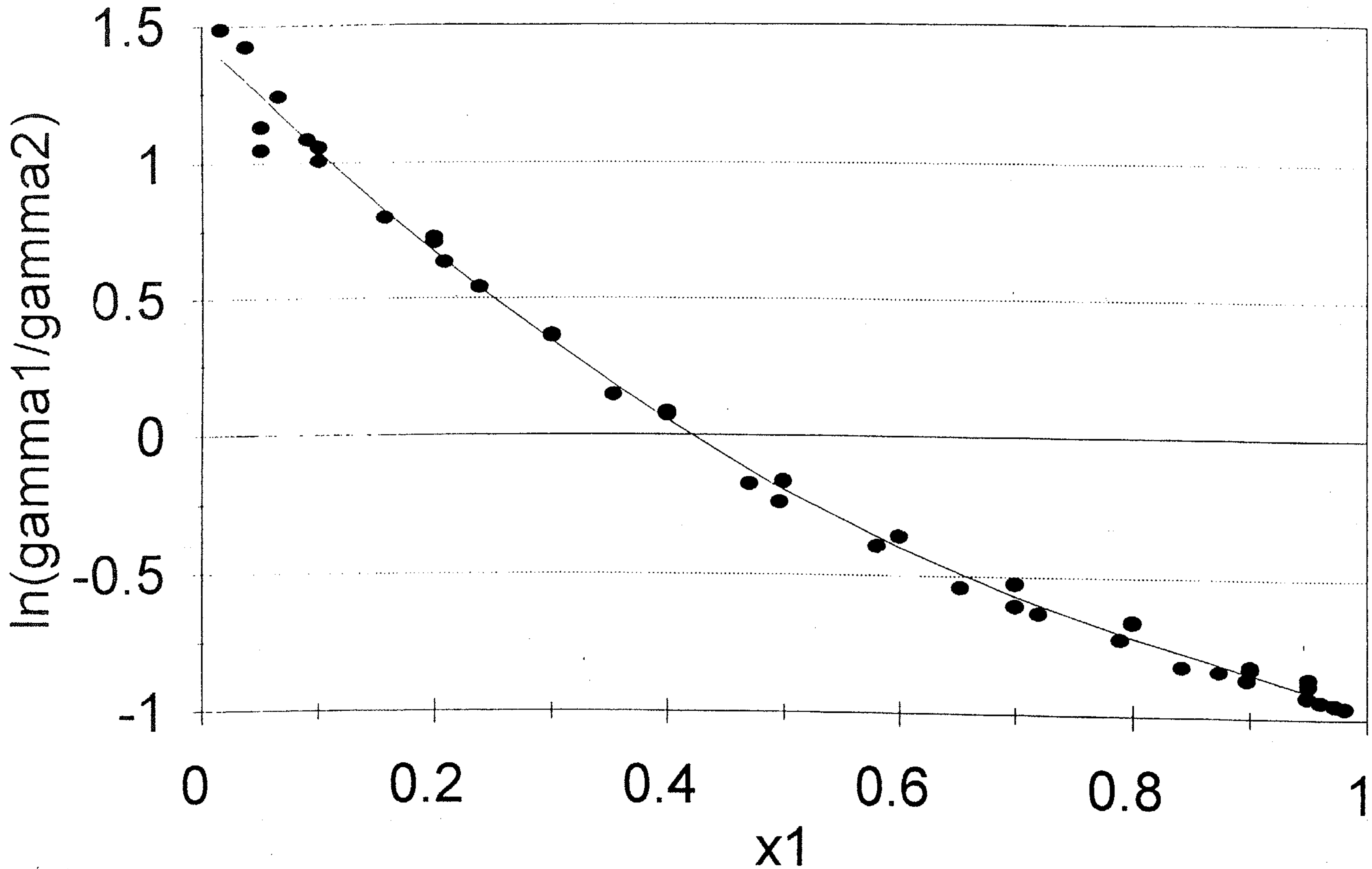
Now:

$$D = 100\left[\frac{0.2809 - 0.3186}{0.2809 + 0.3186}\right] = -6.3\%$$

EtOH/H20 Data of Hirata and Ohe (first four columns)

T (deg.F)	x1	y1	P (psia)	Antoine P1sat (psia)	Antoine P2sat (psia)	$\gamma 1$	$\gamma 2$	ln of $\gamma 1/\gamma 2$	Correlation
143.6	0.016	0.146	3.674	7.485	3.152	4.479	1.012	1.488	1.386
140	0.037	0.2755	3.674	6.852	2.877	3.993	0.961	1.424	1.298
112.82	0.05	0.284	1.934	3.368	1.382	3.261	1.055	1.129	1.245
89.6	0.05	0.271	0.967	1.715	0.692	3.056	1.073	1.047	1.245
134.96	0.065	0.365	3.674	6.042	2.525	3.415	0.988	1.240	1.184
131.54	0.09	0.4125	3.674	5.539	2.308	3.040	1.027	1.085	1.084
84.02	0.1	0.431	0.967	1.443	0.580	2.888	1.054	1.008	1.045
106.79	0.1	0.439	1.934	2.845	1.162	2.984	1.038	1.056	1.045
125.96	0.158	0.5015	3.674	4.795	1.989	2.432	1.094	0.799	0.825
78.35	0.2	0.559	0.967	1.205	0.482	2.242	1.105	0.708	0.674
100.85	0.2	0.56	1.934	2.398	0.975	2.257	1.091	0.727	0.674
127.4	0.209	0.5455	3.674	4.978	2.067	1.926	1.021	0.635	0.642
126.32	0.2385	0.565	3.674	4.840	2.008	1.798	1.045	0.543	0.542
98.78	0.3	0.604	1.934	2.258	0.916	1.724	1.194	0.368	0.343
75.92	0.3	0.606	0.967	1.114	0.445	1.753	1.222	0.361	0.343
122.18	0.3535	0.6045	3.674	4.340	1.794	1.448	1.253	0.145	0.184
74.66	0.4	0.642	0.967	1.069	0.427	1.451	1.351	0.072	0.055
97.61	0.4	0.641	1.934	2.181	0.885	1.421	1.308	0.083	0.055
121.64	0.4705	0.6445	3.674	4.278	1.768	1.176	1.395	-0.171	-0.122
120.02	0.497	0.654	3.674	4.097	1.691	1.180	1.495	-0.236	-0.184
96.62	0.5	0.677	1.934	2.118	0.859	1.236	1.455	-0.163	-0.191
73.58	0.5	0.681	0.967	1.032	0.412	1.276	1.497	-0.160	-0.191
122.9	0.5805	0.6925	3.674	4.424	1.830	0.991	1.472	-0.396	-0.359
95.81	0.6	0.721	1.934	2.068	0.838	1.124	1.610	-0.360	-0.397
72.77	0.6	0.725	0.967	1.005	0.401	1.163	1.658	-0.355	-0.397
119.3	0.6525	0.726	3.674	4.018	1.658	1.017	1.748	-0.541	-0.491
71.96	0.7	0.776	0.967	0.978	0.390	1.096	1.850	-0.524	-0.568
121.64	0.7	0.755	3.674	4.278	1.768	0.926	1.697	-0.605	-0.568
95	0.7	0.773	1.934	2.019	0.817	1.058	1.790	-0.526	-0.568
119.66	0.72	0.7685	3.674	4.058	1.674	0.966	1.814	-0.630	-0.599
122.36	0.7895	0.8152	3.674	4.361	1.803	0.870	1.789	-0.721	-0.699
94.28	0.8	0.836	1.934	1.976	0.799	1.023	1.984	-0.662	-0.713
71.42	0.8	0.839	0.967	0.961	0.383	1.055	2.032	-0.655	-0.713
120.38	0.8416	0.8502	3.674	4.137	1.708	0.897	2.035	-0.819	-0.768
120.2	0.8735	0.879	3.674	4.117	1.699	0.898	2.068	-0.834	-0.809
121.64	0.897	0.899	3.674	4.278	1.768	0.861	2.038	-0.862	-0.839
93.83	0.9	0.9078	1.934	1.949	0.788	1.001	2.261	-0.815	-0.843
70.97	0.9	0.9082	0.967	0.947	0.377	1.031	2.352	-0.825	-0.843
121.1	0.9485	0.9466	3.674	4.217	1.742	0.869	2.187	-0.922	-0.903
70.88	0.95	0.9529	0.967	0.944	0.376	1.028	2.421	-0.857	-0.905
93.65	0.95	0.9512	1.934	1.939	0.784	0.999	2.407	-0.880	-0.905
117.68	0.96	0.958	3.674	3.847	1.585	0.953	2.435	-0.938	-0.918
122.54	0.9719	0.97	3.674	4.382	1.812	0.837	2.164	-0.950	-0.933
119.48	0.9812	0.9798	3.674	4.038	1.666	0.909	2.370	-0.959	-0.945

Thermodynamic Consistency Check



Conclusions

By the criterion of Gmehling ($\leq 2\%$), these data are thermodynamically inconsistent. However, the inconsistency is not great and may simply be due to error in the correlation. Unfortunately, I did not think to do this check until today, the 16th day of the problem. Due to time constraints, I have to keep moving. If there is time at the end, I'll try to find some more consistent data. Somehow, though, my engineering intuition tells me that it won't make too much of a difference either way in the grand scheme of the problem. Or so I hope.

Revised Conclusions

It has just occurred to me (two days later) that when I calculated the activity coefficients to do this check, I was making a major assumption. I calculated the activity coefficients by the equation

$$y_i P = x_i \gamma_i P_i^{sat}$$

This neglects both the fugacity coefficients and the Poynting correction factor! The equation I should have used is:

$$\frac{\hat{\phi}_i}{\phi_i^{sat}} y_i P = x_i \gamma_i P_i^{sat} \exp\left[\frac{V_i (P - P_i^{sat})}{RT}\right].$$

Using this equation would make a slight difference in the calculation and might cause the data to be deemed thermodynamically consistent.

In any case the data are satisfactory enough for this initial design stage.

Characterization of Beer Mixture

Composition of Beer Mixture

Purpose: Beer consists of over 750 known compounds. Obviously an analysis based on all 750 would be nearly impossible. Luckily, two components (H_2O + EtOH) comprise 94% of beer by mass. Hopefully, the other 6% can be adequately characterized. I shall base my analysis on the composition of beer given in Table 4, reference 11, page R-33:

Water	90%
EtOH	4%
Carbohydrate	4%
Inorganic salts	0.8%
Nitrogen compounds	0.3%
Organic acids	0.2%
CO_2	0.5%
"Other compounds"	0.2%

} mass percentages

Clearly, the H_2O, EtOH, and CO_2 are easy to handle. The other, less specific terms are trickier. Let me start with the carbohydrates. These compounds, which consist mainly of glucose and its polymers, are not at all volatile and would tend to go down in a distillation column. Furthermore, they do not exert great influence over the properties of a solvent in which they are dissolved. Also, there is enough water present here to mask any effects. I will illustrate with an example:

The elevation of boiling point by colligative properties of a solution can be approximated by:

$$\Delta T_b \approx \frac{R T_{b0}^2}{\Delta H_v} x$$

For water at 1 psia (a likely distillation condition),

$$T_{b0} = 102^\circ F$$

$$\Delta H_v = 1035.9 \text{ Btu/lbm} \cdot \frac{18.02 \text{ lbm}}{\text{lbmol}} = 18667 \frac{\text{Btu}}{\text{lbmole}}$$

To calculate x, I will assume for a moment that the solution consists solely of 90 parts water and 4 parts monomeric glucose:

$$x = \frac{4 \text{ g glucose} \cdot \frac{1 \text{ mol glucose}}{180.1 \text{ g glucose}}}{90 \text{ g } H_2O \cdot \frac{1 \text{ mol}}{18.02 \text{ g}} + 4 \text{ g} \cdot \frac{1 \text{ mol}}{180.1 \text{ g}}} = 0.0044 \text{ mole fraction}$$

Beer composition [cont'd]

Substituting into the equation:

$$\Delta T_b \approx \frac{1.986 \frac{\text{Btu}}{\text{lbmol}\,^\circ R}\left(102 + 459.67\,^\circ R\right)^2}{18667\ \text{Btu}/\text{lbmole}} \cdot 0.0044$$

$$\approx 0.15\,^\circ R$$

** → Thus, this much carbohydrate changes the boiling point of the water by less than two-tenths of a degree. I believe it is safe to assume, then, that as long as the carbohydrate remains dissolved it will have no great effect on the properties of the solution and can be safely ignored. It will go down in any distillation column. It will of course be included in material balances, but not in VLE calculations.

→ I will apply the same argument to the inorganic salts, which are present in much smaller quantities (~0.8 wt% vs. 4 wt%).

→ The nitrogen compounds, which comprise only 0.3 wt% of the beer, fall into the same category, since they are mostly polypeptides (reference 11, p. R-34).

The only remaining components are "organic acids" and "other compounds." Reference 12 lists many of the organic compounds found in beer. Of the organic acids, the most abundant is pyruvic acid (10-220 ppm). Of the "other compounds" the most abundant is ethyl acetate (13-48 ppm). Therefore, for the sake of simplicity, I shall approximate the components of beer as:

0.2% organic acids = all pyruvic acid

0.2% "other compounds" = all ethyl acetate.

Beer composition (cont'd) — Summary

Summary of assumed beer composition:

call numbers for process simulation:

1. 90% H_2O
2. 4% EtOH
3. 0.5% CO_2

   4% Carbohydrates
   0.8% Inorganic salts
   0.3% Nitrogen compounds
   } largely ignored, except for mass purposes

4. 0.2% Organic acids = all pyruvic acid
5. 0.2% "other compounds" = all ethyl acetate

Now that this composition is established, I can proceed with the problem.

Converting to a H_2O | EtOH | CO_2 | Pyruvic acid | ethyl acetate basis:

$$\text{wt frac } H_2O = \frac{90}{90+4+0.5+0.2+0.2} = 0.9484$$

wt frac EtOH = 4/94.9 = 0.0421

wt frac CO_2 = 0.5/94.9 = 0.0053

wt frac pyruvic acid = 0.2/94.9 = 0.0021

wt frac ethyl acetate = 0.2/94.9 = 0.0021

DENSITY

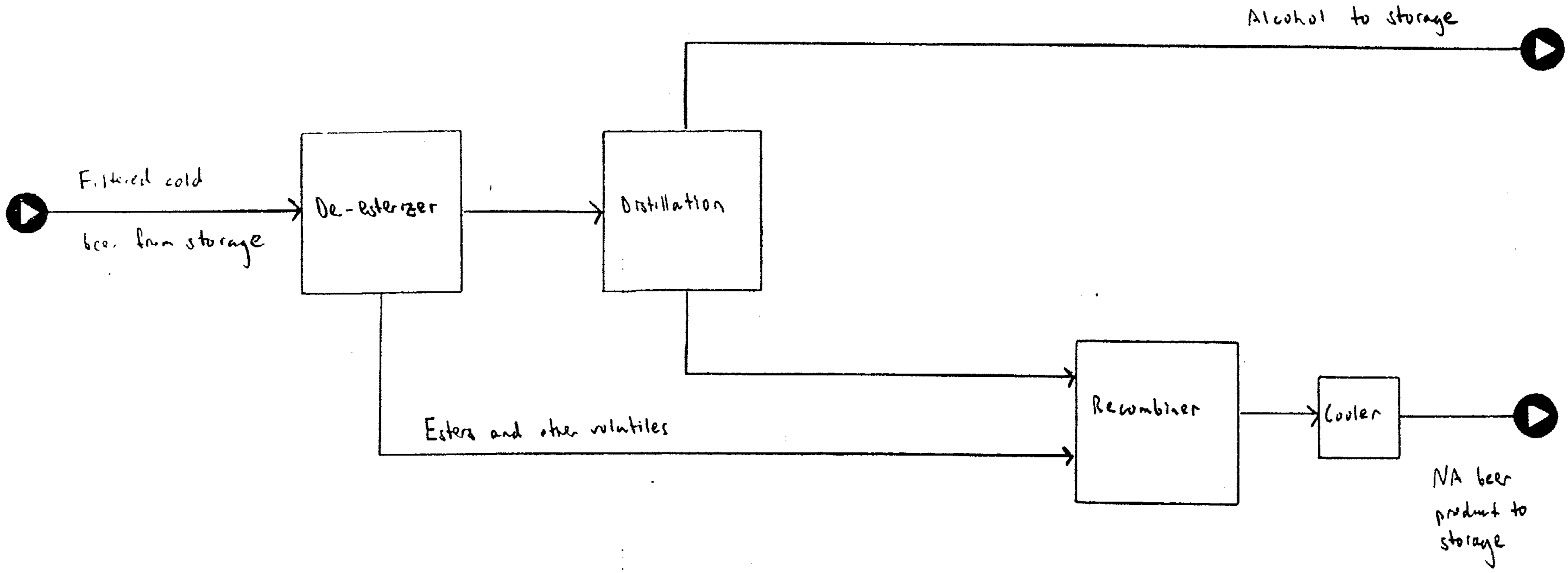
From ChemCad, the density of a mixture with this composition at 70°F is:

61.5 lbm/ft³

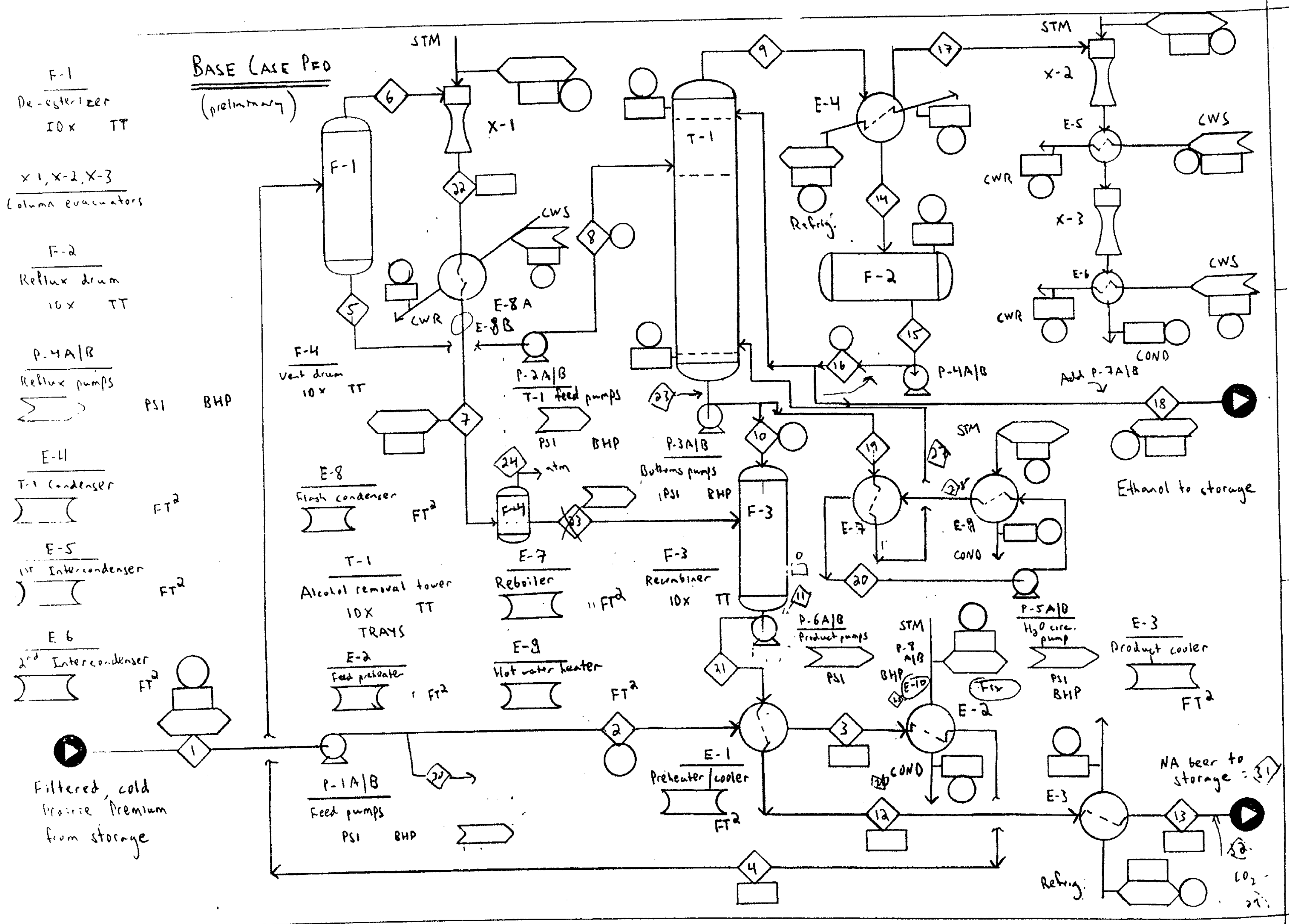
(For future reference)

Initial Simulations and Calculations

Block Flow Diagram



Base Case PFD (preliminary)



Preliminary Simulations / Calculations

I will use ChemCad to perform some of the complicated calculations involved in this design, particularly those concerning distillation columns.

- Column 1: I anticipate using a small column with only a few trays to flash off the CO_2 and organic volatiles before the beer enters a larger column to remove the alcohol (see preliminary flowsheet). So, I'll run a few simulations in order to see what conditions work best.

The Patent article mentions 3" vacuum and a "de-esterizer vessel" (page R-24). To start out, I'll try these conditions. I'll use the components listed on the beer composition page.

... First, though, I'll need data for the phase equilibria between H_2O, EtOH, CO_2, pyruvic acid, and ethyl acetate. I'd like to use the TK Wilson equation throughout the simulation, so I'll generate some data using an equation of state and then regress binary interaction parameters. It would be best to take the data at one pressure, so I'll try the 3" of vacuum.

$$\left(29.92 \text{ in Hg} - 3 \text{ in Hg}\right) \cdot \frac{14.7 \text{ psia}}{29.92 \text{ in Hg}} = 13.226 \text{ psia}$$

The generated data appears on the following pages. I've used the SRK equation. The CO_2 is not a liquid at this condition, so I won't bother with it. I'll have the following pairs:

~~H_2O | Pyruvic acid~~
~~H_2O | Ethyl acetate~~
~~EtOH | Pyruvic acid~~
~~EtOH | Ethyl acetate~~
~~Ethyl acetate | Pyruvic acid~~

Nevermind. The simulator is causing me way too much stress. I'll use SRK for the first column, Wilson for the second.

Some results.

I have simulated a flash vessel on ChemCad. The idea is to remove the ethyl acetate, CO_2, and pyruvic acid but no water or ethanol.

First, I tried flashing to 13.226 psia (3" vacuum) with an inlet temperature of 80° F.

Feed beer at 80°F, flashed from 14.7 to 13.2 psia

Stream No.	1	2	3
Stream Name	Feed	Flash Vapor	Flash Liquid
Temp F	80.0000*	79.7601	79.7601
Pres psia	14.7000*	13.2260	13.2260
Enth MMBtu/h	-0.660356	-0.00351294	-0.657011
Vapor mole fraction	0.00405658	0.584882	0.00000
Total lbmol/h	5.3727	0.0231	5.3496
Total lb/h	100.0000	1.1202	98.8798
Total std L ft3/hr	1.6222	0.0216	1.6006
Total std V scfh	2038.82	8.75	2030.08
Flowrates in lb/h			
Water	94.8400	0.0114	94.8286
Ethanol	4.2100	0.3778	3.8322
Carbon Dioxide	0.5300	0.5204	0.0096
Pyruvic Acid	0.2100	0.0006	0.2094
Ethyl Acetate	0.2100	0.2100	0.0000

As can be seen above, almost all of the CO_2 went out with the vapor, as did all of the ester. Hardly any of the pyruvic acid or water went with the vapor. A small amount of EtOH went over-head, unfortunately.

In order not to have as much EtOH go overhead, I decided to try flashing only down to 14 psia. The results are seen on the next page and are very similar to the first run. Almost all of the CO_2 and ester went overhead and only a small amount of H_2O and pyruvic acid did. A slightly lower amount of EtOH went up.

Feed beer at 80°F, flashed from 14.7 to 14 psia

Stream No.	1	2	3
Stream Name	Feed	Flash Vapor	Flash Liquid
Temp F	80.0000*	79.8937	79.8937
Pres psia	14.7000*	14.0000	14.0000
Enth MMBtu/h	-0.660356	-0.00327349	-0.657083
Vapor mole fraction	0.00405658	1.0000	0.00000
Total lbmol/h	5.3727	0.0223	5.3503
Total lb/h	100.0000	1.0893	98.9107
Total std L ft3/hr	1.6222	0.0210	1.6012
Total std V scfh	2038.82	8.48	2030.34
Flowrates in lb/h			
Water	94.8400	0.0105	94.8295
Ethanol	4.2100	0.3487	3.8613
Carbon Dioxide	0.5300	0.5195	0.0105
Pyruvic Acid	0.2100	0.0006	0.2094
Ethyl Acetate	0.2100	0.2100	0.0000

Next, I tried a lower feed temperature (70 °F) with both flashing pressures - 13.226 psia and 14 psia. As expected, less EtOH went overhead while the rest of the components stayed about the same. These results can be seen below and on the next page.

Feed beer at 70°F, flashed from 14.7 to 13.226 psia

Stream No.	1	2	3
Stream Name	Feed	Flash Vapor	Flash Liquid
Temp F	70.0000*	69.8248	69.8248
Pres psia	14.7000*	13.2260	13.2260
Enth MMBtu/h	-0.661364	-0.00312331	-0.658241
Vapor mole fraction	0.00371111	1.0000	0.00000
Total lbmol/h	5.3727	0.0208	5.3519
Total lb/h	100.0000	1.0250	98.9750
Total std L ft3/hr	1.6222	0.0197	1.6025
Total std V scfh	2038.82	7.91	2030.91
Flowrates in lb/h			
Water	94.8400	0.0072	94.8327
Ethanol	4.2100	0.2877	3.9223
Carbon Dioxide	0.5300	0.5196	0.0104
Pyruvic Acid	0.2100	0.0004	0.2096
Ethyl Acetate	0.2100	0.2100	0.0000

Feed beer at 70°F, flashed from 14.7 to 14 psia

	1	2	3
Stream No.	1	2	3
Stream Name	Feed	Flash Vapor	Flash Liquid
Temp F	70.0000*	69.9225	69.9225
Pres psia	14.7000*	14.0000	14.0000
Enth MMBtu/h	-0.661364	-0.00307151	-0.658293
Vapor mole fraction	0.00371111	1.0000	0.00000
Total lbmol/h	5.3727	0.0203	5.3524
Total lb/h	100.0000	1.0026	98.9974
Total std L ft3/hr	1.6222	0.0193	1.6029
Total std V scfh	2038.82	7.72	2031.11
Flowrates in lb/h			
Water	94.8400	0.0067	94.8333
Ethanol	4.2100	0.2668	3.9432
Carbon Dioxide	0.5300	0.5187	0.0113
Pyruvic Acid	0.2100	0.0004	0.2096
Ethyl Acetate	0.2100	0.2100	0.0000

Conclusion: Essentially all of the CO_2 and ethyl acetate go overhead for all of the attempted conditions. Hardly any water or pyruvic acid goes overhead. Less EtOH goes overhead at higher pressures and lower temperatures.

However, since I anticipate using the feed to provide some heat exchange, I'll use

* $\boxed{\rightarrow \text{Feed at } 80°\text{F, flash to 14 psia}}$ *

This way, I can still get the heat exchange I need without losing too much ethanol.

Distillation Calculations

Purpose: Now that I know what the composition of the streams exiting the de-esterizer are, I'm ready to begin some distillation calculations. First, I'll estimate what kind of column I'll need using Fenske - Underwood - Gilliland methods, then I'll simulate the column to see whether it performs as I think it will.

PRESSURE SELECTION: It is very, very important that I not scorch the beer, so I will use a low pressure so as to keep the temperature low. Based on a ChemCad simulation, I see that at a column pressure of 0.5 psia, the range of possible temperatures is 52.5 - 79.4 °F. This is adequate.

XY data for Ethanol / Water @ 0.5 psia

T K Wilson parameters: -315.361 662.320

T Deg F	P psia	Mole Fractions X1	Y1	Gamma1	Gamma2
79.41301	0.50000	0.00000	0.00000	5.20761	1.00000
68.50969	0.50000	0.05000	0.33578	3.87083	1.00743
63.87804	0.50000	0.10000	0.45365	3.05315	1.02691
61.35881	0.50000	0.15000	0.51450	2.51488	1.05575
59.76714	0.50000	0.20000	0.55319	2.14178	1.09236
58.64079	0.50000	0.25000	0.58163	1.87293	1.13580
57.76830	0.50000	0.30000	0.60499	1.67328	1.18552
57.04472	0.50000	0.35000	0.62585	1.52152	1.24119
56.41441	0.50000	0.40000	0.64568	1.40403	1.30268
55.84758	0.50000	0.45000	0.66536	1.31183	1.36998
55.32811	0.50000	0.50000	0.68552	1.23877	1.44316
54.84801	0.50000	0.55000	0.70660	1.18053	1.52238
54.40404	0.50000	0.60000	0.72898	1.13401	1.60786
53.99596	0.50000	0.65000	0.75300	1.09695	1.69986
53.62560	0.50000	0.70000	0.77899	1.06764	1.79867
53.29619	0.50000	0.75000	0.80731	1.04480	1.90466
53.01213	0.50000	0.80000	0.83836	1.02746	2.01817
52.77904	0.50000	0.85000	0.87257	1.01484	2.13960
52.60332	0.50000	0.90000	0.91048	1.00636	2.26935
52.49266	0.50000	0.95000	0.95270	1.00154	2.40781
52.45598	0.50000	1.00000	1.00000	1.00000	2.55536

Distillation calculations [cont'd]

Before I do anything else, I need to convert to a molar rate than mass basis. The molecular weights are:

H_2O	18.016
EtOH	46.069
CO_2	44.011
(P.A.) Pyruvic Acid	86.063

The liquid stream exiting the flash drum consists of:

94.83 lb H_2O
3.86 lb EtOH
0.01 lb CO_2
0.21 lb Pyruvic acid

98.91 lb total

The composition on a molar basis, then, is:

$$H_2O: \frac{94.83 | 18.016}{\frac{94.83}{18.016} + \frac{3.86}{46.069} + \frac{0.01}{44.011} + \frac{0.21}{86.063}} = \frac{5.26 \text{ lbmol}}{5.35 \text{ lbmol}}$$

$$= 98.39 \text{ mol\% } H_2O$$

$$EtOH: \frac{3.86 | 46.069}{5.35} = 1.566 \text{ mol\% EtOH}$$

$$CO_2: \frac{0.01 | 44.011}{5.35} = 0.0042 \text{ mol\% } CO_2$$

$$P.A.: \frac{0.21 | 86.063}{5.35} = 0.046 \text{ mol\% P.A.}$$

Basis of Calculation:

I'll base everything on 100 $\frac{mol}{h}$ of feed. The light key is EtOH, the heavy key is H_2O. Also, since I know there is an azeotrope around 95% EtOH, I'll specify a distillate purity of 95%. Additionally, since I want to get as much EtOH out of the beer as possible, I'll specify 99% recovery.

Distillation calculations [cont'd]

Now, by material balance on EtOH, determine distillate flow rate:

$$\left(1.566 \frac{mol}{h}\right)\left(0.99\right) = 0.95\ D$$

↑ % recovery   ↑ purity of distillate

$$\Rightarrow \underline{D = 1.632\ mol/h}$$

Therefore, the bottoms flow rate is:

$$B = F - D$$
$$= 100 - 1.632$$
$$\underline{B = 98.368\ mol/h}$$

I will assume the distillate consists only of EtOH and H_2O. 95% is EtOH, therefore

moles H_2O in distillate = $0.05\left(1.632\ mol/h\right) = \underline{0.0816 \frac{mol}{h}}$

Again by material balance,

moles H_2O in bottoms = $98.39 - 0.0816$

$= \underline{98.3084\ mol/h}$

For EtOH:

moles EtOH in distillate = $(1.566)(0.99) = \underline{1.5503\ mol/h}$

moles EtOH in bottoms = $0.01(1.566) = \underline{0.0157\ mol/h}$

Now I have almost enough information to use the famous Fenske equation (Eqn. 12-12 of Henley + Seader, Equilibrium-Stage Separation Operations in Chemical Engineering).

$$N_{min} = \frac{\ln\left[\frac{d_i}{d_j} \cdot \frac{b_j}{b_i}\right]}{\ln \alpha_m}$$

Distillation calculations [cont'd]

I still need the relative volatility. I'll estimate a geometric mean using 3 points:

$x_1 = 0.05$, $x_1 = 0.95$, and $x_1 = 0.50$

From the simulation data provided on the first page of these calculations,

at $x_1 = 0.05$,

$$\alpha = \frac{y_1 | x_1}{y_2 | x_2} = \frac{0.33578 / 0.05}{(1-0.33578) / 0.95}$$

$$= 9.60$$

at $x_1 = 0.95$,

$$\alpha = \frac{0.95270 / 0.95}{(1-0.95270) / 0.05}$$

$$= 1.06$$

at $x_1 = 0.5$,

$$\alpha = \frac{0.68552 / 0.5}{(1-0.68552) / 0.5}$$

$$= 2.18$$

The geometric mean relative volatility is:

$$\alpha_m = \left[(9.60)(1.06)(2.18)\right]^{1/3}$$

$$\underline{= 2.81}$$

Substituting into Fenske equation,

$$N_{min} = \frac{\ln\left[\frac{1.5503}{0.0157} \cdot \frac{98.3084}{0.0816}\right]}{\ln 2.81} \longrightarrow \boxed{N_{min} = 11.3 \text{ stages at total reflux}}$$

(Seems a tad low)

Distillation calculations [cont'd]

Next, assume a bubble point feed (a good assumption in this case since the feed comes from a flash drum) and apply the Underwood equation for minimum reflux

$$q=1, \quad R_{min} = \frac{1}{\alpha - 1}\left[\frac{x_D}{z_F} - \frac{\alpha(1-x_D)}{1-z_F}\right]$$

$$= \frac{1}{2.81-1}\left[\frac{0.95}{0.01566} - \frac{2.81(1-0.95)}{1-0.01566}\right]$$

$$= 33.4 \quad \text{(pretty high)}$$

I'll use $R_{actual} = 1.2\, R_{min}$

$$= 1.2(33.4) = 40.08$$

Now use Molokanov curve fit of Gilliland data:

$$Y = 1 - \exp\left[\frac{1+54.4X}{11+117.2X} \cdot \frac{X-1}{X^{0.5}}\right]$$

where $Y = \dfrac{N - N_{min}}{N+1}$ and $X = \dfrac{R - R_{min}}{R+1}$

In this case,

$$X = \frac{40.08 - 33.4}{40.08+1}$$

$$= 0.1626$$

from the above equation,

$$Y = 0.4935$$

substitute and solve for N:

$$0.4935 = \frac{N - 11.3}{N+1}$$

Distillation calculations [cont'd]

$$0.4935\,N + 0.4935 = N - 11.3$$

$$\Rightarrow \underline{\underline{N = 23.3 \text{ stages}}}$$

Now I need an estimate of the stage efficiency. I'll use the plot in Figure 13.5 of Henley and Seader. First, I need a viscosity-volatility product. I'll use $\alpha = 2.81$ and a viscosity of

$$\mu = \frac{\overbrace{1.2}^{\text{pure EtOH}} + \overbrace{1.0}^{\text{pure } H_2O}}{2} = 1.1 \text{ cP}$$

$$\alpha \cdot \mu = 3.091$$

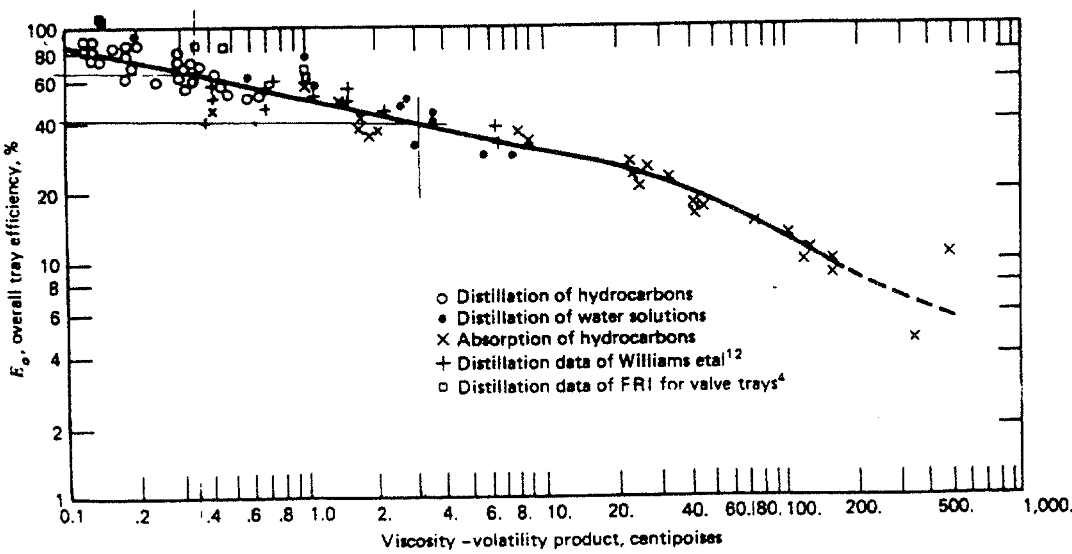


Figure 13.5. Lockhart and Leggett version of the O'Connell correlation for overall tray efficiency of fractionators, absorbers, and strippers. (Adapted from F. J. Lockhart and C. W. Leggett, *Advances in Petroleum Chemistry and Refining*, Vol. 1, eds. K. A. Kobe and John J. McKetta, Jr., Interscience Publishers, Inc., New York, © 1958, 323–326.)

From the plot, the overall tray efficiency is <u>approximately 40%.</u>

Distillation calculations [cont'd]

So, the number of actual stages is:

$$N_{actual} = 23.3/0.40 = 58.3 \longrightarrow \text{round up to } 59$$

Now I'll estimate the feed location using the Kirkbride equation (12-42 of Henley and Seader):

$$\frac{N_R}{N_S} = \left[\left(\frac{z_{HK,F}}{z_{LK,F}}\right)\left(\frac{x_{LK,B}}{x_{HK,D}}\right)^2\left(\frac{B}{D}\right)\right]^{0.206}$$

$$= \left[\left(\frac{98.39}{1.566}\right)\left(\frac{\frac{0.0157}{98.368}}{\frac{0.0816}{1.632}}\right)^2\left(\frac{98.368}{1.632}\right)\right]^{0.206}$$

$\frac{N_R}{N_S} = 0.511 \longrightarrow$ most of the trays will be in the stripping section of the column

So:

59 trays $= N_S + N_R$

$59 = N_S + 0.511 N_S$

$\Rightarrow N_S = 39$
$\Rightarrow N_R = 20$ $\longrightarrow$ So, 39 stages below the feed, 20 above (about 2/3)

** Summary of Distillation Calculations **

59 trays, 39 below feed, 20 above

Reflux ratio = 40

Tray efficiency ≈ 40%

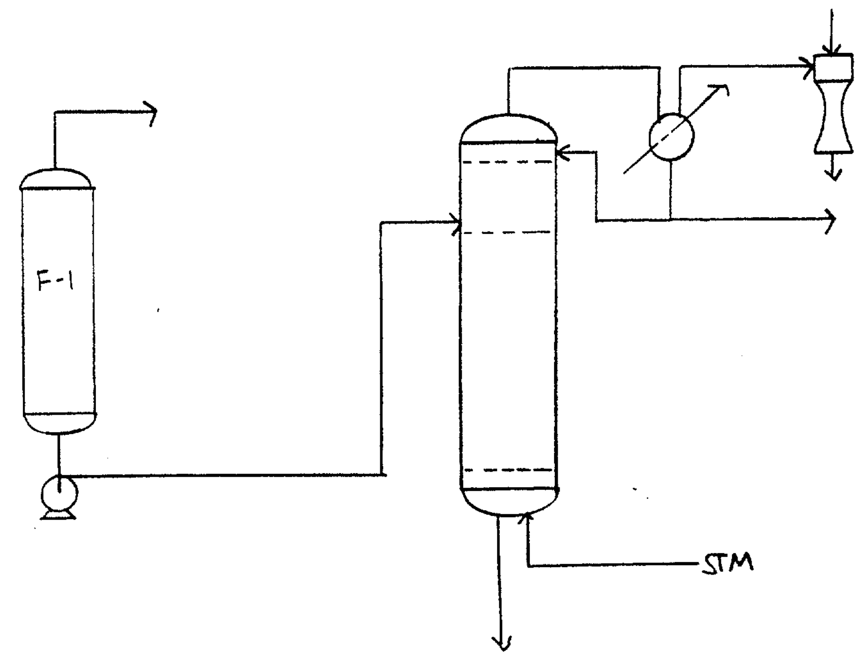
	Bottoms (mol/h)	Distillate (mol/h)	Feed (mol/h)
H_2O	98.308	0.082	98.39
EtOH	0.016	1.550	1.566
CO_2	0	0.0042	0.0042
[illegible] acid	0.046	0	0.046
Total	98.368	1.632	100

Simulations

Purpose: Now that I've completed some approximate calculations for the distillation column, I'll run some simulations on ChemCad to get more exact numbers. Hopefully, the simulation results will support my calculated results.

Model: I'll use the Simultaneous Corrective Distillation Simulation (SCDS) module, which is good for non-ideal systems, such as those that azeotrope.

Conditions: I'll start out with the conditions listed on page 7 of the distillation calculations. The feed to the column will be the same as the liquid outlet of the de-esterizer.



I'll set the overhead pressure of the column at 0.5 psia. Also, I'll do the calculations first as if there is a reboiler present in order to get a heat duty, then I'll use steam injection.

I need to convert the flowrates to lb/h → based on 98.975 lb/h

	Bottoms (lb/h)	Distillate (lb/h)	Feed (lb/h)
H_2O	94.75	0.08	94.83
EtOH	0.10	3.82	3.92
CO_2	0	0.01	0.01
Cyrmic Acid	0.21	0	0.21
Total	95.06	3.91	98.97

Simulations [cont'd]

For my own records, here is the information I have entered into the SCDS module:

top pressure 0.5 psia
Cond pressure drop 5 psi
Coln pressure drop 5.9 psi

No. of stages 61
1st feed stage 22

Top stage efficiency 0.43
Last stage efficiency 0.32

1 EB, 1 MB specification { Reflux ratio 40
Bottom mass flowrate 95 lb/h

Dist. rate 1.632 lbmol/h
Reflux rate 65.28 lbmol/h
T top 72 °F
T bottom 60 °F

ChemCad then calculated condenser duty = -0.081 mmBtu/h
reboiler duty = 0.093 mmBtu/h

The results of the simulation were quite pleasing:

Simulation results with column pressure drop of 5.9 psia

Stream No.	3	4	5
Stream Name	Flash Liquid	Distillate	Bottoms
Temp F	69.8248	-18.1112*	199.2113
Pres psia	13.2260	0.5000*	11.4000
Enth MMBtu/h	-0.658241	-0.0112128	-0.634939
Vapor mole fraction	0.00000	1.6595E-006	0.000171329
Total lbmol/h	5.3519	0.0920	5.2598
Total lb/h	98.9750	3.9749	95.0000
Total std L ft3/hr	1.6025	0.0794	1.5231
Total std V scfh	2030.91	34.93	1995.98
Flowrates in lb/h			
Water	94.8327	0.1703	94.6624
Ethanol	3.9223	3.7942	0.1280
Carbon Dioxide	0.0104	0.0104	0.0000
Pyruvic Acid	0.2096	0.0000	0.2096
Ethyl Acetate	0.0000	0.0000	0.0000

Next simulation

I'm not entirely convinced that the pressure drop in a vacuum system is handled the same way as in a normal system, so I ran one case with no pressure drop through the column (or condenser). Here are the results:

Condenser duty = − 0.077 MMBtu/h
Reboiler duty = 0.078 MMBtu/h

The rest of the results are shown below. They are again quite pleasing, with the added benefit of keeping temperatures lower, i.e. lower possibility of scorching.

Simulation results with zero column pressure drop

Stream No.	3	4	5
Stream Name	Flash Liquid	Distillate	Bottoms
Temp F	69.8248	-13.4406*	79.2609
Pres psia	13.2260	0.5000*	0.5000*
Enth MMBtu/h	-0.658241	-0.0111427	-0.646397
Vapor mole fraction	0.00000	0.00000	0.00000*
Total lbmol/h	5.3519	0.0915	5.2604
Total lb/h	98.9750	3.9750	95.0000
Total std L ft3/hr	1.6025	0.0795	1.5230
Total std V scfh	2030.91	34.72	1996.20
Flowrates in lb/h			
Water	94.8327	0.1537	94.6790
Ethanol	3.9223	3.8109	0.1114
Carbon Dioxide	0.0104	0.0104	0.0000
Pyruvic Acid	0.2096	0.0000	0.2096
Ethyl Acetate	0.0000	0.0000	0.0000

Steam Distillation

Since the simulation was successful, I'll now try to simulate steam distillation. I'll start from the fact that the calculated reboiler duty was 0.078 mmBtu/h. The plant has 100 psig saturated steam. When this is flashed to 0.5 psia at the bottom of the column, it will be superheated. As it gives up its enthalpy, the bottoms liquid will be reboiled.

From steam tables:

H of 100 psig saturated steam = 1189.6 Btu/lbm

H of 0.5 psia saturated liquid (saturation temperature = 80°F, so heat transfer can still occur) = 48.04 Btu/lbm

Solve for the lb of steam required:

$$0.078 \times 10^6 \text{ Btu/h} = \left(1189.6 \frac{\text{Btu}}{\text{lbm}} - 48.04 \frac{\text{Btu}}{\text{lbm}}\right) \dot{m}$$

$$\dot{m} = 68.3 \text{ lb/h}$$

I'll simulate this. Here are the results:

Simulation results with steam feed

Stream No.	3	4	5	6
Stream Name	Flash Liquid	Distillate	Bottoms	Steam feed
Temp F	69.8248	-18.3396*	79.3305	338.0092
Pres psia	13.2260	0.5000*	0.5000*	114.7000*
Enth MMBtu/h	-0.658241	-0.0111635	-1.1120	-0.387399
Vapor mole fraction	0.00000	0.00000	0.00000*	1.0000*
Total lbmol/h	5.3519	0.0917	9.0515	3.7913
Total lb/h	98.9750	3.9825	163.2924	68.3000
Total std L ft3/hr	1.6025	0.0796	2.6180	1.0951
Total std V scfh	2030.91	34.78	3434.84	1438.71
Flowrates in lb/h				
Water	94.8327	0.1540	162.9787	68.3000
Ethanol	3.9223	3.8182	0.1041	0.0000
Carbon Dioxide	0.0104	0.0104	0.0000	0.0000
Pyruvic Acid	0.2096	0.0000	0.2096	0.0000
Ethyl Acetate	0.0000	0.0000	0.0000	0.0000

Comments: the steam feed works again well as far as recovering EtOH is concerned. However, as I should have realized, it also adds a great deal of water to the system. This would result in watered-down beer!

Therefore, I WILL NOT USE STEAM DISTILLATION

Next simulation (particularly pressure drop)

I confess that I don't completely understand all the implications of vacuum distillation, so I need to run another case. I found a source (Perry and Green) that says that using specially designed trays, a column can be operated with a bottom tray pressure of 40-50 torr. This corresponds to:

$$P = 50 \text{ torr} \cdot \frac{14.7 \text{ psia}}{760 \text{ torr}} = 0.97 \text{ psia}$$

The column in question in the reference seemed to have significantly less than 60 trays (it was never stated explicitly), so to be conservative, I'll design for 2 psia at the bottom. Here are the results of this simulation using an overhead pressure of 0.5 psia and a bottoms pressure of 2 psia.

Simulation results with 0.5 psia overhead pressure, 2 psia bottoms pressure

Stream No.	3	4	5
Stream Name	Flash Liquid	Distillate	Bottoms
Temp F	69.8248	-18.3705*	125.7646
Pres psia	13.2260	0.5000*	2.0000*
Enth MMBtu/h	-0.658241	-0.0111573	-0.641982
Vapor mole fraction	0.00000	0.00000	0.00000*
Total lbmol/h	5.3519	0.0916	5.2602
Total lb/h	98.9750	3.9749	95.0001
Total std L ft3/hr	1.6025	0.0795	1.5230
Total std V scfh	2030.91	34.76	1996.15
Flowrates in lb/h			
Water	94.8327	0.1572	94.6755
Ethanol	3.9223	3.8074	0.1149
Carbon Dioxide	0.0104	0.0104	0.0000
Pyruvic Acid	0.2096	0.0000	0.2096
Ethyl Acetate	0.0000	0.0000	0.0000
Glycolic Acid	0.0000	0.0000	0.0000

Condenser duty = -0.077442 mmBtu/h
Reboiler duty = 0.082543 mmBtu/h

Revised Base Case

Sizing Information

Purpose: Now that I have the initial flowsheet drawn and simulations completed, the next step is to size all of the equipment. To do this, I'll need several pieces of physical property information. Rather than do all of the tedious calculations by hand, I'll get values from ChemCad.

Methods: I will use the sizing procedures detailed in Chapter 13 of Henley and Seader. For sizing the large distillation column, I'll take values at various intervals along the column and make an appropriate judgment.

For sizing the de-esterizer:

$L = 98.975$ lb/h
$\rho_L = 61.662$ lb/ft^3

$V = 1.025$ lb/h
$\rho_V = 0.18$ lb/ft^3

* Note: the information on this page is based on a feed rate of 100 lb/h for simulation work and can be converted to actual rates by multiplication of the scale factor, which is 98.43

For sizing the EtOH/H_2O tower:

At tray 59

$L = 172$ lb/h	$V = 163$ lb/h
$M_L = 43.26$ lb/lbmol	$M_V = 43.39$ lb/lbmol
$\rho_L = 50.26$ lb/ft^3	$\rho_V = 0.00395$ lb/ft^3
$\sigma = 24.138$ dyne/cm	

At tray 45

$L = 157$ lb/h	$V = 164$ lb/h
$M_L = 39.16$ lb/lbmol	$M_V = 39.83$ lb/lbmol
$\rho_L = 50.51$ lb/ft^3	$\rho_V = 0.00604$ lb/ft^3
$\sigma = 25.284$ dyne/cm	

At tray 30

$L = 239$ lb/h	$V = 144$ lb/h
$M_L = 25.17$ lb/lbmol	$M_V = 34.03$ lb/lbmol
$\rho_L = 55.11$ lb/ft^3	$\rho_V = 0.00722$ lb/ft^3
$\sigma = 39.682$ dyne/cm	

At tray 15

$L = 237$ lb/h	$V = 143$ lb/h
$M_L = 24.82$ lb/lbmol	$M_V = 33.33$ lb/lbmol
$\rho_L = 55.12$ lb/ft^3	$\rho_V = 0.00908$ lb/ft^3
$\sigma = 39.798$ dyne/cm	

Sizing info [cont'd]

At tray 1

$L = 176$ lb/h
$M_L = 18.11$ lb/lbmol
$\rho_L = 61.51$ lb/ft^3
$\sigma = 66.816$ dyne/cm

$V = 82$ lb/h
$M_v = 18.42$ lb/lbmol
$\rho_v = 0.00581$ lb/ft^3

For sizing the condenser:

Duty = 0.06932 mmBtu/h
Exit T = 52.59 °F
Composition of stream to be condensed (wt frac):

H_2O	0.04
EtOH	0.96

Entry T = 52.~~55~~ °F

Also a heat curve (T vs. q) was plotted and is shown on the next page.

For sizing the reboiler:

Duty = 0.10208 mmBtu/h
Exit T = 125.8
Composition of stream (wt frac):

H_2O	0.997
EtOH	0.001
Pyruvic Acid	0.002

Entry T = 128.1

A heat curve has also been drawn for the reboiler

I also need information about the hot water side of the reboiler. I don't want too great of a temperature difference (so as to avoid scorching the beer). I'll use 150°F water at atmospheric pressure and let it cool down to 136°F (minimum 10°F temperature difference). First I need to know how much water this requires:

from steam table, $H^l_{150} = 118$ Btu/lb
$H^l_{136} = 104$ Btu/lb

using the duty above,

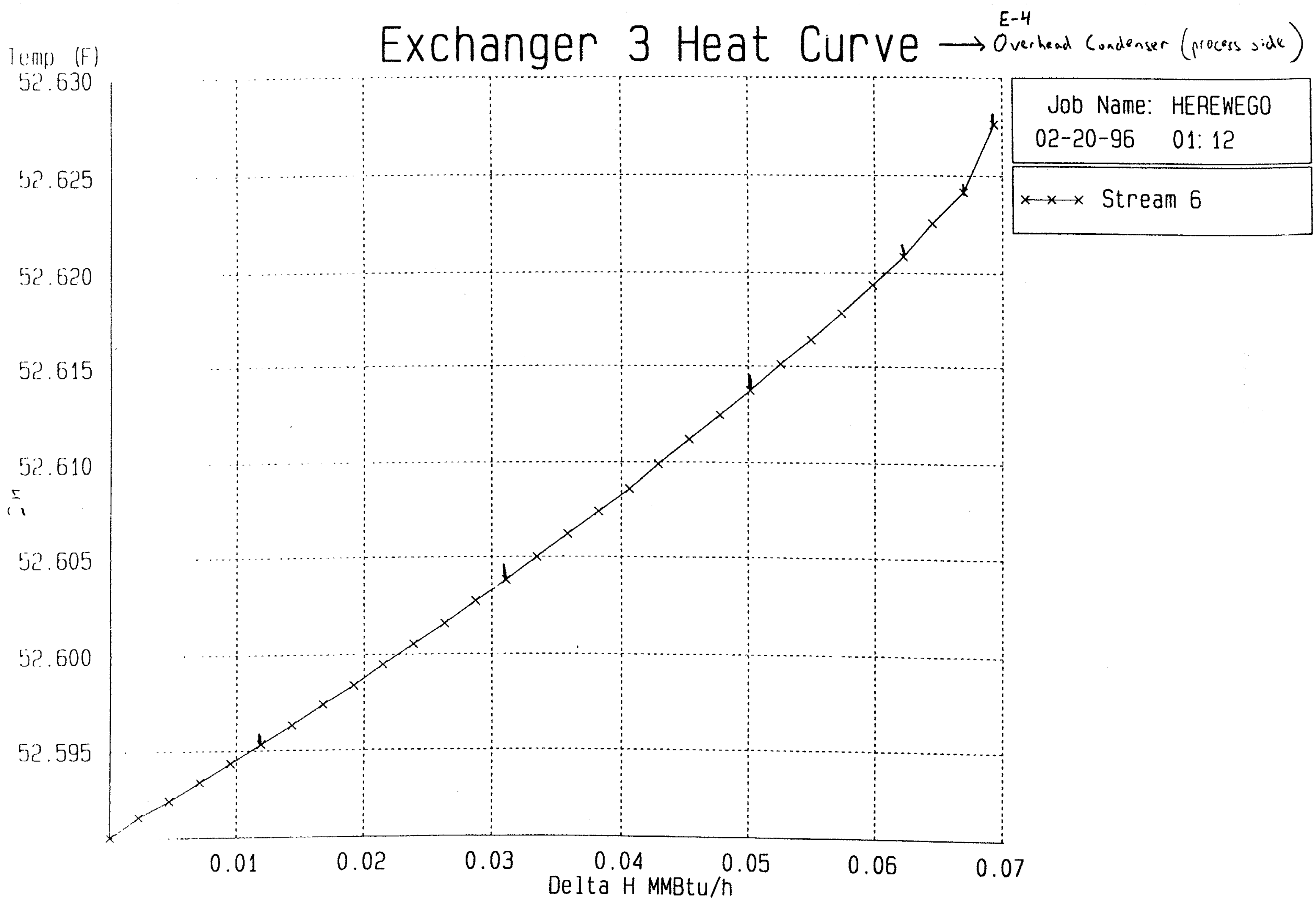
$$0.10208 \times 10^6 \text{ Btu/h} = \left(118 - 104 \ \frac{\text{Btu}}{\text{lb}}\right) \cdot \dot{m}$$

$$\Rightarrow \dot{m} = 7291 \text{ lb/h}$$

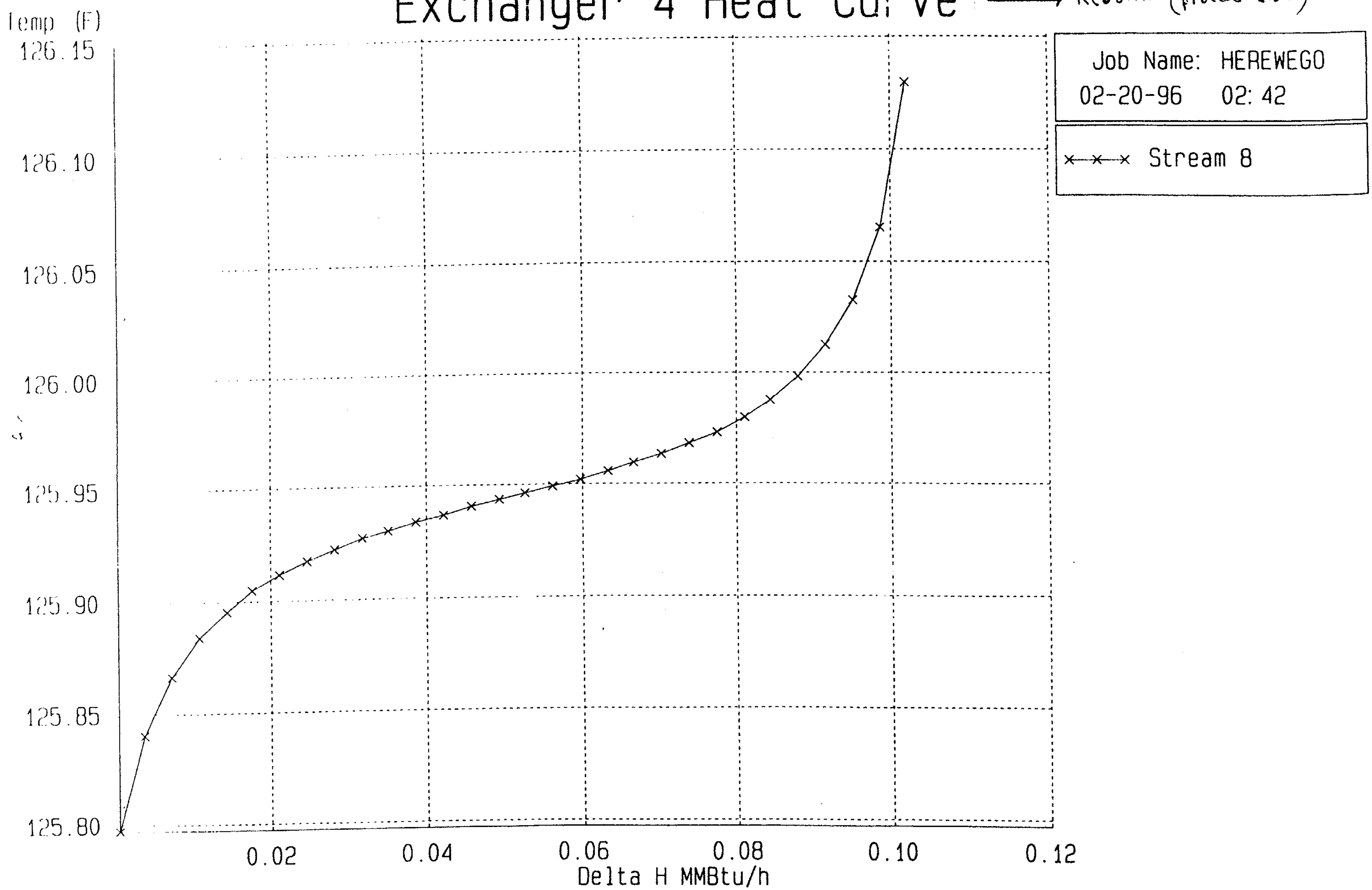
From this information, I can generate a heating curve

Column Sizing Information

Stream No.	1	2	3
Name	Feed	Flash Vapor	Flash Liquid
Overall - -			
Molar flow lbmol/h	5.3727	0.0208	5.3519
Mass flow lb/h	100.0000	1.0250	98.9750
Temp F	70.0000	69.8248	69.8248
Pres psia	14.7000	13.2260	13.2260
Vapor mole fraction	0.0000	0.63824	0.0000
Enth MMBtu/h	-0.66156	-0.0032544	-0.65824
Tc F	693.3578	325.9887	696.3198
Pc psia	3024.3713	1724.5043	3058.2607
Std. sp gr , wtr = 1	0.988	0.833	0.990
Std. sp gr , air = 1	0.643	1.698	0.639
Degree API	11.6686	38.3144	11.3926
Average mol wt	18.6126	49.1775	18.4936
Actual dens lb/ft3	61.508	0.180	61.662
Actual vol ft3/hr	1.6258	5.6845	1.6051
Std liq ft3/hr	1.6222	0.0197	1.6025
Std vap 60F scfh	2038.8240	7.9092	2030.9136
- - Vapor only - -			
Molar flow lbmol/h		0.0133	
Mass flow lb/h		0.6021	
Average mol wt		45.2660	
Actual dens lb/ft3		0.106	
Actual vol ft3/hr		5.6763	
Std liq ft3/hr		0.0116	
Std vap 60F scfh		5.0480	
Cp Btu/lbmol-F		10.1859	
Z factor		0.9934	
Visc cP		0.013644	
Th cond Btu/hr-ft-F		0.0091	
- - Liquid only - -			
Molar flow lbmol/h	5.3727	0.0075	5.3519
Mass flow lb/h	100.0000	0.4228	98.9750
Average mol wt	18.6126	56.0787	18.4936
Actual dens lb/ft3	61.508	51.648	61.662
Actual vol ft3/hr	1.6258	0.0082	1.6051
Std liq ft3/hr	1.6222	0.0081	1.6025
Std vap 60F scfh	2038.8240	2.8612	2030.9136
Cp Btu/lbmol-F	18.2287	30.0650	18.1713
Z factor	0.0010	0.0029	0.0009
Visc cP	1.004	0.89127	1.012
Th cond Btu/hr-ft-F	0.3220	0.0924	0.3274
Surf tens dyne/cm	67.5500	22.5002	69.1073


Exchanger 3 Heat Curve
E-4
Overhead Condenser (process side)
Job Name: HEREWEGO
02-20-96 01: 12
Stream 6
Temp (F)
52.630
52.625
52.620
52.615
52.610
52.605
52.600
52.595
0.01
0.02
0.03
0.04
0.05
0.06
0.07
Delta H MMBtu/h

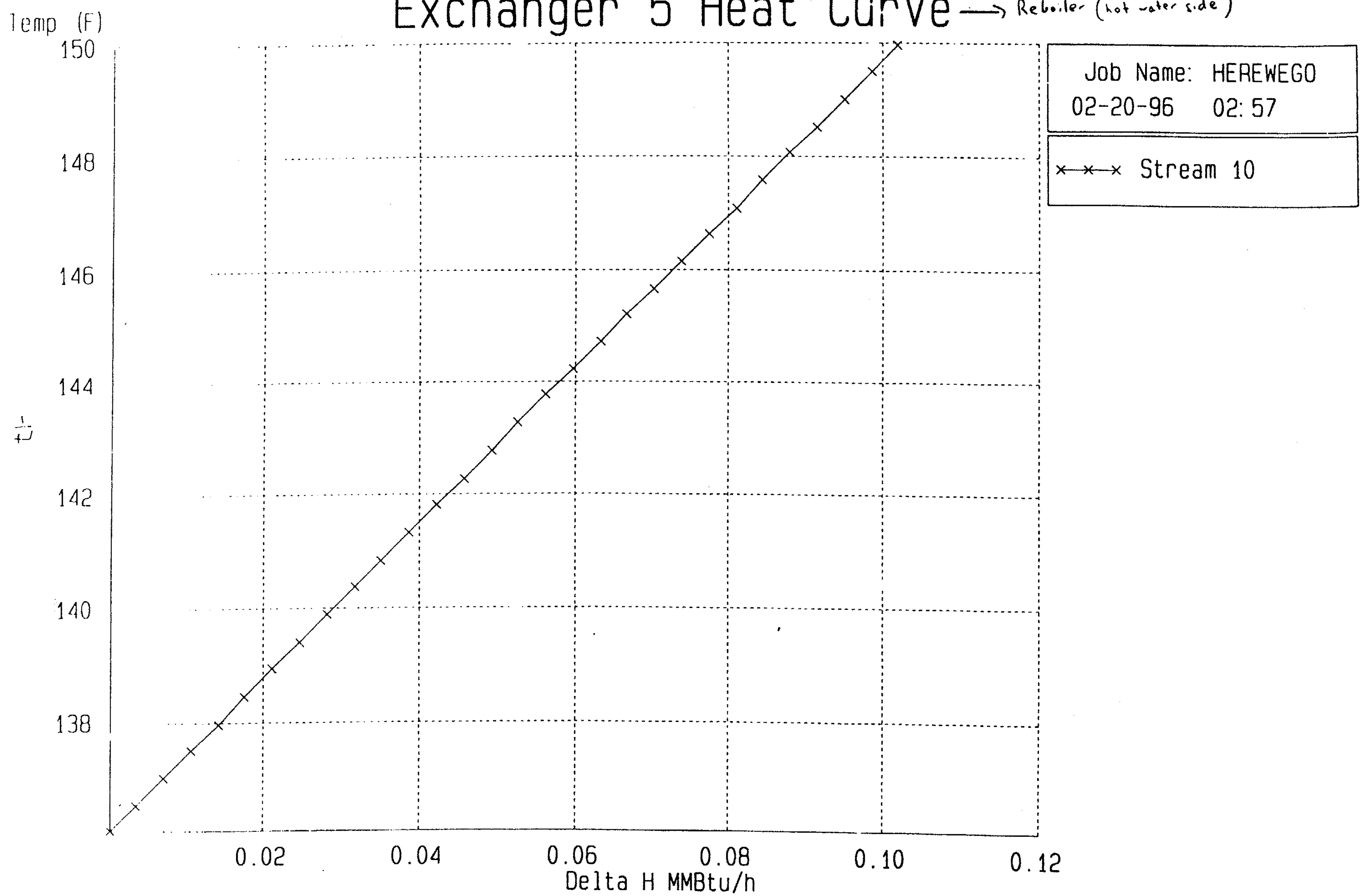
Exchanger 4 Heat Curve → E-7 Reboiler (process side)



Exchanger 5 Heat Curve

E-9 → Hot water reheater (hot water side)

E-7 → Reboiler (hot water side)



Sizing info [cont'd]

Reflux drum sizing: $L = 160.31$ lb/h
$M_L = 43.37$ lb/lbmol
$\rho_L = 50.243$ lb/ft^3

- - - - - - - - - - - - - - - - -

For sizing the flash condenser:

Stream No.	14
Stream Name	
Temp F	142.9626
Pres psia	15.7000
Enth MMBtu/h	-0.428840
Vapor mole fraction	0.718108
Total lbmol/h	3.3747
Total lb/h	131.5299
Total std L ft3/hr	2.4785
Total std V scfh	1280.64
Flowrates in lb/h	
Water	17.0335
Ethanol	34.3230
Carbon Dioxide	51.1344
Pyruvic Acid	0.0590
Ethyl Acetate	20.6700
Air	8.3100

Outlet stream of ejector with combined steam, leaked air, and flash vapor

I want to condense everything but the CO_2 and air, which comprise about 45.2% $\left(\frac{51.13 + 8.31}{131.53}\right)$ of the system. By manipulating the simulator, I should be able to get a decent heating curve.

The result is:

Inlet temp = 143 °F
Outlet temp = 54.1 °F
Duty = -0.0225 mmBtu/h

A heating curve has been constructed.

Sizing flash condenser info [cont'd]

I need to get a heating curve for the cooling water. Cooling H_2O can take me down to 130°F (120°F max, 10° minimum approach temperature), but no further. I will have to use refrigeration also.

From the T vs. q curve, when the process fluid is brought down to 130°F, it loses

$$(2.167 \times 10^{-2} \text{ mmBtu/h} - 1.5 \times 10^{-2} \text{ mmBtu/h}) = 6666.7 \text{ Btu/h of energy.}$$

From the steam tables, $H_{120°} = 58$ Btu/lb
$H_{90°} = 88$ Btu/lb

Solve for mass flow of cooling H_2O:

$$6666.7 \frac{\text{Btu}}{\text{h}} = \dot{m}(88 - 58 \text{ Btu/lb})$$

$\Rightarrow \dot{m} = 222$ lb/h of cooling water needed

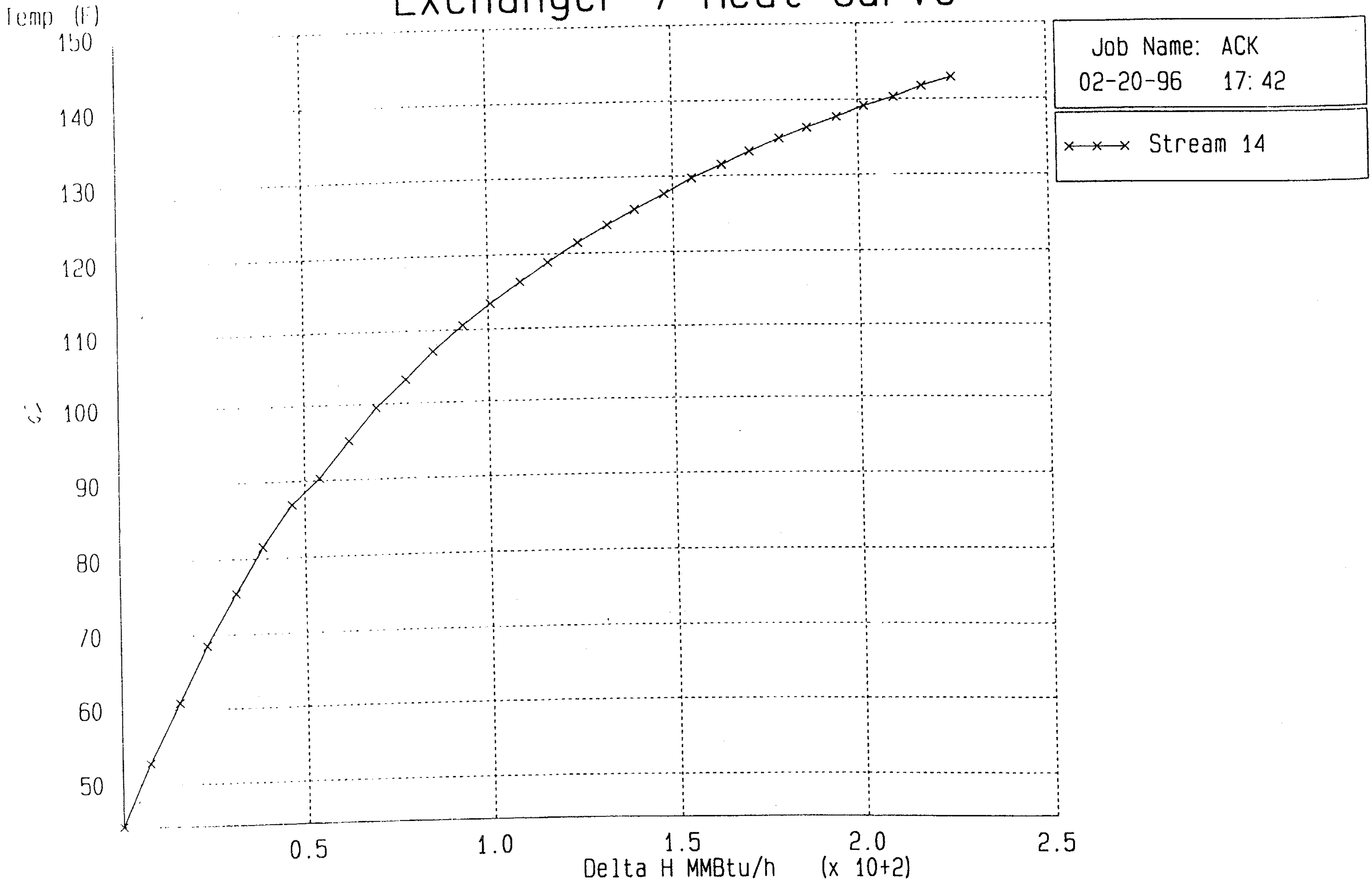
A heating curve has been constructed.

Liquid and vapor formed after heat is removed in after flash condenser:

Stream No.	19	20
Stream Name		
Temp F	44.1179	44.1179
Pres psia	15.7000	15.7000
Enth MMBtu/h	-0.191144	-0.260197
Vapor mole fraction	1.0000	0.00000
Total lbmol/h	1.4249	1.9498
Total lb/h	58.5050	73.0249
Total std L ft3/hr	1.1267	1.3519
Total std V scfh	540.72	739.92
Flowrates in lb/h		
Water	0.1713	16.8622
Ethanol	0.7228	33.6002
Carbon Dioxide	48.6381	2.4963
Pyruvic Acid	0.0000	0.0590
Ethyl Acetate	0.6629	20.0071
Air	8.3098	0.0002
Totals	58.51	73.02

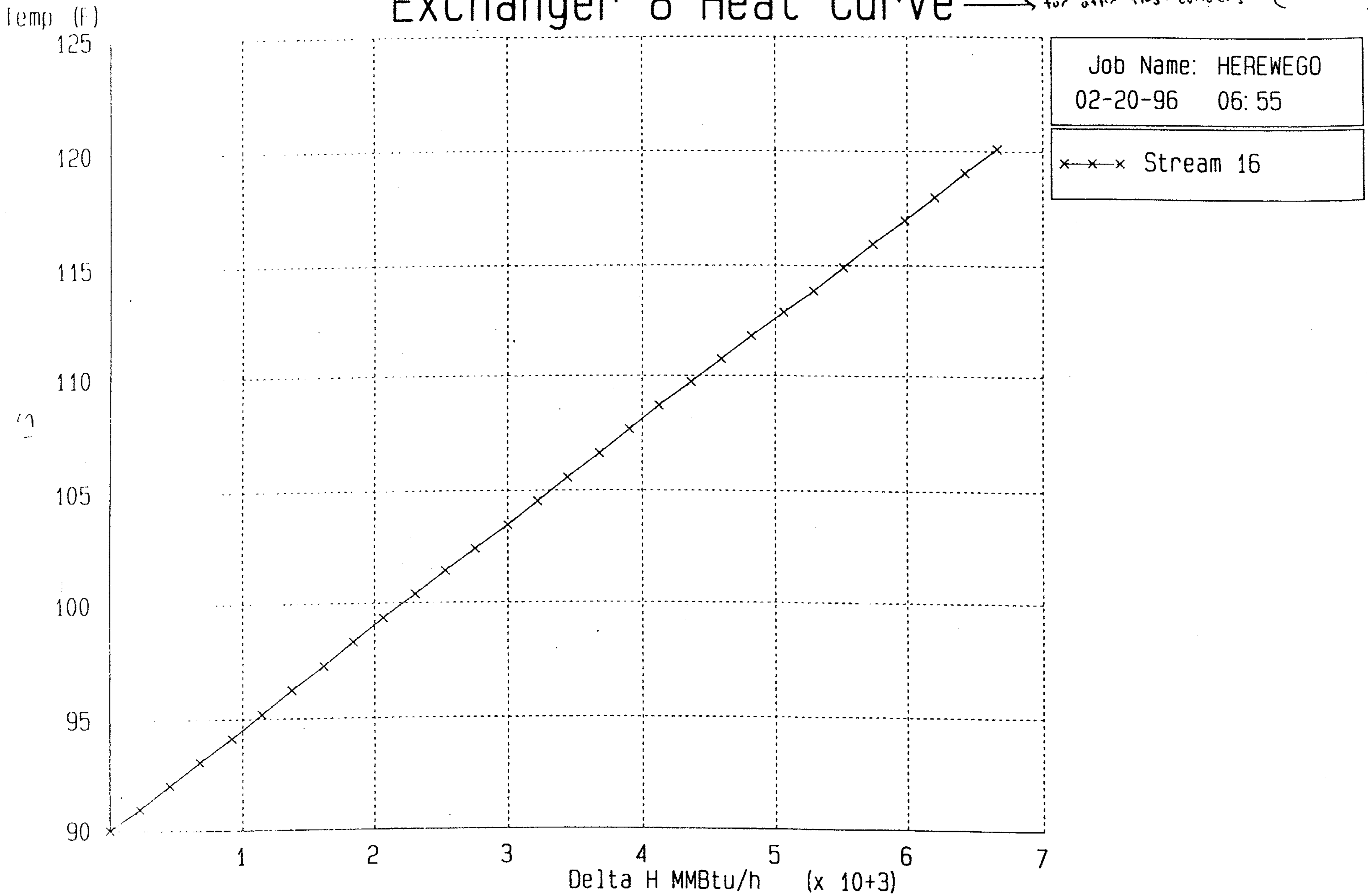
Note that only a small amount of ethyl acetate is lost in the vapor and only a small amount of air is dissolved in the liquid.

Exchanger 7 Heat Curve → for after flash condenser (process side)



Exchanger 8 Heat Curve

for after flash condenser (CW side)



Sizing info [cont'd]

Information needed to size vent drum: $L = 73.0249$ lb/h

(from Chemcad) $\rho_L = 54.580$ lb/ft^3

Information needed to size recombiner drum:

* The recombiner drum combines two streams: distillation bottoms (stream 10) and condensate from de-esterizer (stream 23).

Stream 23	Stream 10
T = 44.1 °F	T = 125.76 °F
P = 15.7 psia	P = 7 psia
16.86 lb/h H_2O	9318.9 lb/h H_2O
33.6 lb/h EtOH	11.3 lb/h EtOH
2.50 lb/h CO_2	—
0.059 lb/h P.A.	20.63 lb/h P.A.
20.01 lb/h Ethyl acetate	—
0.0002 lb/h Air	—

Mixing the streams results in stream 21.

Stream 21 → Product

T = 125.32 °F
P = 7 psia
9335.76 lb/h H_2O
44.9 lb/h EtOH
2.5 lb/h CO_2
20.689 lb/h P.A.
20.01 lb/h Ethyl acetate
0.0002 lb/h air

* MATERIAL BALANCE MOMENT OF TRUTH *

Now I need to do a quick check to make sure that I do not exceed the requirement of < 0.5 vol% EtOH. Adjusting to standard conditions of 68°F and 14.7 psia, Chemcad says that the total volumetric flow rate is:

$$V = 151.3 \text{ ft}^3/\text{h}.$$

The volume % ethanol is:

$$\%\,EtOH = \frac{44.9 \frac{lb}{h}\ EtOH \cdot \frac{1}{(0.8)(62.3 \frac{lb}{ft^3})}}{151.3 \ ft^3/h} \times 100\%.$$

THE NEXT PAGE IS IMPORTANT. IT DISCUSSES THE NEED TO RE-DO SOME CALCULATIONS. UNFORTUNATELY, A MISTAKE WAS MADE IN THE VERY FIRST FLASH CALCULATION.

Material balance moment of truth [cont'd]

= 0.6 vol% ⇒ TOO MUCH!!

Now I have a big decision to make. I can either change the entire process around or I can dilute the stream a little bit with water. I'll go for the first option. Most of the ethanol is coming from the acetate stream and not from the distillate.

Important news! ⟶ Actually, I just redid the initial flash calculation, and it turned out much differently than the first time. (Ah, the pitfalls of using a simulator ...)

Here are the results when feed at 80°F is flashed to 13.226 psia (3" vacuum ⇒ corresponds to Pabst)

Stream No.	1	2	24
Stream Name	Feed	Flash Vapor	Flash Liquid
Temp F	80.0000*	79.2410	79.2410
Pres psia	14.7000*	13.2260	13.2260
Enth MMBtu/h	-0.660584	-0.00200838	-0.658573
Vapor mole fraction	0.00000	1.0000	0.00000
Total lbmol/h	5.3727	0.0124	5.3603
Total lb/h	100.0000	0.5415	99.4585
Total std L ft3/hr	1.6222	0.0105	1.6117
Total std V scfh	2038.82	4.72	2034.10
Flowrates in lb/h			
Water	94.8400	0.0084	94.8316
Ethanol	4.2100	0.0426	4.1674
Carbon Dioxide	0.5300	0.4816	0.0484
Pyruvic Acid	0.2100	0.0000	0.2100
Ethyl Acetate	0.2100	0.0088	0.2012
Air	0.0000	0.0000	0.0000

This is the feed to the column now

Sizing Calculations

Before I size some pieces of equipment, I need to size others. So, before I finish gathering sizing information I'll size a few preliminary pieces of equipment.

→ **Basis of initial sizing:** I know that according to the government, a "large" brewery produces over 60000 bbl of beer a year. Therefore, I'll arbitrarily choose 240000 bbl/yr of feed (4 times this low amount)

To convert this to lb/h, I will also assume 8400 operating hours/yr. I also need a density, which I'll get from ChemCad:

$$\rho_{\text{feed beer at } 80°F} = 61.362 \text{ lb/ft}^3$$

De-esterizer Flash Vessel

The feed rate to this vessel is:

$$240000 \frac{\cancel{bbl}}{\cancel{yr}} \cdot \frac{1 \cancel{yr}}{8400 \text{ hr}} \cdot 42 \frac{\cancel{gal}}{\cancel{bbl}} \cdot \frac{1 \cancel{ft^3}}{7.48 \cancel{gal}} \cdot 61.362 \frac{lb}{\cancel{ft^3}} = 9843 \text{ lb/hr}$$

From the flash simulation on p.3 of the preliminary simulations, 98.91% of this material will remain in the liquid phase. I'll use equation 13-9 of Henley and Seader:

$$V_v = 2Lt/\rho_L$$ → this corresponds to a half-full vessel with a liquid residence time of t

I want a residence time of at least 5 min, so:

$$V_v = 2\left(9843 \frac{lb}{hr}\right)(0.9891)(5 \text{ min})\left(\frac{1 \text{ hr}}{60 \text{ min}}\right) \Big/ 61.362 \text{ lb/ft}^3$$
$$= 26.4 \text{ ft}^3$$

I'll follow the suggestion of Pabst and size the vessel at <u>10 times</u> the required volume so as to ensure no foaming/entrainment problems. Therefore,

$$\text{Required } V = 264 \text{ ft}^3$$

I'll use $H = 4D$. Therefore,

$$D = \frac{H}{4} = \left(\frac{V}{\pi}\right)^{1/3} = \left(\frac{264}{\pi}\right)^{1/3} = 4.4 \rightsquigarrow \text{round to } 4.5 \text{ ft}$$

De-esterizer Flash Vessel [cont'd]

Thus, the height should be 4×4.5 ft $\times 4 = 18$ ft

Summary:	Height = 18 ft
	Diameter = 4.5 ft
	Total volume = 286 ft^3 = 2141 gal

De-esterizer Ejector Sizing

Now I need to determine how much steam the ejector will require. First, I need to calculate the air leakage into the system. I'll use the equation from reference 4:

$$MAL = A \times V_s^{B}$$

In this case, $90 < P < 760$ mm Hg, so $A = 0.1955$, $B = 0.6630$

$$V_s = 286 \text{ ft}^3$$

$$\text{So, } MAL = 0.1955\,(286)^{0.6630}$$

$$= 8.31 \text{ lb/h}$$

Also, the vapor load from the flash is 1.09% of 9843 lb/h. Thus, the total load is

$$8.31 + (0.0109)\,9843 = \underline{115.6 \text{ lb/h}}$$

Now I need to use the procedure on p. 249-250 of reference 3 to size the ejector.

First, average molecular weight = $\dfrac{\underset{\text{air}}{29}\,(8.31) + 0.0109(9843)(52.94)}{8.31 + 0.0109\,(9843)}$

MW of the vapor from Chemcad

$$= 51.2$$

From figure 9-8 of reference 4,

[illegible] = 1.24

De-esterizer ejector [cont'd]

from Fig 9-7, ratio = 1.00 (assuming 80°F gas temperature)

Now I need the value of r in Eq. 9-4.

$$\frac{P_{disch.}}{P_{surge}} = \frac{15.7 \text{ psia (discharge to above atm)}}{14 \text{ psia}} = 1.12$$

Assuming 100 psig motive steam,

$$P_s / P_m = \frac{14 \text{ psia}}{100 + 14.7 \text{ psia}} = 0.122$$

Now, from Fig. 9-9, $\frac{1}{r} \approx 6$ (curve extrapolated)

→ so $r = 0.167$

From before,

$$\text{AEL (air equivalent load)} = \frac{115.6}{(1.24)(1.00)} = 93.2 \text{ lb/h}$$

From Fig 9-10,

SC (size correction) = 1.03

Now use equation 9-4 of reference 4:

$$w_s = (AEL)(r)(SC)$$
$$= (93.2 \text{ lb/h})(0.167)(1.03)$$
$$= \boxed{16.0 \text{ lb/h of 100 psig motive steam required}}$$

Now I have enough information to do gather some more sizing info.

Repeating Distillation Simulation.

With the new flash calculation, the column needs to be simulated again. Here are the results of the simulation:

Newest simulation results

Stream No.	3	4	5
Stream Name	Flash Liquid	Distillate	Bottoms
Temp F	79.2410*	-86.6938*	125.8263
Pres psia	13.226 ~~14.0000~~	0.5000*	2.0000*
Enth MMBtu/h	-0.658573	-0.0126456	-0.641914
Vapor mole fraction	0.00000	0.00000	0.00000*
Total lbmol/h	5.3603	0.1010	5.2593
Total lb/h	99.4586	4.4586	95.0000
Total std L ft3/hr	1.6117	0.0888	1.5229
Total std V scfh	2034.11	38.32	1995.79
Flowrates in lb/h			
Water	94.8316	0.1680	94.6636
Ethanol	4.1674	4.0940	0.0734
Carbon Dioxide	0.0484	0.0484	0.0000
Pyruvic Acid	0.2100	0.0000	0.2100
Ethyl Acetate	0.2012	0.1482	0.0530
Air	0.0000	0.0000	0.0000

By comparison to the earlier simulation results (p. 5 of simulation section), the results are not changed appreciably. Therefore, no recalculation of column properties (for sizing purposes) is warranted.

De-esterizer Ejector Resizing

The ejector, however, does need to be resized. As before, the air leakage is 8.31 lb/h. The vapor load from the flash is different, however. It is:

$$\left(\underset{\text{noncondensable volatiles}}{0.2} + 0.0084 + 0.0426 + 0.4816 + 0.0088\right)\underset{\text{scale-up factor}}{98.43} = 73.0 \text{ lb/h}$$

The total vapor load, then, is $8.31 + 53.3 = 81.3$ lb/h

I need average MW:

$$MW = \frac{\underset{\text{air}}{29}(8.31) + 53.3(\underset{\text{from ChemCad}}{43.54})}{8.31 + 53.3}$$

$$= 41.6$$

Ejector resizing [cont'd]

from Figure 9-8, ratio = 1.15

from Figure 9-7, assuming 80°F, ratio = 1.00

$$\text{So, } AEL = \frac{81.3}{(1.15)(1.0)} = 70.7 \text{ lb/h}$$

$$\text{Also, } \frac{P_{disch}}{P_{suct}} = \frac{15.7 \text{ psia}}{13.226 \text{ psia}} = 1.19$$

$$\frac{P_{suct}}{P_m} = \frac{13.226 \text{ psia}}{100 + 14.7 \text{ psia}} = 0.115$$

from Fig 9-9, $\frac{1}{r} \approx 3.3$

Assume SC = 1.0. Then from equation 9-4,

$$w_s = (AEL)(r)(SC)$$
$$= (70.7 \text{ lb/h})\left(\frac{1}{3.3}\right)(1.0)$$
$$= 21.4 \text{ lb/h}$$

From Fig 9-10, SC = 1.35.
So new w_s = 21.4 (1.35) = 28.9 lb/h

$\boxed{\text{28.9 lb/h of 100 psig motive steam required for this ejector}}$

Recombiner evaluation

Now I'll use ChemCad to combine the steam, leaked air, and process vapor into one stream. Then I'll condense the condensibles and vent off the non-condensibles (air, CO_2). Then I'll put it all into the recombiner drum.

Here's the combined steam, air, and process vapor stream:

Stream No.	22 on PFD ~~14~~
Stream Name	
Temp F	184.2442
Pres psia	15.7000
Enth MMBtu/h	-0.362330
Vapor mole fraction	0.988253
Total lbmol/h	3.1151
Total lb/h	90.4999
Total std L ft3/hr	1.6502
Total std V scfh	1182.12
Flowrates in lb/h	
Water	29.7268
Ethanol	4.1930
Carbon Dioxide	47.4039
Pyruvic Acid	0.0000
Ethyl Acetate	0.8662
Air	8.3100

I need to condense everything but the CO_2 and air if possible.
The simulated result is

Inlet temp = 184.24 °F

Outlet temp = 68.78 °F

Calc. Duty = - 0.035 mmBtu/h

$$\text{Actual duty} = -0.035\left(\frac{0.2(98.43) + 90.4999}{90.4999}\right) = -0.04261 \text{ mmBtu/h}$$

this takes into account the volatiles that CHEMCAD can't take care of

Stream No.	24 on PFD ~~19~~	7 on PFD ~~20~~
Stream Name		
Temp F	68.7844	68.7844
Pres psia	15.7000	15.7000
Enth MMBtu/h	-0.179704	-0.217626
Vapor mole fraction	1.0000	0.00000
Total lbmol/h	1.3661	1.7490
Total lb/h	55.0971	35.4028
Total std L ft3/hr	1.0603	0.5899
Total std V scfh	518.39	663.73
Flowrates in lb/h		
Water	0.5237	29.2031
Ethanol	0.5603	3.6327
Carbon Dioxide	45.6440	1.7599
Pyruvic Acid	0.0000	0.0000
Ethyl Acetate	0.0592	0.8070
Air	8.3099	0.0001

very little air remains

A heating curve has been constructed

Sizing after flash condenser

Once again, I can use CW to take the T down to 130°F, but no further. This corresponds to $(3.5 \times 10^{-2}$ mm Btu/h $- 0.75 \times 10^{-2}$ mm Btu/h$)$

$= 27500$ Btu/h

From steam tables, $H_{120°} = 58$ Btu/lb
$H_{90°} = 88$ Btu/lb

Solve for mass of CW:

$$27500 \text{ Btu/h} = \left(88 - 58 \frac{\text{Btu}}{\text{lb}}\right) \dot{m}$$

$$\underline{\underline{\dot{m} = 917 \text{ lb/h}}} \text{ of CW required}$$

Once again, a heating curve has been constructed

Sizing vent drum

The mass flowrate of the liquid to the vent drum is:

$L = 35.4028$ lb/h
$\rho_L = 59.641$ lb/ft^3

Sizing recombiner

The feeds to the recombiner are the column bottoms (stream 10) and the liquids from the vent drum (~~stream 23~~) now stream 7

Stream ~~23~~ 7	Stream 10
T = 68.78 °F	T = 125.8 °F
P = 15.7 psia	P = 7 psia
29.2 lb/h H_2O	9317.41 lb/h H_2O
3.6 lb/h EtOH	7.2 lb/h EtOH
1.76 lb/h CO_2	0 CO_2
0 PA	20.7 lb/h PA
0.8070 lb/h Ethyl acetate	5.2 lb/h Ethyl acetate
0.0001 lb/h Air	—

Recombiner sizing [cont'd]

The combined stream is #11.

Stream 11	
T = 125.6 °F	→ will be pumped to 28 psia
P = 7 psia	
9346.6 lb/h	H_2O
10.8 lb/h	EtOH
1.76 lb/h	CO_2
20.7 lb/h	Pyruvic acid
6.0 lb/h	Ethyl acetate
0.0001 lb/h	Air

The total flow is: $L = 9385.9$ lb/h
$\rho_L = 61.596$ lb/ft^3

E-1 Sizing info

Now I need to get information to size E-1, the feed preheater / product cooler. On the tube side is stream 21 from the recombiner. It will be pumped up to above atmospheric pressure (28 psia) to allow passage through two exchangers to the storage tanks.

On the shell side is the feed to the process, stream 1:

Stream 1	
T = 34°F	← cold from storage
P = 28 psia	← to go through exchangers
9335.1 lb/h	H_2O
414.4 lb/h	EtOH
52.2 lb/h	CO_2
20.7 lb/h	P.A.
20.7 lb/h	Ethyl acetate
19.7 lb/h	non-characterizable volatiles

maximize heat xfer w/o excessive area requirements

I'll simulate the heat exchanger as two different units. I want a minimum 10°F temperature approach. I'll start out by specifying 80°F as the "cold" beer outlet T, record the heat duty, specify this duty for the "warm" beer and see what the temperature is. I'll repeat this process until I get 10°F difference. Actually, I can get an estimate this way:

$$\dot{m}_{cold} C_{p\,cold} \left(T_f - 34\right) = \dot{m}_{warm} C_{p\,warm} \left(125.6 - \left(T_f + 10\right)\right)$$

Since $\dot{m}_{cold} C_{p\,cold} \approx \dot{m}_{warm} C_{p\,warm}$,

$$T_f - 34 = 125.6 - T_f - 10$$

$$T_f \approx 74.8$$

E-1 sizing info [cont'd]

Setting T_f of cold beer equal to 74.8°F I get:

T_f of cold beer (°F)	Heat duty (mmBtu/h)	T of warm beer (°F)
74.8	0.39412	83.4
73.4	0.38062	84.9
74.0	0.38641	84.3

close enough.

As always, heating curves have been constructed.
Note that the T vs. Q plots are almost perfectly straight lines, since the two streams are almost all water

E-2 sizing info

I need to gather info on this exchanger, which heats the feed on up to 80°F using steam. By simulating it and specifying 80°F as the exit temperature, I'll have the heat duty.

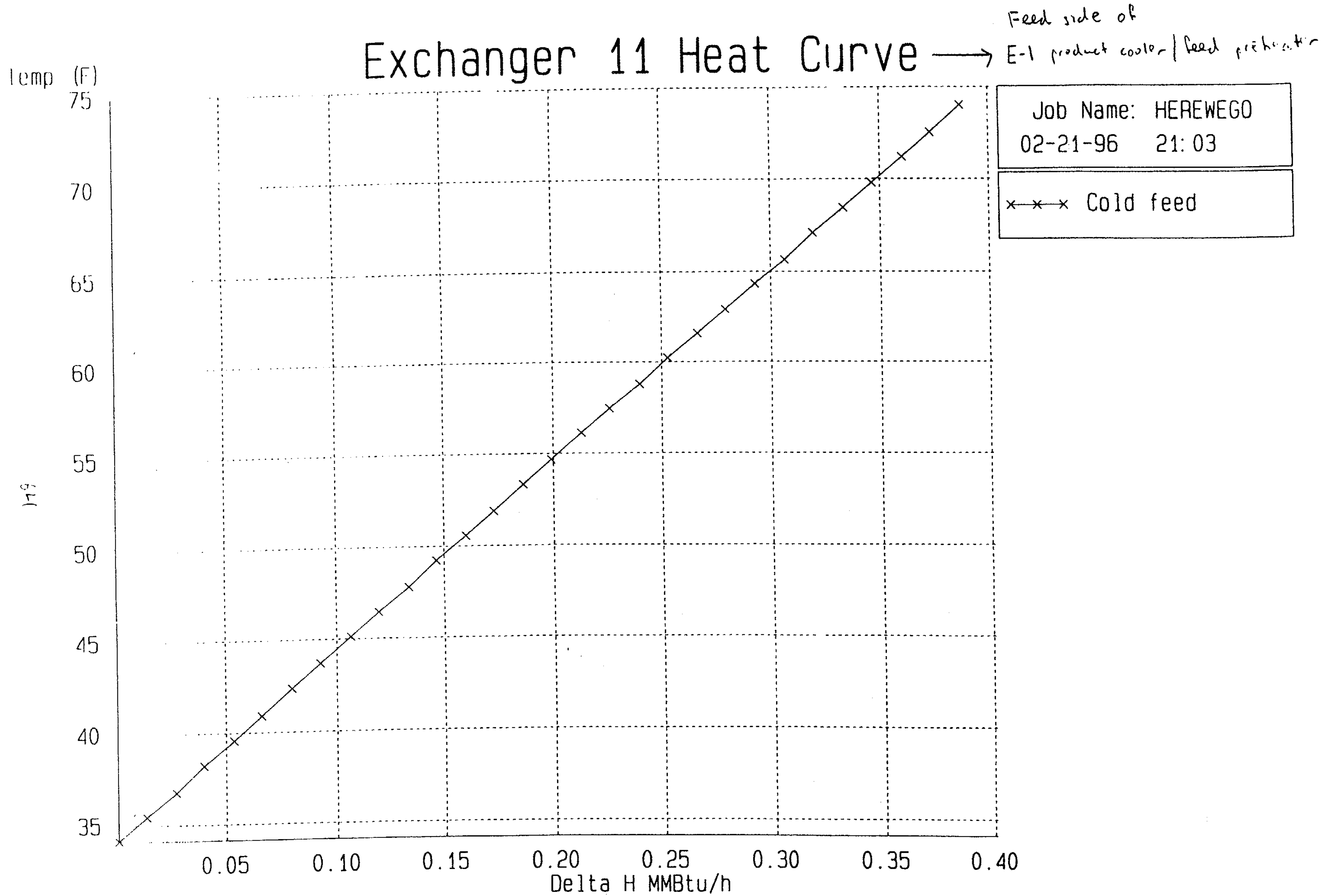
Calculated ChemCad heat duty = 0.057884 mmBtu/h

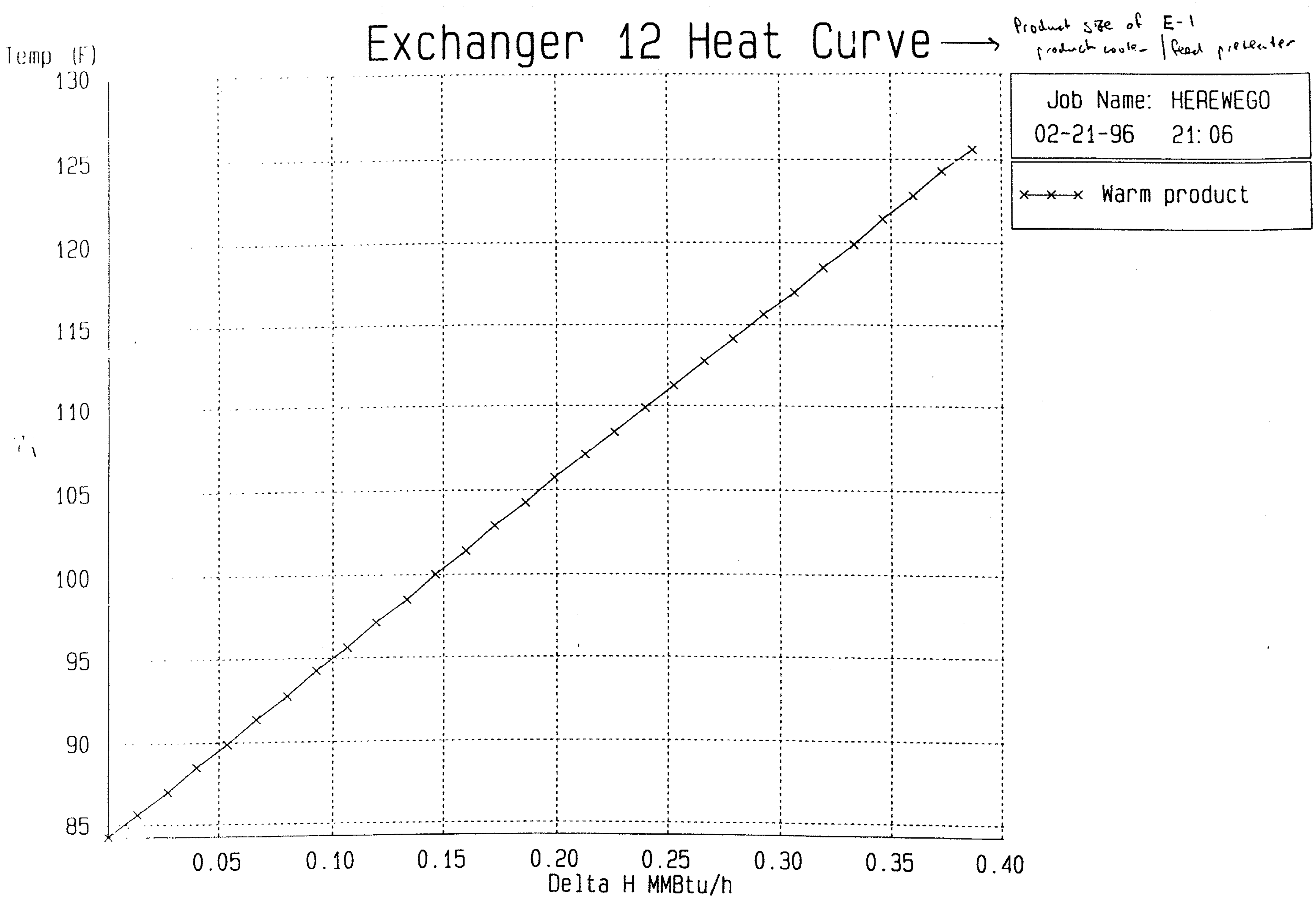
I won't bother to draw a heating curve because I know it will be linear based on the last one.

E-3 sizing info

This exchanger cools the product down the rest of the way to 34°F. I'll simulate it just like the last one.

Calculated ChemCad heat duty = −0.47191 mmBtu/h


Exchanger 11 Heat Curve
Feed side of
E-1 product cooler | feed preheater
Temp (F)
75
70
65
60
55
50
45
40
35
0.05
0.10
0.15
0.20
0.25
0.30
0.35
0.40
Delta H MMBtu/h
Job Name: HEREWEGO
02-21-96 21: 03
Cold feed



Exchanger 12 Heat Curve
Product side of E-1
product cooler | feed preheater
Job Name: HEREWEGO
02-21-96 21:06
Warm product
Temp (F)
130
125
120
115
110
105
100
95
90
85
0.05
0.10
0.15
0.20
0.25
0.30
0.35
0.40
Delta H MMBtu/h

Pump sizing info

Now I need to gather info on my pumps, of which I have six. I'll need to know the specific gravity, volumetric flow rate, and vapor pressure of each stream. For the vapor pressure, I'll assume ideal mixtures for the moment and just use the vapor pressure of ethyl acetate, which is the most volatile (normal boiling point 170.83 °F vs normal bp of EtOH @ 173.03 °F). This vapor pressure is given by Henley and Seader:

$$p^{sat} = 556 \exp\left[6.3307 - \frac{5440.049}{373.48 + T}\right], \quad \text{where } T \,[=]\, {}^\circ F, \; P \,[=]\, \text{psia}$$

The six streams I need info on are 1, 5, 10, 15, 20, and 21

Stream 1: Feed

T = 34° F
Vap. P = 0.50 psia
$\dot{m}$ = 9843 lb/h
ρ = 61.785 lb/ft³
s.g. = 0.988

Stream 5: Flash liquid

T = 79.2 °F
Vap. P = 1.89 psia
$\dot{m}$ = 9790 lb/h
ρ = 61.507 lb/ft³
s.g. = 0.989

Stream 10: Column bottoms

T = 125.8 °F
Vap. P = 5.79 psia
$\dot{m}$ = 18603 lb/h
ρ = 61.613 lb/ft³
s.g. = 1.00

Stream 15: Reflux + distillate

T = 52.6 °F
Vap. P = 0.89 psia
$\dot{m}$ = 15609 lb/h
ρ = 50.239 lb/ft³
s.g. = 0.802

Stream 20: Water recirculation

T = 136 °F
(it's all H_2O) Vap. P = 2.61 psia
$\dot{m}$ = 7.36×10^6 lb/h
ρ = 61.463 lb/ft³
s.g. = 0.987

Stream 21: Recombiner outlet

T = 125.6 °F
Vap. P = 5.76 psia
$\dot{m}$ = 9386 lb/h
ρ = 61.596 lb/ft³
s.g. = 1.00

Actual Sizing Calculations ⟶ And costing too

Now that I have collected all of the information I think I need to size everything, I'll do the actual sizing. I've got all the physical properties now so I can leave the simulator behind (yes!). I'll start with the pumps.

P-1A/B Feed pumps, stream 1 (see info gathered earlier)

Suction P = 14.7 psia
Discharge P = 28 psia (line + exchanger loss)

$$NPSH_A = (\text{suct } P - \text{Vap } P) \times \frac{2.31}{\text{sp. gr.}} = (14.7 - 0.5) \times \frac{2.31}{0.988} = 14.4 \text{ ft} \quad \checkmark \text{OK}$$

$$\text{gpm} = \frac{\text{lb/hr}}{500 \cdot \text{sp. gr.}} = \frac{9843}{500(0.988)} = 19.9 \text{ gpm}$$

$$h = \frac{\Delta P \cdot 2.31}{\text{sp. gr.}} = \frac{13.3(2.31)}{0.988} = 31.0 \text{ ft}$$

From Figure 14-40 of Peters + Timmerhaus, this service requires:

$1\frac{1}{2} \times 1\frac{1}{2}$ pump, $\approx \frac{1}{2}$ hp

Raw cost = \$860

I need to convert the cost to a stainless steel pump in 1999. The SS factor is 1.80, the inflation factor is 1.1646 (see page 9 of economic calculations)

$$1999 \text{ cost} = 860(1.8)(1.1646) = \$1803 \times \overset{\text{spare}}{2} = \boxed{\$3606}$$

$$\text{Operating cost} = 0.5\ \cancel{hp} \cdot 0.746 \frac{kW}{\cancel{hp}} \cdot 8400 \frac{hr}{yr} \cdot \underset{*}{0.0494} \frac{\$}{kWhr} = \boxed{\$154.8/yr}$$

* Note: from a previous course, I estimate the cost of electricity to be about $0.047 \frac{\$}{kWhr}$ in 1997. In 1999, this cost will be

$$0.047 \frac{\$}{kWhr} (1.025)^2 = 0.0494 \frac{\$}{kWhr}$$

P-2A/B, T-1 feed pumps, Stream 5

Suction P = 13.226 psia
Discharge P = 15.226 psia (line loss)

$$NPSH_A = (13.226 - 1.89) \times \frac{2.31}{0.989} = 26.5 \text{ ft} \quad \checkmark OK$$

$$gpm = \frac{9790}{500(0.989)} = 19.8 \text{ gpm}$$

$$h = \frac{2(2.31)}{0.989} = 4.67 \text{ ft}$$

From P+T Figure 14-40 again, this service requires:

$1\frac{1}{4} \times 1\frac{1}{4}$ pump, $\approx \frac{1}{4}$ hp

Raw cost = \$690

Converting to stainless steel, 1999 cost, and a spare,

$$1999 \text{ cost} = 690(1.8)(1.1646)(2) = \boxed{\$2893}$$

Operating cost =

$$0.25 \text{ hp} \cdot 0.746 \frac{kW}{hp} \cdot 8400 \frac{hr}{yr} \cdot 0.0494 \frac{\$}{kWhr} = \boxed{\$77.4/yr}$$

P-3A/B, Column bottoms pumps, Stream 10

Suction P = 2.0 psia
Discharge P = 8 psia (exchanger + line loss)

$$NPSH_A = (2.0 - 5.79) \times \frac{2.31}{1.0} = -8.8 \text{ ft.}$$

This is a potential problem. I need to make sure there is at least 10 ft between the column exit and the pump inlet. Since most columns have a skirt (mine will), this shouldn't be a problem.

$$gpm = \frac{18603}{500(1.0)} = 37.2 \text{ gpm}$$

$$h = 6(2.31)/1.0 = 13.86 \text{ ft}$$

P-3A/B [cont'd]

From P+T Figure 14-40 again, this service requires:

$1\frac{1}{2} \times 1\frac{1}{2}$ pump, $\approx \frac{1}{2}$ hp

Since this is the same as P-1A/B, I know the costs are:

1999 cost = 3606 (1.05)
↑
Temperature factor

$\boxed{\text{1999 cost} = \$3786}$

$\boxed{\text{Operating cost} = \$154.8/\text{yr}}$

P-4A/B Reflux + distillate pumps, Stream 15

Suction P = 0.5 psia
Discharge P = 3.5 psia (line loss)

$$NPSH_A = (0.5 - 0.89) \times \frac{2.31}{0.802} = -1.1 \text{ ft}$$

I just need to ensure that this pump is about 2 ft below the reflux drum. Shouldn't be a problem.

$$\text{gpm} = \frac{15609}{500 \cdot 0.802} = 38.9 \text{ gpm}$$

$$h = 3.0(2.31)/0.802 = 8.6 \text{ ft}$$

From P+T Figure 14-40, this service requires:

$1\frac{1}{2} \times 1\frac{1}{2}$ pump, $\approx \frac{1}{2}$ hp (I'll go up to the next largest)

Raw cost = \$890

1999 cost = 890(1.8)(1.1646)(2) = $\boxed{\$3731}$

For ½ hp, already calculated operating cost = $\boxed{\$154.8/\text{yr}}$

P-5A/B, Hot H_2O recirculation pumps, Stream 20

Suction P = 14.7 psia
Discharge P = 31 psia (3 exchangers + line loss)

$$NPSH_A = (14.7 - 2.61) \times \frac{2.31}{0.987} = 28.3 \text{ ft}$$ ✓ OK

$$gpm = \frac{1.36 \times 10^6}{500(0.987)} = 2749 \text{ gpm}$$

$$h = (31 - 14.7)\, 2.31 \,/\, 0.987 = 38.1 \text{ ft}$$

From P+T Figure 14-40 again, this service requires:

8 × 6 in pump, ≈ 26 hp

Raw cost = \$4150

1999 Cost = 4150 (1.8)(1.1646)(2)(1.05)

stainless steel, inflation, spare, temperature

1999 cost = $\boxed{\$18269}$

Operating cost = $\boxed{\$8049 / yr}$

* see next page

P-6A/B Product pumps, Stream 21

Suction P = 7 psia
Discharge P = 28 psia (2 exchangers + line loss, above atmosphere for storage)

$$NPSH_A = (7 - 5.76) \times \frac{2.31}{1.00} = 2.9 \text{ ft}$$ ✓ OK

$$gpm = \frac{9386}{500(1.0)} = 18.8 \text{ gpm}$$

$$h = 21(2.31)/1.0 = 48.5 \text{ ft}$$

* P-5A|B Horsepower calculation

I've discovered a mistake and I need to calculate the horsepower requirements of P-5A|B. I can get an estimate from the classical thermodynamic equation:

I can't read it off the plot like the others

$$dH = V\,dP$$

$$\Delta H = W_s \approx V\,\Delta P$$

So,

$$\text{hp required} = 2749\ \frac{gal}{min} \cdot 16.3\ \frac{lbf}{\cancel{in^2}} \cdot 144\ \frac{\cancel{in^2}}{\cancel{ft^2}} \cdot \frac{1\ ft^3}{7.48\ gal} \cdot \frac{1\ hp}{33000\ \frac{ft\cdot lbf}{min}}$$

$$= 26\ hp$$

Operating cost =

$$26\ \cancel{hp} \cdot 0.746\ \frac{\cancel{kW}}{\cancel{hp}} \cdot 8400\ \frac{hr}{yr} \cdot 0.0494\ \frac{\$}{\cancel{kWhr}}$$

$$= \$8049/yr$$

P-6A/B [cont'd]

From P+T Figure 14-40 again, this service requires:

1½ × 1½ pump, ≈ ½ hp

Raw cost = \$1220

1999 cost = \$1220 (1.8)(1.1646)(2)(1.05) $\boxed{\$5371}$

As for P-1A/B, Operating cost = $\boxed{\$154.8/yr}$

P-7A/B, Distillate product pumps, Stream 18

These pumps were omitted from the original flowsheet, but are necessary to pump the product to storage. The reason that P-4A/B doesn't do this is that I don't want the reflux pressure to be too high (if I do, the reflux will flash upon entering the column).

Suction P = 2.5 psia

Discharge P = 17 psia (line loss, reach storage slightly above atmospheric)

$$NPSH_A = (2.5 - 0.89) \times \frac{2.31}{0.802} = 4.6 \text{ ft} \quad \checkmark OK$$

$$gpm = \frac{434}{500(0.802)} = 1.1 \text{ gpm}$$

$$h = (17 - 2.5)\, 2.31 / 0.802 = 41.8 \text{ ft}$$

From P+T Figure 14-40, this service requires:

1½ × 1½ pump, ≈ ½ hp (go to next one up)

Raw cost = \$1220

As calculated for P-6A/B, 1999 cost = $\boxed{\$5371}$

Operating cost = $\boxed{\$154.8/yr}$

P-8A/B Protecte hot H_2O circulation pumps, Stream 25

I just realize that I needed to add this pump. The relevant information is:

Stream 25	
T = 120°F	
Vap. P =	1.693 psia
$\dot{m}$ =	5794 lb/h
ρ =	61.73 lb/ft³
s.g. =	0.99

Suction P = 14.7 psia
Discharge P = 26 psia (2 exchangers + line loss)

$$NPSH_A = (14.7 - 1.693) \times \frac{2.31}{0.99} = 30.3 \text{ ft}$$ ✓ OK

$$gpm = \frac{5794}{500(0.99)} = 11.7 \text{ gpm}$$

$$h = (26-14.7)(2.31) / 0.99 = 26.4 \text{ ft}$$

From P+T Figure 14-40, this service requires:

1½ × 1½ pump, ≈ ½ hp

Raw cost = \$860

1999 cost = 860 (1.8)(1.1646)(2)(1.05) = $\boxed{\$3786}$

high T factor

From P-1A/B, Operating cost = $\boxed{\$154.8/yr}$

Driver costs

I also need to determine the costs of the drivers for my 8 pump pairs. Each pump will have its own driver, so with spares, I need 16 drivers. Based on my previous calculations, I need:

12 1/2 hp drivers
2 1/4 hp drivers
2 26 hp drivers

From Figure 14-54 of Peters and Timmerhaus, using alternating current, open, drip proof pump motors, I estimate that the raw cost for both 1/2 + 1/4 hp drivers will be:

Raw cost = \$170 (Lowest cost on figure)

For 14 of them in 1999, the total cost will be:

Small driver costs = 14(\$170)(1.1646)

= \$2772

For the 18 hp driver, from Fig 14-53,

Raw cost = \$1150

1999 cost = (\$1150)(1.1646)(2) = \$2679

Total driver cost = \$5451

Process Vessel Sizing and Costing

Now that the pumps are done, I'll size the drums, of which there are four.

F-2 Reflux drum, Streams 14 + 15

The flow to this vessel (stream 14) is

$$L = 15609 \text{ lb/h}$$
$$(\rho_L = 50.243 \text{ lb/ft}^3)$$

I want a 1/2 full vessel with 5 minute residence time, so I'll use eq. 13-9 of Henley and Seader:

$$V_v = 2Lt / \rho_L$$

$$= \frac{2(15609 \text{ lb/hr})(5 \text{ min})\left(\frac{1 \text{ hr}}{60 \text{ min}}\right)}{50.243 \text{ lb/ft}^3}$$

$$= 51.8 \text{ ft}^3$$

I also want a length that is 4 times the diameter, so:

$$V_v = \frac{\pi}{4} D^2 H = \pi D^3$$

$$\rightarrow D = \frac{H}{4} = \left(\frac{V_v}{\pi}\right)^{1/3}$$

$$= \left(\frac{51.8 \text{ ft}^3}{\pi}\right)^{1/3}$$

$$= 2.54 \text{ ft} \longrightarrow \text{Round to nearest 1/2 foot} = 3 \text{ ft}$$

So $H \sim 10.16$ ft $\longrightarrow$ Round to nearest 1/2 foot = 10.5 ft

Thus, this vessel should be:

10.5 ft in length
3 ft in diameter
Actual volume = 74 ft³

74 ft³ · 7.48 gal/ft³ = 553.5 gal

F-2 [cont'd]

Using Figure 14-56 of Peters and Timmerhaus, the raw cost of this vessel is:

Raw cost = \$22,100 (304 stainless steel "storage" tank. Although it's not really a storage tank, it's the same sort of vessel)

With inflation

$$1999 \text{ Cost} = 22100\,(1.1646) = \boxed{\$25738}$$

F-1 De-esterizer, Streams 5 and 6

The de-esterizer has already been sized earlier in these calculations.

18 ft in height
4.5 ft in diameter
Volume = 286 ft^3 = 2141 gal

Again from Peters + Timmerhaus Fig. 14-56, using curve for 304 stainless steel storage tank,

Raw cost = \$12000

$$1999 \text{ cost} = 12000\,(1.1646) = \boxed{\$13975}$$

F-4 Vent drum, Streams 7, 23, + 24

The flow to this vessel is $L = 35.4$ lb/h

$\rho_L = 59.6$ lb/ft^3

Again allowing for 1/2 full vessel, 5 min residence time,

$$V_v = \left(35.4\,\text{lb/h}\right) 2 \left(5\text{ min}\right)\left(\frac{1\text{ hr}}{60\text{ min}}\right) \Big/ 59.6\text{ lb/ft}^3$$

$$= 0.1\ ft^3$$

$\Longrightarrow$ This is really small. I probably don't need this vessel at all. I can just attach my vent line directly to the after flash condenser (E-8)

F-3 Recombiner, Streams 10, 23, + 21

The combined flow rate to the recombiner is

$$L = 9385.9 \text{ lb/h}$$
$$\rho_L = 61.596 \text{ lb/ft}^3$$

Using the same procedure as before,

$$V_V = 2(9385.9 \text{ lb/h})(5 \text{ min})\left(\frac{1 \text{ hr}}{60 \text{ min}}\right) \Big/ 61.596 \text{ lb/ft}^3$$
$$= 25.4 \text{ ft}^3$$

So,
$$D = \frac{H}{4} = \left(\frac{25.4 \text{ ft}^3}{\pi}\right)^{1/3} = 2.00 \text{ ft}$$
$$\Rightarrow H = 8 \text{ ft}$$

So, this vessel must be:

8 ft in height 2 ft in diameter Volume = 25.4 ft³ = 190 gal

Using Fig 14-56 of P+T for 304 stainless steel mixing tank with agitator,

Raw Cost = \$7000

1999 Cost = \$7000 (1.1646) = $\boxed{\$8152}$

T-1 Alcohol Removal Tower Sizing + Costing

These calculations are a little bit more involved. Information was gathered *earlier for different locations in the column. I will be following the procedures exposed by Henley + Seader, Equilibrium-Stage Separation Operations in Chemical Engineering, Wiley, New York, 1981, pp. 505-510.

The first step in sizing the column is to determine the Sondes and Brown velocity parameter C in the equation

$$U_f = C\left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{1/2} \qquad [13\text{-}3 \text{ of } H+S]$$

where U_f [=] flooding velocity, ft/s
ρ_L [=] liquid density, lb/ft^3
ρ_V [=] vapor density, lb/ft^3

The parameter C is further defined by the empirical relationship

$$C = F_{ST} F_F F_H C_F \qquad [13\text{-}5 \text{ of } H+S]$$

where F_{ST} [=] surface tension factor $= (\sigma/20)^{0.2}$
where σ = surface tension in dyne/cm

F_F [=] foaming factor

$$F_{HA} [=] \begin{cases} 1.0 & \text{for } A_h/A_a \geq 0.10 \\ 5(A_h/A_a) + 0.5 & \text{for } 0.06 \leq \frac{A_h}{A_a} \leq 0.1 \end{cases}$$

where A_h = hole area
A_a = tray active area

C_F [=] read from Fig 13-3 of Henley + Seader

The factor C_F is read as the ordinate of Henley and Seader's Figure 13.3, on which the abscissa is.

$$F_{LV} = \frac{LM_L}{VM_V}\left(\frac{\rho_V}{\rho_L}\right)^{0.5} \qquad [\text{Fig } 13.3 \text{ of } H+S]$$

* See the first page of the "Sizing Information" section

T-1 Sizing [cont'd]

where ρ_V and ρ_L are as before and

L [=] liquid molar flow rate, lbmol/hr
M_L [=] liquid molecular weight

V [=] vapor molar flow rate, lbmol/hr
M_V [=] vapor molecular weight

Once the parameter C is calculated, the flooding velocity is calculated from Eq. 13-3 and the diameter required is calculated according to:

$$D_T = \left[\frac{4VM_V}{(Fl\%)\,U_f\,\pi\,(1 - A_d/A)\,\rho_V}\right]^{0.5} \qquad [13\text{-}7 \text{ of } H+S]$$

where $Fl\%$ [=] percent of flooding, usually 85%

$\frac{A_d}{A}$ [=] ratio of downcomer area to column cross-sectional area, estimated by:

$$\frac{A_d}{A} = \begin{cases} 0.1; & F_{LV} \le 0.1 \\ 0.1 + \dfrac{F_{LV} - 0.1}{9}; & 0.1 \le F_{LV} \le 1.0 \\ 0.2; & F_{LV} \ge 1.0 \end{cases}$$

Procedure: I will follow this procedure at five different points along the column: trays 1, 15, 30, 45, and 59. Then I'll take the maximum calculated diameter and use it as the column diameter.

Assumptions: I plan to use sieve trays with 24" tray spacing. Also, I'll assume a foaming factor of 0.6 (where a non-foaming liquid = 1.0). Most of the foaming agents were removed in the de-esterizer, so this is a bit conservative. However, C.O. Owner did state explicitly to "be conservative in producing a quality product."

Also: My information comes from my earlier simulations based on 100 lb/h of feed. To convert to actual flow rates, I need to multiply by a scale factor of 98.43.

And: Additionally, I'll start out with the standard 85% of flood

T-1 Sizing [cont'd]

Tray 1 Calculations

First, calculate the various factors:

$$F_{ST} = (66.816/20)^{0.2} = 1.273$$

$F_F = 0.6$, as noted earlier

$F_{HA} = 1.0$ (I'll assume that I'm using trays for which $A_h/A \geq 0.10$)

$$F_{LV} = \frac{176 \text{ lb/h } (\cancel{98.43})}{82 \text{ lb/h } (\cancel{98.43})}\left(\frac{0.00581 \text{ lb/ft}^3}{61.51 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 0.021$$

from Fig. 13-3 of H + S, $C_F = 0.39$ ft/s

Thus, from eq. 13-5,

$$C = (1.273)(0.6)(1.0)(0.39 \text{ ft/s})$$
$$= 0.298 \text{ ft/s}$$

And from eq. 13-3,

$$U_f = 0.298 \text{ ft/s}\left(\frac{61.51 - 0.00581 \text{ lb/ft}^3}{0.00581 \text{ lb/ft}^3}\right)^{1/2}$$

$$= \underline{30.66 \text{ ft/s}}$$

And since $F_{LV} \leq 0.1$, $A_d/A = 0.1$

Now:

$$D_T = \left[\frac{4\left(82 \frac{\text{lb}}{\text{h}}\right)\left(\frac{1 \text{ h}}{3600 \text{ s}}\right)(98.43)}{0.85\left(30.66 \frac{\text{ft}}{\text{s}}\right)\pi(1-0.1)\left(0.00581 \text{ lb/ft}^3\right)}\right]^{0.5}$$

$$D_T = \underline{\underline{4.58 \text{ ft}}}$$

T-1 Sizing [cont'd]

Tray 15 Calculations → same procedure as before

$$F_{ST} = (39.798/20)^{0.2} = 1.148$$

$$F_F = 0.6$$

$$F_{HA} = 1.0$$

$$F_{LV} = \frac{237\ lb/h}{143\ lb/h}\left(\frac{98.43}{98.43}\right)\left(\frac{0.00908\ lb/ft^3}{55.12\ lb/ft^3}\right)^{0.5}$$

$$= 0.021$$

from Fig 13-3 of H+S, $C_F = 0.39\ ft/s$

Now, $C = 1.148(0.6)(1.0)(0.39\ ft/s)$

$$= 0.269\ ft/s$$

Then, $U_f = 0.269\ \frac{ft}{s}\left(\frac{55.12 - 0.00908\ lb/ft^3}{0.00908\ lb/ft^3}\right)^{1/2}$

$$= \underline{20.96\ ft/s}$$

And since $F_{LV} \le 0.1$, $A_d/A = 0.1$

Now:

$$D_T = \left[\frac{4(143\ lb/hr)\left(\frac{1\ hr}{3600\ s}\right)(98.43)}{0.85(20.96\ ft/s)\pi(1-0.1)(0.00908\ lb/ft^3)}\right]^{0.5}$$

$$\underline{\underline{D_T = 5.85\ ft}}$$

T-1 Sizing [cont'd]

Tray 30 Calculations → same procedure as before

$$F_{ST} = (39.682/20)^{0.2} = 1.147$$

$$F_F = 0.6$$

$$F_{HA} = 1.0$$

$$F_{LV} = \frac{239 \text{ lb/h}}{144 \text{ lb/h}} \left(\frac{98.43}{98.43}\right) \left(\frac{0.00722 \text{ lb/ft}^3}{55.11 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 0.019$$

from Fig. 13-3 of H+S, $C_F = 0.39$ ft/s

Now, $C = 1.147(0.6)(1.0)(0.39 \text{ ft/s})$

$= 0.268$ ft/s

Next, $U_f = 0.268 \frac{ft}{s} \left(\frac{39.682 - 0.00722 \text{ lb/ft}^3}{0.00722 \text{ lb/ft}^3}\right)^{1/2}$

$= \underline{19.87 \text{ ft/s}}$

Since $F_{LV} \leq 0.1$, $A_d/A = 0.1$

Now:

$$D_T = \left[\frac{4(144 \text{ lb/h})\left(\frac{1 h}{3600 s}\right)(98.43)}{0.85\left(19.87 \frac{ft}{s}\right)(1-0.1)\pi\left(0.00722 \text{ lb/ft}^3\right)}\right]^{1/2}$$

$$\underline{\underline{D_T = 6.76 \text{ ft}}}$$

T-1 Sizing [cont'd]

Tray 45 Calculations → same procedure as before

$$F_{ST} = (25.284/20)^{0.2} = 1.048$$

$$F_F = 0.6$$

$$F_{HA} = 1.0$$

$$F_{LV} = \frac{157 \text{ lb/h}}{164 \text{ lb/h}} \left(\frac{\cancel{98.43}}{\cancel{98.43}}\right)\left(\frac{0.00604 \text{ lb/ft}^3}{50.51 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 0.010$$

from Fig. 13-3 of H + S, $C_F = 0.39$ ft/s

Now, $$C = 1.048 (0.6)(1.0)(0.39 \text{ ft/s})$$
$$= 0.245 \text{ ft/s}$$

So, $$U_f = 0.245 \frac{\text{ft}}{\text{s}} \left(\frac{50.51 - 0.00604 \text{ lb/ft}^3}{0.00604 \text{ lb/ft}^3}\right)^{1/2}$$

$$\underline{= 22.40 \text{ ft/s}}$$

Since $F_{LV} \leq 0.1$, $A_d/A = 0.1$

Now:

$$D_T = \left[\frac{4\left(164 \text{ lb/h}\right)\left(\frac{1 \text{ h}}{3600 \text{ s}}\right)(98.43)}{0.85\left(22.4 \frac{\text{ft}}{\text{s}}\right)\pi(1-0.1)\left(0.00604 \frac{\text{lb}}{\text{ft}^3}\right)}\right]^{1/2}$$

$$\underline{\underline{D_T = 7.43 \text{ ft}}}$$

T-1 Sizing [cont'd]

Tray 59 Calculations $\longrightarrow$ same procedure as before

$$F_{ST} = \left(24.138 / 20\right)^{0.2} = 1.038$$

$$F_F = 0.6$$

$$F_{HA} = 1.0$$

$$F_{LV} = \frac{172\ \text{lb/h}}{163\ \text{lb/h}} \left(\frac{\cancel{98.43}}{\cancel{98.43}}\right)\left(\frac{0.00395\ \text{lb/ft}^3}{50.26\ \text{lb/ft}^3}\right)^{1/2}$$

$$= 0.009$$

from Fig 13-3 of H & S, $C_F = 0.39$ ft/s

Now,

$$C = 1.038\,(0.6)(1.0)(0.39\ \text{ft/s})$$

$$= 0.243\ \text{ft/s}$$

Thus,

$$U_f = 0.243\ \text{ft/s}\left(\frac{50.26 - 0.00395\ \text{lb/ft}^3}{0.00395\ \text{lb/ft}^3}\right)^{1/2}$$

$$\underline{= 27.41\ \text{ft/s}}$$

Since, $F_{LV} \le 0.1$, $A_d/A = 0.1$

Now:

$$D_T = \left[\frac{4\left(163\ \text{lb/h}\right)\left(\frac{1\ \text{h}}{3600\ \text{s}}\right)(98.43)}{0.85\left(27.41\ \frac{\text{ft}}{\text{s}}\right)\pi\,(1-0.1)\left(0.00395\ \frac{\text{lb}}{\text{ft}^3}\right)}\right]^{1/2}$$

$$\underline{\underline{D_T = 8.28\ \text{ft}}}$$

T-1 sizing [cont'd]

Results of Diameter Calculations

Tray	C (ft/s)	U_f (ft/s)	D_T (ft)
1	0.298	30.66	4.58
15	0.269	20.96	5.85
30	0.268	19.87	6.76
45	0.245	22.40	7.43
59	0.243	27.41	8.28

Bottom ↓ Top

Comments and Conclusions: The increasing diameter from the bottom to top of the column is as expected, since at the bottom the pressure is higher and the denser vapor requires less area.

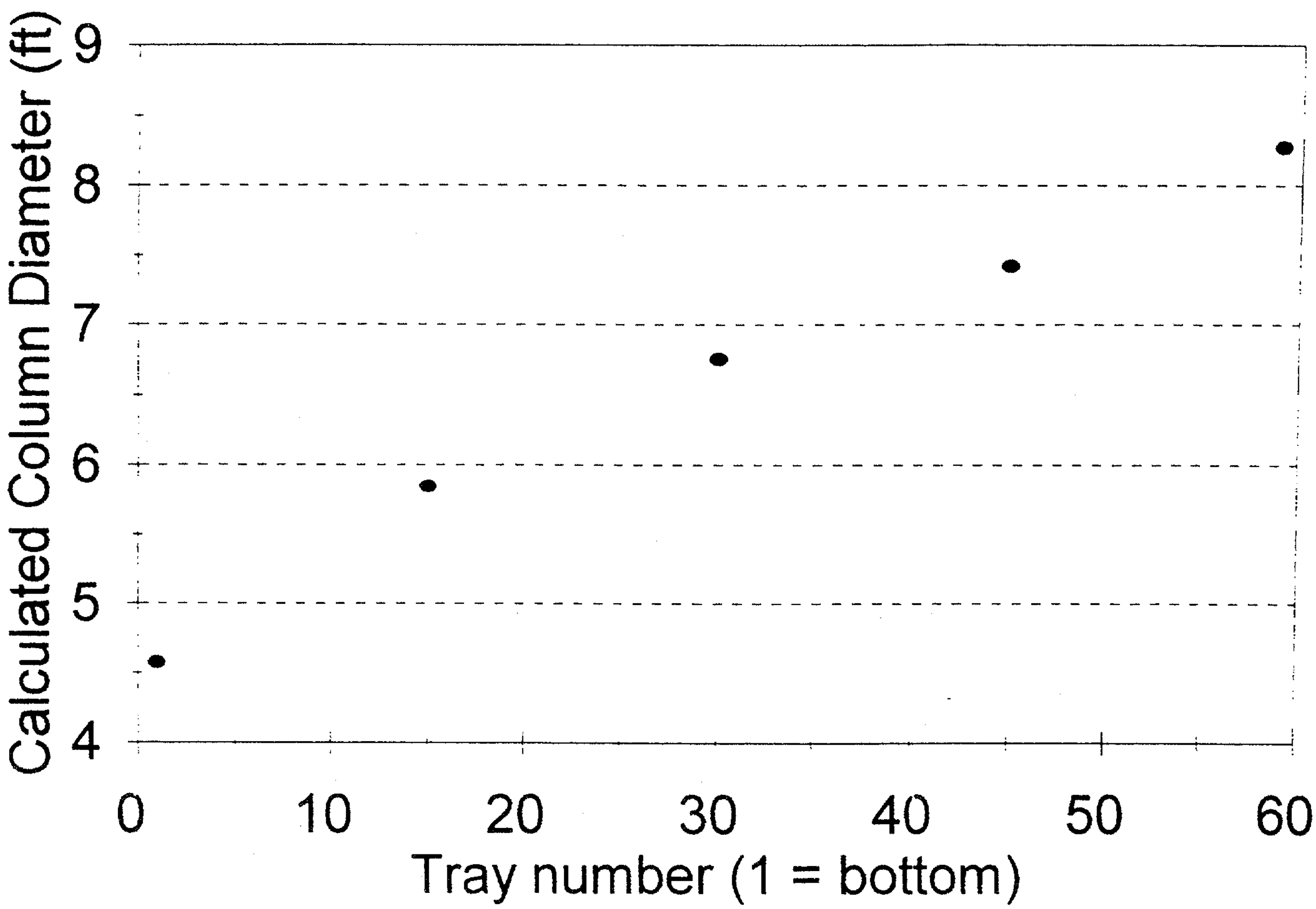
I'm a little concerned, though, that the diameters are not even larger, given that the density of the vapors is so low due to the low overhead pressure. Perhaps the flow rate I am using is too low. I'll find out when I do the optimization.

Sometimes it is possible to have a column with sections of different diameter, but since the change here would only be from 5 ft to 8.5 ft, I won't specify it. Also, having two different diameters would require knowing exactly where to make the switch and I don't feel I know the column dynamics that well.

Therefore, I'll use the largest calculated diameter and specify:

$$\boxed{\text{Column Diameter} = 8.5 \text{ ft}}$$

(round to nearest 1/2 foot)


Calculated Column Diameter (ft)
9
8
7
6
5
4
0
10
20
30
40
50
60
Tray number (1 = bottom)

T-1 Sizing [cont'd]

I also need to know how tall the column will be. Using 24" tray spacing, 6 inches every 20 ft for manways, 6 ft of vapor disengagement space at the top of the column, and 10 ft of reboiler surge space at the bottom, I get:

$$H_T = \underbrace{(59-1)}_{\#\text{ of trays}}(2\text{ ft}) + \underbrace{(59-1)(2\text{ ft})\frac{(0.5\text{ ft})}{20\text{ ft}}}_{\text{For manways}}$$

$$+ \underbrace{10\text{ ft}}_{\text{reboiler surge}} + \underbrace{6\text{ ft}}_{\text{vapor disengagement}}$$

(normally, I'd use 4 ft, but since the potential for foaming exists, I'll use 6 ft)

$H_T = 134.9 \rightsquigarrow$ round to 135 ft

Check aspect ratio: $H_T/D_T = 135/8.5 = 15.9$

✓ OK since ≤ 25

Thus, $\boxed{\text{Column Height} = 135\text{ ft}}$

For a later calculation, I need to know the column volume:

$$V_T = \frac{\pi}{4}(8.5\text{ ft})^2(135\text{ ft}) = \underline{\underline{7661\text{ ft}^3}}$$

T-1 Costing

I'll use Fig. 16-28 of Peters and Timmerhaus to get the cost. I'll use stainless steel sieve trays. The diameter of the tower is 102 inches.

from figure, 1990 cost = \$6000 per foot of height

Since the column is 135 ft tall, the total 1990 cost is:

$$1990 \text{ cost} = \$6000/\text{ft} \cdot 135 \text{ ft} = \$810{,}000$$

In 1999 dollars,

$$1999 \text{ cost} = \$810000 (1.1646)$$

$$\boxed{1999 \text{ Cost} = \$943{,}326}$$

Note that this cost includes installation and auxiliaries, according to Fig. 16-28

Ejector Sizing

Now that I know how big T-1 will be, I have enough information to size the anticipated two ejectors required to maintain the vacuum, X-2 and X-3.

First, I'll calculate the air leakage rate using the equations in reference 4.

The lowest system pressure is 0.5 psia

$$0.5 \text{ psia} \cdot \frac{760 \text{ mm Hg}}{14.7 \text{ psia}} = 25.85 \text{ mm Hg}$$

The coefficients are

$$A = 0.1451$$
$$B = 0.6617$$

The system volume (previously calculated) is 7661 ft^3
Thus,

$$MAL = 0.1451 (7661)^{0.6617}$$
$$= \underline{53.9 \text{ lb/h}}$$

According to reference 3, single stage jets can be used down to 75 torr (= 1.45 psia). Thus the reason for two stages.

In addition to the air leakage, there is some CO_2. From my most recent simulation, the amount of CO_2 is

195.3 lb/h of CO_2

The MW of CO_2 = 44, so the factor from Fig. 9-8, reference 3 = 1.18
The temperature conversion factor from Fig 9-7 is 1.0
The air equivalent load is:

$$AEL = \frac{53.9 \text{ lb/h}}{(1.0)(1.0)} + \frac{195.3 \text{ lb/h}}{(1.0)(1.18)} = 219.4 \text{ lb/h}$$

From Fig. 9-11, the r value (25.85 torr, 2 stages with intercondensers, 90°F CW) is:

$$r \approx 10 \text{ lb steam / lb air}$$

Ejector Sizing [cont'd] + Costing

Now, using eq. 9-5 of reference 3,

$$w_s = (AEL)(r)(MPC)(SC)$$

$MPC = 1.0$ (using 100 psig steam already)
$SC = 1.0$ assumed for time being.

$$w_s = (219.4 \text{ lb/hr})(10)(1.0)(1.0)$$

$$= 2194 \text{ lb/hr}$$

The average steam consumption per stage is

$$2194/2 = 1097 \text{ lb/hr per stage}$$

From Figure 9-10, size correction factor = 1.00

Thus, I will specify:

Two-stage ejector system (X-2 + X-3)
1097 lb/hr 100 psig steam per stage
2194 lb/hr 100 psig steam total

Recall from previous calculations that the other ejector in the process (X-1) required:

Single stage (X-1)
28.9 lb/h 100 psig motive steam

Cost of Ejectors

I will use the cost correlation given as equation 8-1 in reference 3 along with all the parameters associated with it. This book was published in 1986, so I need an appropriate inflation factor:

$$I.F. = \left(\frac{368}{325.3}\right)(1.025)^5 = 1.2799$$

...s of Chemical Engineering Index ... 1994 to 1985

...flation to update to 1999

Ejector Costing [cont'd]

The sizing equation 8-1 from reference 3 is:

$$\text{Installed cost} = \$16{,}000\left[NS + 2(NC)\right]\left(\frac{SCON}{1000}\right)^{0.35}$$

where NS [=] # of ejector stages
NC [=] # of condensers
SCON [=] steam consumption based on 100 psig motive steam, lb/hr

For X-1 :
→ I have already taken care of the condenser elsewhere, so I'll do this calculation as though there is no condenser.

$$1986 \text{ cost} = \$16000\left[1 + 2(0)\right]\left(\frac{28.9}{1000}\right)^{0.35} = \$4628$$

$$1999 \text{ cost} = \$4628\,(1.1)(1.2799)$$

stainless steel factor as given by reference 3 — inflation factor

X-1 1999 cost (not including condenser) = \$6516

For X-2 and X-3 :

$$1986 \text{ cost} = \$16000\left[2 + 2(2)\right]\left(\frac{1997}{1000}\right)^{0.35} = \$99162$$

$$1999 \text{ cost} = \$99162\,(1.1)(1.2799)(1.2)$$

stainless steel — Inflation — factor when using surface condensers *

X-2 and X-3 combined 1999 cost (including condensers) = \$167 531

Ejectors [cont'd]

* Note of explanation: It is possible to use direct contact condensers, which are cheaper. However, in most locations the steam condensate must by law be discharged to waste treatment. By using surface condensers, the cooling water will not be contaminated and the volumetric flow to the waste treatment facility will be reduced by approximately 98% (reference 3, p. 227)

Heat Exchanger Sizing + Costing

The final pieces of equipment to size are the heat exchangers, of which I have ten. I have already gathered most of the information (see the "Sizing Information" section).

Procedure: I'll use the classical design equation:

$$Q = Q_{I} + Q_{II} + \cdots = U_I A_I \Delta T_{lm_I} + U_{II} A_{II} \Delta T_{lm_{II}} + \cdots$$

where the Roman numerals refer to regions in which the temperature profiles are linear (the derivation of the ΔT_{lm} equation assumes a linear T vs. Q profile).

I have already generated the necessary T vs. Q curves.

(for cost) Once the exchangers are sized, I'll refer to Figure 15-15 of Peters and Timmerhaus, using fixed tube sheet exchangers as specified in the problem statement.

Assumptions: For these beginning calculations, I'll assume the following general overall heat transfer coefficients:

$$U = \begin{cases} 10 \ \frac{\text{Btu}}{\text{hr ft}^2\ {}^\circ\text{F}} & \text{for gas cooling} \\ 80 \ \frac{\text{Btu}}{\text{hr ft}^2\ {}^\circ\text{F}} & \text{for liquid-liquid sensible} \\ 160 \ \frac{\text{Btu}}{\text{hr ft}^2\ {}^\circ\text{F}} & \text{for condensing} \end{cases}$$

Source: class notes

Also, since product must be stored around 34°F, I'll assume that the site is already equipped with a refrigeration system. Furthermore, since ammonia refrigeration is common, I'll assume it. Since 34°F is the lowest temperature required, I'll assume that the ammonia is at 58 psia, enough pressure to allow for a 5 psia exchanger pressure drop and still have the evaporation temperature be 24°F.* This corresponds to a minimum 10°F temperature approach.

* U.S. Nat. Bur. Stand. Circ. 142 (1923).

Exchanger Sizing + Costing [cont'd]

E-1: Feed preheater / product cooler

Tube side: warm product, streams 21 and 12

$T_{in} = 125.6\,°F \qquad T_{out} = 84.3\,°F$

Shell side: cold feed, streams 2 and 3

$T_{in} = 34\,°F \qquad T_{out} = 74.0\,°F$

From the heating curves constructed for these two streams, it is seen that the temperature profiles are essentially linear. This makes good intuitive sense since these streams are mostly liquid water.

The calculated heat duty is 0.39412 mmBtu/h

For this case, $U = 80$ Btu/(hr ft² °F) and there is only one region. The equation becomes:

$$Q_I = U_I A_I \Delta T_{lm\,I}$$

$$\Rightarrow A_I = \frac{Q_I}{U_I \Delta T_{lm\,I}}$$

First,

$$\Delta T_{lm\,I} = \frac{(125.6 - 74) - (84.3 - 34)}{\ln \frac{125.6 - 74}{84.3 - 34}}$$

$$= 50.95\,°F$$

So,

$$A_I = \frac{0.39412 \times 10^6 \text{ Btu/hr}}{80 \frac{\text{Btu}}{\text{hr ft}^2 \cancel{°F}} \cdot 50.95 \cancel{°F}} = \boxed{96.7 \text{ ft}^2}$$

From P+T Fig. 15-15, Purchased cost = \$4000 (for 304 stainless steel tubes)

[illegible] cost = \$4000 · 1.1646 = $\boxed{\$4658}$

E-2 Feed preheater

Tube side: feed, streams 3 and 4

$T_{in} = 74.0\ °F \qquad T_{out} = 80\ °F$

Shell side: hot water, streams 25 and 26

$T_{in} = 130\ °F \qquad T_{out} = 120\ °F$

Again, the temperature profiles are linear.

The calculated heat duty is 0.057884 mmBtu/h.

Again I'll use $U = 80 \frac{Btu}{hr\ ft^2\ °F}$

Now,

$$\Delta T_{lm} = \frac{(120-74) - (130-80)}{\ln \frac{120-74}{130-80}} = 48.0\ °F$$

So,

$$A_I = \frac{0.057884\ mmBtu/h}{80 \frac{Btu}{hr\ ft^2\ °F} \cdot 48.0\ °F} = \boxed{15.1\ ft^2}$$

From Fig. 15-15 (extrapolating the curve slightly),

Purchased cost = \$2200

1999 cost = \$2200 · 1.1646 = $\boxed{\$2562}$

Now I'll figure out how much water I need to have circulating.
From the steam tables,

$$H^l_{130} = 97.96\ Btu/lbm$$
$$H^l_{120} = 87.97\ Btu/lbm$$

So: $0.057884\ mmBtu/h = \left(97.96 - 87.97 \frac{Btu}{lbm}\right) \dot{m}$

$\dot{m} = 5794\ lb/h$ hot water must be circulated

While I'm here, I might as well size the hot water reheater, E-10

E-10 [illegible] water preheater

Tube side: [illegible], streams 25 and 26

$T_{in} = $ [illegible] °F $\qquad T_{out} = 130\ °F$

E-10 [cont'd]

Shell side : condensing steam @ 100 psig

$T_{in} = 337.9\ °F \qquad T_{out} = 337.9\ °F$

$$\Delta H^{lv} = (1189.6 - 309)\ \text{Btu/lbm}$$

$$= 880.6\ \text{Btu/lbm}$$

Heat duty = 0.057884 mmBtu/h (same as E-2)

Again, there is just one heating zone.
This time, though, $U = 160 \dfrac{\text{Btu}}{\text{hr ft}^2\ °F}$

Now,

$$\Delta T_{lm} = \frac{(337.9-130)-(337.9-120)}{\ln \dfrac{337.9-130}{337.9-120}} = 212.9\ °F$$

So,

$$A_I = \frac{0.057884\ \text{mmBtu/h}}{160\ \dfrac{\text{Btu}}{\text{hr ft}^2\ °F} \cdot 212.9\ °F}$$

$$= 1.7\ \text{ft}^2$$

This is ridiculously small. I anticipate it increasing to a larger value as the capacity of the process increases. I don't think heat exchangers this small are even manufactured, so I'll use the lower limit of Fig. 15-15 of P+T of 10 ft². I'll just have an oversized exchanger.

$\boxed{10\ \text{ft}^2}$

From the figure (extrapolating slightly),

Purchased cost = \$2100

1999 cost = 2100 (1.1646) = $\boxed{\$2446}$

Steam requirement:

$$0.057884\ \frac{\text{mmBtu}}{\text{h}} = \dot{m}\ (880.6\ \text{Btu/lbm})$$

$\Rightarrow$ $\underline{\dot{m} = 65.7\ \text{lb/hr steam required}}$

E-3 Product cooler

Tube side: warm product, streams 12 and 13
$T_{in} = 84.3\,°F$ $T_{out} = 34\,°F$

Shell side: Ammonia refrigerant
$T_{in} = 24\,°F$ $T_{out} = 24\,°F$, $\Delta H^{lv} = 549.8 \frac{Btu}{lbm}$

Once again, the temperature profiles are linear.

For this case, $U = 160 \frac{Btu}{hr\,ft^2\,°F}$

Also, calculated heat duty = 0.47191 mmBtu/h

Now,

$$\Delta T_{lm} = \frac{(84.3 - 24) - (34 - 24)}{\ln \frac{84.3 - 24}{34 - 24}} = 28.0\,°F$$

So,

$$A_T = \frac{0.47191 \text{ mmBtu/hr}}{160 \frac{Btu}{hr\,ft^2\,°F} \cdot 28.0\,°F} = \boxed{105\ ft^2}$$

From P+T figure 15-15,

Purchased cost = \$4000

1999 cost = 4000(1.1646) = $\boxed{\$4658}$

Now I'll determine how much NH_3 I need to circulate:

$$0.47191 \text{ mmBtu/hr} = 549.8 \frac{Btu}{lbm} \cdot \dot{m}$$

$\Rightarrow$ $\dot{m} = 858$ lb/hr NH_3 must be circulated

E-4 Column Condenser

Tube side: Column overhead (less non-condensibles), streams 9 + 14

T_{in} = 52.59 °F T_{out} = 52.55 °F

Shell side: Ammonia refrigerant

T_{in} = 24°F T_{out} : 24°F, $\Delta H^{lv} = 549.8 \frac{Btu}{lbm}$

Calculated heat duty = $0.06932 \times 98.43 = 6.8232 \frac{mmBtu}{hr}$

scale factor

Again, $U = 160 \frac{Btu}{hr\ ft^2\ °F}$

This time, the temperature profile is not linear. I'll break it into 6 sections which are linear. However, I have just realized what the scale on the ordinate of my heat curve is and see now that there is no reason to do so. One region will suffice.

$$\Delta T_{lm} = \frac{(52.59-24)-(52.55-24)}{\ln \frac{52.59-24}{52.55-24}} = 28.6\ °F$$

So,

$$A_I = \frac{6.8232\ mmBtu/hr}{160 \frac{Btu}{hr\ ft^2\ °F} \cdot 28.6\ °F} = \boxed{1491\ ft^2}$$

From Fig. 15-15 of P + T,

Purchased cost = \$20000

1999 cost = 20000 (1.1646) = $\boxed{\$23292}$

The amount of ammonia required is given by:

$$6.8232\ mmBtu/hr = 549.8 \frac{Btu}{hr}\ \dot{m}$$

$\Rightarrow$ $\dot{m}$ = 12410 lb/hr NH_3 circulation required

E-5 and E-6, Ejector Intercondenser and Aftercondenser

Since the two ejector stages are assumed to be exactly the same, the condensers will be the same.

Tube side: Cooling water
$T_{in} = 90°F$ $T_{out} = 120°F$

Shell side: Condensing steam
$T_{in} = 337.9°F$ $T_{out} = 337.9°F$, $\Delta H^{lv} = 880.6 \frac{Btu}{lbm}$

Flow rate = 289.5 lb/h

Here $U = 160 \frac{Btu}{hr\ ft^2\ °F}$

The temperature profile is linear, so

$$\Delta T_{lm} = \frac{(337.9-120)-(337.9-90)}{\ln\frac{337.9-120}{337.9-90}} = 232.6°F$$

Now,

$$A_I = \frac{\left(1097 \frac{lb}{h}\right)\left(880.6 \frac{Btu}{lb}\right)}{160 \frac{Btu}{hr\ ft^2\ °F} \cdot 232.6°F} = 26\ ft^2$$

So:

$$\boxed{2 \text{ exchangers, } 26\ ft^2 \text{ each}}$$

* Note: The cost of these exchangers was already included in the cost of the ejectors.

Now, I'll find the amount of cooling water required.
From steam tables, $H^l_{90} = 58.02$ Btu/lbm
$H^l_{120} = 87.97$ Btu/lbm

So:

$$(1097\ lb/h)(880.6\ Btu/lb) = \left(87.97 - 58.02 \frac{Btu}{lb}\right)\dot{m}$$

$\Rightarrow$ $\dot{m} = 32254$ lb/h CW required for each

E-7A/B Reboilers

Tube side: T-1 bottoms, streams 19 + 27
$T_{in} = 125.8\ °F$ $\quad T_{out} = 126.1\ °F$

Shell side: Hot water, streams 20 + 28
$T_{in} = 150\ °F$ $\quad T_{out} = 136\ °F$

Calculated heat duty = 18.99 mmBtu/h

Here $U = 80 \ \frac{Btu}{hr\ ft^2\ °F}$

Upon first inspection, the temperature profile does not appear to be linear. However, note that the ordinate on the graph covers a range of only 0.3 °F. Therefore, I will go ahead and use the linear temperature profile assumption.

Now,

$$\Delta T_{lm} = \frac{(136 - 125.8) - (150 - 126.1)}{\ln \frac{136-125.8}{150-126.1}} = 16.1\ °F$$

So,

$$A_I = \frac{18.99 \text{ mmBtu/hr}}{80 \frac{Btu}{hr\ ft^2\ °F} \cdot 16.1\ °F} = 14744\ ft^2$$

Maximum feasible area is 8000 ft^2 (previous class)

Split into two exchangers

$$\boxed{\text{2 exchangers, } 7372\ ft^2 \text{ each}}$$

From Fig 15-15 of P+T,

Purchased cost = \$81000 each

1999 cost = 81000 (1.1646)(2) = $\boxed{\$188,665 \text{ total}}$

Calculate H_2O requirements: $H^l_{150} = 118$ Btu/lb
$H^l_{136} = 104$ Btu/lb

18.99×10^6 Btu/h = (118 − 104 Btu/lb) $\dot{m}$ ⟹ $\dot{m} = 1.36 \times 10^6$ lb/h

water circulation required

E-8A After flash condenser (CW section)

Tube side: Cooling water
T_{in} = 90 °F T_{out} = 120 °F

Shell side: Condensing flash vapor, streams 7 + 22
T_{in} = 184.21 °F T_{out} = 130 °F

Calculated heat duty = 33495 Btu/hr

In this case, $U = 160 \frac{Btu}{hr\ ft^2\ °F}$

This time, the temperature profile is not linear. I can break it into 5 sections which are approximately linear (see T vs. Q plot).

Region	Process Side T_{in}	Process Side T_{out}	CW Side T_{in}	CW Side T_{out}	ΔT_{lm}	$Q\left(\frac{Btu}{hr}\right)$	$A(ft^2)$
I	184.2	175.2	90	100.6	84.4	11766	0.87
II	175.2	161.7	100.6	109.8	63.2	10280	1.01
III	161.7	152.7	109.8	113.8	45.4	4409	0.6
IV	152.7	140.3	113.8	117.7	30.6	4415	0.9
V	140.3	130	117.7	120	16.0	2625	1.03

Total area = 4.41 ft²

Again, I don't think that exchangers this small exist, so I'll specify

$\boxed{10\ ft^2}$

From Figure 15-15,

Purchased cost = \$2100

1999 cost = 2100(1.1646) = $\boxed{\$2446}$

Also, from earlier (p.7 of "Sizing Calculations"),

$\dot{m}_{cw}$ = 917 lb/h

E-8B After flash condenser (NH$_3$ section)

Tube side: Condensing flash vapor, streams 7 + 22
$T_{in} = 130°F$ $T_{out} = 68.8 °F$

Shell side: Evaporating ammonia refrigerant
$T_{in} = 24°F$ $T_{out} = 24 °F$, $\Delta H^{lv} = 549.8$ Btu/lb

Calculated heat duty = 9135 Btu/hr

Again, $U = 160 \frac{\text{Btu}}{\text{hr ft}^2 \text{ °F}}$

In this case, the temperature profile may be approximated as linear. I think I'll find out it doesn't make much of a difference.

Now,
$$\Delta T_{lm} = \frac{(130-24) - (68.8-24)}{\ln \frac{130-24}{68.8-24}} = 71.1 °F$$

Then,
$$A_I = \frac{9135 \text{ Btu/hr}}{160 \frac{\text{Btu}}{\text{hr ft}^2 \text{ °F}} \cdot 71.1 °F} = 0.8 \text{ ft}^2$$

exchangers this small don't exist

Again, I'll specify

$\boxed{10 \text{ ft}^2}$

From Fig 15-15 of P+T,

Purchased cost = \$2100

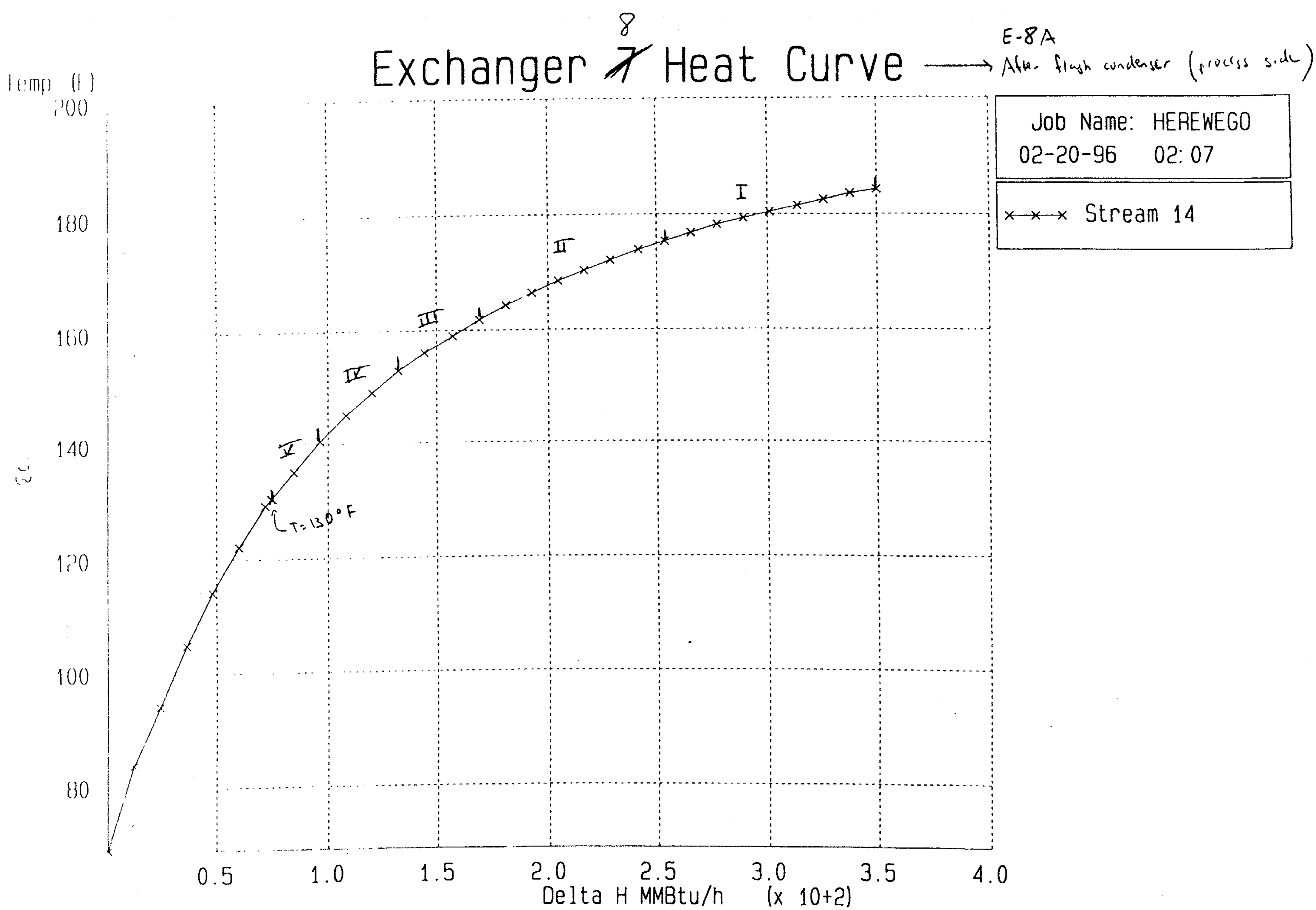
1999 cost = 2100 (1.1646)

= $\boxed{\$2446}$

The amount of NH$_3$ required can be determined from:

$$9135 \text{ Btu/hr} = 549.8 \frac{\text{Btu}}{\text{lb}} \cdot \dot{m}$$

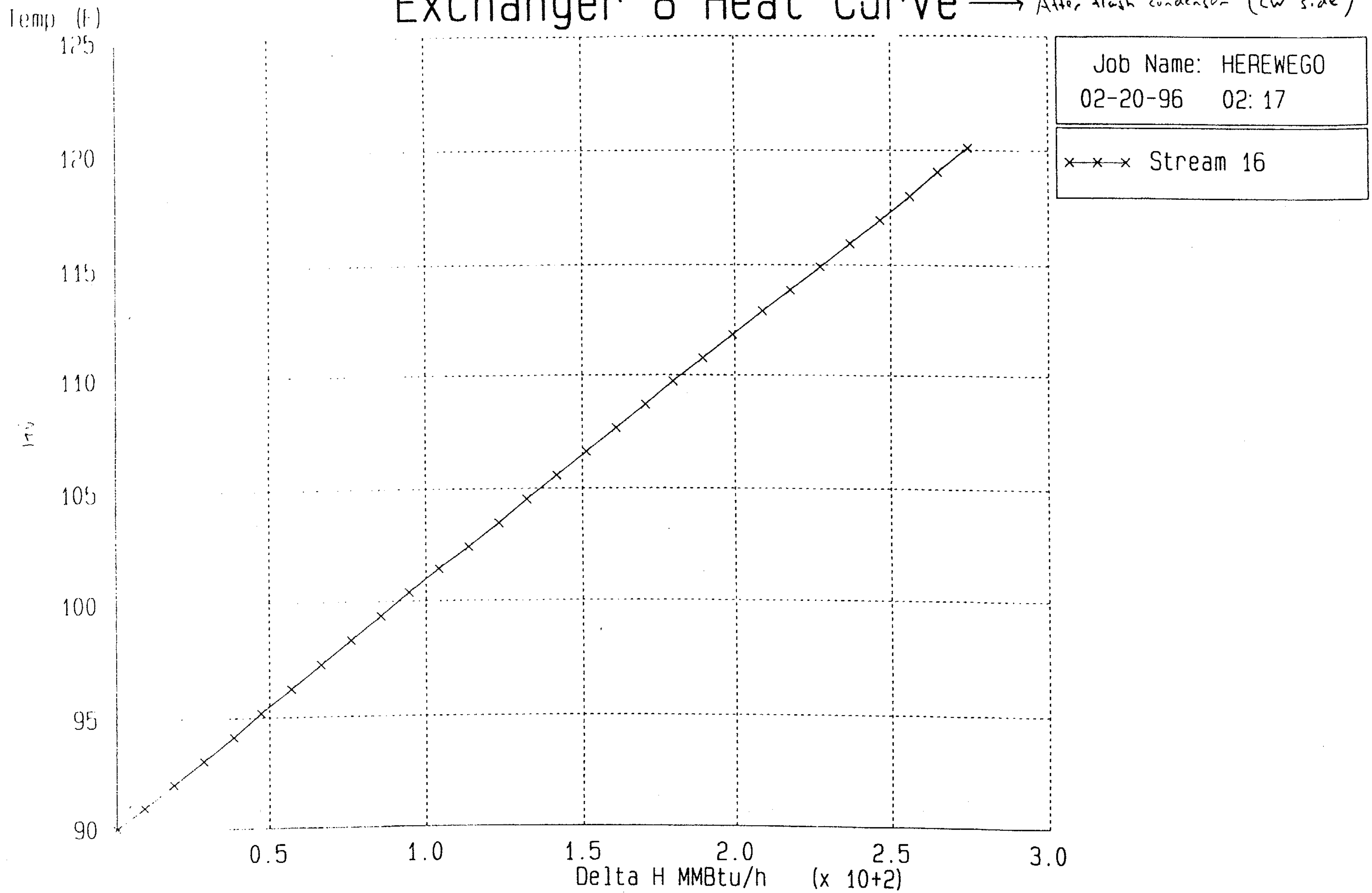
$\Rightarrow \dot{m} =$ 16.6 lb/hr NH$_3$ circulation required



Exchanger ~~7~~ 8 Heat Curve
E-8A
After flash condenser (process side)
Job Name: HEREWEGO
02-20-96 02: 07
Stream 14
Temp (F)
200
180
160
140
120
100
80
0.5
1.0
1.5
2.0
2.5
3.0
3.5
4.0
Delta H MMBtu/h (x 10+2)
I
II
III
IV
V
T=130°F

Exchanger 8 Heat Curve → After flash condenser (cw side)

E-8A



E-9 Reboiler Hot H_2O Reheater

Tube side: heating water, streams 20 and 28
$T_{in} = 136°F$ $T_{out} = 150°F$

Shell side: condensing steam @ 100 psig
$T_{in} = 337.9°F$ $T_{out} = 337.9$

$$\text{Heat duty} = 1.36 \times 10^6 \ lb/h \cdot \left(118 - 104 \ \frac{Btu}{lb}\right)$$

from steam tables @ 136 and 150°F

$$= 19.04 \ mm \ Btu/h$$

Again, $U = 160 \ \frac{Btu}{hr \, ft^2 \, °F}$ and the temperature profile may be considered to be linear. Thus,

$$\Delta T_{lm} = \frac{(337.9 - 150) - (337.9 - 136)}{\ln \frac{337.9 - 150}{337.9 - 136}}$$

$$= 194.8°F$$

$$\text{So, } A_I = \frac{19.04 \times 10^6 \ Btu/hr}{160 \ \frac{Btu}{hr \, ft^2 \, °F} \cdot 194.8 \ °F} = \boxed{611 \ ft^2}$$

From Fig. 15-15 of P + T,
Purchased cost = \$10050

$$1999 \text{ cost} = 10050 \, (1.1646) = \boxed{\$11704}$$

Also, I'll find the required steam flow rate:

$$880.6 \ \frac{Btu}{lb} \ \dot{m} = 1.36 \times 10^6 \ lb/h \ (118 - 104 \ Btu/lb)$$

ΔH^{lv} of 100 psig steam H_2O flowrate ΔH of water

$\Rightarrow \dot{m} = 21621.6$ lb/hr steam required

Storage Tank Sizing & Costing

Purpose: This process will require two storage tanks – one to hold the non-alcoholic beer product and one to hold the distillate ethanol product.

Basis of calculations: I want the tanks to be cylindrical with height = diameter / 3.
I also want them to hold 1 week's worth of product at 90% capacity.

Tank 1: NA beer storage

The flow rate of the NA beer product is that of stream 21,

$$L = 9386 \text{ lb/h}$$
$$\rho_L = 61.596 \text{ lb/ft}^3$$

Thus, for the tank to hold 1 week's worth of product at 90% capacity, the volume will have to be:

$$V_T = 9386 \frac{\text{lb}}{\text{hr}} \cdot \frac{1 \text{ ft}^3}{61.596 \text{ lb}} \cdot 7.48 \frac{\text{gal}}{\text{ft}^3} \cdot \frac{24 \text{ hr}}{\text{day}} \cdot \frac{7 \text{ day}}{\text{week}} \cdot \frac{1}{0.9}$$

$$= 212763 \text{ gal}$$

$$(= 28444 \text{ ft}^3)$$

To determine the height and diameter:

$$28444 \text{ ft}^3 = \frac{\pi}{4} D^2 H = \frac{\pi}{4} D^2 \left(\frac{D}{3}\right) = \frac{\pi D^3}{12}$$

$$\Rightarrow D = 47.7 \text{ ft} \rightarrow \text{round to } 48.0 \text{ ft}$$

So $H = 48.0/3 = 16.0$ ft

$$\text{Actual volume} = \frac{\pi}{4} (48 \text{ ft})^2 (16 \text{ ft})$$

$$= 28953 \text{ ft}^3 = 216568 \text{ gal}$$

Tank 1
48.0 ft in diameter
16.0 ft in height
216568 gal capacity

Tank 1 [cont'd]

From Fig 14-58 of Peters and Timmerhaus, a cone roof tank of this volume has the cost

$$\text{Purchased cost} = \$70000$$

$$1999 \text{ cost} = \$70000\,(1.1646) = \boxed{\$81522}$$

Tank 2: EtOH product storage

The flow rate of the distillate product is that of stream 18:

$$L = 434 \text{ lb/hr}$$
$$\rho = 50.239 \text{ lb/ft}^3$$

Thus, for the tank to hold one week's worth of product at 90% capacity, the volume will have to be:

$$V_T = 434\,\frac{\text{lb}}{\text{hr}} \cdot \frac{1\ \text{ft}^3}{50.239\ \text{lb}} \cdot \frac{7.48\ \text{gal}}{\text{ft}^3} \cdot \frac{24\ \text{hr}}{\text{day}} \cdot \frac{7\ \text{day}}{\text{week}} \cdot \frac{1}{0.9}$$

$$= 12062 \text{ gal}$$

$$(= 1613 \text{ ft}^3)$$

Now I'll determine the length and diameter as before:

$$1613 \text{ ft}^3 = \frac{\pi}{12} D^3$$

$$\Rightarrow D = 18.3 \text{ ft} \rightarrow \text{round to } 18.5 \text{ ft}$$

$$\text{So } H = 18.5 \text{ ft}/3 = 6.2 \text{ ft} \rightarrow \text{round to } 6.5 \text{ ft}$$

$$\text{Actual volume} = \frac{\pi}{4}(18.5 \text{ ft})^2(6.5 \text{ ft})$$

$$= 1747 \text{ ft}^3$$

$$= 13069 \text{ gal}$$

Tank 2
18.5 ft in diameter
6.5 ft in height
13069 gal capacity

Tank 2 [cont'd]

Extrapolating the cone roof tank curve of Figure 14-58 in Peters and Timmerhaus, the raw cost of this tank is:

Purchased cost = $10000

1999 cost = $10000(1.1641) = $\boxed{\$11641}$

Material Balance Calculations

* Note of explanation: Now may seem like an odd time to begin these calculations, considering that I've already sized and costed all equipment for the base case. However, I've actually been doing them all along in a piecewise manner. Otherwise, I wouldn't have had the information to evaluate my equipment needs. Now I'll just make the process formal. Many streams were characterized earlier and I simply need to transfer the information. In some cases, I'll have to calculate a few items. I will record all information on the computer printouts labelled "stream attributes."

* Assumptions: Earlier in the calculations, it was assumed that beer could be represented as:

Component	wt. frac.
H_2O	0.9484
EtOH	0.0421
CO_2	0.0053
Pyruvic Acid	0.0021
Ethyl acetate	0.0021

Upon more reflection, however, I have concluded that there must be a few other components which cannot be easily characterized. These components are more volatile than ethyl acetate, but less volatile than CO_2 – they can be conventionally condensed. I will term these components "Non-characterizable volatiles", or "volatiles" for short. Based on my review of the literature, I estimate that there are 0.2 lb of volatiles per 100 lb of the above mixture. Thus, when dealing with process streams, I will take these volatiles into account (e.g. scaling volumetric flow rates, enthalpies, etc.)

The other major assumption I will use is that of feed rate. I need to start somewhere, so I'll begin with a feed rate of:

$$240000 \frac{bbl}{yr} \cdot \frac{1\ yr}{8400\ hr} \cdot 42 \frac{gal}{bbl} \cdot \frac{1\ ft^3}{7.48\ gal} \cdot 61.362 \frac{lb}{ft^3} = 9843\ lb/hr.$$

from calculated Density of feed at 80°F

I have to be careful with this flow rate because I ran some of my simulations with it and others with a nominal

New mass fractions

Converting the mass fractions of beer w/o volatiles to mass fractions with volatiles, I get:

Component	wt. frac
H_2O	0.9465
EtOH	0.0420
CO_2	0.0053
Pyruvic Acid	0.0021
Ethyl Acetate	0.0021
Volatiles	0.0020

Material balance [cont'd]

flow rate of 100 lb/hr. For the latter case, I'll scale everything by a factor of 98.43

* Calculations: The first calculation involves both the feed to the process and the product stream. When I specified 9843 lb/hr of feed, I meant 9843 lb/hr going to the separation process. In actuality what I plan to do is to take a sidestream from the feed and mix it in with the product. I'll add a very small amount so as to keep the alcohol content low, but enough to enhance the flavor of the product.

Stream 2

Clearly, I need the attributes of stream 2, the feed to the separation. These I'll get from the flash simulation on p. 7 of the "Sizing Information" section. (I need to scale up and include volatiles.

Stream 13

I also need to know about stream 13, which I'll get from my E-3 heat exchanger simulation.

Stream 30

Now that I know these two streams, I'll solve for the amount of regular beer that I can add and still have the mixture be < 0.5 vol% EtOH. (legal criterion for NA beer). Just to be safe, I'll specify 0.4 vol%.

Volumetric flow rate of EtOH already in stream 13:

$$\frac{10.8 \text{ lb/hr}}{500\,(0.8)} = 0.027 \text{ gpm}$$

conversion factor ↗ ↖ sp.gr. of EtOH

Total flow of S_{13} = 18.77 gpm

Therefore,

$$\frac{0.027 + x}{18.77} = 0.4\% = 0.004$$

Solving, $x = 0.04808$ gpm

This is the amount of EtOH I can safely add.

Stream 30 [cont'd]

Converting to mass, $0.04808(500)(0.8) = 19.232$ lb/hr

From my earlier definition of the components of beer, I know it contains

$$\frac{4.21 \text{ lb EtOH}}{100.2 \text{ lb beer}}$$

Therefore, I can add as stream 30

$$19.232 \frac{\text{lb EtOH}}{\text{hr}} \cdot \frac{100.2 \text{ lb beer}}{4.21 \text{ lb EtOH}} = \underline{457.7 \text{ lb beer/hr}}$$

Stream 1

Stream 1 is simply the sum of streams 2 and 30

Stream 31

S-31 is the sum of S-30 and S-13
(density, enthalpy, and MW calculated by ChemCad)

Stream 29

This stream makes up the CO_2 that is lost in the distillation process.

$$\text{Amount } CO_2 \text{ req'd} = \text{Amt. in S-1} - \text{Amt. in S-31}$$

$$= 54.63 \text{ lb/hr} - 4.19 \text{ lb/hr}$$

$$= \underline{50.44 \text{ lb/hr}}$$

I will assume CO_2 is available at 70°F + 100 psia

Streams 3 + 4

S-3 and S-4 are the same as S-2, except at a higher temperature and a slightly lower pressure (due to E-1 and E-2)

Streams 5 + 6

S-5 and S-6 are the de-gasifier flash liquid and flash vapor, respectively, and are calculated by scaling the result of the latest flash simulation (p. 7 of "Sizing Information" section)

I just multiply all of the lb/h flow rates by my scale factor of 98.43 and assume that all of the volatiles go into the vapor.

Here's a check : Total vapor flow rate (w/o volatiles) $= 0.5415 \frac{lb}{hr} (98.43)$

$= 53.3$ lb/hr

Calculated another way by summing up individual components :

Total flow =	0.83	H_2O
	4.19	EtOH
	47.4	CO_2
	0.87	Ethyl acetate
	53.29 lb/hr	

✓ Everything is OK

For the liquid, I'll check the water flow :

H_2O in S-5 = H_2O in S-4 − H_2O in S-6

= 9335.1 − 0.83

= 9334.3 lb/hr

Another way :

H_2O in S-5 $= 94.8316 \frac{lb}{hr} \cdot (98.43)$ ← scale factor

= 9334.27 lb/hr

✓ OK

The same holds true for the rest of the components

** Note: for some reason, the energy balance on S-4, S-5, and S-6 does not quite close, but I'll worry about that later.

Stream 7, 24, and 22

These are the streams after and before the after-flash condenser, respectively. S-22 is the feed to the condenser.
S-24 is the vapor effluent from the condenser
S-7 is the liquid effluent from the condenser

All three streams have been simulated (p. 6 of "Sizing Calculations"). They can be calculated simply by adding in the volatiles and ratio-ing the enthalpies and volumetric flow rates accordingly. No scaling is necessary. Note that more water is now present due to the steam from the ejector.

Stream 32

This stream is simply the product with the addition of the CO_2. I'll just add in the CO_2 and then pull numbers off ChemCad to get density, MW, etc. I still have to ratio for the volatiles.

Stream 8

Just like stream 5 except pumped to a higher pressure.

Stream 9

This is the T-1 overhead. I get it by scaling info from the simulator and then pulling off the properties.

Stage # 2	51.61 F	0.50 psia
	Vap lb/h	Liq lb/h
Water	6.88723	8.27864
Ethanol	167.85460	192.03400
Carbon Dioxide	1.98440	0.00577
Pyruvic Acid	0.00000	0.00000
Ethyl Acetate	6.07710	5.21385
Air	0.00000	0.00000
Total lb/h	182.8033	205.5323

Overhead vapor stream

Streams 10, 19, + 23

Stream 23, the tower bottoms, splits into S-10, the bottoms product, and S-19, which goes to the reboiler.

I can get S-10 by scaling the distillation simulation results (p. 4 of "Sizing Calculations"). Then I'll get S-23 from the simulator, also by scaling:

ıge # 60		124.68 F		1.97 psia
		Vap lb/h		Liq lb/h
Water		91.66531		187.13936
Ethanol		2.31752		0.93759
Carbon Dioxide		0.00000	S-23	0.00000
Pyruvic Acid		0.01203		0.22421
Ethyl Acetate		0.19709		0.20516
Air		0.00000		0.00000
Total lb/h		94.1919		188.5063

Stage # 61		125.83 F		2.00 psia
		Vap lb/h		Liq lb/h
Water		92.47571		94.66364
Ethanol		0.86422		0.07337
Carbon Dioxide		0.00000	S-10	0.00000
Pyruvic Acid	S-27	0.01421		0.21000
Ethyl Acetate		0.15218		0.05298
Air		0.00000		0.00000
Total lb/h		93.5063		95.0000

S-19 is simply the difference of these two streams (S-23 + S-10 need to be scaled however)

Streams 11, 21, + 12

S-11 exits recombiner, is pumped and becomes S-21, exchanges heat and becomes S-12

I have the stream 11 attributes from earlier (p. 8 of "Sizing Calculations"). I'll get the enthalpies from ChemCad. It's all that changes, really.

Streams 14, 15, 16, 17, + 18

S-14 is the condensate to the reflux drum
S-17 is the non-condensible vapors from S-9
S-15 is the effluent from the reflux drum
S-16 is the pumped condensate
S-18 is the distillate product

Procedure: get S-17 by taking CO_2 from S-9 and air leaked into column (calculated as 53.9 lb/h, p.32 of "Sizing Calculations")

get S-14 by taking what's left of S-9, but specify liquid on simulator

S-15 is exactly the same as S-14

S-16 is the reflux. It is the same as S-15 minus S-18

S-18 is the distillate product and can be obtained by scaling simulation results (p.4 of "Sizing Calculations")

* Note: the temperature of S-14 is actually higher than that of S-9. I assume that this is because removing the CO_2 alters the thermodynamics

Streams 20 + 28

These 2 streams are just the hot water circulation for the reboiler. As was calculated earlier (p.43 of sizing calculations),

1.36×10^6 lb/h of water at 136 °F and 150 °F is circulated. From the simulator, I can get the stream attributes easily.

Stream 27

Stream 27 is the vapor returned to the column and can be calculated from the simulator result already presented (p.6 of "Material Balance | Stream Attributes")

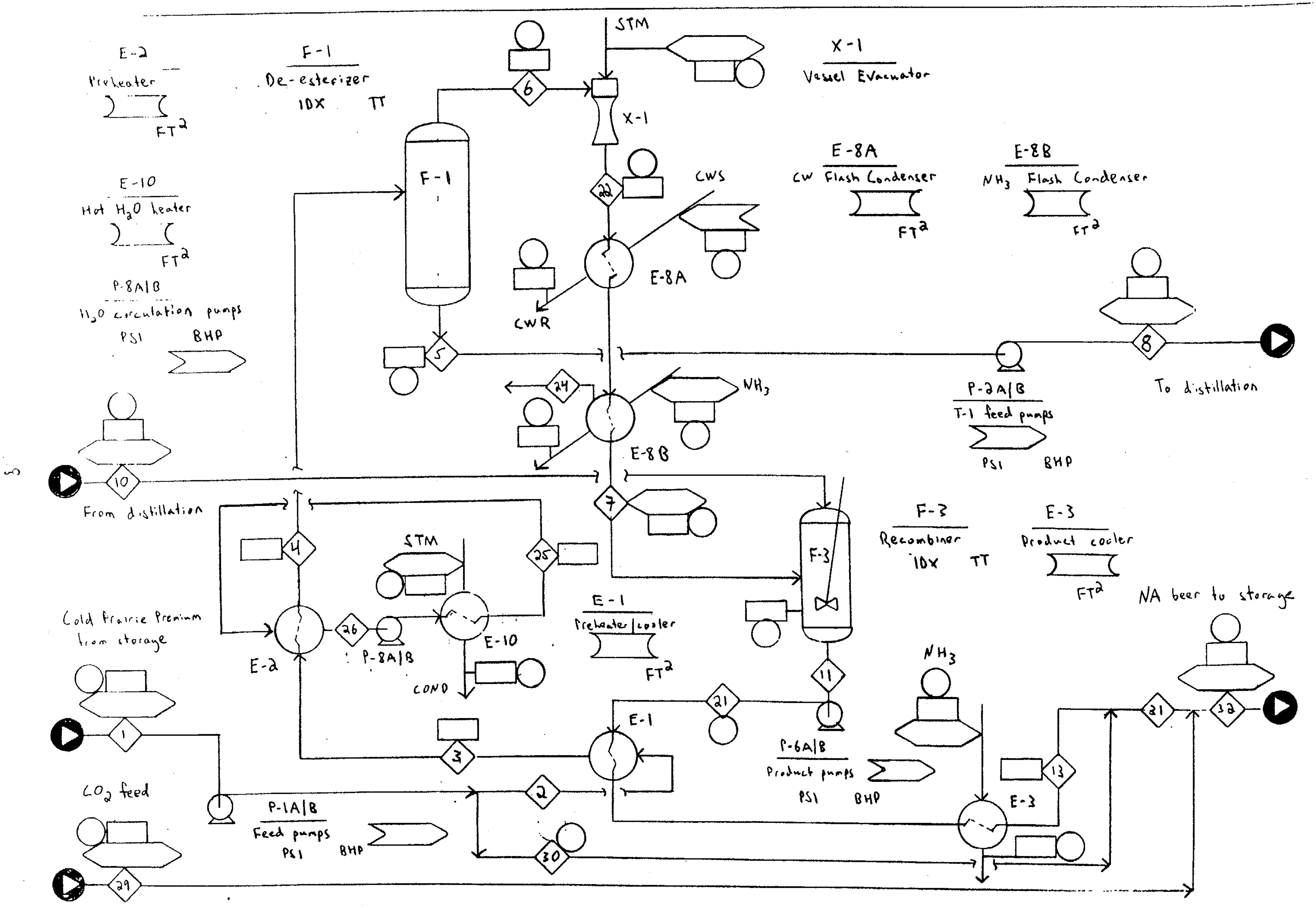
Scaling is necessary, of course.

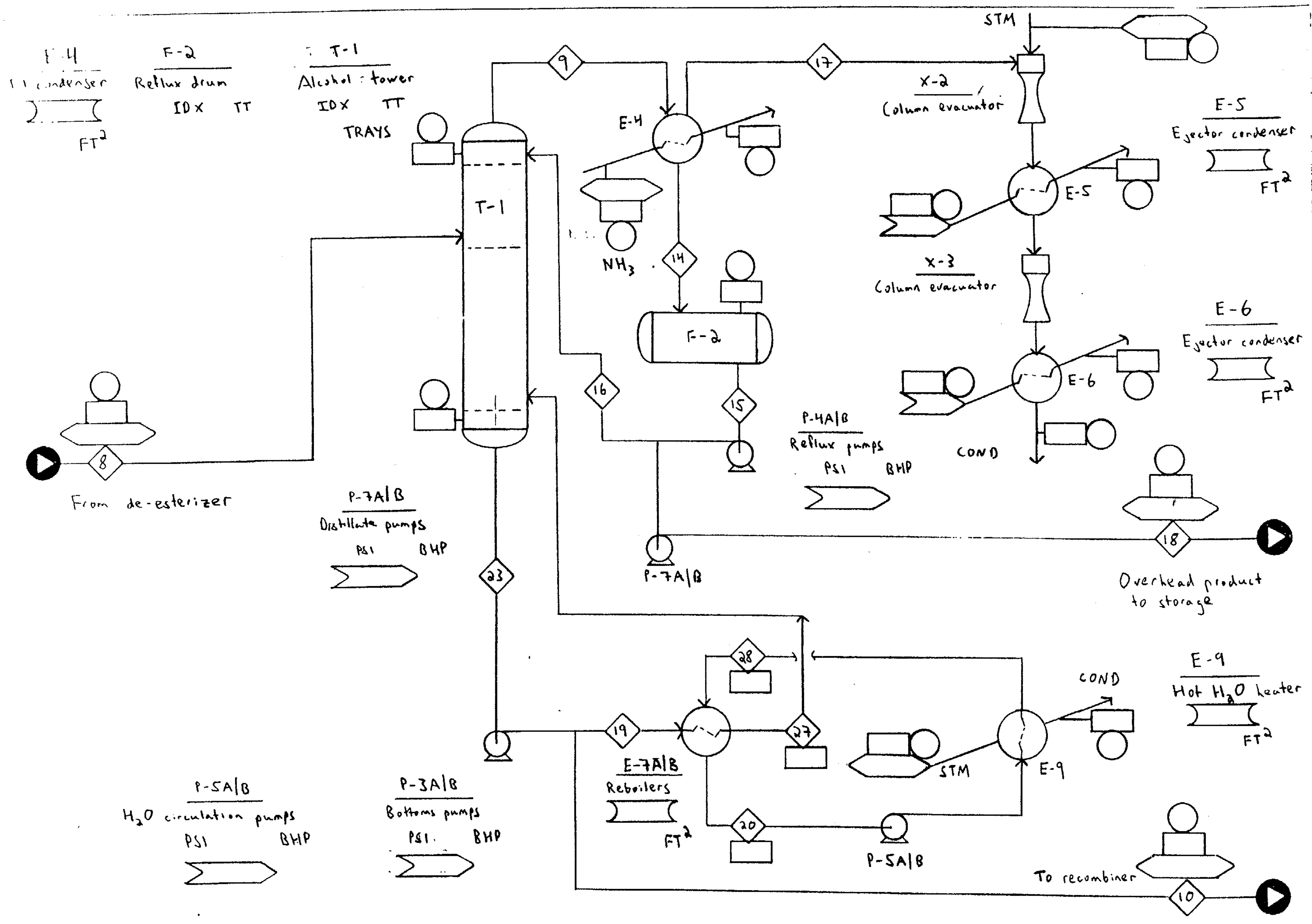
Streams 25 + 26

These are the feed preheater hot water circulation streams. As calculated earlier (p. 38 of "Sizing Calculations"), the required flow rate is 5794 lb/h.

S-25 is at 130°F, while S-6 is at 120°F

→ This concludes the task of assembling this information


E-2
Preheater
FT²
E-10
Hot H₂O heater
FT²
P-8A|B
H₂O circulation pumps
PSI
BHP
F-1
De-esterizer
10X
TT
X-1
Vessel Evacuator
E-8A
CW Flash Condenser
FT²
E-8B
NH₃ Flash Condenser
FT²
STM
CWS
CWR
NH₃
To distillation
P-2A|B
T-1 feed pumps
PSI
BHP
From distillation
F-3
Recombiner
10X
TT
E-3
Product cooler
FT²
NA beer to storage
E-1
Preheater/cooler
FT²
COND
Cold Prairie Premium from storage
CO₂ feed
P-1A|B
Feed pumps
PSI
BHP
P-6A|B
Product pumps
PSI
BHP
1
2
3
4
5
6
7
8
10
11
13
21
22
24
25
26
29
30
31
32


E-4
condenser
FT²
F-2
Reflux drum
ID x TT
T-1
Alcohol tower
ID x TT
TRAYS
T-1
E-4
NH₃
F-2
9
17
14
15
16
8
From de-esterizer
23
STM
X-2
Column evacuator
X-3
Column evacuator
E-5
E-6
COND
E-5
Ejector condenser
FT²
E-6
Ejector condenser
FT²
P-4A|B
Reflux pumps
PSI BHP
P-7A|B
Distillate pumps
PSI BHP
P-7A|B
18
Overhead product to storage
28
27
19
20
E-7A|B
Reboilers
FT²
P-5A|B
STM
E-9
COND
E-9
Hot H₂O heater
FT²
P-5A|B
H₂O circulation pumps
PSI BHP
P-3A|B
Bottoms pumps
PSI BHP
To recombiner
10

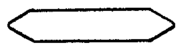
Explanation of Flowsheet Symbols

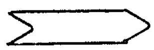
I believe that my flowsheets follow fairly standard conventions, but just in case, here is an explanation of the flow symbols I have used (the pieces of equipment should be obvious):

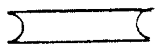
 Stream number

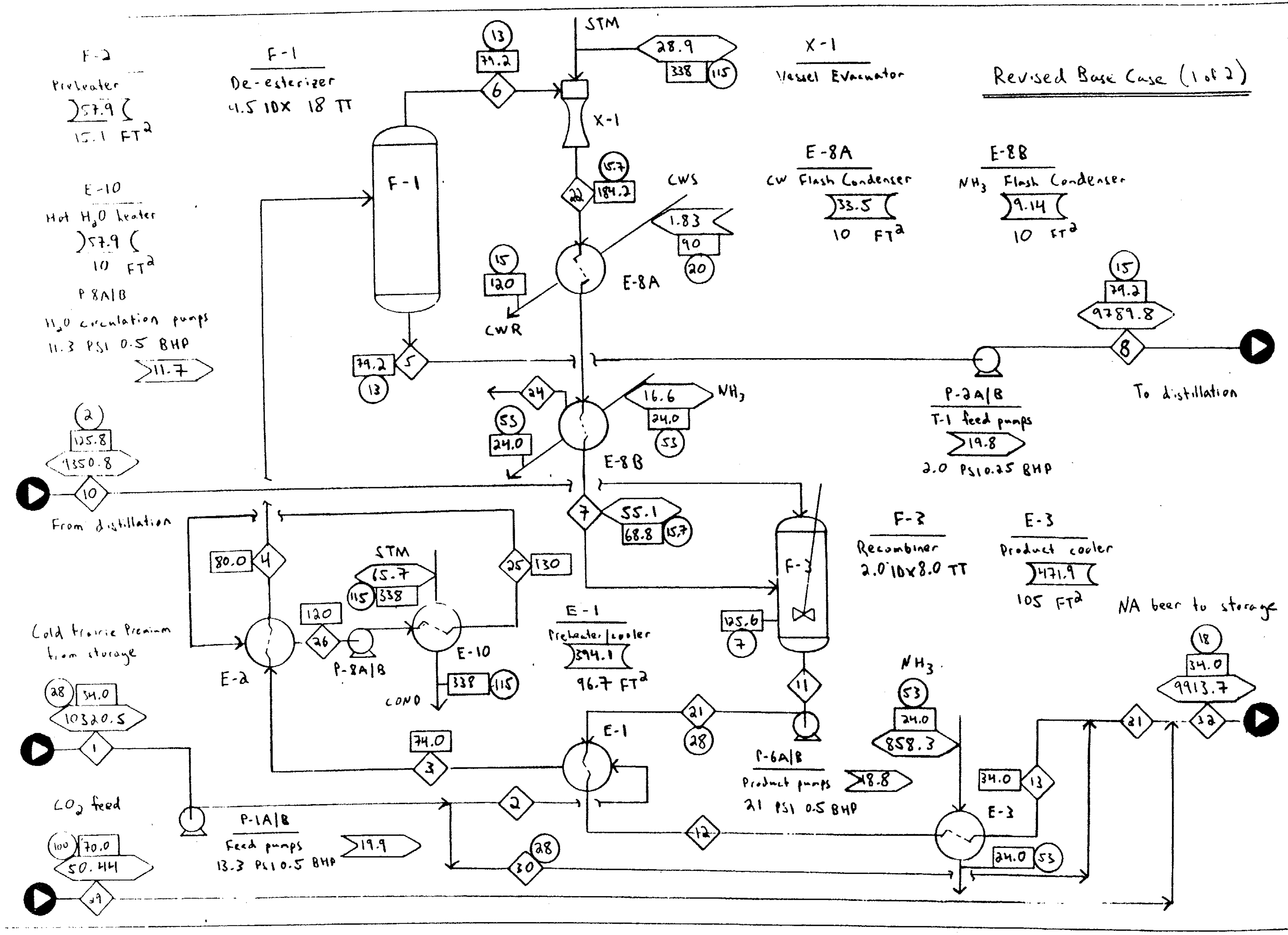
 Pressure, psia

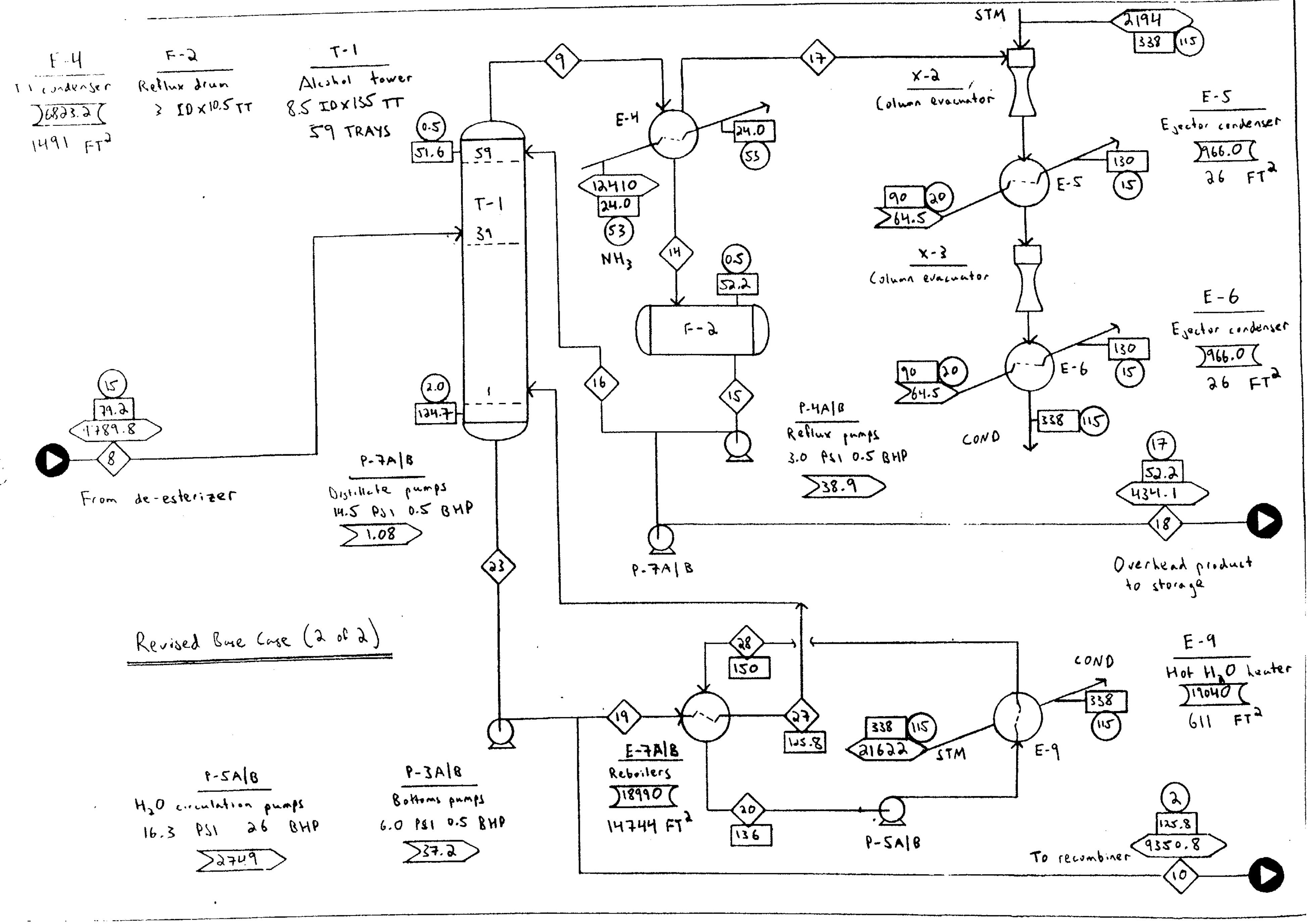
 Temperature, °F

 Mass flow, lb/hr

 Volumetric flow, gpm

 Duty, 10^3 Btu/hr


Revised Base Case (1 of 2)
F-2
Preheater
57.9
15.1 FT²
F-1
De-esterizer
4.5 ID X 18 TT
X-1
Vessel Evacuator
E-10
Hot H₂O heater
57.9
10 FT²
P 8A/B
H₂O circulation pumps
11.3 PSI 0.5 BHP
11.7
E-8A
CW Flash Condenser
33.5
10 FT²
E-8B
NH₃ Flash Condenser
9.14
10 FT²
STM
28.9
338
115
13
79.2
6
X-1
15.7
184.2
22
CWS
1.83
90
20
15
120
CWR
E-8A
F-1
79.2
13
5
24
53
24.0
16.6
24.0
53
NH₃
E-8B
15
79.2
9789.8
8
To distillation
P-2A/B
T-1 feed pumps
19.8
2.0 PSI 0.25 BHP
2
125.8
1350.8
10
From distillation
7
55.1
68.8
15.7
80.0
4
STM
15.7
115
338
25
130
120
26
P-8A/B
E-2
E-10
338
115
COND
E-1
Preheater/cooler
394.1
96.7 FT²
F-3
125.6
7
F-3
Recombiner
2.0 ID x 8.0 TT
E-3
Product cooler
471.9
105 FT²
NA beer to storage
Cold Prairie Premium
from storage
28
34.0
10320.5
1
74.0
3
E-1
21
28
11
P-6A/B
Product pumps
21 PSI 0.5 BHP
18.8
NH₃
53
24.0
858.3
18
34.0
9913.7
32
31
34.0
13
E-3
24.0
53
CO₂ feed
100
70.0
50.44
29
P-1A/B
Feed pumps
13.3 PSI 0.5 BHP
19.9
2
12
28
30


E-4
T-1 condenser
6823.2
1491 FT²
F-2
Reflux drum
3 ID x 10.5 TT
T-1
Alcohol tower
8.5 ID x 135 TT
59 TRAYS
X-2
Column evacuator
X-3
Column evacuator
E-5
Ejector condenser
966.0
26 FT²
E-6
Ejector condenser
966.0
26 FT²
STM
2194
338
115
9
17
E-4
24.0
53
12410
24.0
53
NH₃
14
0.5
52.2
F-2
0.5
51.6
59
31
1
2.0
124.7
16
15
130
15
90
20
64.5
E-5
E-6
130
15
90
20
64.5
338
115
COND
P-4A|B
Reflux pumps
3.0 PSI 0.5 BHP
38.9
15
79.2
1789.8
8
From de-esterizer
P-7A|B
Distillate pumps
14.5 PSI 0.5 BHP
1.08
P-7A|B
17
52.2
434.1
18
Overhead product to storage
23
Revised Base Case (2 of 2)
28
150
19
27
125.8
E-7A|B
Reboilers
18990
14744 FT²
338
115
21622
STM
COND
338
115
E-9
E-9
Hot H₂O heater
19040
611 FT²
20
136
P-5A|B
P-5A|B
H₂O circulation pumps
16.3 PSI 26 BHP
2749
P-3A|B
Bottoms pumps
6.0 PSI 0.5 BHP
37.2
2
125.8
9350.8
10
To recombiner

Stream Attributes – Base Case (1 of 3)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Title	Feed to process	Feed to separation	Partially preheated feed	Preheated Feed	Flash Liquid	Flash Vapor	E-8 liq. effluent	T-1 feed	T-1 overhead	T-1 bottoms to recombiner	Recombiner Outlet	Partially cooled recombined
Water	9768.3	9335.1	9335.1	9335.1	9334.3	0.83	29.2	9334.3	677.9	9317.7	9346.6	9346.6
Ethanol	433.62	414.4	414.4	414.4	410.2	4.19	3.63	410.2	16521.5	7.22	10.8	10.5
Carbon Dioxide	54.63	52.2	52.2	52.2	4.76	47.4	1.76	4.76	195.3	0.0	1.76	1.76
Pyruvic Acid	21.66	20.7	20.7	20.7	20.7	0.0	0.0	20.7	0.0	20.7	20.7	20.7
Ethyl Acetate	21.66	20.7	20.7	20.7	19.8	0.87	0.807	19.8	598.2	5.21	6.0	6.0
Air	0.0	0.0	0.0	0.0	0.0	0.0	0.0001	0.0	0.0	0.0	0.0001	0.0001
Volatiles	20.62	19.7	19.7	19.7	0.0	19.7	19.7	0.0	0.0	0.0	19.7	19.7
Total	10320.5	9862.8	9862.8	9862.8	9789.8	72.99	55.1	9789.8	17992.9	9350.8	9405.6	9405.6
Volumetric	20.81	19.89	20.00	20.03	19.87	12.17	0.115	19.83	756.1	18.93	19.03	18.86
Temperature	34.0	34.0	74.0	80.0	79.24	79.24	68.8	79.24	51.61	125.83	125.6	84.25
Pressure	28.0	28.0	23.0	18.0	13.226	13.226	15.7	15.226	0.50	2.0	7.0	23.0
Phase	L	L	L	L	L	V	L	L	V	L	L	L
Molecular Weight	18.61	18.61	18.61	18.61	18.55	43.54	20.24	18.55	44.16	18.10	18.07	18.07
Density	61.785	61.785	61.455	61.362	61.507	0.100	59.641	61.507	6.00403	61.54	61.596	62.152
Enthalpy	-68174.9	-65151.4	-63525.8	-65151.7	-64953.5	-270.71	-338.73	-64820	-46418.7	-63157.4	-63532.3	-63919.5

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Stream Attributes — Base Case (2 of 3)

Stream Number	13	14	15	16	17	18	19	20	21	22	23	24
Stream Title	Preliminary Product	T-1 Condensate	Reflux from effluent	Pumped reflux	T-1 Air + CO_2	Overhead Product	T-1 bottoms to reboiler	Reboiler H_2O	Pumped recombiner outlet	X-1 Discharge	T-1 total bottoms	E-8 vap. effluent
Water	9346.6	677.9	677.9	661.4	0.0	16.54	9102.4	1.36×10^6	9346.6	29.73	18420.1	0.52
Ethanol	10.8	16521.5	16521.5	16118.5	0.0	403.0	85.07	—	10.8	4.19	92.29	0.56
Carbon Dioxide	1.76	0.0	0.0	0.0	195.3	0.0	0.0	—	1.76	47.4	0.0	45.64
Pyruvic Acid	20.7	0.0	0.0	0.0	0.0	0.0	2.0	—	20.7	0.0	22.07	0.0
Ethyl Acetate	6.0	598.2	598.2	583.61	0.0	14.59	14.98	—	6.0	0.86	20.19	0.059
Air	0.0001	0.0	0.0	0.0	53.9	0.0	0.0	—	0.0001	8.31	0.0	8.3099
Volatiles	19.7	0.0	0.0	0.0	0.0	0.0	0.0	—	19.7	19.7	0.0	0.0
Total	9405.6	17797.6	17797.6	17363.5	249.2	434.13	9204.5	1.36×10^6	9405.6	110.2	18554.7	55.09
Volumetric	18.77	43.96	43.96	42.89	1151.8	1.07	18.64	2758.3	19.03	27.4	37.57	8.19
Temperature	34.0	52.21	52.21	52.21	51.61	52.21	125.83	136.0	125.6	184.2	124.68	68.8
Pressure	18.0	0.5	0.5	3.5	0.5	3.5	2.00	14.7	28.0	15.7	1.97	15.7
Phase	L	L	L	L	V	L	L	L	L	V	L	V
Molecular Weight	18.07	44.16	44.16	44.16	39.56	44.16	18.10	18.016	18.07	29.18	18.10	40.33
Density	62.425	50.44	50.44	50.44	0.0036	50.44	61.54	61.434	61.596	0.067	61.54	0.112
Enthalpy	-64395.1	-48952.3	-48952.3	-47758.3	-730.51	-1194.1	-61698.0	-9194000	-63552.3	-441.2	-124855.4	-179.7

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes — Base Case (3 of 3)

Stream Number	25	26	27	28	29	30	31	32	33	34	35	36
Stream Title	Preheater H_2O	Preheater H_2O	Vapor to T-1	Reboiler H_2O	CO_2 feed	Bypass reg. beer	WAB to storage w/o CO_2	Final NAB Product				
Water	5794.0	5794.0	9102.4	1.36×10^6	0.0	433.2	9779.8	9779.8				
Ethanol	—	—	85.07	—	0.0	19.22	30.02	30.02				
Carbon Dioxide	—	—	0.0	—	50.44	2.43	4.19	54.63				
Pyruvic Acid	—	—	1.40	—	0.0	0.96	21.66	21.66				
Ethyl Acetate	—	—	14.98	—	0.0	0.96	6.96	6.96				
Air	—	—	0.0	—	0.0	0.0	0.0	0.0				
Volatiles	—	—	0.0	—	0.0	0.92	20.62	20.62				
Total	5794.0	5794.0	9203.9	1.36×10^6	50.44	457.7	9863.3	9913.7				
Volumetric	11.70	11.73	26514.8	2769.9	1.05	0.92	19.70	19.81				
Temperature	130	120	125.83	150.0	70.0	34.0	34.0	34.0				
Pressure	26.0	16.0	2.0	26.0	100	28.0	18.0	18.0				
Phase	L	L	V	L	V	L	L	L				
Molecular Weight	18.016	18.016	18.14	18.016	44.01	18.61	18.10	18.15				
Density	61.697	61.536	0.0058	61.177	0.804	61.785	62.395	62.367				
Enthalpy	-39261.6	-39203.8	-52537.8	-9175000	-194.3	-3023.5	-67441.5	-67640.6				

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/nr

Stream Attributes – Base Case (1 of 3)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Title	Feed to process	Feed to separation	Partially preheated feed	Preheated Feed	Flash Liquid	Flash Vapor	E-8 liq. effluent	T-1 feed	T-1 overhead	T-1 bottoms to reboiler	Recombiner Outlet	Partially cooled recombined
Water	9768.3	9335.1	9335.1	9335.1	9334.3	0.83	29.2	9334.3	677.9	9317.7	9346.6	9346.6
Ethanol	433.62	414.4	414.4	414.4	410.2	4.19	3.63	410.2	16521.5	7.22	10.8	10.8
Carbon Dioxide	54.63	52.2	52.2	52.2	4.76	47.4	1.76	4.76	195.3	0.0	1.76	1.76
Pyruvic Acid	21.66	20.7	20.7	20.7	20.7	0.0	0.0	20.7	0.0	20.7	20.7	20.7
Ethyl Acetate	21.66	20.7	20.7	20.7	19.8	0.87	0.807	19.8	598.2	5.21	6.0	6.0
Air	0.0	0.0	0.0	0.0	0.0	0.0	0.0001	0.0	0.0	0.0	0.0001	0.0001
Volatiles	20.62	19.7	19.7	19.7	0.0	19.7	19.7	0.0	0.0	0.0	19.7	19.7
Total	10320.5	9862.8	9862.8	9862.8	9789.8	72.99	55.1	9789.8	17992.9	9350.8	9405.6	9405.6
Volumetric	20.81	19.89	20.00	20.03	19.87	12.17	0.115	19.83	756.1	18.93	19.03	18.86
Temperature	34.0	34.0	74.0	80.0	79.24	79.24	68.8	79.24	51.61	125.83	125.6	84.25
Pressure	28.0	28.0	23.0	18.0	13.226	13.226	15.7	15.226	0.50	2.0	7.0	23.0
Phase	L	L	L	L	L	V	L	L	V	L	L	L
Molecular Weight	18.61	18.61	18.61	18.61	18.55	43.54	20.24	18.55	44.16	18.10	18.07	18.07
Density	61.785	61.785	61.455	61.362	61.507	0.100	59.641	61.507	0.00403	61.54	61.596	62.152
Enthalpy	-68174.9	-65151.4	-63525.8	-65151.7	-64953.5	-270.71	-338.73	-64820	-46418.7	-63157.4	-63552.3	-63919.5

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Stream Attributes — Base Case (2 of 3)

Stream Number	13	14	15	16	17	18	19	20	21	22	23	24
Stream Title	Preliminary Product	T-1 Condensate	Reflux from effluent	Pumped reflux	T-1 Air + CO_2	Overhead Product	T-1 bottoms to reboiler	Reboiler H_2O	Pumped reboiler outlet	X-1 Discharge	T-1 total bottoms	E-8 vap. effluent
Water	9346.6	677.9	677.9	661.4	0.0	16.54	9102.4	1.36×10^6	9346.6	29.73	18420.1	0.52
Ethanol	10.8	16521.5	16521.5	16118.5	0.0	403.0	85.07	—	10.8	4.19	92.29	0.56
Carbon Dioxide	1.76	0.0	0.0	0.0	195.3	0.0	0.0	—	1.76	47.4	0.0	45.64
Pyruvic Acid	20.7	0.0	0.0	0.0	0.0	0.0	2.0	—	20.7	0.0	22.07	0.0
Ethyl Acetate	6.0	598.2	598.2	583.61	0.0	14.59	14.98	—	6.0	0.86	20.19	0.059
Air	0.0001	0.0	0.0	0.0	53.9	0.0	0.0	—	0.0001	8.31	0.0	8.3099
Volatiles	19.7	0.0	0.0	0.0	0.0	0.0	0.0	—	19.7	19.7	0.0	0.0
Total	9405.6	17797.6	17797.6	17363.5	249.2	434.13	9204.5	1.36×10^6	9405.6	110.2	18554.7	55.09
Volumetric	18.77	43.96	43.96	42.89	1151.8	1.07	18.64	2758.3	19.03	27.4	37.57	8.19
Temperature	34.0	52.21	52.21	52.21	51.61	52.21	125.83	136.0	125.6	184.2	124.68	68.8
Pressure	18.0	0.5	0.5	3.5	0.5	3.5	2.00	14.7	28.0	15.7	1.97	15.7
Phase	L	L	L	L	V	L	L	L	L	V	L	V
Molecular Weight	18.07	44.16	44.16	44.16	39.56	44.16	18.10	18.016	18.07	29.18	18.10	40.33
Density	62.425	50.44	50.44	50.44	0.0036	50.44	61.54	61.434	61.596	0.067	61.54	0.112
Enthalpy	-64395.1	-48952.3	-48952.3	-47758.3	-730.51	-1194.1	-61698.0	-9194000	-63552.3	-441.2	-124855.4	-179.7

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Stream Attributes — Base Case (3 of 3)

Stream Number	25	26	27	28	29	30	31	32	33	34	35	36
Stream Title	Preheater H_2O	Preheater H_2O	Vapor to T-1	Reboiler H_2O	CO_2 feed	Bypass reg. beer	NAB to storage w/o CO_2	Final NAB Product				
Water	5794.0	5794.0	9102.4	1.36×10^6	0.0	433.2	9779.8	9779.8				
Ethanol	—	—	85.07	—	0.0	19.22	30.02	30.02				
Carbon Dioxide	—	—	0.0	—	50.44	2.43	4.19	54.63				
Pyruvic Acid	—	—	1.40	—	0.0	0.96	21.66	21.66				
Ethyl Acetate	—	—	14.98	—	0.0	0.96	6.96	6.96				
Air	—	—	0.0	—	0.0	0.0	0.0	0.0				
Volatiles	—	—	0.0	—	0.0	0.92	20.62	20.62				
Total	5794.0	5794.0	9203.9	1.36×10^6	50.44	457.7	9863.3	9913.7				
Volumetric	11.70	11.73	26514.8	2769.9	1.05	0.92	19.70	19.81				
Temperature	130	120	125.83	150.0	70.0	34.0	34.0	34.0				
Pressure	26.0	16.0	2.0	26.0	100	28.0	18.0	18.0				
Phase	L	L	V	L	V	L	L	L				
Molecular Weight	18.016	18.016	18.14	18.016	44.01	18.61	18.10	18.15				
Density	61.697	61.536	0.0058	61.177	0.804	61.785	62.395	62.367				
Enthalpy	-39261.6	-39203.8	-52537.8	-9175000	-194.3	-3023.5	-67441.5	-67640.6				

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Table 13.1 Utility Summary for Revised Base Case

A. Electricity at $0.0494/kWhr

User	Yearly Cost
P-1A/B	$154.80
P-2A/B	$557.20
P-3A/B	$154.80
P-4A/B	$154.80
P-5A/B	$8049.00
P-6A/B	$154.80
P-7A/B	$154.80
P-8A/B	$154.80
Total	**$9535.00**

B. 100 psig Steam at $2.815/1000 lb

User	Amount (lb/h)	Yearly Cost
X-1	28.9	$683.32
X-2 and X-3	2194	$51875.93
E-9	21622	$511240.39
E-10	65.7	$1553.44
Total	**23910.6**	**$565,353.08**

Table 13.1 (cont'd)

C. Cooling Water at $0.117/1000 gal

User	Amount (gpm)	Yearly Cost
E-5	64.5	$3812.67
E-6	64.5	$3812.67
E-8A	1.83	$108.17
Total	**130.83**	**$7733.51**

D. Ammonia Refrigeration at $0.0081/10^3 Btu

User	Amount (10^3 Btu/hr)	Yearly Cost
E-3	471.9	$32,058.53
E-4	6823.2	$463,534.09
E-8B	9.14	$620.93
Total	**7407.24**	**$496,213.55**

Total Utilities Cost = $1,078,835/yr

Table 15.1 Equipment Information Summary for Revised Base Case

A. Pumps

Number	Description	Body Type	Suct. P (psia)	Disch. P (psia)	GPM	Head (ft)	Motor BHP
P-1A/B	Feed	1.5 x 1.5	14.7	28.0	19.9	31.0	0.5
P-2A/B	T-1 Feed	1.25 x 1.25	13.2	15.2	19.8	4.67	0.25
P-3/AB	Bottoms	1.5 x 1.5	2.0	8.0	37.2	13.9	0.5
P-4A/B	Reflux	1.5 x 1.5	0.5	3.5	38.9	8.6	0.5
P-5A/B	Hot water	8 x 6	14.7	31.0	2749	38.1	26
P-6A/B	Product	1.5 x 1.5	7.0	28.0	18.8	48.5	0.5
P-7A/B	Distillate	1.5 x 1.5	2.5	17.0	1.1	41.8	0.5
P-8A/B	Hot water	1.5 x 1.5	14.7	26.0	11.7	26.4	0.5

B. Drums

Number	Description	Orientation	ID (ft)	L (ft)	Capacity (gal)	P (psia)	T (°F)
F-1	De-esterizer	Vertical	4.5	18.0	2141	13.2	79.2
F-2	Reflux	Horizontal	3.0	10.5	555.2	0.5	52.6
F-3	Recombiner	Vertical	2.0	8.0	188.0	7.0	125.6

C. Towers

Number	Description	Trays	Type	Feed to tray	ID (ft)	H (ft)
T-1	Alcohol removal	59	Sieve	39	8.5	135

D. Ejectors

Number	Description	Suction P (psia)	Discharge P (psia)	Stages	Steam/stage (lb/h)
X-1	F-1 Evacuator	13.2	15.7	1	28.9
X-2 and X-3	T-1 Evacuators	0.5	15.7	2	1097

Table 15.1 (cont'd)

E. Heat Exchangers

Number	Description	Shell Side	Tube Side	Type	Duty (10^3 Btu/hr)	Area (ft^2)
E-1	Preheater	Feed	Product	F.T.S.*	394.1	96.7
E-2	Preheater	Hot water	Feed	F.T.S.	57.9	15.1
E-3	Cooler	Ammonia	Product	F.T.S.	471.9	105
E-4	Condenser	Overhead	Ammonia	F.T.S.	6823.2	1491
E-5	X-2 Condenser	Steam	CW	F.T.S.	966.0	26.0
E-6	X-2 Condenser	Steam	CW	F.T.S.	966.0	26.0
E-7A/B	Reboilers	Hot water	Bottoms	F.T.S.	18990	14744
E-8A	CW Flash cond.	CW	Fsh. vapor	F.T.S.	33.5	10.0
E-8B	NH_3 Flash cond.	Ammonia	Fsh. vapor	F.T.S.	9.1	10.0
E-9	Hot water heater	Steam	H_2O circ.	F.T.S.	19040	611
E-10	Hot water heater	Steam	H_2O circ.	F.T.S.	57.9	10.0

*F.T.S. = Fixed-Tube-Sheet

F. Storage Tanks

Number	Description	Type	Capacity (gal)	Diameter (ft)	Height (ft)
V-1	NA beer storage	Cone-roof	216,568	48.0	16.0
V-2	Distillate storage	Cone-roof	13069	18.5	6.5

Material of Construction: All of the above equipment is constructed of 304 stainless steel except P-5A/B, P-8A/B, X-2, X-3, E-5, E-6, E-9, and E-10. These pieces of equipment only see recirculating hot water and never process fluid.

Completion of Process Part of Base Case

Now that I have completed my Stream Attributes Table and have filled in all of the necessary details on my flowsheet, I believe I am done with the process part of my base case. I am ready to move on to the economic analysis using the method I have given an example of on p. 1-11 of the "Economic Calculations" section.

Base Case Economic Calculations

Two additional informational items I need are the production rate of NA beer and the production rate of EtOH byproduct.

The byproduct (stream 18) has a composition of:

from stream attributes ⇝

16.54 lb/hr	H_2O
403.0 lb/hr	EtOH
14.59 lb/hr	Ethyl acetate
434.13 lb/hr	total

$\rho_L = 50.44$ lb/ft³

The weight fraction of EtOH is:

$$X_{EtOH} = \frac{403.0}{434.13} = 0.928$$

This is suitable for sale as fuel grade ethanol

Since my cost data is in \$/gal, I'll convert this flow rate to gal/yr:

$$\dot{m}_{EtOH} = 434.13 \frac{lb}{hr} \cdot \frac{1\ ft^3}{50.44\ lb} \cdot 7.48 \frac{gal}{ft^3} \cdot 8400 \frac{hr}{yr}$$

$$= 540786 \ \ gal/yr$$

The product (stream 32) has a flow rate of

from stream attributes ⇝ $\dot{m}_{NAB} = 9913.7$ lb/hr , $\rho_L = 62.367$ lb/ft³

Base Case Economics [cont'd]

I would like this flow rate in bbl/yr. So:

$$\dot{m}_{NAB} = 9913.7\ \frac{\cancel{lb}}{\cancel{hr}} \cdot \frac{1\ \cancel{ft^3}}{62.367\ \cancel{lb}} \cdot \frac{7.48\ \cancel{gal}}{\cancel{ft^3}} \cdot \frac{1\ bbl}{42\ \cancel{gal}} \cdot 8400\ \frac{\cancel{hr}}{yr}$$

$$= \underline{237{,}800\ bbl/yr}$$

(Note that this is greater than the 60000 bbl per year cutoff for excise taxes had this been regular beer)

Preliminary Economic Calculations

All cost figures in dollars unless otherwise stated

Section I: Capital Investment

Operating Pressure (psia)	Correction
0.08	1.3
0.2	1.2
0.7	1.1
8 to 100	1.0
700	1.1
3000	1.2
6000	1.3

Operating Temperature (C)	Correction
-80	1.3
0	1.0
100	1.05
600	1.1
5000	1.2
10000	1.4

Material	Correction
Carbon steel	1.0
Bronze	1.05
Carbon/molybdenum	1.065
Aluminum	1.075
Cast steel	1.11
Stainless steel	1.28 to 1.5
Worthite alloy	1.41
Hastelloy C alloy	1.54
Monel alloy	1.65
Nickel/inconel alloy	1.71
Titanium	2.0

Note: Correcting factors may have already been taken into account in earlier estimates. For example, inflation factors are shown as 1.0, since inflation of 1.1646 (see Economic Calculations, p. 9) is already included

Base Case

Equipment	Number	Purchase Cost	Inflation Factor	Delivered Cost	Material	Pressure	Temperature	Total Del. Cost	Installation	Controls	Piping	Electrical
Pump	P1-A/B	3606	1	3786.30	1	1	1	3786.30	1779.56	681.53	2498.96	416.49
Pump	P2-A/B	2893	1	3037.65	1	1	1	3037.65	1427.70	546.78	2004.85	334.14
Pump	P3-A/B	3786	1	3975.30	1	1	1	3975.30	1868.39	715.55	2623.70	437.28
Pump	P4-A/B	3731	1	3917.55	1	1.1	1	4309.31	2025.37	775.67	2844.14	474.02
Pump	P5-A/B	18269	1	19182.45	1	1	1	19182.45	9015.75	3452.84	12660.42	2110.07
Pump	P6-A/B	5371	1	5639.55	1	1.1	1	6203.51	2915.65	1116.63	4094.31	682.39
Pump	P7-A/B	5371	1	5639.55	1	1.1	1	6203.51	2915.65	1116.63	4094.31	682.39
Pump	P8-A/B	3786	1	3975.30	1	1	1	3975.30	1868.39	715.55	2623.70	437.28
Drivers	1 to 8-A/B	5451	1	5723.55	1	1	1	5723.55	2690.07	1030.24	3777.54	629.59
Drum	F-1	13975	1	14673.75	1	1	1	14673.75	6896.66	2641.28	9684.68	1614.11
Drum	F-2	25738	1	27024.90	1	1.2	1	32429.88	15242.04	5837.38	21403.72	3567.29
Drum	F-3	8152	1	8559.60	1	1.1	1	9415.56	4425.31	1694.80	6214.27	1035.71
*Tower	T-1	943326	1	943326.00	1	1.2	1	1131991.20	0.00	0.00	0.00	0.00
*Purchase cost correlation includes installation and auxiliaries												
Ejector	X-1	6516	1	6841.80	1	1	1	6841.80	3215.65	1231.52	4515.59	752.60
Ejector system	X-2, X-3, E-5, E-6	167531	1	175907.55	1	1	1	175907.55	82676.55	31663.36	116098.98	19349.83
Exchanger	E-1	4658	1	4890.90	1	1	1.05	5135.45	2413.66	924.38	3389.39	564.90
Exchanger	E-2	2562	1	2690.10	1	1	1.05	2824.61	1327.56	508.43	1864.24	310.71
Exchanger	E-3	4658	1	4890.90	1	1	1	4890.90	2298.72	880.36	3227.99	538.00
Exchanger	E-4	23292	1	24456.60	1	1.2	1.05	30815.32	14483.20	5546.76	20338.11	3389.68
Exchanger	E-5	*Included in X-2 and X-3 ejector system*										
Exchanger	E-6	*Included in X-2 and X-3 ejector system*										
Exchanger	E-7A	94333	1	99049.65	1	1.1	1.05	114402.35	53769.10	20592.42	75505.55	12584.26
Exchanger	E-7B	94333	1	99049.65	1	1.1	1.05	114402.35	53769.10	20592.42	75505.55	12584.26
Exchanger	E-8A	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Exchanger	E-8B	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Exchanger	E-9	11704	1	12289.20	1	1	1.05	12903.66	6064.72	2322.66	8516.42	1419.40
Exchanger	E-10	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Storage tank	V-1	81522	1	85598.10	1.39	1	1	118981.36	55921.24	21416.64	78527.70	13087.95
Storage tank	V-2	11641	1	12223.05	1.39	1	1	16990.04	7985.32	3058.21	11213.43	1868.90
							Totals	1857092.77	340797.74	130518.28	478567.03	79761.17

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Building	Yard	Service	Total Direct Cost	Engineering & Supervision	Construction	Total Indirect
378.63	378.63	2650.41	12570.52	1249.48	1552.38	2801.86
303.77	303.77	2126.36	10085.00	1002.42	1245.44	2247.86
397.53	397.53	2782.71	13198.00	1311.85	1629.87	2941.72
430.93	430.93	3016.51	14306.89	1422.07	1766.82	3188.89
1918.25	1918.25	13427.72	63685.73	6330.21	7864.80	14195.01
620.35	620.35	4342.45	20595.64	2047.16	2543.44	4590.59
620.35	620.35	4342.45	20595.64	2047.16	2543.44	4590.59
397.53	397.53	2782.71	13198.00	1311.85	1629.87	2941.72
572.36	572.36	4006.49	19002.19	1888.77	2346.66	4235.43
1467.38	1467.38	10271.63	48716.85	4842.34	6016.24	10858.58
3242.99	3242.99	22700.92	107667.20	10701.86	13296.25	23998.11
941.56	941.56	6590.89	31259.66	3107.13	3860.38	6967.51
113199.12	113199.12	792393.84	2150783.28	373557.10	464116.39	837673.49
684.18	684.18	4789.26	22714.78	2257.79	2805.14	5062.93
17590.76	17590.76	123135.29	584013.07	58049.49	72122.10	130171.59
513.54	513.54	3594.81	17049.68	1694.70	2105.53	3800.23
282.46	282.46	1977.22	9377.69	932.12	1158.09	2090.21
489.09	489.09	3423.63	16237.79	1614.00	2005.27	3619.27
3081.53	3081.53	21570.72	102306.85	10169.05	12634.28	22803.33
11440.23	11440.23	80081.64	379815.79	37752.77	46904.96	84657.74
11440.23	11440.23	80081.64	379815.79	37752.77	46904.96	84657.74
269.67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
269.67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
1290.37	1290.37	9032.56	42840.15	4258.21	5290.50	9548.71
269.67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
11898.14	11898.14	83286.95	395018.11	39263.85	48782.36	88046.21
1699.00	1699.00	11893.03	56406.93	5606.71	6965.92	12572.63
185709.28	185709.28	1299964.94	4558120.48	612840.61	761408.03	1374248.65

Direct + Indirect	Contractor	Contingency	Fixed Capital Inv.	Working Capital	Total Capital Inv.
15372.38	768.62	1537.24	17678.23	3119.69	20797.92
12332.86	616.64	1233.29	14182.79	2502.84	16685.63
16139.72	806.99	1613.97	18560.68	3275.41	21836.09
17495.78	874.79	1749.58	20120.15	3550.61	23670.76
77880.75	3894.04	7788.07	89562.86	15805.21	105368.07
25186.23	1259.31	2518.62	28964.16	5111.32	34075.49
25186.23	1259.31	2518.62	28964.16	5111.32	34075.49
16139.72	806.99	1613.97	18560.68	3275.41	21836.09
23237.61	1161.88	2323.76	26723.25	4715.87	31439.12
59575.43	2978.77	5957.54	68511.74	12090.31	80602.05
131665.31	6583.27	13166.53	151415.11	26720.31	178135.42
38227.17	1911.36	3822.72	43961.25	7757.87	51719.12
2988456.77	149422.84	298845.68	3436725.28	606480.93	4043206.22
27777.71	1388.89	2777.77	31944.36	5637.24	37581.60
714184.65	35709.23	71418.47	821312.35	144937.47	966249.82
20849.91	1042.50	2084.99	23977.39	4231.30	28208.70
11467.90	573.39	1146.79	13188.08	2327.31	15515.39
19857.05	992.85	1985.71	22835.61	4029.81	26865.43
125110.18	6255.51	12511.02	143876.71	25390.01	169266.72
464473.52	23223.68	46447.35	534144.55	94260.80	628405.36
464473.52	23223.68	46447.35	534144.55	94260.80	628405.36
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
52388.86	2619.44	5238.89	60247.19	10631.86	70879.05
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
483064.32	24153.22	48306.43	555523.97	98033.64	653557.61
68979.56	3448.98	6897.96	79326.49	13998.79	93325.29
5932369.13	296618.46	593236.91	6822224.49	1203921.97	8026146.46

Grand Total Capital Investment:
$ 8.026E+06

Preliminary Economic Calculations

All cost figures in dollars per year unless otherwise stated

Section II: Operating Costs

Operator Positions	Inflation Factor	Operating Labor	Direct Supervision & Clerical Labor	Maintenance & Repairs	Operating Supplies	Laboratory Charges	Local taxes & Insurance	Plant Overhead	Administrative Expenses
1	1.1646	821741.76	143804.81	204666.73	30700.01	123261.26	204666.73	702127.98	175532.00

Steam use (lb/h)	Steam cost
65.7	1553.44
28.9	683.32
2194	51875.93
21622	511240.39
Total	565353.09

Process Water (gpm)	Process Water Cost
0.00	0.00

Cooling Water (gpm)	Cooling Water Cost
64.5	3812.67
64.5	3812.67
1.83	108.17
Total	7733.51

CO_2 (lb/h)	CO_2 Cost
50.44	17640.58

	Electricity Costs
	154.80
	557.20
	154.80
	154.80
	8049.00
	154.80
	154.80
	154.80
Total	9535.00

Refrigeration (1000 Btu/hr)	Cost at $2.00/288KBtu	Inflation factor	1999 Cost
471.9	27527.50	1.1646	32058.53
9.14	533.17	1.1646	620.93
6823.2	398020.00	1.1646	463534.09
		Total	496213.5444

Total Operating Cost
3.5030E+06 $/yr

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Preliminary Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	4558120.48	1374248.65	889855.37	1203921.97				
2000	1					3502977.02	350000	237800	5.00
2001	2						250000	237800	5.00
2002	3						150000	237800	5.00
2003	4						90000	237800	5.00
2004	5							237800	5.00
2005	6							237800	5.00
2006	7							237800	5.00
2007	8							237800	5.00
2008	9							237800	5.00
2009	10							237800	5.00

MISTAKE

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-8026146.46
1189000	1189000	4474444.8	540786	616496.04	0.1429	651355.42	3079272.76	1231709.10	6254.72
1189000	1189000	4474444.8	540786	616496.04	0.2449	1116283.71	3974657.13	1589862.85	3251077.99
1189000	1189000	4474444.8	540786	616496.04	0.1749	797215.27	4293725.57	1717490.23	3223450.61
1189000	1189000	4474444.8	540786	616496.04	0.1249	569309.25	4521631.59	1808652.64	3192288.20
1189000	1189000	4474444.8	540786	616496.04	0.0893	407040.16	4683900.68	1873560.27	3217380.57
1189000	1189000	4474444.8	540786	616496.04	0.0892	406584.35	4684356.49	1873742.60	3217198.24
1189000	1189000	4474444.8	540786	616496.04	0.0893	407040.16	4683900.68	1873560.27	3217380.57
1189000	1189000	4474444.8	540786	616496.04	0.0446	203292.17	4887648.67	1955059.47	3135881.37
1189000	1189000	4474444.8	540786	616496.04			5090940.84	2036376.34	3054564.50
1189000	1189000	4474444.8	540786	616496.04			5090940.84	2036376.34	3054564.50

NPW @ 12%	7.1568E+06
IRR	0.278
NPW @ IRR	-5.3E-10

Net Present Worth at 12 %
$ 7.1568E+06

Internal Rate of Return =
0.278

MISTAKE.
operating costs left out

Mistake!!!

Today, eight days before the deadline, I discovered a major error in all of my economic analyses to date (base case, base case without sale of ethanol, and first optimization). In a moment of gross mental ineptitude, I left out *operating costs* in my economic calculations for project years two through ten. Clearly, this oversight has a significant effect on the overall project economics. I should have realized that there was an error when the economics seemed so favorable. I have attempted to go back through my calculation book and label the pages which contain these errors. Also, the corrected page(s) follow this page. However, it is quite likely that I have not found every place where a correction is needed. I also do not have the time to rewrite large sections of my notes on the calculations. Thus, the reader is advised to keep these facts in mind while reading through my work. Luckily, I discovered this mistake today and not a week from now.

Obviously, the base case is not even close to economical. The first optimization is better, but is still unacceptable. Thus, there is still great need for optimization. I am beginning the capacity optimization now, and the economics will hopefully improve.

Preliminary Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	4558120.48	1374248.65	889855.37	1203921.97				
2000	1					3502977.02	350000	237800	5.00
2001	2					3502977.02	250000	237800	5.00
2002	3					3502977.02	150000	237800	5.00
2003	4					3502977.02	90000	237800	5.00
2004	5					3502977.02		237800	5.00
2005	6					3502977.02		237800	5.00
2006	7					3502977.02		237800	5.00
2007	8					3502977.02		237800	5.00
2008	9					3502977.02		237800	5.00
2009	10					3502977.02		237800	5.00

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-8026146.46
1189000	1189000	4474444.8	540786	616496.04	0.1429	651355.42	3079272.76	1231709.10	6254.72
1189000	1189000	4474444.8	540786	616496.04	0.2449	1116283.71	3974657.13	1589862.85	-251899.03
1189000	1189000	4474444.8	540786	616496.04	0.1749	797215.27	4293725.57	1717490.23	-279526.40
1189000	1189000	4474444.8	540786	616496.04	0.1249	569309.25	4521631.59	1808652.64	-310688.81
1189000	1189000	4474444.8	540786	616496.04	0.0893	407040.16	4683900.68	1873560.27	-285596.45
1189000	1189000	4474444.8	540786	616496.04	0.0892	406584.35	4684356.49	1873742.60	-285778.78
1189000	1189000	4474444.8	540786	616496.04	0.0893	407040.16	4683900.68	1873560.27	-285596.45
1189000	1189000	4474444.8	540786	616496.04	0.0446	203292.17	4887648.67	1955059.47	-367095.64
1189000	1189000	4474444.8	540786	616496.04			5090940.84	2036376.34	-448412.51
1189000	1189000	4474444.8	540786	616496.04			5090940.84	2036376.34	-448412.51

NPW @ 12%	-9.5082E+06
IRR	ERR
NPW @ IRR	ERR

Net Present Worth at 12 %
$ -9.5082E+06

Internal Rate of Return =
N/A

Results of Base Case Economics + Optimization Ideas

I have completed the economic analysis for the base case according to the methods of Peters and Timmerhaus and White, Agee, and Case. The important results are:

Total Capital Investment	$\$8.026 \times 10^6$
Total Operating Cost	$\$3.503 \times 10^6/yr$
Net Present Worth @ 12%	$\$7.~~1566~~ \times 10^6$ -9.508×10^6
Internal Rate of Return	0.~~278~~ N/A

I have used a 12% discount rate to calculate the internal rate of return because the problem statement specifies that 12% < IRR < 20%. Logically, then, the minimum attractive rate of return is 12%.

These results seem very good. However, they may be slightly too good. That is, the internal rate of return of 27.8% is outside the problem guidelines. I assume that the rationale behind the upper limit on IRR is to prevent dumb processes. The value that I really want to maximize is Net Present Worth. It is possible to have a larger NPW but a smaller IRR. Thus, one of my optimization variables will be capacity.

Another major factor in NPW is the sale of the ethanol byproduct. These sales amount to \$616,496 per year. It may be that I won't be able to sell this product after all (I need to do more checking). If I can't sell it, I don't need to bother making a pure distillate stream and can save money on my distillation column. On the other hand, if I can get 190 proof EtOH, I can make a lot more money (190 proof EtOH sells for about twice as much as fuel grade).

Thus, there are some column optimizations I can do with regard to the distillate product. In any normal design project, I would also optimize such variables as column operating pressure (at lower P, $\alpha\uparrow$ and $N_T\downarrow$, but $Q_C\uparrow \Rightarrow$ tradeoff between capital and operating costs). However in this case I want to be very careful about the quality of my product and keep temperatures as low as possible. According to Perry and Green (50th Anniversary edition, p. 13-77) the minimum practical overhead pressure is 10 torr. This is equivalent to 0.2 psia and I have used an overhead pressure of 0.5 psia. So I suppose I could do a little optimization on pressure ($\downarrow P = \uparrow \alpha$, but higher evacuation costs. Is the better product worth it?)

The other optimization I would do involves the column that I have right now. When I started my base case, I did a Fenske-Underwood-Gilliland calculation, entered the appropriate numbers into a simulator, and

Optimization Ideas [cont'd]

simply took the result because it worked. This was the correct step to take at the time because I needed to get the base case done. However, I now need to take a closer look. The reflux ratio I am currently using seems awfully high, for example. I need to explore other combinations of column parameters. This analysis will definitely be worthwhile since the distillation column accounts for 50.3% of the total capital investment ($4.04 × 10^6 out of $8.03 × 10^6).

Also, steam and refrigeration costs for the column account for 30.3% of the yearly operating costs ($1.06 × 10^6/yr out of $3.50 × 10^6/yr).

The final (and very important) optimization variable is capacity. When I did the base case, I arbitrarily selected a capacity. This capacity will almost certainly not be the optimum. Because of different economies of scale, there will be an optimum capacity somewhere and I need to find it. I plan to do a general analysis based on the six-tenths rule (Peters + Timmerhaus, p. 169). As capacity increases, the capital investment will increase in an exponential manner. The operating costs will increase in a linear manner. Somewhere, there is an optimum.

Actually, Peters and Timmerhaus list exponents other than 0.6 depending upon the type of equipment involved. Thus, my economic analysis will be based on the capital investment being given by an equation of the form:

$$\text{Capital investment for capacity } a = \left(\begin{array}{c}\text{cost of heat}\\ \text{exchangers for}\\ \text{case } b\end{array}\right)\left(\frac{\text{capac. } a}{\text{capac. } b}\right)^{0.6} + \left(\begin{array}{c}\text{cost of tower}\\ \text{for capac. } b\end{array}\right)\left(\frac{\text{capac. } a}{\text{capac. } b}\right)^{0.62} + \left(\begin{array}{c}\text{cost of pumps}\\ \text{for capac. } b\end{array}\right)\left(\frac{\text{capac. } a}{\text{capac. } b}\right)^{0.33} + \cdots$$

As stated above, operating costs will be calculated by:

$$\text{Operating costs for capacity } a = \left(\begin{array}{c}\text{operating costs}\\ \text{for capacity } b\end{array}\right)\left(\frac{\text{capacity } a}{\text{capacity } b}\right)$$

Optimization Ideas [cont'd]

Such calculations are relatively simple. Now, I could go through all of the details of the economics (depreciation, taxes, etc.) for each different capacity, but this exercise would be an inefficient use of time. By experience and from prior courses, I know that the quantity

$$\text{Objective Function} = \text{Capital Costs} + 2 \times \text{Operating Costs}$$

is a good indicator of the economic success of a design. Hence, I shall seek to minimize this function. Once I have done so, I can insert the 'six-tenths rule' equipment costs and the linearized operating costs into my economic spreadsheet and come up with a Net Present Worth and Internal Rate of Return. That is, I'll use the above objective function to get close to the optimum, then I'll go to a slightly higher level of detail.

Once I have found what I believe to be the optimum design parameters, I'll actually go through all of the detailed calculations that I did for my base case. I may even do two or three such cases. Then I'll report the results as the optimum design.

To summarize my optimization procedure:

1. Explore the economics of not selling fuel-grade EtOH byproduct
2. Optimize the base case distillation column. As above, the C+20 rule should prove handy
3. Optimize with respect to capacity using a combination of the six-tenths rule and the C+20 rule
4. Once near the optimum, run two or three complete, detailed cases and report the best one in the design report as the optimum design.

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-8026146.46
1189000	1189000	4474444.8	0	0	0.1429	651355.42	2462776.72	985110.69	-363642.90
1189000	1189000	4474444.8	0	0	0.2449	1116283.71	3358161.09	1343264.44	2881180.36
1189000	1189000	4474444.8	0	0	0.1749	797215.27	3677229.53	1470891.81	2853552.99
1189000	1189000	4474444.8	0	0	0.1249	569309.25	3905135.55	1562054.22	2822390.58
1189000	1189000	4474444.8	0	0	0.0893	407040.16	4067404.64	1626961.86	2847482.94
1189000	1189000	4474444.8	0	0	0.0892	406584.35	4067860.45	1627144.18	2847300.62
1189000	1189000	4474444.8	0	0	0.0893	407040.16	4067404.64	1626961.86	2847482.94
1189000	1189000	4474444.8	0	0	0.0446	203292.17	4271152.63	1708461.05	2765983.75
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	2684666.88
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	2684666.88

NPW @ 12%	5.0668E+06
IRR	0.234
NPW @ IRR	2.0E-10

Net Present Worth at 12 %
$ 5.0668E+06

Internal Rate of Return =
0.234

MISTAKE. Operating costs left out

No sale of byproducts

This page represents what the economic analysis would look like if no byproduct EtOH were sold. All other factors have been kept the same. Despite this loss of income, the net present worth is still high and the internal rate of return is still greater than the allowed maximum.

Mistake!!!

Today, eight days before the deadline, I discovered a major error in all of my economic analyses to date (base case, base case without sale of ethanol, and first optimization). In a moment of gross mental ineptitude, I left out *operating costs* in my economic calculations for project years two through ten. Clearly, this oversight has a significant effect on the overall project economics. I should have realized that there was an error when the economics seemed so favorable. I have attempted to go back through my calculation book and label the pages which contain these errors. Also, the corrected page(s) follow this page. However, it is quite likely that I have not found every place where a correction is needed. I also do not have the time to rewrite large sections of my notes on the calculations. Thus, the reader is advised to keep these facts in mind while reading through my work. Luckily, I discovered this mistake today and not a week from now.

Obviously, the base case is not even close to economical. The first optimization is better, but is still unacceptable. Thus, there is still great need for optimization. I am beginning the capacity optimization now, and the economics will hopefully improve.

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-8026146.46
1189000	1189000	4474444.8	0	0	0.1429	651355.42	2462776.72	985110.69	-363642.90
1189000	1189000	4474444.8	0	0	0.2449	1116283.71	3358161.09	1343264.44	-621796.66
1189000	1189000	4474444.8	0	0	0.1749	797215.27	3677229.53	1470891.81	-649424.03
1189000	1189000	4474444.8	0	0	0.1249	569309.25	3905135.55	1562054.22	-680586.44
1189000	1189000	4474444.8	0	0	0.0893	407040.16	4067404.64	1626961.86	-655494.07
1189000	1189000	4474444.8	0	0	0.0892	406584.35	4067860.45	1627144.18	-655676.40
1189000	1189000	4474444.8	0	0	0.0893	407040.16	4067404.64	1626961.86	-655494.07
1189000	1189000	4474444.8	0	0	0.0446	203292.17	4271152.63	1708461.05	-736993.27
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	-818310.14
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	-818310.14

NPW @ 12%	-1.1598E+07
IRR	-1.000
NPW @ IRR	ERR

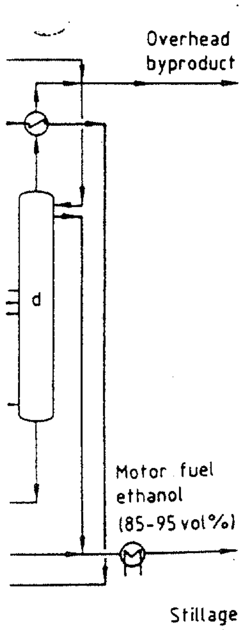
No sale of byproducts

Net Present Worth at 12 %
$ -1.1598E+07

Internal Rate of Return =
N/A

... column, which
ore. ... Here, water is
by use of benzene,
cyclohexane, or some
ne bottom stream from
consists of anhydrous,
s cooled prior to stor-
n from the dehydrating
otrope, which is com-
ors from the entrainer
nbined vapors are con-
ng two phases that are
The upper entrainer
pumped to the top of
hile the lower aqueous
r stripping column for
and ethanol.
n this system is 1.8–
l produced, depending
ion of the mash.

Ethanol. A low energy
d for distillation of the
thanol used in engines
and not gasoline. This
d mainly in Brazil; a
Figure 40. For maxi-
e fermentation feed is



ctifying column (operating

Table 22. Steam consumption and ethanol concentration in the low-energy distillation of ethanol

Parameter	Distillation product			
	Ethanol–water azeotrope	Anhydrous ethanol	Anhydrous motor fuel ethanol	Hydrous motor fuel ethanol
Maximum ethanol concentration in the feed, vol%	10	96	10	10
Ethanol concentration of the product,				
vol%	96	100	99.5	95
proof (U.S.)	192	200	199	190
Steam consumption per liter of ethanol, kg	4.1	1.4	2.2	1.2

fed into two stripper–rectifiers, the first of which operates at a pressure of ca. 0.4 MPa, the second at atmospheric pressure. Steam is used only in the high-pressure column, which takes 55–60% of the fermentation feed. The steam consumption of this system is 1.2–1.5 kg per liter of 85–95 vol% motor fuel ethanol. The overhead and fusel oil byproducts are removed and processed in a fashion similar to that described for production of anhydrous motor fuel ethanol.

5.1.4. Reduction of Energy Costs

Distillation of fermentation ethanol requires large amounts of energy. Energy costs have been cut by reducing steam consumption and by using vapor recompression systems.

Earlier columns for the distillation of ethanol from corn, wheat, or molasses were operated at atmospheric pressure. The development of new multiple-stage, high-pressure systems has reduced steam consumption by 40% compared to previous systems. The new commercial installations are based on a pressure-cascading technique and consume 3.0–4.2 kg of steam for every liter of 96 vol% ethanol produced. Earlier distillation systems required 6 kg of steam per liter of ethanol.

A modern motor fuel ethanol plant has a total energy consumption of 1.1–1.4 MJ per liter of ethanol. Steam consumption figures for the systems described in Sections 5.1.2 and 5.1.3 are summarized in Table 22 [5.6]–[5.8].

Vapor Recompression Systems. Energy costs can be reduced by up to 80% in low-pressure distillation columns by using a vapor recompression system [5.9].

In this system, compressed overhead vapor is used as the heat source for the reboiler, instead of expensive steam. The temperature of the column overhead vapor is increased by compression.

Vapor recompression can be applied to distillation columns operating at or below atmospheric pressure. It is used mainly in the distillation of motor fuel ethanol. Investment costs for ethanol distillation units equipped with a vapor recompression system are almost 50% higher than those for conventional distillation.

5.2. Nondistillative Methods

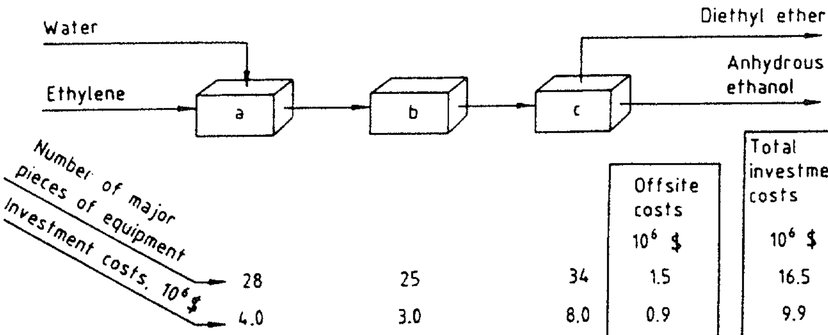
The energy required to remove water from ethanol can be reduced significantly by using methods that do not rely on distillation.

5.2.1. Solvent Extraction

Ethanol dissolves in some liquids that are virtually immiscible with water. These solubility differences can be exploited to recover ethanol from an aqueous solution by means of solvent extraction [5.10].

In the United States, the Energol Corporation employs liquid–liquid extraction with a proprietary solvent to separate ethanol from water. The solvent is then removed by distillation. This method does not require energy-intensive azeotropic distillation and thus has a low energy consumption. The energy budget for the entire plant is 3500–3700 kJ per kilogram of ethanol produced. In 1987, Energol's technique was used in four 40 000 to 45 000-L/d plants which came into operation in the mid 1980s.

The University of Pennsylvania and General Electric have developed a process that uses dibutyl phthalate [*84-74-2*] as a water-immiscible solvent for purifying ethanol. This solvent has a much higher boiling point than ethanol, and ethanol can therefore be separated in a single distillation step; solvent losses are low.



Offsite costs 10^6 $	Total investment costs 10^6 $	Ethanol-plant capacity t/a	L/d
1.5	16.5	50000	193000
0.9	9.9	25000	97000
2.5	27.5	100000	386000
3.3	36.3	150000	579000

Figure 42. Investment costs for a plant producing synthetic ethanol
a) Hydration; b) Purification; c) Distillation

age tanks) with capacities of 25 000, 50 000, 100 000, and 150 000 t/a. The breakdown of the investment for the main processing plant — the total inside battery-limit investment — is as follows:

1) Equipment and machinery, including piping, insulation, painting, and electrical installations, 50 %
2) Construction and erection, 25 %
3) Civil engineering work, 10 %
4) Engineering, costs for equipment purchase and start-up, 15 %

License fees are considered separately.

Production Costs. The annual costs (in 10^6 dollars) of synthetic ethanol produced in a 50 000-t/a plant from ethylene costing $ 350/t are as follows (with the assumption that 0.625 t of ethanol can be obtained from 1 t of ethylene):

Raw materials	11.26	(59 %)
Utilities	4.18	(22 %)
Labor	0.24	(1 %)
Depreciation, interest, maintenance	3.59	(18 %)
Total production costs	19.27	(100 %)

Payback Time. The payback time (in years) for a 50 000-t/a plant can be calculated by using the following figures (given in 10^6 $/a, with the assumption that ethanol sells at $ 450/t and that ethylene costs $ 350/t):

Annual turnover	22.50
Total production costs	19.27
Difference	3.23
Depreciation	1.65
Cash flow	4.88

$$\text{Payback time} = \frac{\text{Investment}}{\text{Cash flow}} = 3.4 \text{ years}$$

For a plant with double this capacity (100 000 t/a), the payback time is reduced from 3.4 to 2.5 years. Figure 43 shows the payback times for synthetic ethanol production in a 50 000-t/a plant as a function of the selling price for ethanol at various ethylene costs. The 1986 prices for naphtha and gasoline are indicated for comparison.

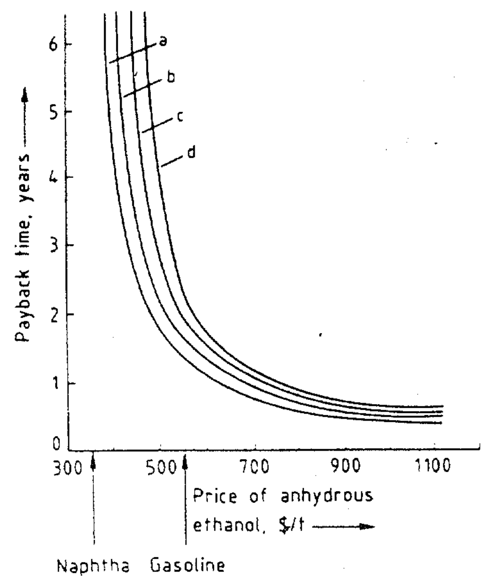


Figure 43. Price of anhydrous ethanol as a function of the payback time in a synthetic alcohol plant
Cost of ethylene (per ton): a) $ 300; b) $ 350; c) $ 400; d) $ 450
Production capacity: 50 000 t/a (19 300 L/d)

6.3. Production Ethanol

The processing tion from regeneral five major steps (for

1) Handling of raw
2) Hydrolysis pretr
3) Fermentation
4) Distillation and
5) Processing of sti tional byproduct

Approximately 1 equipment are requi almost 200 for a wo

Investment Costs mated investment co ethanol plant with (200 000 L/d); costs step and for the enti sites. The total invest plants using sucro containing raw ma cassava/manioc, or

The breakdown o vestment is similar t plants.

Production Costs Materials. The annu

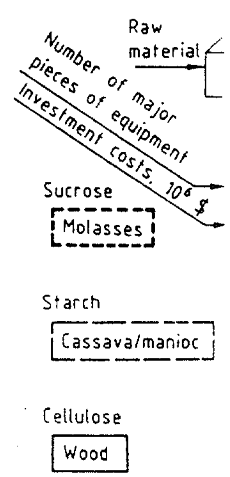


Figure 44. Investment co
a) Handling of raw mat
processing
* Production capacity:

Ethanol Research

I have done a little more looking into the ethanol sales situation, and I believe that to sell my product, it will have to contain at least 95% (by volume) ethanol. Because of the ethyl acetate present, I think I can only sell it as fuel grade EtOH. The next page is from Ullman's Encyclopedia of Industrial Chemistry and it shows some typical EtOH specifications.

I'll convert volume % to wt% assuming a 0.8 specific gravity for ethanol. On a basis of 100 ft^3,

$$95\ ft^3\ EtOH \rightarrow 95\ ft^3 \cdot 62.3(0.8)\frac{lb}{ft^3} = 4734.8\ lb\ EtOH$$

$$5\ ft^3\ H_2O \rightarrow 5\ ft^3 \cdot 62.3\frac{lb}{ft^3} = 311.5\ lb\ H_2O$$

thus,

$$wt.\ frac\ EtOH = \frac{4734.8}{4734.8 + 311.5} = 0.938$$

This is what I have right now based on a feed rate of 99.4586 lb/hr:

Stream No.	3	4	5
Stream Name	Flash Liquid	Distillate	Bottoms
Temp F	79.2410*	-86.6938*	125.8263
Pres psia	14.0000	0.5000*	2.0000*
Enth MMBtu/h	-0.658573	-0.0126456	-0.641914
Vapor mole fraction	0.00000	0.00000	0.00000*
Total lbmol/h	5.3603	0.1010	5.2593
Total lb/h	99.4586	4.4586	95.0000
Total std L ft3/hr	1.6117	0.0888	1.5229
Total std V scfh	2034.11	38.32	1995.79
Flowrates in lb/h			
Water	94.8316	0.1680	94.6636
Ethanol	4.1674	4.0940	0.0734
Carbon Dioxide	0.0484	0.0484	0.0000
Pyruvic Acid	0.2100	0.0000	0.2100
Ethyl Acetate	0.2012	0.1482	0.0530
Air	0.0000	0.0000	0.0000

The purity of the distillate is $\frac{4.094\ lb/h}{4.4586\ lb/h} = 0.918$

If I add a couple of more trays in the rectifying section, maybe I can achieve the 0.938 specification.

Optimization I

New Simulations (trying to hit 95 vol% EtOH)

I'll try to hit 0.938 with a few more trials. Information from Chem-Cad:

Trial 1: 5 more stages (for a total of 64) in rectifying section

$x_{EtOH} = 0.9245$

Trial 2: Move the feed stage down 3 stages (keeping the 5 extra stages)

$x_{EtOH} = 0.92718$

Trial 3: 5 more trays (for a total of 69)

$x_{EtOH} = 0.93107$ (getting close)

Trial 4: Move feed stage down 2 more

$x_{EtOH} = 0.93209$

Trial 5: Move feed down 5 more stages

$x_{EtOH} = 0.93354$ (almost there)

Trial 6: 5 more trays (total of 74)

$x_{EtOH} = 0.93448$

Hmm. Things don't appear to be improving much. I really don't want to raise the reflux any higher (it's already at 40). I'll try to bracket the number of trays I need by adding a lot more all at once.

Trial 7: 20 more trays (total of 94).

$x_{EtOH} = 0.93469$

Trial 8: Move feed stage down 5 trays

$x_{EtOH} = 0.93468$

I give up. I've hit a pinch I think. I've added 35 trays (I started out with 59) and I've experimented with feed stage location, and I still can't get the separation I need. I could spend time trying to get the desired composition, but I don't think this would be an efficient use of time. Based on these simulations, I really don't think the new tower would be economical anyway.

More (and better) simulations

Having decided not to pursue the EtOH route, I can now go the opposite way - I can try to save money on my column. Since now I'm unconcerned about the composition of my distillate, I only need to care about the bottoms product. I want it to have a low alcohol content but keep most of the water. Actually, it's OK for it to lose a little water since the water from my first steam ejector (X-1) is getting mixed in anyway. The amount of steam is 28.9 lb/h. Scaling for my simulation, I can afford to lose: *

$$28.9\ \text{lb/h} \cdot \frac{99.4586}{100 \cdot 98.43} = 0.292\ \text{lb/h of } H_2O$$

Procedure: I'll reset to my original simulation conditions, then start subtracting trays and lowering the reflux ratio to see how much money I can save up until the point where 0.292 lb/h of H_2O are lost. Then I can evaluate the impact on the design

Trial 1

Original conditions
lb H_2O lost = 0.1680

Trial 2

Reflux ratio ↓ to 20
lb H_2O lost = 0.1823

Trial 3

Reflux ratio ↓ to 10
lb H_2O lost = 0.2030

Trial 4

Reflux ratio ↓ to 5
lb H_2O lost = 0.2379

Trial 5

Reflux ratio ↓ to 2.5
lb H_2O lost = 0.7347 Too much!

* If I lose too much, the beer will be over concentrated. If I don't lose enough, it will taste watered-down.

More (and better) simulations [cont'd]

Trial 6

I'll try a reflux ratio of 4.
lb H_2O lost = 0.2577

Trial 7

Reflux ratio ↓ to 3.0
lb H_2O lost = 0.4883 too much!

Trial 8

Reflux ratio ↑ to 3.5
lb H_2O lost = 0.2948 (slightly too much)

* Conclusion: I'll go with the reflux ratio of 4.0, since I'd rather have the beer a little watered down than super-concentrated. However, since this R.R. is so dramatically different from my initial Fenske - Underwood - Gilliland calculation, I'll do a McCabe - Thiele solution to confirm.

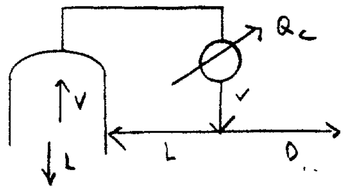
Below are the results of my new simulation for a reflux ratio of 4.0

Stream No.	3	4	5
Stream Name	Flash Liquid	Distillate	Bottoms
Temp F	79.2410*	-85.1061*	125.8405
Pres psia	14.0000	0.5000*	2.0000*
Enth MMBtu/h	-0.658573	-0.0130571	-0.641509
Vapor mole fraction	0.00000	0.00000	0.00000*
Total lbmol/h	5.3603	0.1051	5.2551
Total lb/h	99.4586	4.4586	95.0000
Total std L ft3/hr	1.6117	0.0887	1.5230
Total std V scfh	2034.11	39.89	1994.21
Flowrates in lb/h			
Water	94.8316	0.2577	94.5739
Ethanol	4.1674	4.1124	0.0550
Carbon Dioxide	0.0484	0.0484	0.0000
Pyruvic Acid	0.2100	0.0000	0.2100
Ethyl Acetate	0.2012	0.0401	0.1611
Air	0.0000	0.0000	0.0000

I've been able to reduce the reflux ratio by a factor of 10 and still keep little alcohol in the bottoms. Also, I'm losing less ethyl acetate.

McCabe-Thiele Calculations

External reflux ratio = 4.0 = L



By material balance, I know that:

$$L/D = \frac{L}{V-L}$$ (see p. 328 of Henley + Seader)

$$L/D = \frac{L/V}{1-L/V}$$

Solve for L/V:

$$(1-L/V)4.0 = L/V$$

$$4.0 = 5\,L/V$$

$$L/V = 0.8$$

Now I need to know where to put the feed. I'll approximate the feed as 0.042 wt% EtOH (see "Composition of Beer Mixture") and the rest water. I'll also assume a saturated liquid feed. I know this is not the case, but for this calculation, which is a check, it's close enough.

Convert to mole fractions on a basis of 100 lb of feed:

$$4.2 \text{ lb EtOH} \,\Big|\, 46.069 \text{ lb/lbmol} = 0.09117 \text{ lbmol}$$

$$95.8 \text{ lb } H_2O \,\Big|\, 18.016 \text{ lb/lbmol} = 5.317 \text{ lbmol}$$

$$x_{EtOH,\,Feed} = \frac{0.09117}{5.317} = \underline{0.0171 \text{ lbmol/lbmol}}$$

Now, I'll convert the distillate and bottoms flow rate from my simulation:

$$\text{Distillate} = \frac{4.1124}{4.4586} = 0.9224 \text{ wt\%}$$

$$92.24 \text{ lb} \,\Big|\, 46.069 \frac{\text{lb}}{\text{lbmol}} = 2.00 \text{ lbmol EtOH}$$

$$7.76 \text{ lb } H_2O \,\Big|\, 18.016 \text{ lb/lbmol} = 0.431 \text{ lbmol } H_2O$$

McCabe-Thiele [cont'd]

Mole frac EtOH = $\frac{2}{2+0.431}$ = $\underline{0.822 \text{ lbmol/lbmol}}$

For the bottoms:

$$\frac{0.0550}{95} = 5.789 \times 10^{-4} \text{ wt. frac} = 0.05789 \text{ wt\%}$$

$$0.05789 \text{ lb} \,|\, 46.016 \text{ lb/lbmol} = 0.00126 \text{ lbmol EtOH}$$

$$(10.0 - 0.05789) \text{ lb } H_2O \,|\, 18.016 \text{ lb/lbmol} = 5.5474 \text{ lbmol } H_2O$$

$$\text{Mole frac EtOH} = \frac{0.00126}{5.5474} = 2.27 \times 10^{-4} \sim \underline{0.0 \text{ lbmol/lbmol}}$$

I also need to determine the slope of the operating line below the feed, $\bar{L}/\bar{V}$. I'll use a basis of 100 moles of feed and, of course, the constant molal overflow assumption.

First, solve for D:

$$100 \text{ mol feed} \cdot 0.0171 = 0.0(B) + 0.822(D)$$

$$\Rightarrow D = 2.08 \text{ mol}$$

Using a reflux ratio of 4.0,

$$L = 2.08(4) = 8.32 \text{ mol}$$

By material balance,

$$V = 8.32 + 2.08 = 10.4 \text{ mol}$$

$$\text{So, } L/V = 0.8 = \frac{8.32 \text{ mol}}{10.4 \text{ mol}}$$

When the feed is added, the liquid flow increases according to the CMO assumption. With 100 moles of feed, McCabe - Thiele [cont'd]

$$\bar{L}/\bar{V} = \frac{100 \text{ mol} + 8.32 \text{ mol}}{10.4 \text{ mol}} = \underline{\underline{10.4}}$$

Stepping off stages on the xy diagram on the next page, I come up with:

8 stages above the feed (not accounting for efficiency)

Too small to tell below the feed.

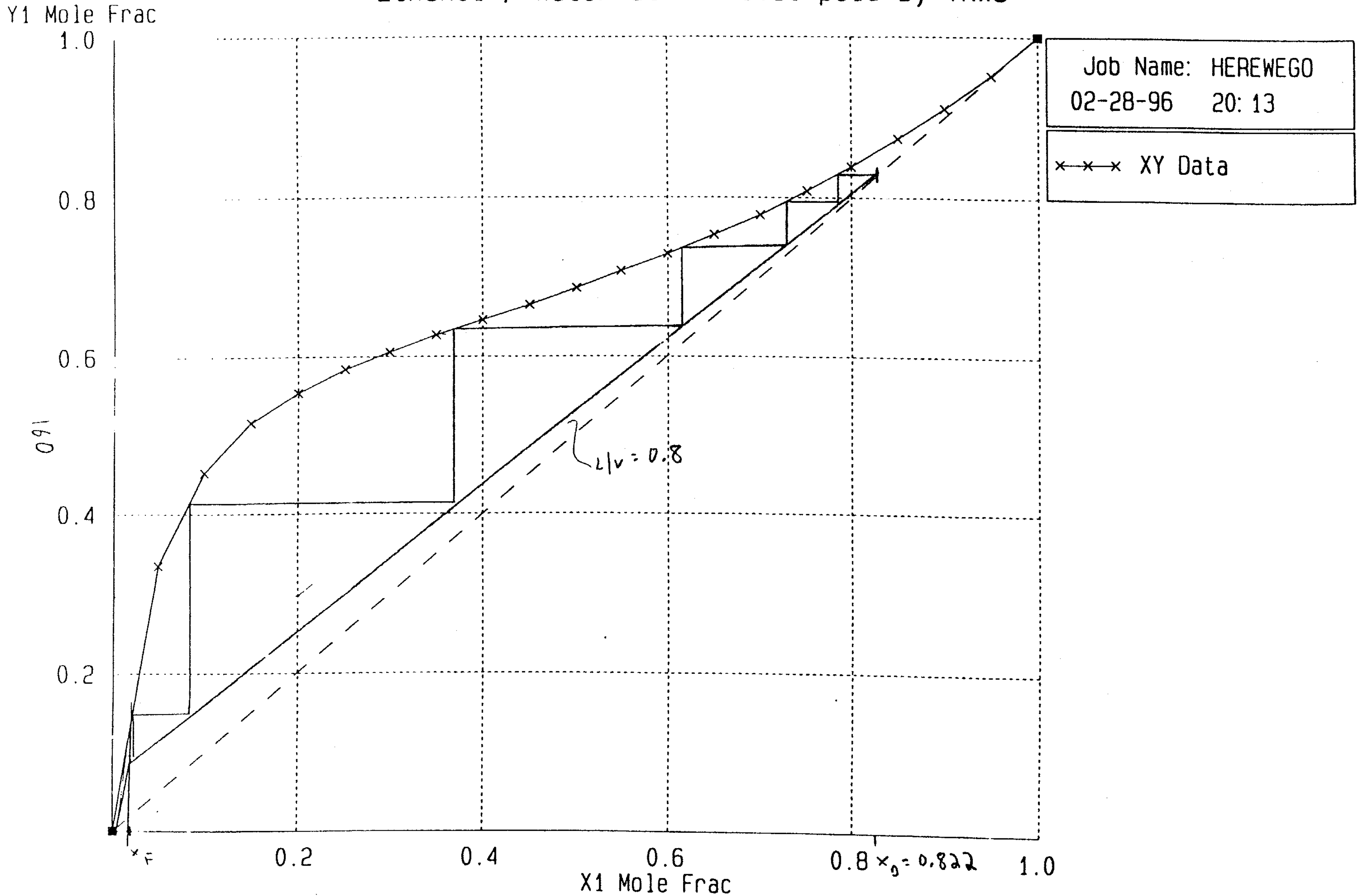
Based on this, I'll declare that <u>my simulation is confirmed</u>.

("Distillation Calculations", p. 6) Using the stage efficiency of 0.43 for the top part of the column that I calculated earlier, this works out to:

$$8/0.43 = 18.6 \text{ stages above the feed.}$$

In my simulation, I used 20. This <u>further confirms my simulation</u>

Ethanol / Water at 0.50 psia By TKWS



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Optimization Case I - Base Case with Better Tower

Now that I've altered my simulation, I can see that this second tower design is much better. The reflux ratio is only 4.0, compared to 40 before. Therefore, I want to use this case rather than the base case. A lower reflux ratio will mean significant decreases in refrigeration costs and significant decreases in reboiler duty. It will probably also lower the tower diameter due to the decrease in vapor rate. This means that I will have to re-do several calculations. Just glancing at the flowsheet, the units affected are:

~~T-1~~	tower
~~F-2~~	reflux drum
~~X-2~~	ejector system
~~X-3~~	
~~E-5~~	
~~E-6~~	
~~E-4~~	Condenser
~~P-4A/B~~	Reflux pumps
~~E-7~~	Reboiler
E-9	Hot H_2O heater
~~P-3A/B~~	Bottoms pumps
~~P-5A/B~~	Hot H_2O circulation pumps
~~F-3~~	Recombiner (slight change)

The streams whose material balance will change are:

~~10~~, ~~11~~, ~~21~~, ~~13~~, ~~31~~, ~~32~~, ~~30~~, 7, ~~23~~, ~~19~~, ~~27~~, ~~9~~, ~~14~~, ~~15~~, ~~16~~, ~~20~~, and ~~28~~

I'll start by collecting the TOWER INFORMATION:

Note: these values are all based on 100 lb/h of feed and can be converted, when necessary to actual values of my case by multiplying by 98.43

At tray 59

$L = 20$ lb/h $\quad V = 22$ lb/h
$M_L = 42.05$ $\quad M_V = 42.41$
$\rho_L = 50.54$ lb/ft³ $\quad \rho_V = 0.00382$ lb/ft³
$\sigma = 24.754$ dyne/cm

At tray 45

$L = 16$ lb/h $\quad V = 21$ lb/h
$M_L = 32.53$ $\quad M_V = 35.53$
$\rho_L = 52.2$ lb/ft³ $\quad \rho_V = 0.00532$ lb/ft³
$\sigma = 29.971$ dyne/cm

Optimization Case I [cont'd]

My earlier calculations for the base case (see "Sizing Calculations", p. 28) showed that the diameter required increased steadily up the column because of: 1. Increasing vapor rate
2. Decreasing vapor density

My new column should follow the same pattern. So I only need to size the diameter once – at the top of the column. Just for my own mental health, though, I'll go through the calculation at tray 45 to confirm the trend.

Following the procedure established earlier:

Tray 45 Calculations

$$F_{ST} = (29.971/20)^{0.2} = 1.084$$

$$F_F = 0.6$$

$$F_{HA} = 1.0$$

$$F_{LV} = \frac{16 \text{ lb/h}}{21 \text{ lb/h}} \left(\frac{\cancel{98.43}}{\cancel{98.43}}\right)\left(\frac{0.00532 \text{ lb/ft}^3}{52.20 \text{ lb/ft}^3}\right)^{0.5}$$

$$= 0.0077$$

From Figure 13-3 of Henley + Seader, $C_F = 0.39$ ft/s

Now,
$$C = 1.084(0.6)(1.0)(0.39 \text{ ft/s})$$
$$= 0.254 \text{ ft/s}$$

So,
$$u_f = 0.254 \frac{\text{ft}}{\text{s}} \left(\frac{52.2 - 0.00532 \text{ lb/ft}^3}{0.00532 \text{ lb/ft}^3}\right)^{1/2}$$

$$\underline{= 25.16 \text{ ft/s}}$$

Since $F_{LV} \le 0.1$, $A_d/A = 0.1$

Optimization Case I [cont'd]

Now:

$$D_T = \left[\frac{4(21 \text{ lb/h})(98.43)\left(\frac{1 \text{ h}}{3600 \text{ s}}\right)}{0.85\left(25.16 \frac{ft}{s}\right)\pi(1-0.1)\left(0.00532 \frac{lb}{ft^3}\right)} \right]^{1/2}$$

$$\underline{\underline{D_T = 2.67 \text{ ft}}}$$

Tray 59 Calculations

$$F_{ST} = (24.754/20)^{0.2} = 1.044$$

$$F_F = 0.6$$

$$F_{HA} = 1.0$$

$$F_{LV} = \frac{20 \text{ lb/h}}{22 \text{ lb/h}} \left(\frac{\cancel{98.43}}{\cancel{98.43}}\right)\left(\frac{0.00382 \text{ lb/ft}^3}{50.54 \text{ lb/ft}^3}\right)^{1/2}$$

$$= 0.008$$

from Fig 13-3 of H+S, $C_F = 0.39$ ft/s.

Now,

$$C = 1.044(0.6)(1.0)(0.39 \text{ ft/s})$$

$$= 0.244 \text{ ft/s}$$

So:

$$U_f = 0.244 \text{ ft/s} \left(\frac{50.54 - 0.00382}{0.00382}\right)^{1/2}$$

$$= \underline{28.06 \text{ ft/s}}$$

Since $F_{LV} \leq 0.1$, $A_d/A = 0.1$

Opt. Case I [cont'd]

Now:

$$D_T = \left[\frac{4(22\ \mathrm{lb/hr})(98.43)\left(\frac{1\ \mathrm{hr}}{3600\ \mathrm{s}}\right)}{0.85\left(28.06\ \frac{\mathrm{ft}}{\mathrm{s}}\right)\pi(1-0.1)\left(0.00382\ \frac{\mathrm{lb}}{\mathrm{ft^3}}\right)} \right]^{1/2}$$

$$D_T = 3.06\ \mathrm{ft}$$

Conclusion: I'll go with the higher value and round to

$$\boxed{\text{Tower diameter} = 3.5\ \mathrm{ft}}$$

I'll recost the tower right now using Fig 16-28 of Peters and Timmerhaus with stainless steel sieve trays. The tower diameter is:

42 inches

The column height is still 135 ft tall. From the figure, the 1990 cost is:

$1900 | ft

1990 cost = $1900 | ft · 135 ft = $256 500

In 1999 dollars,

1999 cost = 256 500 (1.1646)

$$\boxed{1999\ \text{Cost} = \$298\,720}$$

F-3 Reconbiner

The mass flow rate to the reconbiner is the same, and the density is virtually unchanged (61.607 lb/ft3). Thus, it does not need to be resized.

Opt. I [cont'd] → Ejector Sizing, X-2 and X-3

The new column has a smaller volume, a smaller air leakage rate, and will require different ejectors.

$$\text{Volume} = \frac{\pi}{4}(3.5 \text{ ft})^2 (135 \text{ ft}) = 1299 \text{ ft}^3$$

From reference 4,

$$MAL = 0.1451 \, (1299)^{0.6617}$$

$$\underline{= 16.7 \text{ lb/h} \quad \text{Air leakage}}$$

The amount of CO_2 is still 195.3 lb/h.

As before,

$$AEL = \frac{16.7 \text{ lb/h}}{(1.0)(1.0)} + \frac{195.3 \text{ lb/h}}{1.0(1.18)} = \underline{182.6 \text{ lb/h}}$$

from reference 3 → From Fig 9-11, the r value (25.85 torr, 2 stages w/ intercondensers, 90°F cw) is:

$$r \approx 10 \text{ lb steam / lb air}$$

Using eq. 9-5 of reference 3,

$$w_s = (AEL)(r)(MPC)(SC)$$

Assume
$MPC = 1.0$
$SC = 1.0$

$$w_s = (182.6 \text{ lb/h})(10)(1.0)(1.0)$$

$$= 1826 \text{ lb/hr}$$

Per stage = 1826/2 = 913 lb/hr per stage

From Fig 9-10, size correction factor = 1.0

So, I will specify:

$$\boxed{\begin{array}{l}\text{Two-stage ejector system} \\ \quad 913 \text{ lb/hr} \quad 100 \text{ psig steam per stage}\end{array}}$$

Opt. I [cont'd]

Ejector Costs

As before,

$$1986 \text{ installed cost} = \$16000\left[NS + 2(NC)\right]\left(\frac{SCON}{1000}\right)^{0.35}$$

$$= 16000\left[2 + 2(2)\right]\left(\frac{913}{1000}\right)^{0.35}$$

$$= \$92990$$

$$1999 \text{ cost} = \$92990\,\underbrace{(1.2799)}_{\text{Inflation}}\,\underbrace{(1.2)}_{\text{surface condensers}}$$

$$\boxed{\begin{array}{l}\text{X-2 and X-3 combined, installed,} \\ \text{1999 cost} \\ \qquad = \$142821\end{array}}$$

E-5 and E-6 Ejector condenser sizing

Tube side: CW
$T_{in} = 90°F \quad T_{out} = 120°F$

Shell side: Condensing steam
$T_{in} = 338°F \quad T_{out} = 338°F, \quad \Delta H^{lv} = 880.6 \frac{Btu}{lbm}$

Flow rate = 913 lb/hr

Here, $U = 160 \frac{Btu}{hr\ ft^2\ °F}$

The temperature profile is linear, so:

$\Delta T_{lm} = 232.6\ °F$ (from earlier)

Opt. I [cont'd]

Now,

$$A_s = \frac{(913 \ lb/h)(880.6 \ Btu/lb)}{160 \ \frac{Btu}{hr \cdot ft^2 \, °F} \cdot 232.6°F} = 21.6 \ ft^2$$

So,

$$\boxed{2 \text{ exchangers}, \ 21.6 \ ft^2 \text{ each}}$$

Now, from steam tables, $H^l_{90} = 58.02$ Btu/lbm
$H^l_{120} = 87.97$ Btu/lbm

So:

$$(913 \ lb/h)(880.6 \ Btu/lb) = \left(87.97 - 58.02 \ \frac{Btu}{lb}\right)\dot{m}$$

$\Rightarrow$ $\dot{m} = 26844$ lb/h CW for each

E-4 Condenser Sizing

Tube side: Column overhead (less non-condensibles), streams 9 + 14
$T_{in} = 52.8$ °F $\quad T_{out} = 52.6$ °F

Shell side: NH_3 refrigerant
$T_{in} = 24$ °F $\quad T_{out} = 24$ °F, $\Delta H^{lv} = 549.8$ Btu/lb

Calculated heat duty = 0.19414 mmBtu/h
(from ChemCad)

In this case, $U = 160 \ \frac{Btu}{hr \, ft^2 \, °F}$

The temperature profile is close enough to linear for these calculations (see earlier calculations).

$$\Delta T_{lm} = \frac{(52.8 - 24) - (52.6 - 24)}{\ln \frac{52.8 - 24}{52.6 - 24}} = 28.7 \ °F$$

Opt. I [cont'd]

So,

$$A_I = \frac{0.19414 \times 10^6 \ \text{Btu/hr}}{160 \ \frac{\text{Btu}}{\text{hr ft}^2 \ ^\circ\text{F}} \cdot 28.7\ ^\circ\text{F}} = \boxed{42.3 \ \text{ft}^2}$$

From Fig. 15-15 of P+T,

Purchased cost = \$2800

1999 cost = $2800(1.1646) = \boxed{\$3261}$

Also,

$$0.19414 \times 10^6 \ \frac{\text{Btu}}{\text{hr}} = 549.8 \ \frac{\text{Btu}}{\text{hr}} \cdot \dot{m}$$

$\Rightarrow \dot{m} = 353.1$ lb/hr NH_3 circulation required

P-7A/B Reflux Pumps, Stream 15

The flow rate for these pumps is significantly lower now. Everything else is similar.

Suction P = 0.5 psia
Discharge P = 3.5 psia

$$NPSH_A = (0.5 - 0.89) \times \frac{2.31}{0.802} = -1.1 \ \text{ft}$$

I need to ensure that this pump is 2 ft below the reflux drum. Shouldn't be a problem.

$$\text{gpm} = \frac{2170.6}{(500)(0.802)} = 5.41 \ \text{gpm}$$

$$h = 3.0(2.31)/0.802 = 8.6 \ \text{ft}$$

From Fig. 14-40 of P+T, this service requires:

1¼ × 1¼ pump, ≈ $\frac{1}{4}$ hp

Opt. I [cont'd]

Raw cost = \$690

$$1999 \text{ cost} = \underbrace{690}_{}(\underset{\text{stainless}}{1.8})(\underset{\text{inflation}}{1.1646})(\underset{\text{spare}}{2}) = \boxed{\$2893}$$

For 1/4 hp, operating cost calculated from P-2A/B $= \boxed{\$77.4/\text{yr}}$

P-3A/B Bottoms Pumps

The flow rate to these pumps will also change appreciably.
New flow rate = 10853.5 lb/hr, s.g. = unchanged

Suct. P = 2.0 psia
Discharge P = 8.0 psia

$$NPSH_A = (2 - 5.79) \times \frac{2.31}{1.0} = -8.8 \text{ ft}$$

I need to make sure I have 10 ft between column exit and pump inlet. Since most columns have a large skirt, this shouldn't be a problem.

$$gpm = \frac{10853.5}{500(1.0)} = 21.7 \text{ gpm}$$

$$h = (8-2) \times 2.31/1.0 = 13.86 \text{ ft}$$

From Fig. 14-40 of P+T, this service requires:

1½ × 1½ pump, ≈ 1/4 hp (go to next one up)

Raw cost = \$860

$$1999 \text{ cost} = 860(1.8)(1.1646)(2)(\underset{\text{temp. factor}}{1.05}) = \boxed{\$3786}$$

Operating cost, from above for 1/4 hp $= \boxed{\$77.4/\text{yr}}$

Opt. I [cont'd]

P-5A/B, Hot H_2O circulation pumps, Stream 20

The throughput of this pump will also change significantly.

Suct. P = 14.7 psia
Discharge P = 31 psia (3 exchangers + line loss)

$$NPSH_A = (14.7 - 2.61) \times \frac{2.31}{0.987} = 28.3 \text{ ft} \quad \checkmark OK$$

$$gpm = \frac{109370}{500(0.987)} = 221.8 \text{ gpm}$$

$$h = (31 - 14.7)(2.31) \,|\, 0.987 = 38.1 \text{ ft}$$

From Fig. 14-40 of Peters and Timmerhaus, this service requires

3 x 3 pump, ≈ 5 hp

Raw cost = \$1395

$$1999 \text{ cost} = 1395 \underbrace{(1.1646)}_{\text{inflation}} \underbrace{(2)}_{\text{spare}} \underbrace{(1.05)}_{\text{temperature}} = \boxed{\$3412}$$

Operating cost =

$$5 \text{ hp} \cdot 0.746 \frac{kW}{hp} \cdot 8400 \frac{hr}{yr} \cdot 0.0494 \frac{\$}{kWhr}$$

$$= \boxed{\$1547.8/yr}$$

From Fig 14-54 of P+T, driver cost = \$380

$$1999 \text{ cost} = (\$380)(2)(1.1646) = \boxed{\$885}$$

Opt. I [cont'd]

E-9 Hot H_2O Heater

Tube side : heating water, streams 20 and 28

$T_{in} = 136\,°F \qquad T_{out} = 150\,°F$

Shell side : condensing steam @ 100 psig

$T_{in} = 338\,°F \qquad T_{out} = 338\,°F \quad , \Delta H^{v} = 880.6 \frac{Btu}{lb}$

Heat duty = 109370 lb/h $(118 - 104$ Btu/lb$)$

↑ Hot H_2O circulation ↑ enthalpies from steam tables at 136°F + 150°F

= 1,531,180 Btu/hr

Here, $U = 160 \frac{Btu}{hr\text{-}ft^2\,°F}$

The temperature profile is of course linear.

$$\Delta T_{lm} = \frac{(338-150)-(338-136)}{\ln \frac{338-150}{338-136}} = 194.9\,°F$$

So,

$$A_I = \frac{1531180 \text{ Btu/hr}}{160 \frac{Btu}{hr\text{-}ft^2\,°F} \cdot 194.9\,°F} = \boxed{49.1 \text{ ft}^2}$$

From Fig. 15-15 of Peters + Timmerhaus,

Purchased cost = \$3000

1999 cost = $3000\,(1.1646)$ ∴ $\boxed{\$3494}$

I'll also find the required steam flow rate:

$$\left(880.6 \frac{Btu}{lb}\right) \dot{m} = 1,531,180 \text{ Btu/hr}$$

$\Rightarrow$ $\dot{m} = 1738.8$ lb/hr steam required

Opt. I [cont'd]

F-2 Reflux Drum Sizing

The new flow to this vessel (stream 14) is

$$L = 2170.6 \text{ lb/hr}$$
$$\rho_L = 50.487 \text{ lb/ft}^3$$

Allowing for a 1/2 full vessel with 5 minute liquid residence time,

$$V_v = 2Lt/\rho_L$$
$$= 2\left(2170.6 \text{ lb/hr}\right)\left(5 \text{ min}\right)\left(\frac{1 \text{ hr}}{60 \text{ min}}\right) \Big/ 50.487 \text{ lb/ft}^3$$
$$= 7.17 \text{ ft}^3$$

I want $H/D = 4$, so

$$D = H/4 = \left(\frac{V_v}{\pi}\right)^{1/3}$$
$$= \left(\frac{7.17 \text{ ft}^3}{\pi}\right)^{1/3}$$
$$= 1.316 \text{ ft} \rightarrow \text{round to } 1.5 \text{ ft}$$

Therefore, $H = 6$ ft

Actual volume of vessel $= \frac{\pi}{4}\left(1.5 \text{ ft}\right)^2\left(6 \text{ ft}\right) = 10.6 \text{ ft}^3$

$$10.6 \text{ ft}^3 \cdot 7.48 \text{ gal/ft}^3 = 79.3 \text{ gallons}$$

I will specify, then:

6 ft length
1.5 ft diameter
79.3 gallons

Cost: this vessel size is out of the range of figure 14-56 of Peters and Timmerhaus, so I'll use 0.6 rule to scale it to my base case reflux drum, which was 553.5 gallons and cost \$25738 (1999).

$$\text{New cost} = \$25738\left(\frac{79.3}{553.5}\right)^{0.6}$$

$$= \boxed{\$8022}$$

Material Balance Changes (Opt. I)

Now that more water is going overhead, there are a few changes I need to make in the material balance. Streams 10, 11, 14, 15, 16, 20, 23, 27, and 28 have already been changed.

Stream 21

Identical to stream 11, but at a higher pressure

Stream 13

Identical to stream 21, but at a lower temperature

Streams 31, 32, 30, and 1

I'll need to compute again how much regular beer I can divert. As before I'll specify a maximum of 0.4 vol% EtOH in the final product.

First, I'll find the volumetric flow of EtOH already in the preliminary product:

$$\frac{9.01 \text{ lb/hr}}{\underset{\text{conversion factor}}{500}\ \underset{\text{sp.gr. of EtOH}}{(0.8)}} = 0.023 \text{ gpm}$$

Total flow of stream 13 = 18.8 gpm

Therefore,

$$\frac{0.023 + x}{18.8} = 0.4\% = 0.004$$

Solving, $x = 0.0522$ gpm

This is the amount of EtOH I can legally add

Converting to mass,

$$0.0522(500)(0.8) = 20.88 \text{ lb/hr}$$

Material Balance Changes [cont'd] (Opt. I)

Stream 30 [cont'd]

From my earlier definition of the components of beer, I know that it contains

$$\frac{4.21 \text{ lb EtOH}}{100.2 \text{ lb beer}}$$

Therefore, I can add as stream 10

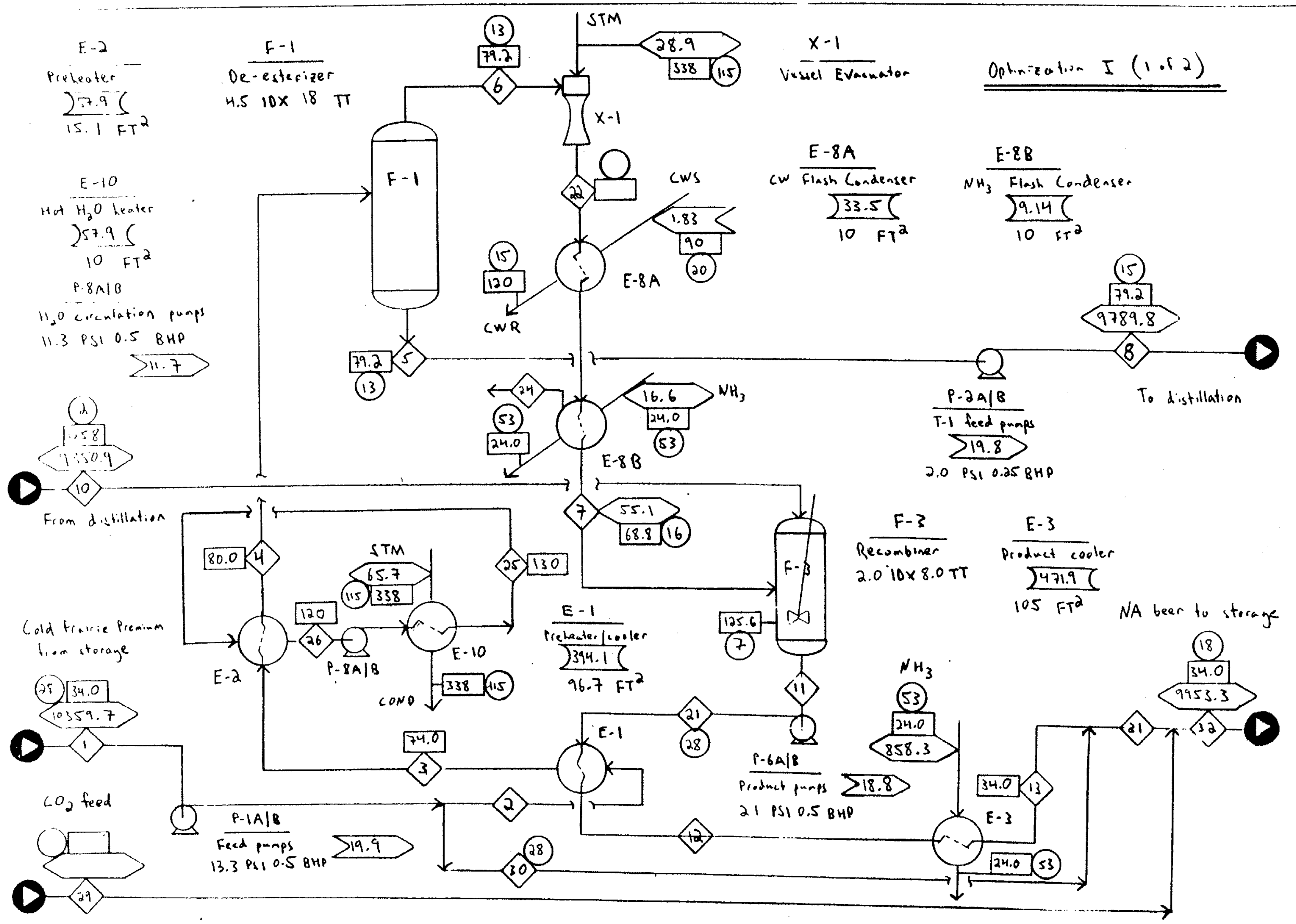
$$20.88 \text{ lb/hr EtOH} \cdot \frac{100.2 \text{ lb beer}}{4.21 \text{ lb EtOH}} = \underline{\underline{497.0 \text{ lb beer/hr}}}$$

Stream 1: Stream 1 is simply Stream 30 + Stream 2, which remains unchanged

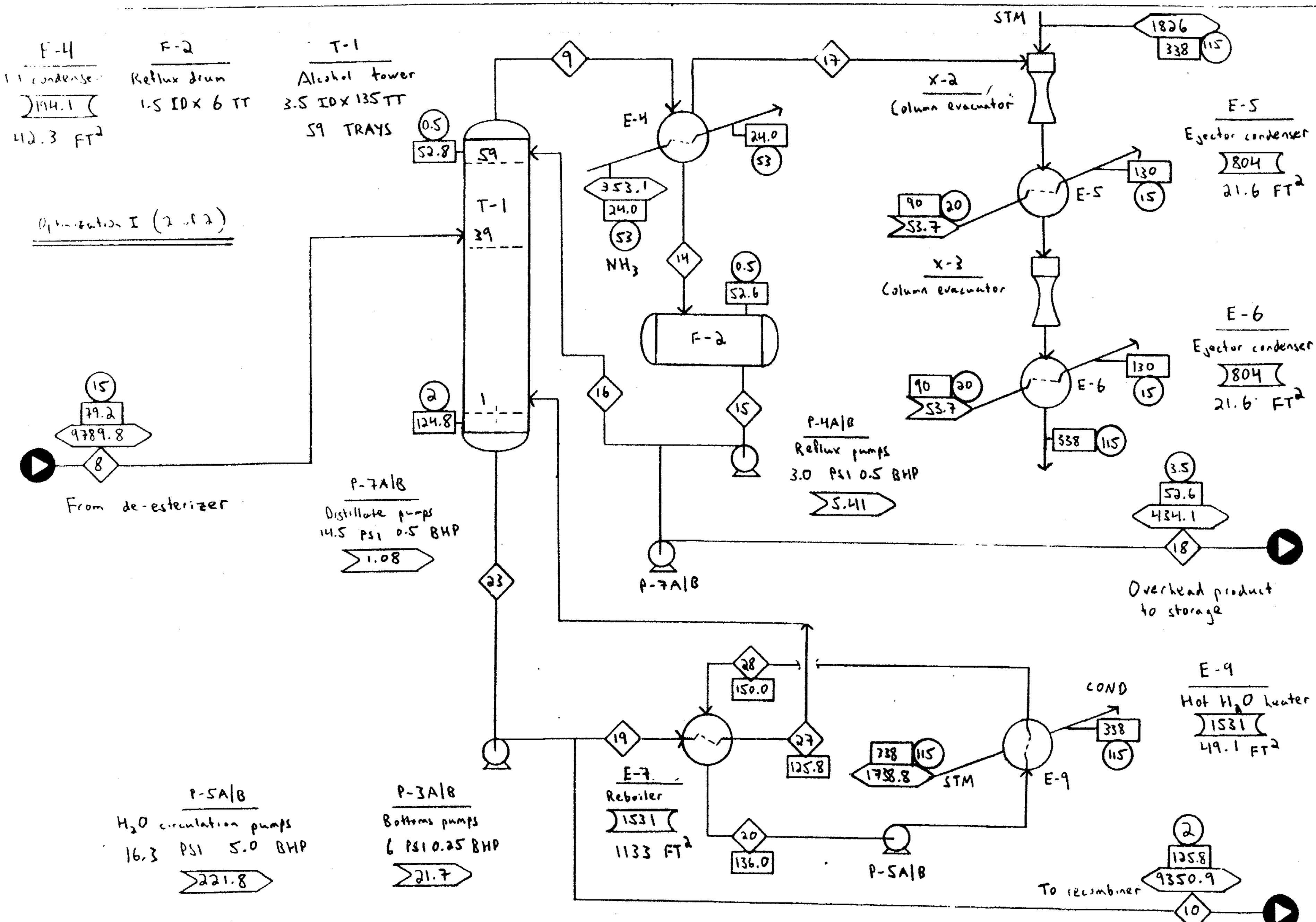
Stream 31: the sum of Stream 30 and 13

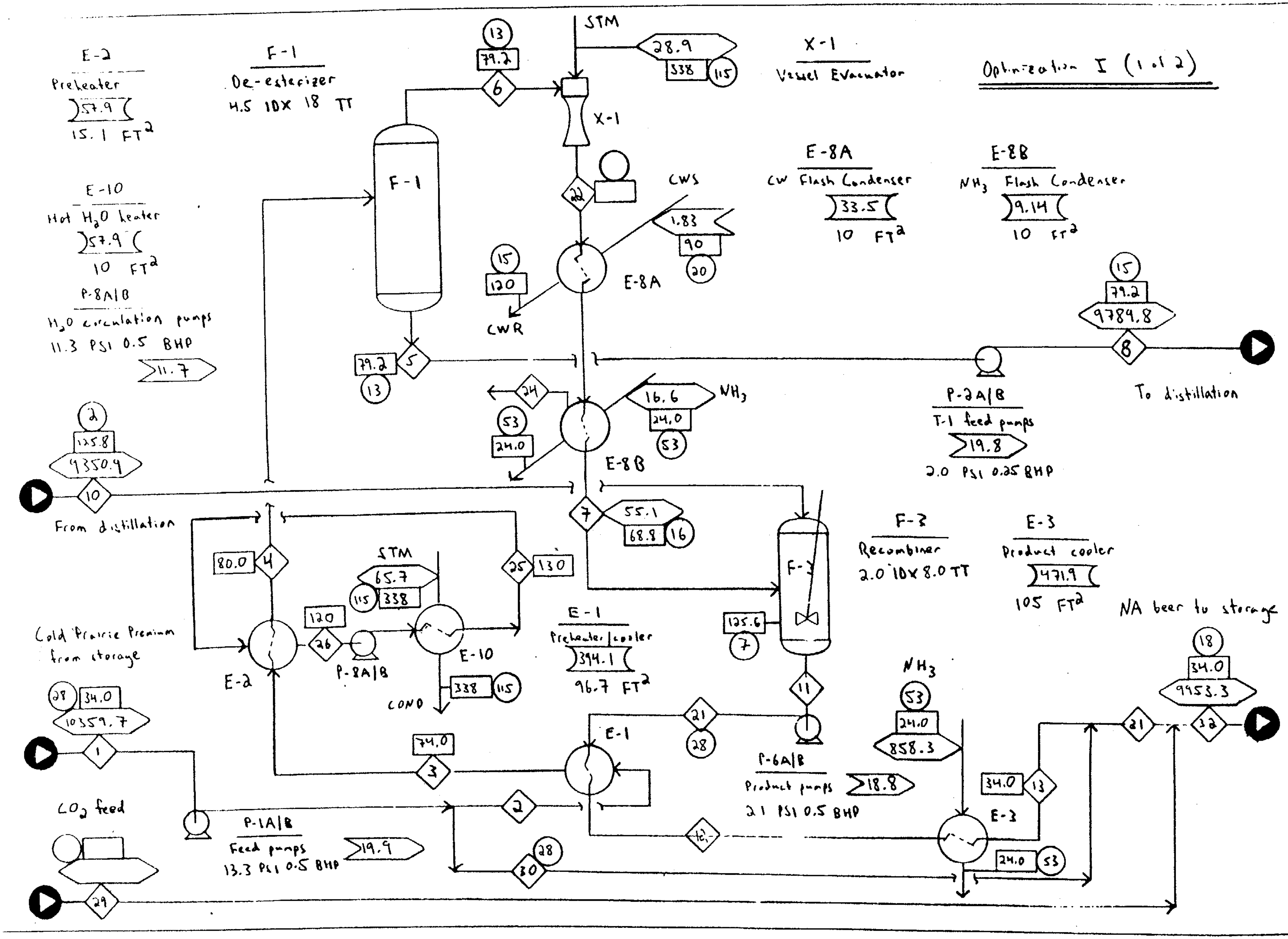
Stream 32: Stream 31 with the CO_2 added

All other streams remain the same!!

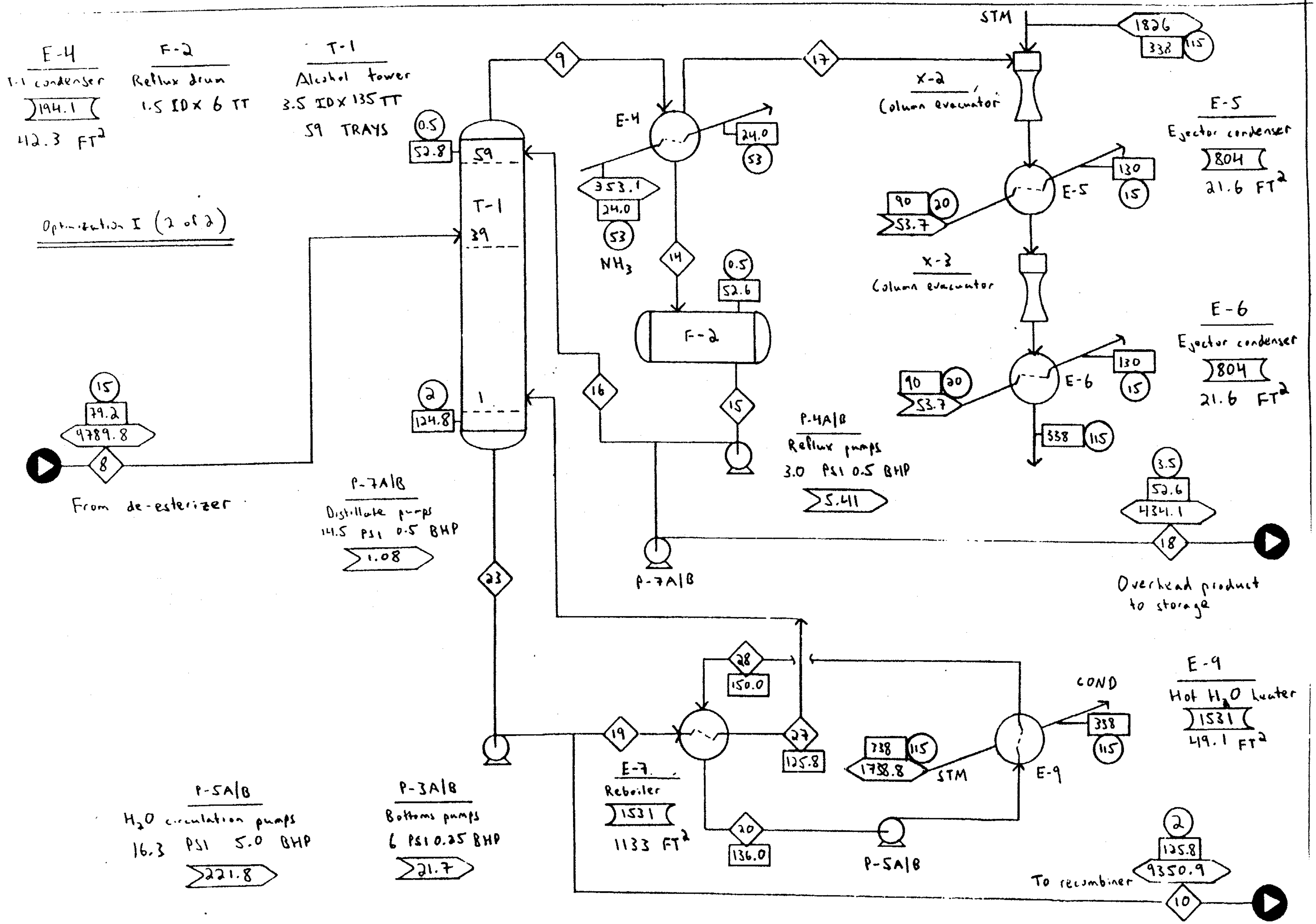


Optimization I (1 of 2)
E-2
Preheater
57.9
15.1 FT²
F-1
De-esterizer
4.5 10X 18 TT
X-1
Vessel Evacuator
E-10
Hot H₂O heater
57.9
10 FT²
P-8A|B
H₂O circulation pumps
11.3 PSI 0.5 BHP
11.7
E-8A
CW Flash Condenser
33.5
10 FT²
E-8B
NH₃ Flash Condenser
9.14
10 FT²
P-2A|B
T-1 feed pumps
19.8
2.0 PSI 0.25 BHP
To distillation
F-2
Recombiner
2.0 10X 8.0 TT
E-3
Product cooler
471.9
105 FT²
NA beer to storage
E-1
Preheater/cooler
394.1
96.7 FT²
P-6A|B
Product pumps
18.8
21 PSI 0.5 BHP
P-1A|B
Feed pumps
19.9
13.3 PSI 0.5 BHP
From distillation
Cold Prairie Premium from storage
CO₂ feed
STM
CWS
CWR
NH₃
COND


F-4
condenser
194.1
112.3 FT^2
F-2
Reflux drum
1.5 ID x 6 TT
T-1
Alcohol tower
3.5 ID x 135 TT
59 TRAYS
Optimization I (2 of 2)
From de-esterizer
8
9789.8
79.2
15
T-1
59
39
1
0.5
52.8
2
124.8
9
E-4
353.1
24.0
53
NH3
24.0
53
14
F-2
0.5
52.6
15
16
P-4A|B
Reflux pumps
3.0 PSI 0.5 BHP
5.41
P-7A|B
Distillate pumps
14.5 PSI 0.5 BHP
1.08
P-7A|B
17
STM
1826
338
115
X-2
Column evacuator
E-5
Ejector condenser
804
21.6 FT^2
E-5
130
15
90
20
53.7
X-3
Column evacuator
E-6
Ejector condenser
804
21.6 FT^2
E-6
130
15
90
20
53.7
338
115
3.5
52.6
434.1
18
Overhead product to storage
23
19
28
150.0
27
125.8
20
136.0
E-7
Reboiler
1531
1133 FT^2
P-5A|B
338
115
1738.8
STM
E-9
COND
338
115
E-9
Hot H2O heater
1531
49.1 FT^2
2
125.8
9350.9
10
To recombiner
P-5A|B
H2O circulation pumps
16.3 PSI 5.0 BHP
221.8
P-3A|B
Bottoms pumps
6 PSI 0.25 BHP
21.7



Optimization I (1 of 2)
E-2
Preheater
57.9
15.1 FT²
F-1
De-esterizer
4.5 ID X 18 TT
X-1
Vessel Evacuator
E-8A
CW Flash Condenser
33.5
10 FT²
E-8B
NH₃ Flash Condenser
9.14
10 FT²
E-10
Hot H₂O heater
57.9
10 FT²
P-8A/B
H₂O circulation pumps
11.3 PSI 0.5 BHP
11.7
STM
28.9
338
115
X-1
CWS
1.83
90
20
E-8A
CWR
120
15
79.2
13
NH₃
16.6
24.0
53
E-8B
24.0
53
To distillation
79.2
9789.8
15
P-2A/B
T-1 feed pumps
19.8
2.0 PSI 0.25 BHP
From distillation
125.8
9350.9
55.1
68.8
16
F-2
Recombiner
2.0 ID X 8.0 TT
E-3
Product cooler
471.9
105 FT²
NA beer to storage
18
34.0
9953.3
80.0
STM
65.7
338
115
130
120
P-8A/B
E-10
338
115
COND
E-2
E-1
Preheater/cooler
394.1
96.7 FT²
125.6
7
Cold Prairie Premium from storage
28
34.0
10359.7
74.0
E-1
28
P-6A/B
Product pumps
18.8
21 PSI 0.5 BHP
NH₃
53
24.0
858.3
34.0
E-3
24.0
53
CO₂ feed
P-1A/B
Feed pumps
13.3 PSI 0.5 BHP
19.9
28



E-4
1-1 condenser
194.1
412.3 FT²
F-2
Reflux drum
1.5 ID x 6 TT
T-1
Alcohol tower
3.5 ID x 135 TT
59 TRAYS
Optimization I (2 of 2)
STM
1826
338
115
9
17
X-2
Column evacuator
E-5
Ejector condenser
804
21.6 FT²
E-4
24.0
53
353.1
24.0
53
NH3
0.5
52.8
59
T-1
39
1
2
124.8
130
15
90
20
53.7
E-5
X-3
Column evacuator
14
0.5
52.6
F-2
E-6
Ejector condenser
804
21.6 FT²
90
20
53.7
E-6
130
15
338
115
16
15
15
79.2
9789.8
8
From de-esterizer
P-4A/B
Reflux pumps
3.0 PSI 0.5 BHP
5.41
P-7A/B
Distillate pumps
14.5 PSI 0.5 BHP
1.08
3.5
52.6
434.1
18
Overhead product to storage
P-7A/B
23
28
150.0
E-9
Hot H2O heater
1531
49.1 FT²
COND
338
115
19
27
125.8
738
115
1738.8
STM
E-9
E-7
Reboiler
1531
1133 FT²
20
136.0
P-5A/B
P-5A/B
H2O circulation pumps
16.3 PSI 5.0 BHP
221.8
P-3A/B
Bottoms pumps
6 PSI 0.25 BHP
21.7
2
125.8
9350.9
To recombiner
10

Stream Attributes

Optimization I (1 of 3)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Title	Feed to Process	Feed to Separation	Partially preheated Feed	Preheated Feed	Flash Liquid	Flash Vapor	E-8 liquid effluent	T-1 Feed	T-1 Overhead	T-1 bottoms to recombiner	Recombiner Outlet	Partially cooled recombiner
Water	9805.5	9335.1	9335.1	9335.1	9334.3	0.83	29.2	9334.3	126.9	9308.9	9338.1	9338.1
Ethanol	435.3	414.4	414.4	414.4	410.2	4.19	3.63	410.2	2024.0	5.41	9.01	9.01
Carbon Dioxide	54.8	52.2	52.2	52.2	4.76	47.4	1.76	4.76	4.76	0.0	1.76	1.76
Pyruvic Acid	21.7	20.7	20.7	20.7	20.7	–	–	20.7	0.0	20.7	20.7	20.7
Ethyl Acetate	21.7	20.7	20.7	20.7	19.8	0.87	0.81	19.8	19.75	15.9	16.667	16.667
Air	–	–	–	–	–	–	0.0001	–	0.0	0.0	0.0001	0.0001
Volatiles	20.7	19.7	19.7	19.7	–	19.7	19.7	–	0.0	0.0	19.7	19.7
Total	10359.7	9862.8	9862.8	9862.8	9789.8	73.0	55.1	9789.8	2175.4	9350.9	9405.9	9405.9
Volumetric	20.9	19.9	20.0	20.0	19.9	12.2	0.12	19.8	9472.9	18.91	19.03	18.9
Temperature	34.0	34.0	74.0	80.0	79.2	79.2	68.8	79.2	52.8	125.8	125.6	84.3
Pressure	28.0	28.0	23.0	18.0	13.2	13.2	15.7	15.2	0.50	2.0	7.0	23.0
Phase	L	L	L	L	L	V	L	L	V	L	L	L
Molecular Weight	18.61	18.61	18.61	18.61	18.55	43.54	20.24	18.55	42.41	18.078	18.085	18.085
Density	61.785	61.785	61.455	61.362	61.507	0.106	59.641	61.507	0.004	61.607	61.589	62.146
Enthalpy	-68434	-65151	-63523	-65152	-64954	-270.7	-338.7	-64820	-5289	-63140	-63493	-63884

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes Optimization I (2 of 3)

Stream Number	13	14	15	16	17	18	19	20	21	22	23	24
Stream Title	Preliminary Product	T-1 Condensate	Reflux Drum effluent	Pumped Reflux	T-1 Air + CO_2	Overhead product	T-1 bottoms to reboiler	Reboiler H_2O	Pumped recombined outlet	X-1 Discharge	T-1 total bottoms	E-8 vapor effluent
Water	9338.1	126.9	126.9	101.5	—	25.37	1484.9	109370	9338.1	29.7	10793.8	0.52
Ethanol	9.01	2024.0	2024.0	1619.2	—	404.8	10.39	—	9.01	4.19	15.8	0.56
Carbon Dioxide	1.76	0.0	0.0	0.0	195.3	0.0	0.0	—	1.76	47.4	0.0	45.64
Pyruvic Acid	20.7	0.0	0.0	0.0	—	0.0	0.2	—	20.7	—	20.9	—
Ethyl Acetate	16.667	19.75	19.75	15.8	—	3.95	7.1	—	16.667	0.86	23.0	0.059
Air	0.0001	0.0	0.0	0.0	16.7	0.0	0.0	—	0.0001	8.31	0.0	8.31
Volatiles	19.7	0.0	0.0	0.0	—	0.0	0.0	—	19.7	19.7	0.0	—
Total	9405.9	2170.6	2170.6	1736.5	212.0	434.12	1502.6	109370	9405.9	110.2	10853.5	55.1
Volumetric	18.8	5.36	5.36	4.29	919.0	1.07	3.04	221.8	19.0	27.4	21.95	8.2
Temperature	34.0	52.6	52.6	52.6	52.8	52.6	124.8	136.0	125.6	184.2	124.8	68.8
Pressure	18.0	0.5	0.5	3.5	0.5	3.5	8.0	14.7	28.0	15.7	2.0	15.7
Phase	L	L	L	L	V	L	L	L	L	V	L	V
Molecular Weight	18.085	42.39	42.39	42.39	42.28	42.39	18.09	18.016	18.085	29.18	18.09	40.33
Density	62.422	50.487	50.487	50.487	0.004	50.487	61.60	61.434	61.589	0.067	61.60	0.112
Enthalpy	-64355	-6170	-6170	-4936	-730.5	-1234	-10141	-739372	-63493	-441.2	-73250	-179.7

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes

Optimization I (3 of 3)

Stream Number	25	26	27	28	29	30	31	32				
Stream Title	Preheater H_2O	Preheater H_2O	Vapor to T-1	Reboiler H_2O	CO_2 Feed	Bypass reg. beer	NAB w/o CO_2	Final NAB Product				
Water	5794.0	5794.0	1484.9	109370	—	470.4	9808.5	9808.5				
Ethanol	—	—	10.39	—	—	20.9	29.9	29.9				
Carbon Dioxide	—	—	0.0	—	50.44	2.63	4.39	54.8				
Pyruvic Acid	—	—	0.2	—	—	1.04	21.7	21.7				
Ethyl Acetate	—	—	7.1	—	—	1.04	17.7	17.7				
Air	—	—	0.0	—	—	—	0.0001	0.0001				
Volatiles	—	—	0.0	—	—	0.99	20.7	20.7				
Total	5794.0	5794.0	1502.6	109370	50.44	497.0	9902.9	9953.3				
Volumetric	11.7	11.7	4324.3	222.8	1.05	1.00	19.8	19.9				
Temperature	130.0	120.0	125.8	150.0	70.0	34.0	34.0	34.0				
Pressure	26.0	16.0	2.0	26.0	100	28.0	18.0	18.0				
Phase	L	L	V	L	V	L	L	L				
Molecular Weight	18.016	18.016	18.09	18.016	44.01	18.61	18.11	18.16				
Density	61.697	61.536	0.006	61.177	0.804	61.785	62.389	62.361				
Enthalpy	-39204	-39262	-8573	-757845	-194.3	-3283.1	-67661	-67861				

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Stream Attributes Optimization I (1 of 3)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Title	Feed to Process	Feed to Separation	Partially preheated Feed	Preheated Feed	Flash Liquid	Flash Vapor	E-8 liquid effluent	T-1 Feed	T-1 Overhead	T-1 bottoms to recombiner	Recombiner Outlet	Partially cooled recombiner
Water	9805.5	9335.1	9335.1	9335.1	9334.3	0.83	24.2	9334.3	126.9	9308.9	9338.1	9338.1
Ethanol	435.3	414.4	414.4	414.4	410.2	4.19	3.63	410.2	2024.0	5.41	9.01	9.01
Carbon Dioxide	54.8	52.2	52.2	52.2	4.76	47.4	1.76	4.76	23.8	0.0	1.76	1.76
Pyruvic Acid	21.7	20.7	20.7	20.7	20.7	–	–	20.7	0.0	20.7	20.7	20.7
Ethyl Acetate	21.7	20.7	20.7	20.7	19.8	0.87	0.81	19.8	19.75	15.9	16.667	16.667
Air	–	–	–	–	–	–	0.0001	–	0.0	0.0	0.0001	0.0001
Volatiles	20.7	19.7	19.7	19.7	–	19.7	19.7	–	0.0	0.0	19.7	19.7
Total	10359.7	9862.8	9862.8	9862.8	9789.8	73.0	55.1	9789.8	2170.6	9350.9	9405.9	9405.9
Volumetric	20.9	19.9	20.0	20.0	19.9	12.2	0.12	19.8	9472.9	18.91	19.03	18.9
Temperature	34.0	34.0	74.0	80.0	79.2	79.2	68.8	79.2	52.8	125.8	125.6	84.3
Pressure	28.0	28.0	23.0	18.0	13.2	13.2	15.7	15.2	0.50	2.0	7.0	23.0
Phase	L	L	L	L	L	V	L	L	V	L	L	L
Molecular Weight	18.61	18.61	18.61	18.61	18.55	43.54	20.24	18.55	42.41	18.078	18.085	18.085
Density	61.785	61.785	61.455	61.362	61.507	0.100	59.641	61.507	0.004	61.607	61.589	62.146
Enthalpy	-68434	-65151	-63523	-65152	-64954	-270.7	-338.7	-64820	-5289	-63140	-63493	-63884

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes

Optimization I (2 of 3)

Stream Number	13	14	15	16	17	18	19	20	21	22	23	24
Stream Title	Preliminary Product	T-1 Condensate	Reflux Drum effluent	Pumped Reflux	T-1 Air + CO_2	Overhead product	T-1 bottoms to reboiler	Reboiler H_2O	Pumped fermentor outlet	X-1 Discharge	T-1 total bottoms	E-8 vapor effluent
Water	9338.1	126.9	126.9	101.5	—	25.37	1484.9	109370	9338.1	29.7	10793.8	0.52
Ethanol	9.01	2024.0	2024.0	1619.2	—	404.8	10.39	—	9.01	4.19	15.8	0.56
Carbon Dioxide	1.76	0.0	0.0	0.0	195.3	0.0	0.0	—	1.76	47.4	0.0	45.64
Pyruvic Acid	20.7	0.0	0.0	0.0	—	0.0	0.2	—	20.7	—	20.9	—
Ethyl Acetate	16.667	19.75	19.75	15.8	—	3.95	7.1	—	16.667	0.86	23.0	0.059
Air	0.0001	0.0	0.0	0.0	16.7	0.0	0.0	—	0.0001	8.31	0.0	8.31
Volatiles	19.7	0.0	0.0	0.0	—	0.0	0.0	—	19.7	19.7	0.0	—
Total	9405.9	2170.6	2170.6	1736.5	212.0	434.12	1502.6	109370	9405.9	110.2	10853.5	55.1
Volumetric	18.8	5.36	5.36	4.29	919.0	1.07	3.04	221.8	19.0	27.4	21.95	8.2
Temperature	34.0	52.6	52.6	52.6	52.8	52.6	124.8	136.0	125.6	184.2	124.8	68.8
Pressure	18.0	0.5	0.5	3.5	0.5	3.5	8.0	14.7	28.0	15.7	2.0	15.7
Phase	L	L	L	L	V	L	L	L	L	V	L	V
Molecular Weight	18.085	42.39	42.39	42.39	42.28	42.39	18.09	18.016	18.085	29.18	18.09	40.33
Density	62.422	50.487	50.487	50.487	0.004	50.487	61.60	61.434	61.589	0.067	61.60	0.112
Enthalpy	-64355	-6170	-6170	-4936	-730.5	-1234	-10141	-739372	-63493	-441.2	-73250	-179.7

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes Optimization I (3 of 3)

Stream Number	25	26	27	28	29	30	31	32				
Stream Title	Preheater H_2O	Preheater H_2O	Vapor to T-1	Reboiler H_2O	CO_2 feed	Bypass reg. beer	NAB w/o CO_2	Final NAB Product				
Water	5794.0	5794.0	1484.9	109370	—	470.4	9808.5	9808.5				
Ethanol	—	—	10.39	—	—	20.9	29.9	29.9				
Carbon Dioxide	—	—	0.0	—	50.44	2.63	4.39	54.8				
Pyruvic Acid	—	—	0.2	—	—	1.04	21.7	21.7				
Ethyl Acetate	—	—	7.1	—	—	1.04	17.7	17.7				
Air	—	—	0.0	—	—	—	0.0001	0.0001				
Volatiles	—	—	0.0	—	—	0.99	20.7	20.7				
Total	5794.0	5794.0	1502.6	109370	50.44	497.0	9902.9	9953.3				
Volumetric	11.7	11.7	4324.3	222.8	1.05	1.00	19.8	19.9				
Temperature	130.0	120.0	125.8	150.0	70.0	34.0	34.0	34.0				
Pressure	26.0	16.0	2.0	26.0	100	28.0	18.0	18.0				
Phase	L	L	V	L	V	L	L	L				
Molecular Weight	18.016	18.016	18.09	18.016	44.01	18.61	18.11	18.16				
Density	61.697	61.536	0.006	61.177	0.804	61.785	62.389	62.361				
Enthalpy	-39204	-39262	-8573	-737845	-194.3	-3283.1	-67661	-67861				

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Table 13.1 Utility Summary for Optimization I

A. Electricity at $0.0494/kWhr

User	Yearly Cost
P-1A/B	$154.80
P-2A/B	$77.40
P-3A/B	$77.40
P-4A/B	$77.40
P-5A/B	$1547.80
P-6A/B	$154.80
P-7A/B	$154.80
P-8A/B	$154.80
Total	**$2399.20**

B. 100 psig Steam at $2.815/1000 lb

User	Amount (lb/h)	Yearly Cost
X-1	28.9	$683.32
X-2 and X-3	1826	$43,174.77
E-9	1738.8	$41,112.98
E-10	65.7	$1553.44
Total	**3659.4**	**$86,524.51**

Table 13.1 (cont'd)

C. Cooling Water at $0.117/1000 gal

User	Amount (gpm)	Yearly Cost
E-5	53.7	$3174.27
E-6	53.7	$3174.27
E-8A	1.83	$108.17
Total	**109.23**	**$6456.71**

D. Ammonia Refrigeration at $0.0081/10³ Btu

User	Amount (10^3 Btu/hr)	Yearly Cost
E-3	471.9	$32,058.53
E-4	194.1	$13,186.18
E-8B	9.14	$620.93
Total	**778.14**	**$45,865.64**

Total Utilities Cost = $141,246/yr

Table 15.1 Equipment Information Summary for Optimization I

A. Pumps

Number	Description	Body Type	Suct. P (psia)	Disch. P (psia)	GPM	Head (ft)	Motor BHP
P-1A/B	Feed	1.5 x 1.5	14.7	28.0	19.9	31.0	0.5
P-2A/B	T-1 Feed	1.25 x 1.25	13.2	15.2	19.8	4.67	0.25
P-3/AB	Bottoms	1.5 x 1.5	2.0	8.0	21.7	13.9	0.25
P-4A/B	Reflux	1.25 x 1.25	0.5	3.5	5.41	8.6	0.25
P-5A/B	Hot water	3 x 3	14.7	31.0	221.8	38.1	5.0
P-6A/B	Product	1.5 x 1.5	7.0	28.0	18.8	48.5	0.5
P-7A/B	Distillate	1.5 x 1.5	2.5	17.0	1.1	41.8	0.5
P-8A/B	Hot water	1.5 x 1.5	14.7	26.0	11.7	26.4	0.5

B. Drums

Number	Description	Orientation	ID (ft)	L (ft)	Capacity (gal)	P (psia)	T (°F)
F-1	De-esterizer	Vertical	4.5	18.0	2141	13.2	79.2
F-2	Reflux	Horizontal	1.5	6.0	79.3	0.5	52.6
F-3	Recombiner	Vertical	2.0	8.0	188.0	7.0	125.6

C. Towers

Number	Description	Trays	Type	Feed to tray	ID (ft)	H (ft)
T-1	Alcohol removal	59	Sieve	39	3.5	135

D. Ejectors

Number	Description	Suction P (psia)	Discharge P (psia)	Stages	Steam/stage (lb/h)
X-1	F-1 Evacuator	13.2	15.7	1	28.9
X-2 and X-3	T-1 Evacuators	0.5	15.7	2	913

Table 15.1 (cont'd)

E. Heat Exchangers

Number	Description	Shell Side	Tube Side	Type	Duty (10^3 Btu/hr)	Area (ft^2)
E-1	Preheater	Feed	Product	F.T.S.*	394.1	96.7
E-2	Preheater	Hot water	Feed	F.T.S.	57.9	15.1
E-3	Cooler	Ammonia	Product	F.T.S.	471.9	105
E-4	Condenser	Overhead	Ammonia	F.T.S.	194.1	42.3
E-5	X-2 Condenser	Steam	CW	F.T.S.	804.0	21.6
E-6	X-2 Condenser	Steam	CW	F.T.S.	804.0	21.6
E-7	Reboiler	Hot water	Bottoms	F.T.S.	1531	1133
E-8A	CW Flash cond.	CW	Fsh. vapor	F.T.S.	33.5	10.0
E-8B	NH_3 Flash cond.	Ammonia	Fsh. vapor	F.T.S.	9.1	10.0
E-9	Hot water heater	Steam	H_2O circ.	F.T.S.	1531	49.1
E-10	Hot water heater	Steam	H_2O circ.	F.T.S.	57.9	10.0

*F.T.S. = Fixed-Tube-Sheet

F. Storage Tanks

Number	Description	Type	Capacity (gal)	Diameter (ft)	Height (ft)
V-1	NA beer storage	Cone-roof	216,568	48.0	16.0
V-2	Distillate storage	Cone-roof	13069	18.5	6.5

Material of Construction: All of the above equipment is constructed of 304 stainless steel except P-5A/B, P-8A/B, X-2, X-3, E-5, E-6, E-9, and E-10. These pieces of equipment only see recirculating hot water and never process fluid.

Optimization I Economic Calculations

All cost figures in dollars unless otherwise stated

Section I: Capital Investment

Operating Pressure (psia)	Correction
0.08	1.3
0.2	1.2
0.7	1.1
8 to 100	1.0
700	1.1
3000	1.2
6000	1.3

Operating Temperature (C)	Correction
-80	1.3
0	1.0
100	1.05
600	1.1
5000	1.2
10000	1.4

Note: Correcting factors may have already been taken into account in earlier estimates. For example, inflation factors are shown as 1.0, since inflation of 1.1646 (see Economic Calculations, p. 9) is already included

Material	Correction
Carbon steel	1.0
Bronze	1.05
Carbon/molybdenum	1.065
Aluminum	1.075
Cast steel	1.11
Stainless steel	1.28 to 1.5
Worthite alloy	1.41
Hastelloy C alloy	1.54
Monel alloy	1.65
Nickel/inconel alloy	1.71
Titanium	2.0

:90

Equipment	Number	Purchase Cost	Inflation Factor	Delivered Cost	Material	Pressure	Temperature	Total Del. Cost	Installation	Controls	Piping	Electrical
Pump	P1-A/B	3606	1	3786.30	1	1	1	3786.30	1779.56	681.53	2498.96	416.49
Pump	P2-A/B	2893	1	3037.65	1	1	1	3037.65	1427.70	546.78	2004.85	334.14
Pump	P3-A/B	3786	1	3975.30	1	1	1	3975.30	1868.39	715.55	2623.70	437.28
Pump	P4-A/B	2893	1	3037.65	1	1.1	1	3341.42	1570.47	601.45	2205.33	367.56
Pump	P5-A/B	3412	1	3582.60	1	1	1	3582.60	1683.82	644.87	2364.52	394.09
Pump	P6-A/B	5371	1	5639.55	1	1.1	1	6203.51	2915.65	1116.63	4094.31	682.39
Pump	P7-A/B	5371	1	5639.55	1	1.1	1	6203.51	2915.65	1116.63	4094.31	682.39
Pump	P8-A/B	3786	1	3975.30	1	1	1	3975.30	1868.39	715.55	2623.70	437.28
Drivers	1 to 8-A/B	3657	1	3839.85	1	1	1	3839.85	1804.73	691.17	2534.30	422.38
Drum	F-1	13975	1	14673.75	1	1	1	14673.75	6896.66	2641.28	9684.68	1614.11
Drum	F-2	8022	1	8423.10	1	1.2	1	10107.72	4750.63	1819.39	6671.10	1111.85
Drum	F-3	8152	1	8559.60	1	1.1	1	9415.56	4425.31	1694.80	6214.27	1035.71
*Tower	T-1	298720	1	298720.00	1	1.2	1	358464.00	0.00	0.00	0.00	0.00
*Purchase cost correlation includes installation and auxiliaries												
Ejector	X-1	6516	1	6841.80	1	1	1	6841.80	3215.65	1231.52	4515.59	752.60
Ejector system	X-2, X-3, E-5, E-6	142821	1	149962.05	1	1	1	149962.05	70482.16	26993.17	98974.95	16495.83
Exchanger	E-1	4658	1	4890.90	1	1	1.05	5135.45	2413.66	924.38	3389.39	564.90
Exchanger	E-2	2562	1	2690.10	1	1	1.05	2824.61	1327.56	508.43	1864.24	310.71
Exchanger	E-3	4658	1	4890.90	1	1	1	4890.90	2298.72	880.36	3227.99	538.00
Exchanger	E-4	3261	1	3424.05	1	1.2	1.05	4314.30	2027.72	776.57	2847.44	474.57
Exchanger	E-5	*Included in X-2 and X-3 ejector system*										
Exchanger	E-6	*Included in X-2 and X-3 ejector system*										
Exchanger	E-7	12811	1	13451.55	1	1.1	1.05	15536.54	7302.17	2796.58	10254.12	1709.02
Exchanger	E-8A	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Exchanger	E-8B	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Exchanger	E-9	3494	1	3668.70	1	1	1.05	3852.14	1810.50	693.38	2542.41	423.73
Exchanger	E-10	2446	1	2568.30	1	1	1.05	2696.72	1267.46	485.41	1779.83	296.64
Storage tank	V-1	81522	1	85598.10	1.39	1	1	118981.36	55921.24	21416.64	78527.70	13087.95
Storage tank	V-2	11641	1	12223.05	1.39	1	1	16990.04	7985.32	3058.21	11213.43	1868.90
							Totals	768025.78	192494.04	73721.12	270310.77	45051.80

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Building	Yard	Service	Total Direct Cost	Engineering & Supervision	Construction	Total Indirect
378 63	378.63	2650.41	12570.52	1249.48	1552.38	2801.86
303 77	303.77	2126.36	10085.00	1002.42	1245.44	2247.86
397 53	397.53	2782.71	13198.00	1311.85	1629.87	2941.72
334 14	334 14	2338.99	11093.50	1102.67	1369.98	2472.05
358 26	358.26	2507.82	11894.23	1182.26	1468.87	2651.12
620 35	620 35	4342.45	20595.64	2047.16	2543.44	4590.59
620 35	620.35	4342.45	20595.64	2047.16	2543.44	4590.59
397 53	397.53	2782.71	13198.00	1311.85	1629.87	2941.72
383 99	383 99	2687.90	12748.30	1267.15	1574.34	2841.49
1467 38	1467.38	10271.63	48716.85	4842.34	6016.24	10858.58
1010 77	1010.77	7075.40	33557.63	3335.55	4144.17	7479.71
941 56	941.56	6590.89	31259.66	3107.13	3860.38	6967.51
35846 40	35846.40	250924.80	681081.60	118293.12	146970.24	265263.36
684 18	684.18	4789.26	22714.78	2257.79	2805.14	5062.93
14996 21	14996 21	104973.44	497874.01	49487.48	61484.44	110971.92
513 54	513 54	3594.81	17049.68	1694.70	2105.53	3800.23
282 46	282 46	1977.22	9377.69	932.12	1158.09	2090.21
489 09	489 09	3423.63	16237.79	1614.00	2005.27	3619.27
431 43	431.43	3020.01	14323.49	1423.72	1768.86	3192.58
1553 65	1553.65	10875.58	51581.31	5127.06	6369.98	11497.04
269 67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
269 67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
385 21	385.21	2696.49	12789.09	1271.20	1579.38	2850.58
269 67	269.67	1887.70	8953.09	889.92	1105.65	1995.57
11898 14	11898.14	83286.95	395018.11	39263.85	48782.36	88046.21
1699 00	1699.00	11893.03	56406.93	5606.71	6965.92	12572.63
76802 58	76802.58	537618.04	2040826.70	253448.51	314890.57	568339.07

Direct + Indirect	Contractor	Contingency	Fixed Capital Inv.	Working Capital	Total Capital Inv.
15372.38	768.62	1537.24	17678.23	3119.69	20797.92
12332.86	616.64	1233.29	14182.79	2502.84	16685.63
16139.72	806.99	1613.97	18560.68	3275.41	21836.09
13566.14	678.31	1356.61	15601.07	2753.13	18354.20
14545.36	727.27	1454.54	16727.16	2951.85	19679.01
25186.23	1259.31	2518.62	28964.16	5111.32	34075.49
25186.23	1259.31	2518.62	28964.16	5111.32	34075.49
16139.72	806.99	1613.97	18560.68	3275.41	21836.09
15589.79	779.49	1558.98	17928.26	3163.81	21092.07
59575.43	2978.77	5957.54	68511.74	12090.31	80602.05
41037.34	2051.87	4103.73	47192.94	8328.17	55521.11
38227.17	1911.36	3822.72	43961.25	7757.87	51719.12
946344.96	47317.25	94634.50	1088296.70	192052.36	1280349.06
27777.71	1388.89	2777.77	31944.36	5637.24	37581.60
608845.92	30442.30	60884.59	700172.81	123559.91	823732.72
20849.91	1042.50	2084.99	23977.39	4231.30	28208.70
11467.90	573.39	1146.79	13188.08	2327.31	15515.39
19857.05	992.85	1985.71	22835.61	4029.81	26865.43
17516.07	875.80	1751.61	20143.48	3554.73	23698.21
63078.35	3153.92	6307.84	72540.11	12801.20	85341.30
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
15639.67	781.98	1563.97	17985.62	3173.93	21159.55
10948.66	547.43	1094.87	12590.96	2221.93	14812.90
483064.32	24153.22	48306.43	555523.97	98033.64	653557.61
68979.56	3448.98	6897.96	79326.49	13998.79	93325.29
2609165.77	130458.29	260916.58	3000540.64	529507.17	3530047.81

Grand Total Capital Investment:
$3,530,048

Optimization I Economic Calculations

All cost figures in dollars per year unless otherwise stated

Section II: Operating Costs

Operator Positions	Inflation Factor	Operating Labor	Direct Supervision & Clerical Labor	Maintenance & Repairs	Operating Supplies	Laboratory Charges	Local taxes & Insurance	Plant Overhead	Administrative Expenses
1	1.1646	821741.76	143804.81	90016.22	13502.43	123261.26	90016.22	633337.67	158334.42

Steam use (lb/h)	Steam cost
65.7	1553.44
28.9	683.32
1826	43174.77
1738.8	41112.98
Total	86524.52

Process Water (gpm)	Process Water Cost
0.00	0.00

Cooling Water (gpm)	Cooling Water Cost
53.7	3174.27
53.7	3174.27
1.83	108.17
Total	6456.71

CO_2 (lb/h)	CO_2 Cost
50.44	17640.58

	Electricity Costs
	154.80
	77.40
	77.40
	77.40
	1547.80
	154.80
	154.80
	154.80
Total	2399.20

Refrigeration (1000 Btu/hr)	Cost at $2.00/288KBtu	Inflation factor	1999 Cost
471.9	27527.50	1.1646	32058.53
9.14	533.17	1.1646	620.93
194.1	11322.50	1.1646	13186.18
		Total	45865.64

Total Operating Cost
2.2329E+06 $/yr

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Optimization I Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	2040826.70	568339.07	391374.87	529507.17				
2000	1					2232901.44	350000	237800	5.00
2001	2						250000	237800	5.00
2002	3						150000	237800	5.00
2003	4						90000	237800	5.00
2004	5							237800	5.00
2005	6							237800	5.00
2006	7							237800	5.00
2007	8							237800	5.00
2008	9							237800	5.00
2009	10							237800	5.00

MISTAKE

Profit on NA beer sales	Opportunity Cost ($5.00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-3530047.81
1189000	1189000	4474444.8	0	0	0.1429	291634.14	3003072.56	1201229.02	690314.34
1189000	1189000	4474444.8	0	0	0.2449	499798.46	3974646.34	1589858.54	2634586.26
1189000	1189000	4474444.8	0	0	0.1749	356940.59	4117504.21	1647001.68	2677443.12
1189000	1189000	4474444.8	0	0	0.1249	254899.25	4219545.55	1687818.22	2696626.58
1189000	1189000	4474444.8	0	0	0.0893	182245.82	4292198.98	1716879.59	2757565.21
1189000	1189000	4474444.8	0	0	0.0892	182041.74	4292403.06	1716961.22	2757483.58
1189000	1189000	4474444.8	0	0	0.0893	182245.82	4292198.98	1716879.59	2757565.21
1189000	1189000	4474444.8	0	0	0.0446	91020.87	4383423.93	1753369.57	2721075.23
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	2684666.88
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	2684666.88
								NPW @ 12%	9.9467E+06
								IRR	0.550
								NPW @ IRR	-3.4E-10

Net Present Worth at 12 %
$9,946,718

Internal Rate of Return =
0.550

Mistake operating costs left out

Mistake!!!

Today, eight days before the deadline, I discovered a major error in all of my economic analyses to date (base case, base case without sale of ethanol, and first optimization). In a moment of gross mental ineptitude, I left out *operating costs* in my economic calculations for project years two through ten. Clearly, this oversight has a significant effect on the overall project economics. I should have realized that there was an error when the economics seemed so favorable. I have attempted to go back through my calculation book and label the pages which contain these errors. Also, the corrected page(s) follow this page. However, it is quite likely that I have not found every place where a correction is needed. I also do not have the time to rewrite large sections of my notes on the calculations. Thus, the reader is advised to keep these facts in mind while reading through my work. Luckily, I discovered this mistake today and not a week from now.

Obviously, the base case is not even close to economical. The first optimization is better, but is still unacceptable. Thus, there is still great need for optimization. I am beginning the capacity optimization now, and the economics will hopefully improve.

Optimization I Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency	Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl
1999	0	2040826.70	568339.07	391374.87	529507.17				
2000	1					2232901.44	350000	237800	5.00
2001	2					2232901.44	250000	237800	5.00
2002	3					2232901.44	150000	237800	5.00
2003	4					2232901.44	90000	237800	5.00
2004	5					2232901.44		237800	5.00
2005	6					2232901.44		237800	5.00
2006	7					2232901.44		237800	5.00
2007	8					2232901.44		237800	5.00
2008	9					2232901.44		237800	5.00
2009	10					2232901.44		237800	5.00

Profit on NA beer sales	Opportunity Cost ($5 00/bbl)	Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
									-3530047.81
1189000	1189000	4474444.8	0	0	0.1429	291634.14	3003072.56	1201229.02	690314.34
1189000	1189000	4474444.8	0	0	0.2449	499798.46	3974646.34	1589858.54	401684.83
1189000	1189000	4474444.8	0	0	0.1749	356940.59	4117504.21	1647001.68	444541.68
1189000	1189000	4474444.8	0	0	0.1249	254899.25	4219545.55	1687818.22	463725.14
1189000	1189000	4474444.8	0	0	0.0893	182245.82	4292198.98	1716879.59	524663.77
1189000	1189000	4474444.8	0	0	0.0892	182041.74	4292403.06	1716961.22	524582.14
1189000	1189000	4474444.8	0	0	0.0893	182245.82	4292198.98	1716879.59	524663.77
1189000	1189000	4474444.8	0	0	0.0446	91020.87	4383423.93	1753369.57	488173.79
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	451765.44
1189000	1189000	4474444.8	0	0			4474444.80	1789777.92	451765.44
								NPW @ 12%	-6.7601E+05
								IRR	0.070
								NPW @ IRR	0.0E+00

Net Present Worth at 12 %
- ($676,011)

Internal Rate of Return =
0.070

Results of Optimization I Economic Calculations → Next Optimization

The economic analysis (following the same procedure as before) is quite good for this first optimization - perhaps too good. I was correct in my assumption about the dominance of the distillation column in both the capital investment and the operating costs. When I did the base case, I just wanted a column that would work. Then, when I optimized, I learned that I could accomplish the same separation with a much cheaper column. My new column has a much lower reflux ratio (it's OK to talk about reflux ratio since the material balance is fixed) and thus a much lower vapor rate is required. The lower vapor rate translates into lower capital costs by dramatically reducing the required column diameter. It translates into lower operating costs by reducing the refrigeration required to condense the overhead and the energy required to reboil the bottoms.

The salient details of the new economic analysis are:

Total Capital Investment	\$3,530,048
Total Operating Cost	\$2,232,900 / yr
Net Present Worth @ 12%	~~\$9,946,718~~ -\$676,011
Internal Rate of Return	0.~~580~~ N/A

Obviously, this IRR is outside of the range of allowable IRR (>20%), meaning that this plant would be too trivial to build - or that I have made a gross calculational error. Hoping that it's the former, I'll scale up my economic calculations using the 0.6 rule for the capital investment. The operating costs are scaled on a linear basis, with the exception of Maintenance + Repairs and Local Taxes + Insurance, which are a certain percentage of Fixed Capital Investment. However, looking back on my earlier economic example, I see that Fixed Capital Investment is related to Total Capital Investment in a linear manner (see p. 3 of the economic example). The relationship is as follows:

Fixed Capital	Working Capital	Total Capital
X	$0.15\left(\dfrac{X}{0.85}\right)$	$X + \dfrac{0.15}{0.85}X$

$$= (1 + 0.176)X$$

$$= 1.176X$$

Next Optimization [cont'd]

Therefore, once I get the Total Capital Investment, I'll just divide it by 1.176 to get the Fixed Capital Investment, which will allow me to calculate the rest of my operating costs. (The other operating costs are related to the base case in a linear manner).

In case I haven't explicitly stated the 0.6 rule for relating capacity to Total Capital Investment, I will here:

$$\text{Cost of equip. a} = \text{cost of equip. b} \left(\frac{\text{capac. equip. a}}{\text{capac. equip. b}} \right)^{0.6}$$

→ page 169 of Peters and Timmerhaus

Unfortunately, for my specific types and sizes of equipment (stainless steel, small sizes), I can't use any of the specific exponents other than 0.6 given on p. 170 of Peters and Timmerhaus. I'll have to go with 0.6.

Also, Peters and Timmerhaus state that this rule should not be used beyond a tenfold range of capacity. I hope the optimum does not lie outside this range.

Also ...

I have just realized that for depreciation purposes, I need to have the Direct Costs, Indirect Costs, the Contractor's Fee, and the Contingency all broken out separately. Once I have the total capital investment, though, I can back calculate the others (see pp. 2-3 of economic example).

Total Delivered Cost	Total Direct Plant Cost	Engineering + Supervision	Construction Expenses
Y	3.32 Y	0.33 Y	0.41 Y

$\underbrace{0.33\,Y \quad 0.41\,Y}_{\text{Total Indirect Costs} = 0.74\,Y}$

Total Indirect Costs = 0.74 Y

Direct + Indirect	Contractor's Fee	Contingency
$(3.32 + 0.74)Y$	$0.05(4.06Y)$	$0.10(4.06Y)$
$= 4.06\,Y$	$= 0.203\,Y$	$= 0.406\,Y$

Next Optimization [cont'd]

Fixed Capital

$4.06Y + 0.203Y + 0.406Y$

$= 4.669Y$

$= X$ (from earlier)

Working Capital

$\frac{0.15}{0.85}(4.669Y)$

$= \frac{0.15}{0.85} X$

Total Capital Investment

$X + \frac{0.15}{0.85} X$

$= 1.176 X$

$=$ TCI calculated from 0.6 rule

Now I can calculate all the other costs in terms of the Total Capital Investment, TCI:

Total Capital Investment $:$ TCI

Fixed Capital $= TCI / 1.176$

Working Capital $= \frac{0.15}{0.85}\left(\frac{TCI}{1.176}\right)$

Total Delivered Cost $= \left(\frac{TCI}{1.176}\right) / 4.669 = Y$

Other costs as given in terms of Y on page 2:

Direct cost $= 3.32Y$
Indirect costs $= 0.74Y$
Fees and Contingency $= (0.203 + 0.406)Y$
$= 0.609Y$

Capacity Correction

When I substituted the new tower, the flow rate of the final product increased a small amount as I was able to add more regular beer. This amount will not make a large difference in the economics, but as long as I'm optimizing anyway, I might as well plug this hole.

From the material balance, the final product stream (32) is:

$$\dot{m}_{NAB} = 9953.3 \text{ lb/hr} \qquad \rho_L = 62.361 \text{ lb/ft}^3$$

Converting to bbl/yr:

$$\dot{m}_{NAB} = 9953.3 \frac{\text{lb}}{\text{hr}} \cdot \frac{1 \text{ ft}^3}{62.361 \text{ lb}} \cdot \frac{7.48 \text{ gal}}{\text{ft}^3} \cdot \frac{1 \text{ bbl}}{42 \text{ gal}} \cdot 8400 \frac{\text{hr}}{\text{yr}}$$

$$= \underline{\underline{238,733 \text{ bbl/yr}}}$$

→ Ready to begin capacity optimization.

I'll start by varying the capacity by even multiples (1, 2, 3 ... 10). Then, as needed, I'll go in between multiples.
Remember, I'm trying to maximize net present worth.

Mistake!!!

Today, eight days before the deadline, I discovered a major error in all of my economic analyses to date (base case, base case without sale of ethanol, and first optimization). In a moment of gross mental ineptitude, I left out *operating costs* in my economic calculations for project years two through ten. Clearly, this oversight has a significant effect on the overall project economics. I should have realized that there was an error when the economics seemed so favorable. I have attempted to go back through my calculation book and label the pages which contain these errors. Also, the corrected page(s) follow this page. However, it is quite likely that I have not found every place where a correction is needed. I also do not have the time to rewrite large sections of my notes on the calculations. Thus, the reader is advised to keep these facts in mind while reading through my work. Luckily, I discovered this mistake today and not a week from now.

Obviously, the base case is not even close to economical. The first optimization is better, but is still unacceptable. Thus, there is still great need for optimization I am beginning the capacity optimization now, and the economics will hopefully improve.

There is no error near this page, but I have included it once again to emphasize that the capacity optimization now takes on added importance.

Capacity Optimization

Results of Capacity Optimization

The next several pages contain the results of the economic calculations based on different multiples of base capacity.

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity = 238733 bbl/yr
Base TCI = $3,530,048
Base Operating = 2052869 $/yr
(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
2	$5,350,552	$4,549,789	4378725

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	3235232.32	721106.00	593450.75
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
802903.95						
	4378725.34	350000	477466	5.00	2387330	2387330
	4378725.34	250000	477466	5.00	2387330	2387330
	4378725.34	150000	477466	5.00	2387330	2387330
	4378725.34	90000	477466	5.00	2387330	2387330
	4378725.34		477466	5.00	2387330	2387330
	4378725.34		477466	5.00	2387330	2387330
	4378725.34		477466	5.00	2387330	2387330
	4378725.34		477466	5.00	2387330	2387330
	4378725.34		477466	5.00	2387330	2387330
	4378725.34		477466	5.00	2387330	2387330

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($5,352,693)
8984000.256	0	0	0.1429	462314.70	7341947.45	2936778.98	$1,318,496
8984000.256	0	0	0.2449	792308.40	8191691.86	3276676.74	$1,078,598
8984000.256	0	0	0.1749	565842.13	8418158.12	3367263.25	$1,088,012
8984000.256	0	0	0.1249	404080.52	8579919.74	3431967.90	$1,083,307
8984000.256	0	0	0.0893	288906.25	8695094.01	3478037.60	$1,127,237
8984000.256	0	0	0.0892	288582.72	8695417.53	3478167.01	$1,127,108
8984000.256	0	0	0.0893	288906.25	8695094.01	3478037.60	$1,127,237
8984000.256	0	0	0.0446	144291.36	8839708.89	3535883.56	$1,069,391
8984000.256	0	0			8984000.26	3593600.10	$1,011,675
8984000.256	0	0			8984000.26	3593600.10	$1,011,675
						NPW @ 12%	$990,293
						IRR	0.165
						NPW @ IRR	-2.2E-10

Net Present Worth at 12 %
$990,293

Internal Rate of Return =
0.165

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
3	$6,824,225	$5,802,912	6506782

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	4126294.57	919716.26	756901.62
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1024043.37						
	6506781.75	350000	716199	5.00	3580995	3580995
	6506781.75	250000	716199	5.00	3580995	3580995
	6506781.75	150000	716199	5.00	3580995	3580995
	6506781.75	90000	716199	5.00	3580995	3580995
	6506781.75		716199	5.00	3580995	3580995
	6506781.75		716199	5.00	3580995	3580995
	6506781.75		716199	5.00	3580995	3580995
	6506781.75		716199	5.00	3580995	3580995
	6506781.75		716199	5.00	3580995	3580995
	6506781.75		716199	5.00	3580995	3580995

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($6,826,956)
13476000.38	0	0	0.1429	589647.49	11706614.78	4682645.91	$1,936,573
13476000.38	0	0	0.2449	1010529.54	12465470.84	4986188.34	$1,733,030
13476000.38	0	0	0.1749	721688.92	12754311.46	5101724.59	$1,717,494
13476000.38	0	0	0.1249	515374.19	12960626.19	5184250.48	$1,694,968
13476000.38	0	0	0.0893	368478.11	13107522.28	5243008.91	$1,726,210
13476000.38	0	0	0.0892	368065.48	13107934.91	5243173.96	$1,726,045
13476000.38	0	0	0.0893	368478.11	13107522.28	5243008.91	$1,726,210
13476000.38	0	0	0.0446	184032.74	13291967.65	5316787.06	$1,652,432
13476000.38	0	0			13476000.38	5390400.15	$1,578,818
13476000.38	0	0			13476000.38	5390400.15	$1,578,818
						NPW @ 12%	$2,963,229
						IRR	0.222
						NPW @ IRR	-2.7E-10

Net Present Worth at 12 %
$2,963,229

Internal Rate of Return =
0.222

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity = 238733 bbl/yr
Base TCI = $3,530,048
Base Operating = 2052869 $/yr
(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
4	$8,109,920	$6,896,191	8625247

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	4903695.23	1092992.31	899503.13
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1216974.82						
	8625247.44	350000	954932	5.00	4774660	4774660
	8625247.44	250000	954932	5.00	4774660	4774660
	8625247.44	150000	954932	5.00	4774660	4774660
	8625247.44	90000	954932	5.00	4774660	4774660
	8625247.44		954932	5.00	4774660	4774660
	8625247.44		954932	5.00	4774660	4774660
	8625247.44		954932	5.00	4774660	4774660
	8625247.44		954932	5.00	4774660	4774660
	8625247.44		954932	5.00	4774660	4774660
	8625247.44		954932	5.00	4774660	4774660

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($8,113,165)
17968000.51	0	0	0.1429	700738.05	16087524.36	6435009.74	$2,557,743
17968000.51	0	0	0.2449	1200914.96	16767085.55	6706834.22	$2,385,919
17968000.51	0	0	0.1749	857656.30	17110344.22	6844137.69	$2,348,615
17968000.51	0	0	0.1249	612471.53	17355528.98	6942211.59	$2,310,541
17968000.51	0	0	0.0893	437899.98	17530100.53	7012040.21	$2,330,713
17968000.51	0	0	0.0892	437409.61	17530590.90	7012236.36	$2,330,517
17968000.51	0	0	0.0893	437899.98	17530100.53	7012040.21	$2,330,713
17968000.51	0	0	0.0446	218704.81	17749295.70	7099718.28	$2,243,035
17968000.51	0	0			17968000.51	7187200.20	$2,155,553
17968000.51	0	0			17968000.51	7187200.20	$2,155,553
						NPW @ 12%	$5,147,449
						IRR	0.265
						NPW @ IRR	5.6E-10

Net Present Worth at 12 %
$5,147,449

Internal Rate of Return =
0.265

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity = 238733 bbl/yr
Base TCI = $3,530,048
Base Operating = 2052869 $/yr
(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
5	$9,271,769	$7,884,157	10737394

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	5606211.50	1249577.26	1028368.31
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

3.5

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1391321.84						
	10737394.42	350000	1193665	5.00	5968325	5968325
	10737394.42	250000	1193665	5.00	5968325	5968325
	10737394.42	150000	1193665	5.00	5968325	5968325
	10737394.42	90000	1193665	5.00	5968325	5968325
	10737394.42		1193665	5.00	5968325	5968325
	10737394.42		1193665	5.00	5968325	5968325
	10737394.42		1193665	5.00	5968325	5968325
	10737394.42		1193665	5.00	5968325	5968325
	10737394.42		1193665	5.00	5968325	5968325
	10737394.42		1193665	5.00	5968325	5968325

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($9,275,479)
22460000.64	0	0	0.1429	801127.62	20479134.91	8191653.96	$3,180,952
22460000.64	0	0	0.2449	1372961.20	21087039.44	8434815.78	$3,037,790
22460000.64	0	0	0.1749	980526.39	21479474.25	8591789.70	$2,980,817
22460000.64	0	0	0.1249	700215.82	21759784.82	8703913.93	$2,928,692
22460000.64	0	0	0.0893	500634.69	21959365.95	8783746.38	$2,938,860
22460000.64	0	0	0.0892	500074.07	21959926.57	8783970.63	$2,938,636
22460000.64	0	0	0.0893	500634.69	21959365.95	8783746.38	$2,938,860
22460000.64	0	0	0.0446	250037.03	22209963.61	8883985.44	$2,838,621
22460000.64	0	0			22460000.64	8984000.26	$2,738,606
22460000.64	0	0			22460000.64	8984000.26	$2,738,606
						NPW @ 12%	$7,470,869
						IRR	0.301
						NPW @ IRR	6.8E-10

Net Present Worth at 12 %
$7,470,869

Internal Rate of Return =
0.301

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
6	$10,343,591	$8,795,571	12844948

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	6254293.04	1394029.17	1147248.33
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1552159.51						
	12844948.23	350000	1432398	5.00	7161990	7161990
	12844948.23	250000	1432398	5.00	7161990	7161990
	12844948.23	150000	1432398	5.00	7161990	7161990
	12844948.23	90000	1432398	5.00	7161990	7161990
	12844948.23		1432398	5.00	7161990	7161990
	12844948.23		1432398	5.00	7161990	7161990
	12844948.23		1432398	5.00	7161990	7161990
	12844948.23		1432398	5.00	7161990	7161990
	12844948.23		1432398	5.00	7161990	7161990
	12844948.23		1432398	5.00	7161990	7161990

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($10,347,730)
26952000.77	0	0	0.1429	893738.48	24878524.19	9951409.67	$3,805,643
26952000.77	0	0	0.2449	1531676.37	25420324.40	10168129.76	$3,688,923
26952000.77	0	0	0.1749	1093875.85	25858124.92	10343249.97	$3,613,803
26952000.77	0	0	0.1249	781161.20	26170839.57	10468335.83	$3,548,717
26952000.77	0	0	0.0893	558508.37	26393492.40	10557396.96	$3,549,656
26952000.77	0	0	0.0892	557882.94	26394117.83	10557647.13	$3,549,405
26952000.77	0	0	0.0893	558508.37	26393492.40	10557396.96	$3,549,656
26952000.77	0	0	0.0446	278941.47	26673059.30	10669223.72	$3,437,829
26952000.77	0	0			26952000.77	10780800.31	$3,326,252
26952000.77	0	0			26952000.77	10780800.31	$3,326,252

NPW @ 12%	$9,895,477
IRR	0.332
NPW @ IRR	-9.9E-11

Net Present Worth at 12 %
$9,895,477

Internal Rate of Return =
0.332

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
7	$11,345,912	$9,647,884	14948956

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	6860350.42	1529114.25	1258419.70
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1702567.83						
	14948956.06	350000	1671131	5.00	8355655	8355655
	14948956.06	250000	1671131	5.00	8355655	8355655
	14948956.06	150000	1671131	5.00	8355655	8355655
	14948956.06	90000	1671131	5.00	8355655	8355655
	14948956.06		1671131	5.00	8355655	8355655
	14948956.06		1671131	5.00	8355655	8355655
	14948956.06		1671131	5.00	8355655	8355655
	14948956.06		1671131	5.00	8355655	8355655
	14948956.06		1671131	5.00	8355655	8355655
	14948956.06		1671131	5.00	8355655	8355655

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($11,350,452)
31444000.9	0	0	0.1429	980344.07	29283918.72	11713567.49	$4,431,477
31444000.9	0	0	0.2449	1680099.82	29763901.08	11905560.43	$4,339,484
31444000.9	0	0	0.1749	1199875.29	30244125.61	12097650.24	$4,247,395
31444000.9	0	0	0.1249	856857.77	30587143.13	12234857.25	$4,170,188
31444000.9	0	0	0.0893	612629.29	30831371.60	12332548.64	$4,162,496
31444000.9	0	0	0.0892	611943.26	30832057.64	12332823.06	$4,162,222
31444000.9	0	0	0.0893	612629.29	30831371.60	12332548.64	$4,162,496
31444000.9	0	0	0.0446	305971.63	31138029.27	12455211.71	$4,039,833
31444000.9	0	0			31444000.90	12577600.36	$3,917,444
31444000.9	0	0			31444000.90	12577600.36	$3,917,444
						NPW @ 12%	$12,398,204
						IRR	0.359
						NPW @ IRR	3.0E-10

Net Present Worth at 12 %
$12,398,204

Internal Rate of Return =
0.359

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III. Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
8	$12,292,340	$10,452,670	17050112

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	7432612.10	1656666.55	1363391.80
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1844588.90						
	17050112.23	350000	1909864	5.00	9549320	9549320
	17050112.23	250000	1909864	5.00	9549320	9549320
	17050112.23	150000	1909864	5.00	9549320	9549320
	17050112.23	90000	1909864	5.00	9549320	9549320
	17050112.23		1909864	5.00	9549320	9549320
	17050112.23		1909864	5.00	9549320	9549320
	17050112.23		1909864	5.00	9549320	9549320
	17050112.23		1909864	5.00	9549320	9549320
	17050112.23		1909864	5.00	9549320	9549320
	17050112.23		1909864	5.00	9549320	9549320

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($12,297,259)
35936001.02	0	0	0.1429	1062120.27	33694142.65	13477657.06	$5,058,232
35936001.02	0	0	0.2449	1820246.70	34115754.32	13646301.73	$4,989,587
35936001.02	0	0	0.1749	1299963.86	34636037.17	13854414.87	$4,881,474
35936001.02	0	0	0.1249	928333.25	35007667.77	14003067.11	$4,792,822
35936001.02	0	0	0.0893	663732.26	35272268.76	14108907.51	$4,776,981
35936001.02	0	0	0.0892	662989.00	35273012.03	14109204.81	$4,776,684
35936001.02	0	0	0.0893	663732.26	35272268.76	14108907.51	$4,776,981
35936001.02	0	0	0.0446	331494.50	35604506.52	14241802.61	$4,644,086
35936001.02	0	0			35936001.02	14374400.41	$4,511,488
35936001.02	0	0			35936001.02	14374400.41	$4,511,488
						NPW @ 12%	$14,963,752
						IRR	0.383
						NPW @ IRR	5.6E-10

Net Present Worth at 12 %
$14,963,752

Internal Rate of Return =
0.383

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
9	$13,192,469	$11,218,086	19148906

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	7976878.58	1777978.96	1463228.63
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
1979662.26						
	19148906.17	350000	2148597	5.00	10742985	10742985
	19148906.17	250000	2148597	5.00	10742985	10742985
	19148906.17	150000	2148597	5.00	10742985	10742985
	19148906.17	90000	2148597	5.00	10742985	10742985
	19148906.17		2148597	5.00	10742985	10742985
	19148906.17		2148597	5.00	10742985	10742985
	19148906.17		2148597	5.00	10742985	10742985
	19148906.17		2148597	5.00	10742985	10742985
	19148906.17		2148597	5.00	10742985	10742985
	19148906.17		2148597	5.00	10742985	10742985

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($13,197,748)
40428001.15	0	0	0.1429	1139895.95	38108367.10	15243346.84	$5,685,748
40428001.15	0	0	0.2449	1953537.56	38474463.59	15389785.44	$5,639,310
40428001.15	0	0	0.1749	1395156.06	39032845.09	15613138.04	$5,515,957
40428001.15	0	0	0.1249	996312.13	39431689.02	15772675.61	$5,416,419
40428001.15	0	0	0.0893	712335.26	39715665.90	15886266.36	$5,392,829
40428001.15	0	0	0.0892	711537.57	39716463.58	15886585.43	$5,392,510
40428001.15	0	0	0.0893	712335.26	39715665.90	15886266.36	$5,392,829
40428001.15	0	0	0.0446	355768.78	40072232.37	16028892.95	$5,250,202
40428001.15	0	0			40428001.15	16171200.46	$5,107,895
40428001.15	0	0			40428001.15	16171200.46	$5,107,895

NPW @ 12%	$17,581,340
IRR	0.406
NPW @ IRR	-3.5E-10

Net Present Worth at 12 %
$17,581,340

Internal Rate of Return =
0.406

Capacity Optimization Economic Calculations

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
10	$14,053,373	$11,950,148	21245699

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	8497427.65	1894004.96	1558714.89
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
2108849.56						
	21245698.85	350000	2387330	5.00	11936650	11936650
	21245698.85	250000	2387330	5.00	11936650	11936650
	21245698.85	150000	2387330	5.00	11936650	11936650
	21245698.85	90000	2387330	5.00	11936650	11936650
	21245698.85		2387330	5.00	11936650	11936650
	21245698.85		2387330	5.00	11936650	11936650
	21245698.85		2387330	5.00	11936650	11936650
	21245698.85		2387330	5.00	11936650	11936650
	21245698.85		2387330	5.00	11936650	11936650
	21245698.85		2387330	5.00	11936650	11936650

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($14,058,997)
44920001.28	0	0	0.1429	1214282.41	42525980.76	17010392.31	$6,313,910
44920001.28	0	0	0.2449	2081020.03	42838981.25	17135592.50	$6,288,710
44920001.28	0	0	0.1749	1486200.10	43433801.18	17373520.47	$6,150,782
44920001.28	0	0	0.1249	1061328.71	43858672.57	17543469.03	$6,040,833
44920001.28	0	0	0.0893	758820.29	44161180.99	17664472.40	$6,009,830
44920001.28	0	0	0.0892	757970.55	44162030.73	17664812.29	$6,009,490
44920001.28	0	0	0.0893	758820.29	44161180.99	17664472.40	$6,009,830
44920001.28	0	0	0.0446	378985.27	44541016.01	17816406.40	$5,857,896
44920001.28	0	0			44920001.28	17968000.51	$5,706,302
44920001.28	0	0			44920001.28	17968000.51	$5,706,302
						NPW @ 12%	$20,243,015
						IRR	0.427
						NPW @ IRR	1.0E-10

Net Present Worth at 12 %
$20,243,015

Internal Rate of Return =
0.427

Capacity Optimization Summary of Results

I have performed the economic analysis on different capacities as multiples of the Optimization I capacity. Here are the important results:

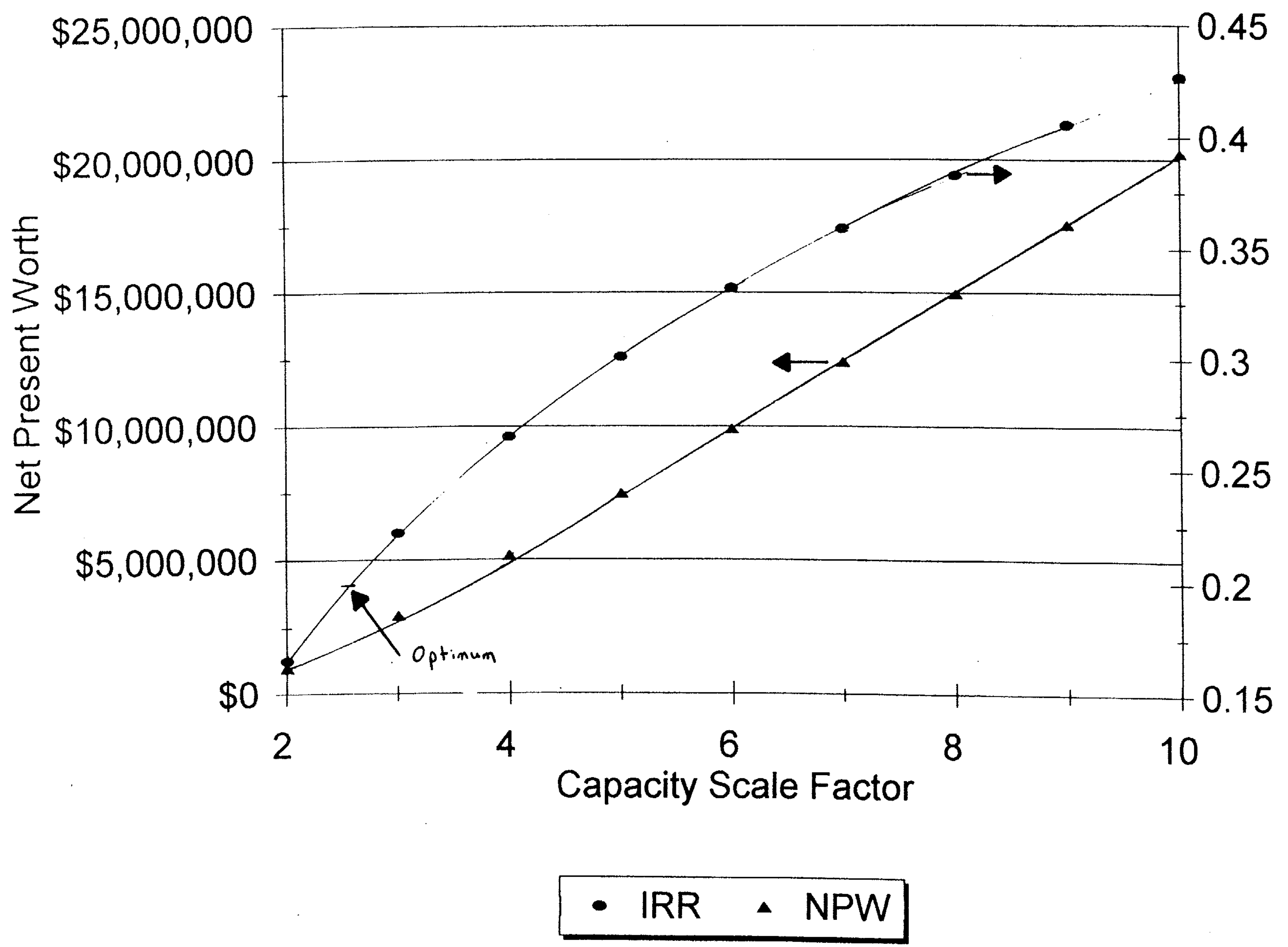
Scale factor = $\frac{\text{new capac.}}{\text{old capac.}}$	IRR	NPW
2	0.165	$990,293
3	0.222	$2,963,229
4	0.265	$5,147,449
5	0.301	$7,470,869
6	0.332	$9,895,477
7	0.359	$12,398,204
8	0.383	$14,963,752
9	0.406	$17,581,340
10	0.427	$20,243,015

I have proceeded up to the 10-fold limit on the 0.6 rule recommended by Peters and Timmerhaus. I realize of course that the IRR's are greater than the problem restriction, but I wanted to see the results. IRR and NPW are graphed on the next page.

Now I'll explore the region between scale factors of 2 and 3. I want the scale factor as high as possible in order to increase the Net Present Worth, but I don't want to go over 20% IRR.

Using the solver on the spreadsheet (I could just as easily have read it off my graph), I obtain:

Scale Factor	=	2.573
Net Present Worth	=	$2,059,015
Internal Rate of Return	=	0.200


$25,000,000
$20,000,000
$15,000,000
$10,000,000
$5,000,000
$0
Net Present Worth
0.45
0.4
0.35
0.3
0.25
0.2
0.15
IRR
2
4
6
8
10
Capacity Scale Factor
Optimum
IRR
NPW

Capacity Optimization Results [cont'd]

The next three pages show the economic calculations for this optimum capacity. Now that I know what the approximate optimum is by using the 0.6 rule, I'll scale up all of my process streams and re-size all equipment. Then I'll do the more rigorous economic analysis, which should turn out similarly.

Notes on Graph

The IRR appears to rise steadily and then approach a maximum. However, I have not gone to that point because the recommended limit on the 0.6 rule is a 10-fold capacity range.

The NPW shows a very slight initial curvature and then becomes a straight line. This trend is not surprising since the NPW can often be directly correlated to the function $f = TCI + 2(oper.)$, which has the form:

$$f = a(\text{Scale Factor})^{0.6} + b(\text{Scale Factor})$$

when related to capacity.

Comparison of Capacities

The optimum capacity, with a scale factor of 2.573 is:

$$\text{Capacity} = 238{,}733 \ \frac{\text{bbl}}{\text{yr}} \times 2.573$$

$$= 614{,}260 \ \text{bbl/yr}$$

To determine how reasonable this value is, I'll compare it to the U.S. industry leader, Anheuser-Busch. According to the A-B annual report, the company sold:

$$88.5 \times 10^6 \ \text{bbl of beer in 1994} \quad (\text{p. R-1})$$

Now, according to Reference 7, p. R-9, NA beer accounts for 1.5 - 2.5% of total beer sales. For Anheuser-Busch, this amounts to:

$$0.015\,(88.5 \times 10^6 \ \text{bbl/yr}) \ \text{to} \ 0.025\,(88.5 \times 10^6)$$

$$= 1{,}327{,}500 \ \text{to} \ 2{,}212{,}500 \ \text{bbl/yr}$$

My yearly production, then is from about 1/4 to 1/2 of that of the industry leader and is thus quite reasonable.

Capacity Optimization Economic Calculations → Best Case

All cost figures in dollars unless otherwise stated

Section III: Cash Flow Table

Base Capacity =	238733	bbl/yr
Base TCI =	$3,530,048	
Base Operating =	2052869	$/yr

(less maintenance and local taxes)

Capacity Scale Factor	New TCI	New Fixed Capital	New Operating
2.573	$6,223,972	$5,292,494	5600052

↑
Best Case

Beginning of year	End of year	Direct Costs	Indirect Costs	Fees & Contingency
1999	0	3763349.43	838818.85	690325.24
2000	1			
2001	2			
2002	3			
2003	4			
2004	5			
2005	6			
2006	7			
2007	8			
2008	9			
2009	10			

Working Capital	Total Operating Costs	Advertising	NA beer sold (bbl)	Price per bbl	Profit on NA beer sales	Opportunity Cost ($5.00/bbl)
933969.45						
	5600051.80	350000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80	250000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80	150000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80	90000	614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967
	5600051.80		614314.6959	5.00	3071573.4797	3071573.47967

Excise taxes Saved	Sales of EtOH (gal)	Income from EtOH sales	MACRS Factor	Depreciation	Taxable Income	Taxes	Net Cash Flow
							($6,226,463)
11558945.32	0	0	0.1429	537782.63	9841424.58	3936569.83	$1,672,324
11558945.32	0	0	0.2449	921644.28	10637301.04	4254920.42	$1,453,973
11558945.32	0	0	0.1749	658209.82	10900735.50	4360294.20	$1,448,599
11558945.32	0	0	0.1249	470042.34	11088902.97	4435561.19	$1,433,332
11558945.32	0	0	0.0893	336067.10	11222878.21	4489151.29	$1,469,742
11558945.32	0	0	0.0892	335690.77	11223254.55	4489301.82	$1,469,592
11558945.32	0	0	0.0893	336067.10	11222878.21	4489151.29	$1,469,742
11558945.32	0	0	0.0446	167845.38	11391099.93	4556439.97	$1,402,454
11558945.32	0	0			11558945.32	4623578.13	$1,335,315
11558945.32	0	0			11558945.32	4623578.13	$1,335,315
						NPW @ 12%	$2,089,015
						IRR	0.200
						NPW @ IRR	6.8E-11

Net Present Worth at 12 %
$2,089,015

Internal Rate of Return =
0.200

Final Optimization

(Final Optimization)

Capacity Material Balance Changes + Resizing + Costing

Now that I have found the optimum capacity of my plant (= 2.573 × Opt. I capacity = 614,260 bbl/yr), I need to go back through the design and re-size all of the equipment in a rigorous manner. I also need to adjust the material balance in a few places. Most of the streams will involve a simple multiplication by 2.573, but a few, like the steam to X-1 will be different since air leakage rates are not linear.

Streams 2, 3, 4, 5, 6, 29, 8, 9, 23, 19, 27, 28, 16, 15, 14, 20, 10, 25, 26

Just multiply Opt. I by $\underline{\underline{2.573}}$ for extensive properties (intensive remain same, obviously)

F-1 De-esterizer

The flow to this vessel is Stream 4.

$L = 25377\ \text{lb/hr}, \quad \rho_L = 61.362\ \text{lb/ft}^3$

Allowing for a 1/2 full vessel with 5 min liquid residence time,

$$V_v = 2Lt/\rho_L$$

$$= 2\left(25377\ \text{lb/hr}\right)\left(5\ \text{min}\right)\left(\frac{1\ \text{hr}}{60\ \text{min}}\right) \Big/ 61.362\ \text{lb/ft}^3$$

$$= 68.9\ \text{ft}^3$$

I'll follow the suggestion of Pabst (p. R-24) and let this vessel be 10 times as large as necessary to alleviate any liquid entrainment and/or foaming problems.
Therefore,

$$V_v = 689\ \text{ft}^3$$

Now:

$$D = \frac{H}{4} = \left(\frac{689\ \text{ft}^3}{\pi}\right)^{1/3} = 6.03\ \text{ft} \rightarrow \text{round to } 6.5\ \text{ft}$$

$$\Rightarrow H = 4(6.5\ \text{ft}) = 26\ \text{ft}$$

Actual volume of vessel $= \frac{\pi}{4}(6.5\ \text{ft})^2(26\ \text{ft}) = 862.8\ \text{ft}^3$

$862.8\ \text{ft}^3 \cdot 7.45\ \text{gal/ft}^3 = 6454\ \text{gal}$

Final Optimization [cont'd]

Therefore, I will specify:

26 ft length
6.5 ft diameter
64554 gallons

From Fig. 14-56 of Peters and Timmerhaus, raw cost = \$11300
(304 stainless steel)

1999 cost = 1.1646(11300) = $\boxed{\$13160}$

X-1 Ejector

The volume of F-1 is 862.8 ft^3. From reference 4, the air leakage rate is.

$$MAL = 0.1955\left(862.8\right)^{0.6630}$$

$$= \underline{17.3 \text{ lb/h air leakage}}$$

In addition to this air, there is an appreciable vapor load (Stream 6) which is 187.8 lb/h, MW = 43.54

From Fig. 9-8 of reference 3, ratio = 1.16
From Fig. 9-7 of reference 3, ratio = 1.0 (assuming 80°F)

So, air equivalent load:

$$AEL = 17.3 + \frac{187.8}{(1.16)(1.0)} = \underline{179.2 \text{ lb/h}}$$

Also,

$$P_{discharge}/P_{suction} = 15.7/13.2 = 1.19$$

$$P_{suct}/P_{motive} = 13.2/114.7 = 0.115$$

from Fig. 9.9 of reference 4, $\frac{1}{r} \approx 3.3$

Assume initially that SC = 1.0

Final Optimization [cont'd]

Thus,
$$w_s = \left(179.2 \text{ lb/h}\right)\left(\frac{1}{3.3}\right)(1.0)$$
$$= 54.3 \text{ lb/h}$$

From Fig. 9-10 of reference 4, SC = 1.08

$$\text{New } w_s = 54.3 \text{ lb/h} (1.08) = \underline{58.6 \text{ lb/h}}$$

Since the air leakage rate is not a linear function, this amount of steam is less than 2.573 times the base case value.

Base Case Value = 28.9 lb/h
× 2.573
74.4 lb/h

However, since my material balance for the rest of the process depends upon this amount of steam being added to the bee and then condensed, I'll specify 74.4 lb/h anyway. This will result in a larger amount of vacuum being pulled, but I don't think it will be too much of a problem. If it ends up causing operational difficulties, the steam rate can be lowered back down to 58.6 lb/h and the bottling section of the plant can simply dilute the final product a little. So:

$\boxed{\text{74.4 lb/h of 100 psig motive steam required for this ejector}}$

The cost of the ejector is given by Eq. 8-1 of reference 3:

$$1986 \text{ installed cost} = \$16000\left[NS + 2(NC)\right]\left(\frac{SCON}{1000}\right)^{0.35}$$
$$= 16000\left(1+2(0)\right)\left(74.4/1000\right)^{0.35} = \$6444$$

$$1999 \text{ cost} = 6444(1.1)(1.2799) = \boxed{\$9073}$$

stainless steel   inflation

Final Optimization [cont'd]

Using this amount of steam means that all of the remaining streams in the process will be simple multiples, too.

F-2 Reflux Drum

Input (Stream 14) = 5585 lb/hr
ρ_L = 50.487 lb/ft^3

Sizing as before,

$$V_v = 2\left(5585\ \frac{lb}{hr}\right)(5\ min)\left(\frac{1\ hr}{60\ min}\right) \Big|\ 50.487\ lb/ft^3$$

$= 18.4\ ft^3$

Then,

$D = H/4 = \left(18.4\ ft^3/\pi\right)^{1/3} = 1.8\ ft \longrightarrow$ round to 2.0 ft

$\Rightarrow$ H = 8.0 ft

Actual volume = $\frac{\pi}{4}(2.0\ ft)^2(8.0\ ft) = 25.1\ ft^3$

$25.1\ ft^3 \cdot 7.48\ gal/ft^3 = 187.7\ gal$

So, I'll specify:

8 ft length
2 ft diameter
187.7 gallons

From Fig. 14-56 of Peters and Timmerhaus,

raw cost = \$3050

1999 cost = 3050(1.1646) = $\boxed{\$3552}$

F-3 Recombine Drum

Combined Flow to this vessel (Streams 7 + 10) = 24201 lb/hr
ρ_L = 61.589 lb/ft^3

Sizing as before,

$$V_v = 2\left(24201\ \frac{lb}{hr}\right)(5\ min)\left(\frac{1\ hr}{60\ min}\right) \Big|\ 61.589\ lb/ft^3$$

$= 65.5\ ft^3$

Final Optimization [cont'd]

Then,

$$D = \frac{H}{4} = \left(\frac{65.5 \text{ ft}^3}{\pi}\right)^{1/3} = 2.75 \text{ ft} \longrightarrow \text{round to } 3.0 \text{ ft}$$

$$\Rightarrow \quad H = 12.0 \text{ ft}$$

$$\text{Actual volume} = \frac{\pi}{4}(3.0 \text{ ft})^2(12.0 \text{ ft}) = 84.8 \text{ ft}^3$$

$$84.8 \text{ ft}^3 \cdot 7.48 \frac{\text{gal}}{\text{ft}^3} = 634.3 \text{ gal}$$

So, I'll specify:

3 ft diameter
12 ft height
634.3 gallons

Using Fig. 14-56 of P+T for 304 stainless steel mixing tank with agitator,

Raw Cost = \$10200

1999 Cost = \$10200 (1.1646)

$= \boxed{\$11879}$

T-1 Distillation Column Re-sizing

The height of the tower remains the same. The diameter, however, will change. I have all the information available from the last resizing (see p. 1 of Optimization Case I). This time, though, my scale factor will be $\underbrace{(98.43)(2.573)}$ rather than 98.43
capacity ratio

Actually, the only place where the flow rate makes a difference is in the final diameter equation (everywhere else it's just a flow ratio). The equation is:

$$D_T = \left[\frac{4V}{0.85\, u_f\, \pi (1 - A_d/A)\, \rho_V}\right]^{0.5}$$

(eq. 3-7 of Henley + Seader)

Final Optimization [cont'd]

Plugging in numbers,

$$D_T = \left[\frac{4\left(22\ \text{lb/h}\right)\left(98.43\right)\left(2.573\right)\ \frac{1\ \text{h}}{3600\ \text{s}}}{0.85\left(28.06\ \text{ft/s}\right)\pi\left(1-0.1\right)\left(0.00382\ \text{lb/ft}^3\right)} \right]^{1/2}$$

** values from pp. 3-4 of Opt. I calculations

$D_T = 4.90$ ft → round up

$$\boxed{\text{Tower diameter} = 5.0\ \text{ft}}$$

To recost the tower, I'll use Fig. 16-28 of Peters and Timmerhaus with stainless steel sieve trays. Diameter = 60 in

$$\boxed{\text{Height} = 135\ \text{ft}\ \text{(from earlier)}}$$

From the figure, raw cost = \$3000/ft

1990 cost = \$3000/ft · 135 ft = 405000

1999 cost = 405000 (1.1646) = $\boxed{\$471663}$

X-2 and X-3 Ejector System

Volume of new tower = $\frac{\pi}{4}\left(5\ \text{ft}\right)^2\left(135\ \text{ft}\right) = 2650.7\ \text{ft}^3$

From reference 4,

$MAL = 0.1451\left(2650.7\right)^{0.6617} = 26.7$ lb/h Air leakage

There is also some CO_2, 12.25 lb/h to be exited.

→ Note: earlier in my calculations I had a value of 195.3 lb/h. This is incorrect, I think. I have no idea where this value came from.

Final Optimization [cont'd]

From Fig. 9-8 of reference 3, ratio = 1.18

So the air equivalent load is:

$$AEL = \frac{26.7 \text{ lb/h}}{(1.0)(1.0)} + \frac{12.25 \text{ lb/h}}{(1.0)(1.18)} = \underline{37.1 \text{ lb/h}}$$

From Fig 9-11 of reference 3, the r value (25.85 torr, 2 stages w/ intercondensers, 90° CW) is:

$$r \approx 10 \text{ lb steam / lb air}$$

Using equation 9-5 of reference 3,

$$w_s = (AEL)(r)(MPC)(SC)$$

MPC = 1.0
Assume for now that SC = 1.0

$$w_s = (37.1 \text{ lb/hr})(10)(1.0)(1.0)$$

$$= 371 \text{ lb/hr}$$

Per stage stm. requirement = 371 / 2 = 185.5 lb/hr per stage

From Fig. 9-10 of reference 3, size correction factor = 1.0

So, I will specify:

$$\boxed{\begin{array}{l}\text{Two-stage ejector system} \\ \quad 185.5 \text{ lb/hr } 100 \text{ psig steam per stage}\end{array}}$$

Note: this is smaller than my previous case due to my earlier error.

Using Eq. 8-1 of reference 3,

$$\text{1986 installed cost} = \$16000\,(NS + 2NC)\left(\frac{SCON}{1000}\right)^{0.35}$$

$$= 16000\left[2 + 2(2)\right]\left(185.5/1000\right)^{0.35}$$

$$= \$53234$$

Final Optimization [cont'd]

$$1999 \text{ cost} = \$53234\,(1.2799)(1.2) = \$81761$$

inflation   surface condensers

$$\boxed{\text{X-2 and X-3 combined, installed, with condensers, 1999 cost} = \$81761}$$

Remaining Sizing Calculations

I'll check these as I go along

My remaining calculations are for heat exchangers and pumps and use several pieces of information already calculated. Rather than recalculate everything, I'll simply refer to my Opt. I calculations, pp. 1-13. I'll also refer to my revised base case calculations, pp. 10-18 and pp. 36-45

E-5 + E-6 Ejector condenser

$\Delta T_{lm} = 232.6\ °F$ 185.5 lb/hr steam

$U = 160\ \frac{Btu}{hr\ ft^2\ °F}$ $\Delta H_{lv} = 880.6$ Btu/lbm

(see Opt. I)

$$\text{Now,}\quad A_s = \frac{(185.5\ lb/h)(880.6\ Btu/lb)}{160\ \frac{Btu}{hr\ ft^2\ °F}\cdot 232.6\ °F}$$

$$= 4.4\ ft^2 \longrightarrow \text{Hard to get an exchanger that small, use}$$

$$\boxed{2 \text{ exchangers, } 10\ ft^2 \text{ each}}$$

(Cost already accounted for with ejectors)

$$\text{Heat duty} = (880.6\ Btu/lb)(185.5\ lb/h)$$

$$= 163.4 \times 10^3\ Btu/hr$$

Cooling water required:

$$m = \frac{163.4 \times 10^3\ Btu/hr}{(77.97 - 58.02)_{Btu/lb}} = 5454.1\ lb/h \text{ CW each}$$

[enthalpy change of H_2O

Final Optimization [cont'd]

E-4 Condenser (see Opt. I)

This is a straight linear scale up. Old $A = 42.3\ ft^2$

New $A = 2.573 \times 42.3\ ft^2$

$= \boxed{109\ ft^2}$

(see Opt. I) From Fig 15-15 of P + T,

Purchased cost = \$4000

1999 cost = \$4000 (1.1646) = $\boxed{\$4658}$

Old $\dot{m}_{NH_3} = 353.1\ lb/hr$

New $\dot{m}_{NH_3} = 353.1\ lb/hr \times 2.573 = 908.5\ lb/hr$ NH_3 circulation

Heat duty $= 0.19414 \times 10^6\ Btu/hr \times 2.573 = 499.5 \times 10^3\ Btu/hr$

E-3 Product cooler (see revised base case)

Another linear scale up.

Old $A = 105\ ft^2$

New $A = (105\ ft^2)(2.573) = \boxed{270\ ft^2}$

From Fig. 15-15 of P + T,

Purchased cost = \$6400

1999 cost = 6400 (1.1646) = $\boxed{\$7453}$

Old $\dot{m}_{NH_3} = 858\ lb/hr$

New $\dot{m}_{NH_3} = 2.573 \times 858\ \frac{lb}{hr} = 2207.6\ lb/hr$ NH_3 circulation required

Heat duty $= 0.47191 \times 10^6\ Btu/hr \times 2.573 = 1214 \times 10^3\ Btu/hr$

Final Optimization [cont'd]

E-2 Feed Preheater (see revised base case)

Heat duty = $0.057884 \times 10^6 \frac{Btu}{hr} \times 2.573$

$= \underline{149 \times 10^3 \ Btu/hr}$

Another linear scale-up.

Old A = 15.1 ft^2

New A = 15.1 ft^2 × 2.573 = $\boxed{38.9 \ ft^2}$

From Fig 15-15 of P+T (extrapolating curve slightly)

Purchased cost = \$2800

1999 cost = 2.573 (\$2800) = $\boxed{\$7204}$

Also, old $\dot{m}_{H_2O}$ = 5794 lb/hr

new $\dot{m}_{H_2O}$ = 2.573 (5794 lb/hr) = 14908 lb/hr

Hot H_2O circulation

E-1 Feed Preheater / Product Cooler (see revised base case)

Heat duty = 0.39412×10^6 Btu/hr · 2.573

$= \underline{1014 \times 10^3 \ Btu/hr}$

Another linear scale-up.

Old A = 96.7 ft^2

New A = 96.7 ft^2 × 2.573 = $\boxed{249 \ ft^2}$

From Fig. 15-15 of P+T,

Purchased cost = \$6000

1999 cost = \$6000 (1.1646) = $\boxed{\$6988}$

E-7 Reboiler (see Opt. I)

Another linear scale-up.

Old A = 1133 ft^2

New A = 1133 ft^2 × 2.573 = $\boxed{2915 \ ft^2}$

Heat duty = $0.015556 \times 10^6 \frac{Btu}{hr} \times 98.43 \times 2.573$ = $\underline{3940 \times 10^3 \ Btu/hr}$

Final Optimization [cont'd]

From Fig. 15-15 of P+T,

Purchased cost = \$33000

1999 cost = \$33000 (1.1646) = $\boxed{\$38432}$

Old $\dot{m}_{H_2O}$ = 109370 lb/hr

new $\dot{m}_{H_2O}$ = 109370 (2.573) = 281409 lb/hr hot H_2O circulation

E-8A After flash condenser (CW section) (see revised base case)

Another linear scale-up. Heat duty = 33495 Btu/hr × 2.573 = 86.2×10^3 Btu/hr

Old A = 4.41 ft^2

New A = 4.41 ft^2 × 2.573 = $\boxed{11.3\ ft^2}$

From Fig. 15-15 of P+T,

Purchased cost = \$2200

1999 cost = \$2200 (1.1646) = $\boxed{\$2562}$

Old $\dot{m}_{cw}$ = 917 lb/h

New $\dot{m}_{cw}$ = 917 lb/h × 2.573 = 2359.4 lb/hr CW

E-8B After flash condenser (NH_3 section) (see revised base case)

Another linear scale-up.

Old A = 0.8 ft^2

New A = 0.8 ft^2 × 2.573 = 2 ft^2 ⟶ they don't make them this small, so I'll use $\boxed{10\ ft^2}$

From this calculation earlier,

1999 cost = $\boxed{\$2446}$

Final Optimization [cont'd]

Old $\dot{m}_{NH_3}$ = 16.6 lb/hr

new $\dot{m}_{NH_3}$ = 16.6 lb/hr × 2.573 = 42.7 lb/hr NH_3 circulation

Heat duty = 9135 Btu/hr × 2.573 = 23.5 × 10^3 Btu/hr

E-9 Hot H_2O heater (see Opt. I)

Another linear scale-up.

Old A = 49.1 ft^2

New A = 49.1 ft^2 × 2.573 = $\boxed{126.3\ ft^2}$

From Fig. 15-15 of P+T,

Purchased cost = \$4400

1999 cost = \$4400 (1.1646) = $\boxed{\$5124}$

Old $\dot{m}_{stm}$ = 1738.8 lb/hr

New $\dot{m}_{stm}$ = 1738.8 lb/hr × 2.573 = 4474 lb/hr steam

Heat duty = 1,531,180 Btu/hr × 2.573 = 3940 × 10^3 Btu/hr

E-10 Hot H_2O heater (see revised base case)

Another linear scale-up.

Old A = 1.7 ft^2

New A = 1.7 ft^2 × 2.573 = 4.4 ft^2 → they don't make them this small, so use

$\boxed{10\ ft^2}$

From this calculation earlier,

1999 cost = $\boxed{\$2446}$

Old $\dot{m}_{stm}$ = 65.7 lb/hr

New $\dot{m}_{stm}$ = 65.7 lb/hr × 2.573 = 169 lb/hr steam

Heat duty = 0.057884 × 10^6 Btu/hr × 2.573 = 149 × 10^3 Btu/hr

Final Optimization [cont'd]

P-1A/B Feed Pumps (see revised base case)

$\Delta P = 13.3$ psi
$NPSH_A = 14.4$ ft

gpm = 53.8 (from Stream Attributes)

h = 31.0 ft

From Fig. 14-40 of P+T, this service requires:

$1\frac{1}{2} \times 1\frac{1}{2}$ pump, $\approx 3/4$ hp

Raw cost = \$890

1999 cost = \$890(1.8)(1.1646)(2) = $\boxed{\$3731}$

stainless steel ↗ inflation ↑ spare ↑

Operating cost:

$$0.75\ \text{hp} \cdot 0.746\ \frac{\text{kW}}{\text{hp}} \cdot 8400\ \frac{\text{h}}{\text{yr}} \cdot 0.0494\ \frac{\$}{\text{hp}} = \boxed{\$232.2/\text{yr}}$$

P-2A/B T-1 Feed pumps

$\Delta P = 2$ psi
$NPSH_A = 26.5$ ft

gpm = 51.2
h = 4.67 ft

From Fig. 14-40 of P+T, this service requires:

2×2 pump, $\approx 1/2$ hp (go to next one up)

Raw cost = \$995
1999 cost = \$995(1.8)(1.1646)(2) = $\boxed{\$4172}$

Operating cost:

$$0.5\ \text{hp} \cdot 0.746\ \frac{\text{kW}}{\text{hp}} \cdot 8400\ \frac{\text{h}}{\text{yr}} \cdot 0.0494\ \frac{\$}{\text{hp}} = \boxed{\$154.8/\text{yr}}$$

Final Optimization [cont'd]

P-3A/B Bottoms pumps (see Opt. I)

$\Delta P = 6$ psi
$NPSH_A = -8.8$ ft (put them 10 ft below column exit)

gpm = 56.5
h = 13.9 ft

From Fig. 14-40 of P+T, this service requires:

2 x 2 pump, ≈ 1/2 hp

From P-2A/B, 1999 cost = \$4172 = $\boxed{\$4381}$

New 1999 cost = 4172(1.05) Operating cost = $\boxed{\$154.8/yr}$
↑ high T factor

P-4A/B Reflux pumps (see Opt. I)

$\Delta P = 3$ psi
$NPSH_A = -1.1$ ft (put them about 2 ft below reflux drum)

gpm = 13.8
h = 8.6 ft

From Fig. 14-40 of P+T, this service requires:

1¼ x 1¼ pump, ≈ 1/4 hp

Raw cost = \$690

1999 cost = \$690(1.8)(1.1646)(2) = $\boxed{\$2893}$

Operating cost = 1/2 of 0.5 hp cost = $\boxed{\$77.4/yr}$

Final Optimization [cont'd]

P-5A/B Hot H_2O pumps (see Opt. I)

$\Delta P = 16.3$ psi
$NPSH_A = 28.3$ ft

gpm = 570.7
h = 38.1 ft

From Fig. 14-40 of P+T, this service requires:

4 × 3 pump, ≈ 10 hp

Raw cost = \$2030

1999 cost = \$2030(1.1646)(2)(1.05) = $\boxed{\$4965}$

Operating cost:

$10\ hp \cdot 0.746\ \frac{kW}{hp} \cdot 8400\ \frac{hr}{yr} \cdot 0.0494\ \frac{\$}{kWhr} = \boxed{\$3096/yr}$

P-6A/B Product pumps (see revised base case.)

$\Delta P = 21$ psi
$NPSH_A = 2.9$ ft

gpm = 48.9
h = 48.5 ft

From Fig. 14-40 of P+T, this service requires:

1½ × 1½ pump, ≈ 1½ hp

Raw cost = \$1220

1999 cost = \$1220(1.8)(1.1646)(2)(1.05) = $\boxed{\$5371}$

Operating costs:

$1.5\ hp \cdot 0.746\ \frac{kW}{hp} \cdot 8400\ \frac{hr}{yr} \cdot 0.0494\ \frac{\$}{kWhr} = \boxed{\$464.3/yr}$

Final Optimization [cont'd]

P-7A|B Distillate product pumps (see revised base case)

$\Delta P = 14.5$ psi
$NPSH_A = 4.6$ ft

gpm = 2.75
h = 41.8 ft

From Fig. 14-40 of P+T, this service requires:

$1\frac{1}{2} \times 1\frac{1}{2}$ pump, $\approx \frac{1}{2}$ hp (go to next one up.)

Raw cost: \$1220

1999 cost = \$1220(1.8)(1.1646)(2) = $\boxed{\$5115}$

Operating costs (as for P-2A|B) = $\boxed{\$154.8/yr}$

P-8A|B, Hot H_2O pumps (see revised base case)

$\Delta P = 11.3$ psi
$NPSH_A = 30.3$ ft

gpm = 30.1
h = 26.4 ft

From Fig. 14-40 of P+T, this service requires:

$1\frac{1}{2} \times 1\frac{1}{2}$ pump, $\approx \frac{1}{2}$ hp

Raw cost = \$860

1999 cost = \$860(1.8)(1.1646)(2)(1.05) = $\boxed{\$3786}$

Operating costs (as for P-2A|B) = $\boxed{\$154.8/yr}$

Final Optimization [cont'd]

Driver costs

Looking back on my pump sizing, I will need a total of:

2 1/4 hp drivers
8 1/2 hp drivers
2 3/4 hp drivers

2 1 1/2 hp drivers

2 10 hp drivers

I'll use Fig. 14-54 of Peters and Timmerhaus with alternating current, open, drip-proof motors. For the 1/4, 1/2, and 3/4 hp drivers, I'll use:

$170 (lowest cost on figure)

For the 1 1/2 hp driver,

Raw cost = $200

For the 10 hp driver,

Raw cost = $600

$$\text{Total raw cost} = 12(\$170) + 2(\$200) + 2(\$600)$$
$$= \$3640$$

$$1999 \text{ Total driver cost} = \$3640(1.1646)$$
$$= \boxed{\$4239}$$

Final Optimization [cont'd]

Storage tanks

I almost to resize the tanks. Actually, it's not very difficult. I can scale-up the capacity linearly.

Tank 1: NA beer storage (see revised base case)

From before, $V_T = 28444 \text{ ft}^3$

$$\text{New } V_T = 28444 \times 2.573 = 73186.4 \text{ ft}^3$$

I'll go with $H/D = 1/2$. This means that:

$$V_T = \frac{\pi}{4} D^2 H = \frac{\pi}{4} D^2 \left(\frac{D}{2}\right) = \frac{\pi}{8} D^3$$

So,
$$D = \left(\frac{8 V_T}{\pi}\right)^{1/3}$$

$$= \left(\frac{8(73186.4 \text{ ft}^3)}{\pi}\right)^{1/3}$$

$$= 57.1 \text{ ft} \rightarrow \text{round to } 57.5 \text{ ft}$$

$$\text{So } H = 57.1 \text{ ft}/2 = 28.6 \text{ ft} \rightarrow \text{round to } 29 \text{ ft}$$

$$\text{Actual volume} = \frac{\pi}{4}(57.5 \text{ ft})^2 (29 \text{ ft}) \cdot 7.48 \frac{\text{gal}}{\text{ft}^3} = 563{,}281 \text{ gal}$$

So, I'll specify:

Tank 1
57.5 ft diameter
29 ft height
563,281 gallons

From Fig. 14-58 of Peters and Timmerhaus, a cone roof tank with this capacity has

Raw cost = \$120000

1999 cost = \$120000 (1.1646)

$$= \boxed{\$139752}$$

Final Optimization [cont'd]

Tank 2: Distilled Product Storage (see revised base case)

From before, the required volume is

$$V_T = 1613 \text{ ft}^3$$

$$\text{New } V_T = 1613 \text{ ft}^3 \times 2.573 = 4150.2 \text{ ft}^3$$

Again using $H/D = 1/2$,

$$D = \left(\frac{8(4150.2 \text{ ft}^3)}{\pi}\right)^{1/3}$$

$$= 21.9 \text{ ft} \rightarrow \text{round to } 22 \text{ ft}$$

$$\text{So } H = 22 \text{ ft}/2 = 11 \text{ ft}$$

$$\text{Actual volume} = \frac{\pi}{4}(22 \text{ ft})^2(11 \text{ ft}) \cdot 7.48 \frac{\text{gal}}{\text{ft}^3} = 31277 \text{ gal}$$

So, I'll specify:

Tank 2
22 ft diameter
11 ft height
31277 gallons

Extrapolating the curve in Fig. 14-58 of Peters and Timmerhaus,

Raw cost = \$21000

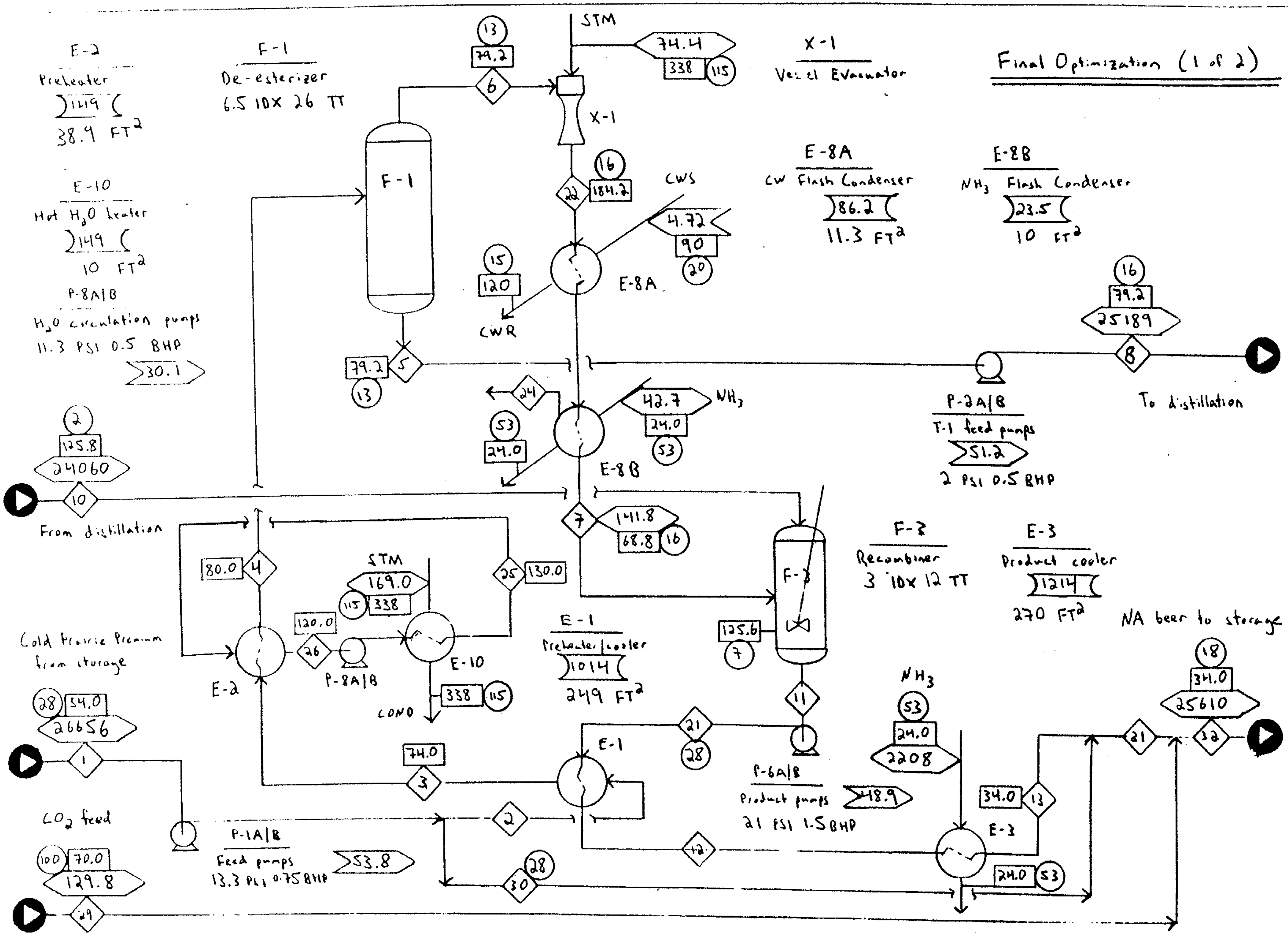
1999 cost = \$21000 (1.1646)

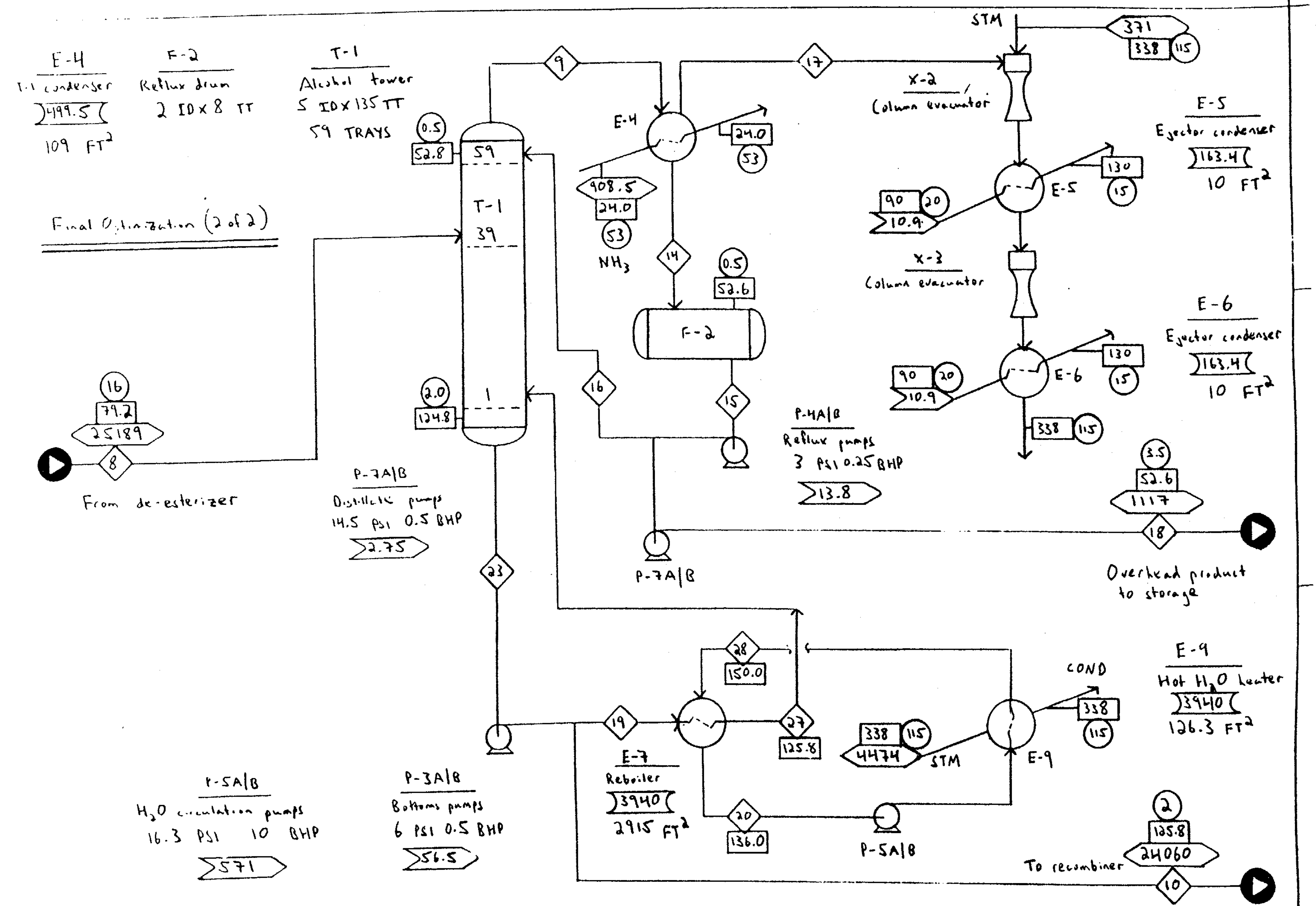
$$= \boxed{\$24457}$$

Final Optimization [cont'd]

Now all that remains to do on this final optimization is to transfer all the information to the flowsheet and do the rigorous economic calculations. I've already done the approximate economics using the 0.6 rule, but I need to do them more exactly. Also, I'll run a case in which the netback on the NA beer is $3 more per barrel than Prairie Premium.

Final Optimization (1 of 2)





E-4
T-1 condenser
499.5
109 FT²
F-2
Reflux drum
2 ID x 8 TT
T-1
Alcohol tower
5 ID x 135 TT
59 TRAYS
Final Optimization (2 of 2)
From de-esterizer
P-7A|B
Distillate pumps
14.5 PSI 0.5 BHP
2.75
P-4A|B
Reflux pumps
3 PSI 0.25 BHP
13.8
X-2
Column evacuator
X-3
Column evacuator
E-5
Ejector condenser
163.4
10 FT²
E-6
Ejector condenser
163.4
10 FT²
STM
NH3
COND
Overhead product to storage
E-7
Reboiler
3940
2915 FT²
E-9
Hot H2O heater
3940
126.3 FT²
P-5A|B
H2O circulation pumps
16.3 PSI 10 BHP
571
P-3A|B
Bottoms pumps
6 PSI 0.5 BHP
56.5
To recombiner

Stream Attributes

Final Optimization (1 of 3)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Title	Feed to Process	Feed to Separation	Partially preheated feed	Preheated Feed	Flash Liquid	Flash Vapor	E-8 liquid effluent	T-1 Feed	T-1 Overhead	T-1 Bottoms to recombiner	Recombiner Outlet	Partially cooled recombined
Water	25230	24019	24019	24019	24017	2.14	75.1	24017	326.5	23952	24027	24027
Ethanol	1120	1066.3	1066.3	1066.3	1055.4	10.78	9.34	1055.4	5207.8	13.9	23.2	23.2
Carbon Dioxide	141.0	134.3	134.3	134.3	12.25	122.0	4.53	12.25	12.25	—	4.53	4.53
Pyruvic Acid	55.8	53.3	53.3	53.3	53.3	—	—	53.3	—	53.3	53.3	53.3
Ethyl Acetate	55.8	53.3	53.3	53.3	50.9	2.40	2.08	50.9	50.8	40.9	42.9	42.9
Air	—	—	—	—	—	—	0.0003	—	—	—	0.0003	0.0003
Volatiles	53.3	50.7	50.7	50.7	—	50.7	50.7	—	—	—	50.7	50.7
Total	26656	25377	25377	25377	25189	187.8	141.8	25189	5597.4	24060	24201	24201
Volumetric	53.8	51.2	51.5	51.5	51.2	31.4	0.31	50.9	24374	48.6	49.0	48.6
Temperature	34.0	24.0	74.0	80.0	79.2	79.2	18.8	79.2	52.8	125.8	125.6	84.3
Pressure	28.0	28.0	23.0	18.0	13.2	13.2	15.7	15.2	0.50	2.0	7.0	23.0
Phase	L	L	L	L	L	V	L	L	V	L	L	L
Molecular Weight	18.61	18.61	18.61	18.61	18.55	43.54	20.24	18.55	42.41	18.08	18.09	18.09
Density	61.785	61.785	61.455	61.362	61.507	0.106	59.641	61.507	0.0041	61.607	61.589	62.146
Enthalpy	-176081	-167634	-163445	-167636	-167127	-696.5	-871.5	-166782	-13609	-162459	-163367	-164374

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Stream Attributes Final Optimization (2 of 3)

Stream Number	13	14	15	16	17	18	19	20	21	22	23	24
Stream Title	Preliminary Product	T-1 Condensate	Reflux Drum Effluent	Pumped Reflux	T-1 Air + CO_2	Overhead Product	T-1 bottoms to reboiler	Reboiler H_2O	Pumped recombiner outlet	X-1 Discharge	T-1 total bottoms	E-8 vapor effluent
Water	24027	326.5	326.5	261.2	—	65.3	3820.6	281409	24027	76.4	27772	1.34
Ethanol	23.2	5208	5208	4166	—	1041.6	26.73	—	23.2	10.78	40.6	1.44
Carbon Dioxide	4.53	—	—	—	12.25	—	—	—	4.53	122.0	—	117.4
Pyruvic Acid	53.3	—	—	—	—	—	0.51	—	53.3	—	53.8	—
Ethyl Acetate	42.9	50.8	50.8	40.7	—	10.16	18.27	—	42.9	2.21	59.2	0.152
Air	0.0003	—	—	—	26.7	—	—	—	0.0003	17.3	—	17.3
Volatiles	50.7	—	—	—	—	—	—	—	50.7	50.7	—	—
Total	24201	5585	5585	4468	39.0	1117	3866	281409	24201	279.4	27926	137.6
Volumetric	48.4	13.8	13.8	11.0	220	2.75	7.82	570.7	48.9	63.4	56.5	20.2
Temperature	34.0	52.6	52.6	52.6	52.8	52.6	124.8	136.0	125.6	184.2	124.8	68.8
Pressure	18.0	0.5	0.5	3.5	0.5	3.5	8.0	14.7	28.0	15.7	2.0	15.7
Phase	L	L	L	L	V	L	L	L	L	V	L	V
Molecular Weight	18.09	42.39	42.39	42.39	32.44	42.39	18.09	18.016	18.085	32.09	18.09	40.83
Density	62.422	50.487	50.487	50.487	0.003	50.487	61.60	61.434	61.589	0.073	61.60	0.114
Enthalpy	-165585	-15875	-15875	-12700	-45.82	-3175	-26093	-1,902,404	-163367	-1331	-188472	-462.2

Mass flows in lb/h
Volumetric flows in GPM for liquids, CFM for gases
Temperature in degrees Fahrenheit
Density in lb/ft^3
Enthalpy in 1000 Btu/hr

Stream Attributes

Final Optimization (3 of 3)

Stream Number	25	26	27	28	29	30	31	32				
Stream Title	Preheater H_2O	Preheater H_2O	Vapor to T-1	Reboiler H_2O	CO_2 Feed	Bypass reg. beer	NAB w/o CO_2	Final NAB Product				
Water	14908	14908	3820.6	281409	-	1210.3	25237	25237				
Ethanol	-	-	26.7	-	-	53.8	76.9	76.9				
Carbon Dioxide	-	-	-	-	129.8	6.76	11.30	141.0				
Pyruvic Acid	-	-	0.51	-	-	2.68	55.8	55.8				
Ethyl Acetate	-	-	18.3	-	-	2.68	45.5	45.5				
Air	-	-	-	-	-	-	0.0003	0.0003				
Volatiles	-	-	-	-	-	2.55	53.3	53.3				
Total	14908	14908	3866	281409	129.8	1278.8	25480	25610				
Volumetric	30.1	30.1	11126	573.3	2.70	2.57	50.9	51.2				
Temperature	130.0	120.0	125.8	150.0	70.0	34.0	34.0	34.0				
Pressure	26.0	16.0	2.0	26.0	100	28.0	18.0	18.0				
Phase	L	L	V	L	V	L	L	L				
Molecular Weight	18.016	18.016	18.09	18.016	44.01	18.61	18.11	18.16				
Density	61.697	61.536	0.006	61.177	0.804	61.785	62.389	62.361				
Enthalpy	-100872	-101021	-22058	-1,898,475	-499.9	-8447.4	-174092	-174606				

Mass flows in lb/h

Volumetric flows in GPM for liquids, CFM for gases

Temperature in degrees Fahrenheit

Density in lb/ft^3

Enthalpy in 1000 Btu/hr

Environmental Considerations

Special Environmental Considerations

Now that I am no longer planning to sell the ethanol byproduct, I will need to dispose of it. The most obvious solution is to discharge it to the municipal sewer. Unfortunately, this stream will create a biological oxygen demand (BOD). According to the 1992 <u>Kirk - Othmer Encyclopedia of Chemical Technology</u>, streams with average BOD level exceeding 600 mg/L are charged extra. Thus, I need to evaluate the BOD level in my distillate product.

From my material balance (for Optimization I), the distillate product (stream 18) contains

25.37	lb/hr	H_2O
404.8	lb/hr	EtOH
3.95	lb/hr	Ethyl acetate
434.1	lb/hr	total

$\rho_L = 50.487\ lb/ft^3$

The volumetric flowrate is:

$$434.1\ \frac{lb}{hr} \cdot \frac{1\ ft^3}{50.487\ lb} \cdot \frac{28.31\ L}{ft^3} = 243.4\ L/hr$$

I will do my calculations on a <u>1 hr basis.</u>

Calculating BOD:

EtOH MW = 46.069 Formula = C_2H_6O

Ethyl acetate MW = 88.107 Formula = $C_4H_8O_2$

I'll start with ethanol:

$$C_2H_6O + 3\ O_2 \rightarrow 2\ CO_2 + 3\ H_2O$$

Thus, 3 moles O_2 are demanded per mole ethanol. The # of moles of EtOH in the stream are:

$$404.8\ lb \cdot \frac{453.59\ g}{lb} \cdot \frac{1\ mole}{46.069\ g} = 3985.6\ gmole$$

Environmental [cont'd]

The amount of O_2 required is:

$$3985.6 \text{ gmole EtOH} \cdot \frac{3 \text{ mole } O_2}{\text{mole EtOH}} = 11956.8 \text{ mole } O_2$$

Converting to mg:

$$11956.8 \cancel{\text{mole}}\ O_2 \cdot 32 \frac{g}{\cancel{\text{mole}}} \cdot 1000 \frac{mg}{g} = \underline{38261760 \text{ mg } O_2}$$

Now for Ethyl Acetate:

$$C_4H_8O_2 + 5\,O_2 \rightarrow 4\,CO_2 + 4\,H_2O$$

Thus, 5 moles O_2 are demanded per mole of ethyl acetate. The moles of ethyl acetate in the stream are:

$$3.95 \text{ lb} \cdot 453.59 \frac{g}{lb} \cdot \frac{1 \text{ mole}}{88.107 \text{ g}} = 20.34 \text{ gmole}$$

Amount of O_2 required is:

$$20.34 \text{ gmole Ethyl acetate} \cdot \frac{5 \text{ mol } O_2}{\text{mole ethyl acetate}} = 101.7 \text{ mole } O_2$$

Converting to mg:

$$101.7 \text{ mole } O_2 \cdot 32 \frac{g}{\text{mole}} \cdot 1000 \frac{mg}{g} = \underline{3254400 \text{ mg } O_2}$$

Environmental [cont'd]

Converting to mg/L :

$$BOD = \frac{382617600 \text{ mg} + 3254400 \text{ mg}}{243.4 \text{ L}}$$

$$\boxed{BOD = 1.58 \times 10^6 \text{ mg/L}} \quad !!$$

Obviously, I need a better way to dispose of this stream. Perhaps I could burn it or sell it to someone who doesn't care about the impurities?

One other consideration

The only other stream I need to be concerned about is Stream #24, the vent from E-8B. From my material balance, I know that this is a vapor stream consisting of:

1.34 lb/h H_2O
1.44 lb/h EtOH
117.4 lb/h CO_2
0.152 lb/h ethyl acetate
17.3 lb/h air

Depending on the specific environmental laws, this stream may require some sort of treatment.