A Financial Analysis For The Recovery of High Purity Propylene from Refinery LPG

by

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Introduction:

In a petroleum refinery, propylene is a by-product of Fluid Catalytic Cracking, Coking and Visbreaking. The FCC Unit is by far the largest source of this material. Propylene is used internally in a refinery to manufacture gasoline via Alkylation or Polymerization or, in some cases, as fuel. The whole propylene/propane (PP) stream may also be sold as a chemical feedstock. There are several large, centralized facilities on the U.S. Gulf Coast that process mixed PPs gathered from surrounding refineries and petrochemical facilities into high purity propylene that is used for feedstock to various chemicals and polypropylene production facilities. There are also a number refineries that have onsite separation facilities to recover high purity propylene. Although conventional distillation equipment including a reboiler utilizing low pressure steam may be used to separate propylene and propane, a number of PP Splitter designs have used a heat pump concept where tower overhead vapors are compressed to a pressure with a corresponding condensing temperature suitable for reboiling the tower. This allows operation at pressures lower than that required to condense the reflux/product stream with water or air, which in turn requires fewer fractionation stages and/or lower energy input to achieve the required propylene product recovery and purity.

Following are the results of a study that was prepared to assess the financial viability of an onsite PP Splitter in a refinery. First, the various uses and the demand history for high purity propylene are presented. Next, the interactions of certain process variables are considered, followed by an analysis of the auxiliary facilities required for feed prefractionation, product treating and storage. Finally, the results of a financial analysis to determine the rate of return for in-plant recovery facilities processing a range of refinery PP production rates are presented.

This study includes the optimization of process variables such as operating pressure, number of trays and reboiler temperature difference. Financial variables that are considered include the value of propylene as a chemical feed and as feed to a refinery Alkylation Unit, utility costs, and capital costs for a new PP Splitter facility. The value of the raw feedstock is also calculated as a function of the high purity propylene product price, and project rate of return is calculated with the alternate value of propylene as alkylation feed as a parameter.

Propylene:

Propylene is sold in the merchant market as refinery grade, chemical grade or polymer grade. Refinery grade propylene specifications are as negotiated between the buyer and seller. The propylene content is usually 65-75%, propane is 20-30%, with the balance butane/butylene (BB) and ethylene/ethane. In the U.S., the price is typically calculated based on a posted price with adjustments for actual composition. Chemical grade specifications usually require a propylene purity of 92%+, while polymer grade is 99.5 wt.% minimum propylene with additional limitations on ethylene/ethane, butane/butylene and other contaminants such as dienes, sulfur and arsine.

Worldwide consumption of propylene in 1996 was about 93 billion pounds. Consumption in North America was around 25 billion pounds. World demand is expected to reach about 110 billion pounds in 2000 while North American demand is projected at around 30 billion pounds.[(1)(2)] Worldwide, non-gasoline demand for propylene is distributed as follows [(3)]:

	Percent
Polypropylene:	52
Acrylonitrile:	12
Oxo Alcohols	10
Propylene Oxide	8
Cumene	6
IPA	4
Acrylic Acid	3
Other	5
	100

Roughly 70% of the worldwide production is as a co-product from the manufacture of ethylene, with 28% from refinery PPs and 2% from propane dehydrogenation. Because of the much larger refinery conversion capacity, refinery PPs account for about 45% of the production in the U.S.

In the U.S., over 50 % of the supply is produced by four companies: Exxon, Lyondell, Ultramar Diamond Shamrock and Shell. Most of this capacity is concentrated in large facilities in the Gulf Coast area with capacities in the range of 500 million to over 1 billion pounds per year. There are a number of new grass-roots facilities and expansions that have either recently been completed or are in design/construction.[(4)]

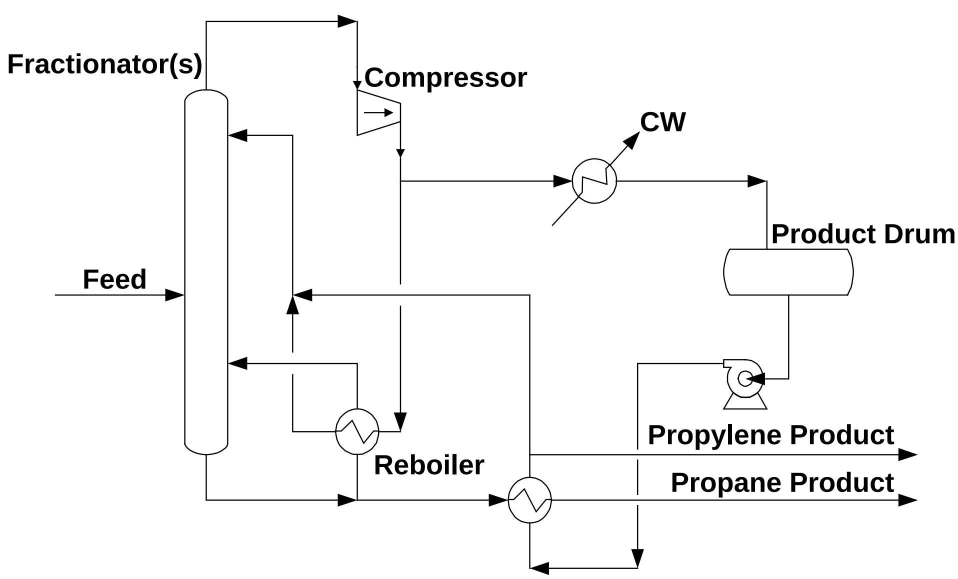
Until the recent financial crisis in the Far East, there was concern that propylene supplies would remain tight. One of the concerns in the U.S. is that a large portion of the new ethylene capacity is based on lighter feedstocks that result in a lower yield of propylene co-product. The long term effects of the downturn in the Far East are unknown at this time, however, the price for polymer grade propylene has dropped to around 15-16 cents per pound early in 1998[5] compared to over 20 cents per pound in the summer of 1997.

PP Splitter Design:

As noted previously, most of the larger PP Splitter facilities utilize a heat pump concept in lieu of a conventional distillation scheme with a low pressure steam reboiler and a condenser. Figure 1 shows a simplified process flow diagram of the heat pump configuration. Overhead vapors are compressed and then a portion is condensed in the reboiler from where the condensate is returned to the top tray for reflux. Net product plus additional reflux are condensed in a parallel exchanger. Depending on the actual column operating pressure, a second compressor stage may be required to increase the pressure of the net product vapors so they may be condensed with air or cooling water. There are numerous other variations where separate cooling is used to remove the heat of compression and/or interchange heat between other warm and cold streams in the system.

Figure 1

Process Flow Diagram
P/P Splitter



There are a number of facilities that use various low level heat mediums (normally low pressure steam) for reboiling instead of a heat pump. Even though most refineries have excess low pressure steam or other low level heat sources available, the magnitude of the requirements for reboiling as well as the cooling water demand for the overhead condensers usually favor the heat pump design. For example, a Splitter operating at 265 psig with 210 trays, processing 6,000 BPSD of FCC PPs (which corresponds to an FCC capacity of around 50,000 BPSD) and designed to produce polymer grade propylene and HD-5 propane would require almost 80,000 lbs/hr of 20 psig steam and 10-15,000 GPM of cooling water. In most refineries this would require new infrastructure facilities to produce these quantities of utilities, increasing the supply costs significantly. In contrast, a comparable heat pump design operating at 85 psig would have a compressor power requirement of around 3,000 HP, which in most cases would be significantly less expensive to accommodate.

Process Variables and Equipment Considerations:

There are a number of process variables that affect the equipment sizes and operating costs. Like all fractionation systems, for a given separation the number of trays required is inversely proportional to the reboiler duty, condenser duty and tower diameter. This effect is more pronounced as the operating pressure is increased. Fig. 2 shows this relationship for a Splitter processing 6,000 BPSD of refinery PPs and yielding a polymer grade propylene product and propane meeting HD-5 specifications. For 150 trays, the required reboiler duty is 58.7 MM Btu/hr at 35 psig operating pressure. For identical fractionation performance, the reboiler duty increases to 67.3 MM Btu/hr at 85 psig, 75.3 MM Btu/hr at 135 psig and 107.8 MM Btu/hr at 265 psig. The tower diameters increase accordingly. It can be seen that this effect is less dramatic as the number of fractionation trays is increased.

Depending on the number of trays, two towers may be required. In a single tower, 150 to 200 trays can be accommodated, depending on the tray spacing. Specialty tray designs, such as UOP MD trays, allow tray spacing as low as 10 in. and are ideally suited for this service because of the relatively high liquid loads. MD trays are slightly less efficient than conventional high capacity trays and are more costly on a per tray basis. However, because of the lower allowable spacing, more trays can be accommodated for a given tower height which will allow a reduction in the tower diameter, compressor horsepower and reboiler area.

A major consideration for the tower mechanical design is the ratio of the length to diameter (L/D) which should generally not exceed 20-25. For the 6,000 BPSD case noted above (150 trays) operating at 85 psig, the required tower diameter is 11'9" and the overall length is almost 250 ft., based on 18 in. tray spacing. This results in an L/D ratio of about 21. For a system capacity of 3,000 BPSD, which corresponds to an FCC capacity of around 25,000 BPSD, the required diameter is reduced to 8' 6" which yields an L/D ratio of almost 30. Depending on the design internal pressure, other factors such as wind or seismic loads may govern the vessel design and increase the overall cost to the extent that a two tower system should be considered.

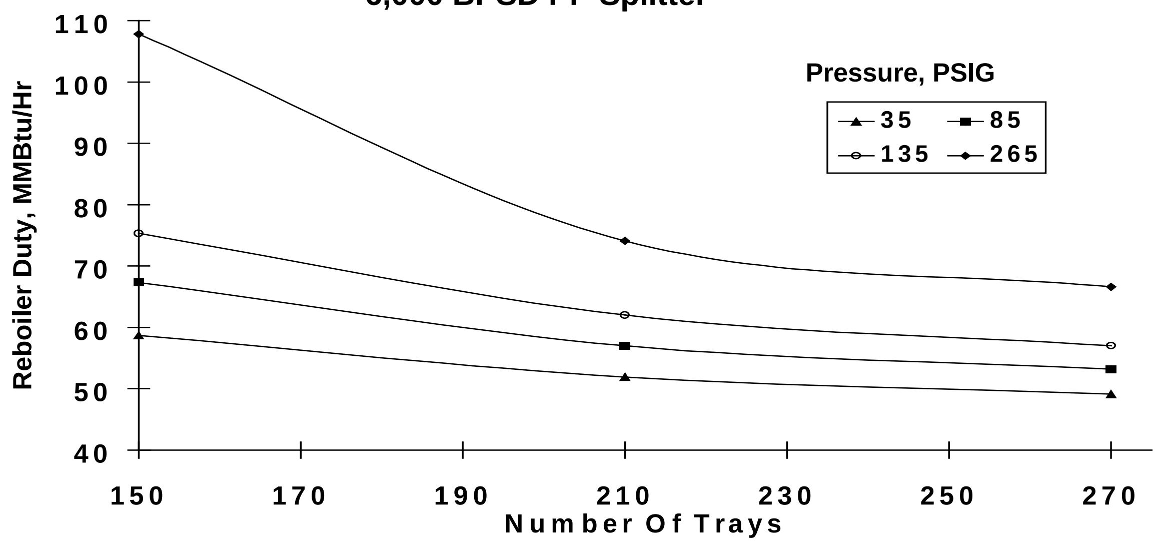
Another important design issue which has a significant impact on the capital cost of a PP Splitter facility is shop versus field fabrication of the tower(s), as limited by the diameter. This is dictated by the capability of the fabrication shop, access of the new PP Splitter facility to water transportation and overland transportation limitations. Generally, towers with diameters of 12-14 ft. can always be shop fabricated and transported to the jobsite. For the cases with tower heights over 120 ft., fabrication in two or more pieces may be required. The above diameter constraints will limit the PP feed to around 8,000-8,500 BPSD for shop fabricated vessels, which corresponds to an FCC capacity of around 70,000 BPSD.

The compressor is a key equipment item in a heat pump configuration. In order to minimize the compressor cost and energy consumption it is important to limit the overall head to about 10,000 ft., which will normally allow the use of a single stage, over-hung centrifugal compressor. For propylene, this corresponds to a compression ratio of around 1.8. For operating pressures lower than 85-100 psig, additional compressor stages are required to increase the pressure of the net propylene product so condensation with air or cooling water can be achieved. The differences in head and flow for the two services (reflux and product) are substantial and may not allow the use of a single, multi-stage compressor body or two bodies with a tandem drive. Two separate compressors and drivers or an integral gear compressor may have to be considered for lower operating pressure designs.

The reboiler temperature difference is also a major variable that affects the compressor design. Lower temperature differences obviously increase the reboiler size, but also reduce the compression ratio and, hence, the size and power consumption of the compressor. Proprietary designs such as UOP high flux tubes yield heat transfer rates two to three times higher than conventional tubes and will allow a lower achievable temperature difference for a constant reboiler surface area. This

Figure 2

Reboiler Duty Vs. Trays
6,000 BPSD PP Splitter



improved performance is more costly from a capital standpoint and has to be evaluated based on the specifics of each project.

Offsite Considerations:

For a refinery considering a PP Splitter to recover high purity propylene, the major offsite issues to consider are prefractionation of the PPs ahead of the splitter, product treating, and product storage. The processing scheme for combined LPG (PPs and BBs) from an FCC and/or Coker will vary depending on the overall refinery configuration. In some cases, the whole stream is fed to a downstream Alkylation or Polymerization Unit, or to both. In other cases, a PP/BB Splitter is used to fractionate the streams for separate processing. Typically, a refinery PP stream will have been amine and/or caustic treated for the removal of hydrogen sulfide. This will also remove any trace of methyl mercaptan that is present and some of the carbonyl sulfide (COS).

Prefractionation:

Typical specifications for polymer grade propylene limit ethylene and ethane concentrations to around 30 and 500 mol ppm respectively and total butane/butylenes to less than 50 mol ppm. Refinery PPs from a PP/BB splitter will normally have 5,000 to 10,000 ppm ethylene/ethane and 1-2 mol % or more BBs. To reduce the feed ethylene/ethane level to that needed for the polymer grade specification requires a Deethanizer with around 50 trays. Depending on the design and performance of the upstream PP/BB Splitter, the Deethanizer design may include additional trays to remove a small BB purge stream from the bottom of the tower while taking the PPs as a vapor side draw. The PP yield from the upstream splitter may also be adjusted to reduce the BB content.

Treating:

Contaminants that are not removed in the upstream treating and fractionation systems and require additional removal facilities are water, oxygenates, trace sulfur including carbonyl sulfide, arsine and phosphine. Typical polymer grade propylene contaminant limits are as follows:

Water, wppm	5
Oxygenates, wppm	5
Total Sulfur,wppm	1
COS , ppb	20
Arsine/phosphine, ppb	20

The water, oxygenates and sulfur compounds can be removed via solid bed absorption with a combination of mol sieves and activated alumina. The removal of arsine and phosphine to 20 ppb is typically accomplished in a vessel with a catalytic adsorbent such as copper oxide.[6]

Product Storage:

The finished product can be stored in bullets, spheres, semi-refrigerated spheres or refrigerated tanks. The factors involved in this decision are the mode of product transport, shipment cycles and sizes, and cost. Bullets and spheres will require design pressures of about 300 psig, and have capacities up to 1,500 and 25,000 Bbl. respectively. Semi-refrigerated spheres will also have capacities up to 25,000 Bbl. and be designed for around 100 psig with a system to recover boil-off vapors. Refrigerated storage tanks are designed for 1-2 psig with capacities usually over 100M Bbls. Systems to recover boil-off vapors, chill incoming product and warm up product send out are required.

Study:

The remainder of this paper is dedicated to the findings of a study that was conducted to determine the feasibility of installing a facility in a refinery to recover polymer grade propylene (99.5 wt% purity). First, a process design evaluation was conducted to determine the operating pressure, number of trays, reboiler duty and reboiler temperature difference for a heat pump system and for conventional distillation using a low pressure steam reboiler. Separate calculations were completed for capacities of 6,000 BPSD and 15,000 BPSD PP feed since it was perceived that field versus shop fabrication for the tower(s) would have a major impact on the results.

Based on the results of the process design calculations, capital investments were developed for PP Splitter capacities (feed) of 3,000, 6,000, 10,000, and 15,000 BPSD. The operating pressure and number of trays selected for the 6,000 BPSD feed case were used for the 3,000 BPSD estimate; similarly, the 15,000 BPSD parameters were used for the 10,000 BPSD case. The cost estimates include all offsite prefractionation, treating and storage equipment. Next, the discounted cash flow rate of return (DCF) was calculated for each capacity case as a function of polymer grade propylene price with the alternate value of propylene as Alkylation feed as a parameter. Finally, the value of the propylene was estimated as a function of polymer grade propylene price for a project rate of return of 15% for each capacity case.

Process Design:

The following procedure and basis was used for determining the equipment sizes and preparing the capital cost estimates for each of the facility capacities noted above.

1. Computer simulations were completed to determine equipment sizes for the following parameters:

   - Feed composition, LV%:

Ethylene:	0.1
Ethane:	1.1
Propylene:	68.6
Propane:	27.8
Butylene:	1.1
Isobutane:	0.8
N butane	0.5

   - Feed Capacities of 6,000 and 15,000 BPSD as noted above
   - Tower top pressures of 35, 85, 135 and 265 psig
   - Fractionation tray efficiency of 80% and 150, 210 and 270 actual trays
   - One column for 150 trays and two columns for 210 and 270 trays
   - For the heat pump cases, reboiler temperature differences of 10, 15, 20 and 25°F were investigated. For the 265 psig cases, the reboiler delta T was set at 90°F based on a maximum flux of 15,000 Btu/hr/ft^2°F.
   - Conventional high capacity trays with 18" tray spacing
   - Kettle reboilers with conventional steel tubes and a maximum bundle area of 35,000 Ft^2
   - Power cost of 4, 6 and 8¢/KWH and low pressure steam costs of $2.50/1000 Lbs.
   - Maximum shop fab diameter of 14'0".

2. A factored cost estimate was prepared for each case including offsite facilities based on a U.S. Gulf Coast location.
3. At each pressure, the evaluated cost was determined with the number of trays and reboiler temperature difference as variants. For this study, evaluated cost is defined as the total facility capital investment plus three years of compressor power or splitter reboiler steam costs. The combination of these variables with the lowest evaluated cost was selected for each pressure.
4. The evaluated costs at each pressure were compared and the lowest chosen as the design point for the subsequent financial analysis.
5. Based on the combination of parameters noted above, equipment sizes and operating costs for 78 cases were prepared.

Process Design Results:

Tables 1 and 2 show the calculation results for the 6,000 and 15,000 BPSD cases, respectively, for power costs of 6¢ /KWH. The minimum evaluated cost options for each pressure level investigated were plotted on Figs. 3 and 4. For a 6,000 BPSD facility, the data shows that two columns with a total of 210 trays operating at 135 psig and with a reboiler temperature difference of 20°F has the lowest evaluated cost. The plots also suggest that the actual minimum cost is at a pressure somewhat higher than 135 psig. This is consistent with the results of previous studies by others.[7][8]

For the 15,000 BPSD case, the indicated minimum evaluated cost is for a single, 150 tray tower operating at 85 psig with a reboiler delta T of 10°F. This result is indicative of the substantially higher cost for the distillation columns that require all field fabrication. The evaluated costs for the conventional distillation system operating at 265 psig with a steam reboiler are substantially higher than all of the heat pump cases. The differences in power cost did not affect the selection of the heat pump design point for any of the cases investigated.

The results of the foregoing process design analysis were then used to prepare additional capital cost estimates for facilities processing 3,000 and 10,000 BPSD of refinery PPs. The key design and cost data for each capacity case is summarized in Table 3.

Figure 3

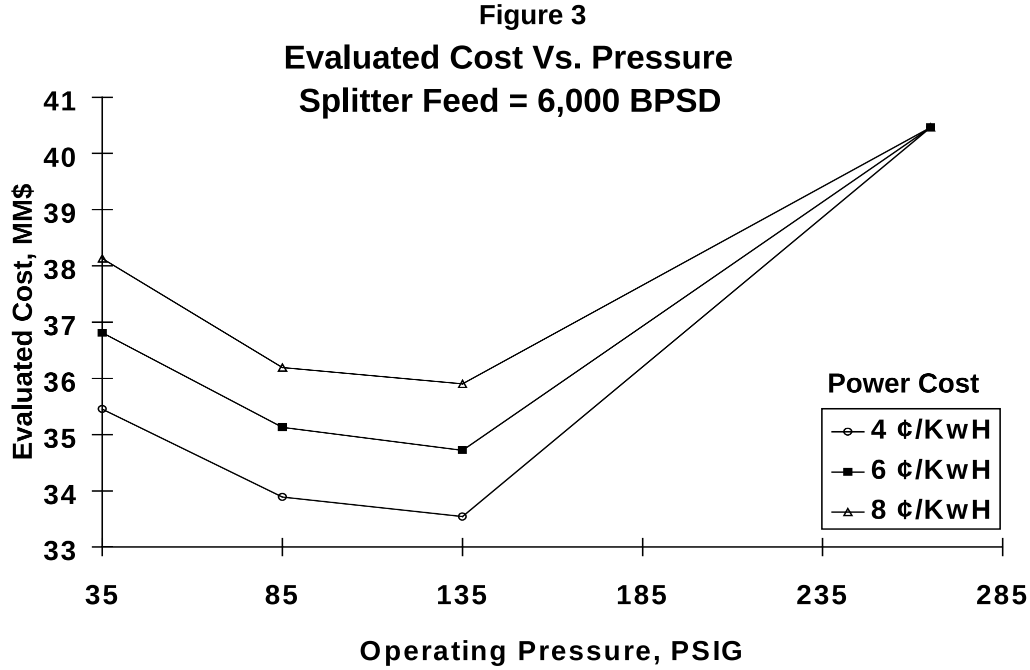


Figure 4

Evaluated Cost Vs. Pressure
Splitter Feed = 15,000 BPSD

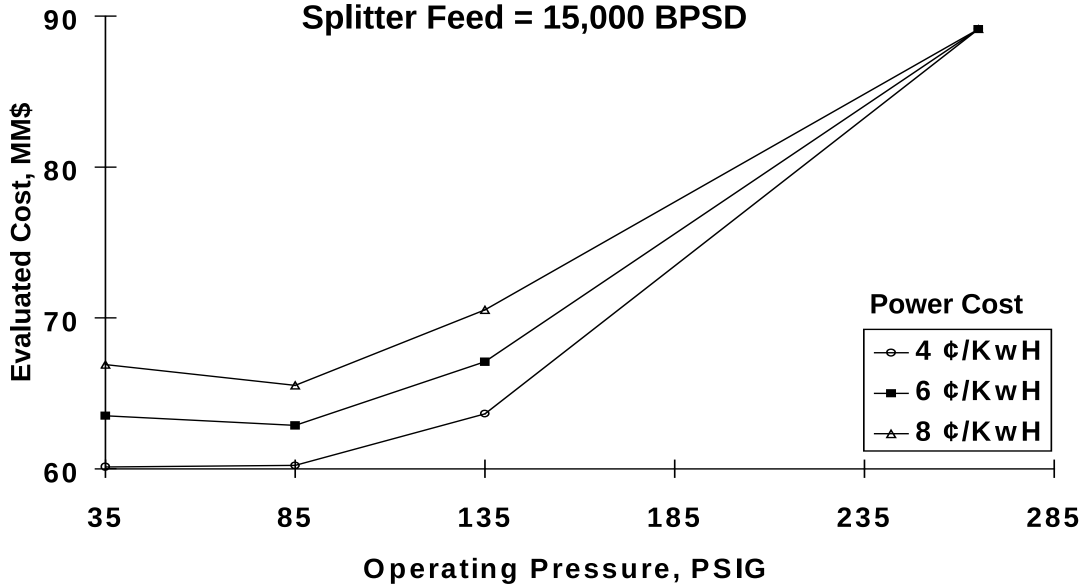


Table 3

Design Cases

Splitter Feed Rate, BPSD	3,000	6,000	10,000	15,000
Operating Pressure, psig	135	135	85	85
Number Of Towers	2	2	1	1
Trays, Total	210	210	150	150
Diameter/Length, Ft	8.5/180	11.5/180	15/250	18.5/250
Compressor Stages	1	1	2	2
Horsepower	1,575	3,150	4,700	7,050
Number Of Reboilers	1	1	3	4
Surface, FT^2 Each	12,350	24,700	28,000	32,000
Miscellaneous HP	100	200	150	210
Total HP	1,675	3,350	4,850	7,260
DeEthanizer Steam, Lb/Hr	11,000	22,000	36,000	55,000
Capital Investment, MM$	24.23	31.17	45.35	54.92

Financial Analysis:

The financial analysis consisted of establishing the value of propylene as an Alkylation Unit feed and then estimating the rate of return for the new facilities over a range of polymer grade propylene prices and feed capacities. The value of propylene as alkylation feed was calculated by assuming alkylate to be worth premium gasoline price and then deducting the value of isobutane and typical operating costs for a Sulfuric Acid Alkylation Unit, all on a per barrel of propylene basis. The results were as follows:

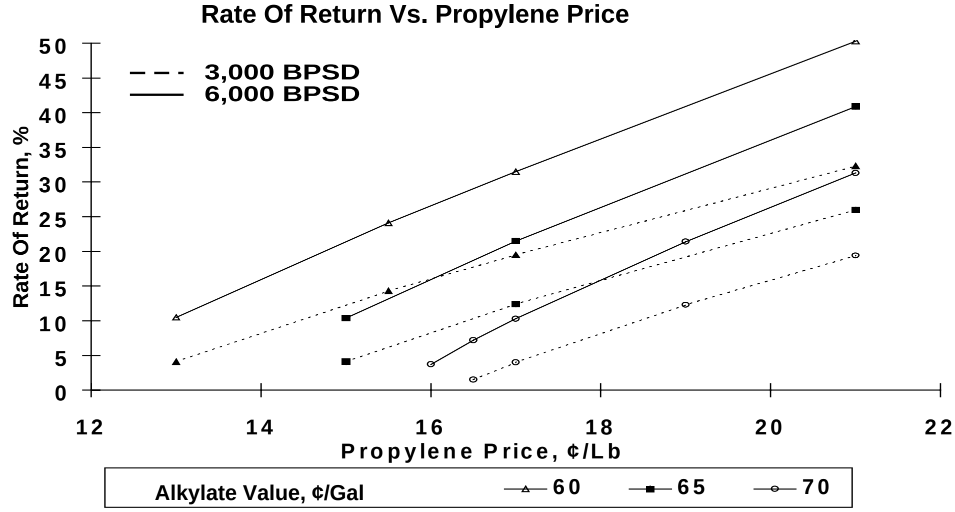
Alkylate Price, ¢/Gal.	60	65	70
Propylene Value as Alky Feed, ¢/Lb.	8.7	10.6	12.7

The rate of return was calculated based on the project cash flow using straight line depreciation and a 15 year life. In addition to utility costs, expenses for labor, maintenance, taxes and insurance for the new facilities were included. It was assumed that there was not an increase in refinery working capital associated with the installation of these facilities. Depending on the facility location, transportation costs to the end user could vary considerably. For this study, a value of 5¢/gal. of propylene product was used.

The calculated rates of return for the various facility sizes are shown on Figs. 5 and 6 as a function of propylene price with the value of Alkylate as a parameter. Figure 7 shows the propylene price required to yield a 15% rate of return as a function of facility capacity. The impact of capacity on project economics is obvious. Assuming a propylene price of 15-16¢/lb.[5] and alkylate worth 65-70¢/gal., only the larger facilities will sustain attractive economics. An alkylate value of 60¢/gal.

Figure 5

Rate Of Return Vs. Propylene Price



or less is required to achieve at least a 15% rate of return for the 3,000 BPSD PP Splitter.

Figure 6

Rate Of Return Vs. Propylene Price

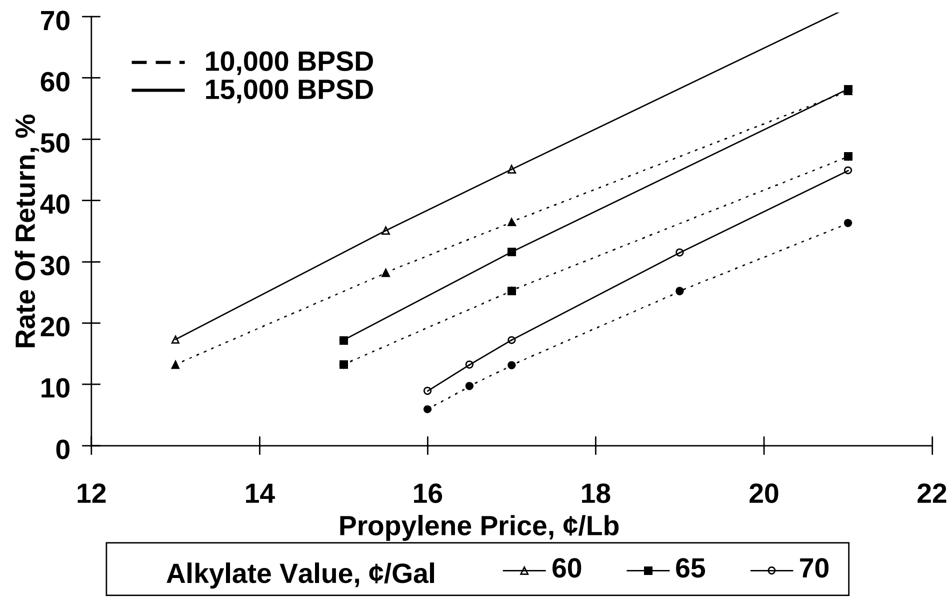
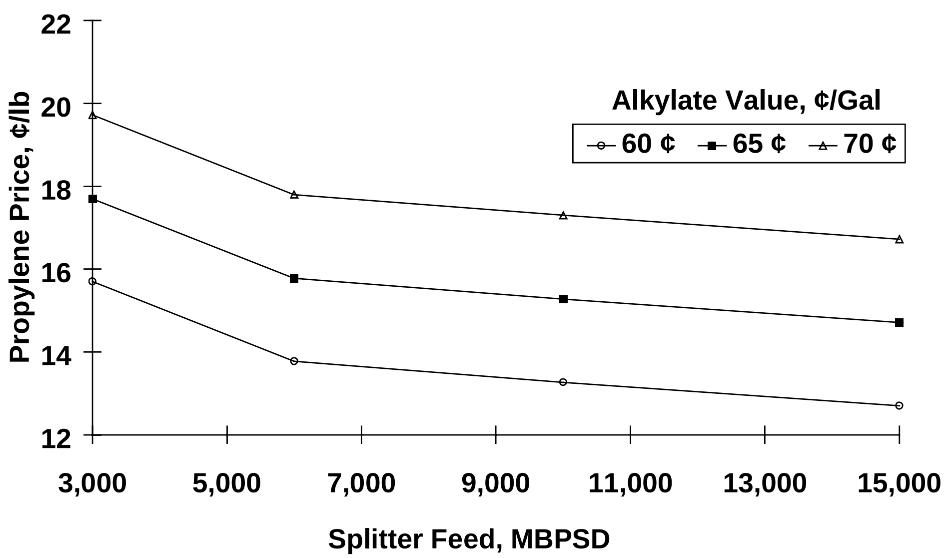


Figure 7

Propylene Value Based On 15% Return On Invested Capital



Conclusions:

Based on the process design calculations and economic evaluations completed for this study, the following may be concluded:

- As expected, the minimum evaluated cost for the 6,000 and 15,000 BPSD cases occur with different system design parameters. The minimum evaluated cost for a 6,000 BPSD facility is for an operating pressure of 135 psig, 210 trays (two towers) and a reboiler temperature difference of 20°F. For the 15,000 BPSD case, the design parameters are 85 psig operating pressure, 150 trays in a single tower and a 10°F reboiler temperature difference. As noted previously, this difference is primarily driven by the higher cost associated with field fabrication of the Splitter Tower for the higher capacity case.

- The design parameters, including facility capacity, resulting in the minimum evaluated cost were not affected by variations in the price of power between 4 and 8¢/KWH.

- The evaluated cost of a conventional distillation system with a steam reboiler was significantly greater than any of the heat pump designs investigated.

- The economics favor the larger capacity facilities, even with propylene as low as 15-16¢/gal. A Splitter designed to process 3,000 BPSD of refinery PPs requires alkylate values of around 60¢/gal. or less.

Table 1
6,000 BPSD
EVALUATED COST ANALYSIS

POWER COST, $/KWH 0.06
LP STEAM COST, $/M LBS 2.50

35 PSIG

	REBOILER	DELTA T =10°F		REBOILER	DELTA T=15°F		REBOILER	DELTA T=20°F		REBOILER	DELTA T=25°F	
ACTUAL TRAYS	150	210	270	150	210	270	150	210	270	150	210	270
COMPRESSOR HP	3612	3347	3432	4008	3787	3837	4443	4250	4223	4911	4650	4633
POWER COST, MM$/YR	1.36	1.26	1.29	1.51	1.42	1.44	1.67	1.60	1.59	1.85	1.75	1.74
CAPITAL COST, MM$	32.74	33.10	38.26	32.46	32.71	37.9	32.46	32.69	38.0	32.35	32.60	37.8
EVALUATED COST, MM$	36.81	36.87	42.13	36.98	36.98	42.2	37.47	37.48	42.7	37.89	37.84	43.0

85 PSIG

	REBOILER	DELTA T =10°F		REBOILER	DELTA T=15°F		REBOILER	DELTA T=20°F		REBOILER	DELTA T=25°F	
ACTUAL TRAYS	150	210	270	150	210	270	150	210	270	150	210	270
COMPRESSOR HP	2822	2570	2533	3308	2947	2882	3747	3341	3248	4207	3726	3627
POWER COST, MM$/YR	1.06	0.97	0.95	1.24	1.11	1.08	1.41	1.26	1.22	1.58	1.40	1.36
CAPITAL COST, MM$	31.95	32.33	36.78	31.40	31.95	36.4	31.23	31.82	36.3	31.13	31.73	36.2
EVALUATED COST, MM$	35.13	35.23	39.64	35.13	35.27	39.6	35.45	35.59	40.0	35.87	35.93	40.3

135 PSIG

	REBOILER	DELTA T =10°F		REBOILER	DELTA T=15°F		REBOILER	DELTA T=20°F		REBOILER	DELTA T=25°F	
ACTUAL TRAYS	150	210	270	150	210	270	150	210	270	150	210	270
COMPRESSOR HP	2687	2330	2447	3164	2725	2612	3662	3150	3011	4199	3602	3440
POWER COST, MM$/YR	1/01	0.88	0.84	1.19	1.02	0.98	1.38	1.18	1.13	1.58	1.35	1.29
CAPITAL COST, MM$	33.68	32.95	38.79	33.06	32.48	38.3	31.73	31.17	37.0	31.54	30.96	36.8
EVALUATED COST, MM$	36.71	35.58	41.32	36.63	35.55	41.3	35.86	34.72	40.4	36.27	35.02	40.7

265 PSIG

	REBOILER	DELTA T =10°F	
ACTUAL TRAYS	150	210	270
REBOILER DUTY,MMBtu/Hr	114.7	78.8	70.8
STEAM COST, MM$/YR	2.41	1.66	1.49
CAPITAL COST, MM$	38.07	35.49	37.20
EVALUATED COST, MM$	45.29	40.46	41.66

____Lowest Evaluated Cost at Pressure

____Lowest Evaluated Cost for 6,000 BPSD Splitter

Table 2
15,000 BPSD
EVALUATED COST ANALYSIS

POWER COST, $/KWH	0.06
LP STEAM COST, $/M LBS	2.50

35 PSIG

	REBOILER DELTA T =10°F			REBOILER DELTA T=15°F			REBOILER DELTA T=20°F			REBOILER DELTA T=25°F		
ACTUAL TRAYS	150	210	270	150	210	270	150	210	270	150	210	270
COMPRESSOR HP	9032	8365	8579	10021	9467	9590	11110	10624	1055	12278	11626	1158
POWER COST, MM$/YR	3.39	3.14	3.22	3.77	3.56	3.60	4.18	3.99	3.97	4.61	4.37	4.35
CAPITAL COST, MM$	53.34	58.04	58.40	52.64	57.19	57.7	52.54	56.82	57.4	52.29	56.82	57.4
EVALUATED COST, MM$	63.52	67.47	68.07	63.94	67.86	68.5	65.07	68.80	69.3	66.13	69.93	70.5

85 PSIG

	REBOILER DELTA T =10°F			REBOILER DELTA T=15°F			REBOILER DELTA T=20°F			REBOILER DELTA T=25°F		
ACTUAL TRAYS	150	210	270	150	210	270	150	210	270	150	210	270
COMPRESSOR HP	7050	6424	6332	8271	7367	7204	9366	8353	8054	10514	9315	9068
POWER COST, MM$/YR	2.65	2.41	2.38	3.11	2.77	2.71	3.52	3.14	3.03	3.95	3.50	3.41
CAPITAL COST, MM$	54.92	58.22	60.04	53.58	57.41	59.2	53.18	56.83	58.7	52.98	56.70	58.6
EVALUATED COST, MM$	<u>62.87</u>	65.46	67.18	62.91	65.72	67.4	63.74	66.25	67.8	64.83	67.20	68.8

135 PSIG

	REBOILER DELTA T =10°F			REBOILER DELTA T=15°F			REBOILER DELTA T=20°F			REBOILER DELTA T=25°F		
ACTUAL TRAYS	150	210	270	150	210	270	150	210	270	150	210	270
COMPRESSOR HP	6716	5826	5617	7910	6812	6530	9154	7867	7526	10497	9005	8601
POWER COST, MM$/YR	2.52	2.19	2.11	2.97	2.56	2.45	3.44	2.96	2.83	3.95	3.38	3.23
CAPITAL COST, MM$	61.37	64.57	68.40	59.85	63.28	67.2	56.77	60.35	64.2	56.34	60.09	63.9
EVALUATED COST, MM$	68.94	71.14	74.73	68.77	70.96	74.5	67.09	69.22	72.7	68.18	70.24	73.6

265 PSIG

	REBOILER DELTA T =10°F		
ACTUAL TRAYS	150	210	270
REBOILER DUTY, MMBtu/Hr	269.4	185.3	166.5
STEAM COST, MM$/YR	6.02	4.14	3.72
CAPITAL COST, MM$	79.84	76.72	78.70
EVALUATED COST, MM$	97.90	89.14	89.86

____ Lowest Evaluated Cost at Pressure
____Lowest Evaluated Cost for 15,000 BPSD Splitter

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* Ms. Andrzejewski is no longer with Mustang.