Evaluation of the Energy Savings Claims of Progressive Distillation

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Abstract

The concept of progressive distillation applied to crude fractionation has started to be mentioned in different circles as a better alternative to conventional schemes that have been around for more than 50 years. Progressive distillation is based on an expired Technip patent claiming that reductions in energy consumption are achieved. We studied this concept as it is explained in the patent and compared it to atmospheric crude columns. We find that progressive distillation can reduce the overall heat utility of a distillation sequence for heavy crude, and even produce more valuable products. It can also reduce the furnace heat utility of a distillation sequence for a light crude, while producing more valuable products as well. Capital costs and heat exchanger networks were not included in the comparison.

1. Introduction

About 2% of the energy content of a crude oil stream is used in the distillation process.[ii] The conventional model was developed in the 1938 and has proven to be useful, efficient, and cost effective.[iv] There are several unconventional distillation designs that have been developed over the last century that merit evaluation as to their potential in energy savings, especially considering recent increases in energy prices. Watkins[v] discussed a few variations to the conventional model such as pump-back reflux and stripping using reboilers. Bagajewicz and Ji (2001)[ii] designed a rigorous procedure for conventional atmospheric crude fractionation units. The concept of prefractionation units was introduced by Burma.[vi] Bagajewicz and Ji (2002)[vii] designed a procedure for crude fractionation units with preflashing or prefractionation. Ji and Bagajewicz (2002)[iii] did research on provided results on stripping-type crude distillation. Because crude fractionation is significantly more complicated than common distillation, there are a variety of approaches to solving the problems that arise.

Crude oil contains a multitude of individual hydrocarbon components, which each have a range of boiling points based loosely on the number of carbon atoms present in the molecule. The crude oil that is most valuable is light crude, which is composed of mostly light components, and therefore has a much higher naphtha and kerosene product potential. On the other end of the spectrum is heavy crude, which is composed mostly of higher boiling point components, and is therefore less valuable by nature. There is a range of crudes available at all compositions between the two as well, and all of them can be separated by distillation into naphtha, kerosene, diesel, and gasoil, with a residual component left over. This residual component can be sent to a vacuum column to recover more gasoil and other useful components. In the conventional model for either, the crude feed is mixed with water and then sent through heat exchangers and into a desalter, which removes impurities and sour water. The resulting more purified crude is then heated in a furnace and sent to the main column.

Conventional distillation is an application of the indirect sequence, (Figure 1a) in which the heaviest components of a crude feed are drawn off the bottom of a column, and what is left (the lighter components) are sent to the next section of the column, where the same procedure is repeated. This is done through the entire height of the column until all that is left is the lightest component.

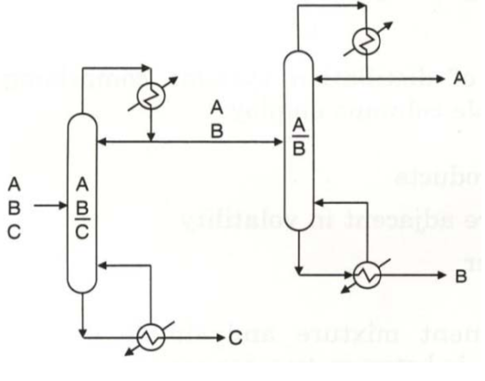


Figure 1a. The Indirect Sequence

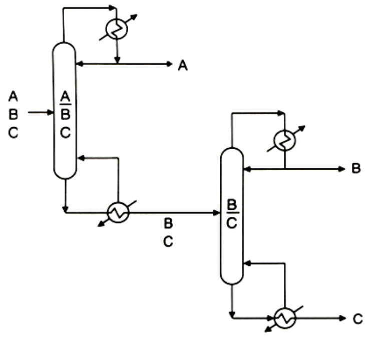


Figure 1b. The Direct Sequence

Progressive distillation is a process that has the potential to reduce this percentage and effectively save the refinery money. The idea of progressive distillation was patented in 1987, with the claim of decreasing the heat utility of a conventional distillation plant. At the time, because energy costs were relatively low and petrochemical product costs were very high, along with the fact that there were very few environmental regulations as compared to the modern day, the idea was not widely used or accepted. As both the cost of energy and the strictness of environmental regulations have increased, interest in the possibility of a lower-utility process has increased. The claim of the patent, and of the company that is currently putting it to use (mainly in Europe), Technip, is that utility costs can be reduced by about 34% for a heavy crude. Our research shows that progressive crude distillation can reduce the heat utility by at least 17% for a heavy crude. This expired patent has potential to dramatically reduce energy cost in refineries.

Progressive Distillation is an application of the direct sequence. The Direct Sequence (Figure 1b) follows an idea similar to that of indirect sequencing. However, the lightest component is separated first instead of the heaviest component. This method, when used alone, does not generally provide good results, as will be discussed later. But because the progressive distillation model discussed contains a theoretical use of the direct sequence, it will be discussed further.

In a direct sequence, the light component alone is taken off of the first column, with the rest of the mixture coming out of the bottom. The way progressive distillation modifies this sequence is that it takes off more than just the lightest component, instead a combination of the lightest component and some of the second lightest, and it then sends that product to a second column. Effectively, this splits the first column into two separate columns. In order to evaluate the claims of the patent, simulations must compare the heat utility of the patent sequence to the heat utility of a simple direct sequence of columns. Additionally, the heat utility of the patent sequence will be compared to heat utility from the conventional model.

Stripping-type distillation operates with the crude being heated to a low temperature and inserted into the top of the column. The crude is warmed by heaters as it travels down the column. The major difference between the conventional distillation and stripping-type distillation is the order in which the crude is heated. Conventional distillation heats all the crude up initially while stripping-type distillation heats the crude up in steps and separates the vapor immediately after it is produced. Stripping-type distillation is similar to progressive distillation in the sense that the lighter ends are separated fist, followed by the heavy ends. Stripping-type crude distillation was shown to have no energy efficiency benefit over conventional distillation when producing similar yields.[iii]

This paper is organized as follows: We first review the theory behind a design of a conventional distillation. We follow with a discussion on progressive distillation and we outline the different possible implementations. We discuss our simulations and our use of pinch analysis as well as the use of demand-supply diagrams. Finally the results of the simulations for each crude (light and heavy) are compared against the conventional design energy requirements.

2. Conventional Design

Figure 2 depicts the basic arrangement of conventional distillation. The crude feed stream enters near the bottom of the column, steam enters the bottom of the column, and draws are sent to side-strippers which use steam to produce better gaps. The three pump-arounds ensure adequate vapor and liquid flowrates and decrease the required heat duty on the condenser. Bagajewicz and Ji (2001)[ii] proved that they are detrimental to separation, that the sum of their load and the condenser is roughly constant and that they are important means to recover energy. Bagajewicz and Ji (2001)[ii] also showed when and how these ought to be activated by introducing demand-supply diagrams to govern their activation.

The basic column was designed in PRO/II using the specifications given later. The feed enters a heater, goes through a desalter, runs through another heater, and enters on tray 29 in a 34 tray column. Steam enters the bottom of the column. Three different product draws are taken off the side of the main column and sent to side-strippers which are fed with steam. The bottom products of the side-strippers are the kerosene, diesel, and gas oil product streams. The two other product streams are the residuals from the bottom of the main column and naphtha from the top of the main column. Also notice that each product stream has a heat exchanger attached. This is required to provide a consistent environment for calculating the hot and cold utilities.

The purpose of the side strippers is to sharpen or loosen the product cuts by adding steam. The steam further separates the product and sends the unwanted components back to the main column. The more steam added, the tighter the cut, however, this steam rate must be accounted for when calculating the overall utility. Also, if too much steam is added, water will collect on the stripper trays and eventually deter the separation process. The steam added to the bottom of the column serves basically the same function, except instead of affecting the gap of a particular product, it instead further separates the heavy components that make it to the bottom of the column.

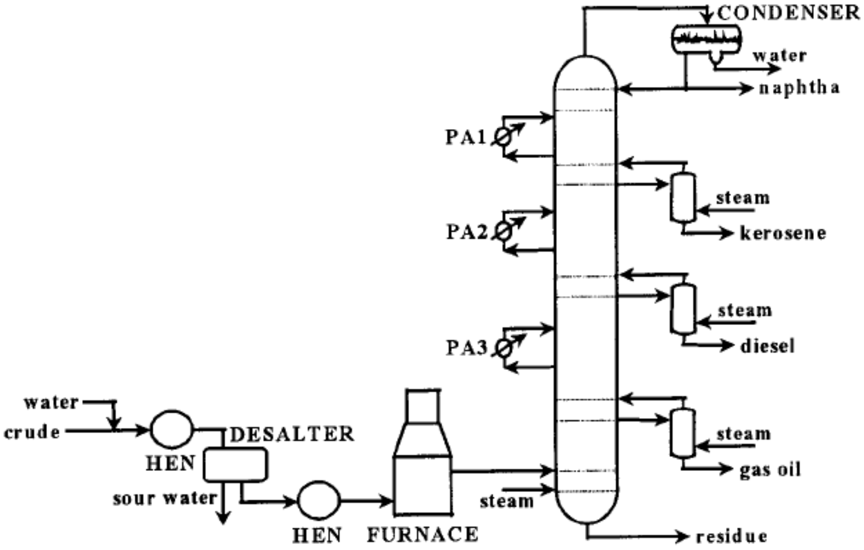


Figure 2. Conventional Distillation

The following is the final heat demand and supply diagram from the conventional case studied by Dr. Bagajewicz and Dr. Ji:

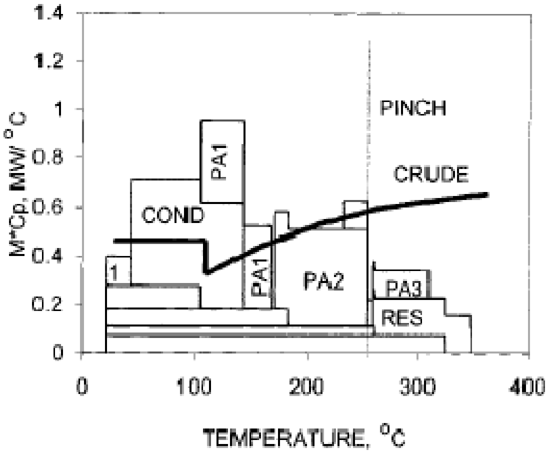


Figure 3. Heat Demand-Supply Diagram for a Light Crude Conventional Case with Three Pumparounds

3. Progressive Distillation

The major concept of progressive distillation is to loosely separate the light components off the mixture in a primary sequence of columns, and to fully separate the resulting mixture in a secondary sequence of columns. In this sense one can call it an extension of the direct sequence, or loosely speaking two direct sequences one feeding the other in a special arrangement.

The now expired US patent no. 4,664,785 states that, "The process consists in successively separating increasingly heavy petroleum cuts at the head of a plurality of columns in [the primary sequence] which feed individually each column of the [secondary sequence]… By carrying out a succession of progressive separations performed in a series of small volume, more efficient utilization of the recovery of heat is achieved. "

This idea is illustrated by figure 4. In the patent, there are seven eventual products, the top group corresponding to naphtha, the next down corresponding to Gasoline and Kerosene, followed by Diesel and Gasoil, and then the Vacuum distillates, and residue. In the complete idea put forward by the patent, there are additional columns including a vacuum column to further separate the reside, and a side stripper after C10. There is no need for a vacuum distillation tower, or for C06, C12, or C11 in figure 4, as the conventional model to be compared only has five eventual product draws, one each for Naphtha, Kerosene, Diesel, Gasoil, and Resid. Both the progressive and conventional models will have five product draws. We also included several other simplifications but were sure to still capture the idea of the patent. The bottom product of C07 was sent to a product stream instead of into the next column of the sequence. The bottom product of C13 is not sent back to C10, because in our simulations C13 is removed.

The technology was first introduced by a French company called Technip. The company claims that progressive distillation will work for all types of crude oil.

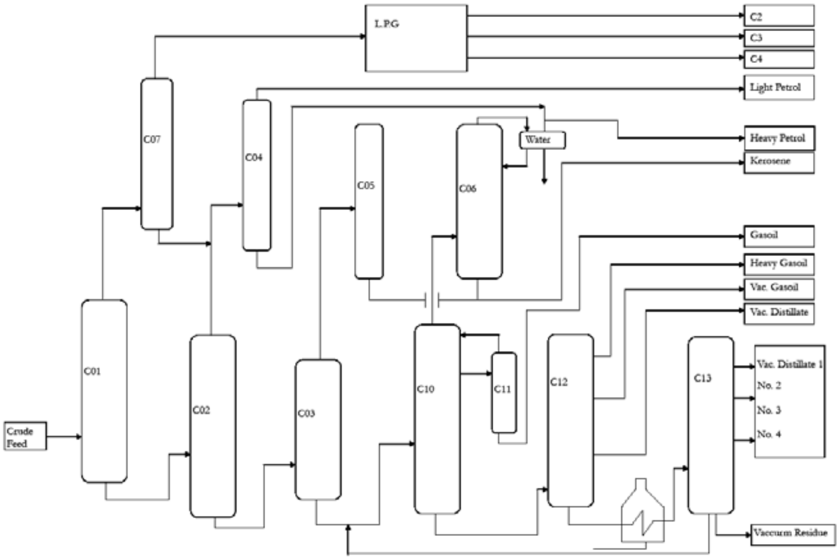


Figure 4. Progressive Model From Patent.

The heat utility benefit of the progressive crude fractionation method of sequential separations depends on the idea that less heat needs to be added to the initial feed stream because columns in the primary sequence produce loose separations. This reduction of heat can then be replaced with stripping steam in the second column to further separate the mixture. Also, the larger number of trays in the progressive fractionation model decreases the reflux ratio, resulting in lower condenser and reboiler heat duty requirements.

Ji and Bagajewicz (2002)[iii] studied a similar version of the direct sequence they called stripping type crude fractionation. This study was prompted by earlier claims by Liebmann and Smith (1995) that this design would be more beneficial. Some of the drawbacks of this design are that in a stripping type distillation column the crude has to be heated eventually to a higher temperature than the conventional case because the carrier effect is no longer present. The carrier effect is no longer present because the bottom of the column only contains the heavy components of the crude feed, and the light components are not present to assist in separation. When compared using the same maximum temperature anywhere in the system, this design proved to be less energy efficient for the same degree of separation.

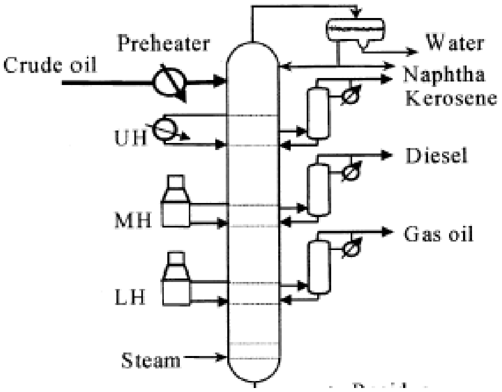


Figure 5. Stripping Type Distillation

To examine those claims in the present context we present a many columns version of the direct sequence. This arrangement will also diminish the carrier effect for the same reason as stated above in the single column model.

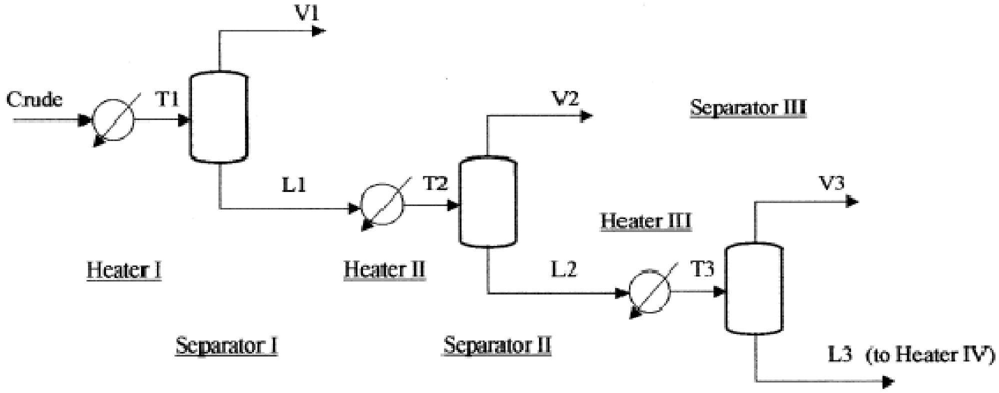


Figure 6. Direct Sequence Model

The temperature of the crude feed in a direct sequence distillation has to be higher than in conventional distillation because the heaters have to be placed lower in the columns, where the liquid composition is weighted more toward the heavy end of the petroleum array. Because the composition is heavier than in conventional distillation, the carrier effect is lessened considerably. The *carrier effect* is the ability of light components to aid in the vaporization of heavy components at lower temperatures through a process similar to adsorption. Steam alone is not enough to overcome this lack of the carrier effect in this configuration.

Because the feed of this type of column is at the top, as the heavy components fall down the column, they exit into the side columns, and have to be removed and recycled into the column, otherwise the flash points of the products will be inconsistent.

Vapor phase withdrawal is partially beneficial. Because the vapor phase removal would be at a higher temperature than that of a similar liquid phase removal, for heat integration, the heat duty of cooling the streams would be at a higher temperature, which can help lower the overall heat duty of the system. Unfortunately, vapor phase withdrawal can't be used without care because if there is any water in the side condensers, corrosion can become a problem.

To ameliorate these previously stated problems a second sequence of columns is introduced. These three columns, fed by the top products of the first the columns in the primary sequence, further separate the intentionally loose cuts of the primary sequence into the desired products. This allows less energy to be put into the first primary series of columns because the separation is not as thorough. This however leaves us with several choices about how to arrange the columns and what means to achieve separation in them. The three choices we have are as follows:

1. *Each column has a reboiler (All Reboiler Model)*
2. *Each column has steam input, but no reboiler (All Steam Model)*
3. *Some columns have steam, others have reboilers (Hybrid Model)*

Later in the results section we will discuss which configuration yielded the best results, and the reasoning behind these results.

In order to quantitatively compare conventional and progressive crude distillation, PRO/II simulations of progressive distillation were also developed. The PRO/II progressive distillation model was created in a number of steps. First the primary sequence was initialized and run in order to separate the components partially. The top products of the first three columns were fed to the secondary sequence of three columns for further separation. Then the product gaps were maintained by controllers set to vary the input steam rate. After these gaps were controlled, the top products of the columns, as well as feed tray locations and feed temperatures of the columns were adjusted until the desired D86 95%-Points and flow rates were achieved. Creating these simulations presented a certain amount of difficulty. Even though there are 13 degrees of freedom, the simulation would result in errors if the initial value of each variable was not within a certain window of values, all interdependent on each other. The simulation is shown in figure 7:

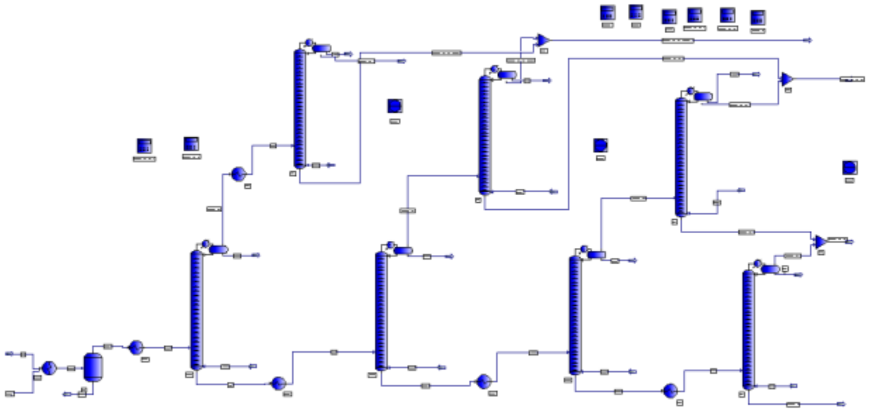


Figure 7. Pro/II Simulation of Progressive Distillation

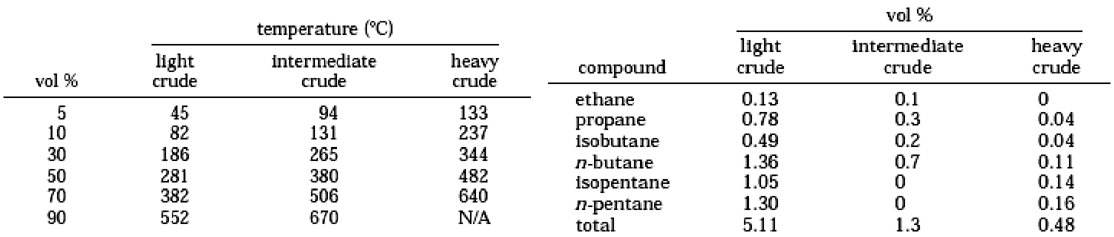
Once the product flow rates and gaps were set, a calculator was used to compute the minimum heat utility based on pinch calculations. Both a Pro/II calculator and an Excel spreadsheet were used to calculate heat utility to ensure the results were not erroneous.

Progressive distillation is considerably different than the previous research on the direct sequence method investigated by Dr. Bagajewicz and Dr. Ji. This is mostly because progressive distillation used neither a direct sequence nor an indirect sequence, but instead a combination of the two. The primary level of columns along the bottom of the simulation follow a direct sequence, but when the columns in the secondary level are added it is no longer a true direct sequence.

4. Crudes and Specifications

We used the TBP parameters for a light and heavy crude as shown in table 1.

Table 1. TBP Data (temperature units in °C)

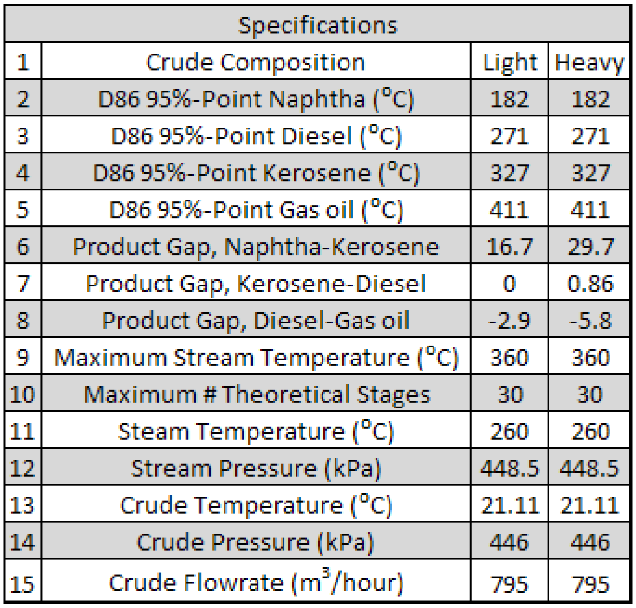


	temperature (°C)		
vol %	light crude	intermediate crude	heavy crude
5	45	94	133
10	82	131	237
30	186	265	344
50	281	380	482
70	382	506	640
90	552	670	N/A

	vol %		
compound	light crude	intermediate crude	heavy crude
ethane	0.13	0.1	0
propane	0.78	0.3	0.04
isobutane	0.49	0.2	0.04
n-butane	1.36	0.7	0.11
isopentane	1.05	0	0.14
n-pentane	1.30	0	0.16
total	5.11	1.3	0.48

D86 95% points, product gaps, and tray specifications are given in table 2 The D86 95% point of gas oil used in the PRO/II simulation was 411 degrees C. D86 95% points of naphtha, kerosene, diesel, and gas oil were specified in the column by varying three product draw flow rates and the condenser duty.

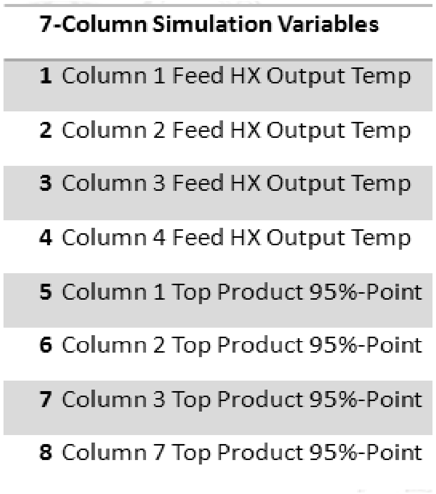
Table 2. Specifications



Specifications			
1	Crude Composition	Light	Heavy
2	D86 95%-Point Naphtha (°C)	182	182
3	D86 95%-Point Diesel (°C)	271	271
4	D86 95%-Point Kerosene (°C)	327	327
5	D86 95%-Point Gas oil (°C)	411	411
6	Product Gap, Naphtha-Kerosene	16.7	29.7
7	Product Gap, Kerosene-Diesel	0	0.86
8	Product Gap, Diesel-Gas oil	-2.9	-5.8
9	Maximum Stream Temperature (°C)	360	360
10	Maximum # Theoretical Stages	30	30
11	Steam Temperature (°C)	260	260
12	Stream Pressure (kPa)	448.5	448.5
13	Crude Temperature (°C)	21.11	21.11
14	Crude Pressure (kPa)	446	446
15	Crude Flowrate (m^3/hour)	795	795

Table 3 summarizes what variables need to be adjusted to arrive at the specifications in both cases.

Table 3. Variables



7-Column Simulation Variables	
1	Column 1 Feed HX Output Temp
2	Column 2 Feed HX Output Temp
3	Column 3 Feed HX Output Temp
4	Column 4 Feed HX Output Temp
5	Column 1 Top Product 95%-Point
6	Column 2 Top Product 95%-Point
7	Column 3 Top Product 95%-Point
8	Column 7 Top Product 95%-Point

This is a very small number of variables compared to the conventional case, where the controlling factors are the input crude feed temperature and steam flow rates to the side strippers. There is also a small amount of variation to the location of the side strippers in reference to the main column.

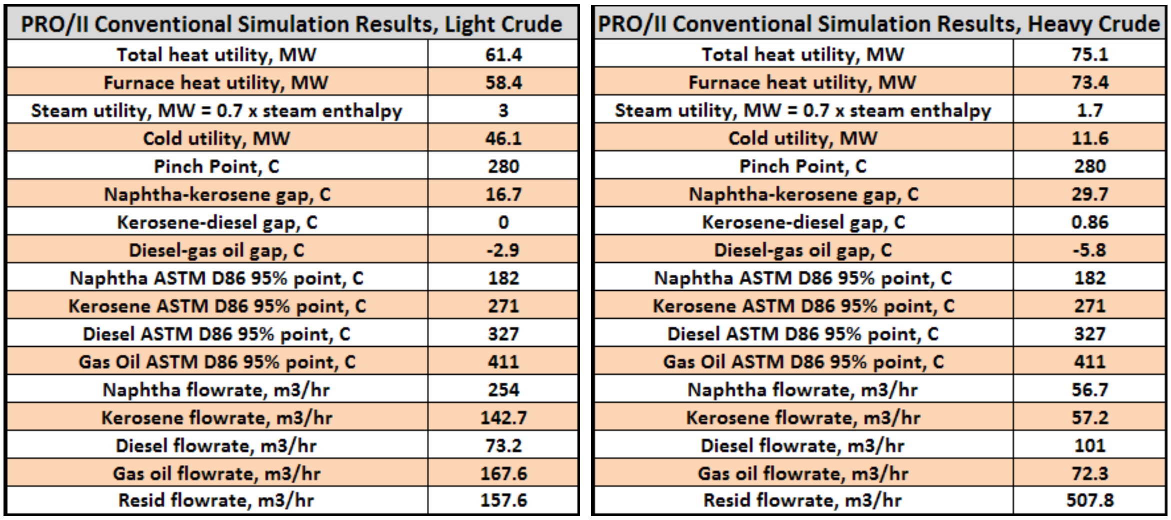
The result of having this large number of independent variables, and a relatively small amount of specifications is that there are a large number of degrees of freedom. This freedom allows the model to take a large number of configurations with respect to the specifications, and a solution has to be found within these limits.

The three different column configurations are controlled in slightly different ways and it is important to note that while the independent variables remain the same for all three cases, they each meet the specifications set forth by altering a different subset of properties. The *All Reboiler Model* reaches the desired product specifications by changing the heat duty of each reboiler in each column. The *All Steam Model* reaches the product specifications by changing the steam flow rates to each of the 7 columns. The *Hybrid Model* adjusts the reboiler duties in those columns that contain reboilers and the steam flow rates in those columns that use steam.

5. Results

Although the conventional design was already run by Ji and Bagajewicz(2001)[ii], we repeated the simulations and obtained the results of Table 4 for a light crude and heavy feed of 120,000 BPD. These match the specs.

Table 4. Conventional Simulation Results



PRO/II Conventional Simulation Results, Light Crude	
Total heat utility, MW	61.4
Furnace heat utility, MW	58.4
Steam utility, MW = 0.7 x steam enthalpy	3
Cold utility, MW	46.1
Pinch Point, C	280
Naphtha-kerosene gap, C	16.7
Kerosene-diesel gap, C	0
Diesel-gas oil gap, C	-2.9
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	411
Naphtha flowrate, m3/hr	254
Kerosene flowrate, m3/hr	142.7
Diesel flowrate, m3/hr	73.2
Gas oil flowrate, m3/hr	167.6
Resid flowrate, m3/hr	157.6

PRO/II Conventional Simulation Results, Heavy Crude	
Total heat utility, MW	75.1
Furnace heat utility, MW	73.4
Steam utility, MW = 0.7 x steam enthalpy	1.7
Cold utility, MW	11.6
Pinch Point, C	280
Naphtha-kerosene gap, C	29.7
Kerosene-diesel gap, C	0.86
Diesel-gas oil gap, C	-5.8
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	411
Naphtha flowrate, m3/hr	56.7
Kerosene flowrate, m3/hr	57.2
Diesel flowrate, m3/hr	101
Gas oil flowrate, m3/hr	72.3
Resid flowrate, m3/hr	507.8

There are a few different ways the progressive columns can be arranged, as stated before. This caused the initial results to be very negative, so most of these results were the reason for developing new configurations.

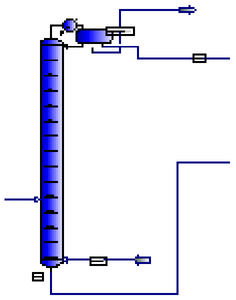


Figure 8a. With Steam Input

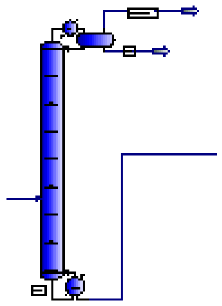


Figure 9b. With a Reboiler

Columns can be fitted with either steam input (Figure 9a) or reboilers (Figure 9b) in the last tray to provide an impetus for separation. In our initial simulation, all of the columns were outfitted with reboilers, which caused the overall heat utility to be very high. This simulation is denoted the *All Reboilers Model.* A heat demand-supply diagram was not created for the all reboiler model as the furnace heat utility was greater than 200 MW, and we decided that this was too much of a deficit to attempt to overcome by simply optimizing the given system.

The next step was to replace the reboilers with steam input in order to try to lower the furnace utility. In the case where all seven columns were fed by steam, the *All Steam Model*, the furnace heat utility was reduced significantly; however, the overall utility was very high as well because of the high steam flow rates. Figure 10 is a heat supply and demand diagram for the all steam model.

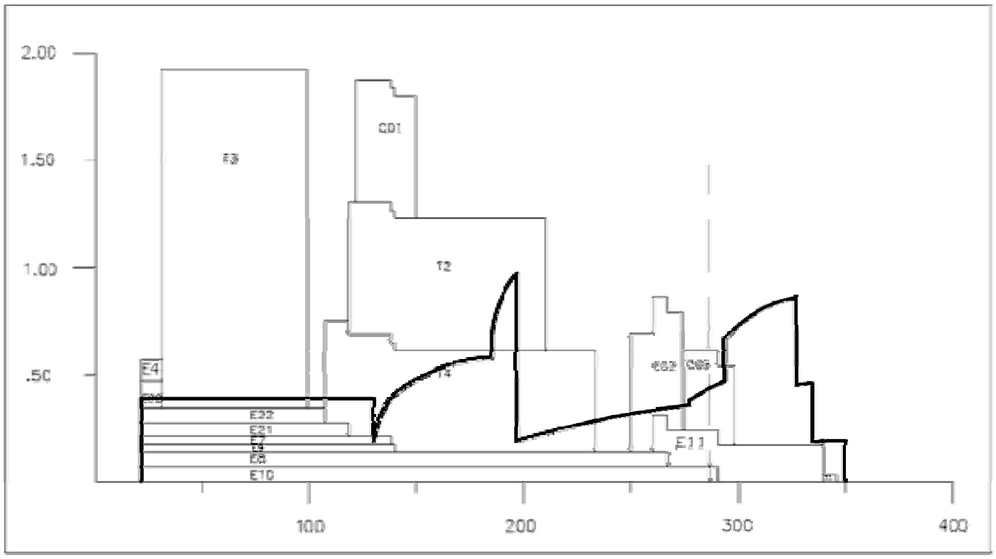


Figure 9. Heat Demand-Supply Diagram, All Steam Model, Light Crude

The results from the A*ll Steam Model* are given in table 5:

Table 5. All Steam Model Results

Light crude feed	**Conventional**	**Progressive**
Total heat utility, MW	**61.5**	**96.7**
Furnace heat utility, MW	**58.5**	**35.5**
Steam utility, MW = 0.7 x steam enthalpy	**3**	**61.2**
Cold utility, MW	**47.2**	**144**
Pinch Point, C	**280**	**306**

From figure 10 it should be noted that there is an excess of heat supply in the low temperature regions that was essentially getting wasted in the configuration as it was. This led to the final step which was to attempt to use a combination of reboilers and steam to overcome this supply/demand problem.

The following simulation, *the Hybrid Model*, was created so that reboilers were added to the 1st, 2nd, and 3rd columns in the primary sequence. This is shown in Figure 11.

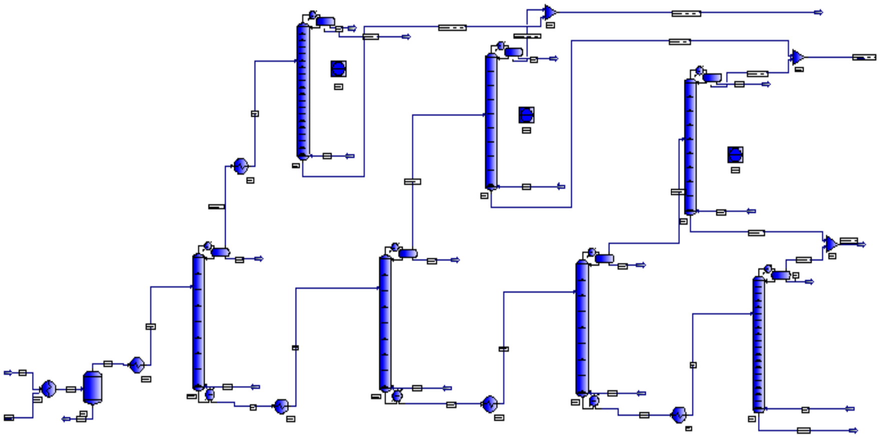


Figure 10. Final Simulation

First as much of the steam load as possible was shifted to the first series of columns, so that when this steam was replaced by reboilers, it would have a significant effect. The first reboiler added was to the 3rd column in the primary sequence. This showed a decrease in overall heat utility, so a second reboiler was added to the 2nd column in the primary sequence. This also caused a reduction in overall heat utility. Again the trend was followed and a reboiler was added to the 1st column in the primary sequence. This caused a small reduction in overall heat utility.

The next logical step was to add a reboiler to the 4th column in the primary sequence and/or the 1st column in the secondary sequence. As the law of diminishing returns would suggest, both of these resulted in an increase in overall heat utility. Because there was already determined a configuration that ended up with a reduction in overall heat utility compared to the conventional case, no more optimization was attempted in the heat utility, instead the flow rates of the products were improved upon to better the economic impact of progressive distillation.

One important item to note is the heat supply-demand diagram for our hybrid model, excess supply can be moved from right to left and from top to bottom, but not in the opposite direction (a direct result of the 2nd law of thermodynamics) and so from figure 12 it is easy to see a reduction in overall utility.

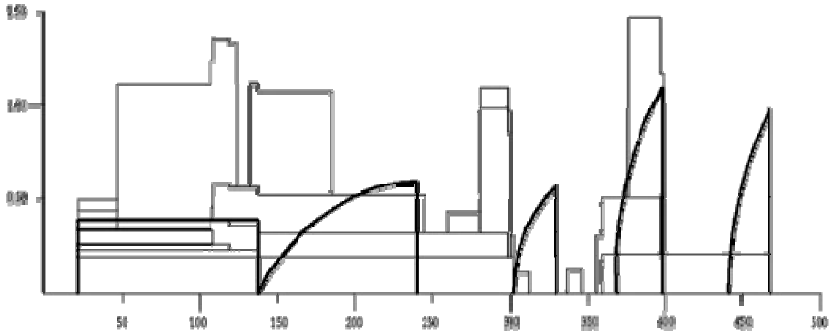


Figure 11. Heat Demand-Supply Diagram, Hybrid Model, Light Crude

Table 6 shows results from the PRO/II progressive simulation using a light crude feed of 120,000 BPD. As with the conventional simulation, notice that the D86 95% points and the product gaps match those in the specifications.

Table 6. Progressive Simulation Results

PRO/II Progressive Simulation Results, Light Crude	
Total heat utility, MW	61.5
Furnace heat utility, MW	49.8
Steam utility, MW = 0.7 x steam enthalpy	11.6
Cold utility, MW	103.5
Pinch Point, C	302
Naphtha-kerosene gap, C	16.7
Kerosene-diesel gap, C	0
Diesel-gas oil gap, C	12.1
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	412
Naphtha flowrate, m3/hr	250.4
Kerosene flowrate, m3/hr	144.5
Diesel flowrate, m3/hr	78.4
Gas oil flowrate, m3/hr	159.9
Resid flowrate, m3/hr	162

PRO/II Progressive Simulation Results, Heavy Crude	
Total heat utility, MW	68.4
Furnace heat utility, MW	63.2
Steam utility, MW = 0.7 x steam enthalpy	5.2
Cold utility, MW	37.6
Pinch Point, C	399
Naphtha-kerosene gap, C	29.7
Kerosene-diesel gap, C	0.86
Diesel-gas oil gap, C	-5.8
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	411
Naphtha flowrate, m3/hr	57.1
Kerosene flowrate, m3/hr	42.6
Diesel flowrate, m3/hr	116.9
Gas oil flowrate, m3/hr	79.6
Resid flowrate, m3/hr	498.8

These results show that furnace heat utility in the progressive simulation using light crude is reduced by 9% when compared with the conventional simulation.

Heavy crude progressive simulation results indicate a 9% decrease in overall heat utility and a 14% decrease in furnace heat utility compared with the similar conventional model results. An overview of the significant results and how they compare in each case is shown in table 7.

Table 7. Heat Utility Comparison of Conventional to Progressive

	Light Crude		**Heavy Crude**	
	Conventional	Progressive	Conventional	Progressive
Overall Utility	61.4	61.5	75.1	68.4
Furnace Utility	58.4	49.8	73.4	63.2
Steam Utility	3	11.6	1.7	5.2

It is important to note that the final product flow rates are changed as a result of using progressive distillation compared to the conventional case. The product rates for each model are given in table 8. This difference in product rates is very important in calculating any potential benefit to using progressive distillation.

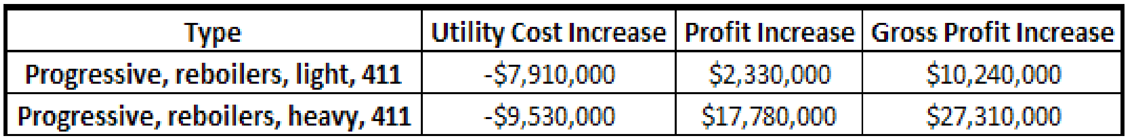
Table 8. Product Rate Comparison of Conventional to Progressive

	Light Crude		Heavy Crude	
	Conventional	Progressive	Conventional	Progressive
Naphtha Rate (m^3/hr)	254	250.4	56.7	57.1
Kerosene Rate (m^3/hr)	142.7	144.5	57.2	42.6
Diesel Rate (m^3/hr)	73.2	78.4	101	116.9
Gasoil Rate (m^3/hr)	167.6	159.9	72.3	79.6
Residual Rate (m^3/hr)	157.6	162	507.8	498.8

Potential Economic Benefit

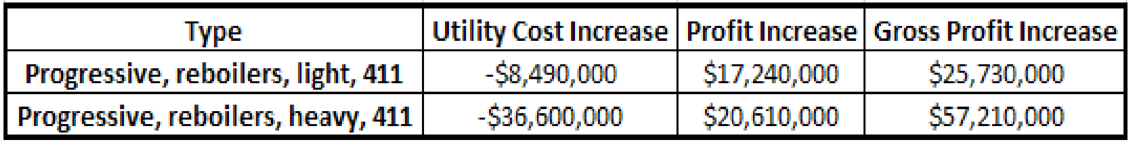
After the overall heat utility was increased while maintaining the same product gaps and D86 95%-points, it was important to analyze what kind of economic impact this could have on a refinery. Using current prices of hydrocarbon products, utility costs, cooling water, and steam generation, an analysis was done on the product sales profit change and the utility cost change in order to determine an overall profit change. In the simulations run, which were based off a crude oil flow rate of 795 m^3/hr, it can be shown that progressive distillation could potentially save money during the refining process. The exact economic benefit is generalized in the following two charts, the first of which is on a refinery which does not include a vacuum distillation unit for the residual products, and the second for a refinery that does employ a vacuum column. For the vacuum column analysis, it was assumed that the gasoil and residue flow rates for conventional and progressive would be identical, where the added benefit came from reducing the overall volume of residual product that has to be reheated before the vacuum column:

Table 9a. Economic Results without a Vacuum Unit



Type	Utility Cost Increase	Profit Increase	Gross Profit Increase
Progressive, reboilers, light, 411	-$7,910,000	$2,330,000	$10,240,000
Progressive, reboilers, heavy, 411	-$9,530,000	$17,780,000	$27,310,000

Table 9b. Economic Results with a Vacuum Unit



Type	Utility Cost Increase	Profit Increase	Gross Profit Increase
Progressive, reboilers, light, 411	-$8,490,000	$17,240,000	$25,730,000
Progressive, reboilers, heavy, 411	-$36,600,000	$20,610,000	$57,210,000

Of course the amount of money saved is all based on the capital investment of the plant itself, and because this changes significantly with each type of crude, length of operation, and each location, no capital cost would be very useful. However, our results do show that an investigation into the capital investment and the overall gross profit change would be a worthwhile exercise.

Accuracy

The Simulations created were consistently in an acceptable range of the specifications given. This is because most of the simulations had these specifications as an important part of the programming instructions that the simulations were based off of. Most notably, the D86 95%-points are each within 1° Celsius of each other, and the Gaps (with the exception of the unspecified gas oil-residue gap) are within 1° Celsius of each other. Because of these specifications matching up so precisely, a useful heat utility comparison can be drawn between conventional and progressive distillation.

It is also important to note that the simulations created may not be optimized completely. It is absolutely possible that an even lower heat utility may be achieved through further tweaking the system. The main goal of the work was to find out if progressive distillation could reduce the heat utility of the process when compared to conventional distillation, and it has been shown that it can.

6. Conclusion

Based on PRO/II simulations using a light crude feed, distillation of petroleum by progressive separations can dramatically reduce the heat utility necessary for distillation when compared with conventional distillation. Specifically, a hot utility savings of 9% was calculated. The main concept of progressive distillation is that loose separations of the lightest components of a crude feed require less energy input than sharp separations of heavier components.

The patent proposed that by cutting certain columns in half or by stacking certain columns together, the separation may require different hot utility inputs. The basic idea of progressive crude distillation may be analyzed and applied to other column sequences. As an industrial application, progressive distillation may reduce overall utility requirements in a heavy crude refinery by 9%, while producing more valuable products, and it may reduce furnace heat utility in a light crude refinery by 16% while producing more valuable products.

7. References

[i] Devos et al. *United States Patent No. 4,664,785*. May 12, 2987.

[ii] Bagajewicz, Miguel and Ji, Shuncheng. "Rigorous Procedure for the Design of Conventional Atmospheric Crude Fractionation Units. Part I: Targeting." *Ind. Eng. Chem. Res.* **2001,** *40*, 617-626.

[iii]Ji, Shuncheng, Bagajewicz, Miguel. "On the Energy Efficiency of Stripping-Type Crude Distillation." *Ind. Eng. Chem. Res.* **2002,** *41*, 5819-5825.

[iv] Miller, W.; Osborne, H. G. History and Development of Some Important Phases of Petroleum Refining in the United States. In *The Science of Petroleum*; Oxford University Press: London, 1938; Vol. 2.

[v] Watkins, R. N. *Petroleum Refinery Distillation*; Gulf Publishing Company: Houston, TX, 1979.

[vi] Brugma, A. J., The Brugma Process. *Refin. Nat. Gasoline Manuf.* 1941, 20 (9), 86.

[vii] Bagajewicz M. and S. Ji. Rigorous Targeting Procedure for the Design of Crude Fractionation Units with Pre-Flashing or Pre-Fractionation. *Ind. Eng. Chem. Res. 2002*, 41, 12, 3003-3011.

FIGURE CAPTIONS

Figure 12a & 1b: The Indirect Sequence/The Direct Sequence

Figure 2: Conventional Distillation.

Figure 3: Heat Demand-Supply Diagram for a Light Crude Conventional Case with Three Pumparounds.

Figure 4: Progressive Model From Patent.

Figure 5: Stripping Type Distillation.

Figure 6: Direct Sequence Model.

Figure 7: Pro/II Simulation of Progressive Distillation.

Figure 9a & 9b: With Steam Input/With a Reboiler

Figure 10: Heat Demand-Supply Diagram, All Steam Model, Light Crude.

Figure 11: Final Simulation.

Figure 12: Heat Demand-Supply Diagram, Hybrid Model, Light Crude.

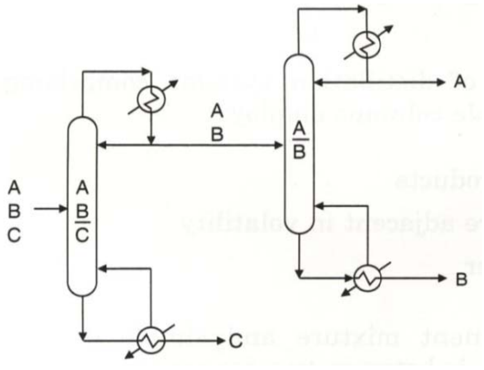


Figure 13a. The Indirect Sequence

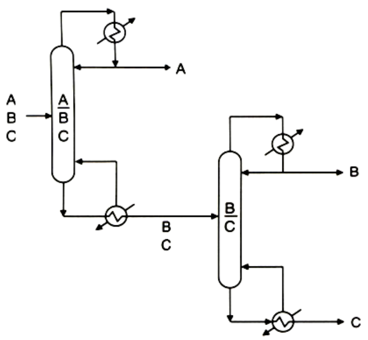


Figure 1b. The Direct Sequence

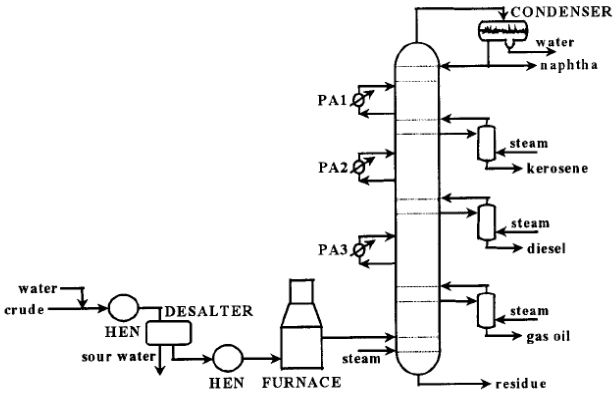


Figure 14. Conventional Distillation

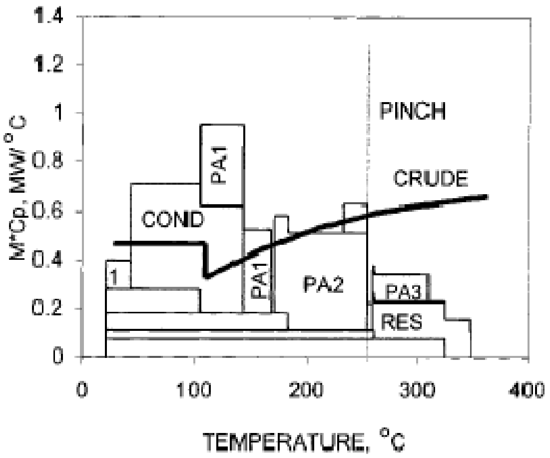


Figure 15. Heat Demand-Supply Diagram for a Light Crude Conventional Case with Three Pumparounds

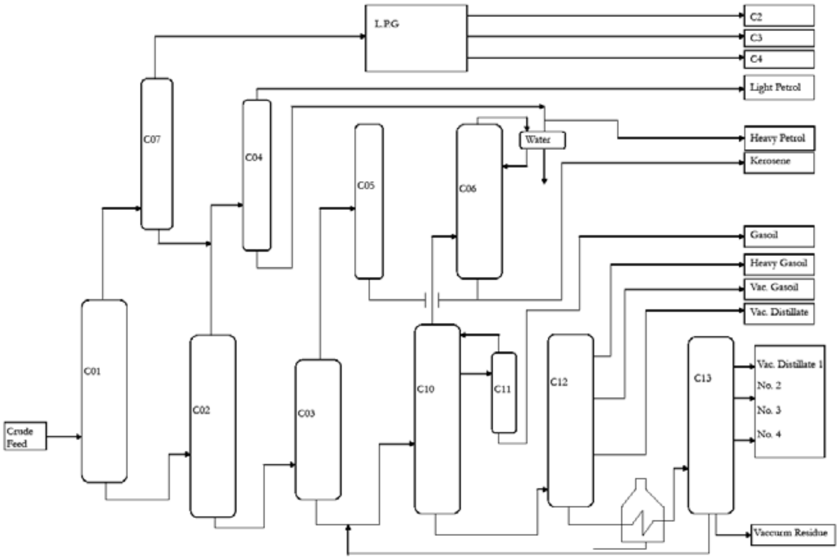


Figure 16. Progressive Model From Patent.

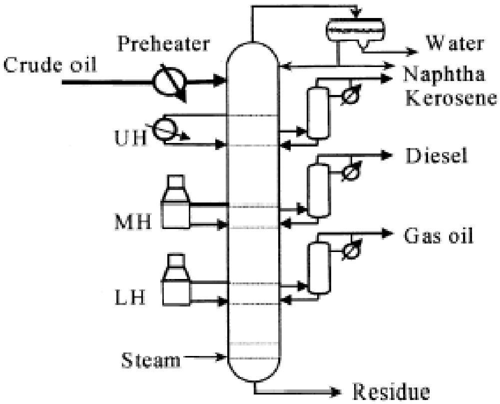


Figure 17. Stripping Type Distillation

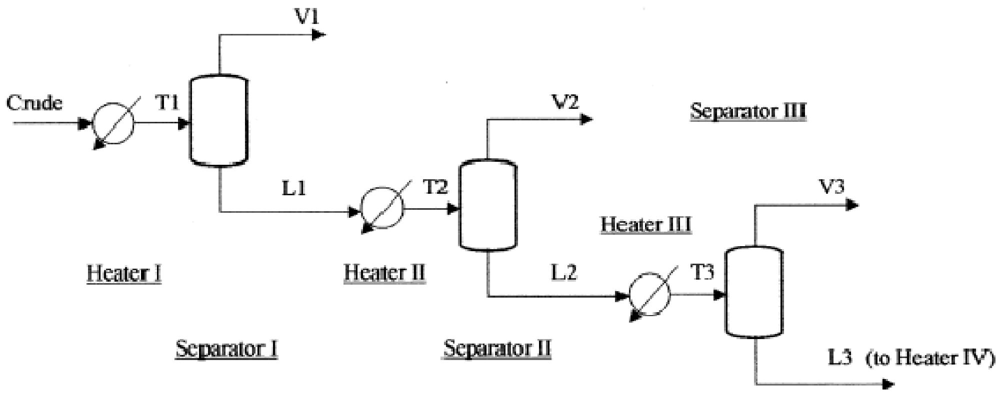


Figure 18. Direct Sequence Model

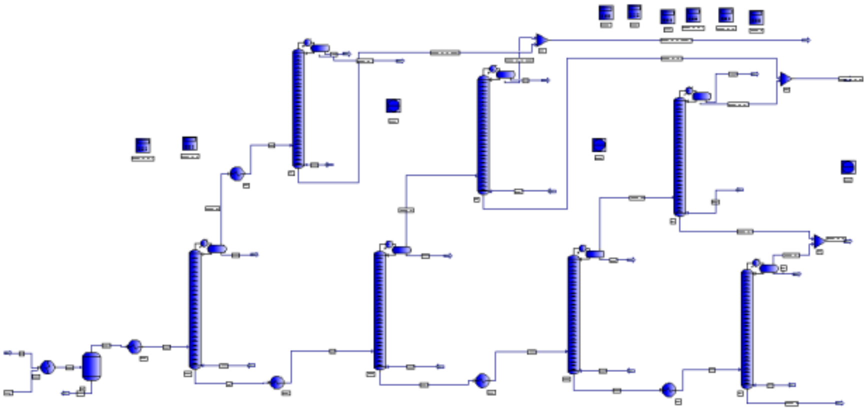


Figure 19. Pro/II Simulation of Progressive Distillation

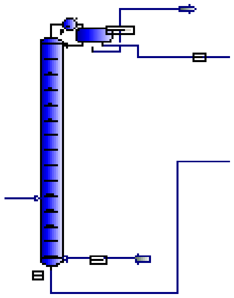


Figure 20a. With Steam Input

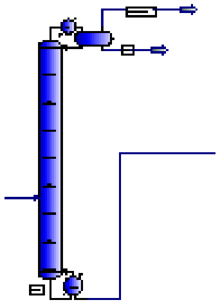


Figure 9b. With a Reboiler

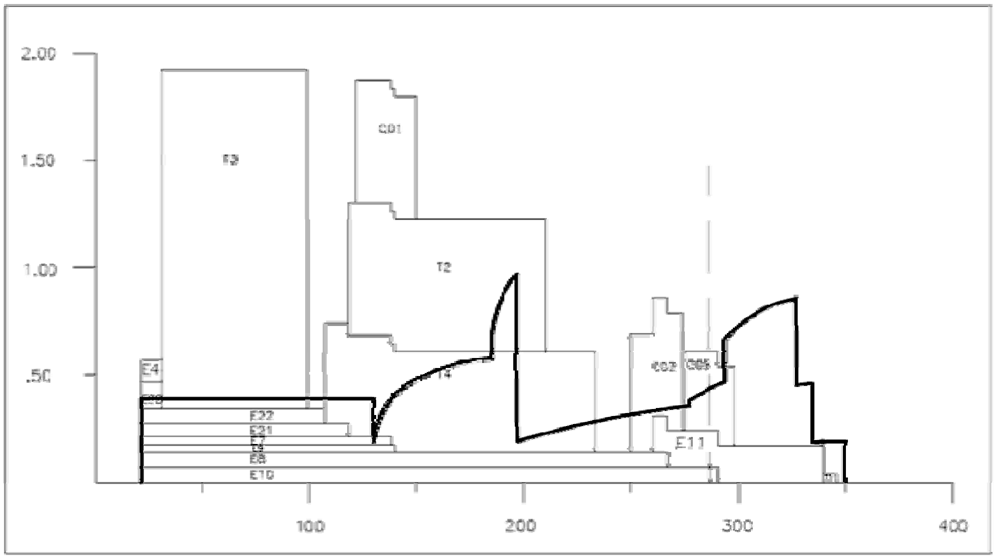


Figure 21. Heat Demand-Supply Diagram, All Steam Model, Light Crude

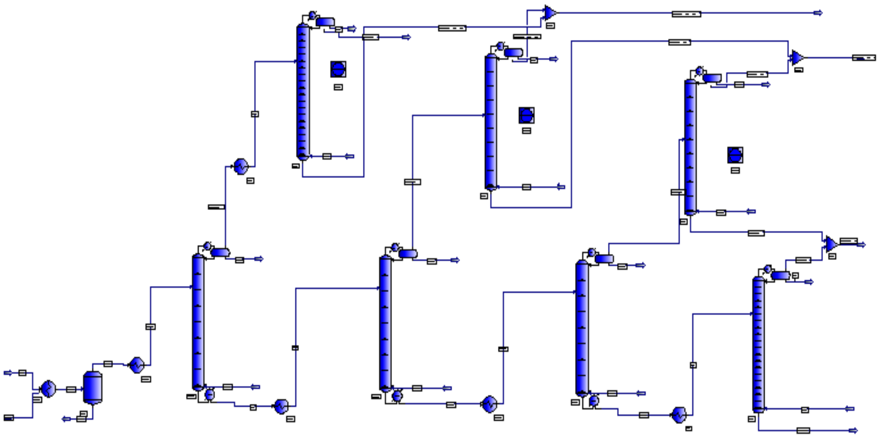


Figure 22. Final Simulation

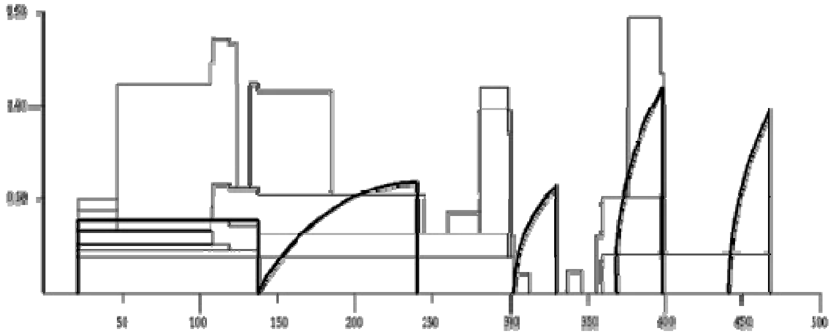


Figure 23. Heat Demand-Supply Diagram, Hybrid Model, Light Crude

TABLE CAPTIONS

Table 1: TBP Data (temperature units in °C).

Table 2: Specifications.

Table 3: Variables.

Table 4: Conventional Simulation Results.

Table 5: All Steam Model Results.

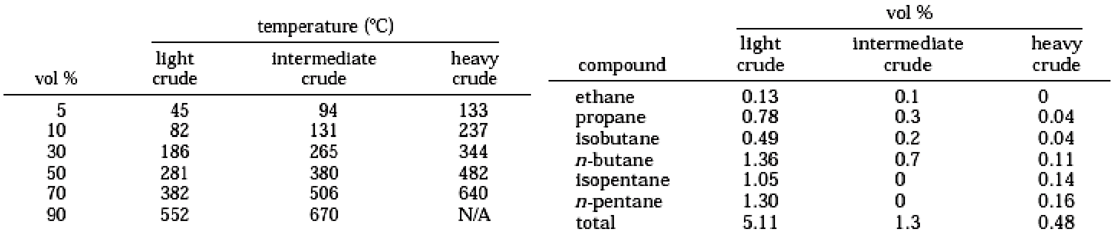
Table 6: Progressive Simulation Results.

Table 7: Heat Utility Comparison of Conventional to Progressive.

Table 8: Product Rate Comparison of Conventional to Progressive.

Table 9a & 9b: Economic Results without a Vacuum Unit/ Economic Results with a Vacuum Unit.

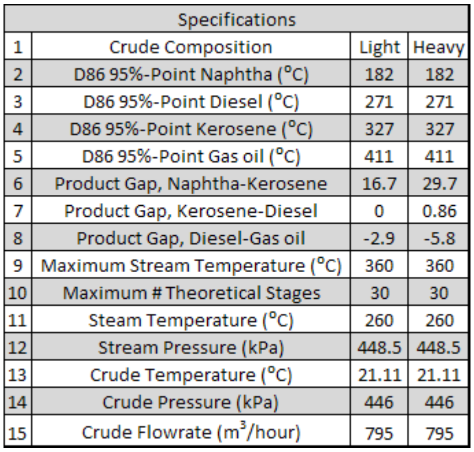
Table 1. TBP Data (temperature units in °C)



vol %	temperature (°C) light crude	intermediate crude	heavy crude
5	45	94	133
10	82	131	237
30	186	265	344
50	281	380	482
70	382	506	640
90	552	670	N/A

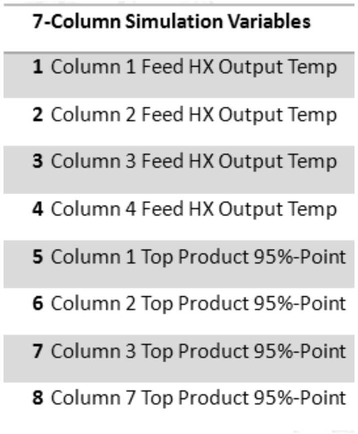
compound	vol % light crude	intermediate crude	heavy crude
ethane	0.13	0.1	0
propane	0.78	0.3	0.04
isobutane	0.49	0.2	0.04
n-butane	1.36	0.7	0.11
isopentane	1.05	0	0.14
n-pentane	1.30	0	0.16
total	5.11	1.3	0.48

Table 9. Specifications



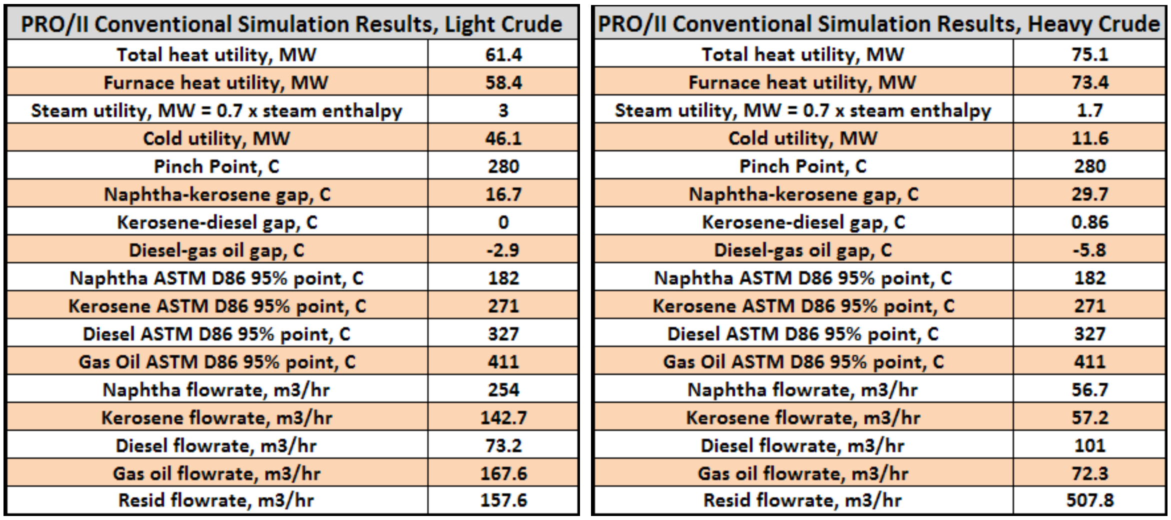
Specifications			
1	Crude Composition	Light	Heavy
2	D86 95%-Point Naphtha (°C)	182	182
3	D86 95%-Point Diesel (°C)	271	271
4	D86 95%-Point Kerosene (°C)	327	327
5	D86 95%-Point Gas oil (°C)	411	411
6	Product Gap, Naphtha-Kerosene	16.7	29.7
7	Product Gap, Kerosene-Diesel	0	0.86
8	Product Gap, Diesel-Gas oil	-2.9	-5.8
9	Maximum Stream Temperature (°C)	360	360
10	Maximum # Theoretical Stages	30	30
11	Steam Temperature (°C)	260	260
12	Stream Pressure (kPa)	448.5	448.5
13	Crude Temperature (°C)	21.11	21.11
14	Crude Pressure (kPa)	446	446
15	Crude Flowrate (m^3/hour)	795	795

Table 3. Variables



7-Column Simulation Variables	
1	Column 1 Feed HX Output Temp
2	Column 2 Feed HX Output Temp
3	Column 3 Feed HX Output Temp
4	Column 4 Feed HX Output Temp
5	Column 1 Top Product 95%-Point
6	Column 2 Top Product 95%-Point
7	Column 3 Top Product 95%-Point
8	Column 7 Top Product 95%-Point

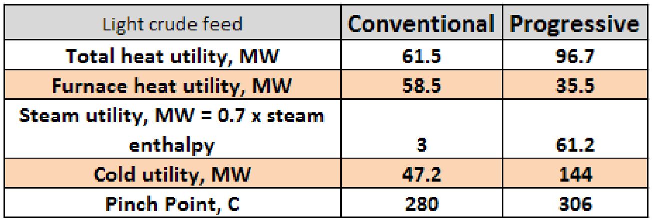
Table 4. Conventional Simulation Results



PRO/II Conventional Simulation Results, Light Crude	
Total heat utility, MW	61.4
Furnace heat utility, MW	58.4
Steam utility, MW = 0.7 x steam enthalpy	3
Cold utility, MW	46.1
Pinch Point, C	280
Naphtha-kerosene gap, C	16.7
Kerosene-diesel gap, C	0
Diesel-gas oil gap, C	-2.9
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	411
Naphtha flowrate, m3/hr	254
Kerosene flowrate, m3/hr	142.7
Diesel flowrate, m3/hr	73.2
Gas oil flowrate, m3/hr	167.6
Resid flowrate, m3/hr	157.6

PRO/II Conventional Simulation Results, Heavy Crude	
Total heat utility, MW	75.1
Furnace heat utility, MW	73.4
Steam utility, MW = 0.7 x steam enthalpy	1.7
Cold utility, MW	11.6
Pinch Point, C	280
Naphtha-kerosene gap, C	29.7
Kerosene-diesel gap, C	0.86
Diesel-gas oil gap, C	-5.8
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	411
Naphtha flowrate, m3/hr	56.7
Kerosene flowrate, m3/hr	57.2
Diesel flowrate, m3/hr	101
Gas oil flowrate, m3/hr	72.3
Resid flowrate, m3/hr	507.8

Table 10. All Steam Model Results



Light crude feed	Conventional	Progressive
Total heat utility, MW	**61.5**	**96.7**
Furnace heat utility, MW	**58.5**	**35.5**
Steam utility, MW = 0.7 x steam enthalpy	**3**	**61.2**
Cold utility, MW	**47.2**	**144**
Pinch Point, C	**280**	**306**

Table 11. Progressive Simulation Results

PRO/II Progressive Simulation Results, Light Crude	
Total heat utility, MW	61.5
Furnace heat utility, MW	49.8
Steam utility, MW = 0.7 x steam enthalpy	11.6
Cold utility, MW	103.5
Pinch Point, C	302
Naphtha-kerosene gap, C	16.7
Kerosene-diesel gap, C	0
Diesel-gas oil gap, C	12.1
Naphtha ASTM D86 95% point, C	182
Kerosene ASTM D86 95% point, C	271
Diesel ASTM D86 95% point, C	327
Gas Oil ASTM D86 95% point, C	412
Naphtha flowrate, m3/hr	250.4
Kerosene flowrate, m3/hr	144.5
Diesel flowrate, m3/hr	78.4
Gas oil flowrate, m3/hr	159.9
Resid flowrate, m3/hr	162

PRO/II Progressive Simulation Results, Heavy Crude	
Total heat utility, MW	68.4
Furnace heat utility, MW	63.2
Steam utility, MW = 0.7 x steam enthalpy	5.2
Cold utility, MW	37.6
Pinch Point, C	399
Naphtha-kerosene gap, C	29.7
Kerosene-diesel gap, C	0.86
Diesel-gas oil gap, C	-5.8
Naphtha ASTM D86 95% point, C	182
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Table 12. Heat Utility Comparison of Conventional to Progressive

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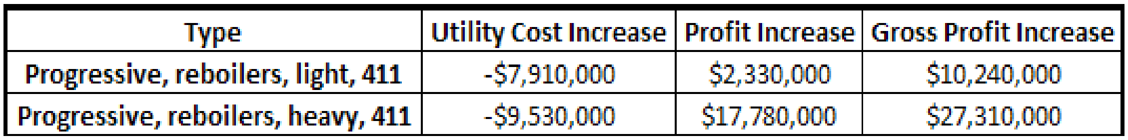
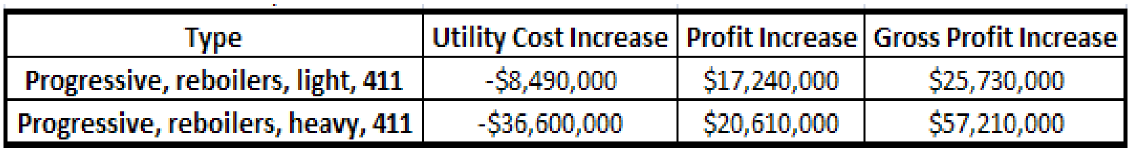
Table 9a. Economic Results without a Vacuum Unit



Type	Utility Cost Increase	Profit Increase	Gross Profit Increase
Progressive, reboilers, light, 411	-$7,910,000	$2,330,000	$10,240,000
Progressive, reboilers, heavy, 411	-$9,530,000	$17,780,000	$27,310,000

Table 9b. Economic Results with a Vacuum Unit



Type	Utility Cost Increase	Profit Increase	Gross Profit Increase
Progressive, reboilers, light, 411	-$8,490,000	$17,240,000	$25,730,000
Progressive, reboilers, heavy, 411	-$36,600,000	$20,610,000	$57,210,000