Process Simulation and Optimization of Cryogenic Operations Using Multi-Stream Brazed Aluminum Exchangers

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ABSTRACT

Since complex brazed aluminum plate exchangers can include as many as 8-10 process streams within a single exchanger, the process design calculations become quite involved and must include incremental duty versus temperature calculations to check for temperature pinch points. The capability to perform the process design for complex plate exchangers has recently been added to a process simulation program called PROSIM. Three case studies involving the simulation and optimization of cryogenic operations for liquids recovery from gases using complex plate exchangers are presented. Due to their low cost and superior heat transfer capabilities, complex plate exchangers can be very useful in reducing costs and optimizing cryogenic operations.

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INTRODUCTION

With the present competitive product markets in the gas processing industry, process optimization to yield the lowest possible capital and operating costs is an absolute necessity. To achieve this level of optimization, detailed process simulation exploring all possibilities is required. In cryogenic operations, some of the major factors to be considered are the heating, cooling, and refrigeration loads. These loads are usually interconnected and are optimized by the use of a network of cross exchangers. The most successful designs are using increasing numbers of cross exchangers with progressively closer approach temperatures to achieve heating and cooling without causing demands on plant utilities. Unfortunately, in many cases the heating and cooling potential is not fully realized because the heat delivery versus temperature curve causes impossible temperature crosses in conventional cross exchangers.

In many cases, complex plate exchangers are being considered to almost fully realize the heat exchange potential within some cryogenic operations. One application has been implemented by MacKenzie and Donnelly[1] for liquids recovery from natural gas using a mixed refrigerant system. As depicted in Figure 1, these exchangers are composed of a series of aluminum plates stacked like a deck of cards and brazed at the edges. They are usually supported at the top and allowed to hang free to accommodate thermal expansion. The fluid flow is vertical with the hot end usually at the top and the cold end at the bottom. Due to their relatively simple construction, the cost is in the range of 8-16 \$/ft of exchange area compared to about 20-30 \$/ft for carbon steel and 100-300 \$/ft for stainless steel (cryogenic) in conventional shell and tube exchangers. The overall heat transfer coefficient in Btu/hr ft^2 ^{o}F for brazed aluminum exchangers is about 150 for gas/gas and 200 for gas/phase change compared to about 60-75 for gas/gas and 80-100 for gas/phase change in shell and tube exchangers. In some units, a given process stream may undergo over 200^{o}F of temperature change within a single exchanger.

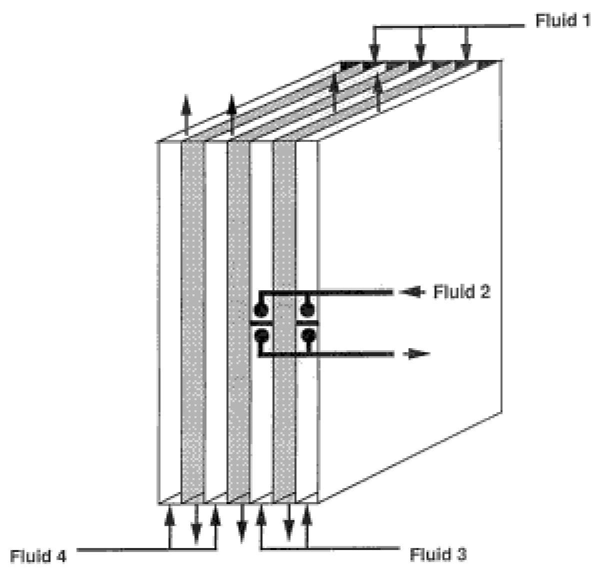

Figure 1. Schematic of complex plate exchanger.

These complex exchangers can be constructed to allow heat exchange to occur simultaneously between as many as 8 or 10 process streams. Each stream physically enters at the point in the exchanger corresponding as closely as possible to the temperature of the incoming stream. Similarly, each stream physically leaves the exchanger at the point corresponding to the desired temperature of the exit stream. The full potential latent and sensible heat may be extracted from or delivered to a series of fluids whose duty versus temperature curves form an overlapping set which prevents simple shell and tube exchangers from being used in series configurations. In effect, this single unit can act like a complex network of simple exchangers in series and parallel.

The overall process design methodology to take advantage of these complex exchangers in cryogenic operations is presented in this work. Three case studies for liquids recovery from gas streams are presented. The first case involves a simple refrigeration system for liquids recovery while the second involves ethane recovery using a mixed refrigerant system and the third involves a turbo expander plant.

PROCESS DESIGN METHODOLOGY

The process design for complex exchangers is based on the same concepts as simple cross exchangers. Before discussing the details of the design procedure, the following definitions are presented to avoid confusion. A cooling load is a process stream which must be cooled or condensed. Similarly, a heating load is a process stream which must be heated or boiled. Thus a refrigerant stream being boiled represents a heat load for the refrigerant. Since a stream with a cooling load is to be reduced in temperature, the cooling load must be at a higher temperature than the stream with which it is cross exchanged. Thus the cooling load must be at a higher temperature than the corresponding heating load at all points within the exchanger.

Obtain Process Heat and Material Balances

The first step is to perform an analysis of the entire process and obtain a heat and material balance for each operation. A process simulation program is obviously the most convenient means to complete this step. The heat and material balances should be performed as if all of the exchanger sides were independent. For this step, refrigeration and external loads are supplied independent of the process and should be ignored. In addition, exchangers in series should be treated as a single exchanger to permit full optimization without worrying about duty constraints.

Tabulate Duties

The second step is to list the inlet and outlet temperatures and duties for all units involving heat exchange.

Match Duties and Temperatures

The third step in performing the process design for complex exchangers is to begin matching duties. The best procedure to follow for most gas processing facilities is to start with the largest cooling loads. If heating loads with temperatures lower than the exit temperature of the cooling load are available, these should be accumulated until the duty approximately matches the cooling load. Any excess cooling load will have to be handled by refrigeration. If no heating load has a starting temperature below that of the cooling load, the "bottom end" of the cooling load will likewise have to be offset with refrigeration. This procedure should be repeated for all cooling loads in the plant. Sometimes, if a large low temperature heat load is present, it will be necessary to group cooling loads against it. However, the usual case for gas processing facilities will have a single large cooling load balanced against several process heating loads plus refrigeration.

Balance Loads

The heating and cooling loads in the complex exchanger should be balanced by separately summing the duties on each side of the exchanger. The residual duty must be reduced to zero by refrigeration or a heat source. The overall process should, obviously, be optimized to reduce any external heating or cooling loads. In many cases, it is possible to almost balance the heating and cooling loads in complex exchangers by adjusting the conditions on one side or another.

Potential Problems

Complex exchangers should be examined to assure that it is physically possible to operate the unit. The major problem which arises with these units is the internal temperature pinch point. This is a condition where, due to phase change or large flow rates across a narrow temperature range, the cumulative cooling load temperature drops below the cumulative heating load temperature. In this case, the heat cannot be delivered, and the unit will not function. This must be checked in all cases by generating heat delivery versus temperature curves for each side of the exchanger and summing the resultant curves to achieve overall heating and cooling delivery versus temperature curves for each exchanger.

This design procedure is of a trial and error nature and involves extended calculations and considerable work. This is the type of problem that is well suited for application on the computer. The most desirable situation is for this complete design procedure to be implemented in a process simulation package.

Process Simulation

The capability to perform the process design for complex exchangers has recently been added to PROSIM®[2], a process simulation program for hydrocarbon processing facilities. In this process simulation program, the process design for a plant using complex exchangers is carried out by using the program's interactive graphics capability to draw the process flow sheet on the computer screen and by filling in pop-up forms for the equipment operating parameters. The pop-up forms are used for each inlet stream and each unit operation in the plant including complex exchangers.

The group of exchanger sides which are to be included in any complex exchanger is identified in the program input. The program then performs the process heat exchange calculations and automatically balances the duty. The program also calculates the incremental duty for each exchanger side and the summed heating and cooling duties (or loads) as a function of temperature. These results are placed in Lotus® compatible files for easy plotting. This program also identifies areas in the complex exchangers which have impossible temperature crosses and issues appropriate warnings.

The capability of PROSIM to handle complex exchangers in the simulation of gas processing facilities is now used to perform three case studies.

CASE STUDIES

Three case studies involving the simulation and optimization of cryogenic operations using complex multi-stream brazed aluminum exchangers is presented. The first case is intended to be a straightforward, easily grasped example of the methodology for the application of complex exchangers while the second case is intended to be a somewhat more complicated example and to show the possibilities of mixed refrigerant systems. The third case is presented to demonstrate the methodology for application of the full capabilities of complex exchangers.

Case 1 LPG Recovery Using Propane Refrigeration

A small liquids recovery unit with a deethanizer and conventional propane refrigeration is used in the first case study. A typical flow diagram employing conventional shell and tube exchangers is shown in Figure 2. The inlet gas is chilled to condense the heavy fractions, which are separated and passed to a deethanizer to remove the methane and ethane. To reduce the refrigeration load, both the sensible heat from the cold vapor and some latent heat from the condensed liquids are used to cool the inlet gas. As shown in Figure 2, this requires three shell and tube exchangers and a flow control valve to properly balance the flow of inlet gas between the parallel exchangers.

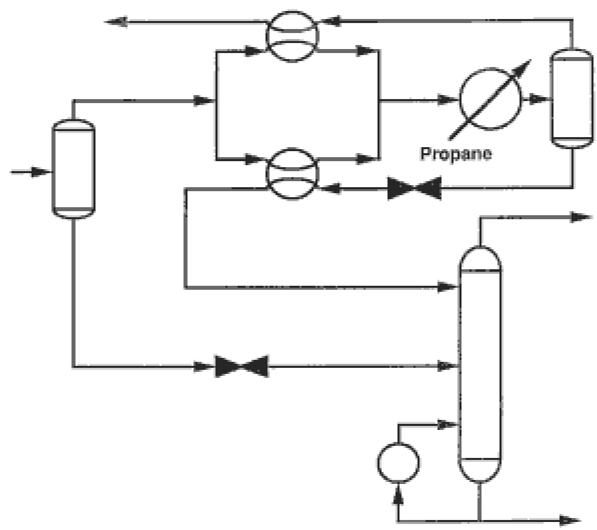

Figure 2. Liquids recovery unit with conventional exchangers for Case 1.

A single complex exchanger with four sides can be used to replace the three shell and tube exchangers as shown in Figure 3. A detail schematic of the complex exchanger is shown in Figure 4. The inlet gas is cooled simultaneously against the cold gas, the cold liquid and the propane refrigerant which means the control valve dividing the inlet gas is no longer necessary. An additional benefit is that a portion of the sensible heat in the propane refrigerant vapor may also be used to cool the inlet gas eliminating an economizer exchanger from the propane refrigeration loop.

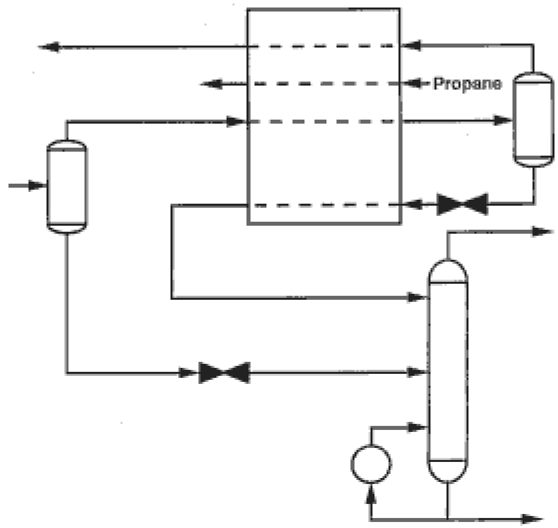

Figure 3. Liquids recovery unit with complex plate exchanger for Case 1.

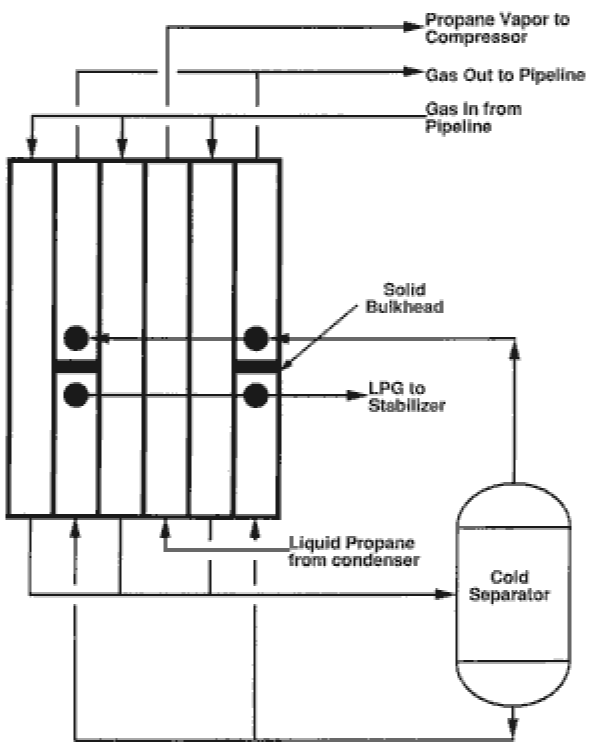

Figure 4. Detail process flow schematic for complex plate exchanger in Case 1.

As shown in Figure 4, the complex exchanger consists of a series of chambers which are formed by plates of aluminum and held apart by spacers made of bent aluminum wire which have the appearance of tiny bridge trusses. In addition to holding the plates apart, these spacers act as extended heat transfer surfaces and induce turbulence to enhance the heat transfer coefficient. The feed gas is charged to the top of the unit into alternating cells and falls out the bottom along with the condensed LPG's. The flashing propane liquid refrigerant is introduced through the bottom of the unit into other cells and boils as it proceeds upward to be removed at the top of the unit as hot low pressure vapor. Each chamber of the exchanger handling the off gas and LPG is divided by a horizontal bulkhead. The LPG's are introduced at the bottom of the unit and removed through the side of the unit as a two phase mixture just below the bulkhead. The off gas is introduced above the bulkhead and is removed from the top of the unit.

The operating parameters for the complex exchangers are based on the following process considerations. The temperature of the cold gas fed to the separator is fixed by the amount of condensed liquid desired along with its composition. The temperature of the residue gas to the pipeline is fixed by the temperature of the feed gas less a 10^{o}F approach. The exit temperature of the cold LPG stream is fixed by the acceptable vapor-liquid split at the deethanizer feed nozzle. When the duties for the above operations are added, the result is the refrigeration load required by the system. The pressure of the boiling propane in the exchanger is set to give an inlet temperature about 10^{o}F below that of the cold separator. The exit temperature of the propane vapor should be about 10^{o}F below the feed gas assuming this causes no problems in the propane compressor. The temperature of the incoming refrigerant liquid may be taken as 120^{o}F before the choke.

These operating parameters may be supplied to the process simulation program to perform heat and material balances for the plant. Based on the simulation results, the operating parameters may be adjusted to optimize the unit.

The last step in the process design for complex exchangers is to ensure that no temperature pinch points occur. A set of incremental duty versus temperature curves is calculated by the program as shown in Table I and placed in a Lotus compatible file for plotting, as shown in Figure 5. The summed heating load curve must remain below the summed cooling load curve to avoid temperature pinch points. The larger the separation between the curves, the higher the effective LMTD and the less area required to achieve the heat exchange. For a simple exchanger such as this, it is trivial to avoid these pinch points. The main cooling duty is delivered at the low end by boiling propane. This accounts for the sudden massive drop in the cumulative heat curve at the left of the chart. The effects of sensible reheating of all three of the cool streams causes the characteristic wedge shape in the curve.

Table I. Simulation program output for complex exchanger for liquids recovery unit for Case 1.

Complex Exchanger Contains Blocks:
BLOCK NUMBER: 24.00 TYPE: EXCH NAME: CB-1 COOL
BLOCK NUMBER: 35.00 TYPE: EXCH NAME: CB-1 HEAT
BLOCK NUMBER: 45.00 TYPE: EXCH NAME: CB-2 HEAT
BLOCK NUMBER: 60.00 TYPE: EXCH NAME: CB-3 HEAT

RESIDUAL DUTY ERROR: -332.6 BTU/HR
EFFECTIVE OVERALL DELTA T: 12.45 DEG F

TEMP DEG F	HEAT 24.00 MBTU/H	COOL 35.00 MBTU/H	HEAT 45.00 MBTU/H	HEAT 60.00 MBTU/H	HEAT TOTAL MBTU/H	COOL TOTAL MBTU/H
-11.9	188.2				324.1	324.4
-5.3	59.9				195.7	324.4
-5.0	59.7	324.0	86.4	49.4	195.6	324.4
1.3	56.3	305.5	81.2	42.9	180.4	305.5
7.9	52.8	285.8	75.8	36.0	164.5	285.8
14.5	49.2	266.1	70.5	28.9	148.6	266.1
21.1	45.6	246.5	65.2	21.7	132.5	246.5
27.7	41.9	226.9	60.0	14.3	116.2	226.9
34.3	38.3	207.3	54.7	6.7	99.8	207.3
40.0	35.1	190.6	50.3	0.0	85.3	190.6
40.9	34.6	187.9	49.6		84.1	187.9
47.5	30.8	168.7	44.4		75.2	168.7
54.1	27.1	149.7	39.3		66.3	149.7
60.7	23.3	131.0	34.2		57.4	131.0
67.3	19.5	112.7	29.1		48.5	112.7
73.9	15.6	94.9	24.0		39.6	94.9
80.4	11.7	77.7	18.9		30.6	77.7

87.0	7.8	61.3	13.8	21.6	61.3
93.6	3.8	46.2	8.7	12.6	46.2
100.0	0.0	33.2	3.8	3.8	33.2
100.2		32.8	3.7	3.7	32.8
105.0		24.2	0.0	0.0	24.2
106.8		21.1		0.0	21.1
113.4		10.3		0.0	10.3
120.0				0.0	0.0

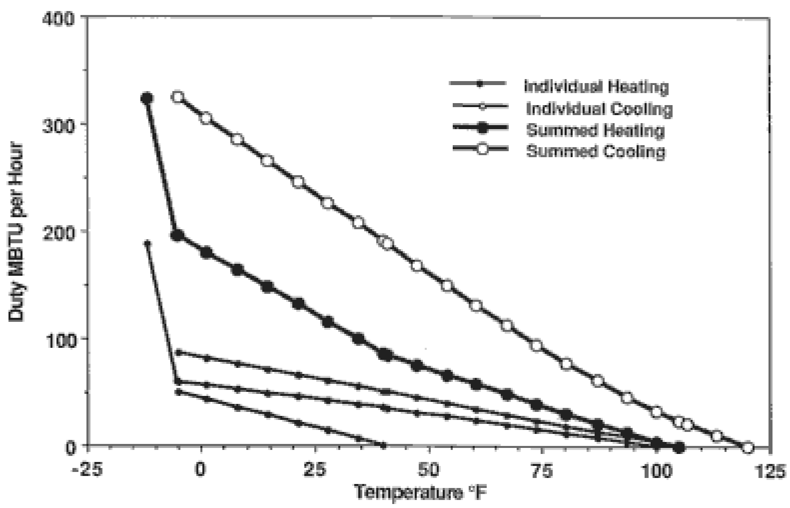

Figure 5. Incremental duties for complex exchanger for liquids recovery unit in Case 1.

Case 2 Ethane Recovery Using a Mixed Refrigerant System

In this case a light refinery gas is processed to extract ethane plus (C_2+). As shown in Figure 6, the incoming gas will be chilled to -100^{o}F by cross exchange with the cold gas and liquid streams and a mixed refrigerant. After cross exchange, the cold liquids are demethanized in a refluxed column. The column overhead condenser is also added into the main complex exchanger making five sides.

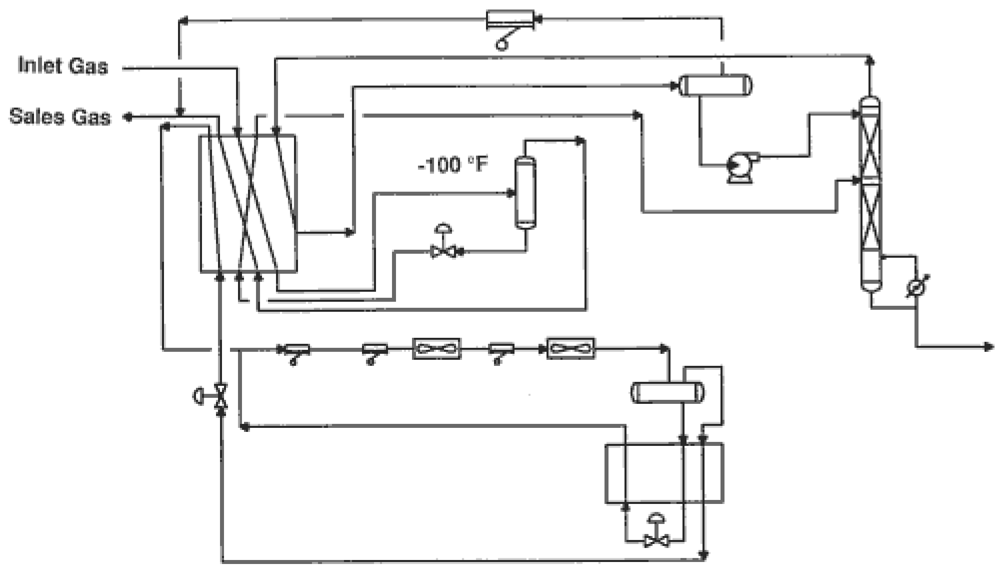

Figure 6. Process flow for ethane recovery using mixed refrigerant system for Case 2.

The mixed refrigerant system uses a blend of methane through butanes to generate a heavy and a light refrigerant liquid. As shown in Figure 6, the heavy liquid is used to liquefy the light refrigerant in a second complex exchanger. It can also be used for any other higher temperature refrigeration requirements. The light refrigerant liquid is then used to achieve the -100°F cooling of the feed gas.

The operating parameters for the main exchanger are fed into the process simulation program which calculates the heat and material balances for the plant. The program also balances the duties in the complex exchanger, calculates the incremental duty versus temperature data and places this data in Lotus compatible files for plotting. Inspection of the curves for the initial process design of the primary complex exchanger plotted in Figure 7 reveals that a major temperature cross occurs in the interior of the exchanger. The program also issues warnings of impossible temperature cross in the interior of the exchanger. This cross is not evident from examination of the end approaches.

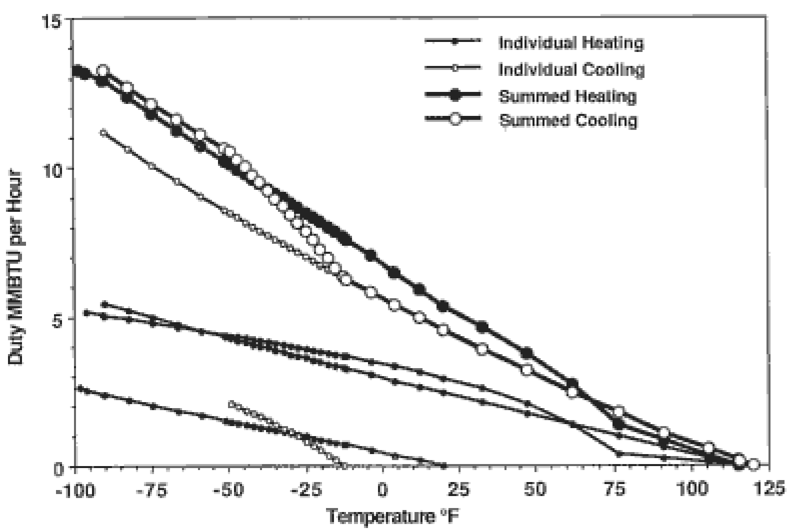

Figure 7. Trial 1: Incremental duties for complex exchanger for ethane recovery unit in Case 2.

To alleviate the temperature pinch point problem, the cooling load curves must be forced upward and/or the heating load curves forced downward. In this case, the easiest change is to increase the methane content in the refrigeration loop. The duty versus temperature curves shown in Figure 8 incorporate this change and reveal that there is still a small pinch point. The refrigeration suction pressure was lowered to produce good duty versus temperature curves throughout the range as shown in Figure 9. The fact should be emphasized that every one of the trials for this case showed a heat balance that was within acceptable limits as well as inlet/outlet temperatures which seemed feasible.

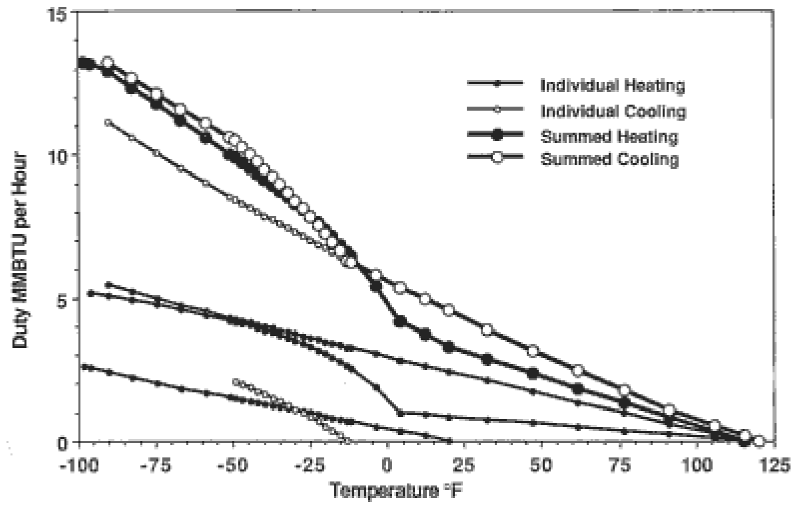

Figure 8. Trial 2: Incremental duties for complex exchanger for ethane recovery unit in Case 2.

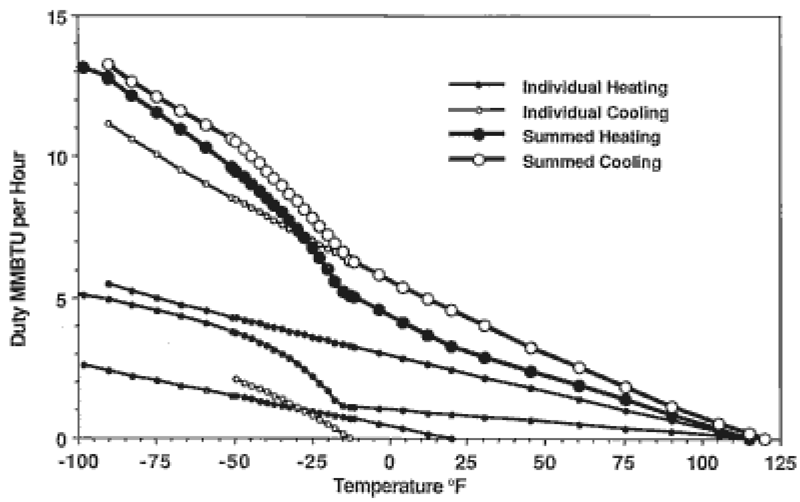

Figure 9. Final trial: Incremental duties for complex exchanger for ethane recovery unit in Case 2.

Case 3 LPG Recovery Using Refrigeration and Turbo Expander

This case is an elaborate expander plant currently under construction and involving a very heavy inlet gas stream. As shown in the process flow diagram in Figure 10, the rich gas is chilled in the first complex exchanger, the liquids are separated and sent to a deethanizer. The cold overhead gas is further cooled in a second complex exchanger, separated from the liquids and passed through a turbo expander to a demethanizer. The third complex exchanger involves a specialized process to subcool the expander bypass gas to use the Joule-Thomson effect to reflux the demethanizer which is patented by Elcor/Ortloff[3].

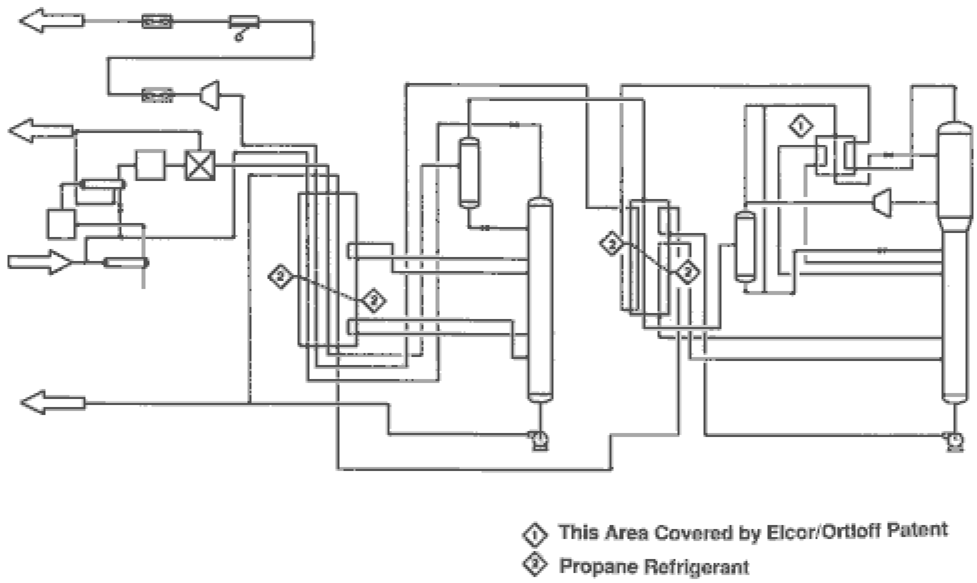

Figure 10. Process flow for liquids recovery from rich gas using refrigeration and expander for Case 3.

The first complex exchanger involves eight process streams including the entire heating and cooling load for the deethanizer. As can be seen from the incremental duty versus temperature curves in Figure 11, adequate temperature difference for good heat exchange is available everywhere. One heat load curve and the summed heat load curve are somewhat peculiar due to the large amount of refrigeration required for the very rich gas.

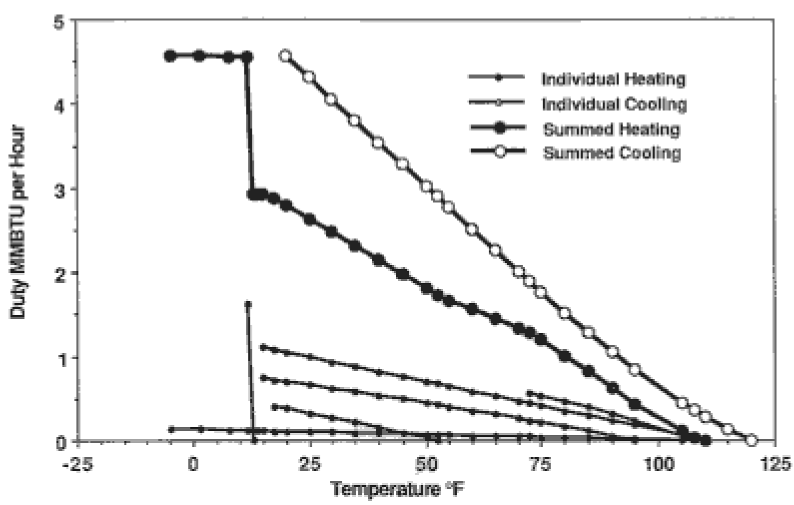

Figure 11. Incremental duties for main complex exchanger for liquids recovery plant in Case 3.

The second complex exchanger is the cryogenic exchanger and involves four process streams and one refrigeration stream. As shown in the duty versus temperature curves in Figure 12, the refrigerant load is delivered at just the right temperature to avoid an impending pinch point. Delivering the refrigerant load at a lower temperature would decrease the area required for this exchanger but would increase the size and operating cost of the refrigerant unit.

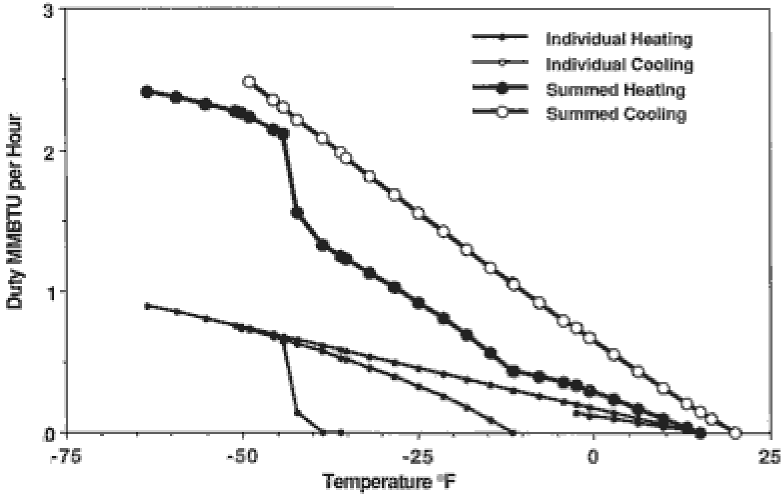

Figure 12. Incremental duties for second complex exchanger for liquids recovery plant in Case 3.

The third complex exchanger involves only three process streams and is quite straightforward. The only concern in this case is to place the demethanizer side reboiler high enough in the column to maintain its exit temperature below that of the incoming feed stream.

SUMMARY AND CONCLUSIONS

Complex brazed aluminum plate exchangers can be used for heat exchange between as many as 8-10 process streams within a single exchanger. The process design of these exchangers requires extended calculations and considerable work due to the number of streams involved and the complexity of the exchangers. The design procedure involves performing heat and material balances, matching duties and temperatures, balancing duties by adding refrigeration or heat and checking for internal temperature pinch points.

The capability to perform the process design for complex exchangers has been recently added to a process simulation program called PROSIM. The program performs the process calculations, automatically balances the duty and calculates incremental duty as a function of temperature through the exchanger.

Three case studies involving the simulation and optimization of cryogenic operations using complex exchangers were presented. The case studies included a simple refrigeration system for liquids recovery, an ethane recovery unit using a mixed refrigerant system and a refrigeration/turbo expander plant. These case studies have shown that the incremental duty versus temperature curves must be closely inspected to ensure that no temperature pinch point exists in any exchanger. If a pinch point does exist, the process or the refrigeration loop must be adjusted to overcome the pinch point. Since these exchangers are considerably less expensive and have heat transfer coefficients on the order of twice those for shell and tube exchangers and since they can act like a complex network of simple exchangers, plate exchangers can be very useful in reducing costs and optimizing cryogenic operations.

REFERENCES

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2. Bryan Research and Engineering, Inc., *PROSIM User’s Manual*, 1988.

3. Elcor/Ortloff, Canadian Patent Numbers 1048397 and 1073804.