CONDENSATE STABILIZATION: HOW TO GET THE MOST FOR YOUR MONEY

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ABSTRACT

Condensate stabilization is one of the most common operations in the gas processing world. The principal considerations affecting the performance of the plant when debottlenecking or designing new condensate stabilization units are the configurations, product quality, energy saving, and equipment size. Sound process design can significantly reduce capital investments and operating costs. The purpose of this paper is to review and illustrate the fundamentals of this process, including flow-sheet configuration and key operating parameters, and to show how they affect the plant performance, including energy consumption and economics.

INTRODUCTION

The incoming gas from offshore or onshore pipelines carries condensates generated during transmission by heat transfer and pressure drop. The two phases are typically separated in the slug catcher or inlet separator. The gas stream is treated before feeding to the cryogenic system for extracting NGL. It is first routed to the acid gas removal unit, then dried in a Glycol unit, and polished in a molecular sieve dehydration unit, (Figure 1). The recovered hydrocarbon condensate is sent to the condensate stabilization unit for vapor-pressure reduction. Stabilized condensate can be stored in floating roof storage tanks, and safely transmitted in pipelines. The stabilized condensate, typically C5+, is sold as final product or as petrochemical feed-stock to olefins and aromatic plants. Figure 1 illustrates how the vapor from a condensate stabilization unit is compressed and recycled to the inlet feed separator gas stream before entering the gas treating system. The presence of contaminants (water, salts and suspended solids) in the slug catcher can cause problems on the condensate stabilization unit performance, including long-term maintenance issues. Water carryover from the slug catcher and hydrocarbon/water separator can lead to off-spec stabilized condensate.

This study (1) evaluates six alternative process configurations and key operating parameters for meeting the stabilized condensate specifications and saving energy, (2) develops optimized utility consumption for the alternatives, (3) develops comparative capital and operating cost appraisals, and (4) presents the optimum alternative in terms of process efficiency and economic analysis. An appropriate process simulator with the right thermodynamic package, optimum flow sheet configuration and equipment sizing is a very important tool for achieving the desired results.

Figure 2 illustrates the integrated engineering and economics work flow. In this paper, HYSYS and Process Economic Analyzer (APEA) is used for simulation and economic analysis. This Aspen Technology, Inc., software is very effective in carrying out comparative evaluation of process designs. Based on a HYSYS flow-sheet simulations, APEA determines the capital expenditures, operating costs, and finally the profitability of the proposed designs. APEA has a system which links process simulation programs to the estimating tool, though a “process engineer’s window” to the “Icarus Estimating engine”. It is used to (1) extend the results of process simulation, (2) generate sizing and cost estimates for process equipment, (3) perform preliminary mechanical designs, and (4) estimate materials and labor costs, indirect costs, total capital investment, and (5) the EPC schedule. [1]

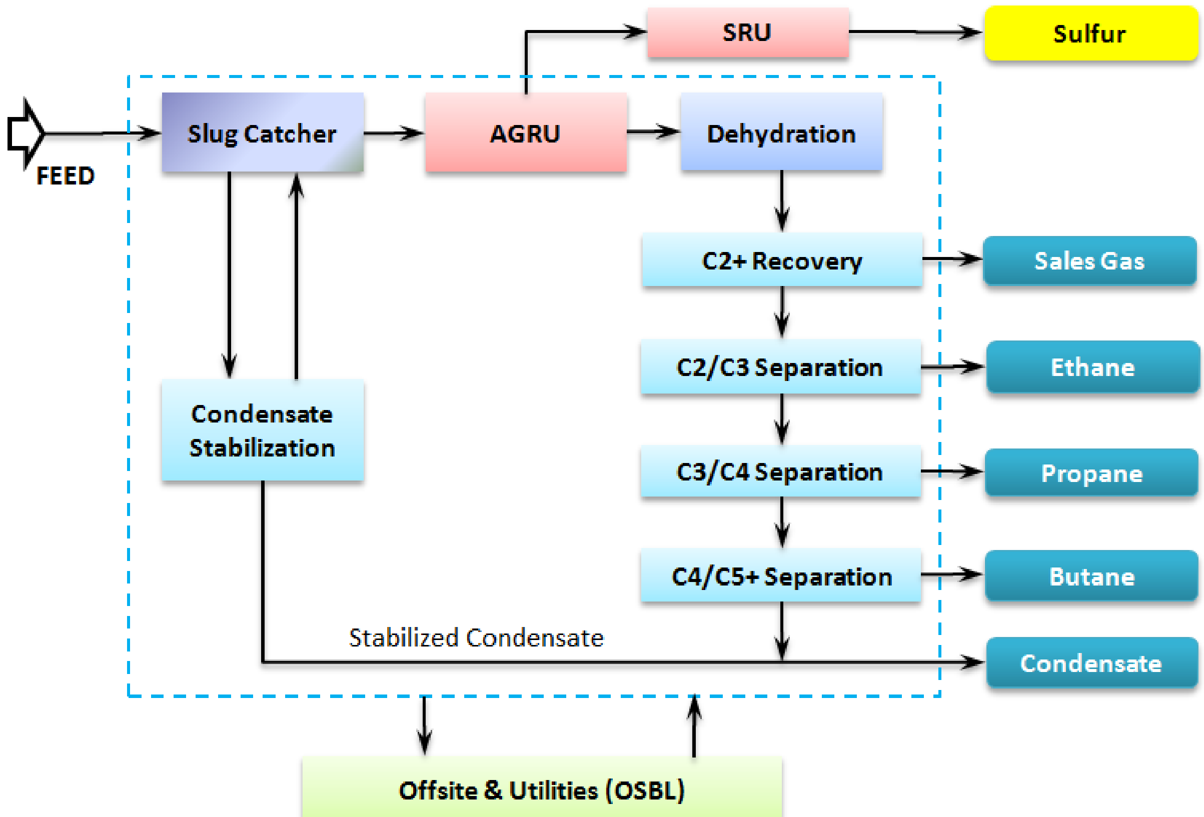

Figure 1 – Typical Block Flow Diagram for a NGL Recovery Plant

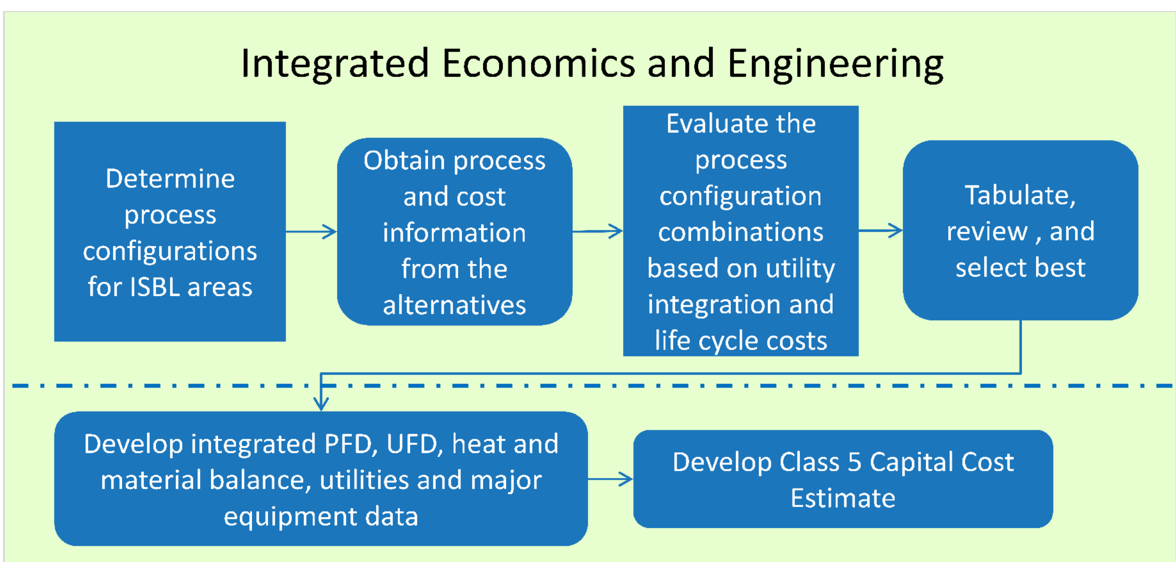

Figure 2 – Process Configuration Screening Work Flow Process

BASIS of DESIGN

Feed characteristics

The single-phase liquid feed properties used for the case studies are shown in Table 1.

Table 1 – Unstabilized Condensate Characteristics

Component	Molar Flows (lbmole/hr)
H_2O	1449.7
N_2	16.1
CO_2	63.5
H_2S	0.0
C1	1317.2
C2	271.3
C3	218.2
i-C4	99.6
n-C4	165.6
i-C5	146.8
n-C5	126.4
C6+	2330.0
Total	6204.3
Pressure, Psig	600.0
Temperature, °F	90.0
RVP, Psia	250.0

Products specifications

The products of condensate stabilization unit are:

- Stabilized condensate meeting the required specifications (Table 2),
- Off-gas compressed and recycled to the gas treating system unit.

Table 2 – Stabilized Condensate Specifications

Component	Specification	Test Method
RVP, Psia	<8.5	ASTM D-323
Color	Not Less than 25 Saybolt	ASTM D-156
Water Content	Nil	

Battery limit conditions

The operating conditions of the major streams to the condensate stabilization unit are:

Table 3 – Battery Limit Conditions

		Operating Conditions	
	Description	Pressure, Psig	Temperature, °F
Feed	Liquids from slug catcher	600	90
Products			
Stabilized condensates	To floating roof storage tank	Atm.	130
Off-gas	2^{nd} stage aftercooler outlet	1100	130

Turndown ratio

The Condensate Stabilization unit is designed to operate 40-110% of the normal continuous flow rate.

CASE STUDIES

The objective of this case study is to evaluate different condensate stabilizer configuration alternatives and operating parameters. Table 4 describes the six process configurations compared in this paper. In all alternatives, the stabilizer is operated at pressure of 145 Psig except for Alternative 4, which is operated at 87 Psig. In all studied cases, 10 theoretical-stages of valve trays are used as additional stages appear to offer no further benefit. The following assumptions are made for the economic analysis based on U.S. Gulf Coast:

- Currency Conversion Rate = U.S. Dollar
- Stabilized condensate = $ 102 per barrel
- Gas Price = $ 3.30/1000ft^3
- Steam cost @ 400 Psia = $ 5/klb
- Electricity cost = $ 0.07/KWh
- Plant availability = 355 days per year
- Annual rate of return = 20%
- Plant life = 25 years

Table 4 – Process Configuration Alternatives for Comparison

	Refluxed Column	Non-Refluxed Column	Non-Refluxed LP Column	Feed Preheating	Feed Splitting	Side Reboiler
Alternative 1		X				
Alternative 2		X		X		
Alternative 3		X		X	X	
Alternative 4		X	X	X	X	
Alternative 5		X		X	X	X
Alternative 6	X			X		

Alternative 1: non-refluxed column, no feed preheating, no split feed & no side reboiler

The condensate from the slug catcher is flashed again at lower pressure in the hydrocarbon/water separator. All hydrocarbon condensate leaving the condensate/water separator is sent to the top of the column. The stabilized condensate leaving the column is cooled before storage, which increases the bottom condensate cooler duty. Stabilizer overhead vapor is compressed by a two-stage reciprocating motor-driven compressor. The first-stage discharge gas is cooled down to 130 °F in an air cooler and is then sent to the three-phase separator for removing any condensate and water. Overhead gas is then mixed with gas from the hydrocarbon/water separator prior to feeding the 2nd stage overhead compressor suction drum. Gas is further compressed by the second stage, cooled in the after-cooler, and sent to the acid gas removal unit for further processing. Figure 9 shows the relationship between the inter-stage compressor pressure and the compressor power.

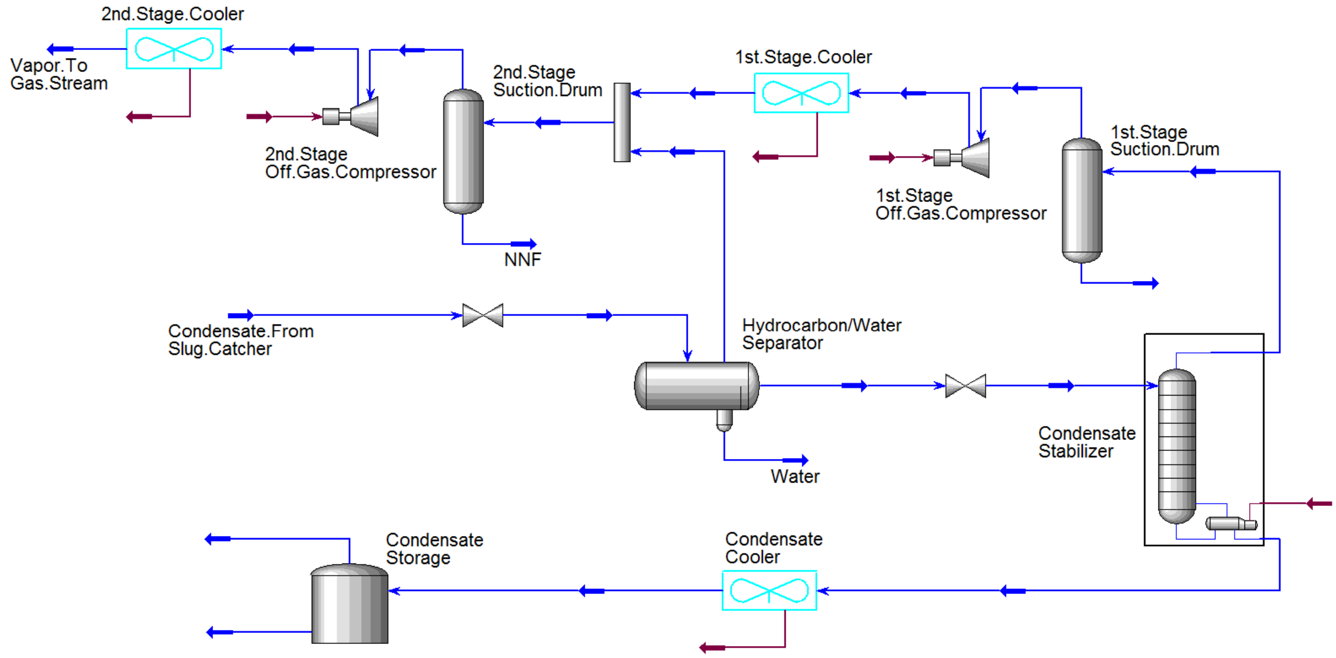

Figure 3 – Alternative 1 Process Flowsheet

Alternative 2: non-refluxed column with feed preheating, no split feed, and no side reboiler

This scheme is similar to Alternative 1. The difference is the presence of a feed/bottom exchanger as a preheating system. The entire hydrocarbon condensate stream from the hydrocarbon/water separator flows through the feed/bottom heat exchanger and enters to the top of the column. The vapor recovery compressors have relatively high power requirements, and reboiler has higher thermal duties. In addition, the bottoms from the 2nd stage suction drum are routed directly to the hydrocarbon/water separator to remove water.

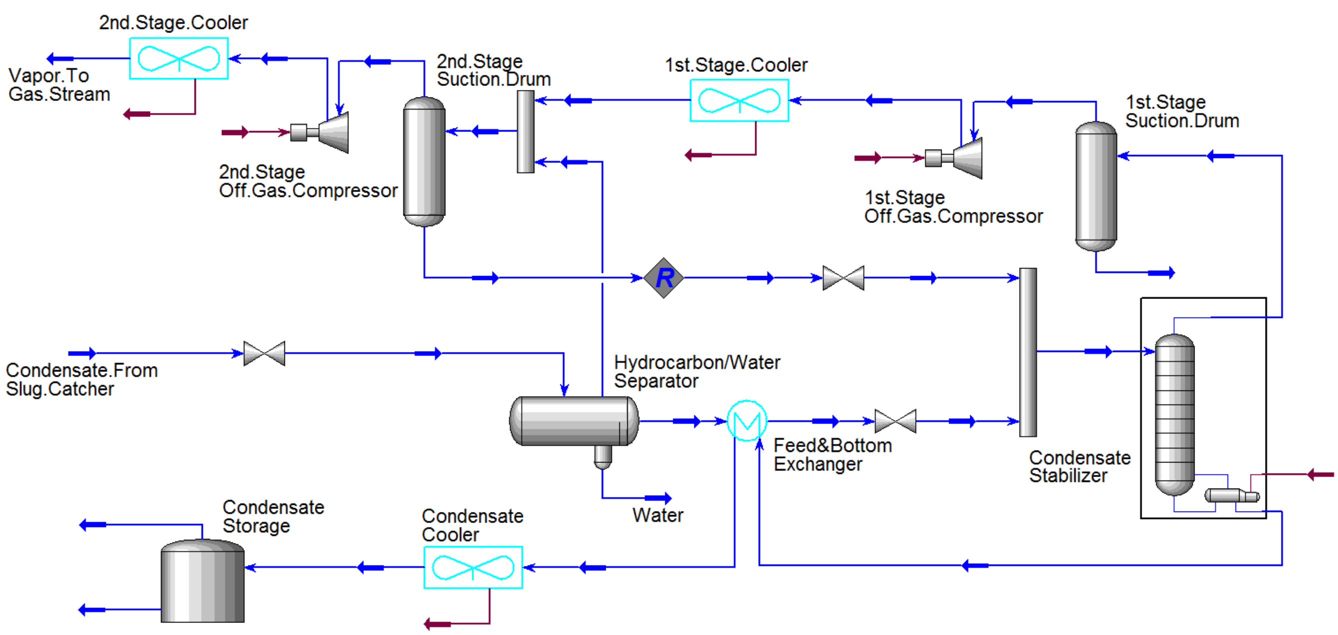

Figure 4 – Alternative 2 Process Flowsheet

Alternative 3: non-refluxed column with feed preheating, split feed, and no side reboiler

The process scheme detailed in Alternative 2 is modified so that the feed is split, with 17% sent cold to the top of the column. The residual feed passes to the feed/bottom exchanger before entering the column at the 4th theoretical stage. Figure 10 illustrates the impact of splitting feed and feed stage location on reboiler duty.

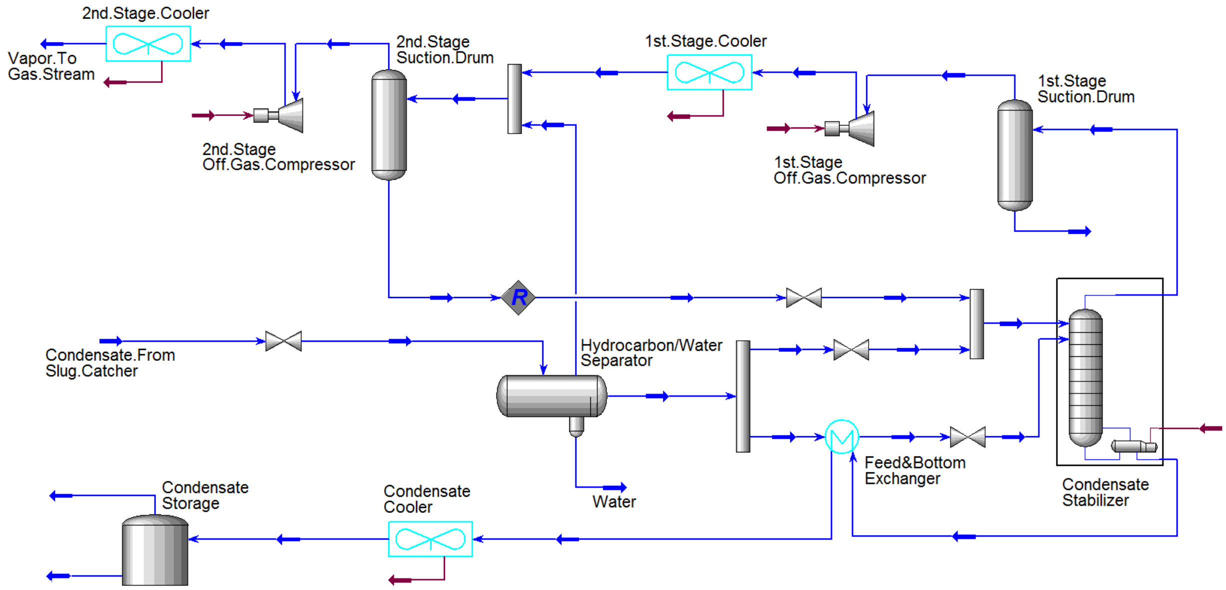

Figure 5 – Alternative 3 Process Flowsheet

Alternative 4: non-refluxed low pressure column with feed preheating & split feed & no side reboiler

This configuration is similar to Alternative 3, except that the stabilizer is operated at lower pressure (87 Psig), which requires an additional recompression stage. In this scheme, the feed is split, with 11% sent cold to the top of the column.

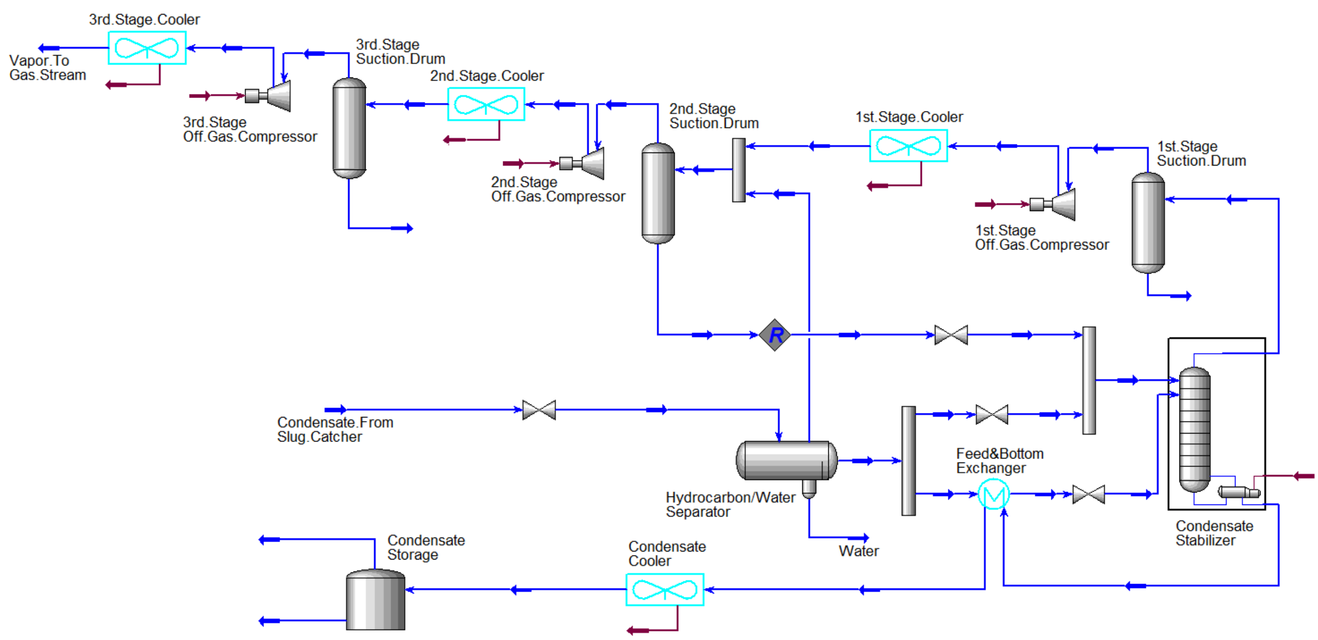

Figure 6 – Alternative 4 Process Flowsheet

Alternative 5: non-refluxed column with feed preheating, split feed, and side reboiler

This scheme of Alternative 4 is modified by the addition of a side reboiler at the 7th theoretical stage. The feed is split, with 10% sent cold to the top of the column. The residual feed is routed through the feed/bottom heat exchanger before entering the column at the 4th theoretical stage.

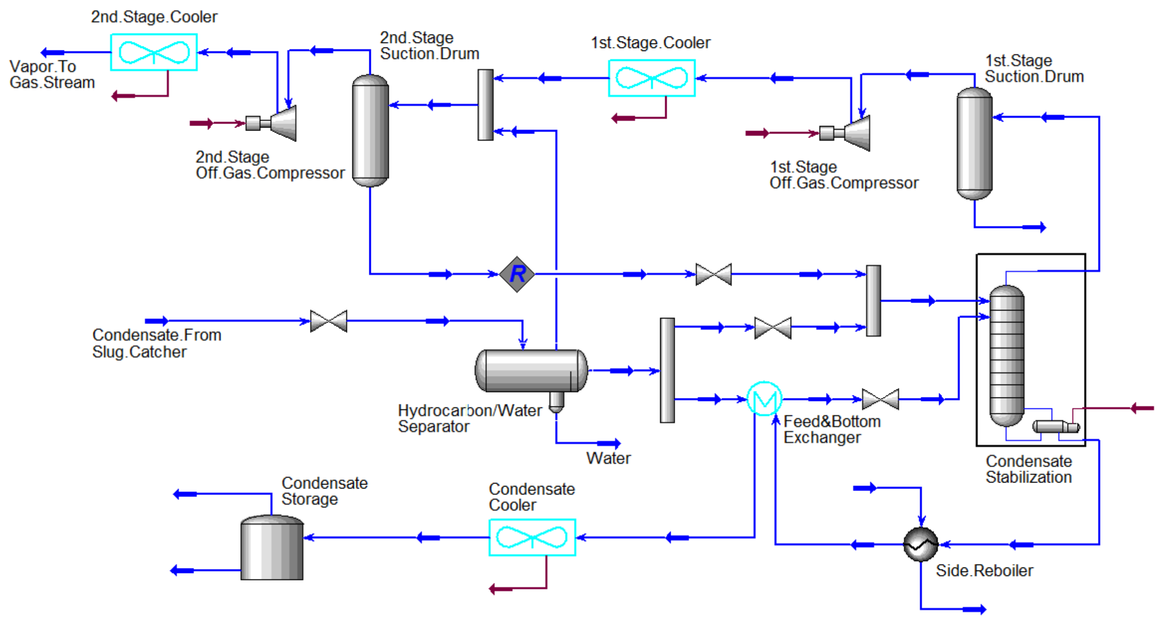

Figure 7 – Alternative 5 Process Flowsheet

Alternative 6: refluxed column with feed preheating, no side reboiler

The condensate leaving the hydrocarbon/water separator is heated using the feed/bottom heat exchanger, and enters the 4th theoretical stage of a conventional distillation column.

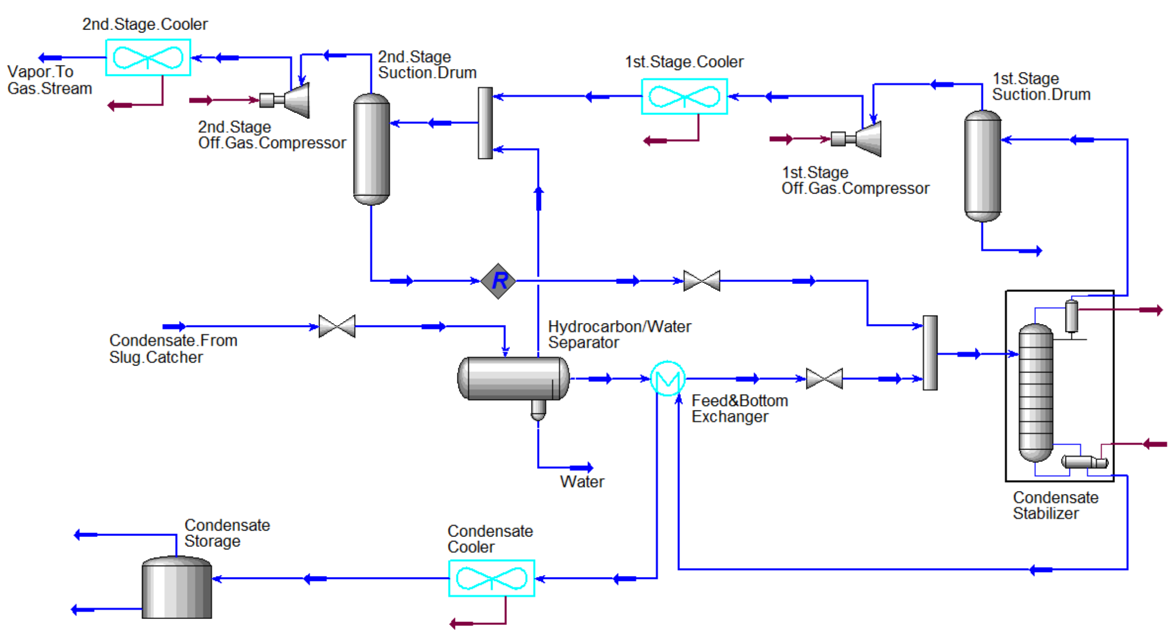

Figure 8 – Alternative 6 Process Flowsheet

RESULTS

Plant designers consider major design parameters for increasing process efficiency and saving energy. The following discussion is limited to the findings of the sensitivity analysis. Sensitivity analyses are performed by investigating and focusing upon various operating parameters, process configurations, and robustness of columns, by observing differences from simulation outputs, and economic analysis outputs. Table 5 presents the optimum alternative in terms of CAPEX and OPEX. Figure 11 illustrates that Alternatives 2 to 6 produce essentially the same stabilized condensate. The provided capital costs are a rough order of magnitude on a U.S. Gulf Coast basis. These estimated costs have been calculated by APEA, and cover engineering, procurement, and installation of process and utility equipment; storage, metering, and loading facilities; plus, construction indirect, spares, engineering, and commissioning. Items such as OSBL costs, owner's costs, land costs, escalation, and contingency are excluded.

Alternative 1 does not have the heat recovery system, and the hydrocarbon condensate with feed operating temperature is sent to the stabilizer. Operational experiences show that the product specifications are achieved with difficulty in terms of water content, because water separation from hydrocarbons is not easy at low temperatures. Figure 11 shows stabilized condensate flow rate, however, the highest reboiler, and condensate cooler duties in comparison to the other cases as shown Figure 13. Figure 9 shows the minimum off-gas compressors power at the lower operating temperature. This process configuration has a lowest capital cost because of the simplest design, fewest pieces of equipment, and less severe operating conditions, however, the operating and utilities costs are the highest, as reported in Table 5.

Alternative 2 can meet the product specifications with an efficient column temperature profile. The reboiler duty is significantly decreased by heat recovery (Figure 13).

Alternative 3 features a traditional design approach in terms of meeting product specifications, and operating cost optimization. Figure 13 shows that the reboiler duty is reduced by feed splitting in comparison to Alternative 2. Figure 10 shows the relationship between feed splitting, location of feed tray, and operating costs, including steam consumption and electricity usage.

Alternative 4 requires an additional compression stage because of the lower column pressure condition. Table 5 reports the highest capital investment; however, the reboiler duty is reduced in comparison to Alternative 3.

Alternative 5 shows an optimum design approach in terms of meeting product specifications, and operating cost optimization. The operating and utilities costs are minimized by the presence of the side reboiler exchanger for further heat recovery; however, the capital cost is slightly increased in comparison to Alternative 3 (Table 5).

Alternative 6 is a conventional distillation column. This case is the most complex system. CAPEX, OPEX and utilities cost are more expensive than Alternative 3 and 5 because of the additional reflux system, reported in Table 5.

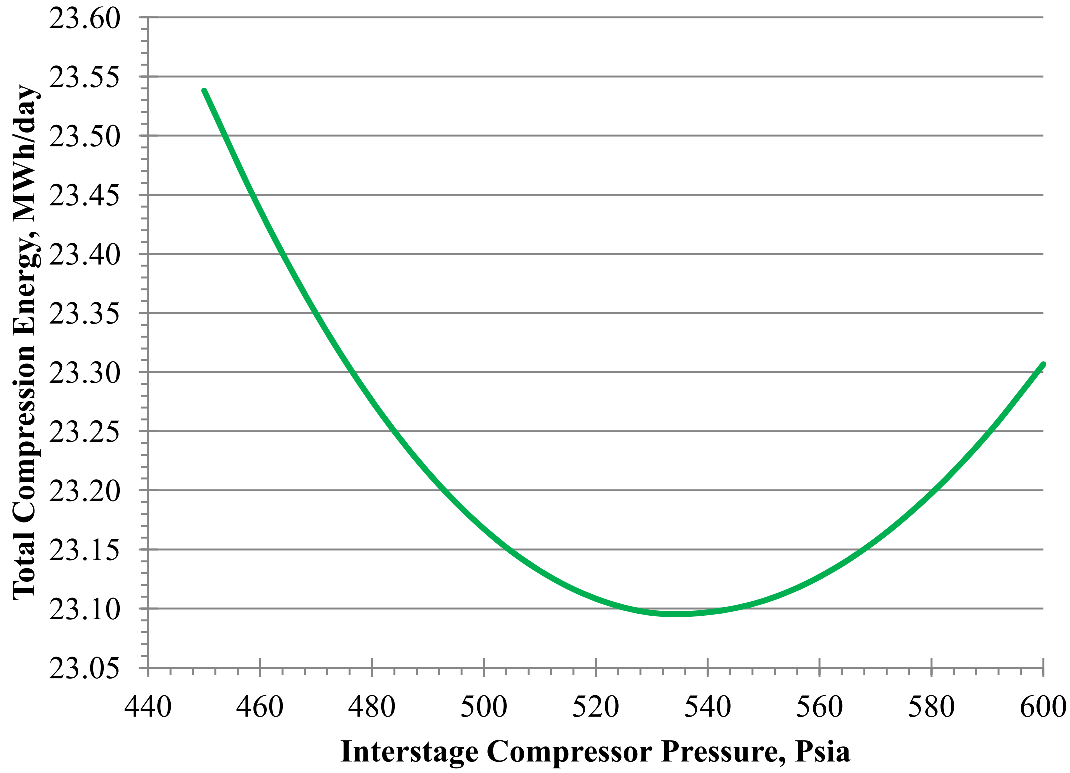

Figure 9 – Off Gas Compressor Power Optimization for Alternative 1

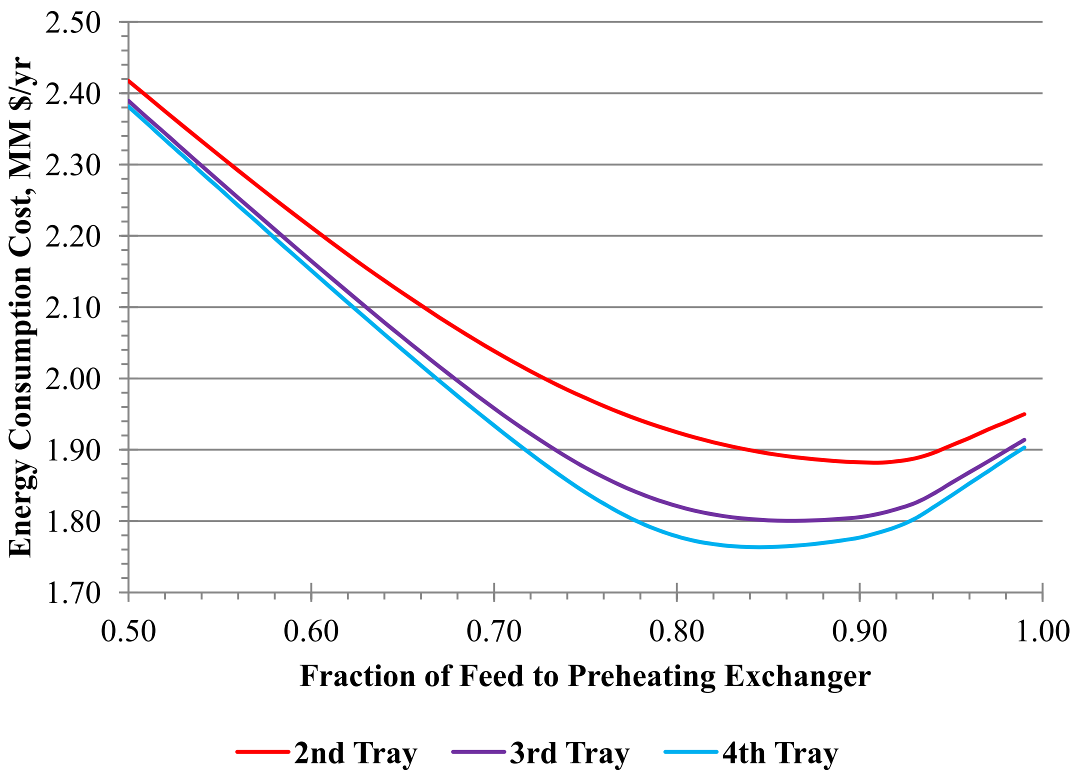

Figure 10 – Reboiler Duty Optimization by Splitting Feed & Feed Stage Sensitivity for Alternative 3

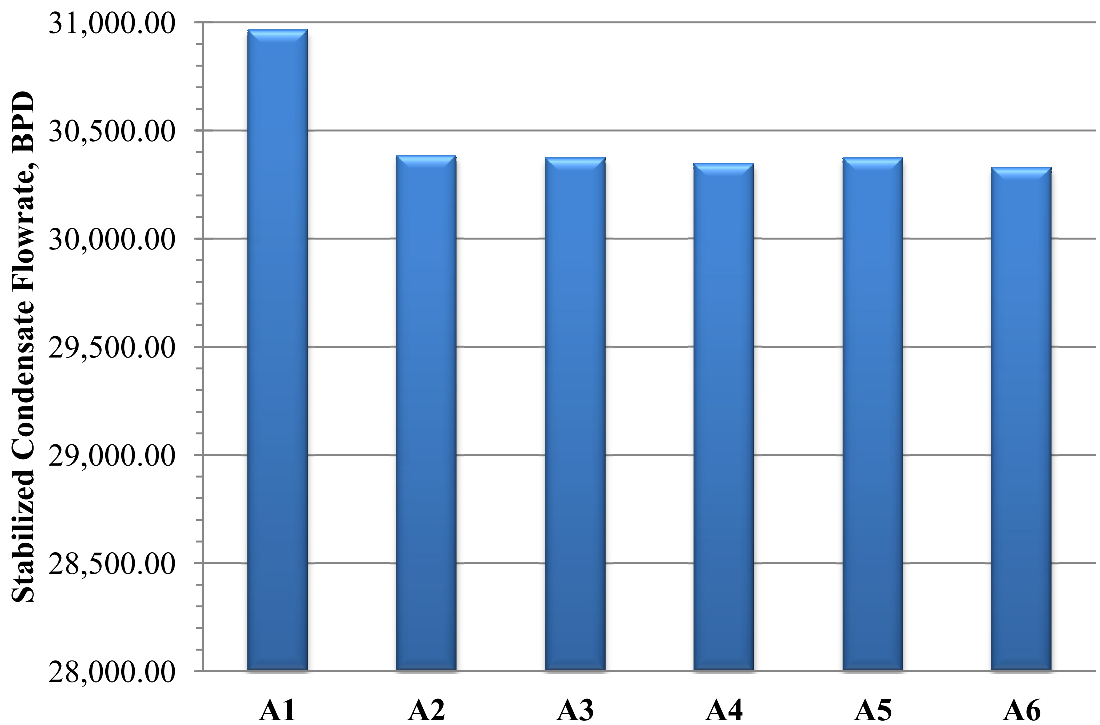

Figure 11 – Produced Stabilized Condensate for Each Process Configuration

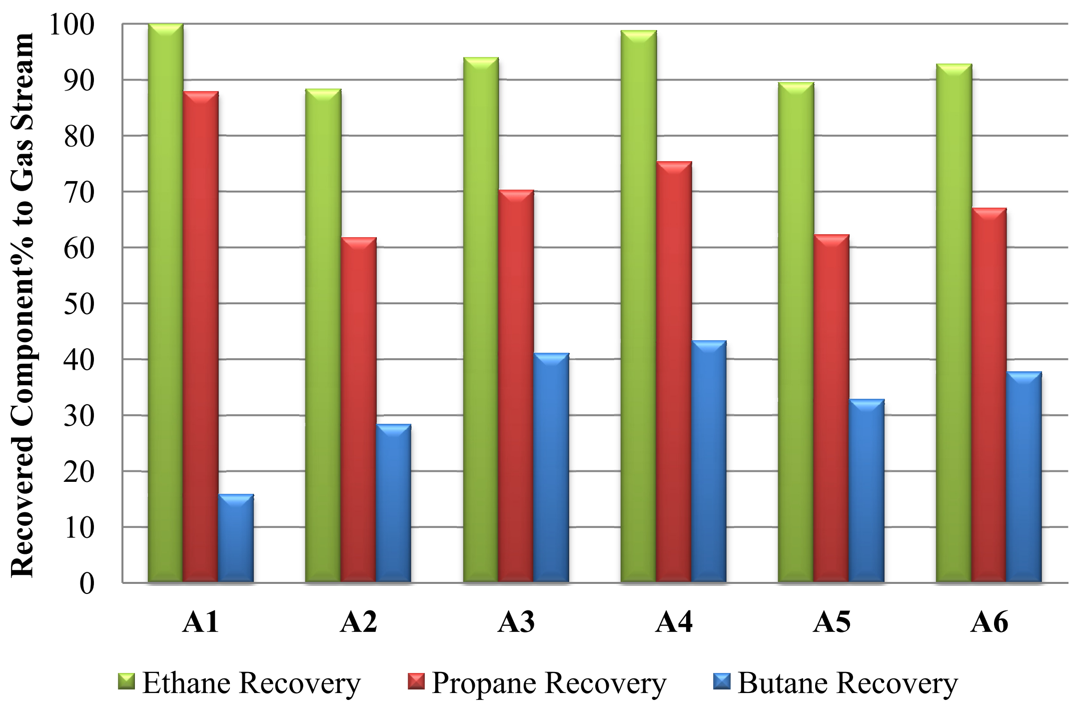

Figure 12 – Products Recovery for Each Process Configuration

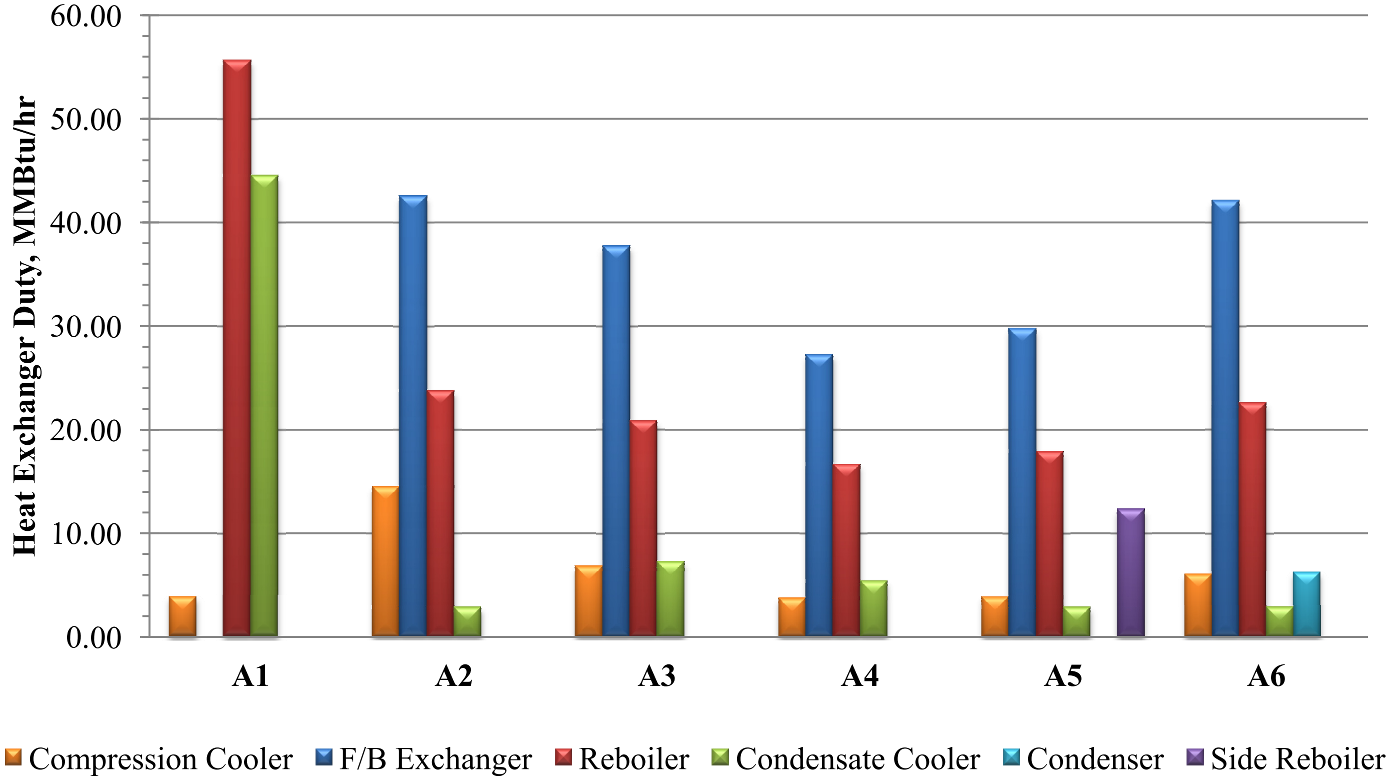

Figure 13 – Heat Exchangers Duty for Each Process Configuration

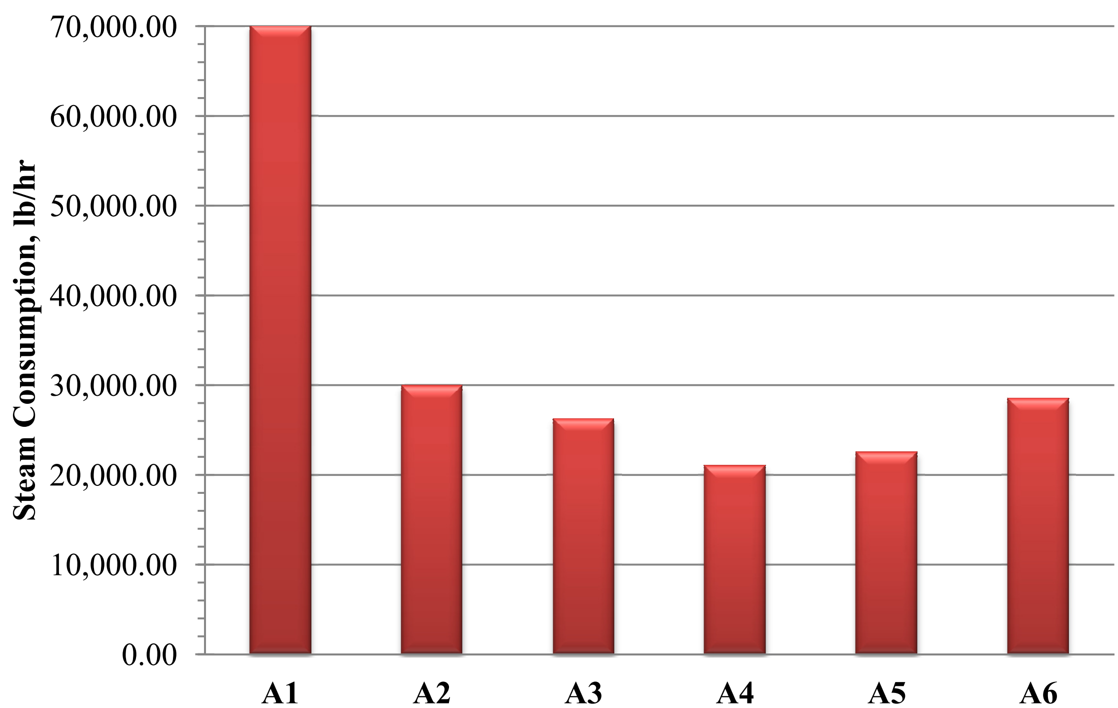

Figure 14 – Steam Consumption for Each Process Configuration

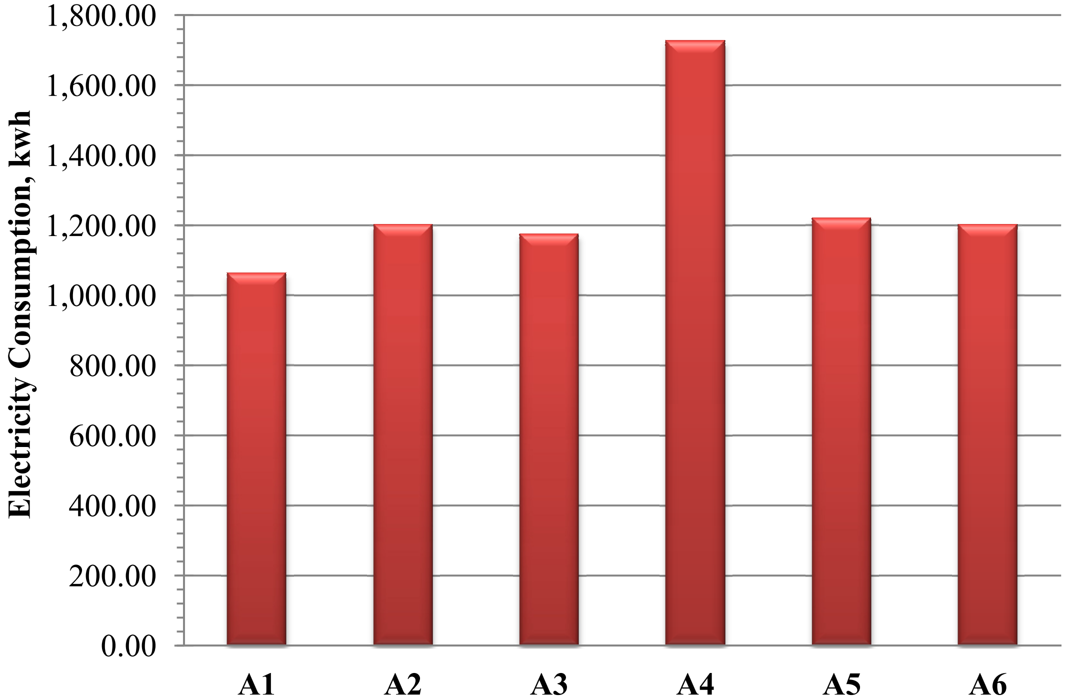

Figure 15 – Electricity Consumption for Each Process Configuration

Table 5 – Cost Comparison for Alternatives

	Capital Cost	Operating Cost (1)	Utilities Cost (2)	Product Sales
Alternative 1 (3)	0.89	1.68	2.16	1.02
Alternative 2	0.89	1.07	1.09	1.00
Alternative 3	0.97	1.06	1.08	1.00
Alternative 4	1.24	1.18	1.07	1.00
Alternative 5	1.00	1.00	1.00	1.00
Alternative 6	1.07	1.08	1.14	1.00

(1) The operating costs cover the raw materials, operating labor costs, maintenance, plant overhead, utility costs, electricity, fuel gas, potable water, and instrument air.
(2) The utilities costs cover the heating and cooling.
(3) Operational experiences show that the stabilized condensate specifications are achieved with difficulty with higher water content.

SUMMARY & CONCLUSIONS

Condensate stabilization units can be optimized by selecting the best process configuration and operating conditions depending on feed characteristics and required product specifications. A number of optimization runs are required using a steady-state simulator in order to determine the appropriate column pressure, column temperature, feed split ratio, and the feed tray location in terms of energy optimization. It is recommended that appropriate equipment design margins and sizing be used in order to ensure acceptable operational flexibility (particularly for water separation from hydrocarbon condensate). Table 5 summarizes the results of this study. Alternative 5 was chosen over the other process configurations through screening following on the following criteria:

- Acceptable capital cost, within the limits of study accuracy
- Lowest operating cost in terms of steam and electricity consumption
- Relatively low process complexity, which should lead to higher availability and lower maintenance costs in the long term, yielding the highest potential return to plant stakeholders.

Figures 9 – 14 illustrate sensitivities of various operating parameters. Based on this analysis and alternatives comparison, the following conclusions can drawn through the selection of the optimum process configuration:

- Optimizing of feed splitting and feed tray location:
  a. Reduces reboiler duty
  b. Reduces tower size
- Off-gas compressor power is optimized by adjusting the interstage pressure
- Off-gas after cooler duty is optimized by adjusting the interstage pressure
- Adding a bypass to the preheat exchanger and changing the feed tray location are obvious zero cost options.

REFERENCES

[1] Warren D. Seider, "Product and Process Design Principles" Second Edition, John Wiley, 2002