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Optimize olefin operations

This operating company used process models to find solutions to poor separation performance

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Bulk petrochemical manufacturing is a highly competitive global industry. When margins are tight, manufacturers seek ways to optimize performance and to reduce costs while maximizing yields and revenue. Optimization options include alternative feeds, plant/process revamps and improved operations to achieve better separation and yields to lower energy consumption, to minimize product loss and to decrease maintenance costs.

Case history. Pequiven is a leading petrochemical company based in Venezuela. Its products include fertilizers (ammonia and urea), chlor-alkali, methanol, methyl tertiary butyl ether (MTBE), aromatics, olefins (ethylene and propylene) and other plastics.

Fig. 1 shows Pequiven's Ana Maria Campos petrochemicals complex, Venezuela. This facility began operating in 1976, and it was expanded in 1992. This petrochemical complex has two olefin plants with a combined capacity of 635,000 metric tpy of ethylene and up to 250,000 metric tpy of propylene for 100% propane feed and uses ethane and propane as feedstocks.

Propane/propylene splitter study. The olefin plant's performance had deteriorated. The conditions resulted in significant propylene losses, higher energy consumption and rising maintenance costs. To improve performance, Pequiven needed a better understanding of process problems and a list of possible cost-effective solutions. Pequiven elected to simulate targeted sections of the olefin plant. Results from the models would provide more insight into the root causes of the poor operating performance. This article discusses the simulation study for the propane/propylene splitters. The study focused on the conceptual design and "what-if" analyses for various revamp options. Using the study results, Pequiven selected the best option to optimize the distillation columns.

Pequiven olefin process. The Olefins I Plant at the Ana Maria Campos Complex was designed to produce 250,000 metric tpy of ethylene and 120,000 metric tpy of propylene, using feedstocks ranging from 100% propane to a mixed feed of 30% propane and 70% ethane. Fig. 2 is the process flow diagram of the Olefins I plant.

The site processing operations are:

- **Pyrolysis.** This plant uses three sets of furnaces. The furnace effluent is first quenched and then cooled to condense the dilution steam, oils and polymers. All are removed by a circulating water system.
- **Process-gas compression.** The process stream is compressed and cooled to separate ethylene and propylene (principal products) from other byproducts and unconverted feed. Five compression stages are used, with acetylene conversion, caustic scrubbing and gas-drying occurring between the fourth and fifth stages. The process gas from the fifth-stage discharge filters is chilled in three stages using refrigerants and a hydrogen/tail-gas stream from the process.

FIG. 1 Pequiven's Ana Maria Campos petrochemical complex.

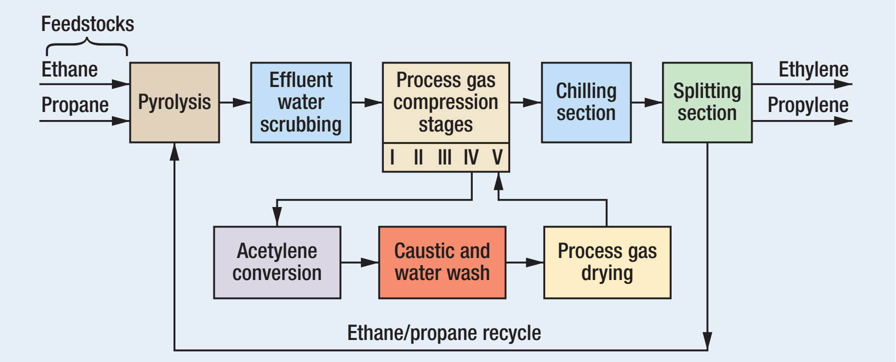

FIG. 2 Process flow diagram of Olefins I plant.

• **Separation.** The cryogenically chilled stream is processed through a series of distillation columns. Several columns are needed to separate out the desired products. This process section consists of a demethanizer, ethane/ethylene and propane/propylene splitters, and a debutanizer, as shown in Fig. 3.

This study focused on revamping the propane/propylene (C_3) splitters to maximize recovery of propane and propylene with greater efficiency and reduced losses. Fig. 4 shows an in-depth description of the C_3 splitter section.

The deethanizer bottoms stream (at approximately 21.3 kg/cm^2g and 62°C) is split into parallel C_3 splitter systems—primary and secondary trains. Each parallel train consists of two splitter columns. The feed is distributed between the two systems. The primary C_3 splitter train receives 60% of the propane feed flow.

The primary train has 277 trays between the first column (124 trays) and second column (153 trays). Both columns use multi-downcomer trays. Feed enters the first column above tray 35 (tray 188 for the combined column) for the propane case, or above tray 51 (tray 204 for the combined column) in the mixed-feed case. The secondary train is configured and operated similarly to the first train with a total of 198 trays between the first column (88 trays) and the second column (110 trays). The secondary system has sieve trays. The feed enters the first column on tray 26 (tray 136 for the combined column) for the propane case or on tray 36 (tray 146 for the combined column) for the mixed-feed case.

The C_3 splitter system was designed to produce 99.6 mol% of propylene in the overhead stream. The bottom stream from the C_3 section is sent to the debutanizer column where the top product, containing propane and butane, is recycled to the pyrolysis furnaces. The heavier components are recovered as a C_5^+ pyrolysis gasoline stream. In the mixed-feed case, there are fewer heavier components to recover.

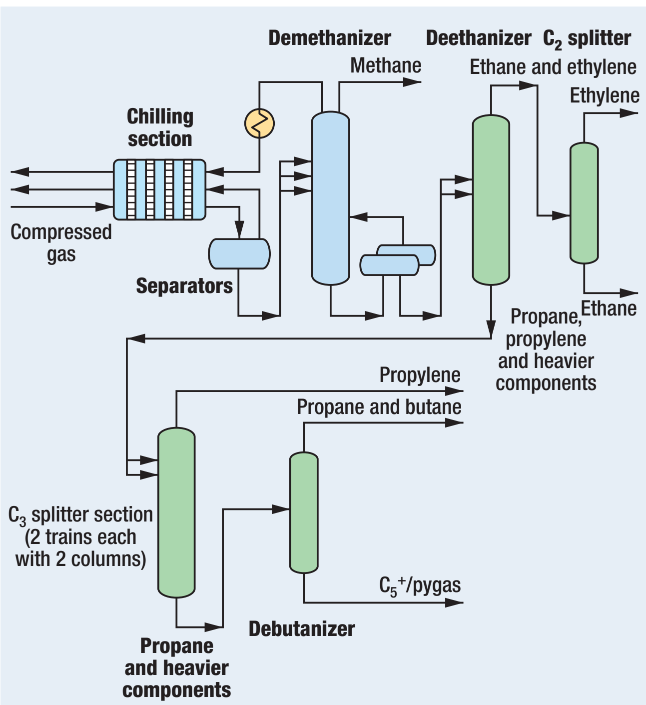

FIG. 3 Separation section of the Olefins I plant.

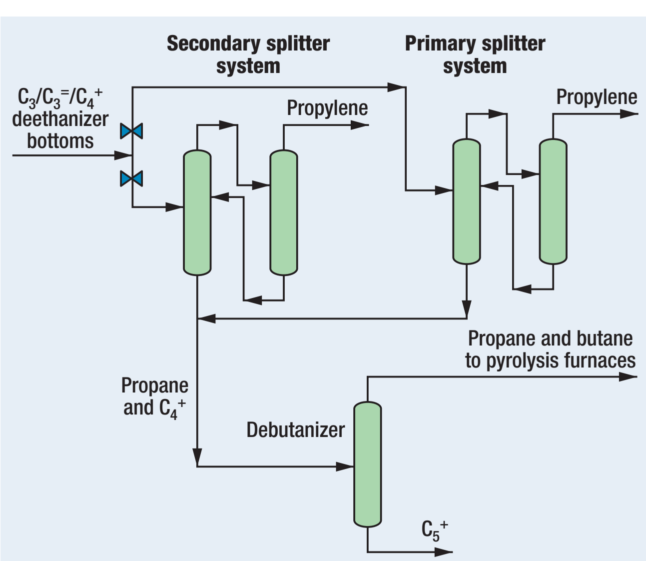

FIG. 4 Propane/propylene splitter section of the existing unit.

Plant operating problems. During 2005–2009, the propane/propylene system experienced several problems. Gradually,

TABLE 1. Results from the Revamp Proposal A simulation modeling study

	Column C (from secondary system) depropanizer	Primary propane/ propylene splitter[1]
Feed flowrate, metric tph	16.98	14.16
Stages	88	277
Feed stream stage	36	171
Distillate rate, metric tph	14.16	6.9
Mol purity propylene, top	0.484	0.998
Mol purity propane, top	0.513	0.001
Bottom rate, metric tph	2.8	7.3
Mol purity propane, bottom	0.0007	0.992
Top pressure, bar	18.9	18.9
Reflux rate, metric tph	22.7	146
Reboiler duty, MW	2.9	12.3

TABLE 2. Results from the Revamp Proposal B simulation modeling study

	Column D (from secondary system) depropanizer	Primary propane/ propylene splitter[1]
Feed flowrate, metric tph	16.9	14.2
Stages	88	277
Feed stream stage	36	171
Distillate rate, metric tph	14.2	6.9
Mol purity propylene, top	0.484	0.998
Mol purity propane, top	0.513	0.001
Bottom rate, metric tph	2.8	7.3
Mol purity propane, bottom	0.0007	0.992
Top pressure, bar	18.9	18.9
Reflux rate, metric tph	22	146
Reboiler duty, MW	2.85	12.3

the facility operating performance worsened. Performance issues included:

- High propylene loss, 25 mol% vs. design < 1 mol%
- Poor separation and high energy usage of the C_3 splitters
- Fouling in the splitter reboilers
- Low propylene and propane recovery, problems with the overhead-product purity and high concentration of unsaturates in the recycle propane to the pyrolysis furnaces.

These problems resulted in significant propylene loss that cumulatively amounted to more than 70,000 metric tons over five years. The lost products were valued at over $75 million. Fouling of reboilers due to using oily water as the hot utility, and coking of the transfer line exchangers from higher propylene content in recycle propane, contributed to higher maintenance costs.

Process simulation study. The objectives of the modeling were to:

- Understand the root causes for these problems
- Develop suitable and cost-effective solutions
- Provide ongoing guidance for troubleshooting
- Improve unit performance.

The simulation model was constructed from design data from the operating manuals and engineering drawings, as shown in Fig. 5. This model was tuned and validated against other data sets. This tuning included comparing different thermodynamic methods and selecting the best with respect to accuracy. The Peng-Robinson (PR) and Soave-Redlich-Kwong (SRK) models were used to describe thermodynamic behavior and equilibrium coefficients. Both methods are commonly used for hydrocarbon systems. For the Olefins I plant, Peng-Robinson provided an accurate fit with the design cases.

Several commercially available simulation programs were used to simulate the C_3 splitters while also considering the existing column geometries and tray efficiencies. This distillation model is a core element. It helped predict column performance and ensured robust initialization and convergence. The rate-based algorithm also significantly improved the model's accuracy compared to the equilibrium-based and first-generation rate-based distillation models.

Simulation results—such as column pressure, operating temperature, reflux ratio, composition, reboiler/condenser duties, column stages, feedrate, overhead and bottoms yield, and tray details—were specified to achieve 99.6% propylene recovery. Propane/propylene (principal products), isobutane, butanes, butenes and heavier components (traces) were also considered in this model. Once the model was tuned, it was used to study a series of conceptual design alternatives, including energy and economic analysis for the different proposals.

Revamp Proposals A and B. The first two options (A and B) were similar. They both involved reconfiguration and using one of the columns in the secondary splitter system as a depropanizer, while taking the other column out of service, as shown in Fig. 6. The simulation model showed that this approach would improve propane/propylene recovery and increase the recycle propane to

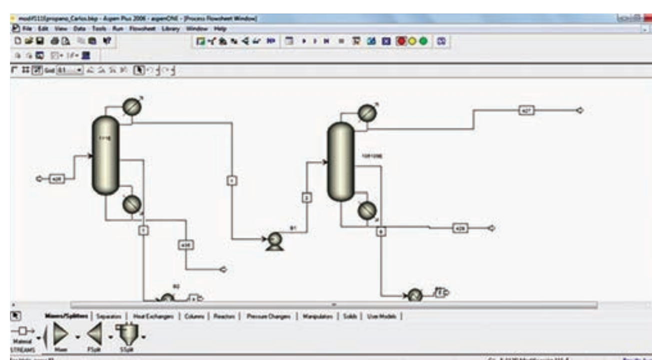

FIG. 5 Simulation model of the propane/propylene splitter system.

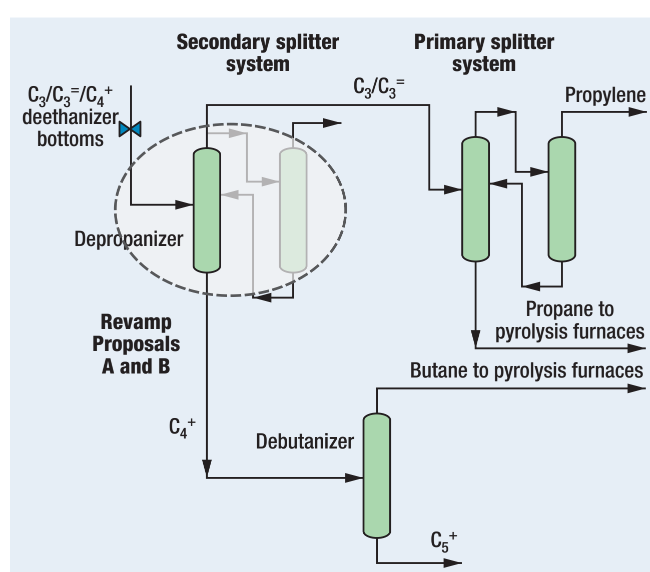

FIG. 6 Proposals A and B: Use one column in the secondary C_3 splitter as a depropanizer.

TABLE 3. Results from the Revamp Proposal C simulation modeling study

	Secondary propane/propylene splitter (depropanizer)	Primary propane-propylene splitter[1]
Feed flowrate, metric tph	17	14.4
Stages	198	277
Feed stream stage	37	172
Distillate rate, metric tph	14.4	6.8
Mol purity propylene, top	0.489	0.999
Mol purity propane, top	0.494	0.0002
Bottom rate, metric tph	2.6	7.6
Mol purity propane, bottom	0	0.996
Top pressure, bar	18.9	18.9
Reflux rate, metric tph	23	182
Reboiler duty, MW	2.9	60.9

TABLE 4. Results from the revamp proposal D simulation modeling study*

	Primary propane/propylene system LP steam	Secondary propane/ propylene system LP steam[1]
Required temperature, °C	128.7	128.7
Required pressure, bar	2.75	2.75
Propylene recovery composition	0.9985	0.9985
Required flowrate, metric tph	24,366	19,035
Total cost, $ million/yr	1.41	1.102

*Based on the original design with propylene losses of less than 1%

the pyrolysis furnaces. Proposal A studied using the first column as the depropanizer, and Proposal B looked at using the second column for this purpose. Tables 1 and 2 summarize the study results.

Findings for Proposals A and B. The operating conditions for Proposals A and B are similar to the original design. A depropanizer in the C_3 splitter system does increase propane and propylene recovery (about 99 mol%). Heating requirements are significantly reduced—15.2 MW vs. 22.8 MW from the original design. Jet flooding is 0.65 (well below the maximum jet flooding limit of 0.85). There is no evidence of overloading in the multi-downcomer trays, in spite of the high reflux rate requirements.[3]

Revamp Proposal C. This option considered using the entire secondary C_3 splitter system (both columns) as a depropanizer, as shown in Fig. 7. The objectives were to improve propane and propylene recovery and to increase recycle propane to the pyrolysis furnaces. Table 3 summarizes results from this processing option.

Findings for Proposal C. The operating conditions are similar to the original design. A depropanizer in the C_3 splitter system increases propane and propylene recovery (about 99.8 mol%). Additional heating is required—60.9 MW vs. 22.8 MW specified in the original design. The risk of jet flooding in multi-downcomer trays in the primary system was identified. Due to tray overloading, this process option was not pursued further.

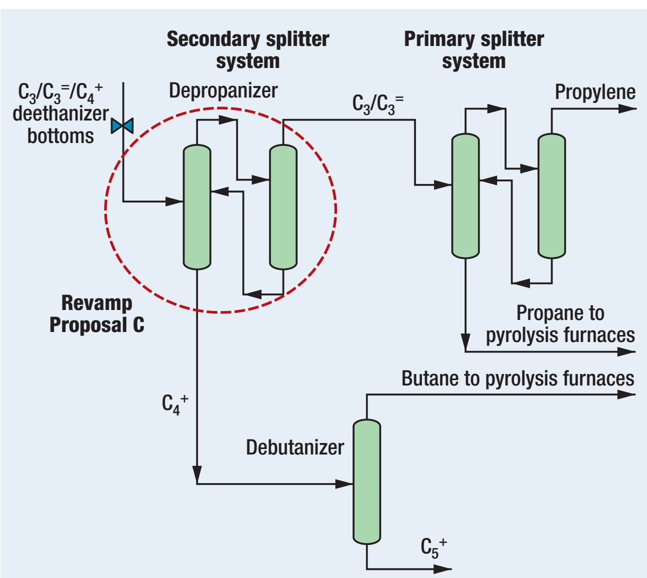

FIG. 7 Proposal C: Use the secondary splitter system as a depropanizer.

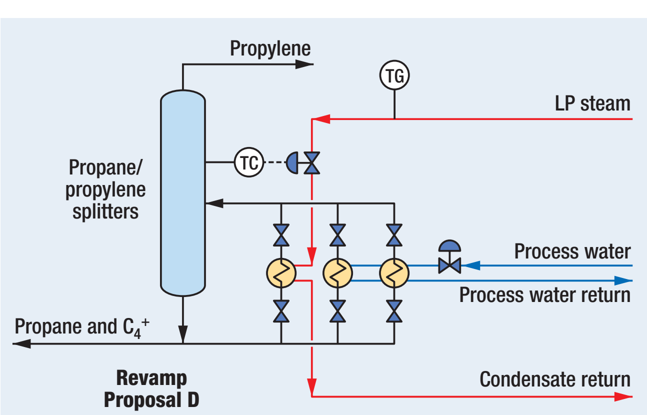

FIG. 8 Proposal D: Using LP steam as the heating medium in reboilers.

Revamp Proposal D. This option evaluated replacing oily water with low-pressure (LP) steam as the heating medium in the C_3 splitter reboilers. The change could reduce fouling on tube surfaces, as shown in Fig. 8. The conceptual design and analysis are based on revamping the original design for the most limiting conditions, as represented by the 100% propane feed case.[2] Table 4 lists the study results.

Annual steam costs are estimated at $1.41 million and $1.102 million, respectively, for the primary and secondary systems. The total steam consumption across the C_3 splitter system is approximately $2.51 million/yr.

Revamp Proposal D project costs. Option D not only addresses exchanger tube-side fouling and maintenance, but it also reduces propylene losses in the splitter bottoms. This will improve propylene recovery from the product and propane for recycle. The economics for this case were evaluated in detail. Table 5 summarizes cost estimates and project economics.

The total capital investment is estimated at $3.025 million, with an annual steam utility cost of $2.51 million as reported earlier. These process improvements are expected to result in 8,915 metric tpy of incremental propylene production. At $1,080/metric ton, the increased production represents $9.62 million of additional annual revenue. This is an excellent return on investment for the project. Switching to LP steam reduces exchanger fouling and enables easier cleaning and maintenance of the thermosiphon reboilers. Annual savings of $500,000 are expected from reduced cleaning and maintenance costs.

Lessons learned and other findings. The overview of the entire study raised several interesting findings:

Proposals A and B. This design delivers the best performance for the C_3 splitter system. The depropanizer aids in increasing product recovery (about 99 mol%) and improves operations for high-purity propylene (approximately 99.6 mol%). This design lowers heating requirements (15.2 MW vs. 22.8 MW for the original design). There is no evidence of overloading (flooding) in the

TABLE 5. Pequiven C_3 splitter revamp proposal D economic analysis[1]

Cost estimates	USD, thousand
Basic and detailed engineering	500
Reboiler modification, process oily water to LP steam	250
Condensate removal system	1,200
Stainless steel pipe, 16 in.	18.5
Stainless steel pipe, 14 in.	13.7
Stainless steel pipe, 12 in.	10.8
Stainless steel pipe, 2 in.	5.3
Isolation	2.8
Installation and manpower costs	1,024
Total	**3,025**
Total investment (CAPEX)	3,025
Steam utilities and operating costs (OPEX)	2,510
Propylene incremental annual production	9,620
% profitability, propylene recovered/CAPEX x 100	318%

multi-downcomer trays, even with high reflux rate requirements.

Proposal C. This alternative is not practical due to a high risk of tray flooding and higher energy requirements.

Proposal D. This design uses LP steam to meet reboiler duty requirements. The switch in heating medium provides easier cleaning and lowers maintenance time for reboilers. Greater recovery of propylene and increased purity of recycle propane are possible. This option improves furnace operations.

This study demonstrated that revamping the C_3 splitter system to use one of the columns from the secondary C_3 splitter as a depropanizer (Proposal A or B) results in propane recovery close to 100%. Heating requirements for the revamped system are lower, with easier cleaning and maintenance of reboilers. Propylene recovery would be 100% while the probability of tray flooding or weeping is low.

Simulation studies also indicated that it is not technically possible to use the primary splitter system or one of its columns as a depropanizer, and the second one as propane/propylene splitter. This arrangement risks overloading trays and has higher heating requirements and reflux rates compared to the original design.

Optimization study. The results from this simulation and engineering study show that Proposals B and D are the optimal revamp alternatives. They deliver improved operability and performance for propane/propylene separation with lower duty requirements, better product recovery and purities and lower utilities and maintenance costs. These options would improve conversion and lengthen the service life for the furnaces, reboilers and distillation columns. However, due to budgetary constraints, only Proposal D is being implemented first—modification of reboilers from wash water to LP steam heating.

Pequiven is executing the project. The scope includes further developing the conceptual design, basic engineering and Class 4 cost estimates (± 20%). Project duration is expected to be around 24 months. When completed, this revamp will deliver 8,915 metric tpy of incremental propylene product valued at $9.62 million/yr, and additional annual savings of $500,000 through reduced reboiler cleaning and maintenance costs.

The process simulation and conceptual estimates in this study were invaluable. Both helped Pequiven gain clearer insight into its olefin plant operations. With such information, Pequiven was able to develop a better understanding of plant and equipment performance problems. **HP**

ACKNOWLEDGMENT

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