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RETROFIT OF NATURAL GAS TURBOEXPANSION PLANTS TO WORK IN DUAL MODE

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Abstract

A MINLP optimization model has been developed to retrofit an ethane extraction plant to work in both ethane production mode and ethane rejection mode. The process model is solved within a sequential process simulator that embeds different natural gas plants designs. Based on an initial Basic Turboexpansion Process, the program determines the optimum process technology that maximizes either ethane or propane recovery and its associated operating conditions.

1. Introduction

Natural gas processing plants are generally based on a cryogenic turboexpansion process. The historically cyclical nature in the market for ethane and propane has demonstrated the need for flexible natural gas liquids recovery plants; i.e., that can be able to work in either ethane or propane recovery mode. The basic turboexpansion process can hardly fulfil this condition and the same applies to propane. However, significant improvement can be achieved by introducing minor plant structural modifications. The presence of discrete decisions to account for different process configurations leads to a Mixed Integer Nonlinear Programming (MINLP) optimization model.

Diaz *et al.* (1997) have studied the design problem for ethane extraction plants and the debottlenecking of existing plants with different optimization objectives. Similar process technologies can be used to obtain propane as the main product and re-inject ethane to pipeline.

The previously developed MINLP model (Diaz *et al.*, 1996) has been extended to embed new natural gas processing schemes where continuous variables represent operating conditions and binary variables are associated to potential units or designs. Equality constraints, which include process model and design equations, are solved within a simulator. Inequality constraints represent equipment capacities, process specifications and carbon dioxide precipitation conditions in the demethanizer column. The optimization program is an extension of the Outer Approximations algorithm (Duran and Grossmann, 1986) that is able to interface the process simulator.

Several process designs proposed in literature have been embedded within the superstructure, but not many of them are suitable for working in both ethane production and rejection modes; or if they do so, considerably lower recoveries are obtained.

As the goal of the optimization task is to determine the best process design that can work in either ethane production or ethane rejection mode, two different objective functions have to be dealt with, maximization of ethane or propane production, with different process specifications. In the first case, a maximum value for the methane/ethane ratio in the bottoms of the demethanizer is imposed. In the second case, the specification is a maximum value for the ethane/propane ratio in the same process stream (the column is a deethanizer in this mode).

At a first step, the program has been run to maximize ethane production. As a second step, the maximization of propane production was performed. Finally, it is important to note that the optimal retrofit design for an existing natural gas plant and its operating conditions have been determined through the application of MINLP techniques associated to a rigorous simulation program. The new design maximizes product recoveries while maximizing operating flexibility.

2. Turboexpansion Processes

Turboexpansion plants have been widely studied since the first plant was built in 1963. Wilkinson and Hudson (1982, 1993) have proposed different turboexpansion plant designs to improve ethane recovery without inlet CO_2 removal. Among them, we can mention the gas subcooled process, the gas-liquid subcooled process, the liquid subcooled process, the overhead recycle and so on.

In a typical ethane extraction plant, inlet gas is filtered and compressed. Thereafter it is air cooled and dehydrated to avoid ice and hydrates formation. After this conditioning, the gas feed is divided into two equal streams, each of which is sent to a different cryogenic train for demethanization. The bottom product from the demethanizers are mixed and sent to a conventional separation process to obtain pure ethane, pure propane, butanes and natural gasoline. After heat exchanging with the entering gases, the top product from the demethanizers, i.e. mainly methane, is recompressed to pipeline pressure and delivered as sales gas.

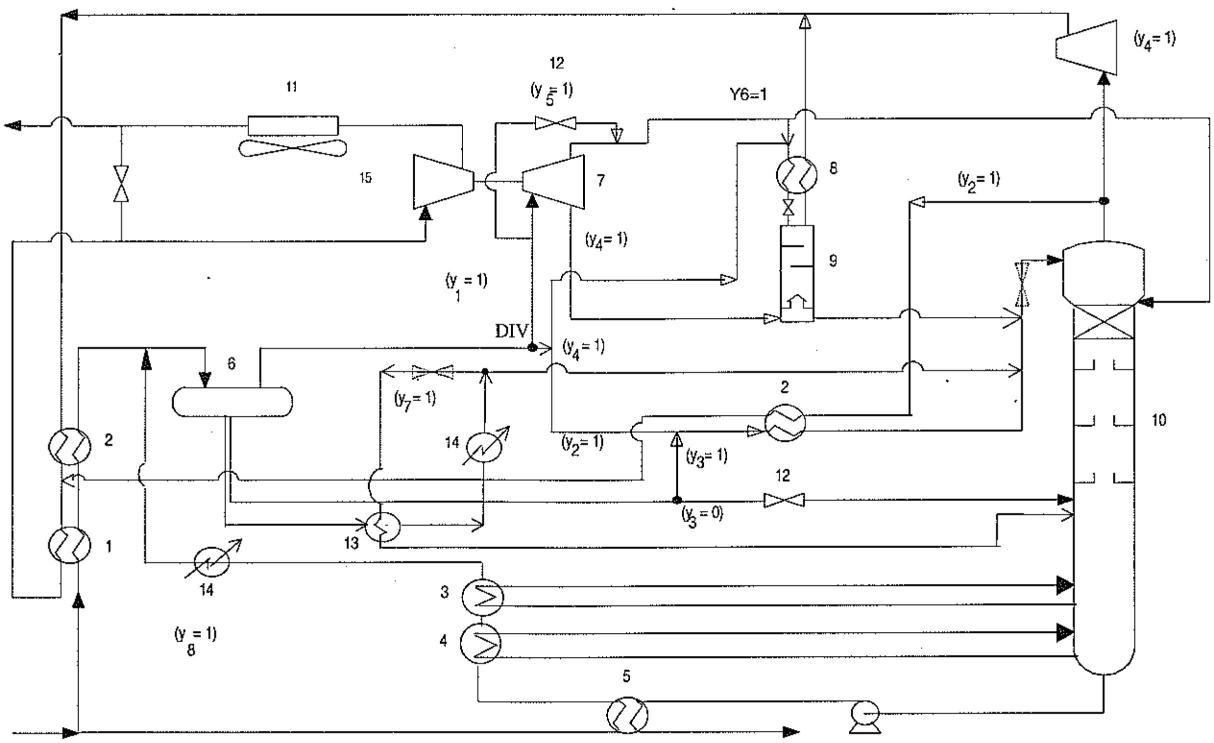

Figure 1 Cryogenic sector superstructure: 1, gas-gas heat exchanger; 2, gas-gas exchanger (subcooler in gas subcooled design); 3, demethanizer side reboilers; 4, bottom reboiler; 5, demethanized product exchanger; 6, cold tank; 7, turboexpander; 8, subcooler; 9, high pressure column (8 and 9 only for two-stage demethanization); 10, demethanizer; 11, air coolers; 12, JT valves

In a propane extraction plant based on turboexpansion (a dual mode turboexpansion plant), the previously described demethanizer columns work as deethanizers and both methane and ethane are re-injected to the pipeline.

The cryogenic sector of a Basic Turboexpansion Process (BTEP) is shown in Fig. 1 in solid lines. The inlet gas is cooled by heat exchange with the residual gas and demethanizer side and bottom reboilers and, if necessary, with external refrigeration. The partially condensed gas feed is then sent to a high pressure separator (cold tank). The vapor is expanded through the turboexpander to obtain the low temperatures required for high ethane recovery and is fed to the top of the demethanizer column. The liquid from the cold tank is directly flashed into the demethanizer at its lowest feed point. Methane and lighter components, such as nitrogen, constitute the top product and ethane and heavier hydrocarbons comprise the main components in the bottoms. Carbon dioxide, which is intermediate in volatility between methane and ethane, is distributed into top and bottom streams. The top product or residual gas cools the inlet gas and is then recompressed to pipeline pressure and delivered as sales gas. The demethanizer bottom product can be further fractionated to produce pure ethane, propane, butanes and natural gasoline. It is important to note that compression and refrigeration constitute the main factors that regulate both capital and operating costs in this type of plant.

In gas subcooled process, a fraction of the vapor from the cold tank is condensed and subcooled by heat exchange with residual gas coming from the demethanizer; the subcooled liquid is then flashed and fed to the top of the demethanizer column, as shown in Fig. 1 for y_2=1. The remaining vapor is expanded through the turboexpander and fed to the middle of the column.

In gas-liquid subcooled process, the vapor that comes from the cold tank is also divided into two streams, one is expanded through the turboexpander and the other is mixed with the liquid stream from the cold tank, as is shown in Fig. 1 for y_3=1. This stream is then condensed and subcooled by heat exchange with residual gas and is fed to the top of the demethanizer.

In a two-stage demethanization process, a major fraction of methane is removed in a predemethanizing column that operates at a higher pressure. The bottoms from this column are sent to a column that operates at a lower pressure, where complete demethanization is achieved. A high ethane recovery is obtained because the demethanizer can work at a lower pressure; there is also a lower energy consumption to recompress the residual gas as about 80% of feed gas constitute the top stream from the higher pressure column.

In the Over Head Recycle, there is also a two stage column scheme in the cryogenic sector, the smaller is an absorber. The entire vapor stream from the cold tank goes through the turboexpander and the expanded vapor enters the bottom of the absorber. The top stream of the demethanizer is subcooled by heat exchange with the top stream of the absorber and enters the absorber.

3. Process Model

The natural gas processing plant model is formulated as a MINLP model where continuous optimization variables are cold tank temperature, demethanizer column pressure and bottoms flowrate and subcooled fraction in alternative designs. Binary variables are related to process configuration. Nonlinear equations constitute the plant mathematical model and are solved within an ad-hoc simulator. Inequalities are process specifications and bounds on equipment capacities and there are also logical conditions.

The simulator embeds a superstructure with different turboexpansion plant designs. At this stage, binary variables have been used related to the different plant flowsheets, shown in Fig. 1, they include basic expander process (y_1=1, flowsheet in solid lines), gas subcooled process (y_2=1), gas-liquid subcooled process (y_2=1,y_3=1), two stage demethanization process (y_4=1), Joule-Thomson expansion instead of turboexpansion (y_5=1), liquid subcooled process (y_7=1), over head recycle (y_6=1), addition of mechanical refrigeration (y_8=1).

Previous work (Diaz *et al.*, 1996) has focused attention on the cryogenic sector to maximize ethane production. In the present paper, new process schemes have been included in the superstructure and the maximization of propane production has been performed while re-injecting ethane to pipeline.

Process specifications and purity requirements are handled as nonlinear constraints for a design problem. For debottlenecking problems and operating optimization, equipment limitations are also taken into account.

4. Solution strategy

The optimizer is an implementation of the Outer Approximation algorithm (Duran and Grossmann, 1986) that can be integrated to a process simulator. The basic algorithm requires successive solution of nonlinear programming (NLP) subproblems and Mixed Integer Linear Programming (MILP) problems that overestimate the feasible region and overestimate (for maximization problems) the objective function. The interaction between simulator and optimizer takes place at the NLP stage; a detailed description of this interface is given in Diaz and Bandoni (1996). For convex problems, the algorithm guarantees convergence to global optimum. For maximization problems, NLP solutions are lower bounds on the original MINLP solution (since they do not correspond to the optimal configuration) and MILP solutions are upper bounds for convex problems. Convergence occurs when both bounds cross. The use of a black box simulator for function evaluation does not guarantee problem convexity and outer approximations may cut off parts of the feasible region and converge to locally optimal solutions. However, results indicate that the quality of nonconvex solutions greatly improve the objective function. In this work, the NLP subproblem has been efficiently solved with subroutine OPT (Biegler and Cuthrell, 1985) and the MILP problem, with program LINDO (Schrage, 1987).

5. Discussion of Results

As a first step, the design problem of a dual mode plant has been studied; i.e., no equipment limitations have been taken into account. The MINLP problem has only purity specifications and operating bounds as nonlinear constraints. In the second part, the entire retrofit problem has been solved, including existing equipment limitations as nonlinear constraints. Moreover, a comparison of main designs variables is given for both ethane production and rejection modes.

5.1 Dual plant: Design problem

The design model for a natural gas plant to maximize propane recovery, while rejecting ethane, is subject to the following nonlinear constraints: a) Purity specifications: Ethane/Propane (c2/c3) ratio in demethanizer bottoms and carbon dioxide mole fraction in residual gas; b) Temperature crosses in gas-gas heat exchangers and subcooler and c) Carbon dioxide precipitation conditions. Feed compositions are shown in Table 1.

Table 1 Feed composition

Component	% molar
Nitrogen	1.44
CO_2	0.65
Methane	90.43
Ethane	4.71
Propane & heavier	2.77

In ethane rejection mode, no heat integration occurs between inlet gas and NGL or column reboiler due to high temperatures in the column bottoms.

The initial configuration (Basic Turboexpansion) optimum is reported in Table 2 and it is compared to the MINLP optimum (Over Head Recycle). Propane recovery can be increased from 82.47% (NLP optimum in BETP) to 98.24%.

The Gas Subcooled Process is a suboptimal scheme for a natural gas plant working in ethane rejection mode, but it has been further analyzed in this work because it is the optimum technology for the maximization of ethane recovery (Diaz *et al.*, 1997). Table 3 shows the optimum operating conditions and product recovery for both ethane production and ethane rejection modes. Temperature profile in the column is quite different when working in one mode or another; only in ethane production mode a heat integration can be done between inlet gas and NGL (natural gas liquids) and with the bottom reboiler.

Table 2 Propane recovery and main variables at the NLP optimum at initial configuration (BTEP) and the MINLP optimum (OHRP)

Variable	BTE Optimum	MINLP optimum(OHR)
$T_{cold\ tank}$ (K)	220.80	223.67
P_{dem} (bar)	18.25	18.51
Bdem (kmol/h)	1003.36.	1111.989
Ttop (K)	192.82	224.63
Tbottom (K)	348.91	348.57
Ethane rec.(%)	0.60	0.46
Prop. Rec.(%)	82.47	98.24
c2/c3 bottoms	0.02	0.02
Binary	10000000	00000100

Table 3 GSP design in ethane and propane production mode. Comparison of product recoveries and main optimization variables

Variable	GSP optimum Ethane production	GSP optimum Ethane rejection
$T_{cold\ tank}$ (K)	218.84	230.00
P_{dem} (bar)	18.00	18.25
Subcooled Fraction	0.30	0.22
Ttop (K)	168.02	187.77
Tbottom (K)	277.50	348.96
Ethane rec.(%)	92.40	0.73
Prop. Rec.(%)	99.15	95.34
c2/c3 bottoms	0.02	0.02
Binary	01000000	01000001

5.2 Dual plant: Debottlenecking problem of a BTE plant

The retrofit of an existing ethane extraction plant has been studied to operate it either in ethane production or rejection mode. In this case there are additional nonlinear constraints that take into account equipment limitations; they are listed below and their bounds are shown in Table 4.

a) Heat capacity in inlet gas heat exchangers (1 and 2): The product of heat transfer coefficient times transfer area must be within equipment limitations
b) Liquid and vapor flowrates in key trays of the demethanizer must be lower than flooding values.
c) Available horsepower in compression trains, as existing gas turbines can provide a certain power.
d) Carbon dioxide precipitation conditions.

When the plant is working in ethane rejection mode, there is no heat integration between natural gas liquids (NGL) and inlet gas; and between the column reboiler and inlet gas due to higher temperatures in column bottoms.

Table 4 Description of nonlinear constraints for the retrofit of an existing plant to work in dual mode

Equipment	Variable	Bound
Demethanizer/ Deethanizer	C2/C3 Bottoms	0.02
Demethanizer/ Deethanizer	CO_2 Mole Fraction Top	0.02
Demethanizer/ Deethanizer	Liquid /Vapor Flowrate	Flooding Values
Demethanizer/ Deethanizer	CO_2 Concentration	Solubility Values
Gas-Gas HE	ΔTmin	10 K
Subcooler	ΔTmin	10 K
Gas-Gas HE	UA (kcal/h)	2800000
Compressor/ Recompressor	HP from Turbines	24000.

Hydraulic limitations on the demethanizer column, which have been taken into account as nonlinear constraints, showed that the existing demethanizer could not allow the higher vapor flowrates that result from changing the turboexpander outlet feed to a lower point. Consequently, the Over Head Recycle process, schematically shown in Fig. 2, with an additional absorption column that works as rectification section, fulfils process specifications while satisfying equipment limitations.

Table 5 Optimal debottlenecking scheme and operating conditions for dual mode plant

Variable	BTE optimum	MINLP Optimum (OHR)
Pcold tank (bar)	57.7	57.7
Tcold tank (K)	227.54	224.98
Pdem(bar)	18.74	18.45
Bdem(kmol/h)	915	1102
Qreboiler(kcal/h)	2300000	4674580
Ethane rec.(%)	0.53	0.473
Prop. Rec.(%)	70.42	97.02
c2/c3 bottoms	0.02	0.015
Binary	0100010	0000011

The Gas Liquid Subcooled Process (GLS) has also been analyzed, it was determined (Diaz *et al.*, 1997) that it is the optimum design for the retrofit of an existing plant to work in ethane production mode. This process performance to work in both ethane production and rejection modes is reported in Table 6.

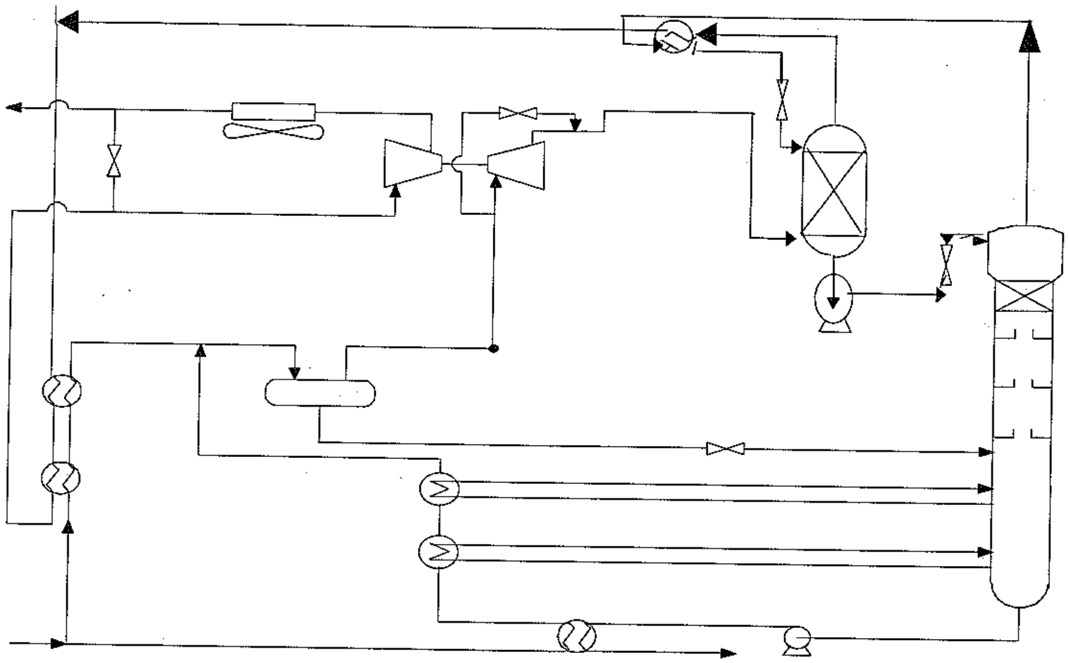

Figure 2 Over Head Recycle process proposed for retrofit of an existing plant to work in ethane rejection mode

Numerical results show that the GLS process is the optimum plant configuration to maximize ethane production and OHR is the MINLP optimum to maximize propane production while rejecting ethane. The optimal dual mode plant must be able to switch from one design to another depending on the economic target. This switch-over is simple between the OHR and the GLS process and it can be done even while the plant is in operation.

Table 6 Performance of Gas Liquid Subcooled process in both ethane production and rejection mode.

Variable	GSLP optimum Ethane production	GSLP optimum Ethane rejection
$T_{cold\ tank}$ (K)	218.84	230.00
P_{dem} (bar)	18.00	18.25
Subcooled Fraction	0.30	0.22
Ethane rec.(%)	92.40	0.73
Prop. Rec.(%)	99.15	95.34
c2/c3 bottoms	0.02	0.02
Binary	01000000	01000001

6. Conclusions

We have studied the retrofit of an existing ethane production plant to increase ethane production and to be able to work in ethane rejection mode. We have integrated a rigorous process simulator to an optimization algorithm and we have solved several MINLP problems to account for different economic scenarios. If ethane demand is high, the plant should work in ethane production mode; the objective function is the maximization of ethane recovery subject to ethane purity specifications, carbon dioxide precipitation conditions (due to low demethanizer column temperatures) and heat integration between column reboilers and inlet gas and NGL and inlet gas. If ethane demand is low, the plant should work in ethane rejection mode and the objective function is now the maximization of propane recovery; there are propane purity specifications, but no heat integration is performed between inlet gas and NGL or bottom column reboiler due to high temperature profile in the column.

The optimization algorithm integrated to a simulation routine has proved to be a powerful tool to tackle different optimization goals and it allowed the analysis of varying sets of constraints that represent different processes and specifications.

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