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Dynamic Modeling of a Deethanizer Column in a Natural Gas Processing Plant

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Abstract. The natural gas industry is of great strategic and economic importance and has become one of the most attractive business opportunities in the petroleum and petrochemical fields. There is a need to produce high quality gas, to reduce product rejection rates and to comply with prevailing laws of environmental and occupational safety. To improve the economics, flexibility, operability, and safety of column-based separation processes, it is important to know the steady-state as well as dynamic behavior of the process. This work aims to develop a dynamic model of the natural gas liquid (NGL) deethanizing process at a natural gas recovery unit using the EMSO (Environment for Modeling Simulation and Optimization) process simulator. Based on available information a dynamic model is developed and model validation is carried out through comparison between real plant data and results and predictions of the dynamic model. Furthermore, simulations are done to determine dynamic responses to process disturbances and other fault sources, as the validated model will be used to simulate process faults and to develop a monitoring system in future works.

Keywords: simulation, dynamic model, column, natural gas processing.

1. Introduction

Mathematical models use a set of equations to describe and simulate a process, thus enabling experiments similar to those which could be carried out in real processes. Simulation allows the generation of scenarios from which it is possible to monitor the process of decision making, to conduct analysis and evaluations of systems and to propose solutions to improve performance. All these procedures can be applied to technical and/or economic parameters.

Distillation is one of the most popular methods of separation in petrochemical and chemical industries. Therefore the dynamic simulation of such a system enables a better understanding of its behavior after changing an input parameter (Haydary and Pavlík, 2009).

The purpose of this paper is to apply the EMSO (Environment for Modeling Simulation and Optimization) process simulator in dynamic simulations of a deethanizer column at a natural gas recovery unit, and to compare simulation results with real measured data. In the column a tray by tray distillation produces a gas product with low C3+ content and a liquid product with low C2 content called as NGL (natural gas liquid). As the gas and liquid column products are the plant's final products, this system is of great importance to the performance of the unit.

In future works, a monitoring system for the deethanizing process will be developed. A process model will be essential, especially if there are insufficient historical data of faults available, as happens in young plants.

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2. Modelling

In order to predict the dynamic behavior of the process a rigorous model for multi-component distillation was implemented in the equation oriented dynamic simulator EMSO (Soares and Secchi, 2003). The physical and thermodynamic properties were obtained from the thermodynamic package VRTherm (VRTech, 2005). The model for each tray is described by the following equations (Staudt *et al.*, 2007):

Molar balance

$$\frac{dM}{dt} = F_{in}.z + F_{in}^{l}.x_{in} + F_{in}^{v}.y_{in} - F_{out}^{l}.x_{out} - F_{out}^{v}.y_{out} \quad (1)$$

where the subscript *in* is used for inlet streams, the subscript *out* for outlet streams, and the superscripts *l* and *v* correspond to liquid and vapor phase, respectively. The feed, liquid, and vapor molar fraction are *z, x* and *y*, respectively.

Energy balance

$$\frac{dE}{dt} = F_{in}.h_{in} + F_{in}^{l}.h_{in}^{l} + F_{in}^{v}.h_{in}^{v} - F_{out}^{l}.h_{out}^{l} - F_{out}^{v}.h_{out}^{v} + Q \quad (2)$$

where *h* is the molar enthalpy, *Q* is the rate of heat supply and *Hr* is the reaction heat.

Molar holdup

$$M = M^{l}.x + M^{v}.y \quad (3)$$

Energy holdup

$$E = M^{l}.h^{l} + M^{v}.h^{v} - P.V_{tray} \quad (4)$$

Chemical equilibrium condition

$$\phi_{liq}.x_n = \phi_{vap}.y_n^* \quad (5)$$

where φ_{liq} and φ_{vap} are the liquid and vapor fugacity coefficient, respectively.

Hydrodynamic equations

$$F_{out}^{l} = 1.84.lw.\frac{\left(\frac{Level-(\beta.hw)}{\beta}\right)^{2}}{v_{liq}} \quad (6)$$

$$F_{in}^{v} = \frac{Ah}{v_{vap}}\sqrt{\frac{(P_{n+1}-P_n)-\rho_{liq}.g.Level}{\alpha.\rho_{vap}}} \quad (7)$$

where F^{l}_{out} is the liquid flow rate leaving the tray, *lw* is the weir length, *hw* is the weir height, β is the aeration fraction, *Level* is the liquid level in the plate, v_{liq} is the liquid molar volume, and, in Equation 7, F^{v}_{in} is the vapor flow rate entering the tray, *Ah* corresponds to the plate total holes area, v_{vap} is the vapor molar volume, α is the dry pressure drop coefficient, ρ_{liq} is the liquid density, *g* is the gravitational constant and $(P_{n+1} - P_n)$ is the tray pressure drop.

The Murphree efficiencies, E_{MV}, are considered known model parameters:

$$E_{MV} = \frac{y_n - y_{n-1}}{y_n^* - y_{n-1}} \quad (8)$$

where y_n is the molar fraction of vapor and y^*_n is the vapor molar fraction in thermodynamic equilibrium with the liquid phase.

The main uncertainties considered in this model are associated with the model parameters α, β and E_{MV}.

3. Column simulation

The deethanizer column is a full stainless steel tower with two trayed sections. The top section is called the absorber (or rectifying) section and contains 10 Sulzer trays. The bottom section is called the deethanizer (or stripping) section and contains 30 valve trays. In order to simplify simulation both sections are assumed as sieve trays.

Heat input to the tower is controlled at two points using side and bottom reboilers, and multiple side draws and returns are utilized to achieve the desired product recovery. This column is equipped with thermowells and temperature transmitters at various heights and a temperature indicator at the bottom liquid section provides an indication of the tower operation. Differential pressure transmitters are located across the upper and lower tower sections.

There are two feed streams, respectively at trays #10 (Feed 1) and #30 (Feed2). They are composed by carbon dioxide, nitrogen, oxygen, methane, ethane, propane, n-butane, isobutane, n-pentane, isopentane, n-hexane, n-heptane and n-octane. Approximately the same volumetric gas flow that is fed to the tower at tray #10 is withdrawn from the deethanizer tower overhead. A 100% flowrate is assumed and Peng-Robinson equation is the thermodynamic model used to calculate the liquid and vapour phase properties. Table 1 summarizes the main column specifications.

Table 1. Steady-state design specifications for the deethanizer column

Feed 1 molar composition (ethane)	0.0822
Feed 2 molar composition (ethane)	0.2026
Top molar composition (ethane)	0.0949
Bottom molar composition (propane)	0.4917
Feed 1 flow rate	4024.53 kmol/h
Feed 2 flow rate	314.37 kmol/h
Top flow rate	3965.69 kmol/h
Bottom flow rate	234.46 kmol/h
Feed 1 pressure	17.4 atm
Feed 2 pressure	17.5 atm
Feed 1 Temperature	198 K
Feed 1 Temperature	308 K
Rectifying stages	10
Stripping stages	30
Rectifying section column diameter	1.8 m
Stripping section column diameter	1.0 m
Reboiler duty	900kW

As a PID controller is employed in the dynamic simulation of this study, the controlled variable is converted into the normalized signal of 0–1.0 based on the transmitter range before it enters the controller. The signal, which varies in the normalized range from 0 to 1.0, is the observable evidence of a variation in the physical variable through the transmitter range. Similarly, the controller output signal, which varies in the range from 0 to 1.0, is converted into physical variable by the control valve and then influences the process. Such a dynamic simulation using a normalized signal and a feedback controller is considered to be more realistic for a practical situation.

For the numerical integration of the complete simulation a step size of 30 s is chosen. When the dynamic simulator is run for a long period (>12h) the steady-state operating conditions can be obtained. Upon reaching the steady state for the simulation system, the simulation time is reset to zero, i.e. t = 0, and dynamic tests are carried out. Since the major disturbances in this system come from Feed 1, two step changes are introduced, one at t = 1000 s and the other at t= 2500 s. They are:

Simulation 1

1. Feed 1 temperature + 5 K: the feed temperature changes from 198K to 203K.
2. Feed 1 temperature - 5 K: the feed temperature changes from 203K to 198 K.

Simulation 2

1. Feed 1 flow +10%: the feed flow rate changes from 4024.53 to 4426.98 kmol/h.
2. Feed 1 flow −10%: the feed flow rate changes from 4426.98 to 4024.53 kmol/h.

4. Results

Figure 1 shows the steady state temperature column profile from the simulation compared with real plant data available.

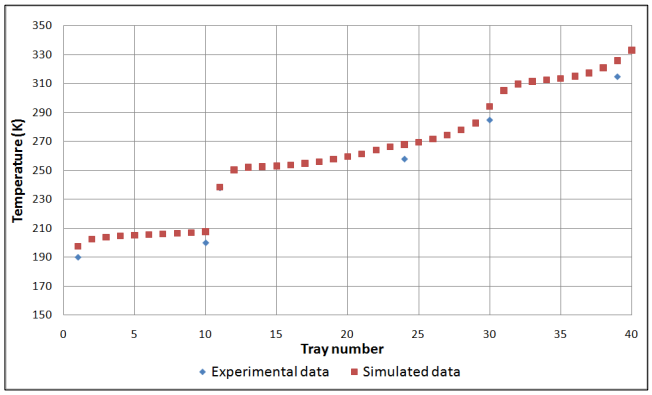
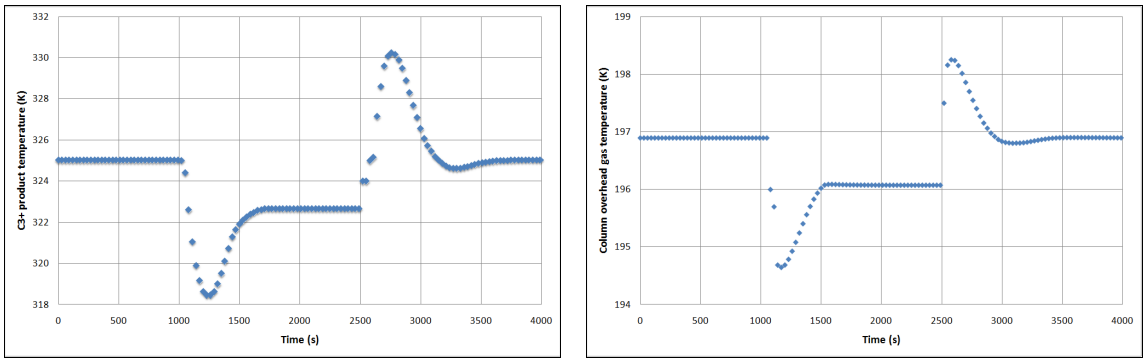


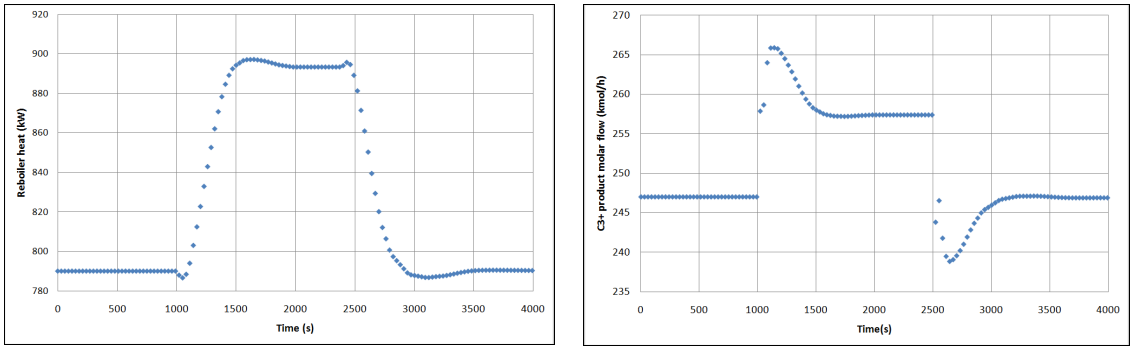
Figure 1 – Column temperature Profile.

The results of the dynamic tests can be seen in Figures 2 and 3, which are in agreement with the experienced on process operation.



(a) (b)

Figure 2 – C3+ product temperature (a) and overhead gas temperature (b).



(a) (b)

Figure 3 – Reboiler heat supply (a) and C3+ product flow rate (b).

Figure 2 shows temperature behavior of overhead gas and C3+ product after changes in Feed 1 temperature and Figure 3 shows how changes in Feed 1 flow rate disturb C3+ flow rate and reboiler heat. As can be seen, the steady-state results are compatible with plant data and the dynamics are well represented. This demonstrates that the developed model is quite representative of the process behavior.

5. Conclusions

In this work a dynamic simulation study of a deethanizer column is developed and implemented in the EMSO simulator. Simulation results are in good agreement with experimental data and dynamic behavior can be reproduced satisfactorily under various disturbances. Therefore this dynamic model can be considered validated, enabling its use in developing a monitoring system in future works.

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