International Journal of Science and Research (IJSR)
ISSN (Online): 2319-7064
Index Copernicus Value (2013): 6.14 | Impact Factor (2014): 5.611

Dynamics and Control Simulation of a Debutanizer Column using Aspen HYSYS

S. Karacan[1], F. Karacan[2]

[1]Ankara University, Engineering Faculty, Department of Chemical Engineering, Tandogan 06100, Ankara, TURKEY

[2]Turkish Atomic Energy Authority, Sarayköy Nuclear Research and Training Center, Istanbul Road 30 km., 06983 Saray, Ankara, TURKEY

Abstract: ***In this work, Aspen HYSYS as efficient computer-aided process engineering tools has been applied on a commercial refinery debutanizer column for the separation of an eight-component hydrocarbon mixture. This conventional distillation unit consists of 15 theoretical stages with a total condenser and a rebolier. The representative column is used to recover butane from an unstabilized naphtha feed having components C_2 to nC_8. The feed is fed at stage 5 (using Aspen HYSYS notation of numbering stages from the reboiler up to the condenser of the column). Both the conventional PID control and the Model Predictive Control (MPC) were applied to the simulation of the Debutanizer column. C_4 composition control for distillate product and temperature control for reboiler were applied to control the debutanizer column. The results show that including level into the MPC controller improves composition control for cases in which the manipulated variable for the reflux flow rate has a significant impact on compositions. Simulation results show that MPC controller perform better than the PID control.***

Keywords:, Debutanizer column, Simulation, Model Predictive Control

1. Introduction

Distillation, which is the workhorse of chemical process industries, is quite energy intensive and accounts for a large part of industrial energy consumption. It is reported that nearly 4% of the total energy requirement in the USA in 1988 is directed to distillation processes. It is a fact that energy consumption in distillation and CO_2 gases produced in the atmosphere are strongly related. The higher the energy demands are, the larger the CO_2 emissions to the atmosphere are. This is because the energy is mostly generated through the combustion of fossil fuel [1]. A debutanizer is a multicomponent distillation column frequently encountered in oil refineries. Debutanizer distillation column is usually used to remove the light components from the gasoline stream to produce Liquefied Petroleum Gas (LPG). This fractionating column has coupled and strong nonlinear dynamics. To maintain the product specifications, it is required to tighten process control, which is really a challenging task for control engineers. The refinery community has recognised the importance of the optimisation of process automation because of the benefits in terms of both profitability and tight control on product quality [2].

The implementation of many linear control strategies to maintain the product specifications of a debutanizer column is reported in literature [3]–[4]. Pashikanti and Liu [5] presented the methodology to develop, validate, and apply a predictive model for an integrated fluid catalytic cracking (FCC) process. They have implemented the methodology with Microsoft Excel spreadsheets and a commercial software tool. The methodology is equally applicable to other commercial software tools. Chun and Kim [6] investigated the design characteristic, cost evaluation and operation difficulty of the divided wall column (DWC) at its utilization in the floating liquefied natural gas (FLNG) plant. The DWC replacing the depropanizer and debutanizer of the conventional distillation system requires 12.5% less investment cost. Jana [7] studied A nonlinear feedback linearizing control (FLC) strategy within the differential geometric framework for temperature control of a refinery debutanizer column. The distillation model is verified by real data. The FLC control algorithm usually consists of a transformer, a state estimator and an external linear controller. Ahmedi et al [8] simulated industrial debutanizer column applying a steady state flow sheet simulator in order to investigate possible sources of low-efficiency separation problem.

In this article is used a distillation column simulated in HYSYS software. Both identification and control algorithm, developed in MATLAB. Both PID and MPC controls are applied to the process and compared.

2. Process Description

A typical conventional debutanizer column (CDBC) is shown in Fig 1 [9]. This conventional distillation unit consists of 15 theoretical stages, including a total condenser and a rebolier. The debutanizer column receives unstabilized naphtha feed having components ranging from C2 (ethane) to C8 (octane) from the crude distiller. This multicomponent distillation fractionates the naphtha such that the lights ends are removed from the top and the debutanized naphtha is removed from the bottom and directed to the splitter/platformer section for further processing. In the overhead section, the condenser liquid is directed to the liquefied petroleum gas section. A portion of the condensed liquid from the overhead is used as a reflux to the column whereas the reboiller provides the heat necessary to partially vaporize the debutanizer bottoms liquid before returning it to the column. The example debutanizer is detailed elsewhere (see [9] and the values of operating parameters and steady state information are reported in Table 1.

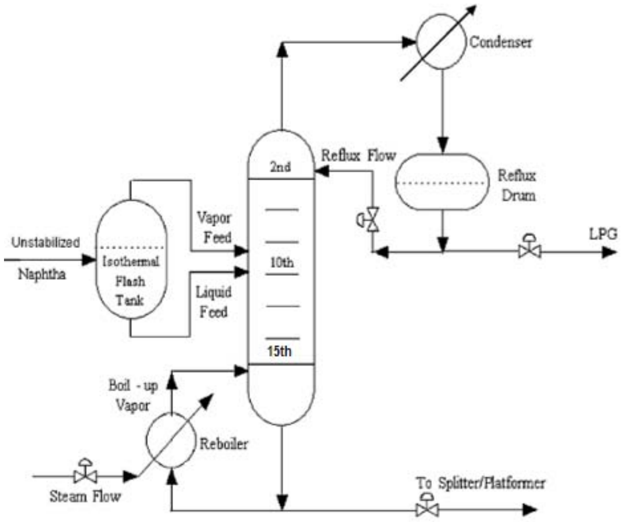

Figure 1: A simplified flow scheme of the refinery debutanizer column.

Table 1: Details of the conventional debutanizer column

Term	Value/condition	Unit
Feed (flashed out) condition	Two phase	-
Feed flow rate	399.42	kmol/h
Reflux ratio	1.4	
Overhead vapor (V_{top}) temperature	33.45	°C
Bottom liquid (L_{bot}) temperature	79.06	°C
Composition of flash drum feed		
C_2	0.00120174	-
C_3	0.0067598	-
iC_4	0.24079915	-
nC_4	0.3151570	-
iC_5	0.12167645	-
nC_5	0.10244855	-
nC_6	0.1315908	-
nC_8	0.0803665	

Aspen HYSYS [10] as efficient computer-aided process engineering tools has been applied on a commercial refinery debutanizer column for the separation of an eight-component hydrocarbon mixture. The flow diagram of this study is given in Fig.2. It consists of 15 theoretical stages with a total condenser and a rebolier. The representative column is used to recover butane from an unstabilized naphtha feed having components C2 to nC8. The feed is fed at stage 5 (using Aspen HYSYS notation of numbering stages from from the condenser down the column). Both the conventional PID control and the Model Predictive Control (MPC) were applied to the simulation of the Debutanizer column. C4 composition control for distillate product and temperature control for reboiler were applied to control the debutanizer column.

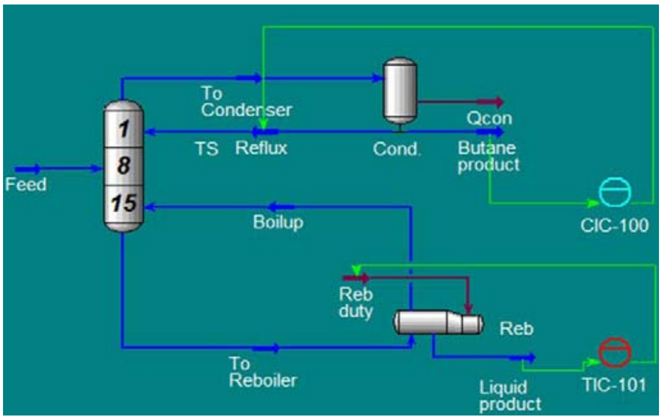

Figure 2: Debutanizer distillation column developed in HYSYS software

Table 2 below shows the specifications that are used to developed the HYSYS model for the debutanizer column. In addition, mole fraction of the component and molar flow rate for the distillate and the bottom product is given at the state-steady simulation results in Table 2.

Table 2: Specification and state steady results of the debutanizer column

Component	Mole Fraction		
	Feed	Distillate	Bottom
C_2	0.0012	0.0022	0.60e-07
C_3	0.0067	0.0125	0.34e-04
iC_4	0.2400	0.4277	0.0245
nC_4	0.3151	0.5143	0.0847
iC_5	0.1216	0.0344	0.2229
nC_5	0.1024	0.0089	0.2106
nC_6	0.1315	0.5e-04	0.2837
nC_8	0.0803	0.8e-07	0.1733
Type of condenser	T. Con.		
Feed Temp. (°C)	82.22		
Overhead Temp.(°C)	57.15		
Bottom Temp. (°C)	104.8		
Feed rate (kmol/h)	399.4		
Reflux rate (kmol/h)	340.2		
Distillate rate (kmol/h)	214.2		
Reboiler duty (kW)	2795		

The process variables chosen are concentration of nButane in butanes stream and temperature of bottom liquid product stream. The manipulated variables chosen are reflux flow rate (manipulating the setpoint of CIC-100) and reboiler heat duty (manipulating the setpoint of TIC-101 in Fig 2.

3. Simulation Results

3.1. Steady State Results

Firstly, the column has been simulated in HYSYS. Plant in the steady state mode. This case will be the steady state model for the debutanizer column. The entire column was divided into 15 stage excluding the condenser and the reboiler and its steady state study was carried out by simulating the prototype plant built using the simulator under the conditions of reflux flow rate of 340.2 kmol/h, the reboiler duty 2795 kW and the feed and the condenser pressure 7.4 atm. The other parameters used for the simulation can be found in Table 2. After the simulation, the temperature and composition profiles obtained are as shown in Fig 3 and 4 respectively.

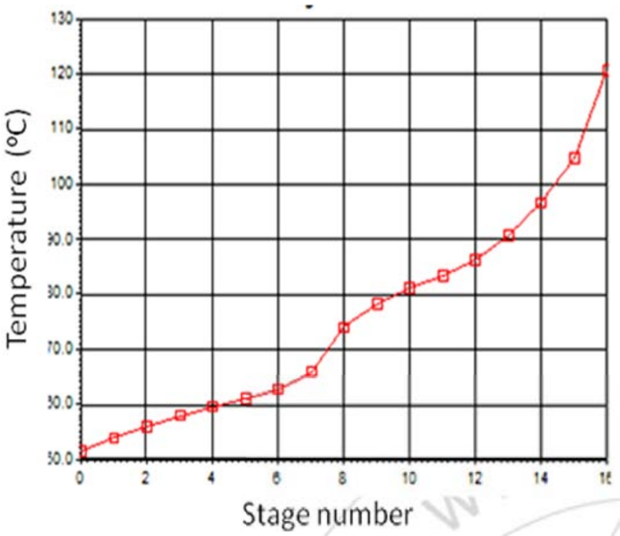

Figure 3: Temperature profiles with stage number of the distillation column

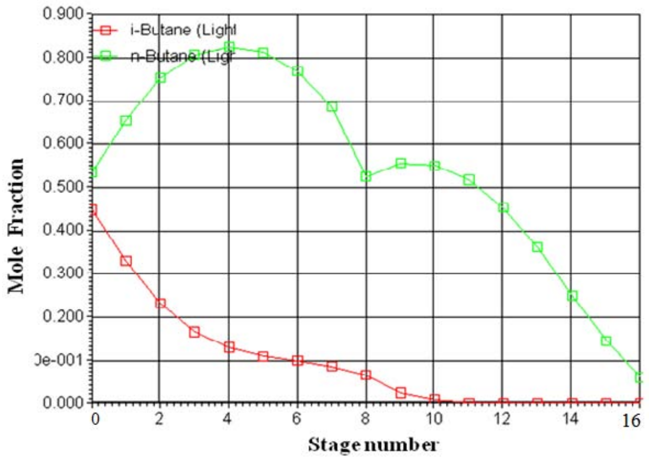

Figure 4: Composition profiles of the Butane with stage number of the distillation column

3.2. System identification procedure

With the two (2) inputs (reflux ratio and reboiler duty) and two (2) outputs (n Butane distillate composition and bottom temperature) chosen as the variables of this process, the results obtained from the dynamic simulation were used to develop the MIMO transfer function models of the process with the aid of System Identification Toolbox of MATLAB [11]. The necessary modifications have been made in order to build the dynamic simulation including the control mechanism which consists of concentration controller and reboiler temperature controller in HYSYS flow diagram. To determine process parameters Eq.(1-2) step test was applied. These results have been shown in Figs 5 and 6 below for the distilate composition of the nButane and the bottom temperature respectively.

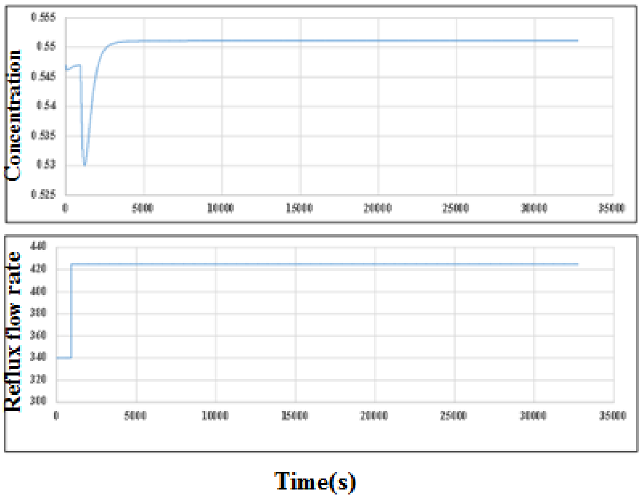

Figure 5: Dynamic simulation response of the distillate compositon of the nButane to step change from 340 kmol/h to 420 kmol/h reflux flow rate

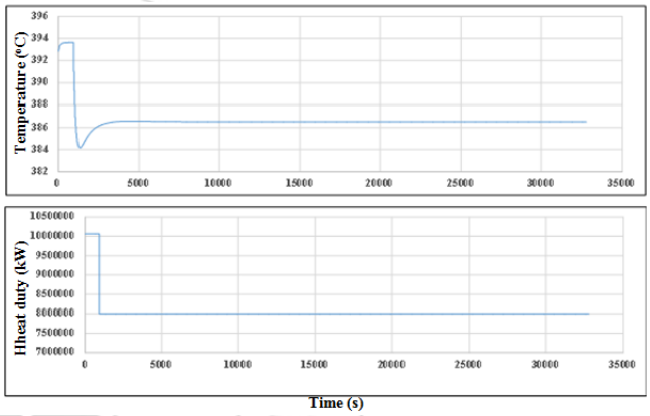

Figure 6: Dynamic simulation response of the bottom temperature of the column to step change from 10e+06 kW to 8.5e+06 kW reboiler heat duty

As can be observed from the results, the application of the inputs resulted in changes in the dynamic responses of the distillate composition of the nButane and and the bottom temperature. This is an indication that the distillate composition of the nButane and the bottom temperature were functions of the inputs. This is, of course, the reason for choosing the inputs as the manipulated variables of the control of this process. System Identification Toolbox was used for process model parameters using dynamic simulation results and model obtained are given in Eq. (1-2). The developed models were then simulated and their simulated results were shown in Fig. 5 and 6.

$$x_{D,nC4}(s)=\frac{5.0e^{(-0.58s)}}{1.041s+1}R(s)+\frac{4.6e^{(-0.28s)}}{2.12s+1}Q_R(s) \qquad (1)$$

$$T_B(s)=\frac{3.25e^{(-5.2s)}}{2.45s+1}R(s)+\frac{3.15e^{(-0.85s)}}{4.34s+1}Q_R(s) \qquad (2)$$

3.3. Tuning of Controllers and Results

Both the conventional PID control and the Model Predictive Control were applied to the simulation of the Debutanizer column. C_4 composition control for distillate product and temperature control for the bottom flow were applied to control of the debutanizer column. The controllers designed for the reactive packed distillation column were tuned using Ziegler-Nichols (Z-N) tuning methods [12], used the MIMO transfer function models developed for PID. With the transfer function of the PID controller given as, eq.(3) the relationships used for the calculation of the tuning parameters of the two techniques are as given in Table 3 below.

$$G_c(s) = K_c\left(1 + \frac{1}{\tau_I s} + \tau_D s\right) \tag{3}$$

Table 3: PID Control model parameters

Parameters	$xD_{,nC4}$-R loop (Z-N) tuning methods	T_B-Q_R loop (Z-N) tuning methods
K_c	16	22
τ_I (min)	7	4
τ_D (min)	2	1.5

Tuning MPC controllers After step tests were conducted for each of the two MPC configurations, step response models for the two MPC controllers were identified. MPC tuning parameters as foloows:

Number of inputs: 2; Number of outputs: 2;
Prediction horizon: 25; Control horizon: 2;
Gamma_U: 0.30; Gamma_Y: 0.25

All two controllers use a control interval of 1 min. The time to steady-state for step response models in all two controllers is 360 min.

The control strategy has been taken directly from original case study while the tuning of these controllers has been made using the PID autotuning function of HYSYS Plant. These tuning parameters are shown in Table 3.

A reference (deviation) of 4.6 % in relation of nC_4 concentration was applied . The process behaviour (read from HYSYS) is showed in Fig. 7 and 9 for PID and MPC, respectively. Other result, changing reference of bottom temperature was implemented (disturbing 12.4% in bottom temperature), is showed in Figure 8 and 10 for PID and MPC, respectively. The right axis of Figure 7 and 8 show the control effort (OP%).

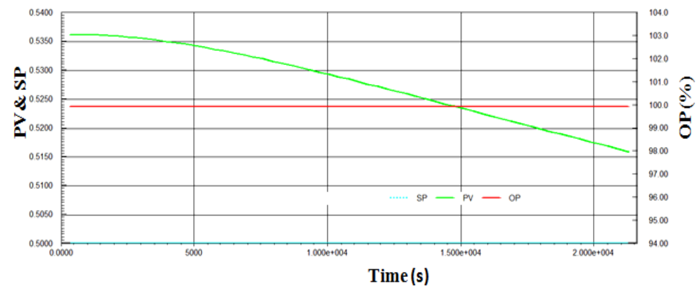

Figure 7: PID control composition responses to decrease set point of n-C_4(from 0.533 to 0.51)

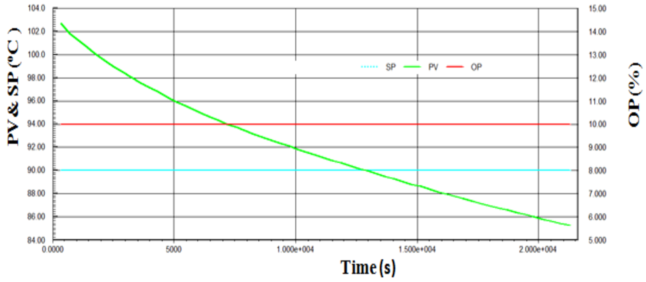

Figure 8: PID control temperature responses to decrease bottom temperature from 104°C to 90°C)

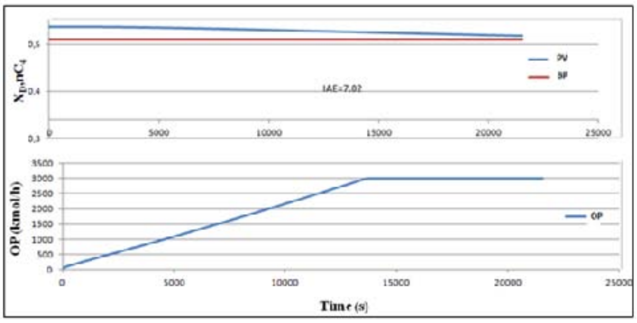

Figure 9: MPC control composition responses to decrease set point of n-C_4 from 0.533 to 0.51

Results show that using Multivariable MPC, time response was decreased around 80%. PID controller the process has response time of 360 minutes and responses were unstable. The responses of MPC controller reach to set point more quickly. Simulation results show that MPC controller perform better than the PID control.

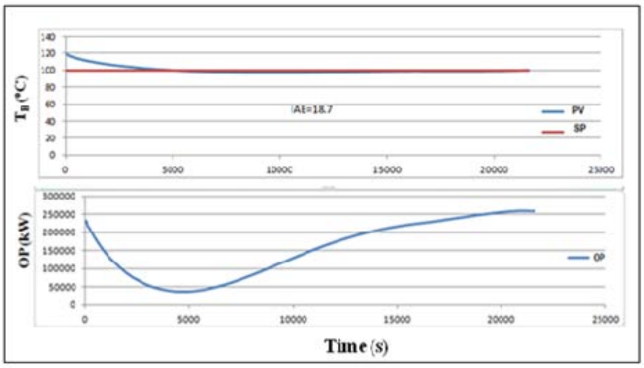

Figure 10: MPC control temperature responses to decrease bottom temperature from 120°C to 100°C

References

[1] J. L. Humphrey, I. I. G. E. Keller, Separation process technology. New York: McGraw-Hill (1997)

[2] N. M. Ramli, M.A.Hussain, B. M. Jan, B. Abdullah, "Composition prediction of a Debutanizer Column using Equation Based Artificial Neural Network Model,"vNeurocomputing, 131, 59-76 (2014)

[3] N. M. A. Jabbar and I. M. Alatiqi, "Inferential-feedforward control of petroleum fractionators: A PNA approach," Comput. Chem. Eng., 21, 255–262 (1997)

[4] E. N. Almeida, M. A. Rodrigues, and D. Odloak, "Robust predictive control of a gasoline debutanizer column," Brazilian J. Chem. Eng., 17, 967– 978 (2000)

[5] K. Pashikanti, Y. A. Liu, "Predictive Modeling of Large-Scale Integrated Refinery Reaction and Fractionation Systems from Plant Data. Part 2: Fluid Catalytic Cracking (FCC) Process," Energy & Fuels, 25, 5298-5319 (2011)
[6] H. H. Chun and . H. Kim, "Application of a divided wall column for gas separation in floating liquefied natural gas plant," Korean J. Chem. Eng., 30(7), 1473-1479, (2013)
[7] A. K. Jana, "A Hybrid FLC-EKF Scheme for Temperature Control of a Refinery Debutanizer Column,"IEEE Transactions on Industrial Informatics, 6, 25-35 (2010)
[8] A. Ahmadi, O. Dehghani, M. Heravi, M.R. "Rahimpour, Performance improvement and efficiency enhancement of a debutanizer column (a case study in South Pars gas field)," J. of Nat. Gas Sci. and Eng., 22, 449-61 (2015)
[9] A. K. Jana, Chemical process modelling and computer simulation (1st ed.). New Delhi: Prentice-Hall, 2008.
[10] Aspen HYSYS 3.2 installation guide. U.S.A: Aspen Technology (2003)
[11] Mathworks,. MATLAB, The Language of Technical Computing. The MathWorks Inc., Natick, Massachusetts, U.S.A (2011)
[12] G. Stephanopoulos. "Chemical Process Control: An Introduction to Theory and Practice", Prentice Hall (1984)

Author Profile

Dr. Süleyman Karacan received his Ph.D. degree from Ankara University, Turkey in 1997. He is a member of Union of Chambers of Turkish Engineers and Architects. He is currently working as a Professor at the Department of Chemical Engineering, Ankara University, Turkey. He is having overall teaching experience of 15 years. His major research interests are in Reactive Distillation, Modelling, Process Control, and Simulation.

Dr. Filiz Karacan received the B.S., M.S. and Pd. D. degrees in Chemical Engineering from Ankara University in 1992, 1997 and 2004, respectively. During 1995-2001, she stayed in Ankara University Chemical Eng. Dept. as a Research Assistant. Her major research interests are energy technology, optimization and Simulation.