Journal of Mining and Earth Sciences Vol. 56 (2016) 20-25

Journal of Mining and Earth Sciences

Website: http://tapchi.humg.edu.vn

Control configurations of distillation culumn

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ARTICLE INFO

Article history:
Received 12 Sep. 2015
Accepted 15 Jul. 2016
Available online 20 Nov. 2016

Keywords:
Control configuration
Distillation model
Disturbance rejection
Level control
Interactions

ABSTRACT

This paper presents some of the important aspects of the control configuration selection for continuous distillation column. The issues covered include level control, disturbances, and gain matrix for various configurations. The treatment is mainly limited to two-product distillation column separating relatively ideal binary mixtures.

1. Introduction

An important task in the design of the control system is to select the control configuration. Designing control configurations refers to all the decisions about the structure include: choosing variable input/output, measured variables, configuration control, controller.

From a control point of view, a two-product distillation column with a given feed, has five degrees of freedom including *L, V,* V_T, *D* and *B*. At steady-state, the assumption of constant pressure and perfect level control in the consender and reboiler, reduces the number of degrees of freedom to two (Shinar, 2007). These two degrees of freedom can then be used to control the two product compositions, x_B and y_D (or some others indicator of the composition, like the stage temperature). It is normal that *L* and *V* are used for the top composition and bottom composition control respectively in two-point control.

A typical two-product distillation column is shown in Figure 1. The most important notation is summarized in Table 1. Index *I* is used to denote the stage number (i =1 is for the top and *i = NT* is for the bottom of column).

In this paper, the C-02 distillation column of Dinh Co plant is considered as an example. The C-02 column data is given in Table 2 and Table 3 (Faanes, 2009)

2. Comparison of diferent control configurations

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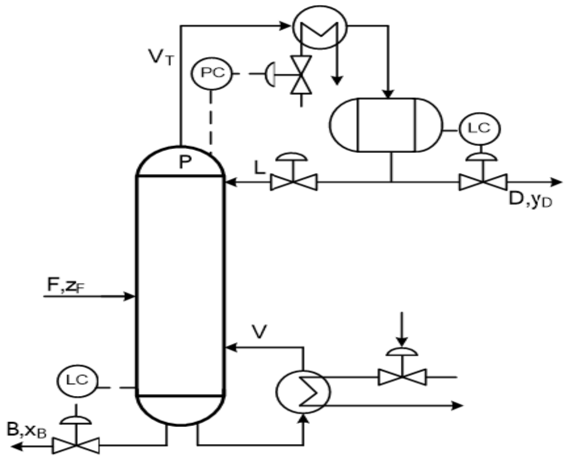

Figure 1. A distillation column controlled with LV-configuration

Table 1. Notation

F	Feed rate [kmol/min]
z_F	Feed composition [mole fraction]
q_F	Fraction of liquid in feed
D and B	Distillate (top) and bottoms product flowrate [kmol/min]
yD and xB	Distillate and bottoms product composition [mole fraction]
$L = L_T = L_{N..}$	Reflux flow [kmol/min]
$V = V_B = V_1$	Boilup [kmol/min]
N	Number of theoretical stages including reboiler
L_i and V_i	Liquid and vapour flow from stage i [kmol/min]
x_i and y_i	Liquid and vapour composition on stage i [mole fraction]
M_i	Liquid holdup on stage I [kmol]
α	Relative volatility between light and heavy component
τ_L	Time constant for liquid flow dynamic on each stage [min]
λ	Constant for effect of vapour flow on liquid flow

Table 2. C-02 data

Height:		24m
Diameter:		2.14m
The number of theoretical stage:		21
Designed Temperature:	MF Mode:	60^0C (top) and 142^0C (bottom)
	GPP Mode:	43^0C (top) and 154^0C (bottom)
Designed pressure		12.5Bar
Operating pressure:		11Bar
Volume:		$83m^3$

Table 3. C-02 operating data

NT	NF	F	Feed composition (%)				L	V	D	B
			C_2H_6	C_3H_8	C_4H_{10}	C_{5+}				
21	10	16.2	1	49.2	29.7	20	19.575	32.625	13.05	3.15

2.1. The model

The model is given by the Matlab code in Table 4. The states are the mole fractions of light component and the liquid holdup. Liquid flow dynamics are included here due to we do not assume constant holdup on the stages (Shinkey, 1994).

$$L_i = L0_i + \frac{M_i - M0_i}{\tau} + (V_{i-1} - V0_{i-1}).\lambda \quad (1)$$

Where *L0i [kmol/min] and M0i [kmol]* are the nominal values for the liquid flow and holdup on stage *i*.

2.2. Comparison of various control configurations

This section will consider the effect of a feed flow disturbance and the effect of level control on various control configurations, more specifically the *LV*, *DB*, *LB* and *(L/D)(V/B)* - configurations.

LV-configuration: An increase in feed rate makes the bottom flow increase which upset in the external material balance a effect on the product composition. But with no level control, the increase in *F* does not have a large effect on the compositions. In general, the column composition response is rather insensitive to actual holdup in the reboiler and consender holdups.

DB-configuration: it cannot be left without adjusting D and B on a long-term basis because otherwise we would fill up or empty the column.

(L/D)(V/B)-configuration: The increased feed rate results in a proportional increase in all streams in the column so the product composition remain almost unchanged.

LB-configuration: The increased feed rate results in D so the response is in the opposite direction of that for the LV-configuration.

Figure 2 shows the respond in top composition to a 1% increase in feed rate. The *LV*-configuration is almost independent of the level control tuning which is very important to other configurations. Take the *DV*-configuration as an example, consider the effect on product compositions of an increase in boilup *V* by 1%. Figure 3 shows response of top composition for a 1% increase in *V* with consender level controller $\Delta L = K\Delta M_D$. With fast consender level control, the increase in boilup goes up the column, but is then returned back as reflux through the action of the level controller and we have an increase in internal flows only. However, with a slow consender level controller, there is no immediate increase in reflux, so the initial response is almost as if we had send the boilup out the top of the column. This might causes an inverse response in product composition, which may make control difficult.

2.3. The relative gain array (RGA)

The control properties of the various configurations may be drastically different, and this is exemplified by studying the steady-sate two-way interations, as expressed by the relative gain array. The relative gain λ_{ij} expresses how the gain g_{ij} changes as we close the other loop(s).

$$\lambda_{11} = \frac{g_{11}}{g_{11} - g_{12}\left(g_{21}/g_{22}\right)} \quad (2)$$

$$G^{LV} = \begin{pmatrix} 0.0761 & -0.0761 \\ 0.0011 & -0.001 \end{pmatrix}; \qquad \lambda_{11} = 11$$

$$G^{DV} = \begin{pmatrix} -0.0761 & 0.0153 \\ -0.1413 & -0.0153 \end{pmatrix}; \qquad \lambda_{11} = 0.35$$

$$G^{(L/D)(V/B)} = \begin{pmatrix} 0.0175 & -0.0307 \\ -0.0149 & -0.043 \end{pmatrix}; \quad \lambda_{11} = 2.56$$

LV-configuration has the biggest interaction between control loops while *DV*-configuration has the smallest.

3. Conclusion

The main problem when selecting the “best” configuration for distillation control is that there are so many considered issues including level control problem and composition control problem

Due to the entered flow (of C-02) varies significantly so it makes the *LV*-configuration attractive (because it is almost independent of the level control tuning and then we can use *D* and *B* for level control for column). Besides, although there is a sensitivity to disturbances (*F*) and strong interactions between control loops, *LV*-configuration is still a suitable choice because it is very simple to implement and possible to achieve fast control by controlling instead two temperatures inside the column, as shown in Figure 4 and 5.

Table 4. Matlab code of dynamic distillation model

```
i=1:NT-1;
y(i)=alpha*x(i)./(1+(alpha-1)*x(i));
y(NT)=x(NT);
i=1:NT-1;
V(i)=VB*ones(1,NT-1);
i=NF:NT-1;
V(i)=V(i) + (1-qF)*F;
i=2:NF;
L(i) = L0b + (M(i)-M0(i))./taul + lambda.*(V(i-1)-V0);
i=NF+1:NT-1;
L(i) = L0 + (M(i)-M0(i))./taul + lambda.*(V(i-1)-V0t);
L(NT)=LT;
i=2:NT-1;
dMdt(i) = L(i+1) - L(i) + V(i-1) - V(i);
dMxdt(i)= L(i+1).*x(i+1) - L(i).*x(i) + V(i-1).*y(i-1) - V(i).*y(i);
dMdt(NF) = dMdt(NF) + F;
dMxdt(NF)= dMxdt(NF) + F*zF;
dMdt(1) = L(2) - V(1) - B;
dMxdt(1)= L(2)*x(2) - V(1)*y(1) - B*x(1);
dMdt(NT) = V(NT-1) - LT- D;
dMxdt(NT)= V(NT-1)*y(NT-1) - LT*x(NT) - D*x(NT);
i=1:NT;
dxdt(i) = (dMxdt(i) - x(i).*dMdt(i) )./M(i);
xprime=[dxdt';dMdt'];
```

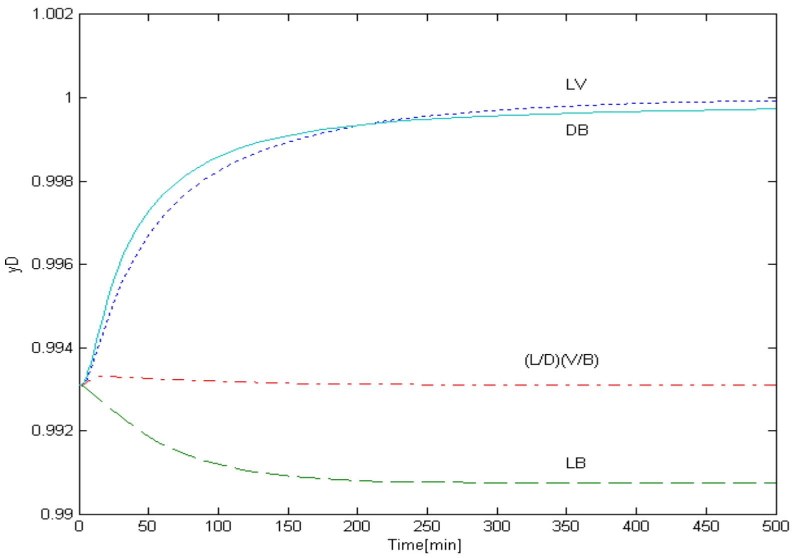

Figure 2. The respond in top composition to a 1% increase in feed rate

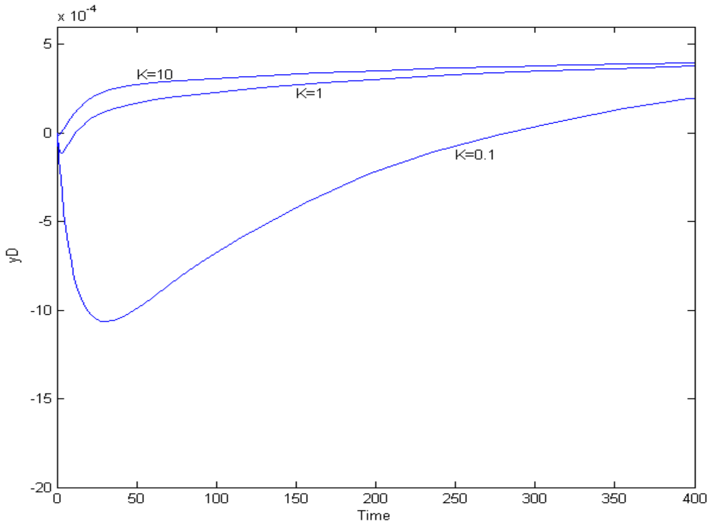

Figure 3. Response of top composition for a 1% increase in V with consender level controller $\Delta L = K\Delta M_D$

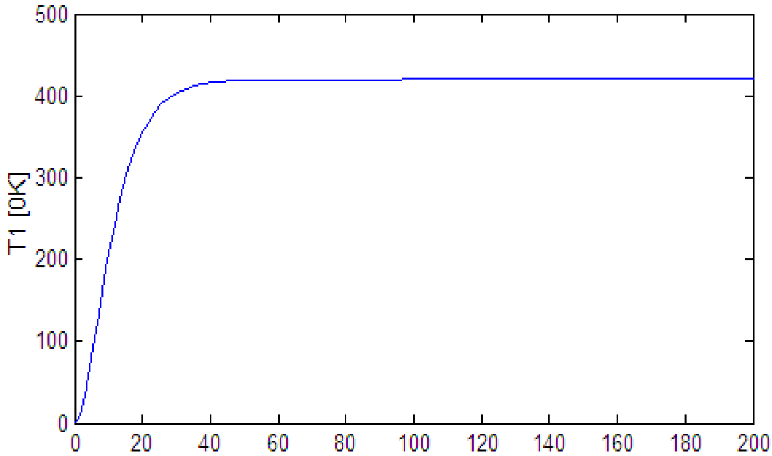

Figure 4. Response of top temperature

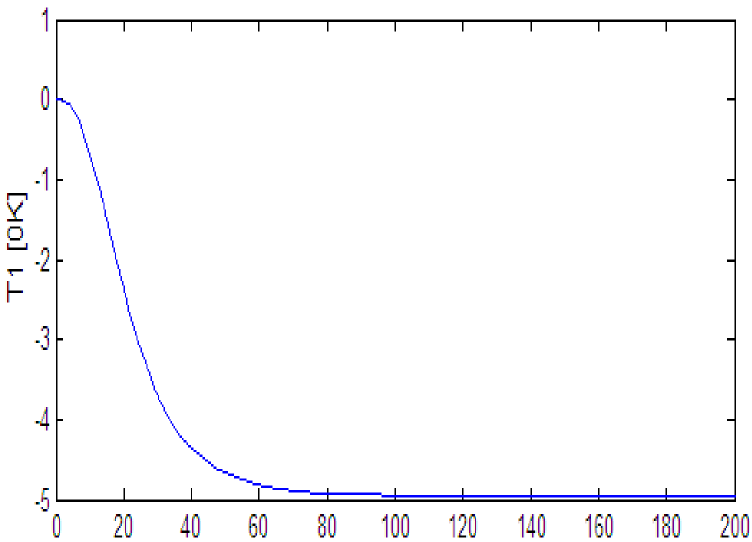

Figure 5. Response of bottom temperature

It can be clearly seen that there is not a single “best” configuration, and this explains why there sometimes seem to be conflicting rankings given in the literature: some focuses on level control, another on composition control. So, if you know what you want, then one can probably find a good configuration to fit your needs.

References

Anderson, B.D.O., and Moore, J. B., 1990. *Optimal control: Linear Quadratic Methods*. Prentice Hall, New Jersey.

Arbel, A., Rinard, I.H., and Shinnar, R., 1996. Dynamics and control of fluidized catalytic crackers. *Industrial & Engineering Chemistry Research* 34(4), 1228–1243.

Faanes, A., 2009. *Operating Manual Unit 102* - Distillation Unit (DU).

Harve, K., 1998. *Studies on controllability analysis and control structure design*. Unpublished PhD Thesis, Norwegian University of Science and Technology, Trondheim.

Hoang, M.S., 2009. *Basics for process control*, Hanoi Technology Publishing House, Hanoi.

Johnson, A.M. and Mohammad, H.M., 2005. *PID Control - New Identification and Design Methods*. Springer-Verlag, London.

Nguyen, P.Q., 2002. *Matlab & Simulink for automatic control Engineer*. Hanoi Science and Technology Publishing House, Hanoi

Shinkey, F.G., 1994. *Distillation Control*. 2 edition, McGraw-Hill Book Company, Columbus.

Shinnar, R., 1981. Chemical reactor modelling for purposes of controller design. *Chemical Engineering Communications* 9(1), 73-99.