MODELLING AND ADVANCED PROCESS CONTROL (APC) FOR DISTILLATION COLUMNS OF LINEAR ALKYLBENZENE PLANT

Xiaoming Jin, Gang Rong and Shuqing Wang

National Key Lab of Industrial Control Technology, Institute of Advanced Process Control,
Zhejiang University, 310027, Hangzhou, China
email: xmjin@iipc.zju.edu.cn

Abstract: This paper introduces industrial application of model predictive control (MPC) for the series of columns in a linear alkylbenzene (LAB) complex. The APC system that is consisted of twelve controlled variables, twelve manipulated variables and eight disturbance variables is used to deal with the constrained multivariable control problem of the distillation columns. Firstly, process modelling that includes experimental test and process identification is presented. Then, the construction and implementation of the APC system for the distillation columns are discussed. Industrial application results show that the APC system can maintain the best operation for a long time and realize ultimate operating potential of the distillation columns. *Copyright © 2003 IFAC*

Keywords: Model predictive control, distillation process, series of columns

1. INTRODUCTION

Distillation columns have been widely used for separation processes in the petroleum and chemical industries. These columns are not only the most energy-intensive operations, but also determine the quality of products of those industries and many times limit process product rates (Kister, 1990). Recent progresses in the control theory, computer and communication have made possible extensive application of APC in process industries (Wang et al, 2001). The large economic benefit of applying APC system to the distillation columns arises from reducing the deviations of crucial process variables and pushing the steady-state operation to a better operating point (Riggs, 1998).

Distillation control has been studied extensively and poses many challenging problems since a distillation column is complex, highly non-linear, multivariable process (Shinskey, 1984, Luyben, 1992, Lundstrom and Skogestad, 1995). Recent efforts have contributed to the analysis of control properties of distillation sequences (Jimenez, et al, 2001) and comparison of control strategies for the distillation columns (Huang and Riggs, 2002). However, the APC system that deals with whole distillation columns rather than single column via a MPC controller is lack of report in open literature.

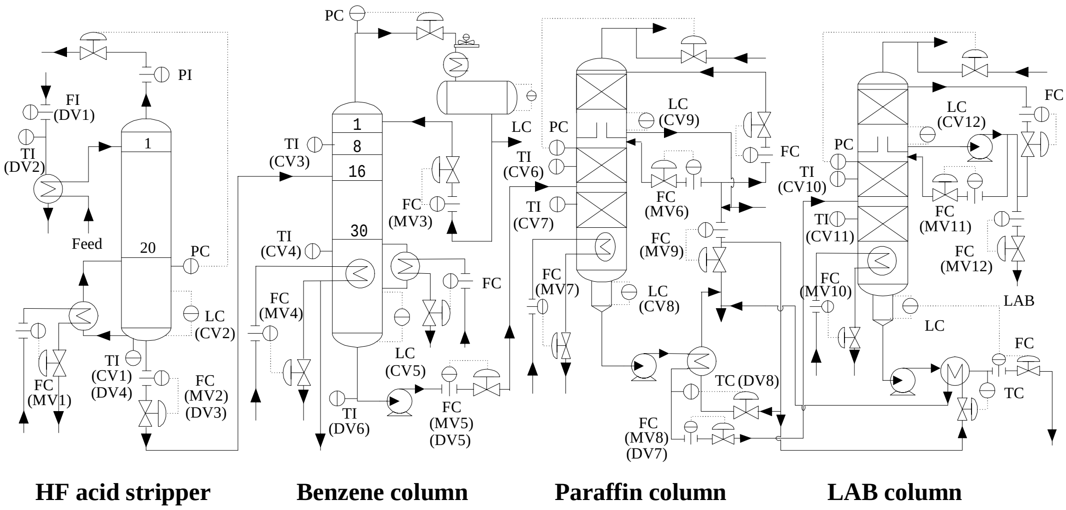
This paper introduces industrial application via a commercial software of MPC for the distillation columns in a LAB plant, which consists of four distillation columns operated in series: HF acid stripper, benzene column, paraffin column, and LAB column. A MPC system that is constructed of twelve controlled variables, twelve manipulated variables and eight disturbance variables is developed to deal with the constrained multivariable control problem of the distillation columns. The MPC system is divided to four subsystems according to the distillation column. Each subsystem is relatively independent with the others and mainly takes charge one column. Industrial application results show that the APC system can maintain the best operation for a long time and realize ultimate operating potential of the distillation system by reducing the consumption of energy, improving product purity, and minimizing operating cost.

2. PROCESS DESCRIPTION

LAB now accounts for nearly all of the worldwide production of alkylbenzene sulfonates (LASs) that are frequently used as raw material of biodegradable household detergents. A LAB complex consists of two major steps: production of normal paraffins, and production of LAB from normal paraffins. The straight run kerosene from a refinery is used to produce normal paraffins through kerosene pre-fractionation, distillate unionfining process and Molex process. Then, the normal paraffins are dehydrogenated to corresponding mono-olefins over a highly selective and active catalyst. Lastly, benzene is alkylated with mono-olefins to LAB using hydrofluoric (HF) acid as the catalyst in alkylation process.

The alkylation process includes two major sections: alkylation section and distillation section. The distillation columns are researched in this paper. The process diagram of the distillation columns with the basic regulatory loops is shown in Figure 1. The controlled variables (CVs) , manipulated variables (MVs) and disturbance variables (DVs) of the APC system are also given in Figure 1. The distillation columns include a HF acid stripper, a benzene column, a paraffin column, and a LAB column. The columns are operated to separate multi-component mixtures of LAB stream from upstream section. The separations of various distillation columns are key variables influencing the economic performance of the alkylation process since the series of columns directly affect product quality, product rate, and utility usage.

The feed of the distillation columns is a mixed LAB stream with HF, benzene and paraffin from alkylation section, which passes through a feed heat exchanger enters at the top tray of the HF stripper. The HF vapor is vented to the HF recovery system. The hot oil to the reboiler is on flow control. The bottom level is directly controlled by adjusting bottom product flow to the benzene column. The bottom temperature is a key variable that reflects fractionation effect. The HF stripper is typically not a bottleneck, but it can be disturbed by operation of alkylation section and flooded if overloaded.



HF acid stripper **Benzene column** **Paraffin column** **LAB column**

Fig. 1 Process diagram of the distillation in LAB plant

The benzene column has a whole condenser. The pressure is controlled via a hot vapor bypass around the overhead condenser. The accumulator level is controlled by adjusting top benzene product flow rate. There is a flow limit for top benzene product in that it is returned to alkylation section. The hot oil to the reboiler and reflux to column are on flow control. The bottom level is directly controlled by adjusting bottom product flow to the paraffin column. There are two important temperatures in benzene column: upper tray temperature and lower tray temperature, which reflect distillation effect and are controlled variables. The benzene column is very sensitive to feed composition disturbance. When HF stripper is overloaded or operated unsteadily the upper tray temperature of benzene column can vary acutely and the lower tray temperature may exceed the lower limit, then influence the operation of the paraffin column.

The paraffin column is a typical packed column that is operating at a pressure slightly lower than atmospheric pressure. A jet pump is used to create and maintain the vacuum. The accumulator level is controlled by adjusting top paraffin product flow that is returned to dehydrogenation process of paraffin. The hot oil to the reboiler and hot reflux to column are on flow control, but the cool reflux to column is fixed. The bottom level is directly controlled by adjusting bottom product flow to the LAB column.

There are two important temperatures in paraffin column: upper temperature and lower temperature, which reflect distillation effect and are used to direct normal operation. The paraffin column is very important to LAB productivity of the plant since any LAB in top paraffin product will suffer losses. The paraffin column is the throughput bottleneck in the distillation columns, and can be easily overloaded.

The LAB column is also a packed column and its operation is similar to the paraffin column's. The top product is LAB that can be used to produce LASs. There are also two important temperatures in LAB column: upper temperature and lower temperature, which reflect distillation effect and are used to instruct normal operation. The LAB column is a column of finished product in which any LAB in bottom heavy LAB product will suffer losses. It is important for the LAB column to maintain steady operation.

In conclusion, the columns need to be maintained close to optimal operating conditions lay at some constraints. Nevertheless, regulatory controls of distillation column are difficult to achieve better control performance all the time for moving the column to its optimal operating point and rejecting disturbances on the controlled variables. Therefore, for the control of distillation column, especially a series of columns affected by many constraints, MPC can be used to improve control performance characterized by a reduction in the variability of the controlled variables through information gathering, process analysis, and constrained multivariable optimisation.

3. MODELLING AND MPC OF THE COLUMNS

There exist several difficulties in terms of controlling the columns of the LAB plant: 1) there are some constraints on equipment capacities and process operations. 2) The flow rate and properties of feed to distillation columns are not constant because it is dependent on the operating conditions of the upstream alkylation section. 3) There are couples among various manipulated variables of each columns, especially two-end temperature control of the column is adopted. 4) There are different time-delays in the responses of CVs versus MVs and some channels show non-linearity. The skilled operators can operate the columns based on their experience and knowledge, but they need to work intensively and can't optimise the columns. Therefore, it is necessary for the distillation columns to implement MPC system in order to realize steady operation and process optimisation. The modelling procedure and results of distillation columns and the structure and implementation of MPC system are given below.

3.1 Modelling of the columns

Dynamic models play a central role in MPC system. It has been shown that modelling is the most difficult and time-consuming work in an MPC project. Generally, the classical black-box model identification methodology is used for MPC controllers. Nevertheless, a good knowledge of the process is still required to bring out possible model structures and define an experiment design.

The modeling procedure of the distillation columns may be summarized as the following steps:
1) studying and understanding the columns operation from collected data sets,
2) determining the model structures of the columns and designing test signals,
3) implementing open-loop multivariable tests of the columns and dealing with the relevant data,
4) estimating the open-loop step response (or impulse response) models of the columns and editing these models,
5) validating and assessing the estimated models of the columns.

The MVs, DVs and CVs of the MPC controller for the distillation columns are listed in Table 1. The choice of CVs was fixed by specifications of the columns including tray temperatures and levels of the reboilers and/or the accumulators.

A dynamic model was developed with adding test signals in the manipulated and disturbance variables on the distillation columns and estimating model parameter from collected data. The sample time was taken to be 2 min. The data collections lasted between 48 and 72 hours. Each time, 98 variables were measured and selected. Data pretreatment and model identification were achieved using a professional off-line identification tool.

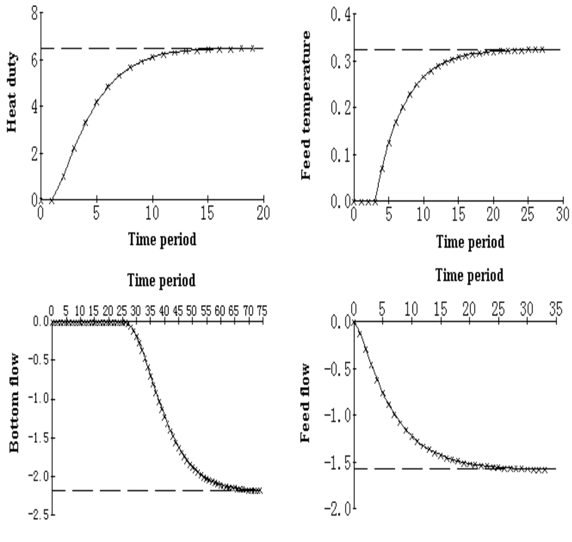


Fig. 2 Open-loop model of bottom temperature in HF stripper

The estimated model of the columns describes the multivariable relationship between 20 inputs and 12 outputs. Here, the open-loop step response models of the bottom temperature in HF stripper are shown in Figure 2 as examples. To demonstrate model validation, we use fresh data to test how well the model output agrees with the measured bottom temperature. Figure 3 shows the test result obtained over more than 900 time period.

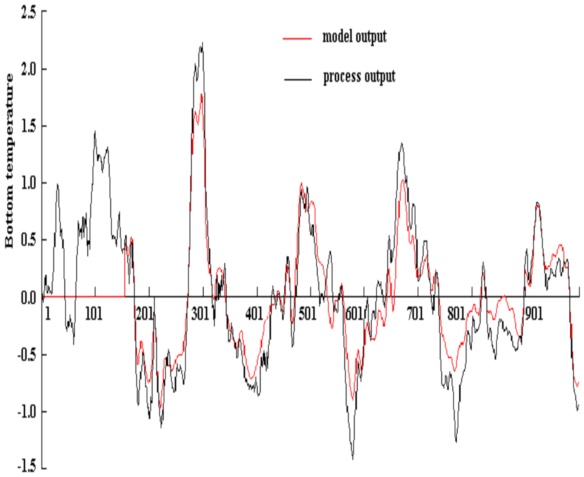


Fig. 3 Model validation of bottom temperature in HF stripper

3.2 MPC of the columns

MPC is a control strategy which predicts the future behavior of process in a control envelope from the past moves of MVs with using the dynamic models of the CVs versus MVs and determines the future moves of MVs in order that CVs will match the target values as close as possible in each control cycle. We adopted Model Algorithmic Control (MAC) to realize the overall control of the distillation columns.

The relationship between CVs and MVs of the columns is given in impulse response that can be achieved by process modelling:

$$y_m(k) = \sum_{j=1}^{N} \hat{h}_j u(k-j) \qquad (1)$$

Where, $\hat{h}_j$ is the coefficients of a impulse response, y is CV, u is MV, and N is number of impulse. The subscript m denotes model output. Model vector $\mathbf{h}=[\hat{h}_1 \cdots \hat{h}_N]^T$ is often saved in host computer and names internal model.

The closed-loop predictive model is obtained in terms of equation (2) as follows:

$$\begin{aligned} y_p(k+i) &= y_m(k+i) + [y(k) - y_m(k)] \\ &= y(k) + [y_m(k+i) - y_m(k)] \\ &= y(k) + \sum_{j=1}^{N} \hat{h}_j [\Delta u(k+i-j) + \Delta u(k+i-j-1) + \cdots + \Delta u(k+2-j) + \Delta u(k+1-j)] \end{aligned}$$

$$i = 1, 2, \cdots, P \qquad (2)$$

Where the subscript m denotes model predictive output, $y(k)$ is current process output.

The reference trajectory that is used to smooth expected output from $y(k)$ to setpoint y_{sp} adopts first order exponential form in MAC as follows:

$$\begin{cases} y_r(k+1) = \alpha^i y(k) + (1-\alpha^i) y_{sp} & i = 1, 2, \cdots, P \\ y_r(k) = y(k) \end{cases} \qquad (3)$$

Where, $\alpha = \exp(-T/\tau)$, T is sample time and τ is constant time of reference trajectory.

To meet the above control goals, the objective function can be formulated as follows:

$$J = [\mathbf{y}_p(k+1) - \mathbf{y}_r(k+1)]^T \mathbf{Q} [\mathbf{y}_p(k+1) - \mathbf{y}_r(k+1)] + \mathbf{u}_2^T(k) \mathbf{R} \mathbf{u}_2(k) \qquad (4)$$

Where,

$$\mathbf{Q} = diag[q_1, q_2, \cdots, q_P]$$

$$\mathbf{R} = diag[r_1, r_2, \cdots, r_M]$$

$$\mathbf{y}_r(k+1) = [y_r(k+1), y_r(k+2), \cdots, y_r(k+P)]^T$$

We have combined our process control experience, design skills and process modelling capabilities into the MPC application. The MPC system shown in Figure 4 is implemented via a MPC controller and divided to four subsystems according to distillation column process. Each subsystem turns on/off independently with the others and mainly takes charge one column.

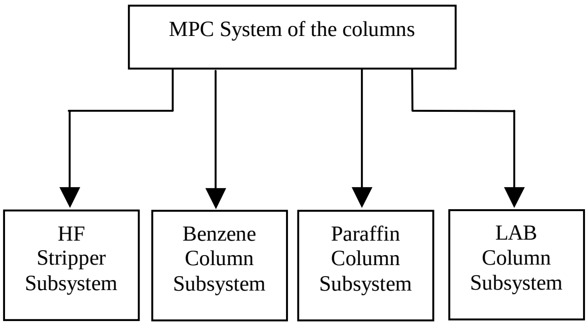


Fig.4 MPC system structure of distillation columns in LAB plant

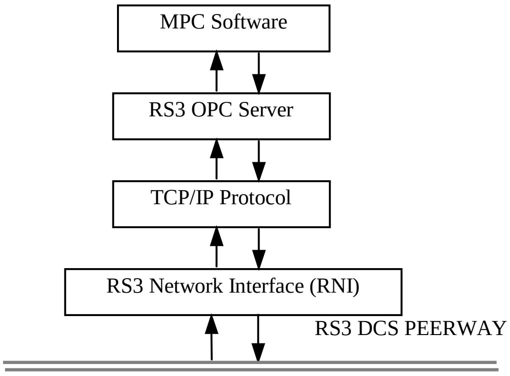


Fig.5 Connection between MPC system and DCS system

After four subsystems turn on, the over MPC system simultaneously adjusts the setpoint values for 12 regulatory loops of the distillation columns, while continuously checking the status of process and equipment limits. To handle the interactions between the different subsystems, the HF stripper bottom temperature is used as a CV in the HF stripper subsystem and its measurement as a DV in the subsequent Benzene column subsystem, the HF stripper bottom flow is a MV in the HF stripper subsystem and its measurement as a DV in the Benzene column subsystem, the downstream columns may be deduced by analogy. By integrating controls into a single controller, interactions across entire distillation columns can be better managed.

The MPC software named APC-Adcon is provided by Zhejiang Supcon Software Ltd.. APC-Adcon has two kinds of optimising functions. One is steady optimising function, which solves the steady optimal values of a CV and MV as target values for dynamic optimising. The other is the dynamic optimising function, which calculates the optimal future path of an MV. The MPC software is equipped in the host computer of an existing distributed control system (DCS) RS3 made in Fisher-Rosemount and the communication between host computer and DCS is seamless connection based on RNI for RS3 and OPC interface. Figure 5 shows the connection between MPC system and DCS system.

Table 1 Configuration for distillation columns MPC system

Controlled Variables	Manipulated Variables	Disturbance Variables
CV1-HF stripper bottom temperature	MV1-HF stripper reboiler duty	DV1-Reactor feed flow
CV2-HF stripper reboiler level	MV2-HF stripper bottom flow	DV2-Reactor feed temperature
CV3-Benzene column upper tray temperature	MV3-Benzene column reflux flow	DV3-HF stripper bottom flow
CV4-enzene column lower tray temperature	MV4-Benzene column reboiler duty	DV4-HF stripper bottom temperature
CV5-Benzene column reboiler level	MV5-Benzene column bottom flow	DV5-Benzene column bottom flow
CV6-Paraffin column upper tray temperature	MV6-Paraffin column hot reflux flow	DV6-Benzene column bottom temperature
CV7-Paraffin column lower tray temperature	MV7-Paraffin column reboiler heat duty	DV7-Paraffin column bottom flow
CV8-Paraffin column reboiler level	MV8-Paraffin column bottom flow	DV8-Paraffin column bottom temperature
CV9-Paraffin column accumulator level	MV9-Paraffin column top paraffin flow	
CV10-LAB column upper tray temperature	MV10-LAB column hot reflux flow	
CV11-LAB column lower tray temperature	MV11-LAB column reboiler heat duty	
CV12-LAB column accumulator level	MV12-LAB column top LAB flow	

4. PERFORMANCE OF THE APC SYSTEM

After the MPC controller was tuned, it had been tested on-line for three months. The comparison of the performance before and after the MPC implementation is summarized. The comparison of product quality of LAB plant and LAB in recycle paraffin is given in Table 2. The result of control performance comparison for LAB column MPC subsystem is shown in Figure 6 and Figure 7.

The MPC system is superior to human supervisory control in follow aspects:

1) The skill of most high experienced operator has been implemented by the MPC system.
2) Operation and compensation is executed at fairly frequent intervals. In case of a skilled operator, operation frequency is about 20 minutes, while, in case of the MPC system, it is 2 minutes.
3) Operator control manipulates the process variables in sequence. Nevertheless, the MPC control can manage several process variables in parallel.

After the implementation of the MPC system, the deviation of the main process variables became one half of that before implementation. As a result, The over economical merit from this implementation is approximately over one million Yuan by reducing the consumption of energy, improving product purity, and minimizing operating cost.

Table 2 Comparison of product quality of LAB plant

Quality Index of LAB Plant	Average Value		Standard Deviation	
	Before	After	Before	After
LAB TNP %	Base	-0.01%	0.094	0.048
LAB BR	Base	-0.37	1.580	0.820
LAB in recycle paraffin	Base	-0.01%	0.280	0.071

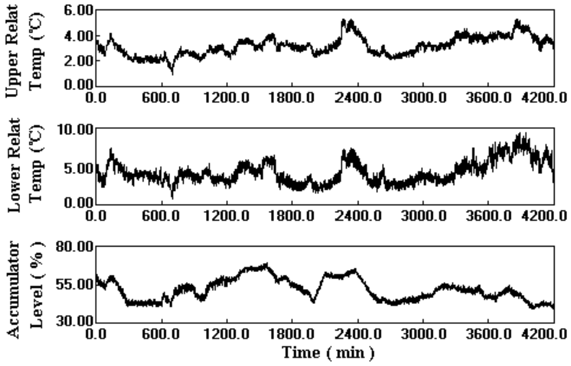


Fig. 6 Performance of the CVs of LAB column before MPC implementation

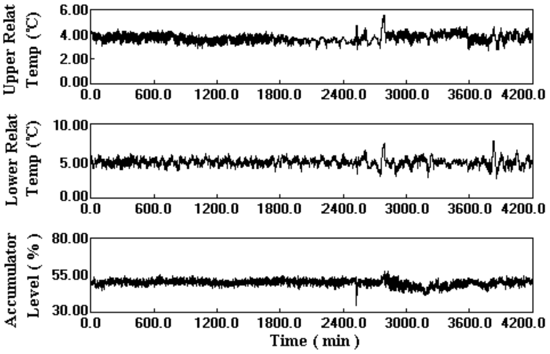


Fig. 7 Performance of the CVs of LAB column after MPC implementation

5. CONCLUSION

This paper presents process modelling and APC system of the distillation columns in LAB plant. Industrial application results show that the APC controller is superior to conventional control. The APC control can maintain the best process operation for a long time and realize ultimate operating potential of the distillation columns. Economic benefits are achieved by using the APC system when the process model is constructed correctly and MPC controller is set compatibly.

REFERENCES

Kister, H.Z.(1990). *Distillation Operation*, McGraw-Hill, New York.

Wang, S.Q. et al, (2001). *Advanced Process Control and its Applications*, Chemical Industry Press (in Chinese), Beijing.

Riggs, J.B. (1998) Improve distillation column control. *Chemical Engineering Progress*, 10: 31-47.

Shinskey, F.G. (1984) *Distillation control*, 2nd Edition, McGraw-Hill, New York.

Luyben, W.L. (1992) *Practical distillation control,* Van Nostrand Reinhold, New York.

Lundstrom, P., S. Skogestad, (1995) Opportunities and difficulties with 5 × 5 distillation control. *Journal of Process Control*, 5(4): 249-261.

Jimenez, A. et al, (2001) Analysis of control properties of conventional and nonconventional distillation sequences. *Ind. Eng. Chem. Res.*, 40: 3757-3761.

Huang, H.T., J.B. Riggs, (2002) Comparison of PI and MPC for control of a gas recovery unit. *Journal of Process Control*, 12: 163-173.