MODEL PREDICTIVE CONTROL OF VAPORIZER IN VINYL ACETATE MONOMER PROCESS

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I declared that this thesis entitled "*Model Predictive Control of Vaporizer in Vinyl Acetate Monomer Process*" is the result of my own researched excepted as cited in the references. The thesis has not been accepted for any degree and is not concurrently submitted in candidature of any other degree.

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Date : 29 APRIL 2008

To my beloved mother and father

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ABSTRACT

A Model Predictive Control (MPC) algorithm is developing on the Vinyl Acetate Monomer process provided by Luyben and Tyreus. Stress on the vaporizer, consideration only for the 'Single Input Single Output' (SISO) strategy with two control objective. Three closed loop have been applying with involve the vaporizer pressure and level. The MPC algorithm is implemented to the control loop by using Simulink/MATLAB 7.0. The limitation on the set-point using conversional controller would be overcome by using an advanced control method that is MPC. In direct comparison, we show that MPC algorithm would give better performance due to future prediction potential compare to conventional controller that is PI. The capabilities of MPC have been tested with three different inputs which represent the real mechanistic behaviors of chemical plant through the Simulink/MATLAB 7.0 simulations. The MPC algorithm is believed can capture the nonlinear and robust of Vinyl Acetate Monomer process that refers to real chemical industries.

ABSTRAK

Kawalan peramalan model (MPC) dibangunkan dan diuji pada proses *Vinyl Acetate (VAc) Monomer* yang dibangunkan oleh *Luyben dan Tyreus(1997)*. Dalam kajian ini, kami hanya menumpukan perhatian kepada *vaporizer*. Kajian hanya tertumpu pada cara'Satu Input Satu Output (SISO) dengan dua objektif kawalan. Dua pembolehubah kawalan, dua pembolehubah manipulasi dan dua pembolehubah penganggu dikenalpasti bagi menghapuskan perubahan pada *output*. Secara keseluruhannya MPC dibangunkan pada gelung kawalan dalam persekitaran *Simulink/MATLAB 7.0*. Kekangan pada input boleh diatasi atau dihapuskan dengan mengunakan MPC strategi. Dengan melakukan perbandingan , MPC dikenalpasti menberikan tindakbalas yang lebih positif berbanding dengan pengawal konversional iaitu pengawal berkadar terus dan kamiran (PI).Kemampuan MPC diuji mengunakan tiga input yang berbeza bagi mewakili ketidakstabilan industri kimia yang sebenar mengunakan *Simulink/Matlab 7.0* simulator.Selain itu,MPC juga diakui boleh mengatasi masalah dinamik , tidak linear dan tidak sekata proses *Vinyl Acetate Monomer* yang merujuk kepada industri kimia yang sebenar.

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a_i	-	*i*th unit step response coefficient
A	-	Dynamic matrix.
CV	-	Control variable
d	-	Disturbance prediction.
DMC	-	Dynamic matrix control.
DV	-	Disturbances variable
e	-	Predicted error.
$\tilde{e}$	-	Vector of predicted errors
i	-	Index
I	-	Identity matrix
IDCOM	-	Identification and command
j	-	Time index
LQG	-	Linear quadratic Gaussian (LQG)
M	-	Control horizon (number of controller output moves)
MPC	-	Model predictive control.
MV	-	Manipulated variable
N	-	Current samples
N	-	Model horizon.
P	-	Prediction horizon
PI	-	Proportional integral
QDMC	-	Quadratic dynamic matrix control
Q, R	-	Weighting matrices.
SISO	-	Single input single output
VAc	-	Vinyl Acetate monomer process
u	-	Controller output variable
y_o	-	Initial steady state of process variable
$\tilde{y}$	-	Predicted process variable.

y_{sp}	-	Process variable set point.
λ	-	Move suppression coefficient (controller output weight).
Δt	-	Sampling period.
α_i	-	Reference rejection.

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CHAPTER 1

INTRODUCTION

1.1 Overview of Research.

In recent year, chemical industries face a lot of problem due to the economical considerations in the process industry. Generally, all chemical industries are inherently nonlinear. This, together with higher product quality specification, the increasing productivity demands and tightens environmental regulations. A further complication is that modern plants have become more difficult to operate because of the trend toward complex and highly integrated process. For such plant, it's difficult to prevent disturbances from propagating from one unit to other interconnected units (Seborg *et al*, 2003).

This make computer based process become very important to operate modern plants safely and profitably. Besides that, a new generation of process control is very important to reduce variability in the end product, which ensures a consistently high-quality product. By reducing variability, its also can save money by reducing the product padding to meet required product specifications. Padding refers to the process of making a product of higher-quality than it needs to meet some specifications. Furthermore, its can increase efficiency because some processes need to be maintained at a specific point to maximize efficiency. Otherwise, process control can ensure safety because sometimes a run-away process, such as an out-of-control nuclear or chemical reaction, may result if we do not maintain precise control of all the process variables. The proportional-integral (PI) controller has been widely used in the industry because of its simple structure and robust performance in a wide range of operating conditions. PI

The proportional-integral (PI) controller has been widely used in the industry because of its simple structure and robust performance in a wide range of operating conditions. PI control has reigned as the industrial standard, and for good reason: such as simple, fast, versatile, flexible, and a sensible design that the underlying algorithm hasn't changed one bit in all these years. But, the widespread use of PI, the technologies inherent its weaknesses which are:

A. PI controls have difficulty handling process delays, nonlinear processes, and noisy process signals. This leads to suboptimal control and increased tuning effort.
B. PI is not as robust as alternatives, often delivering higher process variability.
C. PI tuning is not easy to handle. Effective tuning requires experience, extensive training, and an investment in tuning software.
D. PI transfers process signal noise directly to its controller output. This accelerates valve wear and increases energy usage.

Considering to the pad simulation, the major problem using the conventional controller is the limitation on variability of set point tracking. For example, the PI controller for vaporizer pressure only let the range of 125 -130 psi. This limitation would make the controller cannot perform well when the changes of set point that is pressure. These weaknesses add up over time, with the net impact being PI use may actually increase process variability, decrease production and product quality, and ultimately increase operating and maintenance costs. Besides that, other factor that encourages the replacement of PI is:

A. The evolution of control systems from pneumatics to distributed control system (DCS) to process knowledge systems.
B. The convergence of hardware and software technologies
C. Wider industry acceptance of advanced control technologies

Due to its high potential for global optimization, advanced controls such as Model Predictive Control (MPC) it's preferred as an alternative controller. Objective of MPC control calculation is to determine a sequence of moves by manipulates input changes so that predicted response moves to the set point in optimal manner.

MPC displays improved performance because the process model allows current computations to consider future dynamic events. For example, this provides benefit when controlling processes with large dead times or nonminimum phase behavior (Dougherty and Cooper, 2001). In MPC, the outputs would be a control variables and input is manipulated variables. Furthermore, the main element in MPC is predictor and optimizer. This combination would make MPC as a perfect controller rather than PI controller.

In this study, the MPC will be implementing on Vinyl Acetate Monomer process by Luyben and Tyreus (1998) in MATLAB 7.0 environment. The process has 10 unit operations, which includes a vaporizer, a catalytic plug flow reactor, a feed-effluent heat exchanger (FEHE), a separator, a gas compressor, an absorber, a carbon dioxide (CO2) removal system, a gas removal system, a tank for the liquid recycle stream, and an azeotropic distillation column with a decanter. But, this research only concentrates to control one unit operation that is a vaporizer.

1.2 Research Objective.

The objective of this research is to implement the model predictive control at the vaporizer in Vinyl Acetate Monomer process.

1.3 Scope of Study.

A. To develop test-bed platform for nominal condition.

B. Dynamic response on test-bed.

C. Set-point tracking and disturbance rejection on Vinyl Acetate monomer process.

D. Implementation of MPC on vaporizer pressure and level control loop for Vinyl Acetate monomer process.

E. The tuning process for MPC controller.

F. Performance comparison between the MPC and PI controller.

CHAPTER 2

LITERATURE REVIEW.

2.1 Overview of Vinyl Acetate Monomer Process.

In 1997, Luyben and Tyreus was published a new plantwide control test problem based on the Vinyl Acetate monomer (VAC) process. Generally, this process would the real plant wide problem for plant wide design, optimization, and control study with:

I. A realistically large process flow sheet containing standard chemical unit operation.
II. A process with the typical industrial characteristic of recycles stream and energy integration.
III. A real Nonideal chemical components.

In the VAC process, there are 10 basic unit operations, which include a vaporizer, a catalytic plug flow reactor, a feed-effluent heat exchanger (FEHE), a separator, a gas compressor, an absorber, a carbon dioxide (CO2) removal system, a gas removal system, a tank for the liquid recycle stream, and an azeotropic distillation column with a decanter as shown in figure 2.1. There are seven chemical components in the VAC process that is Ethylene (C_2H_4), pure oxygen (O_2), and acetic acid (HAc) are converted into the vinyl acetate (VAc) product, and water (H_2O) and carbon dioxide

The following reactions take place:

$$C_2H_4 + CH_3COOH + 1/2O_2 \rightarrow CH_2 = CHOCOCH_3 + H_2O \tag{2.1}$$
$$C_2H_4 + 3O_2 \rightarrow 2CO_2 + 2H_2O \tag{2.2}$$

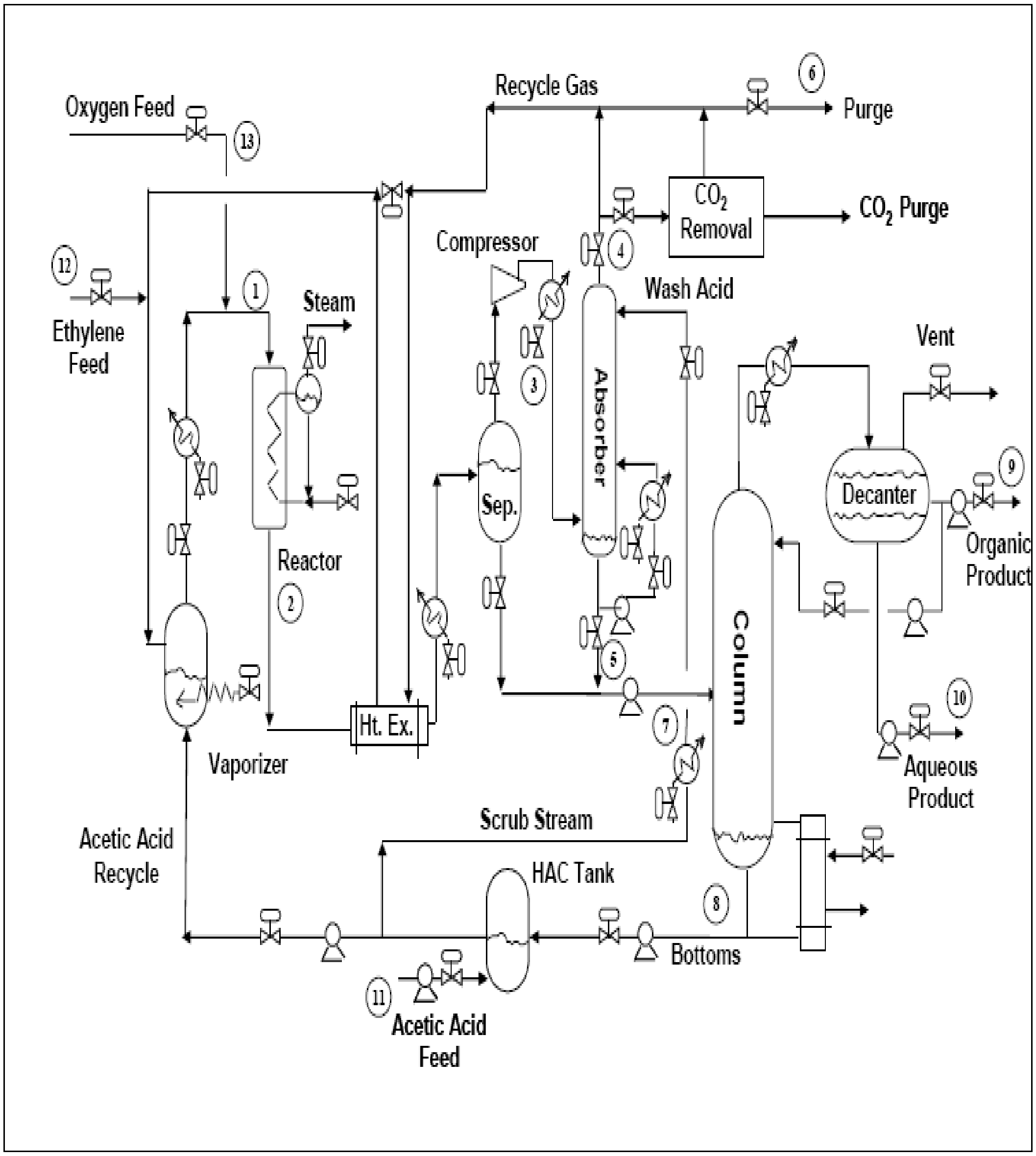

Figure 2.1: Flow sheet of the Vinyl Acetate Monomer Process (Luyben and Tyreus, 1997).

2.1.1 Vinyl Acetate monomer Process Modeling

This section discusses design assumptions, equipment data, and modeling formulations for each unit operation. For each unit, the state and manipulated variables are identified.

2.1.1.1 The Vaporizer

The vaporizer is implemented as a well-mixed system with seven components. It has a gas input stream (F_1), which is a mixture of the C_2H_4 feed stream and the absorber vapor effluent stream. It also has a liquid input stream (F_2), which comes from the HAc tank. There are 8 state variables in the vaporizer, including the liquid level, the mole fractions of O_2, CO_2, C_2H_4, VAc, H_2O, and HAc components in the liquid, and the liquid temperature. The liquid level is defined by the ratio of the liquid holdup volume over the total working volume. Since the dynamics of the vapor phase are ignored, total mass, component and an energy balance are used to calculate the dynamics in the liquid as:

$$r_L^{VAP} \dot{V}_L^{VAP} = F_1^{VAP} MW_1^{VAP} + F_2^{VAP} MW_2^{VAP} - F_V^{VAP} MW_V^{VAP}$$

(2.3)

$$M_L^{VAP} \dot{x}_{L,i}^{VAP} = F_1^{VAP}(X_{1,i}^{VAP} - x_{L,i}^{VAP}) + F_2^{VAP}(x_{2,i}^{VAP} - x_{L,i}^{VAP}) - F_V^{VAP}(y_{V,i}^{VAP} - x_{L,i}^{VAP})$$

(2.4)

$$Cp_L^{VAP} M_L^{VAP} \dot{T}_L^{VAP} = F_1^{VAP}(h_1^{VAP} - h_L^{VAP}) + F_2^{VAP}(h_2^{VAP} - h_L^{VAP}) - F_V^{VAP}(H_V^{VAP} - h_L^{VAP}) + Q^{VAP}$$

(2.5)

Vapor liquid equilibrium (VLE) is assumed in the vaporizer, and as a result, the vaporizer pressure and the vapor compositions are determined by a bubble point calculation. Two manipulated variables (Q^{VAP} and F_V^{VAP}) are available in the vaporizer. In the base operation, the liquid holdup, V_L^{VAP}, is 2.8 m^3, which is 70% of the working level volume. The vaporizer is followed by a heater, and the heater duty is a manipulated variable. In the base operation, the heater exit temperature is specified to be 150 ^{o}C.

2.1.1.2 The Catalytic Plug Flow Reactor

The reactor is implemented as a distributed system with ten sections in the axial direction. Two irreversible exothermic reactions, given by Eq. (2.1) and (2.2), take place. In the MATLAB model, the following assumptions are made for the purpose of model simplification:

§ Plug flow is assumed so that there are no radial gradients in velocity, concentration, or temperature. Diffusion occurring in the axial direction is considered negligible compared to the bulk flow. Potential and kinetic energy and work are considered negligible in the energy balance calculation.

§ It is assumed that the mass and heat transfer between the fluid and catalyst are very fast and therefore the concentrations and temperatures in the two phases are always equal.

§ Pressure drop is assumed linear along the length of a tube, and it is time-independent. Eqn.2.6 is used to calculate the pressure drop in each section:

$$\Delta P / \Delta Z = f * r_1^{RCT} * (u_1^{RCT})^2 \tag{2.6}$$