NONLINEAR MODEL BASED CONTROL OF A RECYCLE
REACTOR PROCESS

by

RAVI SHANKAR KANADIBHOTLA, B.Sc.CH.E.

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CHAPTER 1

INTRODUCTION

Many process operations in the refining and petrochemical industries are comprised of more than one operation. Examples include extraction/distillation combinations (e.g. aromatic recovery in a refinery and supercritical fluid extraction of a speciality chemicals process), a reactor/catalyst regeneration system (fluidized catalytic cracker in a refinery), and a low conversion reactor with a separation system and feed recycle (e.g. an adipic acid process used in nylon production). Each of these examples is made up of separate unit operations with significant interaction among the individual units. While all of these processes provide substantial steady state operational advantages, they are, in general, more difficult to control than their individual units. That is, coupling among the individual units significantly complicates the overall process control problem.

Most industrial control applications for coupled, multiunit processes use traditional proportional-integral-derivative (PID) controllers. Since PID controllers, in general, use one manipulated variable to control one output variable neglecting coupling and also neglecting process nonlinearity, they do not provide satisfactory control performance for many of these processes. Improved control for

many coupled, multiunit processes would translate into very significant economic savings as the result of utility savings and/or increased production rates and/or improved product uniformity since these processes operate at high production rates.

Advanced control strategies can be classified into two main subgroups, strategies based on linear models and strategies based on nonlinear models. Among the linear model based strategies are Internal Model Control (IMC) (Garcia and Morari, 1982), Dynamic Matrix Control (DMC) (Cutler et al., 1983), Model Algorithmic Control (Richalet et al., 1979). Among the nonlinear models are Generic Model Control (GMC) (Lee and Sullivan, 1988b), Nonlinear IMC (NLIMC) (Economou et al., 1986) and Nonlinear Model Predictive Control (Parrish and Brosilow, 1988).

The technique used in this work is nonlinear model based control and it uses the GMC law. The nonlinear models used are approximate models, phenomenological models derived from constitutive and conservation relationships. Since there is model/process mismatch with changes in operating states or under the effect of disturbances and changes in setpoint, certain parameters are adjusted to account for model/process mismatch. These parameters are chosen so as to provide accurate information about the most uncertain process quantities. Nonlinear PMBC has been successfully applied to processes such as coal gasification (Pandit et al., 1989),

binary distillation with a sidestream draw (Riggs, 1989), wastewater neutralization (Choi and Rhinehart, 1986; Williams et al., 1989) and batch polymerization (Zheng, 1990). The control applications make nonlinear PMBC an attractive proposition for multiunit process control since the nonlinear models directly consider the nonlinearity of the process, and the coupling between the various units of a multiunit process can also be considered directly. In addition, the models used for control can also be used for on-line optimization.

1.1.Process Description

The basis of this study is a multiunit process simulator provided in the Tennessee Eastman Challenge problem reference (Fig 1.1). The Tennessee Eastman problem was developed by their control engineers Ernie Vogel and James Downs and it is based on an actual process. This was posed as a challenge problem to academic researchers in process control at the 1990 AIChE Annual National meeting in Chicago. The process is a recycle reactor process which is open loop unstable. Gaseous reactants are fed to the reactor where they undergo exothermic catalytic reactions. The products from the reactor are cooled and flashed in the separator. The vapor from the separator is largely recycled (except for the purge) back to the reactor and the remaining liquid serves as feed to the stripper. The stripper uses gaseous feed and gaseous material to remove by-products from the heavier final products. The heat for the

Fig 1.1. The Tennessee Eastman process schematic

stripping is supplied by steam. The vapor stream leaving the stripper is combined with the compressed vapor stream (which originally came from the overhead of the separator) and is returned to the reactor.

There are four reactions, two simultaneous gas/liquid exothermic reactions and two additional by-product reactions. All reactions are irreversible and exothermic. They are as follows.

The gaseous reactants are fed to the reactor where they form liquid products. The gas-phase reactions are catalyzed by a nonvolatile catalyst dissolved in the liquid phase. The reactor has an internal cooling bundle for removing the heat of reaction. The reactor agitator speed affects only the heat transfer coefficient. The large recycle stream contains a large amount of unreacted feedstock.

1.2.Control Objectives

The control objectives in this process are the same as most processes, that is, to maintain product compositions at their setpoints and to minimize variation of manipulated variables. The primary control objective is to minimize the variability of the bottom draw from the stripper while

maintaining product specifications and minimizing the manipulation of the feeds.

Since the feeds are the products of upstream processes and have very little holdup, variation of these streams is to be minimized. The conditions which cause open-loop instability have to be identified and made stable in the closed loop. The reactor is the source of open loop instability. The small heat transfer area to volume ratio of the exothermic reactor requires sufficiently rapid heat removal to stabilize operation. Moreover, reactor temperature and the reactant ratios in the feed determine selectivity. If reactor temperature and reactor pressure are maintained at their setpoints, most of the unstable nature of the system is removed. Excessive accumulation of noncondensable reactants in the recycle loop has to be avoided since these components would drive up the pressure.

Control strategies are not general and are unique in the sense that they evolve out of a good comprehension of the process, especially the interaction among the various variables involved. The Tennessee Eastman problem has been formulated to test various algorithms in vogue today and the effort in this work has been directed towards applying the nonlinear model based control approach effectively to a multivariable open loop unstable process with constraints.

CHAPTER 2

LITERATURE SURVEY

There has been a proliferation of process control literature recently. However, the bulk of literature pertains to single-input, single-output (SISO) processes, and multi-input, multi-output (MIMO) systems have remained on the fringes of control literature. A variety of control techniques has been proposed in the recent past to deal with MIMO processes.

The multivariable techniques in are, among others,

- (1) relative gain arrays (Bristol, 1966);
- (2) optimal control with state space models (Wu et al., 1983);
- (3) singular value decomposition (Tomizuka et al., 1979).

There have been several advanced control strategies put forward over the last decade or so.

2.1.Linear Model Based Control

The linear model based strategies, include among others,

- (1) Internal Model Control (Garcia and Morari, 1982);
- (2) Model Algorithmic Control (Richalet et al., 1979);
- (3) Dynamic Matrix Control (Cutler, 1983);
- (4) IDCOM (Setpoint, 1979).

Internal Model Control employs a variable order plus deadtime model using Laplace transfer functions. The manipulated variable is calculated for the process to reach

its setpoint and a filter applied to the feedback signal to remove process/model mismatch and to improve robustness. The setpoint is then automatically biased by the process/model mismatch to remove offset.

IdCom uses an impulse response model instead of a step response model and uses time series models to find a series of manipulated action that minimizes the sum of squared deviations from a pre-set path to setpoint.

Dynamic Matrix Control has been utilized extensively in industrial applications. It uses a step response linear model for each input/output pair. DMC uses a time horizon in the sense that a hypothetical disturbance level is used at each time step to remove process/model mismatch. This "residual" is used to bias the setpoint. The major benefits are the constraint handling capabilities.

The models are linear approximations of the process and are derived from step responses. The models do not account for process nonlinearity and process/model mismatch occurs. They also may not be effective for nonstationary processes because they accurately model for the process only around the operating point.

2.2.Nonlinear Model Based Control

Nonlinear model strategies have been proposed and among them are:

(1) Generic Model Control (Lee and Sullivan, 1988);

- (2) Nonlinear Predictive Model Control (Parrish et al, 1988);
- (3) Reference System Synthesis (Bartusiak, 1987).

Generic Model Control uses a simple nonlinear model of the process which captures the essential phenomenological aspects of the process. Let us assume a first-order SISO system described by

$$\frac{dy}{dt} = f(y, u, \underline{d}_o, k) \quad (2.1)$$

where y is the output variable, u is the manipulated variable, $\underline{d}_o$ is a vector of measured disturbances, and k is a vector of parameters. A forward difference approximation leads to

$$\frac{y_{sp} - y_o}{\tau} = f(y_o, u, \underline{d}_o, k) \quad (2.2)$$

where y_o is the current process variable, y_{sp} is the setpoint, τ is the time interval and a tuning parameter. Thus this equation has to be solved for u , the control action.

$$f(y_o, u, \underline{d}_o, k) + K_1(y_o - y_{sp}) + K_2 \int (y - y_{sp}) dt = 0. \quad (2.3)$$

Nonlinear Internal Model Control was used by Economou et al. so that nonlinear models could be used by Internal Model Control. However, it is implicit and requires an iterative integration of the model to arrive at the control action.

Nonlinear predictive model control uses the internal model control structure but removes the process/model mismatch by making adjustments or by parameter updates.

Reference System Synthesis employs a similar approach to NPMBC in that it forces an nth-order process to behave like a lower order process in closed loop by forcing the system to follow a trajectory set by the reference system.

A variety of nonlinear models has been investigated using the differential geometric approach. Nonlinear decoupling via feedback using this approach has been explored by Isidori et al. (1981). An overview of nonlinear geometrical methods also been summarized by Kantor (1987). Henson and Seborg (1989) also discuss a unified differential geometric approach to process control. Ha and Gilbert (1983) outline a complete characterization of decoupling laws for a general class of nonlinear systems.

The technique we employ is nonlinear model based control and it uses the Generic Model Control law to generate the control trajectory. The nonlinear models used are approximate models, phenomenological models derived from constitutive and conservation relationships. Certain parameters are adjusted to account for model/process mismatch following changes in operating states, disturbances or setpoint changes. These parameters are chosen so as to provide accurate information about uncertain process quantities. Nonlinear PMBC has been successfully applied to processes such as coal gasification (Pandit et al., 1988), pH (Williams et al., 1990), (Choi et al., 1986), (Mahuli et al., 1992), commercial distillation (Riggs et al., 1989), wastewater neutralization (Rhinehart,

1989). Most of the applications have been to single units. The control applications make nonlinear PMBC an attractive proposition for multiunit process control.

2.3. Multiunit Process Control Overview

The main objectives in multiunit process control are to decouple interactions among loops, handle deadtimes and smoothly recover from both setpoint changes and disturbances. Most of the literature published in the area of multivariable control has been very theoretical and there have been a number of mathematical techniques like the two block structured singular value measures (Nett et al., 1981), loop transfer recovery (Doyle et al., 1981), the linear quadratic gaussian procedure and stable factorization technique (Vidyasagar et al., 1985). Most of these techniques incorporate linear models.

The multivariable techniques used in chemical engineering are the relative gain arrays and decouplers, optimal control with state space models and singular value decomposition. The relative gain array proposed by Bristol (1966) was a steady-state interaction measure and does not account for dynamic effects, which makes it ineffective for an open-loop unstable process. SVD has been used in conjunction with linear models and does not account for process nonlinearity. There have been applications of nonlinear predictive control and dynamic matrix control to processes exhibiting nonlinearity.

Nonlinear process model based control was investigated on a simulated multiunit supercritical fluid extraction process (Ramachandran et al., 1991) which consisted of a highly coupled extractor, distillation column system. The controlled variables were the composition of the raffinate, the overhead and bottom product compositions and the manipulated variables were the solvent flow rate, the reflux and boilup rates.

2.4.Applications to Recycle Reactors

There have been applications of dynamic matrix control to several multiunit processes, particularly to recycle reactors. However, most of these processes were open loop stable. A multivariable controller has been implemented to control reaction severity in a FCCU feed hydrotreater, (Scott et al., 1988). Four manipulated variables were used to control three dependent variables in the DMC controller which incorporated an additional three feedforward variables, and constraints on three other dependent variables. The system consisted of oils from three crude units and a delayed coker at the feed surge drum, which were combined with hydrogen and heated to reaction temperature in a fired heater before entering the reactor which contains two fixed catalyst beds with interstage hydrogen quench. Effluents flow through hot and cold flash drums where unreacted hydrogen is compressed and recycled to the reactor. Dynamic Matrix Control was also applied to a hydrocracker second stage reactor, which used 12

manipulated variables to predict the behavior of 16 controlled variables(Cutler et al.,1988). The process included in the scope was a second stage hydrocracker reactor, vapor liquid separators and hydrogen compressors,two major towers in the fractionation train and a recycle loop. Quadratic dynamic matrix control has been applied to hydrocracking reactors, (Kelly et al., 1988). The system consists of two multibed reactors, a first stage hydrotreater, a second stage hydrocracker and a recycle loop for unreacted hydrogen and makeup hydrogen.

2.5.Open-Loop Unstable Responses

Not much literature has been published in the application of the control strategies to multiunit processes, much less in the area of multivariable open-loop unstable processes with constraints. Georgiou et al. (1989) investigated control of a multivariable open loop unstable process, a network of two CSTR's with recycle and separator using pole placement techniques. The whole system was open loop unstable because of the limited heat removal due to the small ratio of area to volume. Yamuna Rani et al. (1989) investigated the application of Dynamic Matrix Control using logarithmic and exponential transformations of the controlled variable to a multivariable open-loop unstable process with least-squares minimization for constraints. The main problem with open-loop unstable processes is process identification. Process identification

becomes an uncomfortable task because open loop step tests cause the process to become unstable. Most multivariable techniques, such as relative gain arrays involve steady state analysis of open loop responses. A sensitivity analysis leads to a comprehensive understanding of coupling involved and permits design of a proper control system.

Chapter 3

PROCESS DESCRIPTION

The process description is mentioned in the introduction (Section 1.1) in general terms. Some other points are listed in the section that follows.

The condenser condenses the reactor products. It is equipped with an internal cooling bundle to cool the reaction products to the separator temperature. Noncondensed components recycle through a centrifugal compressor. The compressor internal surge protection is through a mechanical bypass arrangement.

The condensed components from the separator move to a stripping column to remove the remaining reactants by stripping with the feed C/A. Products are moved to a downstream refining section.

3.1.Constraints

The process operating constraints are on the reactor pressure, reactor temperature, reactor level, separator level and stripper level. There is no lower limit on the reactor temperature and reactor pressure. These constraints are listed in Table 3.1. The shutdown conditions are listed along with the lower and upper constraints on the reactor temperature, reactor pressure and the levels.

Table 3.1 Process Constraints

VARIABLE	OPERATING LIMIT	SHUTDOWN LOW LIMIT	SHUTDOWN HIGH LIMIT
REACTOR TEMPERATURE	150 DEG C	NONE	175 DEG C
REACTOR PRESSURE	2895 kPa	NONE	3000 kPa
REACTOR LEVEL	50-100 %	0%	100%
SEPARATOR LEVEL	30-100	0%	100%
STRIPPER LEVEL	30-100	0%	100%

3.2.Disturbances

There are twenty process disturbances and four setpoint changes against which the control strategy responses have to be evaluated. The disturbances are activated when the process disturbance variable in the program is changed from 0 to 1. The step disturbances are applied to A/C ratio with B composition constant in the A&C feed, B composition with A/C ratio kept constant in the A&C feed, D feed temperature, reactor and condenser cooling water inlet temperatures, A feed loss and a C header pressure loss. Random variations are also applied to A,B,C compositions in the A&C feed, D feed temperature, C feed temperature, reactor cooling water inlet temperature, and the condenser cooling water inlet temperature. A slow drift in the reaction kinetics and sticking cooling water valves were investigated. These are listed in Table 3.2.

3.3.Setpoint changes

The setpoint changes for the base case are changes to the production rate, the reactor pressure, a product grade change and the purge gas composition of component B. These are listed in Table 3.3.

3.4.Assumptions

The main assumptions that have been used in implementing this model are as follows:

Table 3.2 Process Disturbances

1	A/C Feed Ratio, B Comp Const (Stream 4)	Step
2	B Comp, A/C Ratio Constant (Stream 4)	Step
3	D Feed Temperature (Stream 2)	Step
4	Reactor Cooling Water Inlet Temperature	Step
5	Condenser Cooling Water Inlet Temp	Step
6	A Feed Loss (Stream 1)	Step
7	C Header Pressure Loss (Stream 4)	Step
8	A,B,C Feed Composition (Stream 4)	Random Variation
9	D Feed Temperature (Stream 2)	Random Variation
10	C Feed Temperature (Stream 4)	Random Variation
11	Reactor Cooling Water Inlet Temperature	Random Variation
12	Condenser Cooling Water Inlet Temperature	Random Variation
13	Reaction Kinetics	Slow Drift
14	Reactor Cooling Water Valve	Sticking
15	Condenser Cooling Water Valve	Sticking
16	Unknown	Unknown
17	Unknown	Unknown
18	Unknown	Unknown
19	Unknown	Unknown
20	Unknown	Unknown

Table 1.2. Setpoint Changes

Production Rate Change	Step	-15 %
Product Mix Change	Step	50 G/50 H to 40 G/60 H
Reactor Operating Pressure Change	Step	-60 kPa
Purge Gas Composition of Component B Change	Step	+2 %

- (1) The vapors all behave ideally;
- (2) The vapor pressures are calculated from the Antoine equation and the vapor liquid equilibrium follows Raoult's law;
- (3) The vessels are well mixed, hence all models are lumped parameter models;
- (4) The flow variables which have units kscmh, kg/hr, m³/hr are not functions of either upstream or downstream pressure. The flow variables which are listed as percentage valve position are functions of pressure;
- (5) The speed of reactor agitation affects only the heat-transfer coefficient;
- (6) The flow through the compressor and the inlet-outlet pressure difference follow a typical centrifugal compressor curve;
- (7) The noise on the measurement variables is the gaussian type with typical standard deviation.

CHAPTER 4

DESCRIPTION OF THE DYNAMIC PROCESS SIMULATOR

The Tennessee Eastman process simulator is a dynamic simulator of the process. In spite of many assumptions, it captures the dynamic behavior accurately enough to be a valid testing ground for control studies. The simulator that Tennessee Eastman has supplied is not very rigorous in the sense that it leaves out certain fast acting dynamic elements such as transmitter lags. However, it retains valve lags. Thus, there is a tradeoff between capturing the essential dynamics and having a complicated stiff system of differential equations to solve.

The reactor, the condenser, the separator, the recycle compressor and the stripper are very important from a dynamic standpoint.

4.1.The Reactor

The feed to the reactor consists of the recycle stream, the stripper overhead, and the A,D,E feeds. An input/ output scheme for the reactor is illustrated in Fig 4.1. The separator overheads are compressed, recycled and join the A&C feed. Both these streams contain all eight components, with A and C being present in the largest proportion. The A,D,E streams are almost totally pure component streams but each contains a small amount of inert component. The reactor model

Figure 4.1 Reactor input/output scheme

consists of eight component balance equations, an overall material balance and an energy balance around the reactor. The component balance equations are

$$\frac{dM_A}{dt} = F_6 \times C_{6A} - F_7 \times C_{7A} - RR(1) - RR(2) - RR(3); \quad (4.1)$$

$$\frac{dM_B}{dt} = F_6 \times C_{6B} - F_7 \times C_{7B}; \quad (4.2)$$

$$\frac{dM_C}{dt} = F_6 \times C_{6C} - F_7 \times C_{7C} - RR(1) - RR(2); \quad (4.3)$$

$$\frac{dM_D}{dt} = F_6 \times C_{6D} - F_7 \times C_{7D} - RR(1) - 1.5 \times RR(4); \quad (4.4)$$

$$\frac{dM_E}{dt} = F_6 \times C_{6E} - F_7 \times C_{7E} - RR(2) - RR(3); \quad (4.5)$$

$$\frac{dM_F}{dt} = F_6 \times C_{6F} - F_7 \times C_{7F} + RR(3) + RR(4); \quad (4.6)$$

$$\frac{dM_G}{dt} = F_6 \times C_{6G} - F_7 \times C_{7G} + RR(1); \quad (4.7)$$

$$\frac{dM_H}{dt} = F_6 \times C_{6H} - F_7 \times C_{7H} + RR(2); \quad (4.8)$$

where F_6 - the molar flow rate of the stream entering the reactor,

F_7 - the molar flow rate of the stream leaving the reactor,

$RR(1), RR(2),$
 $RR(3), RR(4)$ - the molar reaction rates of the four reactions on page 5,

$C_{6A}, C_{6B},$
 $C_{6C}, C_{6D},$
 $C_{6E}, C_{6F},$
 C_{6G}, C_{6H} - the mole fractions in the reactor inlet stream,

$C_{7A}, C_{7B},$
 $C_{7C}, C_{7D},$
 $C_{7E}, C_{7F},$
 C_{7G}, C_{7H} - the mole fractions in the reactor outlet stream,

$M_A, M_B,$
 $M_C, M_D,$
 $M_E, M_F,$
 M_G, M_H - molar holdups of each component in the reactor.

The reaction rates are expressed as an Arrhenius expression and are also functions of the partial pressures of each component in the reactor. The partial pressure exponents show the nonlinear dependence of reaction rates on each of the

components. They are expressed by the following equations.

$$RR(1) = K1 \times \exp\left(-\frac{E1}{R \times T}\right) \times PA^a \times PC^c \times PD^d \times RV; \quad (4.9)$$

$$RR(2) = K2 \times \exp\left(-\frac{E2}{R \times T}\right) \times PA^{a'} \times PD^{d'} \times PE^e \times RV; \quad (4.10)$$

$$RR(3) = K3 \times \exp\left(-\frac{E3}{R \times T}\right) \times PA^{a1} \times PD^{d1} \times RV; \quad (4.11)$$

$$RR(4) = K4 \times \exp\left(-\frac{E4}{R \times T}\right) \times PA^{a1'} \times PE^{e1'} \times RV; \quad (4.12)$$

where $RR(I)$ - the rate of reaction I,

$K1, K2,$

$K3, K4$ - the specific reaction rate constants of the four reactions,

$E(I)$ - the activation energy of reaction I,

T - the reactor temperature,

PA - the partial pressure of component A,

PC - the partial pressure of component C,

PD - the partial pressure of component D,

PE - the partial pressure of component E,

$a, c, d,$

$e, a1, d1$ - the exponents of the partial pressures.

Since the streams A,B,C are noncondensable throughout the system, the partial pressures of these components at the reactor exit are modelled by the ideal gas equation

$$P(I) \times V = N \times R \times T \quad (4.13)$$

where $P(I)$ - the component partial pressure,

T - the reactor temperature,

N - the total number of moles leaving the reactor,

R - the universal gas constant,

V - the reaction volume.

The reaction volume as modeled in the simulator is the vapor phase volume. It is obtained from the subtracting the reactor liquid volume from the reactor volume. The reactor level is calculated from the reactor liquid volume. The reactor liquid holdup is the sum of the individual liquid holdups of the condensable components, that is D,E,F,G and H.

$$F_T = F_C + F_D + F_E + F_F + F_G + F_H \quad (4.14)$$

where F_T is the total number of moles in the liquid phase,

F_C - the total number of moles of C in the liquid phase,

F_D - the total number of moles of D in the liquid phase,

F_E - the total number of moles of E in the liquid phase,

F_F - the total number of moles of F in the liquid

phase,

FG - the total number of moles of G in the liquid
phase,

FH - the total number of moles of H in the liquid
phase.

The individual liquid holdups are obtained by integrating the component balance equations. Thus the mole fractions in the liquid phase are

$$X(I) = \frac{F(I)}{F_T} \quad (4.15)$$

where $F(I)$ - the component holdup,

F_T - the total liquid holdup.

The vapor phase compositions are obtained by applying Raoult's law.

$$P(I) = X(I) \times PVAP(I) \quad (4.16)$$

$$Y(I) = \frac{P(I)}{P_T} \quad (4.17)$$

where $P(I)$ - the partial pressure,

$PVAP(I)$ - component vapor pressure,

P_T - the reactor pressure.

The reactor pressure is the sum of the partial pressures of the noncondensable and the condensable portions, the noncondensable components being A, B, and C.

$$PT = PA+PB+PC+PD+PE+PF+PG+PH , \quad (4.18)$$

The energy balance equation around the reactor is

$$\rho \times C_p \times V \times \frac{dT}{dt} = HS6 \times FM6 - HS7 \times FM7 + HTR + F \times C_{pw} \times (T_{IL} - T_{OL}) \quad (4.19)$$

- where ρ - the density,
- C_p - the specific heat of the gaseous mixture,
- V - the reaction volume,
- $HS6$ - the enthalpy at the reactor inlet,
- $HS7$ - the enthalpy at the reactor outlet,
- $FM6$ - the molar inlet flow,
- $FM7$ - the molar outlet flow,
- HTR - the heat of reaction,
- F - the flow of cooling water,
- C_{pw} - the specific heat of cooling water,
- T_{IL} - the temperature at the cooling water flow inlet,
- T_{OL} - the temperature at the cooling water flow outlet.

The molar density is calculated from the reaction volume, the reaction temperature, and the compositions. The specific heats are calculated at the reaction temperature. The products from the reactor move to a condenser and to a vapor liquid separator. No friction losses have been modelled and the condenser temperature is assumed to be operating at the same

temperature and pressure as the separator. The condenser partially condenses the reactor overheads and therefore is modeled as a partial condenser.

4.2.The Condenser

The condenser is simply modeled by an energy balance for the condenser cooling water. The heat lost to the cooling water appears in the energy balance for the separator. The energy balance for the cooling water is

$$QC = F \times C_{pw} \times (TC_{inlet} - TC_{outlet}) \quad (4.20)$$

where F - the cooling water flow rate,
 C_{pw} - the specific heat of the cooling water,
 TC_{inlet} - the condenser cooling water inlet temperature,
 TC_{outlet} - the cooling water outlet temperature.

4.3.The Separator

The lighter unreacted components from the separator are recirculated while the heavier liquid products are sent to the stripper. The vapor/liquid split of the condensable components are calculated from Raoult's law. The separator temperature is calculated from an energy balance. The separator pressure is calculated as a summation of the partial pressures of all the components. The dynamic component balances around the separator are given by the following equation.

$$\frac{dM_i}{dt} = F_8 \times C_{8i} - F_{pr} \times C_{pri} - F_r \times C_{ri} - F_{10} \times C_{10i} \quad (4.21)$$

where $i = A, B, C, D, E, F, G, H$

F_8 - the separator inlet molar flow rate,

F_p - the purge molar flow rate,

F_r - the recycle molar flow rate,

F_{10} - the separator underflow molar flow rate,

C 's - the compositions of the particular

component in the corresponding stream.

The energy balance around the separator takes into account the heat lost to the condenser cooling water.

$$\rho_s \times C_v \times V \times \frac{dT}{dt} = F_{m8} \times H_{s8} - F_p \times H_{sp} - F_{m_r} \times H_{sr} - F_{10} \times H_{s_{10}} - QC \quad (4.22)$$

where ρ_s - the density,

C_v - the specific heat,

V - the holdup,

H_{s8} ,

H_{s10} - the enthalpies of the corresponding streams,

F_{m8} - the flow rate at the reactor outlet,

QC - the condenser heat load. (Sec 4.2.)

The separator underflow enters the stripper where it is stripped by the A&C feed. The separator overheads are compressed and recycled. The separator liquid density is calculated from the separator temperature and the separator

underflow liquid compositions. The specific heats are evaluated from the empirical relations.

4.4.The Stripper

The stripper is modelled by an energy balance and component material balances. An input output scheme for the stripper is shown in Fig 4.2. The stripper temperature is calculated by integrating the energy balance equation. The stripper pressure is modelled as the sum of partial pressures. The liquid holdup in the stripper is obtained by determining the amount of stripping by the A&C feed. The separation factor is the ratio of component in the stripper overhead to that in the feed. It is given by an empirical relation of the gas/liquid ratio entering the stripper and the stripper temperature. The empirical coefficients are different for each component.

$$U = \frac{G}{L} \times \frac{a}{(b-T)} \quad (4.23)$$

$$S(i) = \frac{k_i \times U}{(1.+k_i \times U)} \quad (4.24)$$

where a,b - empirical coefficients for temperature
 correlation,
 k_i - empirical coefficient for ith component,
 T - stripper temperature,
 G/L - gas/liquid ratio.

Figure 4.2 Stripper input/output scheme

The compositions in the stripper overhead are obtained by multiplying the inflow of each component by the separation factor. The energy balance around the stripper reduces to the following.

$$\rho \times C_v \times V \times \frac{dT}{dt} = F_{m_{10}} \times H_{s_{10}} + F_{m_4} \times H_{s_4} - H_{s_5} \times F_{m_5} - H_{s_{11}} \times F_{s_{11}} \quad (4.25)$$

where ρ - the density of the stripper holdup,
 C_v - the specific heat of the stripper holdup,
 V - the stripper holdup,
 $F_{m_{10}}$ - the separator molar underflow rate,
 F_{m_4} - the A&C feed molar flow rate,
 F_{m_5} - the stripper overhead molar flow rate,
 $F_{m_{11}}$ - the stripper underflow molar flow rate,
 $H_{s_4}, H_{s_{11}},$
 $H_{s_5}, H_{s_{10}}$ - the enthalpies of the corresponding streams.

4.5.The Compressor

The compressor is modelled by an empirical equation which relates the inlet/discharge pressure ratio to the flow through the compressor, following a typical centrifugal compressor curve. The inlet pressure is the separator pressure and the outlet pressure is the reactor pressure since there are no frictional or other losses modelled. The compressor work is a measured variable and is modelled as follows.

$$pr = \frac{p_{sep}}{prct} \quad (4.26)$$

$$CW = a' \times (1. - pr^{**}b') \quad (4.27)$$

where CW is the compressor work,
 pr is the pressure ratio,
 a', b' are empirical coefficients,
 psep is the separator pressure,
 prct is the reactor pressure.

4.6.Mixing of Streams

Another dynamic material balance describes the mixing of the A, D, E feed streams, the stripper overhead and the recycle stream before the reactor inlet.

$$\frac{dM_i}{dt} = F_r \times C_{ri} + F_{sto} \times C_{stoi} + F_A \times C_{Ai} + F_D \times C_{Di} + F_E \times C_{Ei} - F_6 \times C_{6i} \quad (4.28)$$

where F_r - the molar flow rate of the recycle stream in the mixing stream,
 M_i - molar holdup for ith component,
 C_{ri} - the mole fraction of component in recycle stream,
 F_{sto} - the molar flow rate of the stripper overhead,
 C_{stoi} - the mole fraction of component in the stripper overhead.

CHAPTER 5

APPROXIMATE MODELS

Approximate models are developed to capture essential aspects of the process while being based upon available process data. The general approach to the development of approximate models is accompanied by the two steps of model development and validation. A review of nonlinear approximate models for the dual purposes of process control and optimization is presented by Riggs and Rhinehart (1989). They stress the fact that although a large variety of nonlinear control laws and model parameter adaptation techniques are available, the validation of approximate models determines the failure or success of its implementation. The model should be simple enough to maintain and be computationally efficient. Approximate models were developed for the reactor and stripper only.

The approach to reactor model development is to model it using steady-state material and energy balances, while modeling the kinetics over the entire range to describe the nonlinear nature. The holdups are also modelled and the reactor pressure is calculated as a sum of the partial pressures. The manipulated variable is the reactor cooling water flow rate and is calculated from an energy balance.

The approach to the stripper development was to model it using steady-state material and energy balances. This was

accomplished by defining a separability factor for the condensable components D,E and F and modeling the overheads as this factor times the net amount of each component entering the system. The manipulated variable for the stripper is the stripper steam flow estimated from an energy balance.

5.1.The Reactor

Development of approximate model equations: The reactor model consists of eight component balance equations, an overall material balance and an energy balance around the reactor. The assumptions used in modeling the reactor were

- (1) The noncondensable components follow the ideal gas law.
- (2) The partial pressures for the condensable components follow Raoult's law.
- (3) The heats of reaction are constant and the heats of the byproduct reactions are negligible.
- (4) The reaction volume is the vapor phase volume.

The model equations were developed using the following procedure. First, the partial pressures were estimated from the reactor inlet compositions. The reaction rates then were calculated from the partial pressures and the reactor temperature. The heat of reaction was estimated from the reaction rates and in turn, the rate of heat removal from the reactor was calculated from a steady-state energy balance. The inlet and outlet enthalpies themselves were calculated as

empirical functions of compositions and temperature and the steady-state component balance equations are as follows.

$$F_7 \times C_{7A} = F_6 \times C_{6A} - RR(1) - RR(2) - RR(3) \quad (5.1)$$

$$F_7 \times C_{7B} = F_6 \times C_{6B} \quad (5.2)$$

$$F_7 \times C_{7C} = F_6 \times C_{6C} - RR(1) - RR(2) \quad (5.3)$$

$$F_7 \times C_{7D} = F_6 \times C_{6D} - RR(1) - 1.5 \times RR(4) \quad (5.4)$$

$$F_7 \times C_{7E} = F_6 \times C_{6E} - RR(2) - RR(3) \quad (5.5)$$

$$F_7 \times C_{7F} = F_6 \times C_{6F} + RR(3) + RR(4) \quad (5.6)$$

$$F_7 \times C_{7G} = F_6 \times C_{6G} + RR(1) \quad (5.7)$$

$$F_7 \times C_{7H} = F_6 \times C_{6H} + RR(2) \quad (5.8)$$

where F_6

- the molar flow rate of the stream
entering the reactor;

F_7

- the molar flow rate of the stream
leaving the reactor;

$RR(1), RR(2),$

$RR(3), RR(4)$ - the molar reaction rates;

$C_{6A}, C_{6B},$

$C_{6C}, C_{6D},$

$C_{6E}, C_{6F},$

C_{6G}, C_{6H} - the molar flow rates in the reactor inlet stream;

$C_{7A}, C_{7B},$

$C_{7C}, C_{7D},$

$C_{7E}, C_{7F},$

C_{7G}, C_{7H} - the molar flow rates in the reactor outlet stream.

The reaction rates are expressed in the following manner.

$$RR(1) = K1 \times \exp\left(-\frac{E1}{R \times T}\right) \times PA^a \times PC^c \times PD^d \times RV \quad (5.9)$$

$$RR(2) = K2 \times \exp\left(-\frac{K2}{R \times T}\right) \times PA^{a'} \times PC^{c'} \times PE^e \times RV \quad (5.10)$$

$$RR(3) = K3 \times \exp\left(-\frac{E3}{R \times T}\right) \times PA^{a1} \times PD^{d1} \times RV \quad (5.11)$$

$$RR(4) = K4 \times \exp\left(-\frac{E4}{R \times T}\right) \times PA^{a1'} \times PE^{e1'} \times RV \quad (5.12)$$

where $RR(I)$ - the rate of reaction I ;

$K_1, K_2,$
 K_3, K_4 - the specific reaction rate constants of the four reactions;
 $E(I)$ - the activation energy of reaction I;
 T - the reactor temperature;
 P_A - the partial pressure of component A;
 P_C - the partial pressure of component C;
 P_D - the partial pressure of component D;
 P_E - the partial pressure of component E;
 $a, c, d,$
 e, a_1, d_1 - the exponents of the partial pressures.

Since the streams A,B,C are noncondensable throughout the system, the partial pressures of these components at the reactor exit are modelled by the ideal gas equation

$$P(I) \times V = N \times R \times T \quad (5.13)$$

where $P(I)$ - the component partial pressure;

T - the reactor temperature;
 N - the total number of moles of each component in the reactor holdup;
 R - the gas constant;
 V - the reaction volume.

The total number of moles leaving the reactor is estimated from a material balance since we know that all of the noncondensables are either recycled or purged. The reaction volume is the vapor phase volume. It is obtained from

subtracting the reactor liquid volume from the reactor volume. The reactor liquid volume is calculated from the measured reactor level. The reactor liquid holdup is the sum of the individual liquid holdups of the condensable components, that is D,E,F,G and H.

$$FT = FD + FE + FF + FG + FH \quad (5.14)$$

where FT - the total number of moles in the liquid phase;

FC - the total number of moles of C in the liquid phase;

FD - the total number of moles of D in the liquid phase;

FE - the total number of moles of E in the liquid phase;

FF - the total number of moles of F in the liquid phase;

FG - the total number of moles of G in the liquid phase;

FH - the total number of moles of H in the liquid phase.

The individual liquid holdups are first-order filtered each time the controller is called. During the transient phase, the component balances will not be at steady state. By calculating the new holdups and filtering, an approximation of the holdups is done periodically. In this way, an approximation for the compositions is made up for the transient phase. In the test case when there is a product grade change, the compositions of each individual component in the liquid phase is different. Thus, the mole fractions of each component in the liquid phase are given by

$$X(I) = \frac{F(I)}{F_T} \quad (5.15)$$

where $F(I)$ - the component holdup,

F_T - the total liquid holdup.

The vapor-phase compositions are obtained by applying Raoult's law.

$$P(I) = X(I) \times PVAP(I) \quad (5.16)$$

$$Y(I) = P(I) \frac{1}{P_T} \quad (5.17)$$

where $P(I)$ - the partial pressure

$PVAP(I)$ - component vapor pressure

P_T - the reactor pressure.

The reactor pressure is the sum of the partial pressures of the noncondensable and the condensable portions.

$$P_T = P_A + P_B + P_C + P_D + P_E + P_F + P_G + P_H \quad (5.18)$$

5.1.1. Evaluation of fixed parameters

The fixed parameters, the exponents of the partial pressures in the kinetic equations, were determined in the following manner. A series of runs was first conducted with isothermal operation of the reactor for different component partial pressures. The measured exit compositions were carefully filtered. The reactor was maintained at constant

temperature for this series of runs. The reaction rates RR1 and RR2 were estimated. Since the rate of production of both the products G and H are these reaction rates, an overall material balance around the system gives us these reaction rates.

$$RR(1) = X_{PG} \times F_{PG} + X_{PRG} \times F_{PR} \quad (5.19)$$

$$RR(2) = X_{PH} \times F_{PH} + X_{PRH} \times F_{PR} \quad (5.20)$$

where X_{PG} - the G composition in the purge,
 X_{PH} - the H composition in the purge,
 F_{PG} - the purge molar flow rate,
 X_{PRG} - the G composition in the product,
 X_{PRH} - the H composition in the product,
 F_{PR} - the product molar flow rate.

The series of runs was conducted at different PA's, PC's, PD's and PE's. These partial pressures were estimated from equations (5.13) and (5.16). Since running the reactor at different partial pressures would lead to different holdups in the reactor, there has to be a correction for the reactor volume. About 12 such runs were performed with the reaction rates corrected for reaction volume change. The compositions of G and H in the purge and product were filtered with a first order filter and extreme care was taken to ensure that these were accurate values. The process was also run at different

product grades 50/50, 40/60, 43/57, and 47/53 which basically meant that D & E feeds were run at different ratios to the C/A feed. A multiple linear regression of the following equations yielded the partial-pressure indices a, c, d, and e. The data are given in Tables 5.1. and 5.2.

$$LN\left(\frac{RR(1)}{RR10} \times \frac{RVO}{RV}\right) = a \times \left(\frac{PA}{PAO}\right) + LN\left(\frac{PC}{PCO}\right) + LN\left(\frac{PD}{PDO}\right) \quad (5.21)$$

where RR10, and RVO are reaction rates and reaction volume at base case, and PAO, PCO, PDO, and PEO are base case partial pressures. The exponents are calculated similarly for the second equation. To estimate the reaction rate constants and the activation energy constants, the reactor was run at five different temperatures while maintaining the feed compositions constant by manipulating the feed rates. The reaction rates were estimated using overall material balances on G and H with correction for reaction volume. The partial pressures were noted and since the indices were calculated, a plot of the logarithm of the reaction rate versus 1/T yielded as the slope $-E_1/R$, and as the intercept $\ln(K_1)$ and $\ln(K_2)$. The activation energy calculated from the slopes were -40000.81 and -19999.95. The constants are 31.578 and 3.00. The plots are shown in Figs. 5.1 and 5.2.

The energy balance equation around the reactor is used to calculate the heat load which in turn is used to calculate the cooling water flow rate, F.

Table 5.1 Isothermal Runs

PA	PC	PD	RV	RR(1)
5095.6	3559.4	225.99	818.3	240.41
5100.	3678	225.90	834.2	245.35
5150.51	3721.98	226.345	787.5	229.45
5201.99	3784.70	226.649	777.6	237.43
5323.98	3794.64	226.32	799.3	241.85
5382.95	3873.48	245.	758.3	267.27
5398.10	3916.35	226.7	758.00	246.19
5405.83	3827.53	226.86	757.43	250.16
5406.7	3932.21	226.876	757.74	250.47
5530.93	3993.02	225.91	764.20	258.91
5630.2	4023.01	126.00	741.75	242.13

Table 5.2 Isothermal Runs Continued

PA	PC	PE	RV	RR(2)
5090.	3560.32	3530.19	814.3	198.41
5102.31	3634.678	3576.90	824.2	205.46
5245.60	3721.8	3595.35	737.5	192.77
5301.99	3780.1	3630.65	748.6	201.18
5323.99	3790.5	3670.32	759.3	207.49
5382.95	3816.4	3710.00	758.3	213.87
5398.1	3873.4	3725.7	758.0	214.07
5405.84	3920.4	3730.59	759.4	217.65
5406.7	3935.21	3730.88	759.7	217.92
5530.93	3995.01	3730.91	764.2	226.25
5923.1	4015.44	3730.00	741.7	143.68

Table 5.3 Regressed Parameters

PARAMETER	REGRESSED VALUES
a	1.120
c	.410
d	1.120
e	1.120
a1	1.000
d1	1.000
e1	1.000

Figure 5.1 Plot of logarithm of first reaction rate versus reciprocal of the temperature.

Figure 5.2 Plot of logarithm of second reaction rate versus reciprocal of the temperature.

$$Q = -(HS6 \times FM6 - HS8 \times FM8 + HR1 \times RR1 + HR2 \times RR2) \quad (5.22)$$

where Q - the rate of heat removal from the reactor,

$HS6$ - the enthalpy at the reactor inlet,

$HS8$ - the enthalpy at the reactor outlet,

$FM6$ - the molar inlet flow,

$FM8$ - the molar outlet flow,

$HR1$ - the heat of reaction 1,

$HR2$ - the heat of reaction 2.

The heat of reaction is the sum of the individual heats of reaction for the first two reactions. The enthalpies are calculated as empirical functions of the compositions of the streams and the reactor temperature. Q calculated in equation 5.22. is used in equation 5.23 for the flow of cooling water.

$$F = \frac{Q}{C_{pw} \times (T_{inlet} - T_{outlet})} \quad (5.23)$$

where F - the flow of cooling water,

C_{pw} - the specific heat of cooling water,

T_{inlet} - the temperature at the cooling water inlet,

T_{outlet} - the temperature at the cooling water flow outlet.

5.1.2. Parameterization of the approximate model

The approximate model must adapt to changes caused by the various disturbances and setpoint changes and the resulting

process/model mismatch has to be eliminated. A model that is not true to the process behavior may not yield good controller performance. Suitable adjustable parameters have to be chosen in order to adapt the process to changing process conditions. The parameter with the highest degree of uncertainty, the specific reaction rate constant of the first reaction, is chosen as the adjustable parameter for the approximate model of the reactor. The technique of online incremental parametrization discussed by Riggs and Rhinehart (1990) for steady state models and especially applied to an open-loop unstable CSTR is employed. The difference between the temperature calculated from the model, that is, the steady state energy balance equation and the measured temperature from the simulator, is to be driven to zero.

$$DEV = TCAL - TMEAS. \quad (5.24)$$

The specific reaction rate constant is adjusted according to the following equation.

$$CK1 = CK1 + \alpha \times \frac{DEV}{\frac{\partial T}{\partial CK1}} \quad (5.25)$$

$$\frac{\partial T}{\partial CK1} = -E1 \times \frac{CK1}{R \times T^2}, \quad (5.26)$$

where α is the relaxation factor to compensate for Newton linearization. The variation of the adjustable parameter with

changes in process conditions is shown in the graphs along with the control results. Fig 5.3 shows variation of the reactor model parameter with product grade setpoint change. The reaction rate constant for the first reaction decreases since we are making a lower amount of G and there is a sharp undershoot before it levels off to its steady state value. Fig 5.4 shows the variation to be much smoother. Since there is a decrease in the pressure setpoint, there is again a decrease in the reaction rate constant to drive the temperature difference calculated from the model and the measured process temperature to zero. Since selectivity has to be maintained, the specific reaction rate of the second reaction is varied as the base case ratio of the reaction rates times the specific reaction rate of the second reaction. Since it is difficult to calculate the flow rate needed to maintain a certain reactor pressure, this approach was not used. Moreover, it is desirable to choose a model that uses as few parameters as possible.

5.2.The Stripper

The stripper is modelled by an energy balance and a component balance on D,E and F. The assumptions used while modelling the stripper were the following.

- (1) The vapors are ideal gases.
- (2) The condensable components follow Raoult's law.
- (3) The temperature and pressure are constant throughout the column.

Figure 5.3 Variation of reactor model parameter with product grade setpoint change

Figure 5.4 Variation of reactor model parameter with pressure setpoint change

(4) The flow rates of liquid and vapor are constant throughout the column.

The stripper is modeled by component balances on the condensable components D,E and F and an energy balance. The general approach has been to derive a linear relationship between an index of separation (the separation factor S), and the gas/liquid ratio. This gives us an estimate of the overhead composition and also the temperature adjustment required for a changing gas/liquid ratio. The steam flow required is estimated from an energy balance around the stripper. The enthalpies are calculated from empirical relationships with the temperature and compositions.

Since the noncondensable components leave in the same amount as they enter and the heaviest products G and H leave the stripper bottom being negligibly stripped, an approximation is done by neglecting them. The D,E and F in the stripper overhead is obtained by determining the amount of stripping by the A&C feed. The separation factor is an empirical relation of the gas/liquid ratio entering the stripper and the stripper temperature. The empirical coefficients are different for each component. Since this factor is dependent on the gas/liquid ratio entering the stripper and the stripper temperature, we had to develop an approximate equation for the separation factor/temperature dependence. The separation factor is estimated by dividing the molar flow rates of each individual component in the stripper

overhead by the total inflow of each component from both the separator underflow and C/A feed. The empirical coefficients were obtained from a linear approximation of the separation factor and temperature relationship. For the operating range of the stripper and for a constant gas/liquid ratio, the factor shows a linear dependence on the temperature

$$S = \frac{G}{L} \times [a(i) + b(i) \times T], \quad (5.27)$$

where a_i , b_i are the linearized coefficients. These coefficients are the intercept and slope of plots of $S/(G/L)$ versus the stripper temperature. The stripper is run at different G/L ratios. The plots and data are shown in Figs. 5.5 to 5.7 and Tables 5.4 to 5.7. The compositions in the bottoms flow were obtained from subtracting the molar flow rates of each individual component from the total molar flow rates of each component entering the stripper. The energy balance of the stripper reduces to the following equation

$$q_s = -(F_{m10} \times H_{s10} + F_{m4} \times H_{s4} - F_{m5} \times H_{s5} - F_{m11} \times H_{s11}) \quad (5.28)$$

where F_{m10} is the separator underflow, F_{m4} is the A&C feed, F_{m5} is the stripper overhead molar flow rate, F_{m11} is the stripper underflow molar flow rate, H_{s4} , H_{s11} , H_{s5} , H_{s10} are the enthalpies of the corresponding streams.

Since the reactor pressure is controlled by the gas feed to the stripper, we have to compensate for the varying vapor/liquid ratio by the temperature, since the separation is

Figure 5.4 Data for D Component

Stripper Temperature	G/L	S/(G/L)
65.0	1.608	.105052
67.5	1.608	.014827
70.0	1.608	.014293
72.5	1.608	.013991
66.0	1.708	.014107
68.5	1.708	.014311
70.0	1.708	.014008
67.5	1.808	.014643
69.5	1.808	.014123

Figure 5.5 Data for E Component

Stripper Temperature	G/L	S/ (G/L)
65.0	1.608	.105144
67.5	1.608	.014772
70.0	1.608	.014245
72.5	1.608	.013754
66.0	1.708	.015113
68.5	1.708	.014561
70.0	1.708	.014249
67.5	1.808	.014891
69.5	1.808	.014356

Figure 5.6 Data for F Component

Stripper Temperature	G/L	S/(G/L)
65.0	1.608	.105135
67.5	1.608	.014577
70.0	1.608	.014062
72.5	1.608	.013580
66.0	1.708	.014930
68.5	1.708	.014383
70.0	1.708	.014077
67.5	1.808	.014714
69.5	1.808	.014189

**Table 5.7 Tabulated values for
for a & b**

Component	a	b
D	-.00020	.028007
E	-.00021	.028699
F	-.00020	.028027

Figure 5.5 Plot of separation factor versus G/L for component D.

Figure 5.6 Plot of separation factor versus G/L for component E.

Figure 5.7 Plot of separation factor versus G/L for component F.

an empirical function of the vapor/liquid ratio and the temperature. The amount of separation in turn is set by a low gain feedback controller on the stripper byproduct F composition. The amount of steam needed is calculated by the energy balance around the stripper.

There is no adjustable parameter for the stripper model that is driving the difference between model calculated values and measured value to zero. It is used more in a feedforward sense that it anticipates disturbances and the temperature is adjusted. The steam flow required to maintain this stripper temperature is calculated from an energy balance. The feedback control on the stripper F product composition to maintain it at setpoint is achieved by using the separation factor as a manipulated variable. This loop acts in a slower fashion and has a much lower gain compared to the faster adjustment to the temperature. The energy balance is used to calculate the amount of steam flow required.

CHAPTER 6

CONTROL STRATEGIES

The main control objective is to maintain product compositions to within 5% absolute variation about the product specifications. Since the C/A feed and other feeds are the product of upstream processes and have very little holdup, variation of more than 5% over a 5 hr period is to be avoided.

6.1. Base Level Controls

The control algorithms use both model based control and PI controls (for the level and the flow controllers for the feeds). The level of the separator is controlled by the separator underflow using a PI controller, and the purpose of this controller is to absorb short term variations in liquid flow to the stripper. An averaging inventory control is required in order to absorb upstream upsets and not pass them on to downstream equipment. The stripper level is controlled by the stripper underflow. This controller has to be tuned in such a way that the fluctuations in the stripper underflow are held to a minimum. The stripper is coupled with the reactor since changes in reactor pressure adjust the C/A feed to the stripper. The stripper temperature is adjusted accordingly, resulting in varying steam flow which affects the stripper level. Thus, the stripper level controller has to be detuned accordingly.

The feed flows are controlled by the feed flow valves. PI flow controllers and the setpoints to these controllers are constant base-case ratios to the C/A flow. The controllers act on the filtered flow measurements.

6.2.Reactor Control Strategies

There are two controlled variables on the reactor, the reactor temperature and pressure. If reactor temperature and reactor pressure were maintained at their setpoints, most of the unstable nature of the system would be removed. Each of the disturbances and setpoint changes affects the controlled variables in various ways.

6.2.1.Reactor temperature control

The reactor temperature is controlled best by the reactor cooling water as shown in Fig 6.1. The control law is developed from the essentials of nonlinear process control. Let the reactor temperature at the present conditions be T and the desirable operating temperature is T_{sp} and we want to move the process from T to T_{sp} in a time interval τ . The forward difference yields

$$\frac{dT}{dt} = \frac{T_{sp} - T}{\tau}. \quad (6.1)$$

The controller calculates a steady-state target value based on the error $T_{sp} - T$. The PMBC law is given by the following

Figure 6.1 Schematic of reactor temperature control

equation. Since we are using a steady state material balance equation, we use this to calculate the reactor cooling water flow rate using the temperature set up by the trajectory.

$$T_{ss} = T + K_1 \times (T_{sp} - T) . \quad (6.2)$$

The model develops a steady state offset between model parametrizations and in order to remove this offset, addition of an integral term is required. The control law becomes

$$T_{ss} = T + K_1 \times (T_{sp} - T) + K_2 \times \int (T_{sp} - T) dt . \quad (6.3)$$

The integral term is analogous to the PI integral term and the purpose is to eliminate offset. The above expression is an extension of the PMBC law, and is called Generic Model Control law (Lee and Sullivan, 1988). T_{ss} is used in the steady state energy balance to calculate the manipulated value. The reactor inlet and outlet enthalpies, the heat of reaction are calculated at this temperature.

$$Q_{CW} = -HTR - F_{INL} \times H_{INL} + F_{OUT} \times H_{OUT} \quad (6.4)$$

where Q_{CW} - the heat lost to cooling water,
 HTR - the heat due to reaction,
 F_{INL} - the molar flow rate at the inlet,
 H_{INL} - the enthalpy of the inlet stream,
 F_{OUT} - the reactor exit molar flow rate,
 H_{OUT} - the reactor exit stream enthalpy.

The heat of reaction is calculated as the sum of the heats of reaction of the first two reactions as in equation 5.23 in the chapter on approximate models. It uses online composition measurements, the temperature measurement, and the level measurement for reaction volume. The cooling water flow rate is calculated by a steady state energy balance around the heat exchanger. The inlet temperature to the heat exchanger is estimated from the heat removal and a filtered heat generation rate.

$$T_{INL} = T_{OUT} - \frac{Q_{CW}}{F_{CW} \times C_{PW}} \quad (6.5)$$

where Q_{CW} is the heat removal rate, T_{OUT} is the cooling water outlet temperature, F_{CW} is the cooling water flow rate and C_{PW} is the specific heat. The outlet temperature is a measured variable. The flow rate of cooling water is calculated from the following equation

$$F_{CW} = \frac{Q_{CW}}{C_{PW} \times (T_{OUT} - T_{INL})} \quad (6.6)$$

where F_{CW} is the flow rate of the cooling water, T_{OUT} is the heat exchanger outlet temperature, T_{INL} is the heat exchanger inlet temperature. Thus, changes in the cooling water inlet temperature are compensated by the above algorithm.

6.2.2.Reactor pressure Control

The composition disturbances cause the reactor pressure to reach shutdown conditions and also reduce reaction rates affecting the level. Since A,B,C are noncondensable, excessive accumulation of these in the recycle loop has to be avoided since they would drive up the pressure.

A sensitivity analysis of the system identified that the reactor pressure was most sensitive to the C/A feed. The reactor pressure does not show much variation under some of the test scenarios but reacts immediately to composition upsets. Composition upsets affect selectivity and reaction rates. The reactor pressure is thus controlled by the C/A feed, and to maintain stoichiometry, the A, D, & E flows are maintained at base case ratios to the C/A feed as shown in Fig 6.2. The C/A feed has the quickest effect on the reactor pressure and can maintain the reactor pressure at its setpoint under the most severe composition disturbances. Thus, the gain directionality is good in the sense that small input changes can be reflected in large changes in output. The reactor pressure is the sum of the partial pressures of the noncondensable components and the partial pressures of the condensable components. By adjusting the C/A ratio, both parts are adjusted. The reactor pressure is a pure PI feedback controller and it sets the C/A feed flow by a PI velocity algorithm. In order to maintain stoichiometrically correct amounts, the other feed rates are kept at base case ratios to

Figure 6.2 Schematic of reactor pressure control.

the C/A feed. These ratios multiplied by the C/A flow rate provide the setpoints for the respective feed flow controllers.

Since our system has forty-one measured variables and twelve manipulated variables, the solution to the problem is to look at the best pairing. This can be best understood by studying the interaction involved. Several alternative manipulated variables were studied to understand whether they could serve as a better choice than C/A feed.

The maximum purge flow rate is too small for controlling the reactor pressure under some of the tests. The recycle flow is internal, and since the flow is proportional to the square root of the difference in pressures between the compressor and the reactor, a change in recycle causes a back pressure to develop with the result that the separator and stripper pressures go up. This has the effect of making the system unstable. The condenser cooling water does not work well for the composition disturbances. The reactor pressure is thus controlled by the C/A feed, and to maintain stoichiometry, the A, D, and E flows are maintained at base case ratios to the C/A feed.

6.3.Stripper Control Strategies

The important function of the stripper model based controller is to maintain the product composition of component F at its setpoint. Since the condensable components are

stripped differently, and since F is a middle fraction and an undesirable product, it really determines the amount of separation we need. Inventory control of the stripper is important to reducing fluctuations in stripper product flow.

The reactor and stripper become coupled because changes in reactor pressure change the amount of C/A feed which enters the system via the stripper because of the reactor pressure control. Fluctuations in the C/A feed change the vapor/liquid ratio which alters the separability in the stripper. Since the separability is a function of both the vapor/liquid ratio and the temperature, the stripper temperature is adjusted for the changing vapor/liquid ratio in order to maintain constant separability. How much the separability itself should be is dictated by the byproduct composition. Thus, a slow feedback PI loop controls F product composition by manipulating its separation factor. A model predicting how much stripper steam should be used to compensate for a variable liquid/vapor ratio forms the basis of our control system for the stripper as shown in Fig 6.3. The separability is expressed by the following expression in terms of the vapor/liquid ratio and the stripper temperature

$$SRF = \frac{G}{L} \times (a(i) + b(i) \times T) \quad (6.7)$$

where $a(i)$, $b(i)$ are the coefficients for each of the components. For a constant gas/liquid ratio, the separation factor is approximated as a linear function of temperature.

Figure 6.3 Schematic of stripper composition controller

The temperature required for a different vapor/liquid ratio can be calculated from this empirical equation. Using this temperature, the amount of steam needed can be calculated from a steady state energy balance around the stripper. The steady state energy balance around the stripper

$$QS = -FS4 \times HS4 + FS5 \times HS5 - FS11 \times HS11 + FS13 \times HS13 \quad (6.8)$$

where QS is the heat incoming from the steam, HS4, FS4 are the enthalpy and molar flow rate of the incoming A/C feed, HS5, FS5 are the enthalpy and molar flow rate of the stripper overheads, HS11, FS11 are the enthalpy and molar flow rate of the separator underflow, and HS13, FS13 are the enthalpy and molar flow rate of the stripper underflow. The amount of steam is calculated by

$$FS = \frac{QS}{LH} \quad (6.9)$$

where FS is the molar flow rate of steam, LH is the latent heat of steam. The manipulated variable here is the steam flow, but both the flow rate of A/C feed and the composition of F in the product determine the steam flow required.

The control configuration as described above does well for most of the disturbances and setpoint changes except for two disturbances, a step change affecting a loss in A feed, disturbance no 6., and a step change in the C/A ratio with B composition constant, disturbance no 1. Thus, there is a slow feedback controller on the A/C ratio in the purge which uses the ratio of the A feed to the C/A feed.

6.4.Override Control Schemes

For a step change in the A/C ratio, the A/C ratio in the purge would open up the A valve. Since the A valve saturates, there is an override scheme which would now start opening up the purge valve to maintain the A/C ratio. The important thing here is the tuning of the individual loops concerned. The pressure loop cannot act as tightly as it does under normal circumstances since it would start cutting down on the C/A feed too fast for the other loops to act in concert. Thus the pressure controller and the A, D and E controllers are switched to lower gains. The three loops have to be tuned accordingly since the tuning constants determine whether closed loop stability has been achieved.

In the case when there is a loss in the A feed, the C/A ratio cannot put in more A so the A/C ratio in the purge acts on the purge valve in order to maintain the A/C ratio (Fig 6.4.). In this case, the pressure controller and the feed controllers are again switched to lower gains.

Figure 6.4 Schematic of the override controller

CHAPTER 7

CONTROL RESULTS

The control results illustrate the response of the main controlled variables and the variation of manipulated variables in the face of disturbances and the setpoint changes. The setpoint changes are made to the production rate, the reactor pressure, the purge B composition, and a product grade change. The disturbances are both of the step type and the random variation type.

7.1.Tuning Of Controllers

There is hierarchy of control, the model-based reactor temperature and stripper composition controls at the upper level and at the lower level, the PI flow and level controllers.

The inventory controls were tuned first, since these controllers do not require as sensitive a tuning as the other controllers and they could be adjusted later. The separator level controller is a loose averaging type controller but it has to be tuned keeping in mind that the fluctuations in the separator underflow are kept to a minimum. It is a low-gain and low reset controller. Higher gains tended to produce fluctuations in the separator underflow. The stripper controller is a moderate gain, moderate reset controller. The criteria were that we maintained levels close to setpoint

while not producing fluctuations in the manipulated variables, the outlet flows.

The reactor model based temperature controller was tuned next. The setpoint of the reactor temperature was first raised by 1 degree C and lowered by 1 degree C. The response of the model based controller was observed on a graph in terms of the temperature response. A gain of 1.0 was tried first. The response appeared to be sluggish. Higher values of gain were tried and a value of 3.0 appeared to be suitable for temperature setpoint changes. The criterion for tuning was that temperature setpoint be maintained very tightly, in a band of about 0.4 degrees about the setpoint.

Because of certain composition changes, the reactor pressure reacts almost immediately. Lower gains for the reactor pressure PI controller were tried out first. Pressure setpoints changes of +60kPa and -60kPa were made. The A,D and E flow controllers were tuned first and the pressure controller next. A gain of 1.0 for all the controllers seemed to give a very good response. The criteria were that the pressure be maintained in a tight band of about +5kPa and -5kPa about the setpoint and the flows of the feeds were not varied too frequently.

The stripper composition controller is a very low-gain controller since the gain is scaling the composition output to correct the F composition. A value of $1E-5/\text{mole\%}$ was used. The pressure controller was retuned to correct for maintaining

uniform C/A feed flow. The level controllers seemed to work fine with the original tuning parameters.

The choice of gains on the individual controllers was based on an understanding of the process. The gains of each of the control loops with respect to each other were important criteria for the process to attain stability. For example, if the pressure controller gains were increased, they would start cutting back on the feed too fast for the override controller to have any effect. In addition, the stripper steam valve would have to make too many high frequency adjustments for the changing gas/liquid ratios. The choice of tuning parameters was found through trial. A list of tuning parameters is given in Tables 7.1 and 7.2.

7.2.Setpoint Changes

The product grade change, from a 50/50 G/H grade to 40/60 G/H grade, changes the flows of D and E because the ratios of D and E to C/A are adjusted to meet new product specifications. The product grade change causes a change in selectivity, an increase in one reaction rate corresponding to the other. There is a lowering of the reactor level, although the reactor pressure and temperature are smoothly maintained at setpoint as shown. The C/A feed is cut back slightly to meet new product specifications.

The pressure setpoint change from 2705 kPa to 2645 kPa is very smooth in that the pressure easily moves to the new

Table 7.1 Controller Tuning parameters

CONTROLLER	GAIN	RESET
TEMPERATURE CONTROLLER	3.0	1.0 /sec
PRESSURE CONTROLLER	1.05 (/kpag)	5.0 /sec
A FEED CONTROLLER	1.0 (hr/kmol)	5.0 /sec
D FEED CONTROLLER	1.0 (hr/kmol)	5.0 /sec
E FEED CONTROLLER	1.0 (hr/kmol)	5.0 /sec
SEPARATOR LEVEL CONTROLLER	.10	10.0
STRIPPER LEVEL CONTROLLER	.65	5.0
A/C CONTROLLER	.01	7.0

Table 7.2 Override Tuning parameters

CONTROLLER	GAIN	RESET
A/C CONTROLLER	3.00	5.00 /sec
REACTOR PRESSURE CONTROLLER	0.01/kPag	5.00 /sec
A FEED CONTROLLER	0.01hr/kmol	5.00 /sec
D FEED CONTROLLER	0.01hr/kmol	5.00 /sec

setpoint pressure. The product compositions show almost no change as shown. The feed is cut back fast enough by the pressure controller to enable this smooth change. There is a small initial dip in temperature but it recovers smoothly to setpoint.

The production rate change causes a cut in the feed, thus lowering the amount of A, D, E feed to the reactor. In the transient phase, there is a mismatch between reactor model and simulator because of reaction rates. Thus, initially there is a change in the reactor level. However, the reactor temperature and pressure are smoothly maintained at setpoint and the compositions do not deviate from specifications. The setpoint change in purge inert component B composition, a change of +2% causes a building up of inert and thus a buildup of pressure. The feed is cut back slightly and the product compositions show almost no change. The reaction rates show almost no change.

7.3. Disturbance Rejection

The effect of the disturbances is described in the following cases to describe how each individual disturbance affects the process variables and to understand what the controller action is to retain controlled variables at setpoint.

The increase in A/C ratio, (disturbance no. 1) lowers reaction rates and causes the reactor pressure to increase

after the disturbance is activated because of accumulation of noncondensables and lowered reaction rates. The reactor temperature goes down as a result since the heat produced by the exothermic reactions is not enough to sustain reactor temperature. Lower temperature further reduces reaction rates. Reduction of reactor temperature lowers separator temperature and stripper temperature thus sending more reflux to the stripper with the stripper not retaining enough separating power. So the controller increases the amount of A and cuts back on the feed while the temperature controller decreases the cooling water flow rate. The stripper composition controller increases the separation factor needed and increases the amount of steam. Although the reactor temperature and reactor pressure are maintained at setpoint, the individual holdups in the liquid phase are not predicted dynamically very well. To maintain the separating power in the stripper, the steam shows a corresponding increase. Although the override control works, the tuning constants of the purge loop, the A feed loop and the pressure loop have to be adjusted correspondingly but the reactor level shows a large change from base case value.

The B change in the C/A feed (disturbance no. 2) would cause an accumulation of inerts and the reactor pressure to increase and the reactor temperature to decrease. The pressure controller cuts back the feed slightly and the reactor temperature cuts back on the reactor cooling water slightly.

The reactor level is not affected significantly and the reaction rates are affected slightly. The reactor temperature and pressure are maintained at setpoint and the compositions are maintained. The C/A feed is cut back to counter the increasing inert composition. The results are shown.

A step change in D feed temperature (disturbance no. 3) increases feed enthalpy slightly, increases the amount of heat into the reactor, the reactor pressure falls slightly, and the reactor temperature increases. The temperature controller increases cooling water and the pressure controller increases feed before cutting it cut back slightly and recovers to the base case value.

The change in reactor cooling water inlet temperature (disturbance no. 4) causes the reactor pressure to fall dramatically, the reactor level falls and the reactor temperature increases. Decreased heat transfer as a result of smaller temperature differential leads to a higher reactor temperature. Increased reaction rates contribute to more G,H formation and the noncondensable portion of the total pressure suffers a decrease. The temperature controller increases cooling water and the pressure controller cuts back slightly on the feed before it recovers to the base case value which indicates that the model based temperature controller starts taking control and cause decreased heat transfer to the reactor, thus leading to higher reactor temperatures. But the model does rather well.

The change in condenser cooling water temperature (disturbance no. 5) causes the reactor pressure to fall as a result of decreased heat transfer due to a smaller temperature differential on the condenser. The reactor level falls and the reactor temperature rises. The separator temperature goes up and thus recycle is increased and liquid feed to the stripper is decreased. The stripper temperature and the G/L ratio increase, thereby increasing the separation factor. The temperature controller increases the cooling water rate and the pressure controller leads to a slight reduction in the feed rate which indicates that the model based temperature controller starts taking control before the PI pressure controller can change the feed significantly. The reactor temperature and pressure are maintained at setpoint and the product compositions are tightly controlled.

The drop in the feed rate of A (disturbance no. 6) affects selectivity and reaction rates, which leads to a rapid pressure increase and loss of reactor temperature. Both separator temperatures and stripper temperatures decrease. Thus stripping is less and the products have unwanted reactants and more byproduct. The A/C ratio in the purge increases the purge while the reactor feed is cut simultaneously to maintain reactor pressure. The compositions show a change initially but are brought back to their setpoints smoothly because the composition controller increases the separating power of the stripper.

The unavailability of C (disturbance no. 7) similarly affects the process conditions leading to a rapid increase in pressure and a loss of temperature because of reduction in the reaction rates. The feed is cut slightly before being restored to the base case value. The temperature controller decreases cooling water to maintain reactor temperature.

Random variations in the A, B, C compositions (disturbance no. 8) cause reaction rates to vary, a slight fluctuation in the reactor level and correspondingly more variation in the C/A feed. The reactor pressure does go down initially although not significantly. A random variation in A,B,C compositions in the C/A feed is one of the stiffer upsets. The reactor temperature controller maintains reaction temperature and pressure is maintained at setpoint. The C/A feed fluctuates in order to control the reactor pressure which is affected because of the varying liquid holdups.

The results for random variation in D feed temperature (disturbance no. 9) and C feed temperature (disturbance no 10) are also shown.

There are random variations in both the reactor cooling water inlet temperature and condenser cooling water inlet temperature (disturbances nos. 11 and 12) causing uneven loads. But there is no significant deviation from setpoints of either temperature or pressure and the compositions behave rather smoothly.

The slow drift in reaction kinetics (disturbance no. 13) induces a mismatch but product compositions do not show large variation. The reactor cooling water and condenser cooling water sticky valves (disturbances nos. 14 and 15) add more to the valve inertia but do not significantly affect the controller performance. Random variations in the reactor cooling water inlet temperature and condenser cooling water inlet temperature show effects on the C/A feed and thus they affect the stripper steam in the same way.

The other disturbances are of an unknown nature which are coupled with the known disturbances but again do not significantly affect controller performance.

Thus, the model-based temperature controller coupled with a feedback PI controller seems to predict process gains acceptably and act to respond rather quickly. This is a very sensitive system and even very small perturbations in manipulated values make it unstable. Thus, the control configuration put in place for these different scenarios seems to close the loop effectively. The results are shown in Figs. 7.1 through 7.42.

**Figure 7.1 Product grade Setpoint change
(Setpoint change no 2)**

Figure 7.2 Product grade setpoint change.
(Setpoint change no 2)

Figure 7.3 Reactor pressure setpoint change
(Setpoint change no 4)

Figure 7.4 Reactor pressure setpoint change.
(Setpoint change no 4)

Figure 7.5 Step change in A/C ratio(Stream 4).
(Disturbance 1)

Figure 7.6 Step change in A/C ratio(Stream 4).
(Disturbance 1)

Figure 7.7 Step change in B composition. (Stream 4)
(Disturbance 2)

Figure 7.8 Step change in B composition(Stream 4).
Disturbance 2.

Figure 7.9 Step change in D feed temperature.
(Disturbance 3)

Figure 7.10 Step change in D feed temperature.
(Disturbance 3)

Fig 7.11 Step change in reactor cooling water inlet temperature. (Disturbance 4)

Figure 7.12 Step change in reactor cooling water inlet temperature. (Disturbance 4)

Figure 7.13 Step change in condenser cooling water inlet temperature. (Disturbance 5)

Figure 7.14 Step change in condenser cooling water inlet temperature. (Disturbance 5)

**Figure 7.15 A loss effected by a step change.
(Disturbance 6)**

**Figure 7.16 A loss effected by a step change.
(Disturbance 6)**

**Figure 7.17 Step change in C header pressure loss
(Disturbance 7)**

**Figure 7.18 Step change in C header pressure loss
(Disturbance 7)**

**Figure 7.19 Random variations in A,B,C compositions
(Disturbance 8)**

**Figure 7.20 Random variations in A,B,C compositions.
(Disturbance 8)**

**Figure 7.21 Random variation in D feed temperature.
(Disturbance 9)**

**Figure 7.22 Random variation in D feed temperature
(Disturbance 9)**

**Figure 7.23 Random variation in C feed temperature
(Disturbance 10)**

**Figure 7.24 Random variation in C feed temperature
(Disturbance 10)**

Figure 7.25 Random variation in reactor cooling water inlet temperature.(Disturbance 11)

Figure 7.26 Random variation in reactor cooling water inlet temperature. (Disturbance 11)

Figure 7.27 Random variation in condenser cooling water inlet temperature. (Disturbance 12)

Figure 7.28 Random variation in condenser cooling water inlet temperature. (Disturbance 12)

**Figure 7.29 Slow drift in reaction kinetics.
(Disturbance 13)**

**Figure 7.30 Slow drift in reaction kinetics.
(Disturbance 13)**

Figure 7.31 Disturbances 4 and 16.

Figure 7.32 Disturbances 4 and 16.

Figure 7.33 Disturbances 5 and 17.

Figure 7.34 Disturbances 5 and 17.

Figure 7.35 Disturbances 8 and 18.

Figure 7.36 Disturbances 8 and 18.

Figure 7.37 Disturbances 9 and 19.

Figure 7.38 Disturbances 9 and 19.

**Figure 7.39 Step change in A/C ratio(Stream 4).
(Disturbance 1)**

Figure 7.40 A loss effected by a step change.
(Disturbance no 6)

Figure 7.41 Random variations in A, B, C compositions.
(Disturbance no 6)

Figure 7.42 Product grade setpoint change.
(Setpoint change no 2)

CHAPTER 8

CONCLUSIONS AND RECOMMENDATIONS

The main purpose of this work has been to devise a control configuration for the Tennessee Eastman problem and to test it with the set of disturbances and setpoint changes for the base case posed in the problem. In doing so, the nonlinear model based approach is also put to the test since it is at the center of the control strategy. Approximate models have been developed for the reactor and the stripper.

The disturbance rejection was found to be very good. the setpoint changes were also effected very smoothly. The technique of online incremental parameterization was found to be very effective in removing the process model/mismatch as process conditions change. Despite this, the models are really the heart of the control.

The models have to predict process gain changes very well and in order to do that, have to describe the process fairly accurately. This work was limited to the implementation of controllers for each separate unit only.

The stripper model was used mainly as a feedforward scheme. The GMC law was not applied to the stripper since it was necessary to implement the temperature compensation on a faster scale than the slower composition feedback loop. The model based control was therefore strictly used in a feedforward sense.

8.1.Recommendations

This problem has a lot of work still left to be done. The work discussed has dealt with the base case only. There are two other operating conditions for the process, one at 10/90 G/H and the other at 90/10 G/H. The control configuration has to move the process from the base case to these two conditions. This is a huge setpoint change and involves different ratios of D/E in the feed. In addition, the disturbances and setpoint changes test the control configuration at these two operating conditions.

Operation at these two conditions involves a different holdup distribution in the reactor. Since holdups determine the partial pressures and as a result, the reaction rates, an accurate estimate of the holdups is necessary. Dynamic component and energy balances have to be studied.

The other interesting control problem would be to understand fully the effects of the reactor agitator control. Since agitation speed affects only the heat transfer coefficient, it could serve as a useful device for removing heat to keep the reactor from going unstable. Since the liquid phase seems to act as a heat sink, the agitator could prevent excessive liquid-phase temperatures and prevent the reactor level from falling as a result.

Optimization determines the economically most useful operating conditions at the base case and the two other operating states. On the other hand, operating at the optimum

states could be a more difficult control problem. Optimization involves the steady-state solution of this set of highly nonlinear equations. Convergence methods, such as Newton's method, have not been very effective with the base case values as initial guesses in the initial attempts. Optimization studies would lead to an integrated control strategy.

The problem is complex and still offers a great deal of challenge for the control researcher. So, there are many possibilities for a radically new approach to be tried.

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