7. CONTROL SYSTEM SYNTHESIS AND PLANT-WIDE CONTROL OF THE TENNESSEE-EASTMAN CHALLENGE PROBLEM

7.1. Introduction

In this chapter, we will address the problem of control system synthesis for chemical plants consisting of continuous processing steps. We restrict the control system synthesis discussion to plants for which the flowsheet and the equipment design is available. Thus, the objective of this chapter is to provide a procedure and some guidelines to: (i) convert the process flowsheet into a process and instrumentation (P & I) diagram; and (ii) implement a multi-level, plant-wide control scheme.

The chapter is organized as follows: Process control system synthesis problem is defined in section 7.2. Since the synthesis problem for an entire plant is complex, we recommend the decomposition of the flowsheet into modules. Control system synthesis tasks associated with a module are described in section 7.3. Some general guidelines, applicable to most chemical plants, are provided. In section 7.4, the guidelines are applied to develop a plant-wide control scheme for the Tennessee-Eastman (T-E) problem.

A number of researchers have presented solutions to the plant-wide control of the T-E problem. McAvoy and Ye (1994) give a decentralized, multiloop control scheme. However, they do not evaluate their control strategy by moving the process from the nominal operating mode to other operating modes. Ricker (1993) gives the optimal steady state operating conditions for different modes of operation. In a later publication, Ricker and Lee (1994a, b) demonstrate the use of non-linear model predictive control to move the plant to different operating modes.

In section 7.5, we develop two plant-wide control schemes: (i) Using multiloop, decentralized Proportional-Integral (PI) controllers; (ii) Using linear DMC controller along with PI controllers. Both the control schemes are evaluated by subjecting the plant to the recommended setpoint and disturbance changes. In section 7.6, we determine operating conditions for the maximum production of product mix 50G/50H by solving a non-linear optimization problem. The performance of the decentralized controllers and linear DMC controller while moving the process to the new operating conditions are compared.

We conclude with some remarks on our experience on developing a plant-wide control system for the T-E problem. Topics that need further investigation are identified.

7.2. Synthesis of Process Control Systems

The control system synthesis for a chemical plant consists of the following tasks (see Figure 7.1):

T1. Modularize the processing steps in the plant.

T2. Define the control objectives for the plant based on: process understanding, constraints imposed by process equipment, and the source and nature of disturbances entering the plant.

T3. Determine the manipulated variables.

T4. Determine the variables to measure, and select the controlled variables.

T5. Design the interconnecting structure, the *Control System*, between the controlled and manipulated variable so that the control system is:

- *Economical*: The plant operates safely and profitably while satisfying certain objectives and respecting all operational constraints. The objectives could be the product quality and production rate specifications.
- *Reliable*: The plant operates safely and profitably in the presence of: (a) varying market conditions, (b) changing raw material quality, (c) different product specifications, (d) measured and unmeasured disturbances, and (e) when a few degrees of freedom are lost.

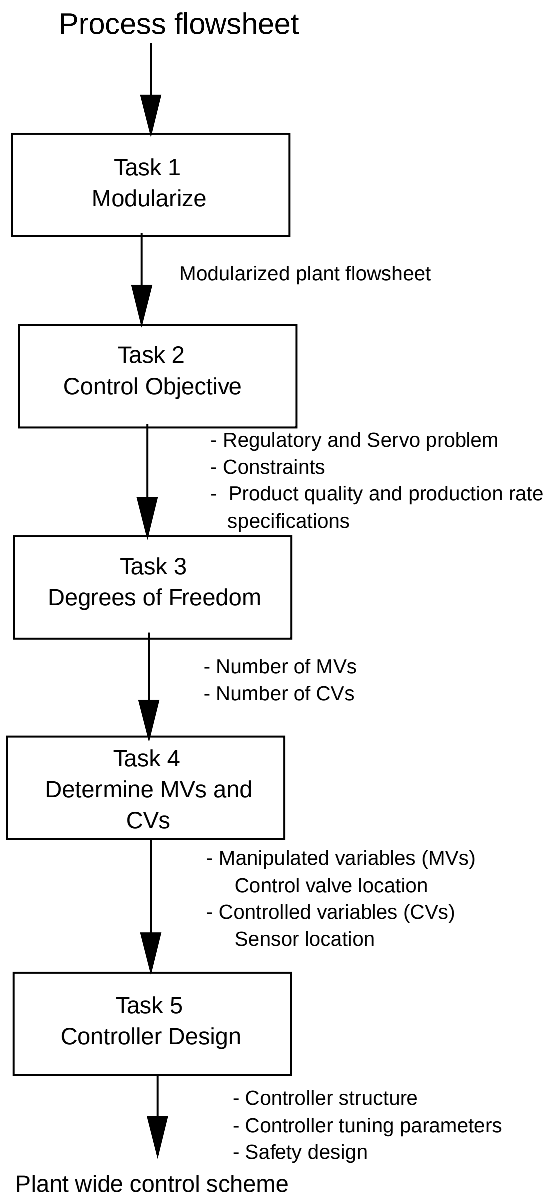

Figure 7.1 Steps in control system synthesis of a chemical plant.

- *Safe*: The plant moves to a safe state if the operating conditions become unsafe for plant equipment or personnel.

Each of the above tasks comprise of a number of subtasks which will be described in section 7.3.

A successful operation of a chemical plant requires a large number of measurements. A few of these variables, called the *controlled* or *dependent variables*, are maintained at a definite value, called the *setpoint value*, using *manipulated* or *independent variables*. Since the chemical plant consists of a large number of measurements, controlled variables, and manipulated variables, addressing the synthesis problem for the entire plant is a formidable task. To make the problem tractable, the plant is usually decomposed into smaller group(s) of processing step(s) - called *Modules*, and the control system synthesis of each module is addressed separately. The plant is decomposed so that the disturbances entering one module does not significantly affect the processing step(s) in other modules. Storage tanks between processing steps typically filter disturbances, and therefore can sometimes be used to identify modules (Buckley, 1992; Shinskey, 1988; Stephanopoulos, 1984). Since there is little interaction between modules, tasks T1-T5 can be addressed separately for each module. After which, the modules can be concatenated, and any redundancy eliminated.

The modules can be further decomposed into submodules. Submodules consist of a group of unit-operations that are tightly coupled due to a recycle stream, energy integration, or operational requirements. Some examples of submodules are: (i) Reactor-separator system with a recycle stream; (ii) Distillation column-reboiler-condensor system; and (iii) Reactor-regenerator system. Depending on the processing steps upstream and downstream of a submodule, it may be possible to address all or few of the synthesis tasks T1-T5 for each submodule separately.

7.3. General Guidelines for Control System Synthesis

T1. *Modularize the processing steps in the plant.*

Given a process flowsheet, the first task is to identify and trace the major process streams starting from the raw-materials to the products and byproducts leaving the plant. The objective is to understand the processing steps involved in the transformation of the raw-materials to the products and byproducts, and to identify modules in the plant for which control system synthesis can be addressed separately.

T2. *Define control objectives.*

A proper understanding of the processing steps helps in clearly defining the control system objectives. The control objectives can be classified into the following subgoals:

- *Regulatory control objectives.*

A list of disturbances entering the plant are identified and their effect on product quality and production rate is studied using cause-and-effect relationships. The objective of regulatory control is to reject fast disturbances and to operate the plant close to a nominal steady state.

- *Servomechanical control objectives.*

This involves being able to move the plant from one steady state to another. It also includes achieving the product quality and production rate specifications, as well as controlling the composition of the byproduct and purge streams. Controllers are implemented to satisfy the material and energy balance while considering various changes in production rate and product quality demands under which the plant might be operated.

- *Process and equipment constraints.*

Define the constraints that the control system should respect. The process constraints are typical limits on product quality deviations, production rate deviations, stream flow rates, and equipment operating range. Constraints can be either hard or soft. Hard constraints should be respected at all times, while soft constraints may be violated.

- *Optimal operation objectives*.

  Identify process variables that should be optimized to minimize the operating cost and maximize profitability while satisfying constraints. Examples of this include: equipment pressure and temperature, purge flow rate, feed rate of non-limiting reactants, flash drum temperature, and compressor work.

T3. *Select Manipulated variables.*

Manipulated variables control the flow of energy and material entering and leaving the plant using flow control valves. The speed of a rotary machine (e.g., a motor, compressor, or pump) may also be used as a manipulated variable. Following are some guidelines on placing control valves and some desired characteristics of these valves (Buckley, 1964, 1992; Stephanopoulos, 1984; Luyben, 1989; Seborg, 1989):

1. Control valves should be placed on streams through which the process interacts with the surrounding environment or a downstream/upstream process (Buckley, 1992). For instance:
    a) Control valves should be placed on the inlet feed streams (if these streams are from an upstream storage tank), and on the products and byproducts streams leaving the plant.
    b) Control valves should be placed on all utility streams entering the process. The valve may be placed on the inlet side or the exit side, but not both.
    c) Control valves should be placed on the purge and make-up streams.
2. A control valve should be used whenever there is a capacity in the system to respond to the control valve changes. For example, a liquid outlet stream from a storage tank should have a control valve. The reflux and distillate flow out of a condenser should have control valves. Exceptions to this rule are:
    a) Control valves are usually not placed on streams with large flow rate leaving a tank as moving such control valves is difficult, and

  b) Control valves are placed on a stream split even though there is no capacitance present.

3. Control valves should not be placed on streams for which control is difficult. For instance:

   a) A control valve on a gas stream with a very large flow rate should be avoided. This is because controlling this flow by moving the valve position is difficult. For example, a control valve is not place on the vapor stream to the condenser in a distillation column.

   b) Metering pumps should be used, instead of flow control valves, when the flow must be controlled very accurately.

   c) Screw pumps are used to control the solids flow.

4. While handling hazardous process streams, or in critical unit operations, additional control valves are often made available to facilitate manual control. Also, emergency control valves are placed on strategic streams for equipment and operational safety.

5. Control valves should be sized to give a reasonable variability in the manipulated variable from the nominal steady state. Oversized or undersized control valves will lead to valve saturation and loss of control.

6. The characteristics of the control valve (linear, square-root, equal percent, etc.) should be chosen to give a linear relationship between flow rate and valve position.

7. Valve positioners should be used (except in flow control loops) to overcome problems such as valve stickiness, valve hysterisis, etc. Flow control loops are generally too fast for use of valve positioners.

Each control valve in effect represents a manipulated variable. The number of manipulated variables (denoted by m) in a plant should always be greater than or equal to the number of independent controlled variables (denoted by n). If $m<n$, then there are a few controlled variables which will have a steady state offset. On the other hand, if $m>n$, the extra degrees of freedom can be utilized to operate the process economically.

T4. ***Measured and Controlled Variables.***

Process variables that are important for control or for process monitoring should be measured if possible. The measured variables could be a primary or a secondary variable. Primary variables are those that are measured directly (e.g., flow rate measurement); whereas, the secondary variables are measured to estimate some unmeasured process variables (e.g., the tray temperature in a distillation column can be used to estimate the composition). Secondary variables are chosen when measurement of certain process variables is slow, not possible, or unreliable. Measurements are often added to monitor the status of a process and only a subset of measured variables are used as controlled variables. Following are some guidelines for selecting process measurements:

1. Flow rate of all streams should be measured if possible.

2. In an unit operation with liquid hold-up, the level in the vessel should always be measured.
3. Pressure should be measured in all contained vessels. Gas pressure in a system should always be monitored. Pressure of certain liquid phase streams are sometimes important and have to be measured. For example, in a polymerization unit the pressure at the exit of the extruder is measured in order to maintain sufficient flow of the polymer to the downstream spinning and drawing operations.
4. Temperature measurements should be available whenever there is exchange of energy in an unit operation. Temperature measurements are inexpensive and are hence used extensively. These measurements are often used as secondary measurements.
5. Composition measurements are usually slow, expensive, and off-line. Such measurements should be used conservatively and when necessary. Product quality and reactant feed composition are sometimes crucial from operational standpoint. In such situations an accurate measure of composition is necessary.
6. Measurement of other physical quantities like viscosity, density, humidity, pH, conductivity, thickness, rpm, etc., may be required for control or monitoring in some processes. Such measurements may also serve as secondary indicators of quality or composition.

Guidelines for placing sensors used for process variable measurement and some desired features of these sensors are (Buckley, 1992; Luyben, 1989; Seborg *et al*, 1986; Stephanopoulous, 1984):

1. The optimal location and number of sensors are important design decisions. Sensors should be located so that a "true" measure of the process variable is obtained with little time-delay. Sensors should be located in close vicinity of unit operations that use the measurements for control/monitoring. Within a unit, sensors should be located so that external disturbances can be detected easily.
2. Redundant measurements should be used for crucial process variables and in noisy measuring environment.
3. Sensors should be placed to measure important disturbances that may affect the process (if possible).
4. The span of the sensor should be selected so that sufficiently accurate measurements are obtained. It may be required to change the zero of the span to measure a variable over a wider range. The compromise between span and range is important. For example, during startup the process variable changes over a large range and hence the span of the sensor should be large. However, after steady state is reached the process variable does not change as much. The span of the sensor at steady state can be much smaller so that more accurate and sensitive measurements are obtained.

5. The sensor should be located so that the effect of a disturbance is detected before it affects any downstream variable. For example, the inlet temperature of a cooling water or a feed stream should be monitored.
6. Calibration curves (or conversion formulas) relating the measured variables and the sensor output variables should be updated periodically. A linear relation between these variables in the operating region is desirable. Nonlinear transformations are often used to linearize the relationship between the measured variable and the sensor output. Improper calibration is a recurring problem in many operating plants.

T5 *Control System Design*

After tasks T1-T4 are resolved, the design of a plant-wide control system is considered. Decisions concerning the controller structure and controller tuning parameters are addressed. A multi-layered approach consisting of the following tasks is usually used (Price and Georgakis, 1992; Stephanopoulos, 1984; Prassinos *et al*, 1984; Morari, 1981) :

1. Design controllers to reject local disturbances.
2. Add stabilizing controllers.
3. Design decentralized multi-loop/multivariable predictive controllers.
    a) Production rate controller.
    b) Product quality controller.
4. Add feedforward and override controllers.

5. Design on-line optimization scheme.

Controllers to reject local disturbances

Local disturbances such as turbulence in the process stream, noise in measurements, changes in inlet cooling water temperature, changes in ambient temperature, changes in steam supply pressure, etc., entering a process are fast acting and local. To operate the plant near the steady state, it is desirable to reject such disturbances before they affect other process variables. Fast acting PI controllers, which are also known as *Inner Loop Cascade Controllers* (ILCC), are typically used for this purpose. Flow controllers between flow rate measurement and the flow control valve are used to reject local disturbances such as pressure fluctuations. Temperature controllers are used to reject disturbances in the inlet temperature or pressure of a utility stream. Flow controllers with a proportional bandwidth of 150% and an integral time of about 0.1 minutes are typically used (Luyben, 1989). Once the ILCCs are in place and properly tuned, the plant can be operated at the nominal steady state. The *Outer Loop Cascade Controllers* (OLCC) or *Decentralized Controllers* are used to move the plant to new steady states. The setpoint of the ILCCs are used as manipulated variables by the OLCCs. The ILCCs take more frequent control actions and they have a faster settling time as compared to the OLCCs.

The outer loop controllers are implemented in an hierarchical manner. The stabilizing controllers are implemented first. These controllers prevent accumulation of mass and energy in the system and typically consist of level and pressure controllers for mass balance and temperature controller for energy balance. The product quality and production rate control are implemented next. Ratio controllers, split-range controllers, adaptive controllers, over-ride controllers, etc. are used as needed.

Stabilizing Controllers

Chemical plants that are self-regulating (i.e., in which mass and energy cannot accumulate) do not require stabilizing controllers. Process model for a self-regulating process can be obtained by subjecting the manipulated variables to known changes. The model can then be used to pair controlled and manipulated variables and to tune the decentralized controllers.

For most chemical plants, level controllers, pressure controllers, and temperature controllers, are sufficient to stabilize the plant. Depending on the process all or a few of these controllers may be used to stabilize the process. Controlled and manipulated variables are paired based on an intuitive understanding of the process. A process model is generally not available at this stage since conducting any tests on the plants may make it unstable. Controller tuning is based on rules of thumb.

Proper pairing of controlled and manipulated variables and the tuning of level controllers is extremely important, and in fact is sometimes a crucial and a difficult decision. While pairing controlled and manipulated variables the feasibility of operating the plant at different modes should be considered. It is desirable that the same controller configuration is used at all operating conditions. Another important decision while tuning level controllers is its speed. There are two types of level controllers. One is the averaging type of level control where the tank capacity is used to dampen out process flow variations. In this case the controller is tuned very loosely to minimize the variations in the inlet or outlet flow. The second type of level control is used when strict level control is required. For example, the reactor level can influence other process variables and a strict level control may be necessary. Tuning the controller to respond fast can be detrimental for the control of downstream processes. This is because such control actions will act as high frequency disturbances entering the downstream process. On the other hand, slow level controllers can lead to poor product quality and production rate control. A careful design of these controllers is therefore necessary.

Total material balance of a plant/module consists of mass balance and individual component balances. Material balance is said to be satisfied when there is no accumulation of mass, i.e., the rate at which material enters the plant/module is equal to the rate at which it leaves. For a process involving only liquid phase reactants and products, controlling the level in all vessels with a liquid holding capacity will satisfy the material balance. However, if a process involves both gas phase and liquid phase reactants/products, material balance is satisfied by controlling the level in all vessels and by controlling independent pressure measurements. Vapor inventory (measured in terms of pressure) is controlled by adjusting the heat input or removal. For instance, in a distillation column, pressure is controlled by adjusting the condenser duty. In a boiler, the pressure is controlled by adjusting the heat input to the boiler. Pressure controllers are usually tuned to be fast responding.

For safe and successful operation of a plant there should be no accumulation of energy in any unit operation. This objective can be achieved by controlling the independent temperature measurements in unit operations involving exchange of energy. If the temperature in all unit operations are held constant, then the energy in streams entering the plant and that leaving the plant is equal.

After the stabilizing controllers are in place, the need for using advanced control algorithms like ratio control, over-ride control, split-range control, constrained multivariable predictive control, non-linear control, adaptive control, etc., can be investigated. Steady state and dynamic Relative Gain Array, Niederlinski index, and Singular Value Decomposition are some tools that can be used for such investigations. A reasonably accurate process model is required for these analyses.

Decentralized Controllers

Multiloop decoupled Proportional-Integral-Derivative (PID) control constitute the decentralized controllers. Such controllers can be used when there is little or no interaction between variables or when it is possible to design a decoupler that yields a system with little interaction. The ILCC setpoints are used as the manipulated variables. If ratio controllers are necessary, then the setpoint of the ratio controllers are used as the manipulated variables. Constraints on controlled variables and the knowledge of the process model are not explicitly used in these controllers. Constraints on manipulated variables are usually implemented using select switches.

Production Rate and Product Quality Control

Production rate can be controlled directly or indirectly by changing the amount of raw material entering the plant. For a plant with multiple raw material streams, there are two typical configurations for production rate control (Shinskey, 1988):

1. Change the setpoint for the primary raw material stream. The primary stream has a large flow rate and does not change substantially during product quality changes. The primary stream is ratioed to other raw material streams (see Figure 7.2). In the configuration shown in Figure 7.2, the primary raw material is stream C. Changing the production rate results in change in stream C flow rate. Stream C is ratioed with streams A and B to maintain the production rate and product quality.

2. Ratio the production rate to all the raw material streams (see Figure 7.3). Here all streams vary substantially during product quality and production rate changes. As a result, streams A, B and C are ratioed directly with the production rate.

Product quality can be controlled by: (i) adjusting the feed composition or feed ratios, and (ii) adjusting operating conditions of units such as reactor, separator, etc. The ratios of the stream flow rates are the manipulated variables used to achieve different product quality (see Figures 7.2 and 7.3). Operating conditions like temperature, pressure, etc. in unit operations can also be changed independently to achieve desired product purity.

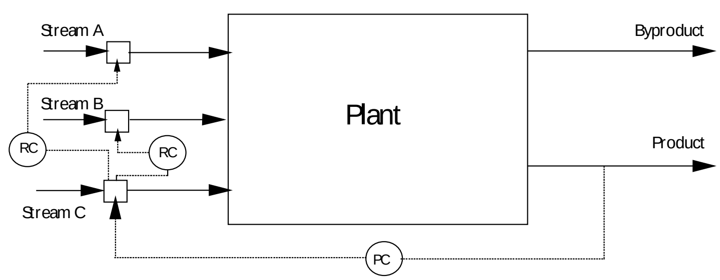

Figure 7.2 Production rate control approach I for multiple stream plant.

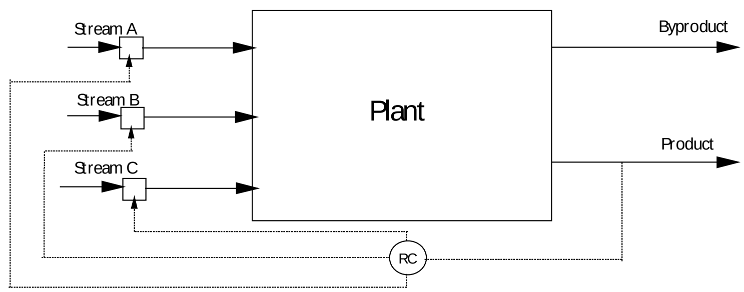

Figure 7.3 Production rate control approach II for multiple stream plant.

Multivariable Predictive Controllers

Processes with operating constraints and strong interacting output variables can be controlled effectively using multivariable predictive controllers. Dynamic Matrix Control (DMC) is one of the most widely used multivariable model predictive control algorithm in the chemical process industry. The DMC controllers use a linear model of the process to predict the future controlled variable values. The deviation of the predicted values from a desired setpoint trajectory is minimized using a quadratic objective function to obtain future manipulated variable changes. The optimization is subject to manipulated and controlled variable constraints. The first manipulated variable changes are implemented. The predicted controlled variable values are updated and the optimization is repeated.

The multivariable predictive control algorithm accepts optimal target values for controlled and manipulated variables from a linear programming (LP) problem that minimizes the operating cost (Harkins, 1991; Brosilow and Zhao, 1988; Cutler, 1983). It has been found in practice that operating the equipment close to the constraints results in maximum economic benefits (Cutler and Perry, 1982). The LP problem minimizes an economic objective function which reflect the cost of moving various manipulated variables. The constraints in the optimization are the upper and the lower limits on the manipulated variables and the controlled variables. The following are some advantages of using the LP optimization along with the predictive controller:

1. The optimal target values for the manipulated variables obtained using on-line steady state optimization (discussed later in this chapter) may not be the true optimum. The LP problem updates the targets and drives them closer to the true optimum.
2. In the event of the loss in degrees of freedom in control, which could be due to valve saturation or sticky valves, the number of manipulated variables can be less than the controlled variables. For this situation, the LP problem determines the least expensive controlled variables that can have a steady state offset.
3. If the number of manipulated variables are more than the number of controlled variables, the LP determines the optimum target values making use of the extra degrees of freedom.

The weight in the LP objective function reflect the cost of moving the controlled/manipulated variables. The bounds on the controlled and the manipulated variables are updated by the operator depending on the process operation.

Feedforward Controllers

The disturbances affect the controlled variables before the feedback controllers react to it. Feedforward controllers, on the other hand, measure the disturbance and take corrective action to nullify the influence of the disturbance on the controlled variable. To design a feedforward control, a direct measure of the disturbance should be available, and a model relating the disturbances and the controlled variables should be known.

On-Line Optimization

On-line or real time optimization is used to identify the most profitable plant operating conditions while respecting equipment and process constraints. The operating conditions are then passed to the predictive control/decentralized control layer for implementation. The optimization is based on a rigorous steady state model of the plant. An open-equation or a closed-equation based simulation is used in the optimization. The advantage of open-equation based approach over the closed-equation based approach is that the same model can be used for data reconciliation, parameter estimation, and optimization. However, there are some difficulties in obtaining initial guesses for the variables and computing the gradients in the open-equation based approach. The steps involved in on-line optimization are:

1. Steady state detection.
2. Data validation and reconciliation.
3. Parameter estimation.
4. Optimization.
5. Setpoint implementation.

A statistical test is used to determine if the plant is at steady state. The composite test for steady state detection has been developed by Narasimhan *et al* (1986, 1987). If some key plant measurements satisfy the statistical criteria, then the plant is considered to be at steady state. After verifying steady state operation, the plant data is collected for parameter estimation.

Before using the plant data for parameter estimation, the plant measurements are validated using steady state data reconciliation. In data reconciliation, the measurements are first screened to eliminate any outliers. The measurements are then adjusted using a constrained weighted least squares optimization to close the material and energy balance. Usually the number of measurements are such that the least squares problem has sufficient redundancy. The weights used in the least squares problem are assumed to be inversely proportional to the variance of that measurement. Larger the variance of a measurement, more is the margin for adjustment. After the data is reconciled, the data is used to estimate unmeasured parameters such as heat transfer coefficients, fouling factors, pressure drops, tray efficiency, etc.

The plant model updated with the estimated parameters is then used to solve the optimization problem. A Sequential Quadratic Programming (SQP) optimization is commonly used to obtain optimal operating conditions that maximizes plant profitability.

Some key plant measurements are again collected to see if the plant is still at steady state. If the plant is at the same steady state when the data was collected, then the optimal operating conditions are passed to the control system for implementation. The optimization is run periodically to improve the plant economics. Cutler and Perry (1982) state that: "*Improvements in the range of 5% to 10% of the value added by the process have been obtained for closed loop real-time optimization and constrained multivariable control systems applied to a number of oil and chemical units*". The payback time on the optimization and control system is sometimes as short as a year.

7.4. The Tennessee-Eastman Problem

For a thorough evaluation of the synthesis technique described in sections 7.2 and 7.3, one must apply the technique to real chemical plants. However, it is difficult in an university to conduct full scale experiments for this purpose. For this reason academicians have constantly expressed a need for the simulation of an industrial problem of sufficient complexity that could be used as a test bed.

William-Otto plant (William and Otto, 1960) is one such simulation that was made available to academicians. The William-Otto plant consists of a cooled stirred tank reactor, a cooler, a decanter to separate valuable products, and a product separator. A recycle stream from the separator bottom to the reactor contains unreacted reagents. The William-Otto plant has been used in a number of optimization and control studies. Morari *et al* (1980) have applied their synthesis procedure on this plant.

The Shell Process Control Problem (Prett and Morari, 1987) is another control problem that was put forth by industry. The Shell problem addresses the control of a heavy oil fractionator with significant interaction and hard constraints. The transfer function models and the control objectives are provided. The problem is suited only to apply process control techniques around a single unit operation. Model Predictive Control was a favorite choice among the solutions presented (Prett *et al*, 1990).

More recently, Downs and Vogel (1993) have posed an Industrial Challenge Problem in Process Control, referred to hereafter as the Tennessee- Eastman (T-E) problem. This problem simulates an industrial chemical process consisting of a non-linear, multi-component, 2-phase reactor, a condenser, a vapor-liquid separator, and a stripper (see Figure 7.4). The T-E problem is quite different from the previous industrial challenge problems. Its purpose is to stimulate study, development, and evaluation of the required process control technology. The process is well suited for a wide variety of studies, including plant-wide control, multivariable control, model predictive control, non-linear control, optimization, estimation, adaptive control, and process diagnosis. Downs and Vogel have also defined the control objective and constraints for this problem. However, the process model equations have not been made available. In this section, we describe the unique features of this problem. The control problem is also defined.

7.4.1. Process Description

Figure 7.4 shows the flowsheet of the T-E process. The process has five major units: a reactor, a product condenser, a vapor/liquid separator, a recycle compressor, and a product stripper. The process produces two products (G,H) from four reactants

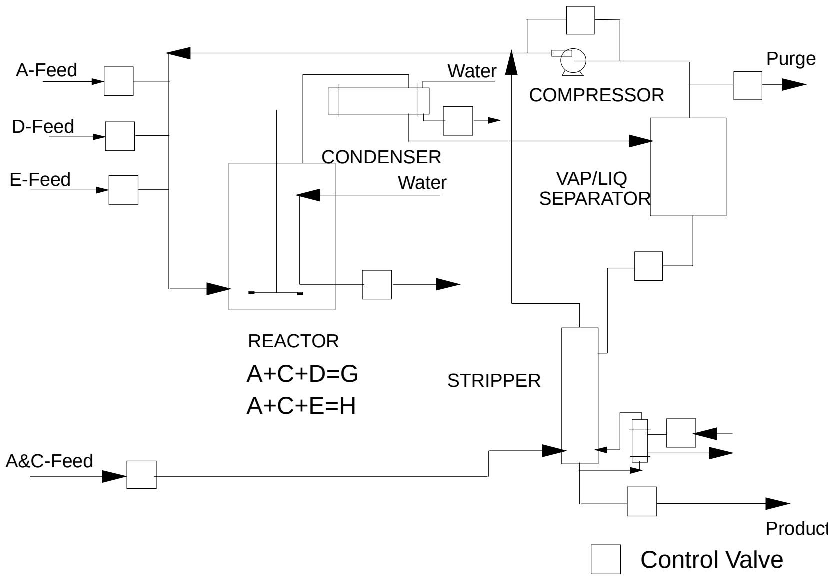

Figure 7.4 Eastman Tennessee Problem Process Flowsheet (Downs and Vogel, 1993).

(A,C,D,E). Also, there is an inert (B) and a byproduct (F). The reactions occurring in the reactor are :

$$A_{(g)} \quad C_{(g)} \quad D_{(g)} \qquad G_{(l)} \qquad \text{(R1)}$$
$$A_{(g)} \quad C_{(g)} \quad E_{(g)} \qquad H_{(l)} \qquad \text{(R2)}$$
$$A_{(g)} \quad E_{(g)} \qquad F_{(l)} \qquad \text{(R3)}$$
$$3\,D_{(g)} \qquad 2\,F_{(l)} \qquad \text{(R4)}$$

All reactions are irreversible and exothermic. The reaction rates are a function of temperature through the Arrhenius expression. The reaction to produce G has a higher activation energy resulting in greater sensitivity to temperature. The reactions are approximately first order with respect to reactant concentration.

The gaseous reactants are fed to the reactor where they form volatile liquid products. The reactor has cooling bundles to remove the heat produced during the reaction. The reactor product stream consists of unreacted non-condensable reactants and condensable products. The reactor product stream passes through a cooler that condenses the products. This stream then passes through a vapor/liquid separator. Non-condensable components are recycled through a compressor to the reactor feed, while the condensable components move to a product stripper. To avoid accumulation of inert B, part of the recycle stream is purged. Products G and H exit the system from the product stripper.

7.4.2. Control Problem Definition

Given 41 measurements (see Table 7.1), 12 manipulated variables (see Table 7.2), and a number of control objectives and constraints, the aim is to develop and test a plant-wide control scheme.

The control objectives are :

1. Maintain process variables at desired value.
2. Keep the process operating condition within safe limits.
3. Minimize variability of product rate and product quality during disturbance.
4. Minimize movement of valves, which affect other processes.

5. Recover quickly and smoothly from disturbances, production rate changes, or product mix changes.

The constraints are:

1. The variability of the product stream should be less than or equal to 5% at a frequency of 8-15 hr^{-1} .
2. Composition variation should be less than 5 mole % G at a frequency of 6-10 hr^{-1}.
3. Minimize flow variability of stream 4 in the frequency range 12-80 hr^{-1}.
4. Minimize flow variation of streams 1 and 2 in the frequency range 8-16 hr^{-1}.
5. The dynamic performance tests to be made are :
    a) Setpoint changes :
        i) -15% to variable used to set process production rate.
        ii) make step change to the variable used to ensure correct product composition.
        iii) make step change so that the reactor pressure changes from 2705 to 2645 kPa.
        iv) make a step change of +2% to the composition of B in purge.
    b) Disturbances :

        A list of disturbance changes for which the controllers implemented should be tested are given in Downs and Vogel (1993).

Table 7.1 List of Measured Variables.

Variable Name	Variable number
A Feed	xmeas(1)
D Feed	xmeas(2)
E Feed	xmeas(3)
A and C Feed	xmeas(4)
Recycle Flow	xmeas(5)
Reactor Feed Rate	xmeas(6)
Reactor Pressure	xmeas(7)
Reactor Level	xmeas(8)
Reactor Temperature	xmeas(9)
Purge Rate	xmeas(10)
Product Separator Temperature	xmeas(11)
Product Separator Level	xmeas(12)
Product Separator Pressure	xmeas(13)
Product Separator Underflow	xmeas(14)
Stripper Level	xmeas(15)
Stripper Pressure	xmeas(16)
Stripper Underflow	xmeas(17)
Stripper Temperature	xmeas(18)
Stripper Steam Flow	xmeas(19)
Compressor Work	xmeas(20)
Reactor Cooling Water Outlet Temp	xmeas(21)
Separator Cooling Water Outlet Temp	xmeas(22)
Reactor Feed Analysis A	xmeas(23)
B	xmeas(24)
C	xmeas(25)
D	xmeas(26)
E	xmeas(27)
F	xmeas(28)
Purge Gas Analysis A	xmeas(29)
B	xmeas(30)
C	xmeas(31)
D	xmeas(32)
E	xmeas(33)
F	xmeas(34)
G	xmeas(35)
H	xmeas(36)
Product Analysis D	xmeas(37)
E	xmeas(38)
F	xmeas(39)
G	xmeas(40)
H	xmeas(41)

Table 7.2 List of Manipulated Variables

Manipulated Variable	Variable number
D Feed Flow	xmv(1)
E Feed Flow	xmv(2)
A Feed Flow	xmv(3)
A and C Feed Flow	xmv(4)
Compressor Recycle Valve	xmv(5)
Purge Valve	xmv(6)
Separator Pot Liquid Flow	xmv(7)
Stripper Liquid Product Flow	xmv(8)
Stripper Steam Valve	xmv(9)
Reactor Cooling Water Flow	xmv(10)
Condenser Cooling Water Flow	xmv(11)
Agitator Speed	xmv(12)

7.4.3. Distinct Features of the T-E Problem

Following are some features that make the T-E problem challenging:

1. The gaseous reactants and gas phase reactions result in fast dynamics, making control of reactor pressure, production rate, and product quality difficult.
2. The products leave the reactor as vapor, and the outlet stream from the reactor has no control valve. Intuitively, we would expect a stream to draw the liquid products from the reactor; however, this stream does not exist in this process since the catalyst is dissolved in the liquid phase. Also, the composition measurements at the reactor outlet are not available.
3. The pressure difference between the reactor and the separator is the driving force for the flow between these units. Further, the pressure in the separator is dictated by the pressure in the reactor.
4. The reactor pressure is very sensitive to changes in reactant inlet flow rates. Accumulation of any gaseous reactant results in the reactor pressure to reach 3000 kPa, the shut-down limit.
5. The product quality depends on the reactant concentration in the reactor. However, at the given steady state, a change in concentration of one reactant cannot be achieved by increasing/decreasing the inlet flow rates. Such a change would result in accumulation of that reactant or some other reactant, which would result in reactor pressure reaching the shut-down limit.
6. The recycle stream recirculates the non-condensable components, mainly reactants, back to the reactor feed. This recirculation results in process variables, especially the

reactor pressure, reactor level, reactor temperature and the reactor feed compositions, to be tightly coupled. The recycle ratio is high (recycle rate/production rate is 5.69 (kgmol/hr)/(kgmol/hr) at the base case operation), and the yield of the product in the reactor is low.

7. The process does not remain at steady state when the simulation is run at the given base case values. Process noise causes accumulation of reactants, which results in increase in pressure, and within 2-3 hours the pressure reaches the shut-down limit.
8. Step/Impulse tests cannot be conducted without having controllers for variables that are not self-regulating. If such a test is conducted without these controllers, the pressure reaches the shut-down limit.
9. The reactor dynamics are non-linear and fast. Therefore it is difficult to control the variables associated with the reactor.
10. The purge flow rate, compressor work, reactor level, reactor temperature, and the reactor pressure are important optimizing variables. Cost of each of the manipulated variables is available and setpoints for minimum operating cost can be obtained.

The T-E problem is a realistic problem and has many features which make it difficult to control. Therefore, this problem can be used as a test-bed for the study and evaluation of the process control theory.

7.5. Synthesis of Control System

7.5.1. Manipulated and Measured Variables Selection

The T-E process flowsheet cannot be modularized into sub-modules as the gaseous reactants involved in the process result in a tightly coupled process. Therefore, the control system synthesis of the entire plant is addressed. The control objective, and the manipulated and measured variables for the T-E process are provided by Downs and Vogel (1993) (see Tables 7.1 and 7.2). Here we verify the selection of manipulated and measured variables using the rules presented in section 7.3.

Manipulated Variables Selection

1. *Rule: Control valves should be place on all streams through which the process interacts with the surrounding environment.* This justifies placing control valves on the following streams: A-Feed, D-Feed, E-Feed, A&C-Feed, Reactor Cooling Water, Condenser Cooling Water, Purge, and Stripper Underflow.
2. *Rule: A control valve should be used whenever there is a capacity in the system to respond to the control valve changes.* This suggests that a control valve should be placed on the separator underflow. A control valve is already placed on the stripper underflow.

3. *Rule: Control valves should not be place on streams for which control is difficult*. Since the stream leaving the reactor is a vapor stream with a large flow there is no control valve.
4. *Rule: Control valves should be placed for equipment safety*. The compressor recycle valve is placed to protect the recycle gas compressor.

Measured Variables Selection

1. *Rule: If possible, flow rate for all streams should be measured*. The vapor steam leaving the reactor involves a two-phase mixture with a high volumetric flow rate. Reliable measurements are difficult to obtain.
2. *Rule: Place level measurements in equipment with liquid hold-up*. Therefore, level should be measured in the reactor, separator, and stripper.
3. *Rule: Place pressure measurements if gaseous phase is present in an unit operation*. Reactor, separator and stripper pressures should therefore be measured.
4. *Rule: Place temperature measurements whenever there is exchange of energy in an unit operation*. The reactor, the separator, and the stripper temperature should be measured along with the temperatures of the cooling water outlet streams.
5. Composition of the product stream, purge stream and the reactor feed stream are crucial.
6. The compressor work is one of the key operating variable. It should therefore be measurement.

7.5.2. Inner Loop Cascade Controllers

The inner loop cascade controllers are fast acting and they control variables that are in close vicinity of the manipulated variables. The control actions influence the controlled variables with little time lag.

The inner loop controllers installed in the T-E plant are shown in Figure 7.5 and the tuning parameters are summarized in Table 7.3. Proportional gain for these controllers are computed from steady state gains (which is computed using the nominal controlled variable value and the nominal valve position) assuming a proportional band of 100%. Integral time for flow controllers was assumed to be 0.1 min. (Luyben, 1989). The tuning parameters for the temperature controllers are obtained using a trial-and-error procedure. We start with a large controller gain and integral time and reduce it till satisfactory performance is obtained. The controller sampling time is 1 second.

7.5.3. Stabilizing Controllers

The T-E process has a few variables that are not self-regulatory, this makes it difficult to develop a process model by subjecting the manipulated variables to known changes. It was therefore inevitable to use some amount of intuition and some ingenuity to get appropriate pairing between the manipulated and the controlled variables which stabilized the process. This section explains the rationale behind the pairing used in the implementation of the stabilizing PI controllers.

Table 7.3 Tuning parameters for Inner Loop Cascade Controllers.

Controlled Variable	Manipulated Variable	Prop. Gain	Int. Time (min.)
A-Feed xmeas(1)	A-Feed Flow xmv(3)	100 (%/kscmh)	0.1
D-Feed xmeas(2)	D-Feed Flow xmv(1)	0.0172 (%/kg/hr)	0.1
E-Feed xmeas(3)	E-Feed Flow xmv(2)	0.008 (%/kg/hr)	0.1
A & C-Feed xmeas(4)	A & C Feed Flow xmv(4)	6.5573 (%/kscmh)	0.1
Purge Rate xmeas(10)	Purge Valve xmv(5)	118 (%/kscmh)	0.1
Prod Sep Underflow xmeas(14)	Separator Pot Liq. Flow xmv(6)	1.2 ($\%/m^3/hr$)	0.1
Stripper Underflow xmeas(17)	Stripper Liq. Prod. Flow xmv(7)	1.2 ($\%/m^3/hr$)	0.5
Strip Steam Flow xmeas(19)	Stripper Steam Valve xmv(9)	0.206 (%/kg/hr)	0.1
Reac. Cooling Water Outlet Temperature xmeas(21)	Reac. Cooling Water Flow xmv(10)	-5 ($\%/^oC$)	0.3
Sep. Cooling Water Outlet Temperature xmeas(22)	Cond. Cooling Water Flow xmv(11)	-3 ($\%/^oC$)	0.3

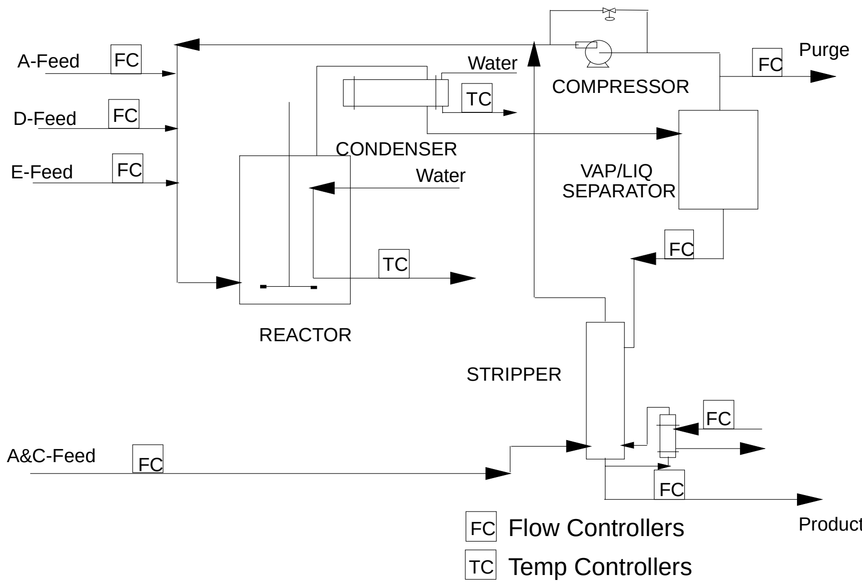

Figure 7.5 Inner loop cascade controllers.

7.5.3.1. Level Controllers

Successful implementation of material balance controllers for the T-E problem is a rather complex task as implementing Inner Loop Controllers does not stabilize the T-E plant, and therefore no dynamic tests can be performed. Bearing the objective of implementing material balance controllers (i.e., assuring no accumulation of reactants/products in the plant), and making some simple observations help in determining suitable candidates for material balance control that can stabilize the plant. The following arguments are used to arrive at the material balance control:

1. At steady state, if material balance is satisfied, the reactor level, product separator level, and product stripper level should not change. Furthermore, there should be no reactant or product accumulation in the gaseous phase.
2. Product separator level controller can be cascaded to product separator underflow controller, and the product stripper level can be cascaded to stripper underflow controller. The pairing is justified as these manipulated variables have direct and fast effect on the controlled variables. Reactor level can be controlled by cascading it to one of the following inner loop controllers: (i) E-feed flow, (ii) D-feed flow, (iii) A & C-feed flow, (iv) A-feed flow, (v) Reactor cooling water outlet temperature (RCWOT), (vi) Condenser cooling water outlet temperature (CCWOT). The first four configurations control the material entering the reactor, while the last two control the material that leaves the reactor. A few step tests were performed to decide the manipulated variable for reactor level control.
3. Reactions (R1) and (R2) suggest that moles of reactants A and C determine the production rate. Most of the reactants A and C enter through stream 4 (xmeas(4)=9.3477 kscmh and xmeas(1)=0.25052 kscmh). Mole fraction of A and C in stream 4 are 0.485 and 0.51 respectively - suggesting that stream 1 is a make-up stream for inadequate reactant A entering in stream 4. Hence, the production rate should be controlled by stream 4. Cascading production rate control to reactor level control is not justified as both the controlled variables have comparable response time. Further, if the production rate control is cascaded to reactor level control, then

at maximum production the reactor level will reach the maximum value. The gas volume in the reactor will be very small under this condition, and the reaction rate will therefore be small. The maximum achievable production is not favored by this configuration for reactor level and production rate control.

4. If a positive step change in A and C feed flow rate (steam 4) is introduced, then the reactor pressure increases and reaches the shut-down limit in about 30 minutes. This occurs because the excess of gaseous reactants A and C accumulate over time and increase the reactor pressure. The accumulation results in further increase in reactor pressure as the reaction rates decrease when the concentrations of reactants D and E decrease. Further, the purge rate is a small fraction of the recycle flow and cannot purge out the excess reactants. For a negative step change in A and C feed rate, less amount of reactants are converted to products, and therefore less amount of heat is generated. Consequently, the reactor temperature starts decreasing, which in turn results in further decrease in reaction rate. Eventually gaseous reactants accumulate and cause the pressure to reach the shut-down limit. Similar dynamic responses are observed for positive and negative step changes in D and E feed flow rates. These observations suggest the following :

    (a) To satisfy material balance excess reactants being recycled should be consumed to produce liquid products. Excess reactants are not purged out as the purge rate is small.

(b) Changing reactor cooling water outlet temperature or condenser cooling water outlet temperature does not result in consumption of excess reactants. Cascading the reactor level control to the reactor temperature control would result in unwarranted change in product quality.

(c) Excess reactants A and C can be consumed by changing D or E feed flow rates. Excess D (or E) can be consumed by changing E (or D) flow rate. Furthermore, reactor level determines the amount of product formed.

5. Feed flow rates of D or E seem to be a reasonable choice for reactor level control, the other will be used for product quality control. Streams 1 and 2 cannot be varied in the frequency range 8 to 16 hr^{-1} (see section 7.4.2). Since reactor level control is a fast responding controller, we use E-feed flow for reactor level control. We label this selection as *Configuration I* (see Figure 7.6a).

6. Condenser cooling water outlet temperature can be used for reactor level control. If the outlet temperature is decreased (less heat is removed), the recycle flow increases (separator underflow decreases) and the reactor level increases as less product leaves the recycle loop. On the other hand, if the outlet temperature is increased (more heat is removed), the recycle flow decreases and there is more product leaving the recycle loop resulting in the reactor level to decrease. We label this selection as *Configuration II* (see Figure 7.6b).

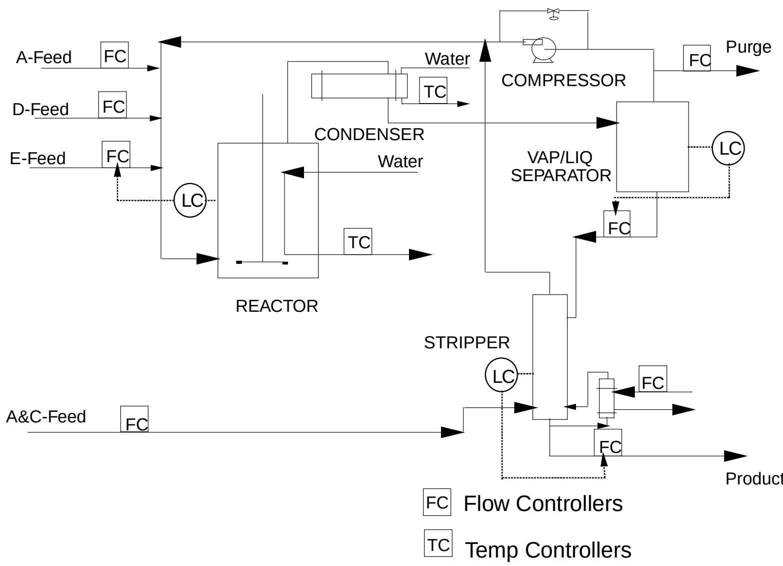

Figure 7.6a Level controllers -Configuration I.

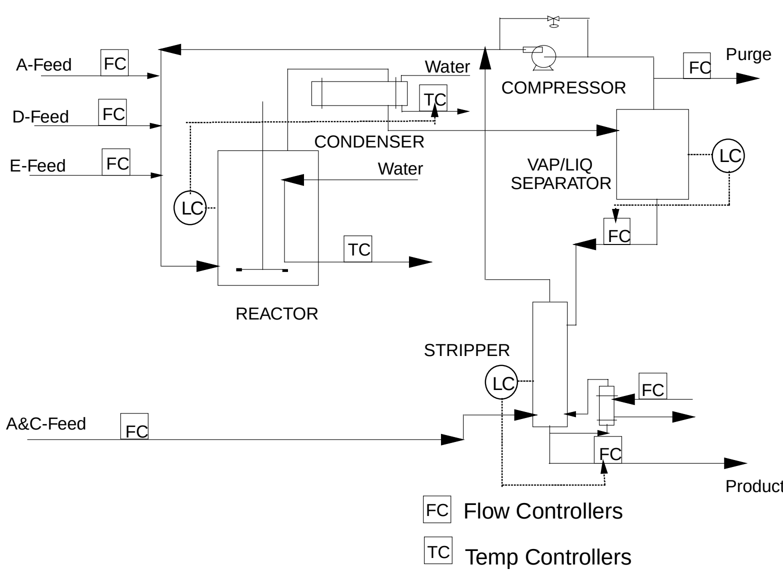

Figure 7.6b Level controllers -Configuration II.

7.5.3.1.1. Tuning Material Balance Controllers

Configuration I

The tuning of material balance controllers is done as follows:

a *Product Separator Level Control* : The inner loop separator pot liquid flow controller setpoint was changed by ±10%. The product separator liquid level exhibits an integrating response - (-18.04 (%/m^3/hr))/*s* (where *s* is the Laplace operator). A PI controller with gain 2/(k_p $_c$) and integral time 2 $_c$ (Rivera *et al*, 1986); where, k_p is the process gain and $_c$ is the close-loop time constant, is used. Using $_c$=0.4 hr, we obtain k_c=-0.301 and τ_I=0.8 hr.

b *Stripper Level Control* : The inner loop stripper liquid product flow controller setpoint was changed by ±10%. The product separator liquid level exhibits an integrating response -- (-21.331 (%/m^3/hr))/*s*. Using $_c$=0.5, we obtain k_c=-0.1789 and τ_I=1 hr.

c *Reactor Level Control* : The E-feed flow setpoint was changed by ±10. The reactor level response can be modeled as: 0.03992(%/kg/hr)/*s*. The controller tuning parameters obtained using $_c$=0.15 are k_c=300 kg/hr/% and τ_I=0.32 hr.

A sample time of 30 seconds is used for the level controllers.

Configuration II

Tuning of the separator level and stripper level controllers is the same as in Configuration I. The reactor level is controlled using the condenser cooling water outlet temperature (CCWOT). A 10% change in CCWOT controller was introduced. The reactor level exhibits an integrating response that can be modeled as: 2.8634(%/^{o}C)/*s*. Using IMC rules and $_{c}$=0.08 hr, the controller tuning parameters are: k_c=0.1916 kg/hr/% and $_{I}$ 100 min. A sample time of 30 seconds is used.

Configuration I is stable when the inlet stream feed rate is changed around the nominal steady state value. The excess reactants are consumed by changing the E-feed rate. Configuration II, however, is not stable when the inlet feed rates are changed. The reactor pressure reaches a shutdown limit of 3000 kPa due to accumulation of unreacted reagents. Hence, only Configuration I is considered in the rest of this chapter.

7.5.4. Decentralized Control

The decentralized or outer loop PI controllers were synthesized after the plant is stabilized using Configuration I level controllers. The pairing of the manipulated variables with the controlled variables is selected using steady state Relative Gain Array (RGA). Dynamic models, assuming the plant to be linear time-invariant (LTI) around the nominal steady state, are developed by considering the plant to be a black-box (see Figure 7.7). Step changes of magnitude 2-5% of the steady state value were introduced in the setpoints of the inner loop controllers. Changes in both the positive and the negative direction were introduced. The input-output data collected was used to identify linear time-invariant, continuous dynamic models using graphical techniques. The models are summarized in Table 7.4. Table 7.5 lists the process gain obtained for step changes in the positive and negative directions. Note from Table 7.5 that the response of the reactor pressure, the B-composition in purge, and the stripper temperature are non-linear.

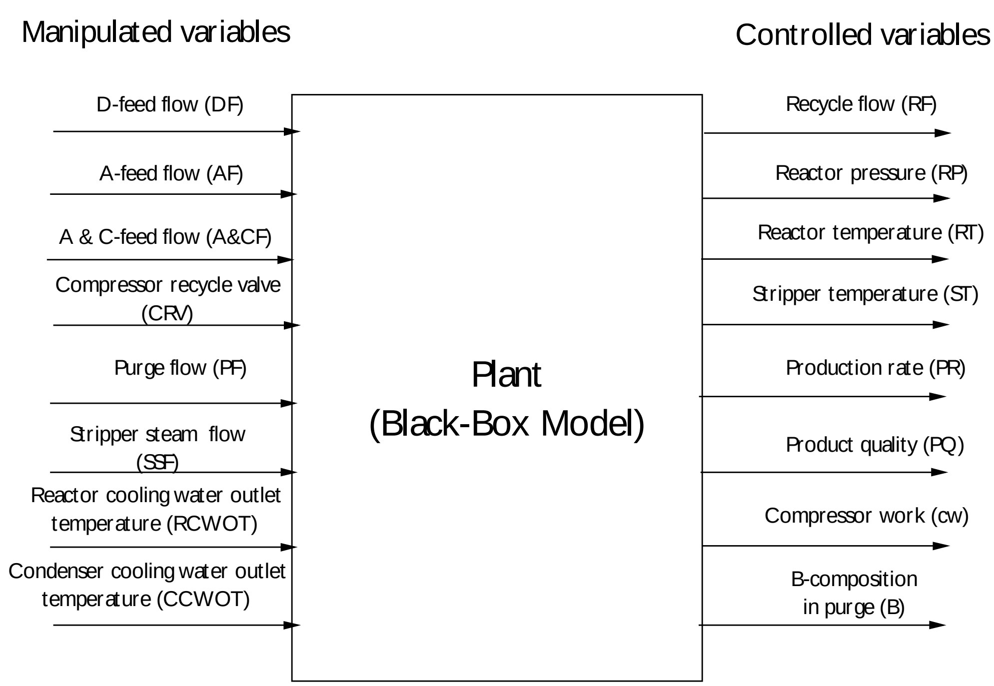

Figure 7.7 Black-box process identification approach.

Table 7.4 Nominal Model for the T-E problem

	Recycle Flow (RF)	Reac. Pressure (RP)	Reactor Temperature (RT)	Stripper Temperature (ST)	Production Rate (PR)
D Feed Flow (DF)	$\frac{-0.00273}{0.25s+1}$	$\frac{-0.27}{12s+1}$	$\frac{0.00218}{0.00089s^2 \quad 0.01705s \quad 1}$	$\frac{0.0095524}{0.3478s \quad 1}$	0
A Feed Flow (AF)	$\frac{-7.8237}{0.15s+1}$	$\frac{3083.6}{16.615s \quad 1}$	$\frac{6.386}{14s \quad 1}$	$\frac{67.8585}{14.995s \quad 1}$	$\frac{4.068}{0.1667s+1}$
A & C Feed Flow (A&CF)	$\frac{1.7384}{0.12s \quad 1}$	$\frac{287.895}{5.75s \quad 1}$	$\frac{1.63135}{0.0743s^2 \quad 0.13145s \quad 1}$	$\frac{6.225}{4s \quad 1}$	$\frac{2.67443}{0.875s \quad 1}$
Compressor Recycle Valve (CRV)	0	$\frac{7.654}{0.098s^2+0.217s+1}$	0	0	0
Purge Flow (PF)	$\frac{-16.255}{4.55s+1}$	$\frac{-949.22}{11.75s+1}$	$\frac{0.5928}{18s \quad 1}$	$\frac{20.468}{13s \quad 1}$	$\frac{-2.906}{0.2s+1}$
Stripper Steam Flow (SSF)	0	$\frac{0.0868}{5s+1}$	0	$\frac{0.03689}{0.174s+1}$	0
Reactor cooling water outlet temperature (RCWOT)	$\frac{0.6857}{0.2836s \quad 1}$	$\frac{-35.23e^{-0.9s}}{8s+1}$	$\frac{1.11}{0.01747s^2+0.0237s+1}$	$\frac{1.938}{0.0676s^2 \quad 0.26s \quad 1}$	$\frac{0.01726}{0.75s+1}$
Condenser cooling water Outlet temperature (CCWOT)	$\frac{0.2587}{0.4782s \quad 1}$	$\frac{13.7}{8.5s+1}$	$\frac{0.02156}{0.0656s^2 \quad 0.1505s \quad 1}$	$\frac{0.44165}{0.0945s^2 \quad 0.03095s \quad 1}$	0

Table 7.4, continued.

	Product Quality (PQ) G/H ratio	Compressor Work (CW)	B Composition in Purge (B)
D Feed Flow (DF)	$\frac{7.762\times10^{-4}e^{-0.25s}}{2s+1}$	$\frac{-0.016375}{0.1952s+1}$	0
A Feed Flow (AF)	0	$\frac{125.734e^{-2.96s}}{10.275s+1}$	0
A & C Feed Flow (A&CF)	$\frac{-0.2885e^{-0.25s}}{2s+1}$	$\frac{13.3722}{0.2s \quad 1}$	$\frac{0.0287e^{-0.1s}}{25s+1}$
Compressor Recycle Valve (CRV)	0	$\frac{2.9268}{0.02215s^2+0.1141s+1}$	0
Purge Flow (PF)	0	0	$\frac{-31.79e^{-0.1s}}{13.9s+1}$
Stripper Steam Flow (SSF)	0	0	0
Reactor cooling water outlet temperature (RCWOT)	$\frac{4.01\times10^{-4}e^{-0.25s}}{2s+1}$	$\frac{4.4926}{0.096s^2 \quad 0.20415s \quad 1}$	0
Condenser cooling water Outlet temperature (CCWOT)	0	$\frac{2.47969}{0.11s^2 0.2109s \quad 1}$	$\frac{0.0752e^{-0.1s}}{s+1}$

Table 7.5 Process gain for positive and negative step changes.

Note: Gain for negative step change (Gain for positive step change).

Manipulated variable (Mag. of step)	RF	RP	RT	ST	PR	PQ	CW	B
DF (5%)	-2.71e-3 (-2.7e-3)	-0.27 (-0.26)	2.18e-3 (2.2e-3)	10.e-3 (9.1e-3)	-8.24e-4 (-8.1e-4)	7.76e-4 (7.8e-4)	-0.017 (-0.016)	0
AF (5%)	-7.82 (-7.82)	-3273 (-2874)	6.39 (6.39)	72.57 (63.15)	7.9 (4.07)	-0.288	126.14 (125.34)	0
A&CF (2%)	1.62 (1.86)	281.6 (294.19)	1.66 (1.60)	-5.86 (-6.58)	2.40 (2.95)		12.99 (13.75)	0.039 (0.019)
CRV (5%)	0	7.20 (8.10)	0	0	0	0	3.09 (2.764)	0
PF (5%)	-11.74 (-20.76)	-1008.5 (-889.9)	0.59 (0.59)	22.01 (18.92)	-3.02 (-2.91)	0	0	-27.45 (-36.13)
SSF (5%)	0	0.087 (0.087)	0	0.037 (0.036)	0	0	0	0
RCWOT (3%)	-0.81 (-0.56)	-49.33 (-21.14)	1.13 (1.09)	2.195 (1.68)	0.017 (0.018)	4.e-4 (4.e-4)	-4.96 (-4.03)	0
CCWOT (3%)	0.22 (0.30)	12.29 (15.09)	0.021 (0.021)	0.46 (0.42)	0	0	2.34 (2.617)	-0.042 (-0.108)

The nominal process gain obtained by averaging the gain for positive and negative step changes are used for steady state RGA analysis. The relative gain array obtained using the nominal process gains is:

$$
\Lambda = \begin{array}{c}
\left[\begin{array}{cccccccc}
-0.0034 & -0.0452 & 0.0055 & 0 & -0.167 & 0 & 0.0891 & \mathbf{1.1205} \\
-0.0423 & \mathbf{0.8553} & 0.1138 & 0.1482 & 0.0184 & 0.0523 & -0.0317 & -0.1139 \\
0.0044 & -0.0237 & 0.0058 & 0 & 0.0035 & 0 & \mathbf{0.9510} & 0.0541 \\
-0.0033 & 0.0443 & 0.0058 & 0 & 0.0009 & \mathbf{0.9471} & -0.0041 & 0.0087 \\
0 & 0.0781 & \mathbf{0.9166} & 0 & 0.0192 & 0 & -0.0139 & 0 \\
\mathbf{1.0384} & 0 & -0.0373 & 0 & 0 & 0 & -0.0010 & 0 \\
0.0063 & 0.0912 & -0.0138 & \mathbf{0.8518} & 0 & 0 & 0.0106 & 0.0540 \\
0 & 0 & -0.0013 & 0 & \mathbf{1.1246} & 0 & 0 & -0.1233
\end{array}\right]
\begin{array}{c} RF \\ RP \\ RT \\ ST \\ PR \\ PQ \\ CW \\ B \end{array} \\
\begin{array}{cccccccc} DF & AF & A\&CF & CRV & PF & SSF & RCWOT & CCWOT \end{array}
\end{array}
$$

The RGA analysis suggests the pairing of the manipulated and controlled variables listed in Table 7.6 for decentralized control (also see Figure 7.12). Internal Model Control (IMC) rules for tuning decentralized controllers are used (Rivera *et al*, 1986). The PI tuning parameters for the controllers are summarized in Table 7.7. The filter time constants are tuned by subjecting the plant to the recommended dynamic step tests and disturbance changes. The controllers were tuned with a sample time of 1 sec. After the tuning parameters were obtained, the controller sample time was increased till the controller performance (especially the reactor pressure response) does not degrade significantly. Using this procedure, a controller sample time of 45 sec. was obtained.

Table 7.6 Pairing between the manipulated and the controlled variables.

Manipulated Variable	Controlled Variable	Controller Number
D Feed	Product Quality	1
A Feed	Reactor Press.	2
A&C Feed	Production Rate	3
Recycle Flow	Compressor Work	4
Purge Flow	B-Composition	5
Steam Flow	Stripper Temperature	6
Reactor CWOT	Reactor Temperature	7
Condenser CWOT	Recycle Flow	8

Table 7.7 PI controller tuning parameters for decentralized controllers.
(Sampling Time: 45 sec.)

Controller Number	Filter Time Constant	Proportional Gain	Integral Time (min.)
1	0.5	5516.7	2.059
2	5.0	-0.0011	16.615
3	2.0	0.1636	0.875
4	1.0	3.903	0.1141
5	1.4	-4.9969	13.9
6	0.02	235.84	0.174
7	0.01	2.135	0.0237
8	5.0	0.3697	0.4782

7.5.4.1. Setpoint Changes

The decentralized controllers are tested for the setpoint changes recommended by Downs and Vogel (1993) (see section 7.4.2). Figures 7.8-7.11 graph the controller responses to setpoint changes in product rate, product quality, reactor pressure, and B-composition. Following are some comments based on these responses:

1. **Production Rate Change**: The production rate is reduced by 15%, i.e. stripper underflow is changed from 14228 to 12094 kg/hr. Figure 7.8 shows the responses obtained from this test. Note that:
   a) Since the production rate is reduced, the amount of G and H produced decreases. Therefore, flow rate of D, E, A and C decrease.
   b) The other operating conditions remain unchanged.
2. **Product Mix Change:** In this test the product quality is changed from 50G/50H to 40G/60H. The responses obtained by making this change are shown in Figure 7.9 and the following points can be observed:
   a) Since the amount of G produced decreases, the amount of fresh D feed to the reactor is reduced.
   b) The amount of H produced is increased, so E feed flow increases.
   c) The production rate does not change and therefore the flow of A and A and C do not change.

3. **Reactor Operating Pressure Change:** A setpoint change in reactor pressure is made so that the pressure changes from 2705 to 2645 kPa. The responses to this test are shown in Figure 7.10.

   a) The flow of various reactants does not change as the production rate and the product quality are the same.

4. **Purge Gas Composition of Component B change:** The responses obtained by changing the composition of component B in the purge from 13.82 mole % to 15.82 mole % are shown in Figure 7.11.

   a) The inlet flow rates of the reactants remains unchanged; the product quality and production rate are also the same.

   b) The composition of B in the purge increases and the purge flow rate decreases.

7.5.4.2. Disturbance Response

Among the 20 process disturbances, IDV(1) through IDV(20), listed in Downs and Vogel (1993), IDV(6) - A feed loss - results in an interesting scenario. As stated before, A-feed (stream 1) is a make-up stream for the deficiency of A entering in stream 4. Reactants A and C are consumed in equimolar quantities in the reactor. If there is loss of A-feed, then there will be excess of reactant C which if not purged completely will accumulate. To minimize the production loss when A feed is lost, the purge valve should

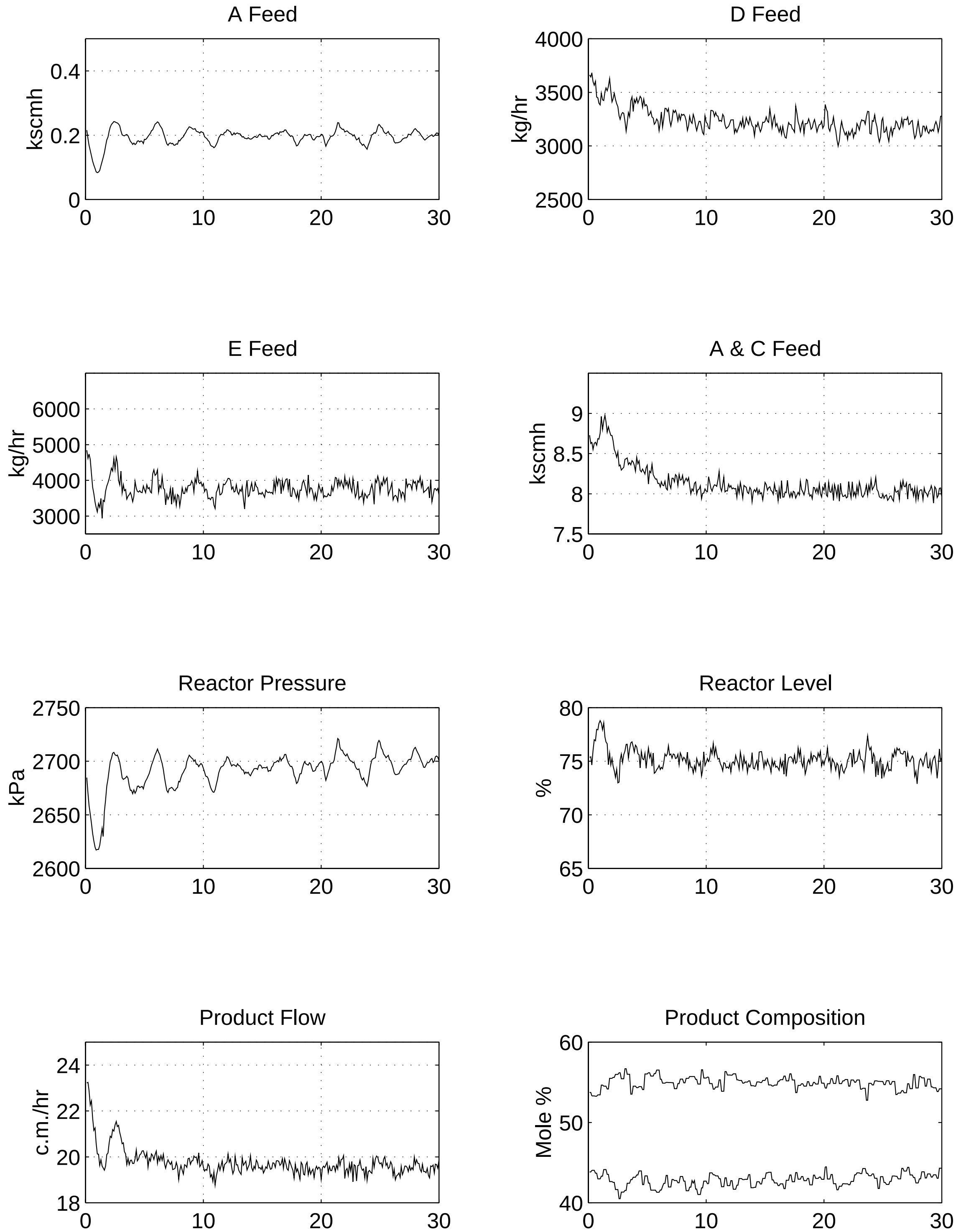

Figure 7.8 Responses for production rate step change by -15%.
(x-axes units in hr)

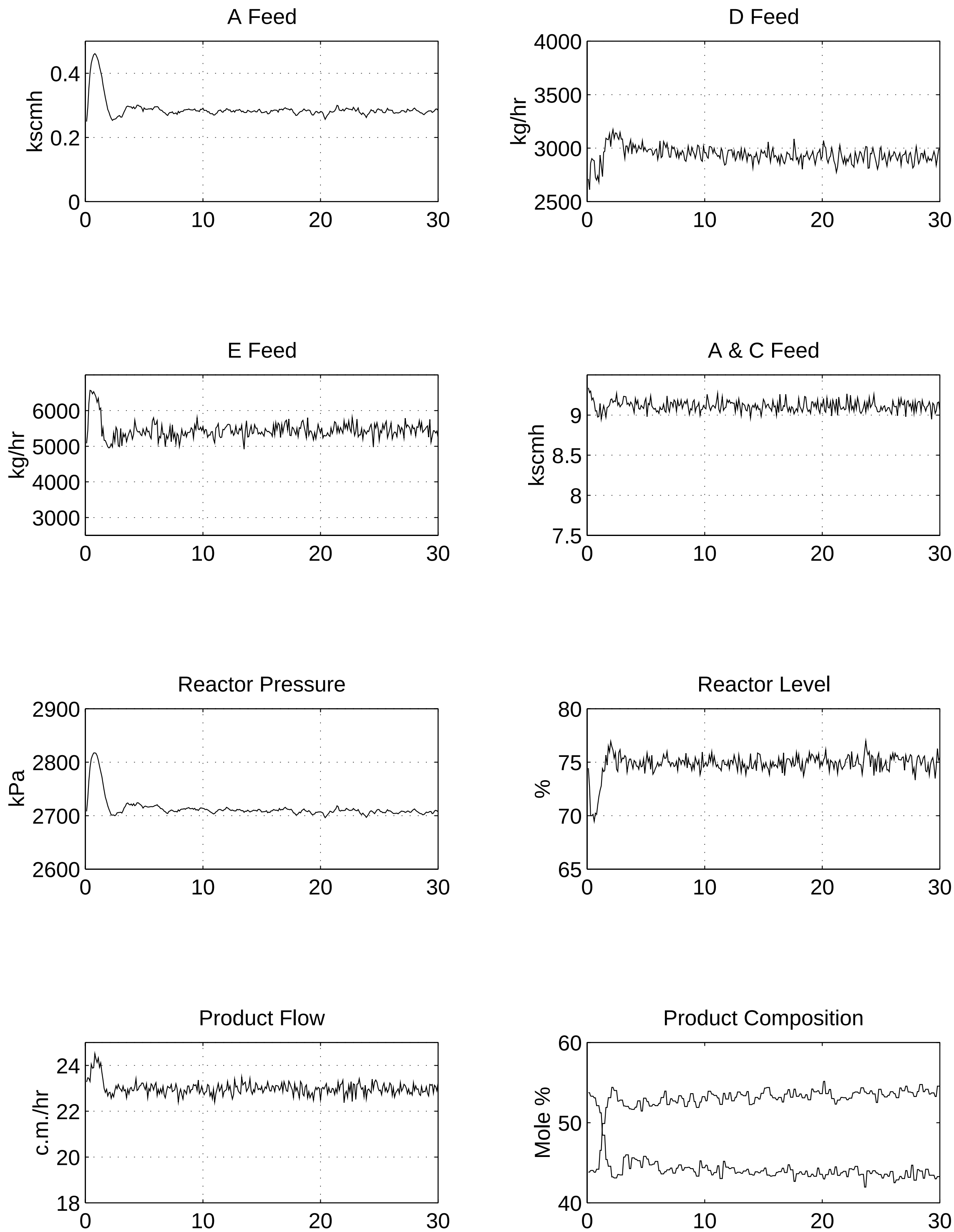

Figure 7.9 Responses for product mix step change from 50G/50H to 40G/60H.
(x-axes units in hr)

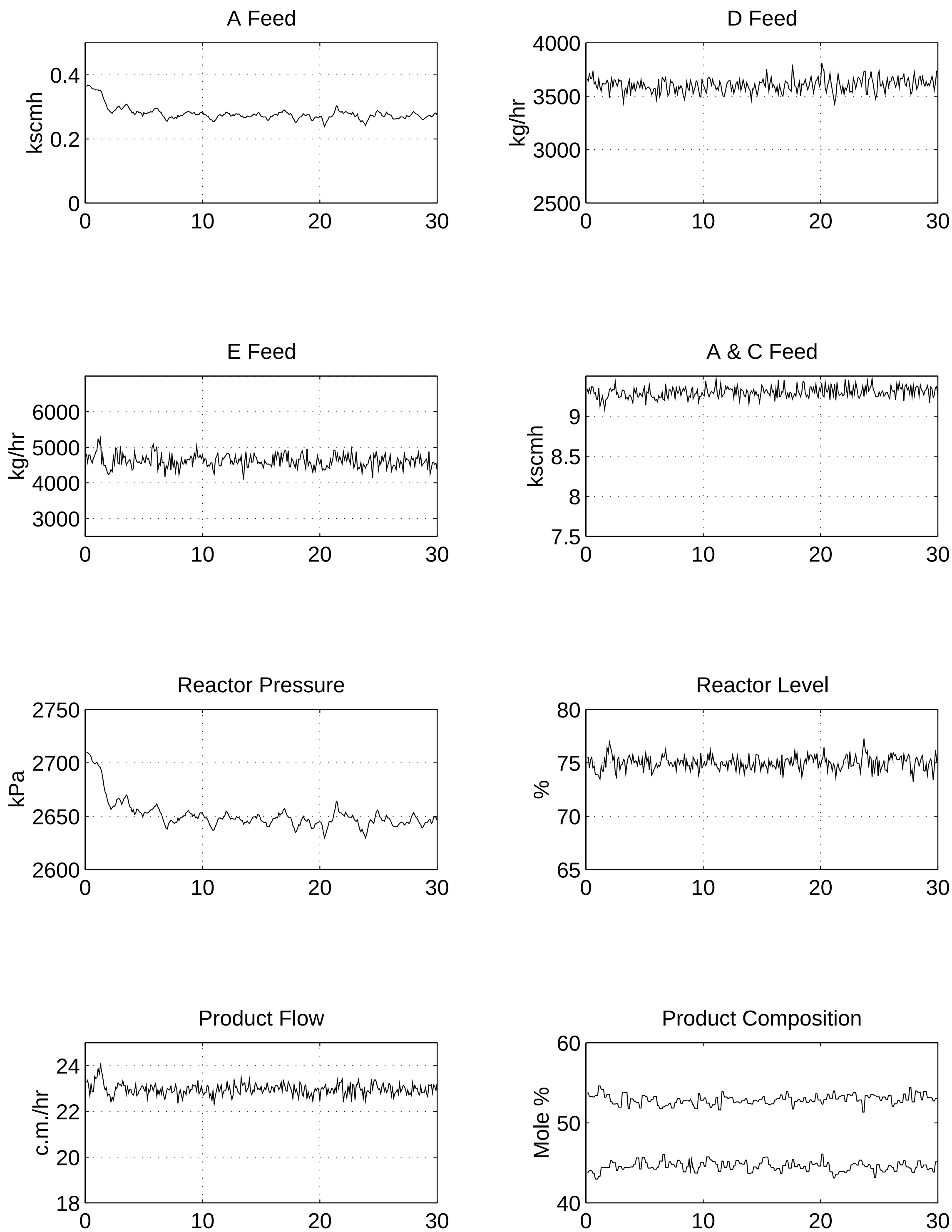

Figure 7.10 Responses for reactor pressure step change by -60 kPa.
(x-axes units in hr)

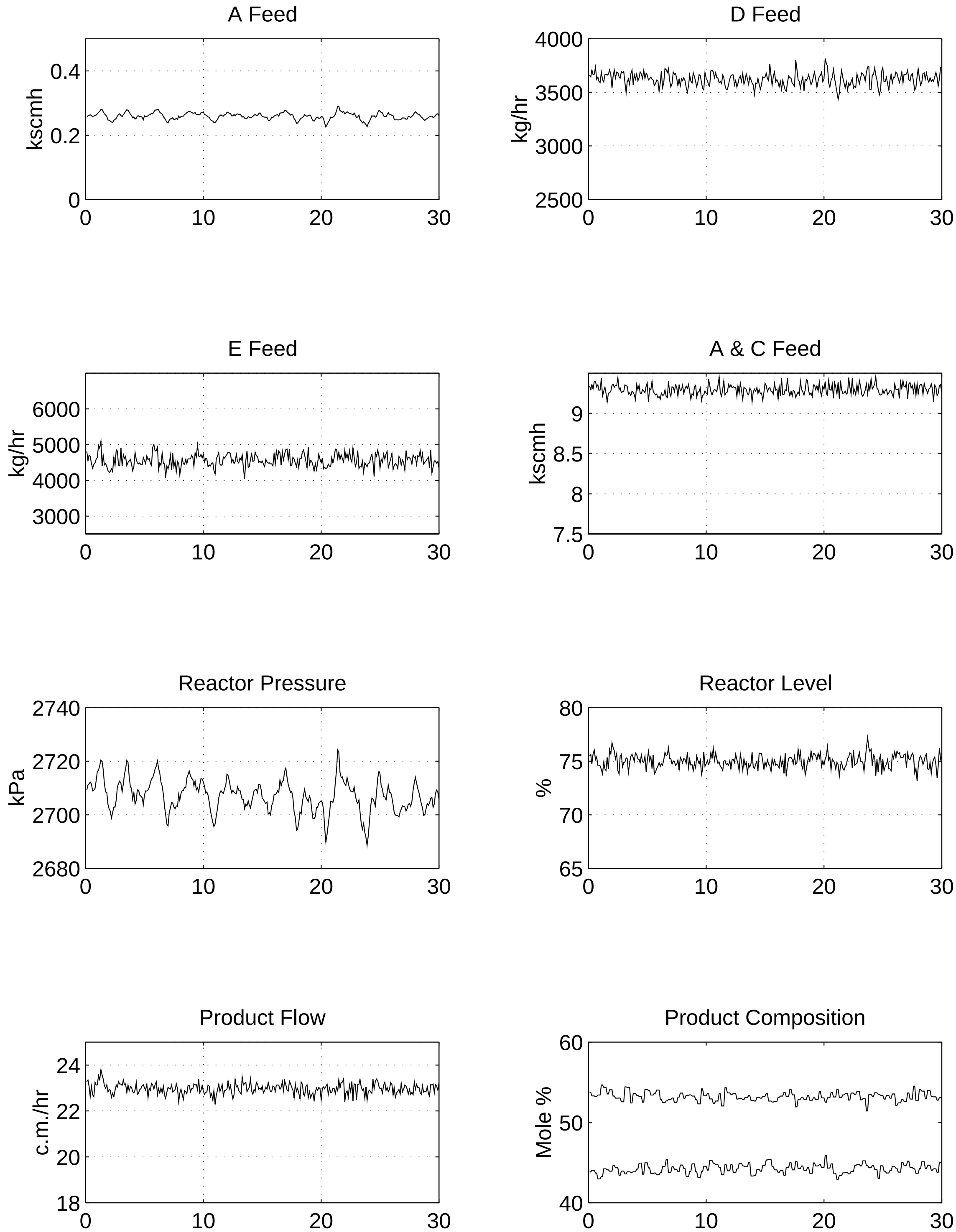

Figure 7.11 Responses for purge gas composition of component B step change by +2%.
(x-axes units in hr)

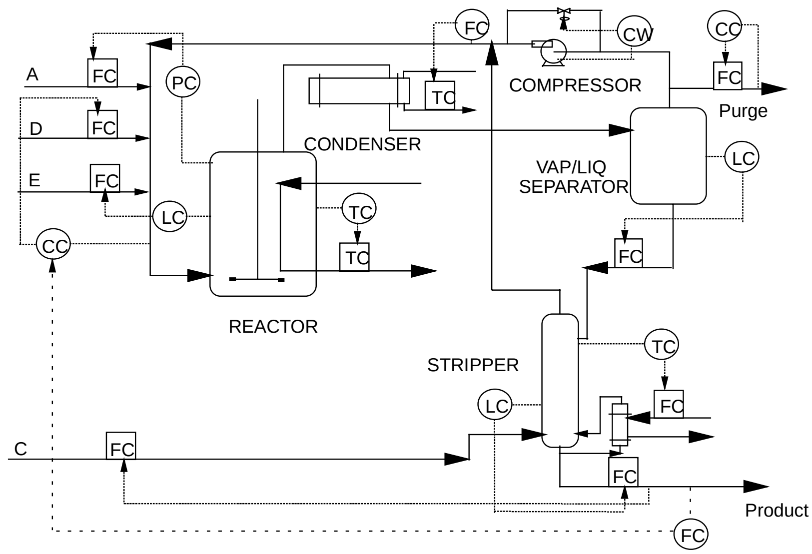

Figure 7.12 Decentralized control of the T-E plant.

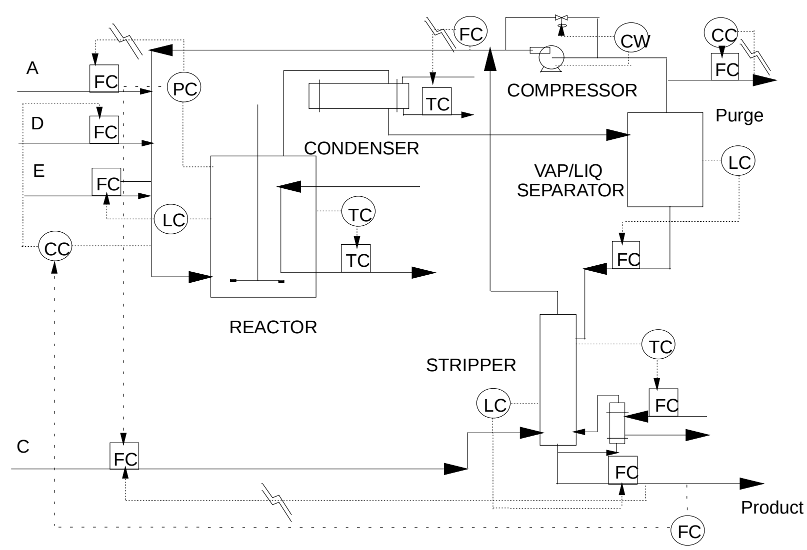

Figure 7.13 Over-ride controllers necessary to handle IDV(6) - loss in A-feed.

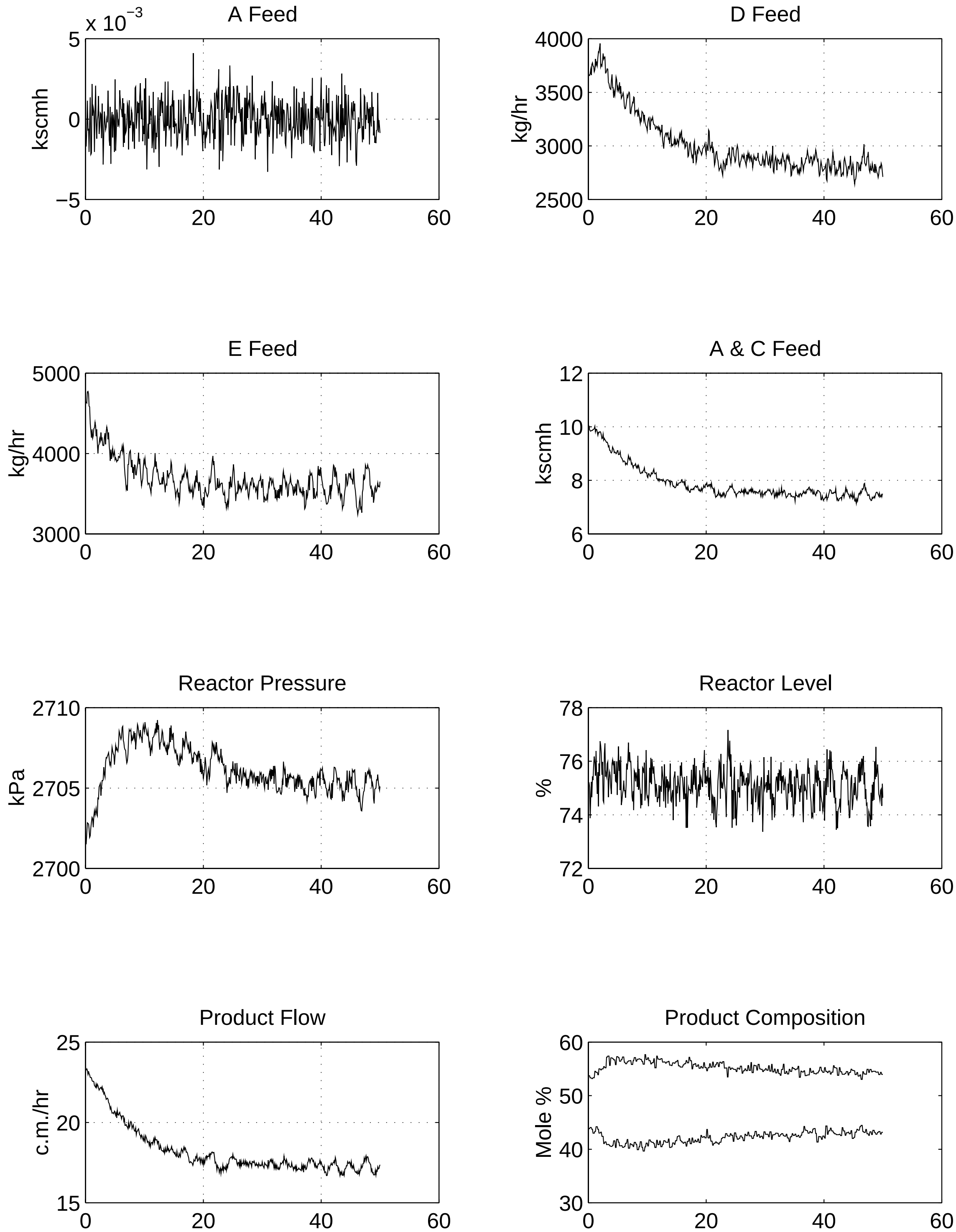

Figure 7.14 Response for IDV(6) (loss in A-feed) using over-ride controllers.

be opened completely. Simulation of this disturbance suggests that in spite of opening the purge valve completely there is accumulation of component C. This suggests that the production rate setpoint should be reduced. We implement this by introducing an over-ride control for the production rate control. The over-ride controller opens the production rate control loop, and then controls the reactor pressure using A-C feed stream. The recycle flow setpoint is not known when A-feed is lost and is therefore left open. The reactor pressure setpoint may be changed to the maximum allowable value (2895 kPa) so that there is minimum loss in production. The over-ride controllers are shown schematically in Figure 7.13 and the responses are plotted in Figure 7.14.

7.5.5. Multivariable Unconstrained Control

Dynamic Matrix Control (DMC) is a multivariable control algorithm that uses a process model to predict P future values of error between the controlled variables and the desired setpoint values. The prediction is based on the past manipulated variable changes. M future manipulated variable changes are computed by solving the following least square problem that minimize the future error between the controlled variables and the setpoints:

$$\underset{\Delta\mathbf{u}}{Min} \left\| \mathbf{y}^p - \mathbf{y}^s \right\|_{2,\Gamma}^2 + \left\| \Delta\mathbf{u} \right\|_{2,\Lambda}^2$$

P and M are called the prediction and the control horizons respectively. Λ and Γ are diagonal matrices and are respectively called the move suppression parameter and the output weight. Of the M future manipulated variable changes computed, the first move is applied at each sample time. The corrective action results in the controlled variable predictions to change and therefore the predicted error changes. As a result, the manipulated variable changes computed at the previous sample time become sub-optimal. Hence, the above optimization problem is solved at every sample time.

Dynamic tests performed on the decentralized controllers discussed in section 5.2 suggest the following:

1. There is significant interaction between the controlled variables (see Figure 7.8-7.11).
2. The reactor temperature and the separator temperature have fast response times. They have insignificant interaction with other controlled variables.
3. Setpoint change in B-composition in the purge indicates that it has little influence on other controller variables (see Figure 7.11).

Therefore, the reactor temperature and the separator temperature along with B-composition are controlled using decentralized, multi-loop PI controllers. The manipulated variables used for control of these variables are those corresponding to the pairing listed in Table 7.6. The remaining variables (in Table 7.6) are controlled using unconstrained DMC. The DMC algorithm utilizes the linear time-invariant models given in Table 7.5. A sample time of 1 minute is used. The other tuning parameters are summarized in Table 7.8. The output weights were determined by first scaling all controlled variables to 1, and then introducing weights that indicate the relative importance of the controlled variable. The move suppression parameters were determined using a trial-and-error tuning procedure and these values were evaluated by subjecting the plant to the recommended step changes. Figures 7.15, 7.16 and 7.17 show the responses for step changes in production rate, product quality and reactor pressure respectively. The responses are smoother and have less interaction compared to the corresponding decentralized controller responses (see Figures 7.8, 7.9 and 7.10). The linear DMC controller has been tested only for the setpoint changes. The recommended disturbance changes have not been tested.

Table 7.8 DMC Tuning Parameters

Sample Time=1 min., P=60, M=15

Controlled Variable	**Output Weight**
Recycle Flow	0.2
Reactor Pressure	0.0025
Product Quality (G/H)	140
Stripper Underflow	0.4
Compressor Work	0.05

Manipulated Variable	**Move suppression**
D-Feed	5
A-Feed	75
A&C-Feed	40
Comp. Recycle valve	20
Separator CWOT	10

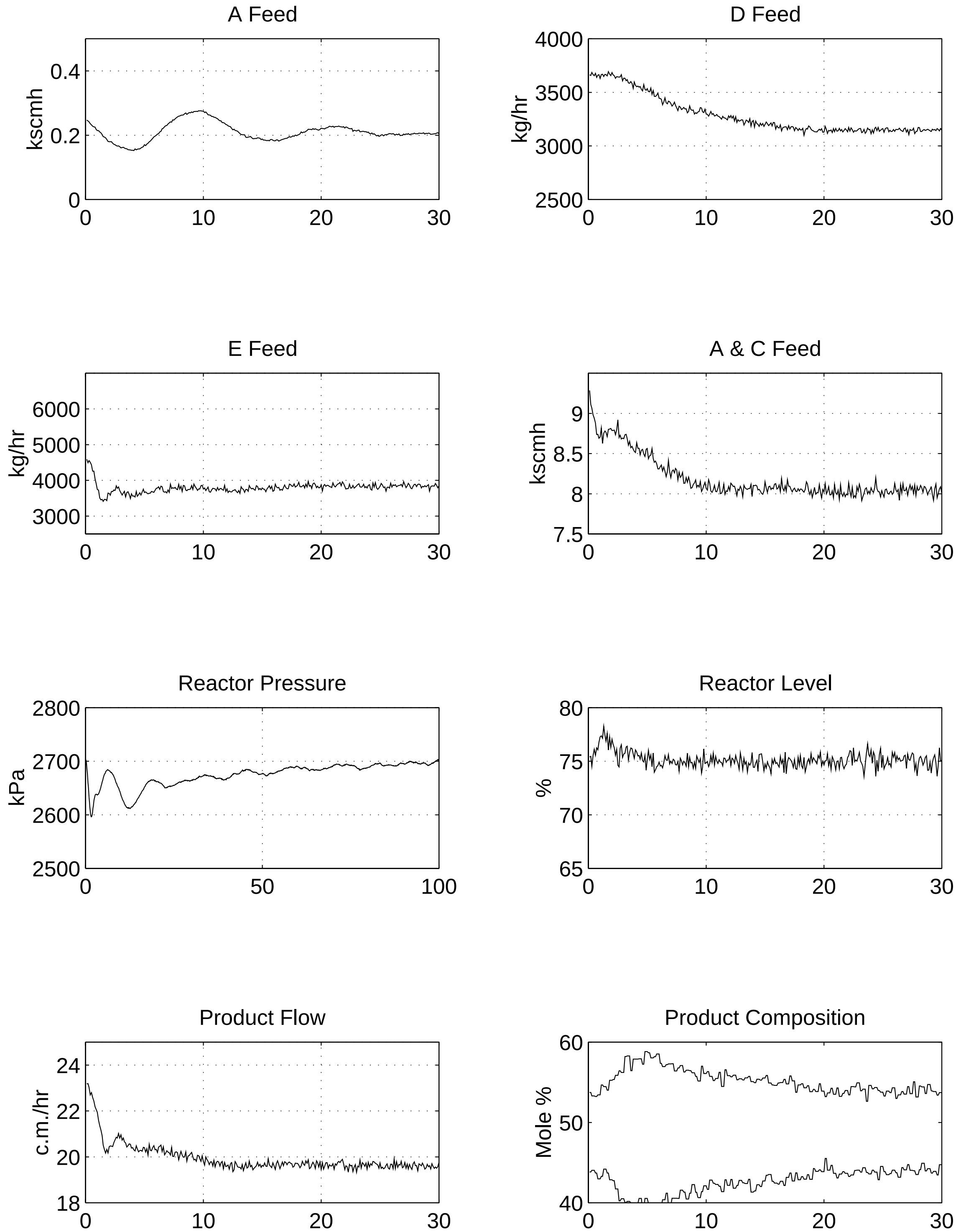

Figure 7.15 DMC responses for production rate step change by -15%.
(x-axes units in hr)

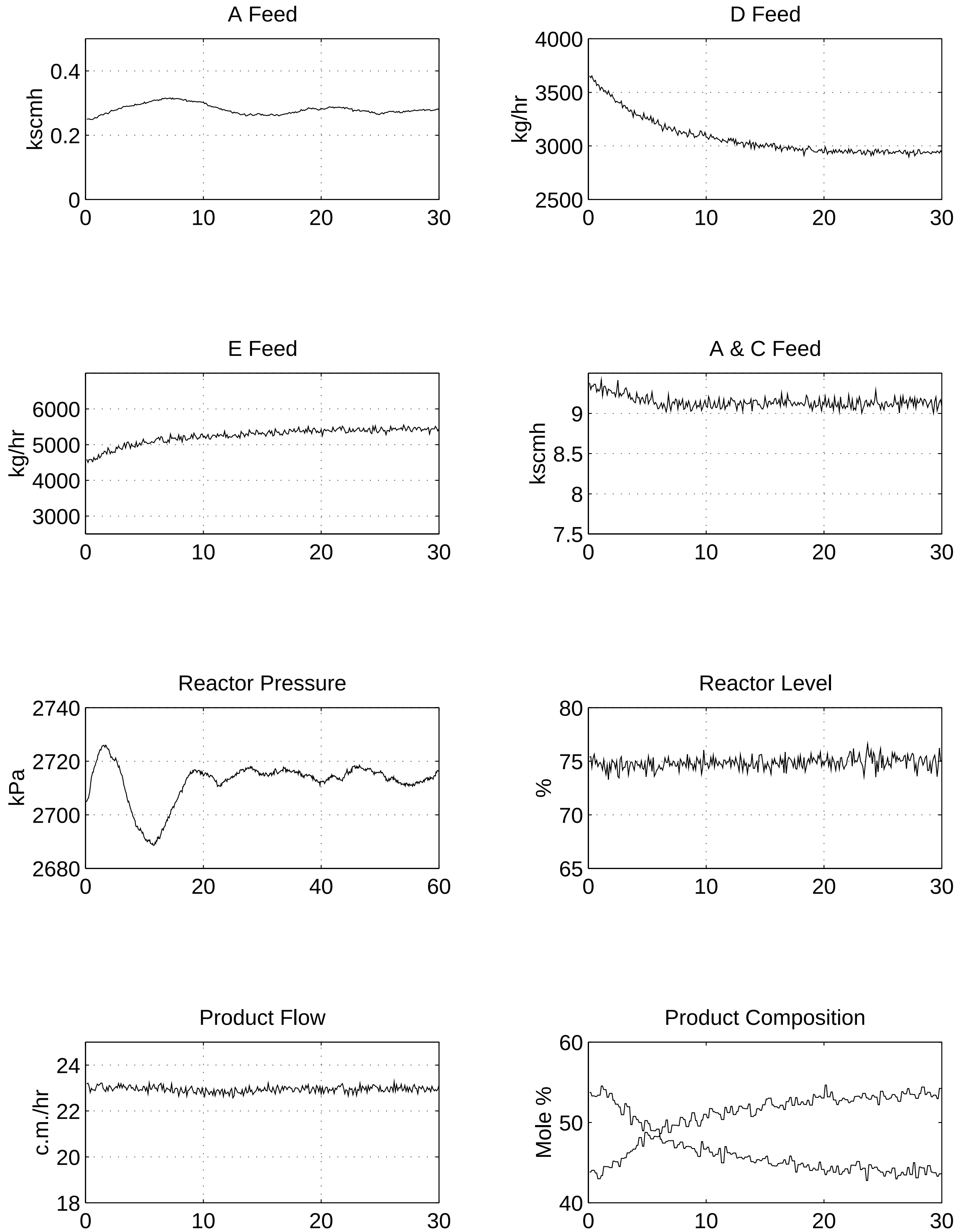

Figure 7.16 DMC responses for product mix step change from 50G/50H to 40G/60H. (x-axes units in hr)

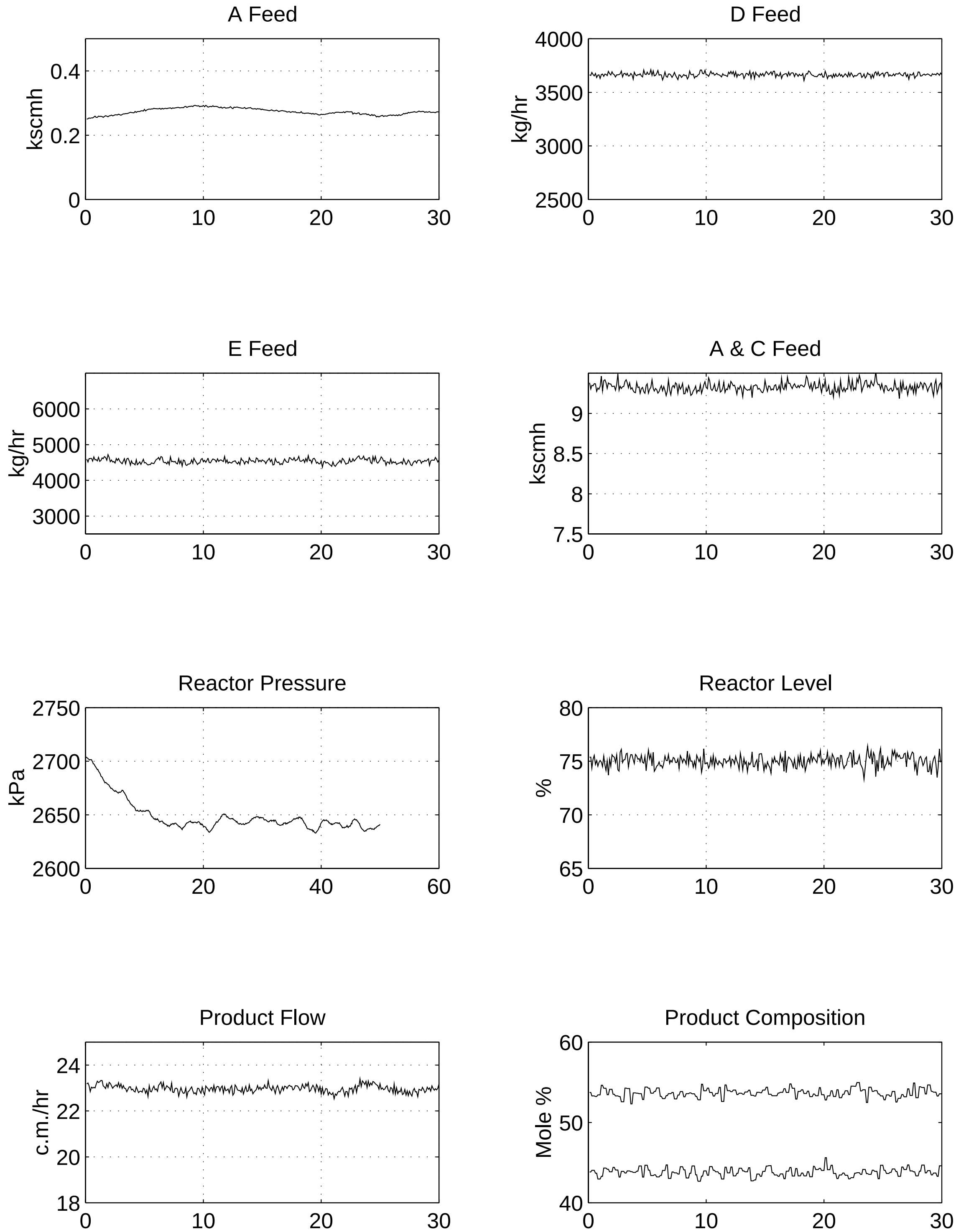

Figure 7.17 DMC responses for reactor pressure step change by -60 kPa.
(x-axes units in hr)

7.6. Off-line Optimization

Downs and Vogel (1993) have listed six operating modes for the T-E plant. There are three different product mix with G/H mass ratios 50G/50H, 10G/90H and 90G/10H. The production rate is either fixed or maximum. The steady state values for the 41 measured variables at the nominal operating mode (defined by a product mix of 50G/50H and a production rate of 14076 kg/hr) are available in Downs and Vogel (1993). The measured variables at other operating modes are not provided. In this section, we present the methodology to determine the measured variables at 50G/50H, maximum production operating mode using off-line optimization.

Downs and Vogel have not provided the model equations simulating the process. Instead, a FORTRAN subroutine, TEFUNC, that simulates the plant has been made available. The input arguments to this subroutine are the current state variables (x), the current manipulated variables (u), and the current time (t). The output from the subroutine is the derivative of the states (x). The subroutine can be used for both steady state and dynamic process simulation. The input argument t is set equal to 0 for steady state simulation. No time delay nor measurement noise is introduced in this case. Using TEFUNC, our objective is to determine the state and the measured variables at the new operating mode.

The state variables *x* and the measured variables *xmeas* at the new operating mode can be determined by solving the following constrained non-linear optimization problem:

$$\underset{x}{Min}\ \phi(x)$$

Subject to:

(C1) $x = f(x, t = 0) = 0$ - Steady State Condition.
(C2) $xmeas_{\min} \le xmeas \le xmeas_{\max}$ - Process Operating Constraints.
(C3) Product Quality Constraint.
(C4) $u_1, \ldots, u_{12} = x_{39}, \ldots, x_{50}$ - Independent Variables.
$0 \le u_1, \ldots, u_{12} \le 100$ - Manipulated Variable Constraints.

The objective function $\phi(x)$ is the production rate. There are 50 optimization variables, $x(1{:}50)$, in this problem and a number of process and production constraints. At steady state, the derivatives of the states, x, must all be equal to zero. Since the objective in this study is to drive the process to a new steady state, we introduce constraint (C1) that imposes $x = 0$. The manipulated variables, *u*, are included in the state vector *x* and therefore only $x(1{:}38) = 0$ should be considered explicitly. The measured variables should be within safe limits, such constraints are specified as in constraint (C2). The product quality is specified as an equality constraint (C3). The manipulated variable constraints (C4) specify that at steady state these variables should lie between 0 and 100.

The degree of difficulty, defined as: degree of difficult=number of variables - number of equality constraints, of this constrained non-linear optimization problem is 50-(38+1)=11 (50 optimizing variables, 38 equality constraints (C1), and 1 product quality constraint (C3)). A few simplifying assumptions (like, fixed reactor, separator and stripper level, and fixed reactor pressure) are made to reduce the degree of difficulty to 7. The optimization problem is then solved using a FORTRAN code for the Generalized Reduced Gradient (GRG) method (Abadie and Carpentier, 1969) called GRG2 (Lason *et al*, 1980).

7.6.1. The Non-Linear Optimization Problem

GRG solves a non-linear optimization problem of the form:

$$\begin{array}{rll} & \underset{x}{Min}\ g_{m+1}(x) & \\ \text{subject to:} & g_i(x) = 0 \qquad i = 1,2,\ldots,m_1 \geq 0 & \\ & 0 \leq g_j(x) \leq ub(n+j) \qquad j = m_1+1,\ldots,m \geq m_1 & \\ & lb(k) \leq x_k \leq ub(k) \qquad k = 1,\ldots,n & \end{array} \tag{P1}$$

where $x=\{x_1, x_2, ..., x_n\}$ is a vector of n variables. The scalar function $g_{m+1}(x)$ is called the *objective function*, and the functions g_i (i=1,2,...,m) consists of m_1 equality and m-m_1 inequality constraints. The functions g_i are assumed to be differentiable. $lb(k)$ and $ub(k)$ are the lower and the upper bound of variable x_k.

GRG2 solves problem (P1) by decomposing it into a sequence of reduced problems. GRG2 first determines the active equality constraints, say equal to *nb*, and uses it to solve for *nb* of the *n* variables in *x* in terms of the *n-nb* remaining variables and the slack variables (slack variables are introduced to convert the inequality constraints into equality constraints). These *nb* variables are called the *basic variables*, and are denoted by *y*. The rest are called the *non-basic variables*, denoted by z. The problem is thus reduced to the following optimization problem

$$\begin{aligned} &\underset{z}{Min}\, g_{m+1}(y,z) = F(z) \\ &\text{subject to: } \; l < z < u \end{aligned} \qquad \text{(P2)}$$

The reduced problem is then solved by a gradient method. At each iteration a search direction is formed from the gradient and a one-dimensional search is initiated. The search is repeated at every iteration till the Kuhn-Tucker conditions are met, or if the user defined terminating criteria are satisfied.

The optimization problem stated in problem (P1) for 50G/50H product mix and maximum production is formulated as follows

$$\underset{x}{Min}\, g_{45}(x) = \text{-}xmeas(17)\times 9.21\times (xmeas(40)\times 62 + xmeas(41)\times 76)$$

$$\text{Subject to : } g_{1,2,...,38}(x) = yp(1{:}38) = x(1{:}38) \qquad \text{(C1)}$$

$$\left.\begin{aligned} g_{39}(x) &= 1 - \frac{xmeas(7)}{RP} \\ g_{40}(x) &= 1 - \frac{xmeas(8)}{RL} \\ g_{41}(x) &= \frac{xmeas(9)}{150} \\ g_{42}(x) &= 1 - \frac{xmeas(12)}{50} \\ g_{43}(x) &= 1 - \frac{xmeas(15)}{50} \end{aligned}\right\} \qquad \text{(C2)}$$

$$g_{44}(x) = 1 - \frac{xmeas(40) * 62}{xmeas(41) * 76} \qquad \text{(C3)}$$

$$0 \le u_{1,...12} \le 100 \qquad \text{(C4)}$$

where, RL and RP are the desired reactor level and reactor pressure respectively. The objection function $g_{45}(x)$ computes the production rate in kg/hr. The molar density of the product stream is assumed to be 9.21 kgmol/m^3 (Downs and Vogel, 1993). Since the maximum of the objective function is sought, $g_{45}(x)$ is multiplied by -1. For the fixed production problem, both the reactor level and the reactor pressure were set equal to their values at the 50G/50H nominal case (the reactor level was fixed at 75% and the reactor pressure was fixed at 2705 kPa.). Both variables were fixed at these values until the final portion of the production maximization procedure, when each was varied to determine their effect on the maximum obtainable production rate. Decreasing the reactor level and increasing the reactor pressure increased the maximum production rate. In addition, in constraints g_{42} and g_{43}, the stripper level and the separator level were fixed at the base case value of 50%. Varying the separator and the stripper levels at the optimal solution did not have any significant influence on the maximum production. The mass flow rate of products G and H in the product stream are equal for the 50G/50H product quality. g_{44} specifies this constraint. The lower and upper bounds on the manipulated variables u_1, ..., u_{12} were specified as 0 and 100 respectively. All constraints were scaled to give each constraint equal importance.

7.6.2. Convergence Procedure

Proper scaling of the variables and the functional constraints is extremely important for successful convergence to an optimal solution. Lasdon *et al* (1980) recommend scaling all variables and functions to have absolute value less than 100 but much larger than 10^{-4}. Further, they recommend that the scaling should be such that a unit change represents a small but significant change in that variable.

The vector x at the nominal steady state is known. It is extremely difficult, if not impossible, to determine *a priori* the expected change in each state after successful optimization. We therefore recommend optimization without scaling till the new state variables vector is reasonably close to the optimal value. The states are then scaled as follows:

$$xx = \frac{x - x_i}{x_f - x_i} \times 100$$

where, xx are the scaled states, x_f are the states near the optimal value, x_i are the states at the nominal operating condition.

The following procedure was followed to obtain an optimal solution:

1. Using material balance equations involving the raw-material inlet and the product streams, initial guesses for raw-material inlet feed rates were computed. The states $x(39:50)=u(1:12)$ are fixed based on these guesses. The lower bound on fractional change in objective function EPTOP is set equal to 0.001.
2. On executing GRG2, if the objective function change is smaller than EPTOP, then we update the initial guesses for the states as the final result of the last run, decrease

EPTOP by a factor of 10, and repeat the optimization run. This procedure is continued till EPTOP is 10^{-6}.

3. The states are scaled using the initial states as the ones at 50G/50H nominal case and the final state as those obtained in step 2. The initial guess for the scaled states, *xx*, is set between 90% and 100%. EPTOP is set equal to 0.001.
4. After each run of GRG2, the initial guesses are updated, EPTOP is reduced by a factor of 10 till EPTOP is 10^{-6}.
5. The states when no constraints are in violation are considered to be an optimal solution.

Table 7.8 lists the measured variables at 50G/50H nominal operation (column 2), and 50G/50H maximum production (column 3). We compare our results with those reported in Ricker (1993) for 50G/50H maximum production and minimum cost (see Table 7.8, column 4). The maximum production we obtain is 36.61 m^3/hr and that reported by Ricker is 36.01 m^3/hr. The discrepancy could be due to a lower value of upper bound for reactor pressure, a higher value of lower bound for reactor level, the minimum cost objective considered, and a different optimization routine used by Ricker.

From the steady state measured variables at maximum production (Table 7.9, columns 3 and 4), the following observations can be made:

1. D-feed is the limiting reactant. The D-feed valve saturates to the maximum value of 100%.

2. The purge valve also saturates to 100%, suggesting maximum purge flow rate.

3. Reducing the reactor level to the lower bound and increasing the reactor pressure to the upper bound, increases the production by a small amount. This suggests that maximum possible gas space in the reactor is favorable to achieve maximum production. However, to operate the plant with a safe margin from the bounds, we fix the reactor pressure at 2705 kPa and the reactor level at 55%.

4. The amount of byproduct F in the purge increases from 2.26 mole% to 12.51 mole%. By including the minimum cost objective the value is reduced to 5.72 mole%. This suggest that raw material waste in the form of by-products reduces to minimize cost.

5. The composition of B in purge decreases from 13.82 to 8.36 mole % when the production rate is maximized. This indicates that the purge flow is increased to maintain a lower composition of the inert. With the minimum cost objective, the B composition in purge increases to 15.90 mole %, i.e., the loss of raw material in the purge stream is reduced.

Table 7.9 Steady state operating conditions at 50G/50H operating mode.

Measured Variable	Mode 1 50G/50H	Mode 2 Max Prod 50G/50H	Ricker's result
A Feed	0.25052 kscmh	0.623	0.503
D Feed	3664 kg/hr	5811	5811
E-Feed	4509.3 kg/hr	7212.1	7244
A&C Feed	9.3477 kscmh	14.96	14.73
Recycle Flow	26.902	27.79	29.22

	kscmh		
Feed rate	42.339 kscmh	52.54	53.76
Reactor Pressure	2705 kPa	2895	2800
Reactor.Level	75 %	50.0	65.0
Reactor Temperature	120.4 Deg C	136.40	128.2
Purge rate	0.33712 kscmh	0.89	0.462
Separator Temp	80.109 Deg C	91.58	74.1
Separator Level	50 %	50.0	50.0
Separator Pressure	2633.7 kPa	2789.6	2699
Separator underflow	25.16 m^3/hr	40.325	40.06
Stripper level	50 %	50.0	50.0
Stripper pressure	3102.2 kPa	3482.4	3365
Stripper undeflow	22.949 m^3/hr	36.61	36.04
Stripper temperature	65.731 Deg C	74.9	51.5
Steam flow	230.31 kg/hr	355.55	6.87
Compressor work	341.43 kW	250.51	263.2
Reactor CWOT	94.599 Deg C	102.64	96.6
Condenser CWOT	77.297 Deg c	85.69	73.5
Reactor Feed Analy.			
A	32.188	35.90	36.4
B	8.8933	4.56	8.78
C	26.383	26.22	22.36
D	6.8820	7.84	7.95
E	18.776	11.29	17.01
F	1.6567	7.94	3.88

Table 7.9, continued

Measured Variable	Mode 1 50G/50H	Mode 2 Max Prod 50G/50H	Ricker's results
Purge Gas Analy.			
A	32.958	39.65	40.94
B	13.823	8.36	15.90
C	23.978	22.39	15.68
E	1.2565	0.19	0.68
E	18.579	7.27	15.41
F	2.2633	12.51	5.72
G	4.8436	6.49	3.85
H	2.2986	3.14	1.82

Product Analysis			
D	0.01787	3.221e-3	0.02
E	0.8357	0.214	1.21
F	0.09858	0.36	0.44
G	53.724	53.91	53.35
H	43.828	43.98	43.52

7.6.3. Moving the Plant from One Operating Mode to Another

In this section the procedure adopted to move the plant from 50G/50H nominal operation to 50G/50H maximum production is explained. To move the plant to the new operating conditions from the nominal conditions without causing the reactor pressure to reach the shut-down limit, the following guidelines were used:

1. The production rate should be increased to the final value in the shortest possible time. A fast rate of change results in the reactor pressure to reach the shut-down limit. For this example, the production rate is increased by 10-15% every 5 hr, if possible.
2. The reactor level should be decreased gradually (by 10-15% every 5 hr.) to the final value of 55%. Decreasing the reactor level initially may cause the reactor pressure to reach the shut-down limit due to accumulation of reactants A and C. If such a situation arises, the reactor level is increased initially to make more reactant E available. The reactor level is then gradually decreased.
3. The reactor temperature and the stripper temperature are fast responding. The reactor temperature is changed if reactant A and C accumulate in spite of the availability of reactants D and E.

4. The compressor work is fast responding and is decreased to the final value in one step.
5. The recycle flow is decreased when the output response suggests that there is an offset in this controlled variable.
6. The B-composition in purge is decreased whenever the recycle flow has an offset.

The same procedure is used to determine the setpoint changes for the decentralized controllers and the linear DMC controller.

7.6.3.1. Moving to New Operating Conditions Using Decentralized Control

The following scheme for setpoint changes was arrive at using the above procedure:

1. ***At Time=0***: Product flow is increased to 25.24 m^3/hr.
2. ***At Time=5 hr***: Product flow is increased to 28.68 m^3/hr. Recycle flow is decreased to 26 kscmh. Reactor level is increased to 85%. Reactor temperature is increased to 128 Deg C. Compressor work is increased to 237.19 kW. B composition in purge stream is decreased to 10 mole %.
3. ***At Time=10 hr***: Product flow is increased to 32.42 m^3/hr. Reactor temperature is increased to 133.9 ^oC.
4. ***At Time=20 hr***: Product flow is increased to 34.42 m^3/hr. Recycle flow is decreased to 26.11 kscmh. B composition in purge stream is decreased to 8.73 mole %.
5. ***At Time=25 hr***: Reactor level is decreased to 75%. Product flow is increased to 36.1 m^3/hr.

6. ***At Time=35 hr***: Reactor level is decreased to 66%. The stripper temperature is decreased to 73.45 ^{o}C.
7. ***At Time=40 hr***: Reactor level is decreased to 57%.
8. ***At Time=45 hr***: Reactor level is decreased to 55%.

The controller responses obtained by making the above setpoint changes are shown in Figure 7.18.

7.6.3.2. Moving to New Operating Conditions Using DMC Control

The DMC controller was unstable when the production rate was increased by 157.3% in one step. The controller is stable, however, when the setpoints are changed gradually using the procedure described before. The following setpoint changes were made:

1. ***At Time=0***: Product flow is increased to 25.24 m^3/hr. The reactor level is decreased to 70%.
2. ***At Time=5 hr***.: Product flow is increased to 27.54 m^3/hr. The reactor level is decreased to 65%.
3. ***At Time=10 hr***.: Product flow is increased to 29.83 m^3/hr. The reactor level is decreased to 60%. The stripper temperature is decreased to 61 ^{o}C.
4. ***At Time=20 hr***.: Product flow is increased to 32.13 m^3/hr. The reactor level is decreased to 57%. The recycle flow is decreased to 23 kscmh. The reactor temperature is increased to 132 ^{o}C.

5. ***At Time=30 hr***.: Product flow is increased to 34.42 m^3/hr. The compressor work is increased to 360 kW. Reactor temperature is increased to 133.9 ^oC. The reactor level is decreased to 55. Composition of B in purge is decreased to 7.18 mole %.
6. ***At Time=35 hr***.: Product flow is increased to 36.03. Reactor level is decreased to 55%.

The responses obtained using this approach are shown in Figure 7.19.

The following observations can be made by comparing the decentralized controller responses (Figure 7.18) and the DMC controller (Figure 7.19) responses while moving the process to the new operating conditions:

1. The DMC controller being a multivariable algorithm results in less interaction in the controlled variables than the decentralized controllers. Therefore, the DMC controller yields a much smoother transition to the new operating mode.
2. When the same setpoint changes as the decentralized controllers are used to move the process to the new operating mode, the DMC controller was unstable. A key to successful transition to the new operating conditions is providing proper setpoint targets for the controlled variables. A linear programming problem that provides manipulated variable setpoints should yield a better and a gradual transition of the process. This is a topic that needs further investigation.

7.7. Conclusions

In this chapter a modular and a hierarchical procedure for the implementation of a plant-wide control scheme was presented. The procedure was applied to the T-E challenge problem. Decentralized controllers and unconstrained multivariable controllers were developed and tested. An off-line optimization using the non-linear process model was used to obtain optimal operating conditions. The decentralized and the unconstrained DMC controllers were then used to move the process to the new operating mode.

While we succeeded in implementing and testing the plant-wide control scheme for the T-E problem, a number of aspects remain to be investigated:

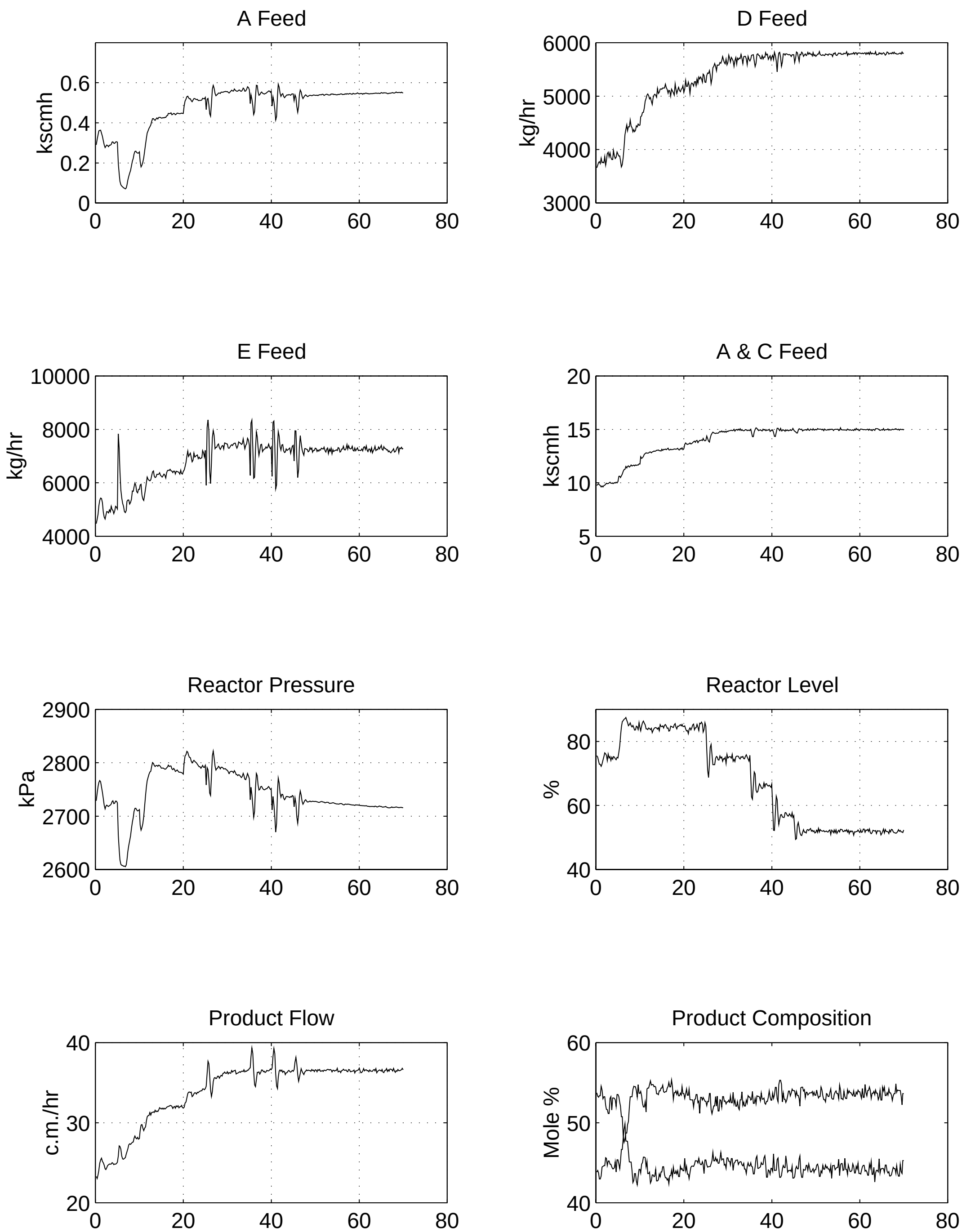

Figure 7.18 Decentralized controller response obtained when moving from nominal operating mode to maximum production with product mix 50G/50H.
(x-axes units in hr)

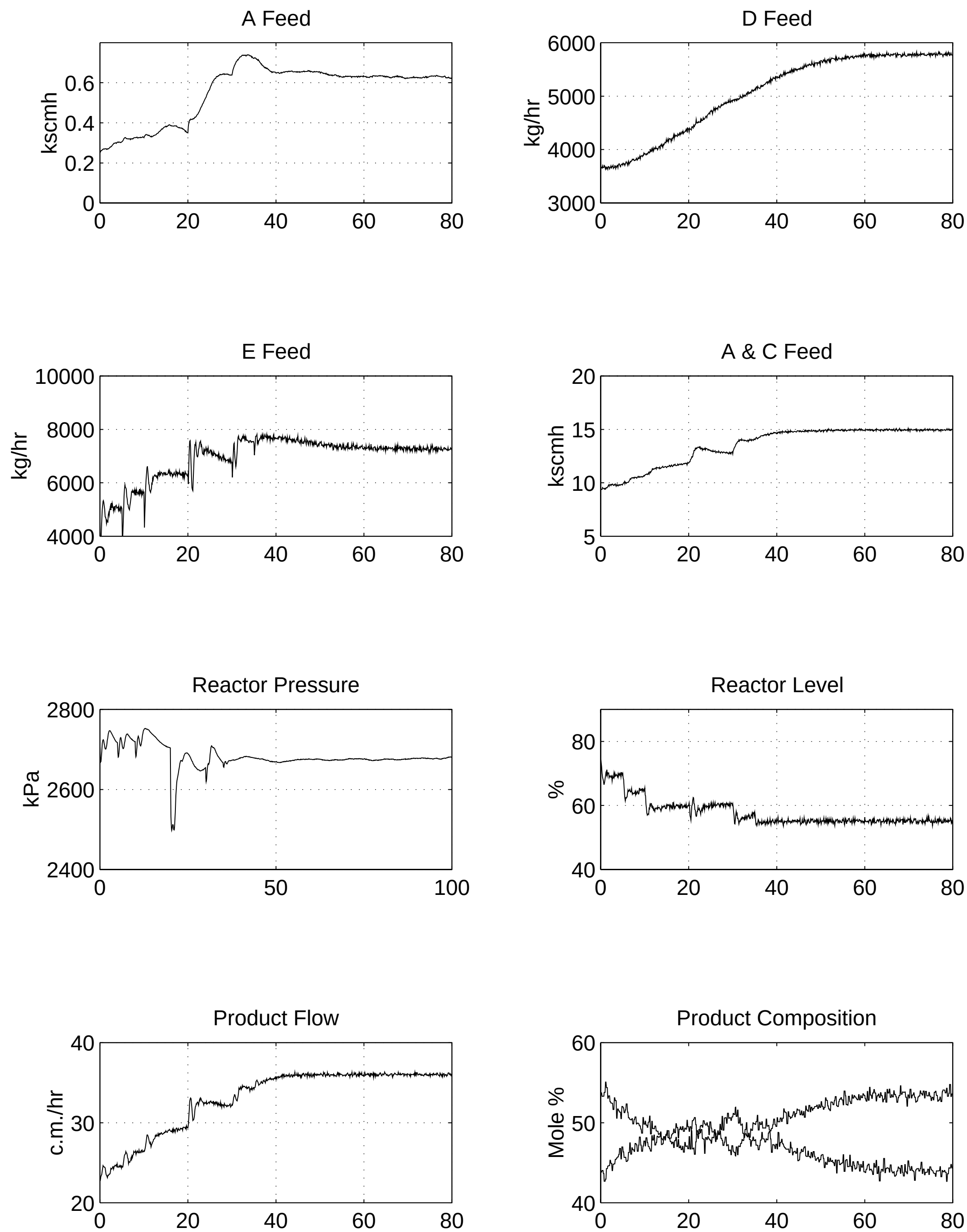

Figure 7.19 DMC controller response obtained when moving from nominal operating mode to maximum production with product mix 50G/50H.
(x-axes units in hr)

1. *Synthesis*

a) Configuration I for level controllers was used in this study. How Configuration II will perform compared to Configuration I needs to be investigated.

b) The controlled variable selection was based on process understanding. It is difficult to perform any rigorous analysis till the plant is stabilized. Now that we have designed stabilizing controllers, Principal Component Analysis and Statistical techniques should be used to justify the selection of the controlled variables.

c) We assumed that there are 11 manipulated and 11 controlled variables. Theoretical analysis to prove that there are 11 degrees of freedom is necessary.

2. *Identification*

a) Step response models were used in this study. Using these model, rigorous and better identification tests should be conducted to evaluate the predictions obtained when all manipulated variables are moved simultaneously.

3. *Control*

a) Constrained DMC controller should be installed to incorporate constraints on controlled variables and on rate of manipulated variable changes.

b) The multivariable controllers were not evaluated for the recommended disturbance changes. To achieve good disturbance rejection, disturbance models should be identified for feedforward controller design.

4. *Off-line Optimization*

a) Solving the non-linear optimization to determined the new operating conditions for other modes of operation is difficult. We recommend the reformulation of the optimization so that the gradients are steeper and the convergence to the solution is faster.

b) Setpoints to move the process to new operating conditions should be computed using a systematic approach (e.g., the LP optimization described earlier).

Implementing plant-wide control schemes is an art as well as a science. In addition to the mathematical analysis of control systems, many heuristics based on experience and process knowledge are used to developing a plant-wide control scheme. The T-E problem served as a good test bed to investigate the proposed plant-wide control system design scheme. There are a number of aspects of the problem that were not addressed in this chapter. A lot more work remains to be done.