PART

3

Industrial Examples

CHAPTER

8

Eastman Process

8.1 Introduction

Now that we have laid the groundwork by looking at the control of individual unit operations, we are ready to return to the plantwide control problem. In the next four chapters we illustrate the application of the nine-step design procedure with four industrial process examples.

We begin with a fairly simple process consisting of a reactor, condenser, separator, compressor, and stripper with a gas recycle stream (Fig. 8.1). This process was developed and published by Downs and Vogel (1993) as an industrial plantwide control test problem. A FORTRAN program is available from them that does the derivative evaluations for the process. The user must write a main program that initializes the simulation, does the controller calculations, performs the numerical integration, and plots the results.

A detailed description of the process in this book is unnecessary since one was provided in the original paper. We summarize here only some of the essential and unusual dynamic features. A small amount of an inert noncondensible component B is introduced in a feed stream and must be purged from the process. There are four fresh gas feed streams: F_{oA}, F_{oD}, F_{oE}, and F_{oC}. The first three are mixed with the recycle gas and fed into the bottom of the reactor. The last fresh feed F_{oC} is fed into the bottom of the stripper.

There are two main reactions, both of which are irreversible and exothermic:

$$A + C + D \rightarrow G \tag{8.1}$$

$$A + C + E \rightarrow H \tag{8.2}$$

Two additional irreversible and exothermic side reactions produce by-

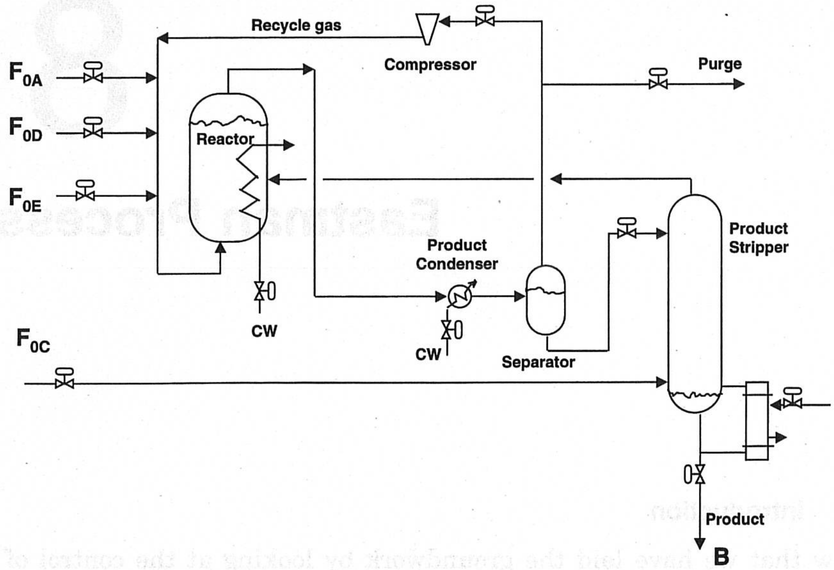


Figure 8.1 Eastman process flowsheet and nomenclature.

product F. The reactions are approximately first-order with respect to reactant concentrations. The reactor, which is open-loop unstable, contains both liquid and vapor phases, but no liquid stream leaves the reactor. Vapor from the reactor flows through a partial condenser and into a separator drum. Liquid from the drum is fed to the top tray of a stripping column. Vapor from the drum is compressed, a small portion is purged, and the remainder is recycled back to the reactor. The stripper has two sources of vapor: a small reboiler and the F_{oC} fresh feed.

Both gas pressure and liquid level in the reactor are integrating phenomena, and the choice of manipulated variables to control them is somewhat clouded. Temperature, pressure, and liquid level in the reactor all interact and their behavior is nonlinear. The gas purge stream from the process is very small, so its effectiveness in controlling pressure is doubtful.

The four fresh reactant feed streams must be managed in an appropriate way to satisfy overall component balances. Fortunately, composition analyzers are available. Figure 8.2 gives a sketch of the process with nomenclature and the values of flowrates, compositions, temperatures, and pressures at the initial steady state (Mode 1).

Several different control structures have been published in the literature for the Eastman process: Ricker (1993), McAvoy and Ye (1994), Price et al. (1994), Lyman and Georgakis (1995), Ricker and Lee (1995), Banerjee and Arkun (1995), Kanadibhotla and Riggs (1995), McAvoy

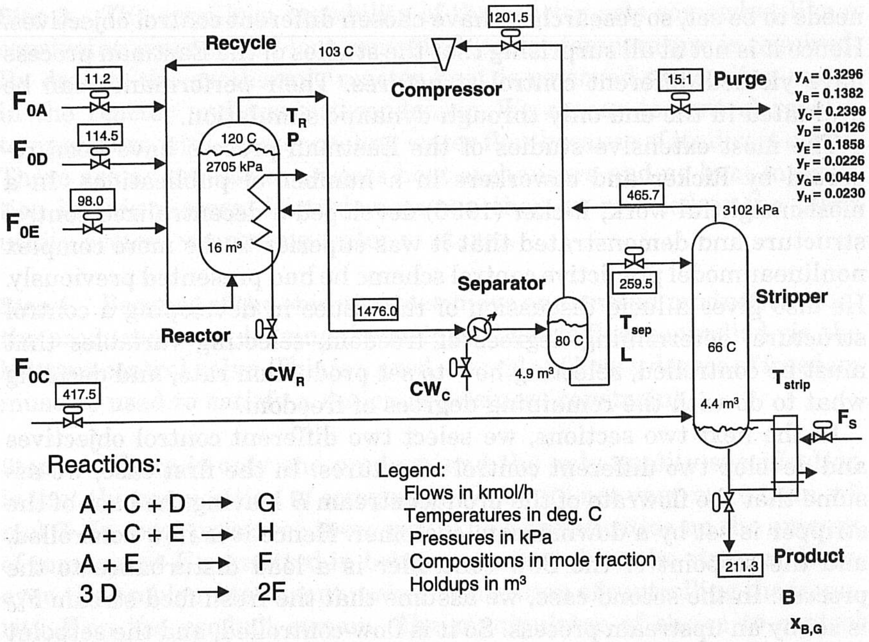


Figure 8.2 Eastman process base steady-state conditions for Mode 1.

et al. (1996), and Ricker (1996). Several common loops appear in nearly all of these control strategies. Reactor temperature is typically controlled with cooling water to the reactor, but in some structures the setpoint of this reactor temperature controller is changed via a cascade configuration. Most of the published structures control liquid levels in the separator and stripper by manipulating the liquid streams leaving those vessels. The fresh feed F_{oA} is generally used to control the composition of component A in the system and the purge to control component B composition.

However, many differences arise among the schemes when looking at how production rate is set and how liquid level and pressure in the reactor are controlled. For example, production rate is set in various strategies via fresh feed F_{oD} flow, condenser cooling water flow, separator liquid flow, stripper base flow, or fresh feed F_{oC} flow. Reactor level is controlled by fresh feeds F_{oD} and F_{oE}, separator temperature setpoint, compressor recycle valve, or fresh feed F_{oC} flow. Reactor pressure is controlled by reactor cooling water flow, purge flow, or F_{oA} feed flow. In one strategy reactor pressure is uncontrolled and allowed to float.

Throughout this book we have repeatedly stressed that the plantwide control problem is open-ended, which means there is no unique correct solution. For this process it is also not clear where production rate needs to be set, so researchers have chosen different control objectives. Hence it is not at all surprising that the studies of the Eastman process have yielded different control structures. Their performance can be evaluated in the end only through dynamic simulation.

The most extensive studies of the Eastman process have been reported by Ricker and coworkers in a number of publications. In a most insightful work, Ricker (1996) developed a decentralized control structure and demonstrated that it was superior to the more complex nonlinear model predictive control scheme he had presented previously. He also gives a lucid discussion of the issues in developing a control structure: determining degrees of freedom, selecting variables that must be controlled, selecting how to set production rate, and deciding what to do with the remaining degrees of freedom.

In the next two sections, we select two different control objectives and develop two different control structures. In the first case, we assume that the flowrate of the product stream B leaving the base of the stripper is set by a downstream customer. Hence it is flow-controlled, and the setpoint of the flow controller is a load disturbance to the process. In the second case, we assume that the fresh feed stream F_{oC} is set by an upstream process. So it is flow-controlled, and the setpoint of the flow controller is a load disturbance to the process. These two cases demonstrate that different control objectives produce different control strategies.

8.2 Case 1: On-Demand Product

8.2.1 Regulatory control strategy

Step 1. We are assuming in this section that the product stream from the bottom of the stripper is set on the demand of a downstream user. The bottoms stream from the stripper is flow-controlled and so we set the position of the control valve, XMV(8), on this stream (B). The rest of the liquid level controls must be chosen to accommodate this first-priority choice. Note that we could put a flow controller on this stream if necessary, but this was not done in the simulations described later. The quality specification is that component G in the product should not vary more than ± 5 mol %.

Step 2. This process has 12 degrees of freedom. One of these is agitation rate, which we simply hold constant. This leaves 11 degrees of freedom: four fresh feeds F_{oA}, F_{oD}, F_{oE}, and F_{oC}; purge valve; gas recycle valve; separator base valve; stripper base valve; steam valve; reactor cooling water valve; and condenser cooling water valve.

Step 3. The open-loop instability of the reactor acts somewhat like a constraint, since closed-loop control of reactor temperature is required. By design, the exothermic reactor heat is removed via cooling water in the reactor and product condenser. We choose to control reactor temperature with reactor cooling water flow because of its direct effect. There are no process-to-process heat exchangers and no heat integration in this process. Disturbances can then be rejected to the plant utility system via cooling water or steam.

Step 4. Because of the objective to achieve on-demand production rate, the product stream leaving the stripper base is flow-controlled via the bottoms control valve. This is a good example of how a degree of freedom must be used to satisfy a design or business constraint.

Step 5. There is only one product, and the only quality specification is that the composition of component G should not vary more than ± 5 mol %. In most processes there would be a specification on the amount of component E permitted in bottoms product from the stripper. However, the problem statement makes no mention of controlling the impurity E in the product stream. The manipulator of choice to control product quality ($x_{B,E}$) is stripper steam flow (F_S) because of its fast response. Stripper temperature can be used to infer product composition. It does a good job in keeping most of the light components from being lost in the product. There are only small changes in $x_{B,E}$ for the disturbances specified by Downs and Vogel (1993) (± 0.5 percent). Another manipulated variable that directly affects stripper bottoms purity is the flowrate of feed F_{oC}. However, this fresh feed makeup stream affects the component balances of A and C in the system, while steam does not, and we would have recognized this at Step 7. Therefore, we choose reboiler steam to control product purity.

High reactor temperature (175°C) is one safety constraint. Reactor cooling water flow has previously been selected to control reactor temperature. The only other known safety constraint for this process is pressure, which must not exceed the shut-down limit of 3000 kPa. The gas fresh feed streams, cooling water streams, reboiler steam flow, and purge directly affect pressure. Any of these could be used to control pressure. Of course the reaction rate also affects pressure, so a variable that changes the reaction rate could potentially be used to control pressure indirectly. Reactor cooling water flow and steam have already been selected. Condenser cooling rate is smaller than reactor cooling rate, so it may not be very effective in controlling pressure. This leaves one of the gas flows.

The purge stream is only 15.1 kmol/h, while the largest fresh feed makeup stream, F_{oC}, is 417 kmol/h. The vapor holdup in the reactor, separator, and stripper is estimated to be about 15 m^3. This gives a time constant of about 2 minutes if F_{oC} is used to control pressure. If the purge flow is used, the time constant is about 60 minutes. Because fresh feed F_{oC} is the largest of the gas flows by far, we choose it to control pressure.

Step 6. Three liquid levels need to be controlled: reactor, separator, and stripper base. We must use the Buckley strategy of *level control in the reverse direction to flow* since the stripper base product B is fixed by production rate. Therefore liquid flow from the separator (L) must be used to control stripper base level. To control level in the separator, we select the cooling water flow to the condenser (CW_C).

Now we must decide how to control the liquid level in the reactor. This liquid consists of mostly the heavy products, components G and H. The more fresh reactant components D and E are fed into the process, the more products will be produced. So we select the two fresh feed flowrates F_{oD} and F_{oE} to control reactor liquid level. We ratio one to the other depending upon the desired split between components G and H in the final product. Simple flow ratios should be accurate enough to maintain the desired product distribution without any feedback of product compositions. So on-line analyzers on the product streams should not be required.

Step 7. A light inert component B enters in one of the feed streams. It can be removed from the process only via the purge stream, so purge flowrate is used to control the composition y_B in the purge gas stream. Stripper temperature control keeps the volatile gas reactants within the gas recycle loop. Components D and E are accounted for via reactor level. The component balance for C is maintained via pressure, assuming we can control the composition of the other major component A in the gas loop. There must be some feedback mechanism to guarantee that precisely the correct number of moles of this component are fed into the system to react with the number of moles of component C. The only manipulator available to satisfy the component balance for A is the fresh feed stream F_{oA}. So we select this flow to control the composition of A in the purge gas stream y_A. The compositions of either the purge gas or the reactor feed could be used, but both are not necessary.

Only two analyzers are required to run this process: one measuring the amount of A in the system and one measuring the amount of B. The latter could be eliminated if the amount of inert B coming into the system does not vary drastically. The purge stream is very small, and variations in the concentration of B in the system should have only a minor effect on controllability. However, Ricker (1996) has shown that the purge does have a significant economic impact.

Step 8. Control of the individual unit operations has been established and all of the control valves have been assigned except for the gas recycle valve.

Step 9. Of the original 11 degrees of freedom, we have used one for production rate, one for reactor temperature control, one for product quality, one for pressure control, three for liquid levels, and two for compositions. An additional one was used to set the G/H ratio. This leaves one degree of freedom to be specified. This is the valve that controls the flowrate of the gas recycle. We fix this valve to be wide open, based on the Douglas heuristic (Fisher et al., 1988) that gas recycle flows should be maximized to improve yields.

We have used the reactor cooling water valve to stabilize the system by controlling reactor temperature. However there is no specific temperature at which the reactor must operate. The best way to manage the reactor temperature setpoint is not immediately obvious. It might be used in conjunction with the production rate controller, i.e., higher temperatures may be needed to increase throughputs. It might be adjusted to maximize yields and suppress undesired by-products.

However, after making some simulation runs with several of the disturbances suggested in the original paper, it became apparent that the temperature in the separator was changing quite substantially and adversely affecting the stripper. Low separator temperature drops too many light components into the stripper, and the reboiler steam has trouble maintaining product quality. Therefore a separator temperature controller was added, whose output signal is the setpoint of the reactor temperature controller. The final basic regulatory control structure (Fig. 8.3) is simple, effective, and easily understood by operating personnel.

8.2.2 Override controls

The basic regulatory control structure outlined above was able to hold the process at the desired operating point for most of the disturbances. However, when manipulated variables hit constraints it was unable to prevent a unit shutdown. The disturbance IDV(6) that shuts off the fresh feed flowrate F_{oA} is probably the most drastic. The resulting imbalance in the stoichiometric amounts of components A and C drives the concentration y_A down quite rapidly. The reaction rate slows up, reactor temperature drops, and the process shuts down on high pressure. Since one degree of freedom has been removed by this disturbance, the control structure must be modified with overrides to handle the component balances.

The F_{oC} stream contains more C than A, so the excess C must be

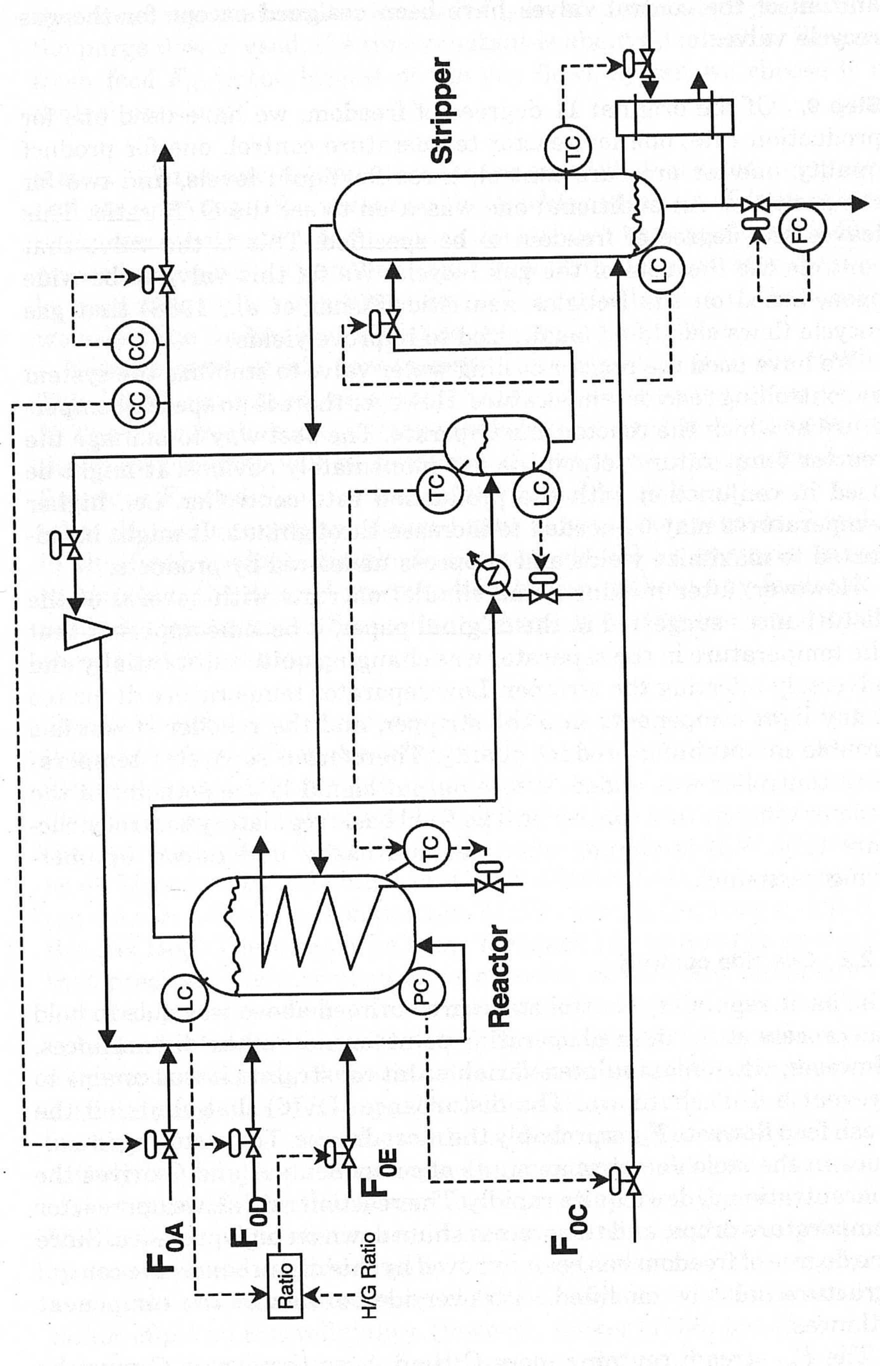


Figure 8.3 Control structure for on-demand product flow.

removed from the system. The only place available is the purge stream. Therefore, a low F_{oA} flow override controller is used to open the purge valve (Fig. 8.4). The other action that must be taken is to prevent the concentration of component A in the system from dropping down too low and reducing reaction rates. This is achieved by using a low y_A concentration override controller to pinch the fresh feed flowrate F_{oD} to slow up the rate of consumption of A. Of course F_{oE} is also reduced through the ratio.

Now the liquid level loops must also be modified, since we no longer can specify production rate and reactor level control cannot use F_{oD}. This is easily accomplished by using low level override controllers on each of the three levels. Low stripper level pinches product base product flowrate B. Low separator level pinches separator liquid flowrate L. Low reactor level pinches the condenser cooling water flowrate CW_C. In an override situation the level control structure has been reversed from the basic structure and now levels are held *in the direction of flow.*

8.2.3 Simulation results

Dynamic simulations were run with the proposed control structure for all disturbances proposed by Downs and Vogel (1993). Only the Mode 1 operation was studied. Section 8.5 gives the FORTRAN program used. Figures 8.5 to 8.8 show results for several disturbances.

Figure 8.5 shows how large changes in production rate are handled. Product stream B *immediately* changes, and the rest of the process flows adjust appropriately. At time equals 1 hour, the valve position on the production rate (B) is dropped 50 percent. At time equals 10 hours, it is increased back to its base case value. The process follows these changes in B by gradually and smoothly reducing fresh feeds through the level controllers.

Figure 8.6 gives results for changing the G/H split in the product. The two fresh feed streams F_{oD} and F_{oE} are immediately changed to the appropriate new values. The reactor level controller output signal is sent to a flow controller on F_{oD}, and the bias value on this flow controller is changed to the desired value. At the same time the ratio between the two flowrates is set to the new desired number. Product composition $x_{B,G}$ and $x_{B,H}$ change to their new values in about 4 hours.

Figure 8.7 shows how the process rides through the loss of F_{oA}, disturbance IMV(6). The override controller takes action almost immediately, and production rate is reduced after about 5 hours when liquid levels drop.

Figure 8.8 shows the responses to IMV(1), a change in the composition of A and C in the F_{oC} stream. As the amount of A in the system drops, the override controller cuts feed streams F_{oD} and F_{oE}. When reactor

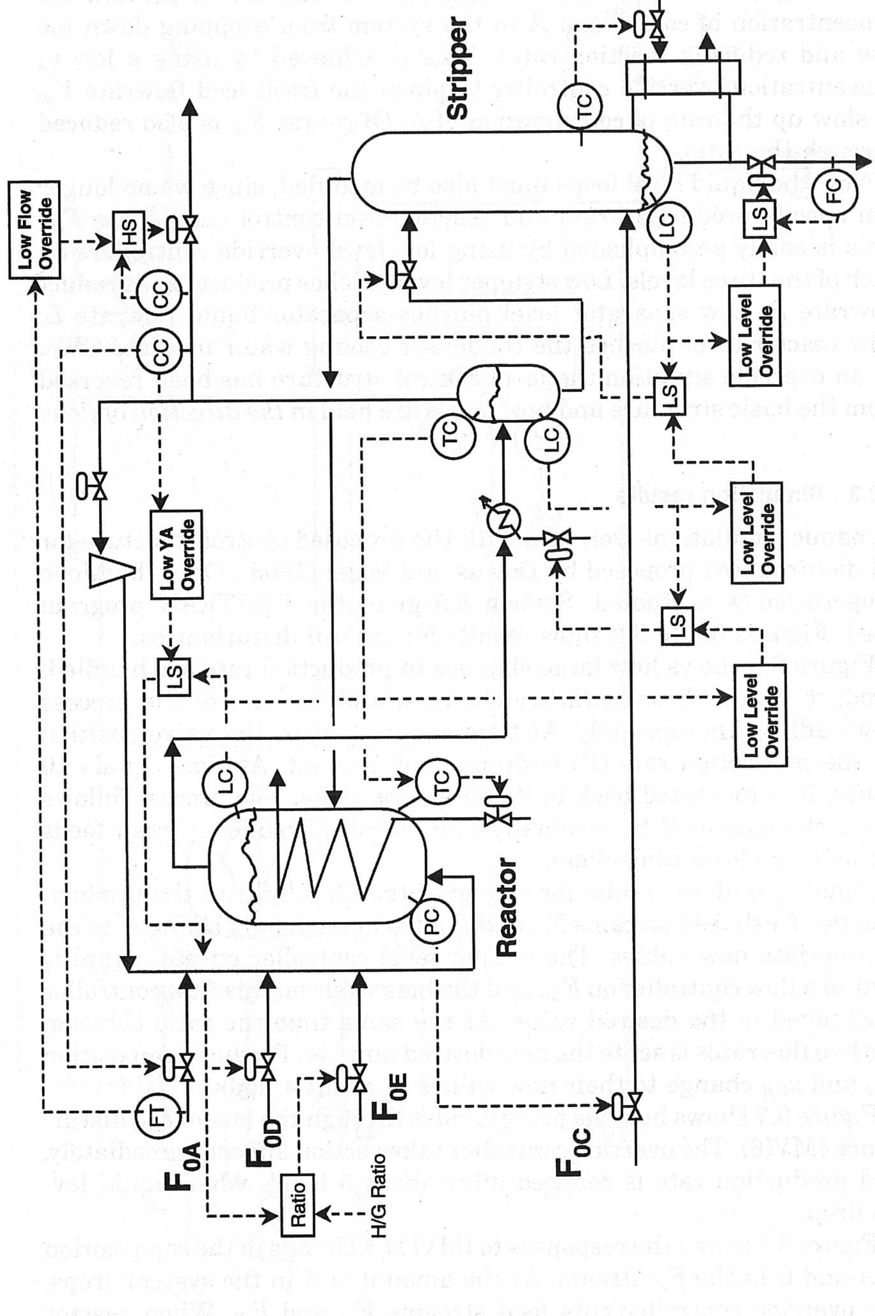


Figure 8.4 Control structure for on-demand product flow with overrides.

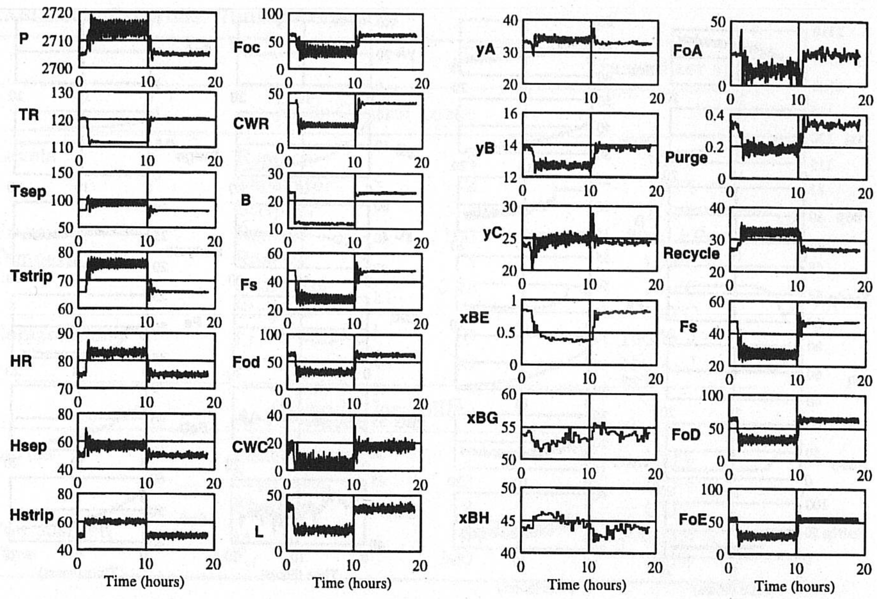


Figure 8.5 Dynamic response for 50 percent change in product B flow.

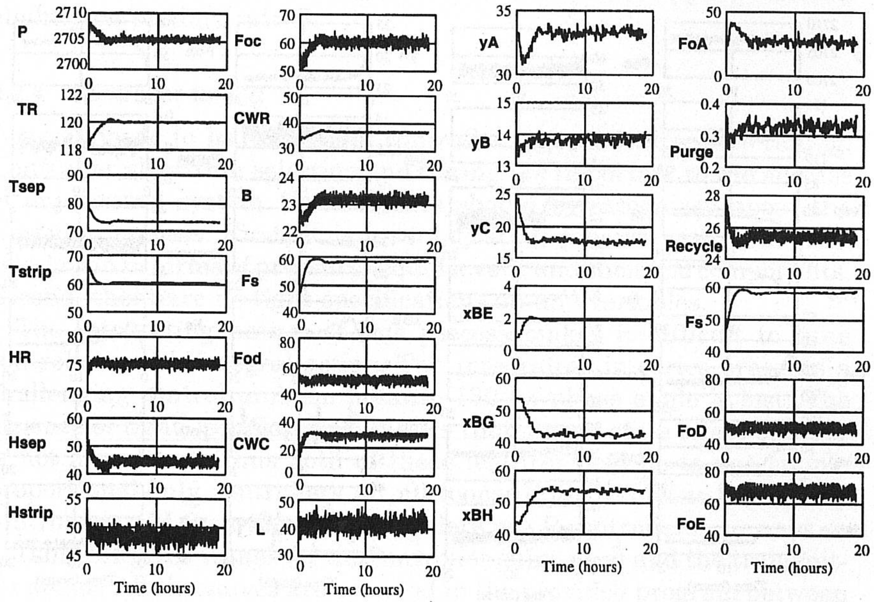


Figure 8.6 Dynamic response for change in G/H split.

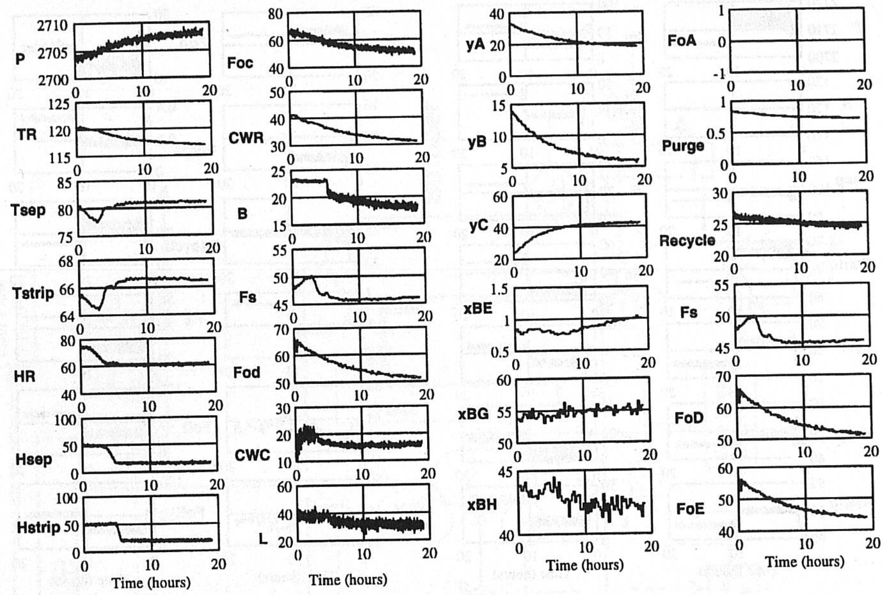


Figure 8.7 Dynamic response for loss of fresh feed F_{oA}.

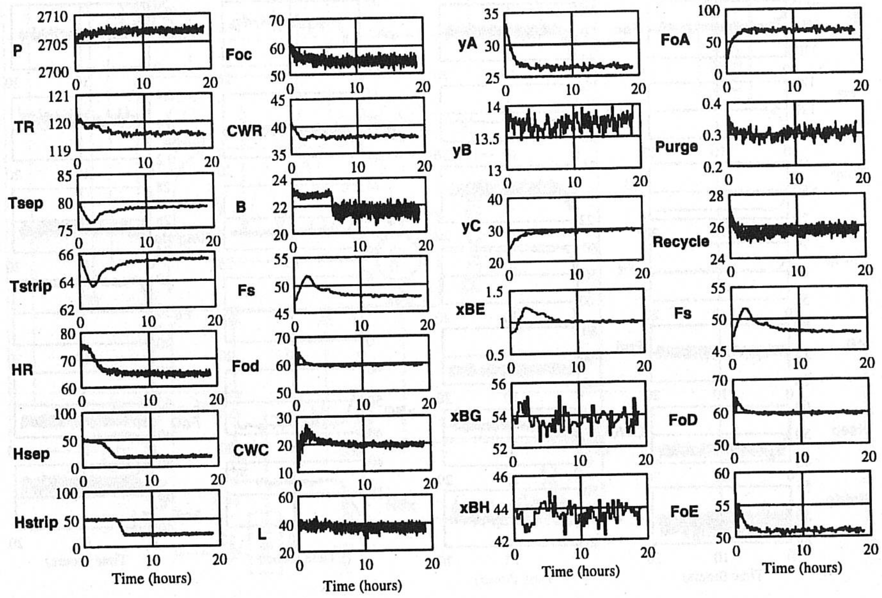
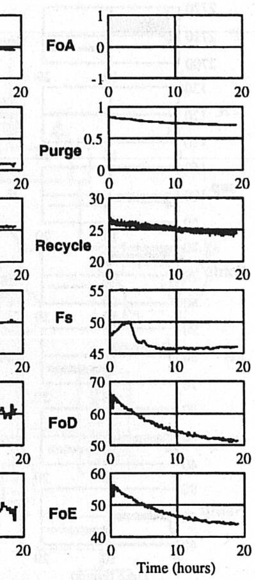


Figure 8.8 Dynamic response for change in F_{oC} feed composition.



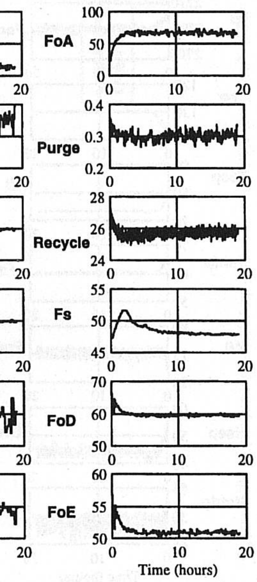


TABLE 8.1 Controller Tuning Constants

		K_c	Transmitter span
	Basic Control Loops		
Levels	Reactor	4	100%
	Separator	2	100%
	Stripper	2	100%
Pressure	Reactor	100	3000 kPa
Temperatures	Reactor	3	100°C
	Separator	0.15	100°C
	Stripper	2	100°C
Compositions	y_B	16	100 mol %
	y_A	10	100 mol %
	Override Controllers		
Levels	Reactor	1	100%
	Separator	2	100%
	Stripper	2	100%
Composition	y_A	1	100 mol %
Flow	F_{oA}	100	100%

holdup drops, the override controller cuts condenser cooling. When separator level drops the override controller cuts separator liquid L. Finally, after about 25 hours, the low level in the stripper cuts back slightly on product rate B.

8.2.4 Controller tuning

A word needs to be said about controller type and controller tuning. Controller algorithm selection and tuning are important to the success of any control system. Two features should be recognized about the Eastman process. First, it is an integrating process with little self-regulation in terms of pressure, liquid levels, and chemical components. Second, there are no tight specifications on any variables.

The integrating nature of this process makes it difficult to tune controllers with integral action. Two integrators in series presents a challenging control problem because 180° of phase angle is lost. The absence of tight specifications implies that steady-state offset or error is not a problem. Thus both of these features lead us to use simple proportional-only controllers on all loops. Both the basic regulatory controllers and the overrride controllers are P-controllers.

Table 8.1 gives values for the controller gains used and the transmitter spans. All the valves are spanned in the provided program between 0 and 100 percent. The level controller gains ranged from 1 to 4 and required little tuning. The loops that required a little empirical tuning were the three temperatures, the pressure, and the two compositions.

Tuning was performed by increasing the controller gain and testing the dynamic response to a step change in setpoint until the loop became too oscillatory. Reactor temperature was tuned first, followed by pressure, separator temperature, stripper temperature, component A composition, and component B composition. No claim is made that these are the best settings, but they give adequate control and required little time to tune.

A plantwide control design procedure was used to develop a simple but effective regulatory control system for the Eastman process with an on-demand product control objective. With this strategy, control of production rate is essentially instantaneous. Drastic upsets and disturbances are handled by simple proportional-only overrides.

8.3 Case 2: On-Supply Reactant

8.3.1 Regulatory control strategy

Step 1. In this section we alter the control objective relating to production rate. Instead of flow controlling the product stream from the bottom of the stripper, we assume that an upstream process sets the flow of the F_{oC} stream and the process must take whatever amount is fed into it. Most of the steps in the design procedure are the same as the previous section, but the control of liquid levels is now in the direction of flow. The product quality criterion is the same.

Step 2. The same number of degrees of freedom exist.

Step 3. This is constructed as above.

Step 4. Production rate is set by fresh feed F_{oC}.

Step 5. Reboiler steam controls product purity. Now the fresh feed F_{oC} cannot be used to control pressure. The purge stream is so small that effective pressure control is unlikely. Reactor cooling water flow is used to control reactor temperature. Therefore, the logical choice for pressure control is the cooling water flow to the condenser CW_C. The controller gain of this loop was empirically set at 25 (with a pressure transmitter span of 3000 kPa).

Step 6. We use the Buckley strategy of *level control in the direction of flow* to regulate two liquid levels. The stripper base level is controlled by manipulating stripper bottoms flow. The separator level is controlled by manipulating the liquid flowrate from the separator to the stripper. As before, the liquid level in the reactor is controlled by the two fresh feed flowrates F_{oD} and F_{oE}. We ratio one to the other depending upon the desired split between components G and H in the final product.

Step 7. Purge flow controls the composition y_B in the recycle gas stream. The fresh feed F_{oA} controls the composition y_A in the recycle gas stream.

Step 8. All control valves have been assigned.

Step 9. Separator temperature is controlled by changing the setpoint of the reactor temperature controller. The controller gain of the separator temperature controller was empirically set at 0.5 (with a temperature transmitter span of 100°C).

8.3.2 Control scheme and simulation results

The control system is shown in Fig. 8.9. The override controller on low F_{oA} flow was used to open the purge valve. The override controller on low y_A composition was used to pinch the fresh feed flowrate F_{oD}.

Typical simulation results are shown in Fig. 8.10. At time equals zero, the fresh feed flowrate F_{oC} is reduced by 25 percent from its base-case value. The process responds to this change by gradually cutting back on the other feed streams and the product leaving the unit. The new steady-state conditions are attained in a little over 1 hour. The control structure also successfully handled the other disturbances.

8.4 Conclusion

Developing a plantwide control system for the Eastman process is fairly straightforward. There are only five unit operations and one gas recycle stream. No energy integration is present. So the major feature of this process from a plantwide viewpoint is the problem of accounting for the multiple component inventories.

We have shown how different control objectives lead to different control structures. Although the two strategies handle disturbances differently, they both work without showing a clear advantage of one compared with the other.

In the next three chapters, we consider more challenging processes with many more unit operations and multiple recycle streams.

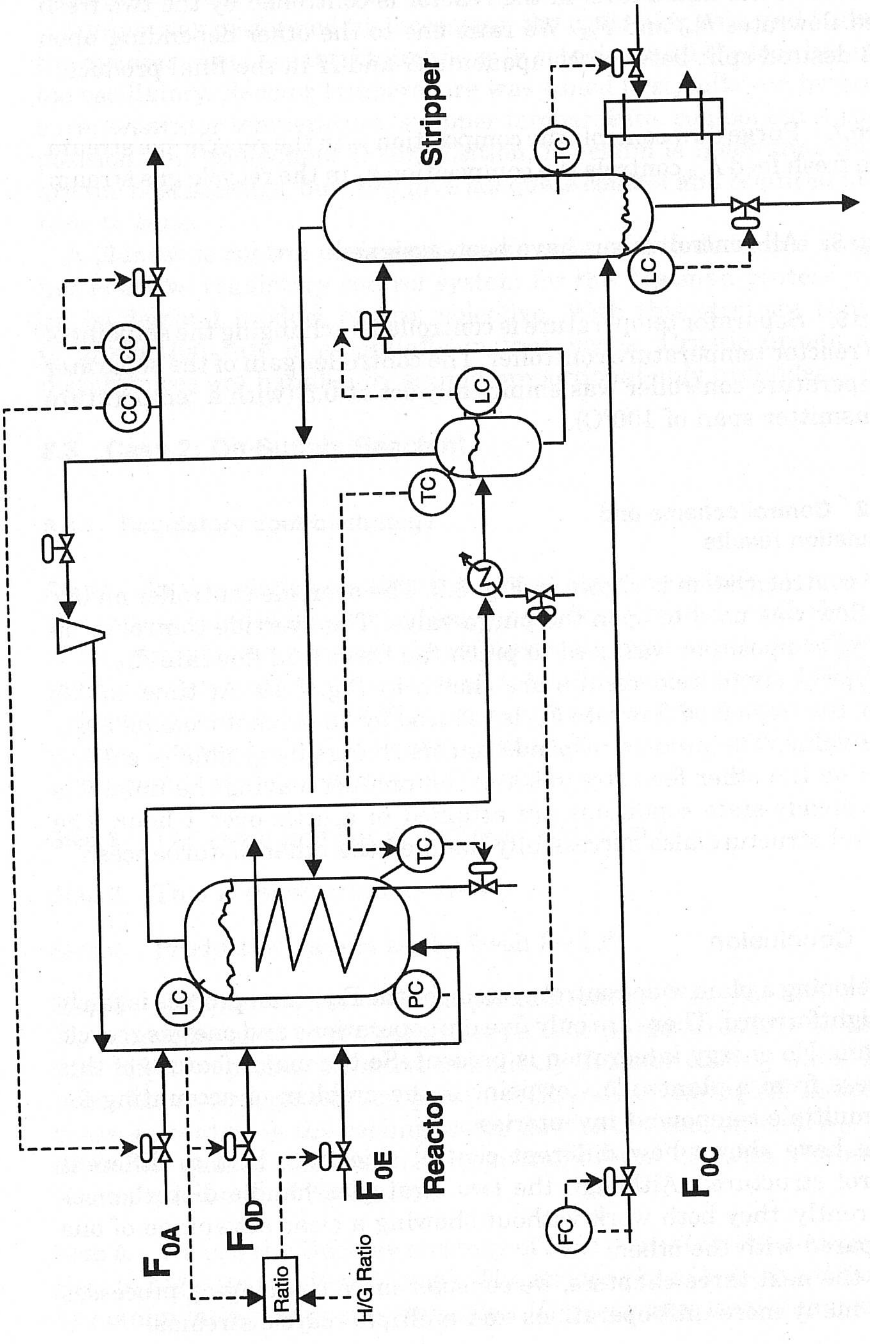


Figure 8.9 Control structure for fixed fresh feed F_{0C}.

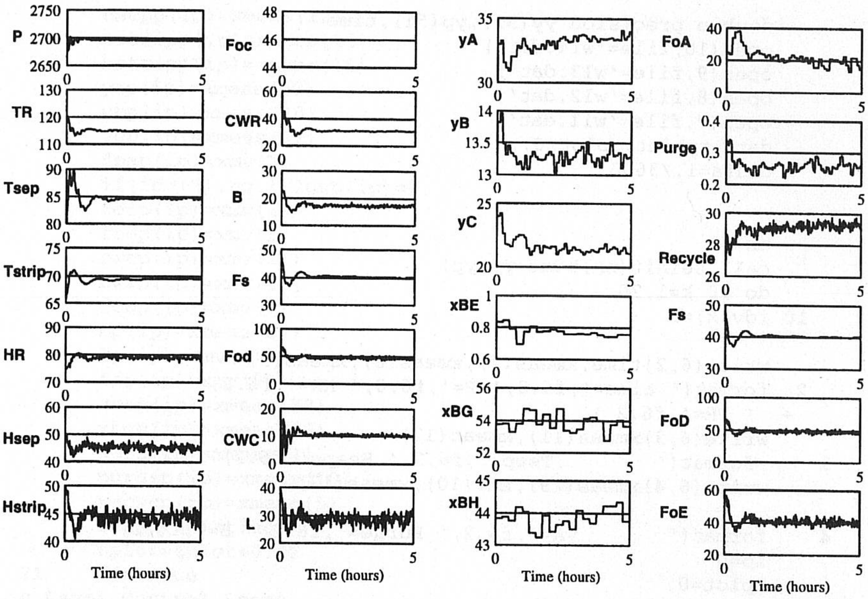


Figure 8.10 Dynamic response for 25 percent reduction in F_{oC} feed flow.

8.5 FORTRAN Program for Eastman Process

On-Demand Product

```
c
c Control structure 1: "eastcs1.for"
c Production rate (B) is flow controlled
c TC of Tsep added, changing TR
c Overrides added on pressure to purge
c                    stripper level to B
c                    separator level to L
c                    yA to FoD
c                    reactor level to CWC
      DOUBLE PRECISION XMEAS,XMV,SETPT
      COMMON/PV/XMEAS(41),XMV(12),SETPT(21)
      INTEGER IDV
      real kc1,kc2,kc3,kc4,kc5,kc6,kc7,kc8,kc9,kc10
      real lp(5000)
      COMMON/DVEC/IDV(20)
      double precision gain(21),reset(21)
      dimension trp(5000),prp(5000),hrp(5000),timep(5000)
      dimension tsepp(5000),hsepp(5000),tstripp(5000)
      dimension hstripp(5000),yap(5000),ybp(5000),ycp(5000)
      dimension fodp(5000),cwcp(5000),focp(5000)
      dimension cwrp(5000),foap(5000),purgep(5000)
      dimension foep(5000),bp(5000),fsp(5000)
      dimension xbep(5000), xbgp(5000), xbhp(5000)
      dimension recycp(5000),tlagp(5000)
```

```
      double precision yy(51),yp(51),time
      open(10,file='wl4.dat')
      open(9,file='wl3.dat')
      open(8,file='wl2.dat')
      open(7,file='wl1.dat')
      data tprint,tplot/0.,0./
      delta=1./3600.


      nn=51
      call teinit(nn,time,yy,yp)
      do 10 k=1,20
   10 idv(k)=0

      write(6,2)time,xmeas(7),xmeas(8),xmeas(9)
    2 format(' time=',f6.3,' P=',f6.0,' HR=',f6.2,
     + ' TR=',f6.2 )
      write(6,3)xmeas(11),xmeas(12)
 3     format('        Tsep=',f6.2,' Hsep=',f6.2)
      write(6,4)xmeas(29),xmv(10),xmeas(17)

 4    format('        yA=',f6.2,' Purge=',f6.2,' B=',f6.2)
      ip=0
      tplot=0.
      tstop=19.
c Specify disturbance number

      idv(1)=1
c***************************************************
c Disturbance is 15% reduction in production rate
c     xmv(8)=xmv(8)*0.85
c     xmv(8)=xmv(8)*1.2
c***************************************************
      xmv8o=xmv(8)
      xmv8base=xmv8o
c  Base case is fodo=63.053
       ratio=53.98/63.053
       fodo=63.053
c Disturbance is switch from 50/50 G/H to 1/3 G/H
c      fodo=50.434
c      ratio=68.38/50.434
c*********************************************
 100  continue
c Put in series of rate changes
c      if(time.gt.1.)xmv8o=xmv8base*0.5
c      if(time.gt.10.)xmv8o=xmv8base
      if(time.lt.tprint)go to 20
      write(6,2)time,xmeas(7),xmeas(8),xmeas(9)
      write(6,3)xmeas(11),xmeas(12)
      write(6,4)xmeas(29),xmeas(10),xmeas(17)
      tprint=tprint+.1
20    if(time.lt.tplot)go to 21
      ip=qip+1
      timep(ip)=time
      trp(ip)=xmeas(9)
      prp(ip)=xmeas(7)
      hrp(ip)=xmeas(8)
      tsepp(ip)=xmeas(11)
```

```
      hsepp(ip)=xmeas(12)
      tstripp(ip)=xmeas(18)
      hstripp(ip)=xmeas(15)
      yap(ip)=xmeas(29)
      ybp(ip)=xmeas(30)
      ycp(ip)=xmeas(31)
      foap(ip)=xmv(3)
      if(idv(6).eq.1)foap(ip)=0.
      focp(ip)=xmv(4)
      fodp(ip)=xmv(1)
      cwcp(ip)=xmv(11)
      cwrp(ip)=xmv(10)
      foep(ip)=xmv(2)
      bp(ip)=xmeas(17)
      lp(ip)=xmv(7)
      fsp(ip)=xmv(9)
      xbep(ip)=xmeas(38)
      xbgp(ip)=xmeas(40)
      xbhp(ip)=xmeas(41)
      purgep(ip)=xmeas(10)
      recycp(ip)=xmeas(5)
      tlagp(ip)=tlag
      tplot=tplot+0.02
21    continue
c Level Control Loops
c LC 1: hr(8) controlled by fod(1)
      kc1=4.
      err1=75-xmeas(8)
      xmv(1)=fodo+kc1*err1
c*********************************************
c Low yA override pinches fod
      yakc=1.
      yaerr=30.-xmeas(29)
      yafod=fodo-yakc*yaerr
      if( yafod.lt.xmv(1))xmv(1)=yafod
c***********************************************
c Ratio foe to fod
      xmv(2)=xmv(1)*ratio
c LC 2: hsep(12) controlled by CWc(11)
      kc2=2.
      err2=50.-xmeas(12)
      xmv(11)=18.114+kc2*err2
c**********************************************
c Low reactor level pinches condenser CWc
      hrkc=1.
      hrerr=75.-xmeas(8)
      hrcwc=30.-hrkc*hrerr
      if( hrcwc.lt.xmv(11))xmv(11)=hrcwc
      if(xmv(11).lt.0.)xmv(11)=0.
      if(xmv(11).gt.1000.)xmv(11)=100.

c**********************************************
c*******************************************8
c Low stripper level pinches B
      xmv(8)=xmv8o
      bor=100.+2.*(xmeas(15)-50.)
      if( bor.lt.xmv(8))xmv(8)=bor
c******************************************
```

```
c LC 3: hstrip(15) controlled by L(7)
      kc3=2.
      err3=50.-xmeas(15)
      xmv(7)=38.1+kc3*err3
c*******************************************
c Low separator level pinches L
      xlor=100.+2.*(xmeas(12)-50.)
      if( xlor.lt.xmv(7))xmv(7)=xlor
c*******************************************
c PC 4: pr(7) controlled by foc(4)
c Ramp up pset
c      pset=2705.+pramp*time
c      if(pset.gt.pmax)pset=pmax
c Disturbance in pset: 2705 to 2645.
     pset=2705.
     kc4=100.
     err4=(pset-xmeas(7))/30.
     xmv(4)=61.302+kc4*err4
     if(xmv(4).lt. 0.)xmv(4)=0.
c Temperature Control Loops
c Tsep (11) controlled by trset
      tsepkc=0.15

      tsepset=80.109

      tseperr=tsepset-xmeas(11)
      trset=120.4+tsepkc*tseperr
c TC 5: TR (9) controlled by CWR (10)
      kc5=3.
      err5=(trset-xmeas(9))
      xmv(10)=41.106-kc5*err5

c TC 6: Tstrip(18) controlled by Fs(9)
      tstrpset=65.731
      kc6=2.
      err6=tstrpset-xmeas(18)
      xmv(9)=47.446+kc6*err6
      if(xmv(9).lt.0.)xmv(9)=0.
      if(xmv(9).gt.100.)xmv(9)=100.


c Composition Control Loops
c CC 7: yA (23) controlled by Foa (3)
      kc7=10.
      err7=32.188-xmeas(23)
      xmv(3)=24.644+kc7*err7
      if(xmv(3).gt.100.)xmv(3)=100.
      if(xmv(3).lt.0.)xmv(3)=0.
c CC 8: yB (24) controlled by purge (6)
      kc8=16.
      err8=13.823-xmeas(30)
      xmv(6)=40.064-kc8*err8
c***********************************************
c Loss of FoA opens purge valve

      if(idv(6).eq.1)xmv(6)=100.
c************************************************
c Integration
```

```
      call tefunc(nn,time,yy,yp)
      do 50 k=1,51
      yy(k)=yy(k)+yp(k)*delta
50      continue

      time=time+delta
      if(xmeas(7).gt.2950.)go to 89
      if(time.lt.tstop)go to 100
   89 do 90 k=1,ip
      write(7,91)timep(k),trp(k),prp(k),hrp(k),tlagp(k)
      write(8,91)tsepp(k),hsepp(k),tstripp(k),hstripp(k)
     + ,yap(k),ybp(k),ycp(k)
      write(9,91)foap(k),focp(k),fodp(k),cwcp(k),cwrp(k)
     + ,recycp(k)
      write(10,91)foep(k),bp(k),lp(k),fsp(k),xbep(k)
     + ,xbgp(k),xbhp(k),purgep(k)
91      format(8(1x,f12.5))
90      continue
      stop
      end
```

8.6 References

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NOMENCLATURE

B = flow rate of stripper bottoms

CW_C = flow rate of cooling water to condenser

CW_R = flow rate of cooling water to reactor

F_{oj} = flow rate of fresh feed, $j = A, C, D, E$

F_S = flow rate of steam to stripper

HR = reactor holdup, percent level

H_{sep} = separator holdup, percent level

H_{strip} = stripper holdup, percent level
L = flow rate of liquid from separator to stripper
P = pressure
Purge = purge gas flow rate
Recycle = flow rate of gas recycle to reactor
TR = reactor temperature
T_{sep} = separator temperature
T_{strip} = stripper temperature
x_{Bj} = composition of stripper bottoms, mole fraction component j
y_j = composition of purge gas, mole fraction component j