Chapter 23

23.1

Option (a):

- Production rate set via setpoint of w_A flow controller
- Level of R_1 controlled by manipulating w_C
- Ratio of w_B to w_A controlled by manipulating w_B
- Level of R_2 controlled by manipulating w_E
- Ratio of w_D to w_C controlled by adjusting w_D

Options (b)-(e) are developed similarly. See table below.

Option	Production Rate Set With	Control Loop #	Type of Controller	Controlled Variable	Manipulated Variable
a	w_A	1	Flow	$w_{A,m}$	w_A (V1)
	w_A	2	Ratio	$w_{B,m}$	w_B (V2)
	w_A	3	Level	H_{R1}	w_C (V3)
	w_A	4	Ratio	$w_{D,m}$	w_D (V4)
	w_A	5	Level	H_{R2}	w_E (V5)
b	w_B	1	Flow	$w_{B,m}$	w_B (V2)
	w_B	2	Ratio	$w_{A,m}$	w_A (V1)
	w_B	3	Level	H_{R1}	w_C (V3)
	w_B	4	Ratio	$w_{D,m}$	w_D (V4)
	w_B	5	Level	H_{R2}	w_E (V5)
c	w_C	1	Flow	$w_{C,m}$	w_C (V3)
	w_C	2	Ratio	$w_{B,m}$	w_B (V2)
	w_C	3	Level	H_{R1}	w_A (V1)
	w_C	4	Ratio	$w_{D,m}$	w_D (V4)
	w_C	5	Level	H_{R2}	w_E (V5)
d	w_D	1	Flow	$w_{D,m}$	w_D (V4)
	w_D	2	Ratio	$w_{C,m}$	w_C (V3)
	w_D	3	Level	H_{R1}	w_A (V1)
	w_D	4	Ratio	$w_{B,m}$	w_B (V2)
	w_D	5	Level	H_{R2}	w_E (V5)
e	w_E	1	Flow	$w_{E,m}$	w_E (V5)
	w_E	2	Ratio	$w_{D,m}$	w_D (V4)
	w_E	3	Level	H_{R2}	w_C (V3)
	w_E	4	Ratio	$w_{B,m}$	w_B (V2)
	w_E	5	Level	H_{R1}	w_A (V1)

- Subscript m denotes “measurement”.

In options c, d, and e, valves 1 and 2 can be used interchangeably. Thus, a total of 8 options can be developed.

Advantages and Disadvantages:

Each option is equivalent in the sense that 5 control loops are required: 1 flow, 2 level, and 2 ratio. Since there is no cost or complexity advantage with any option, the production rate should be set via the actual product rate, w_E , i.e. option e.

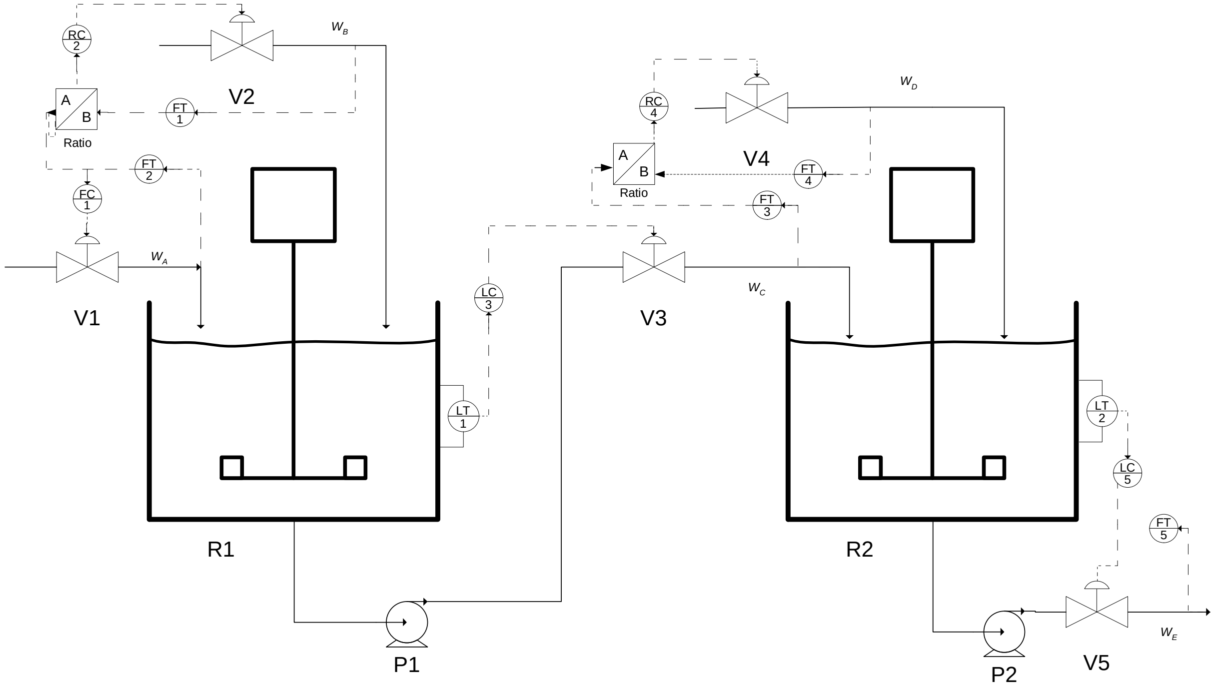

Figure S23.1. *Solution for option a)*

23.2

a) The level in the distillate (H_D) will be controlled by manipulating the recycle flow rate (D), and the level in the reboiler (H_B) via the bottoms flow rate (B).

Thus, H_D and H_B are pure integrator elements.

Closed-loop TF development assuming PI controller:

$$G_{CL}(s) = \frac{\tau_I s + 1}{(\frac{\tau_I}{K_c K_p})s^2 + \tau_I s + 1}$$

Relations:

$\frac{\tau_I}{K_c K_p} = \tau^2 \qquad \tau_I = 2\zeta\tau \qquad K_p = -1$ for both H_D and H_B loops.

Let $\zeta = 1$ for a critically damped response

Initial settings:

$$K_c = -0.4$$
$$\tau_I = 10$$

Final tuning: changed to proportional control only to obtain a faster response

$$K_c = -1$$

b) The distillate composition (x_D) will be controlled by manipulating the reflux flow rate (R), and the bottoms composition (x_B) via the vapor boilup (V).
Use a step response to determine an approximate first-order model (calculations are shown on last page).

$$\frac{x_D}{R} = \frac{0.0012}{2.33s + 1}$$
$$\frac{x_B}{V} = \frac{-0.000372}{2.08s + 1}$$

Using the Direct Synthesis method:

$$G(s) = \frac{K}{\tau s + 1}$$

$$G_c(s) = \frac{\tau}{\tau_c K}\left(1 + \frac{1}{\tau s}\right)$$

Choosing $\tau_c = \frac{1}{4}\tau$, the settings are:

x_D - R loop $\left\{ \begin{array}{l} K_c = 3333.3 \\ \tau_I = 2.33 \end{array} \right.$

x_B - V loop $\left\{ \begin{array}{l} K_c = -10649.6 \\ \tau_I = 2.08 \end{array} \right.$

c) The reactor level (H_R) will be controlled by manipulating the flow from the reactor (F).

H_R is a pure integrator element.

Using the same relations as in part a, initial controller settings are:

$$K_c = -0.4$$
$$\tau_I = 10$$

After tuning:

$$K_c = -1$$
$$\tau_I = 5$$

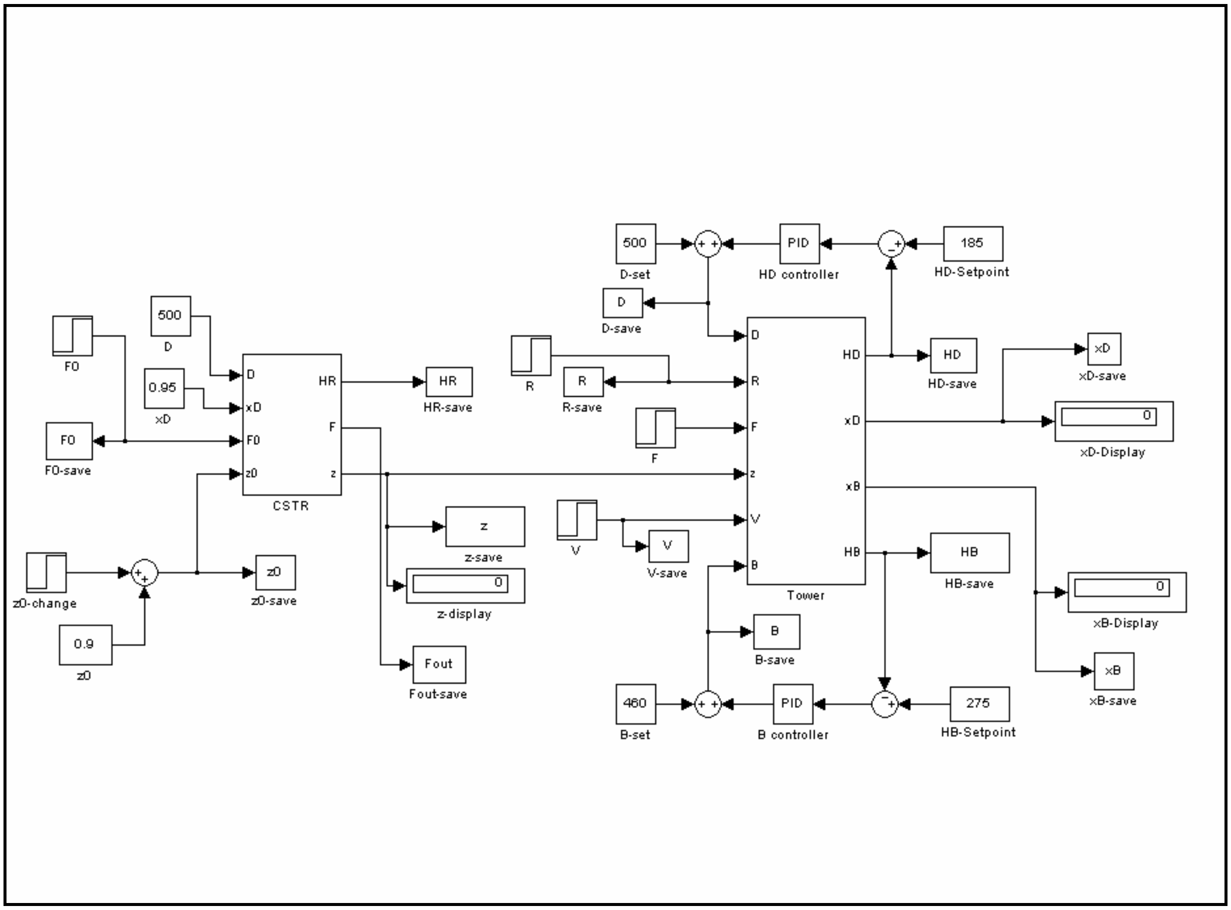

Figure S23.2a. *Simulink-MATLAB block diagram for case a)*

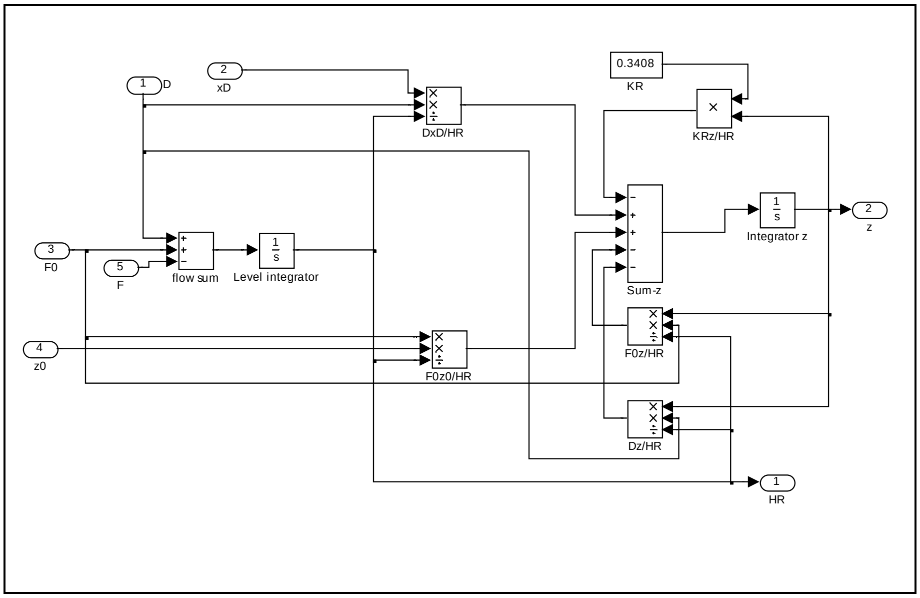

Figure S23.2b. *Simulink-MATLAB block diagram for the CSTR block*

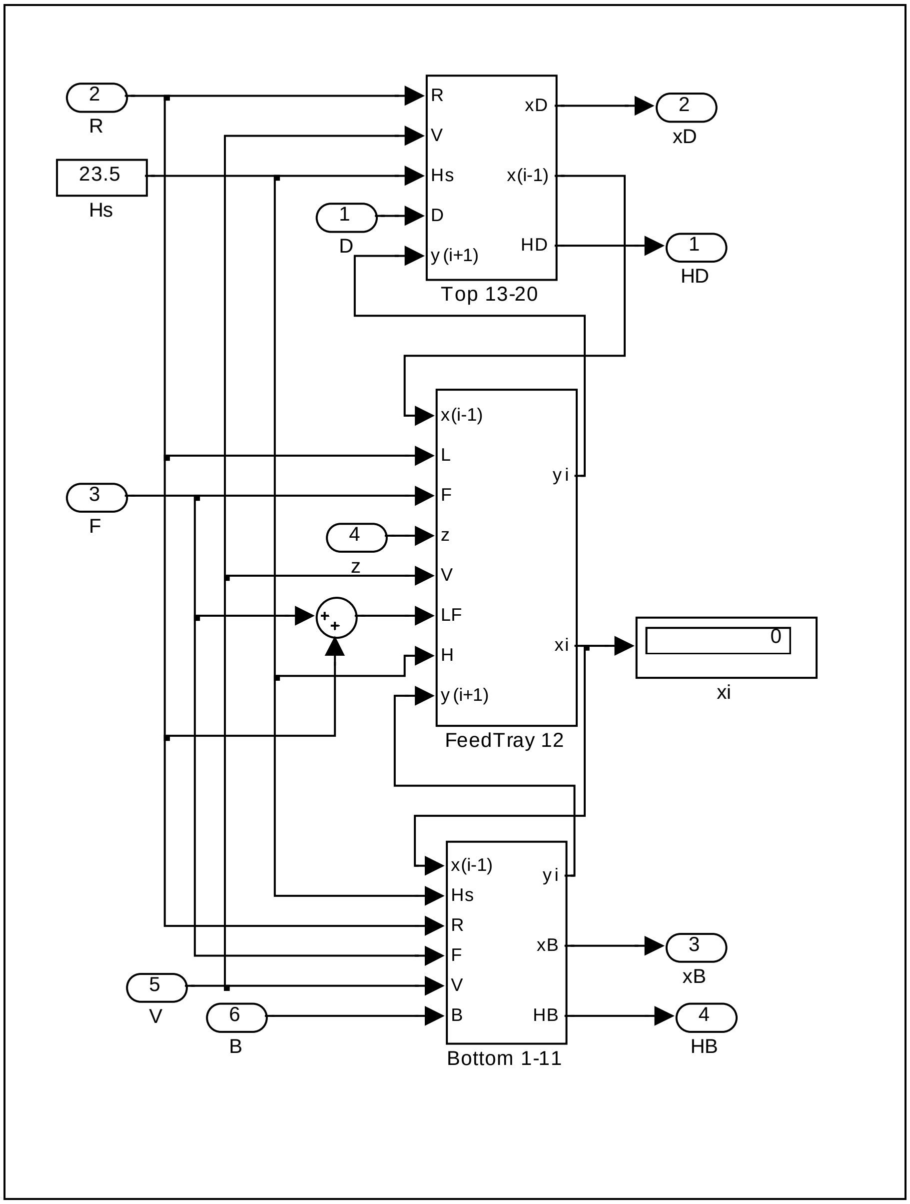

Figure S23.2c. *Simulink-MATLAB block diagram for the Tower block*

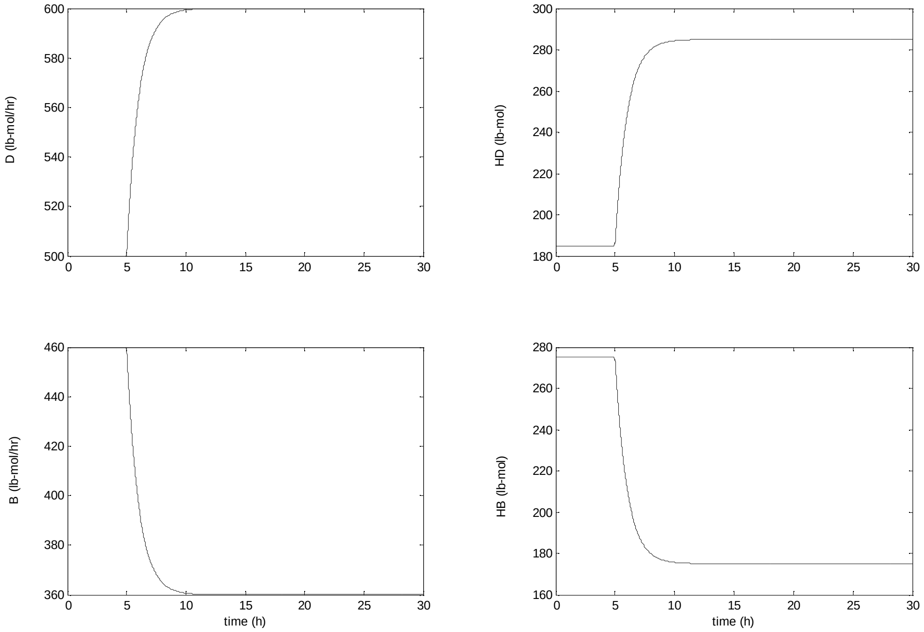

Figure S23.2d. *Step change in V (1600-1700) at t=5*

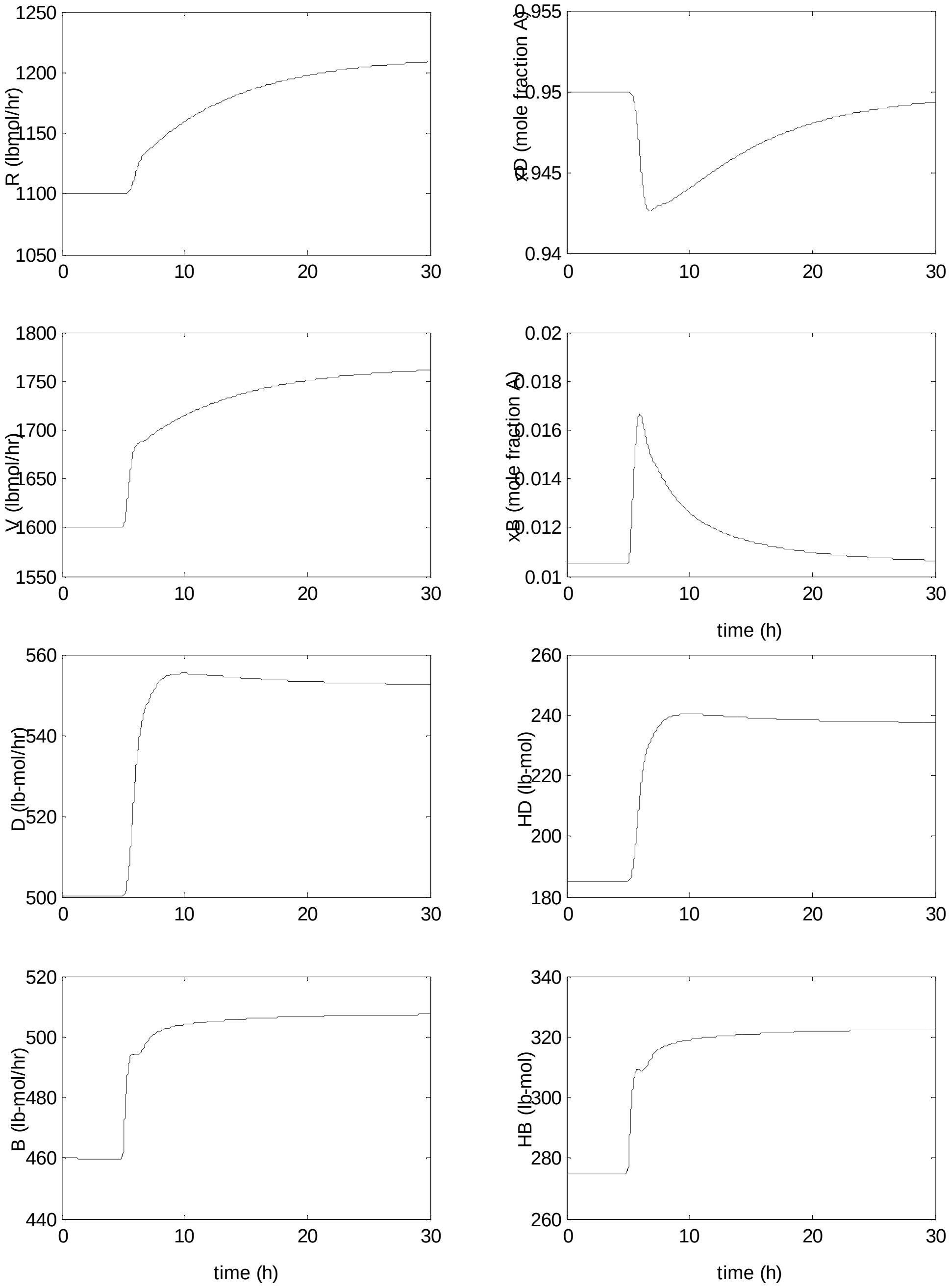

Figure S23.2e. *Step change in F (960-1060) at t=5*

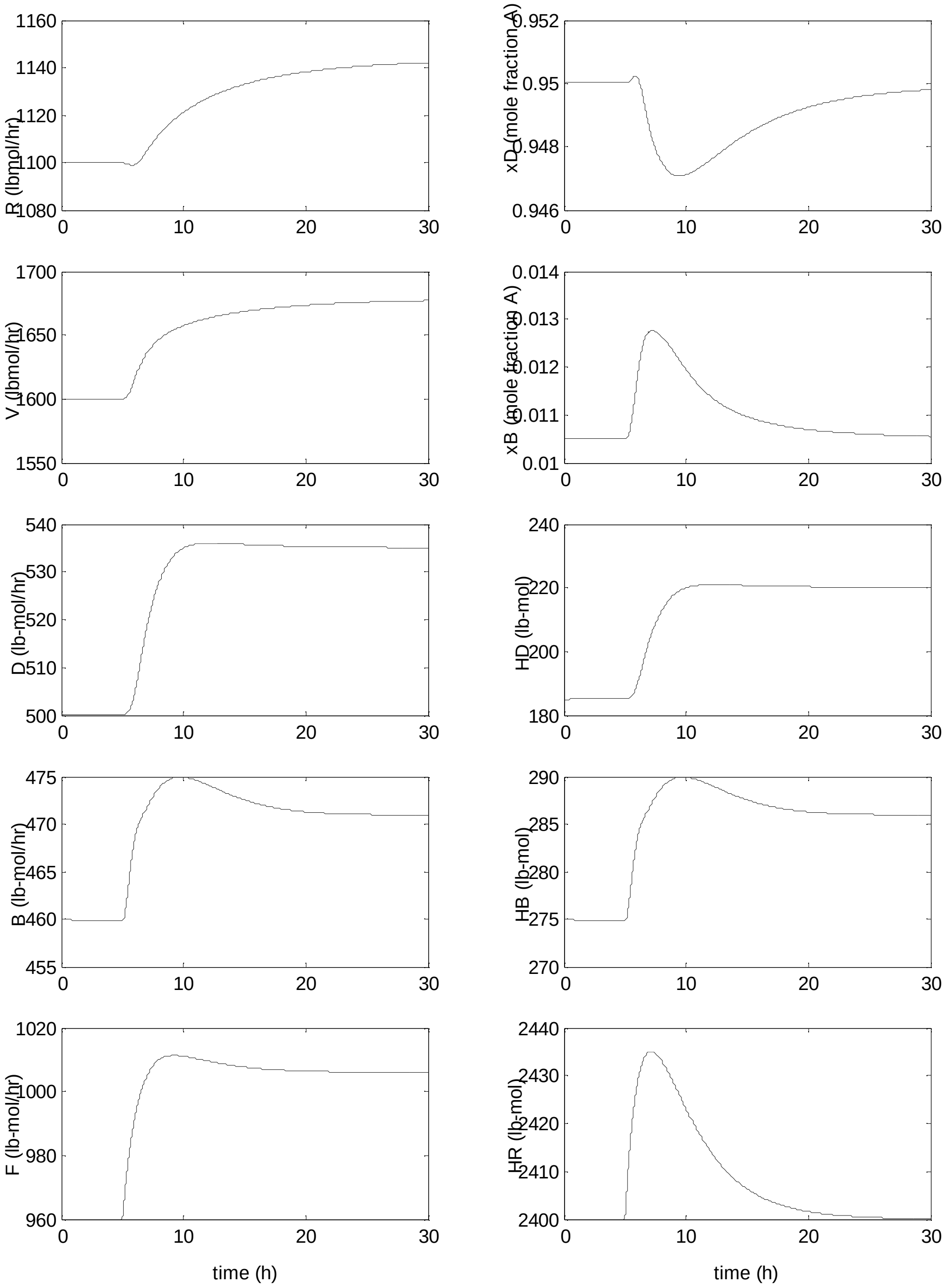

Figure S23.2f. *Step change in F_0 (460-506) at t=5*

* Calculation of First-Order Model Parameters for x_D and x_B Loops

x_D -*R*:

A step change in the reflux rate (*R*) of +10 lbmol/hr is made and the resulting response is used to fit a first-order model:

$$G(s) = \frac{K}{\tau s + 1}$$
$$K = \frac{\Delta x_D}{\Delta R} = \frac{0.9624 - 0.950}{10} = 0.0012$$

Use 63.2% of the response to find τ

$$0.632(\Delta x_D) = (0.632)(0.012) = 0.007584$$
$$\tau = time(x_D = 0.957584) = 12.33 - 10 = 2.33$$

x_B -*V*:

Similarly, a step change in the vapor boilup (*V*) of +10 lbmol/hr is made:

$$K = \frac{\Delta x_B}{\Delta V} = \frac{0.00678 - 0.0105}{10} = -0.00372$$

Use 63.2% of the response to find τ

$$0.632(\Delta x_B) = (0.632)(-0.00372) = -0.00235$$
$$\tau = time(x_B = 0.00815) = 12.08 - 10 = 2.08$$

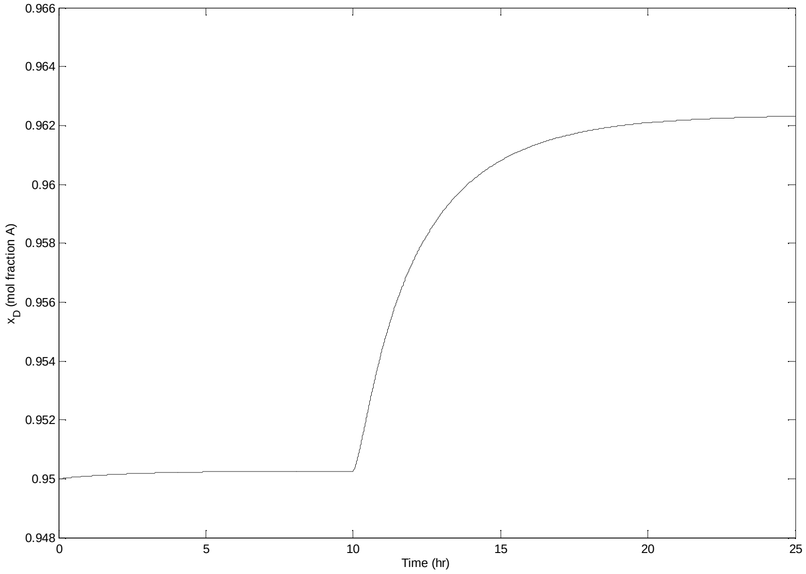

Figure S23.2g. *Responses for step change in the reflux rate R*

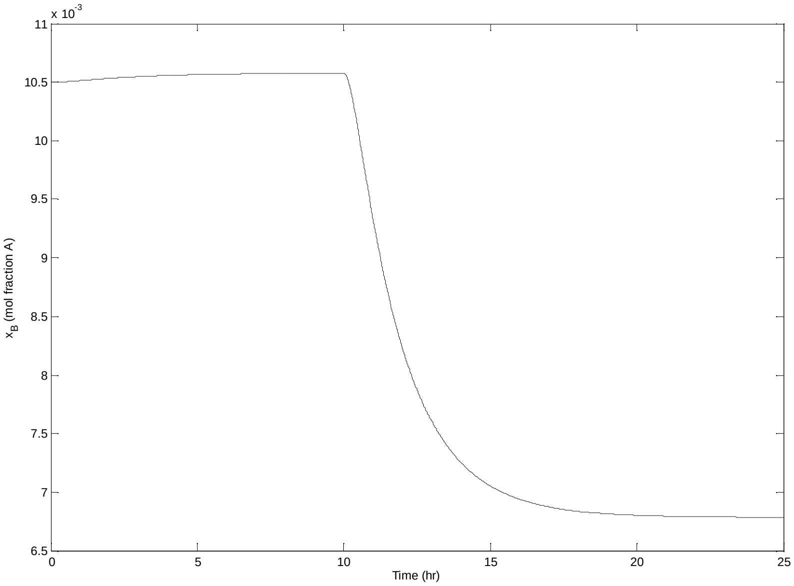

Figure S23.2h. *Responses for step change in the vapor boilup V.*

23.3

The same controller parameters are used from Exercise 23.2

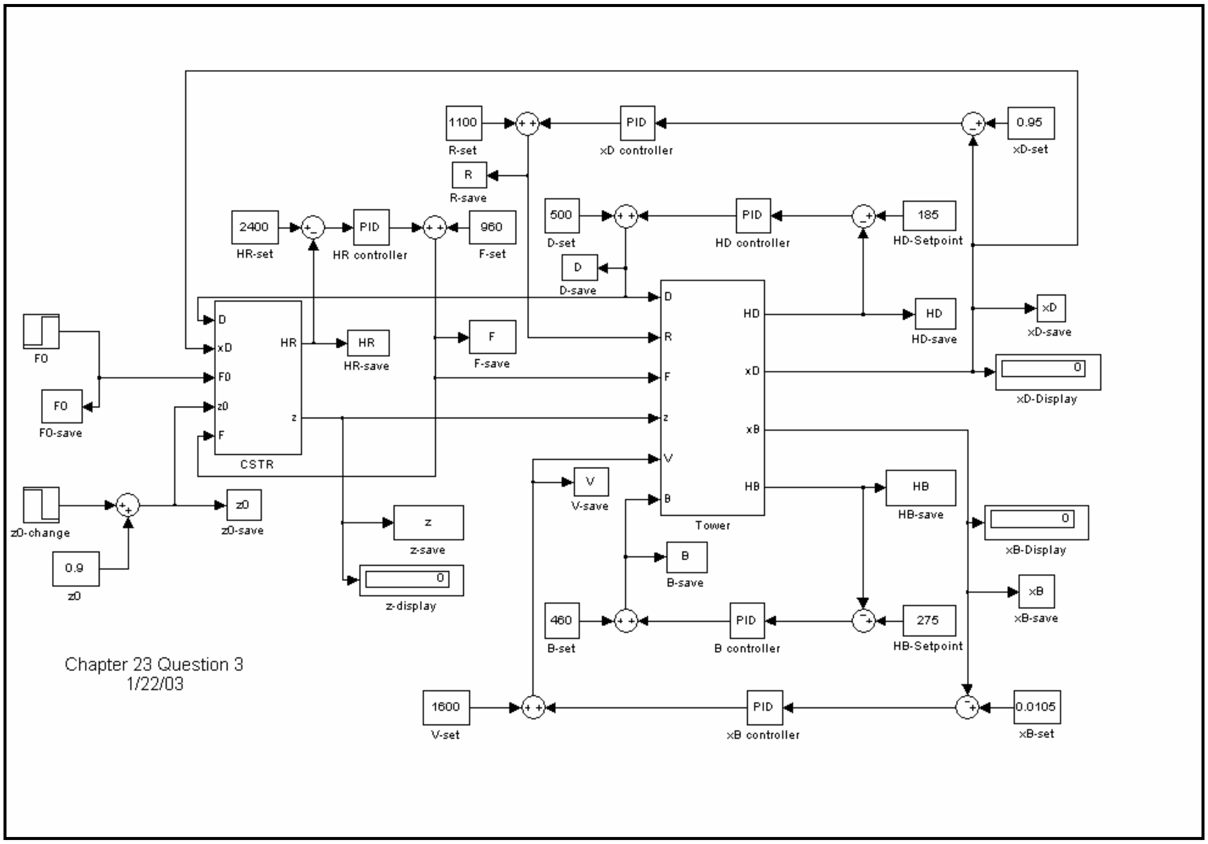

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Figure S23.3a. *Simulink-MATLAB block diagram*

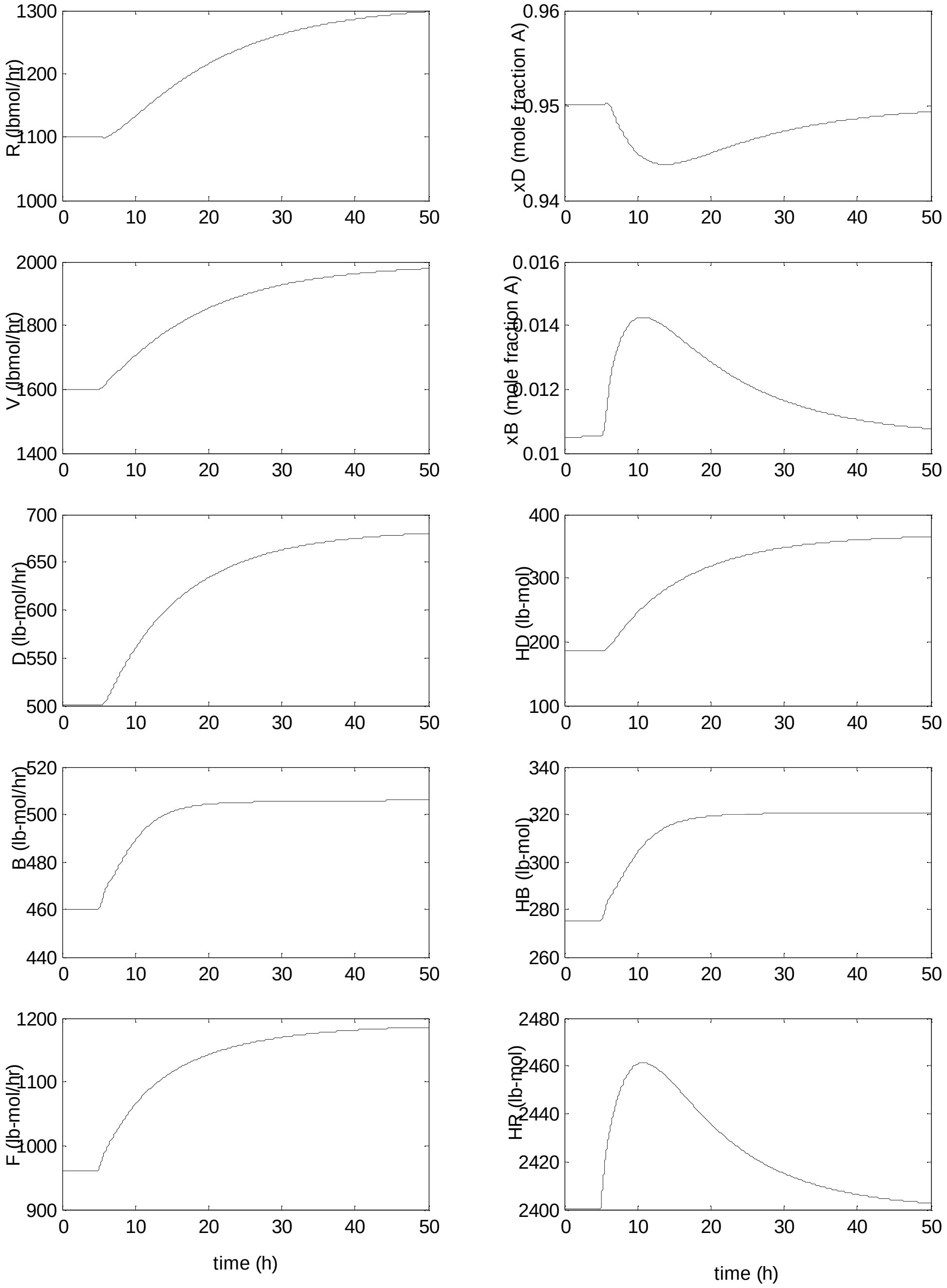

Figure S23.3b. *Step change in F_0 (+10%) at t=5*

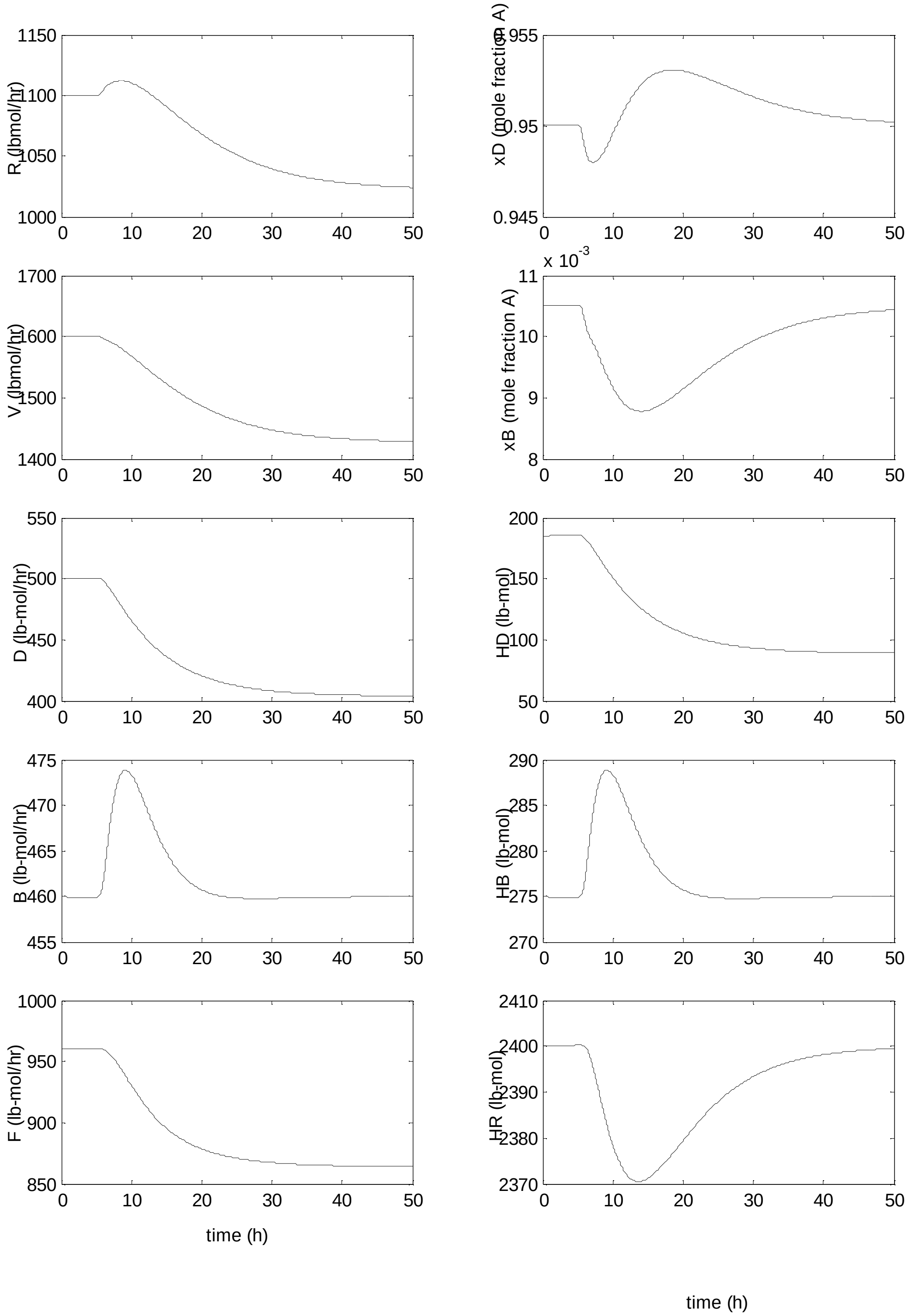

Figure S23.3c. *Step change in z_0 (-10%) at t=5*

23.4

a) The flow controller on *F*, the column feed stream, should be simulated in MATLAB as a constant flow. The controller parameters used are taken from those derived in Exercise 23.2.

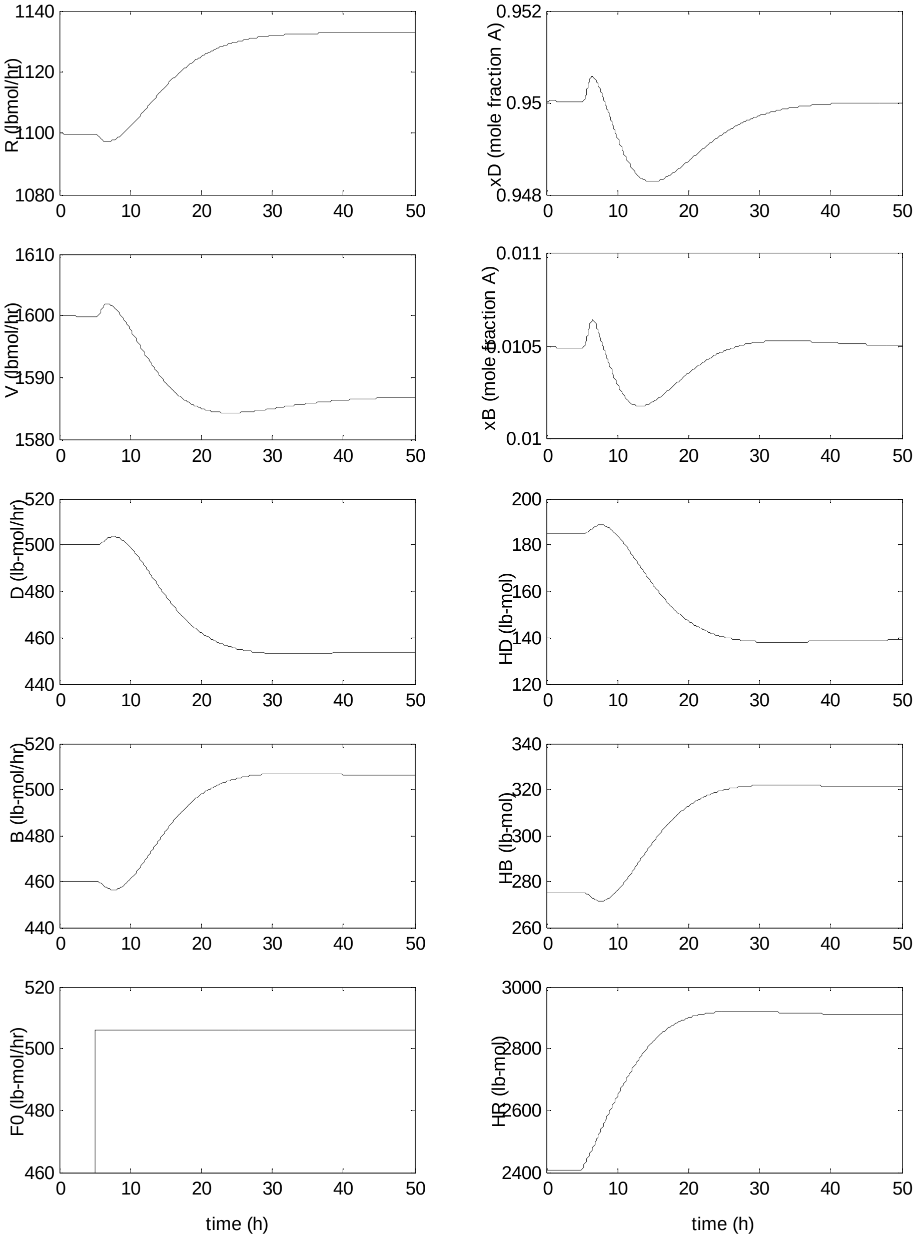

Figure S23.4a. *Step change in F_0 (+10%) at t=5*

b) Using the approximate relation (23-17), a +10% step change in F_0 will result in a reactor holdup of:

$$\bar{H}_R \approx \frac{\bar{z}_0}{k_R\left(\frac{1}{\bar{F}_0}-\frac{1}{\bar{F}}\right)} = \frac{0.90}{0.34\left(\frac{1}{506}-\frac{1}{960}\right)} = 2800\,\text{lbmol}$$

Using the exact relation (23-6, rearranged):

$$\bar{H}_R = \frac{\bar{F}_0\bar{z}_0 - \bar{B}\bar{x}_B}{k_R\bar{z}} = \frac{(506)(0.9-0.0105)}{(0.34)(0.455)} = 2910\,\text{lbmol}$$

The value taken from the graph (2910) matches up with the expected value from the equation without the approximation.

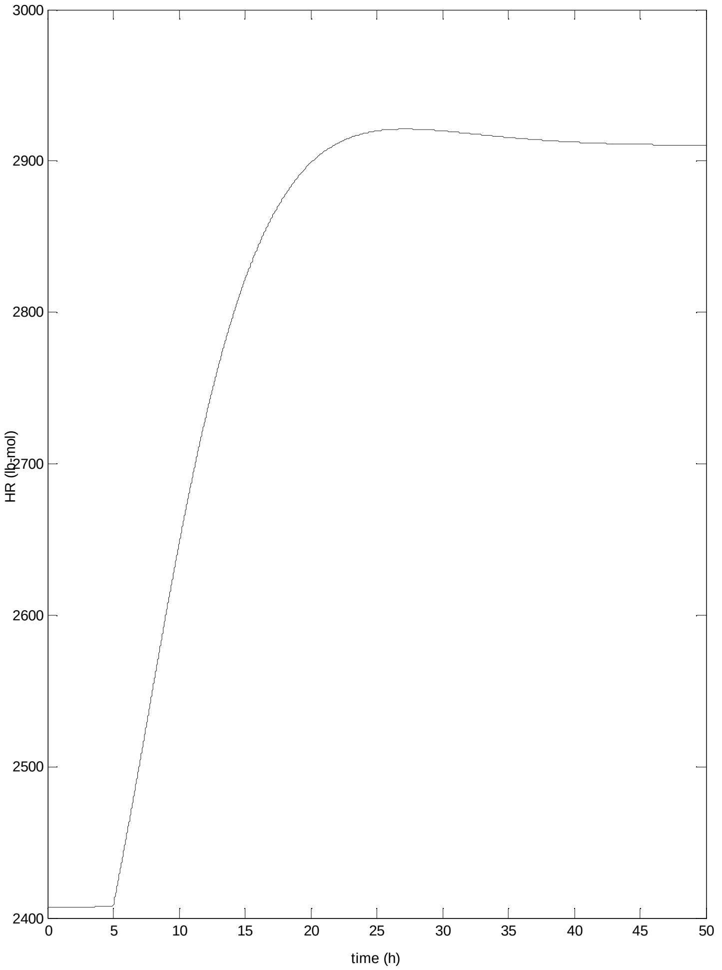

Figure S23.4b. *Step change in F_0 (+10%) at t=5*

23.5

a),b) Feedforward control is implemented using the H_R-setpoint equation:

$$H_R(t) \approx \frac{z_0(t)}{k_R\left(\frac{1}{F_0(t)} - \frac{1}{\overline{\overline{F}}}\right)}$$

Empirical adjustment of the feedforward equation is required because it is not exact:

$$H_R(t) \approx \frac{z_0(t)}{k_R\left(\frac{1}{F_0(t)} - \frac{1}{\overline{\overline{F}}}\right)} + 70$$

This adjustment matches the initial values of H_R (i.e., with and without feedforward control).

Parts a and b are represented graphically.

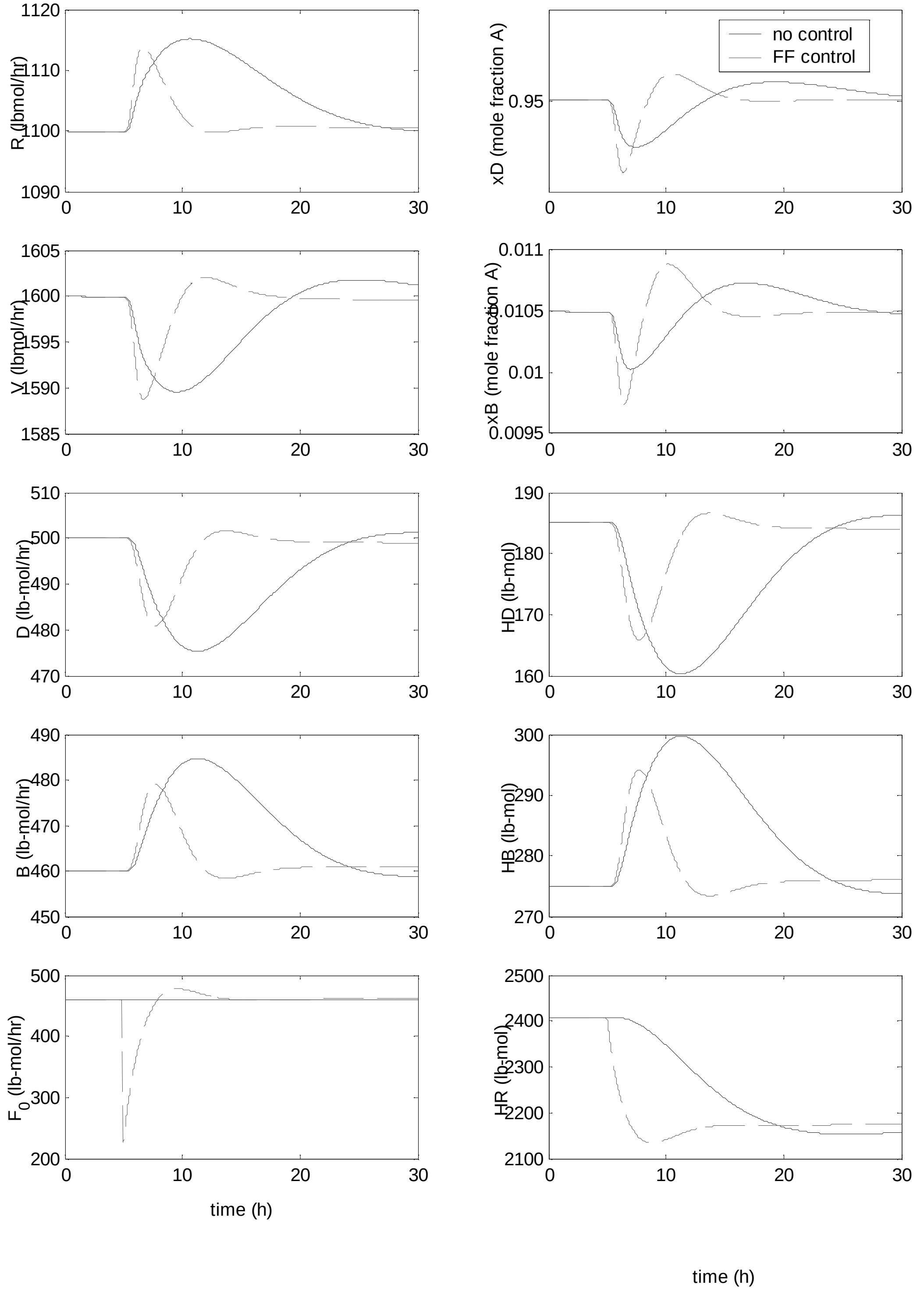

Figure S23.5a. *Step change in z_0 (-10%) at t=5*

c) The controlled plant response is much faster with the feedforward controller (~10 hours settling time versus ~20 hours without it).

d) Advantage: Faster response.
Disadvantage: Have to measure or estimate two flow rates and one concentration, therefore significantly more expensive.

23.6

a) Use a flow controller to keep F constant (make F a constant in the simulation).

b) Use ratio control to set F. The ratio should be based on the initial steady state values (960/460). Therefore, as F_0 changes, F will be controlled to the corresponding value set by the ratio.

Parts a) and b) show very different results for the two alternatives. With alternative # 3, feedforward control is necessary to keep the level in the distillate receiver from integrating. However, in alternative # 4, the control structure without the feedforward loop is superior to that with feedforward control.

Responses are displayed with controlled variables adjacent to their corresponding manipulated variable.

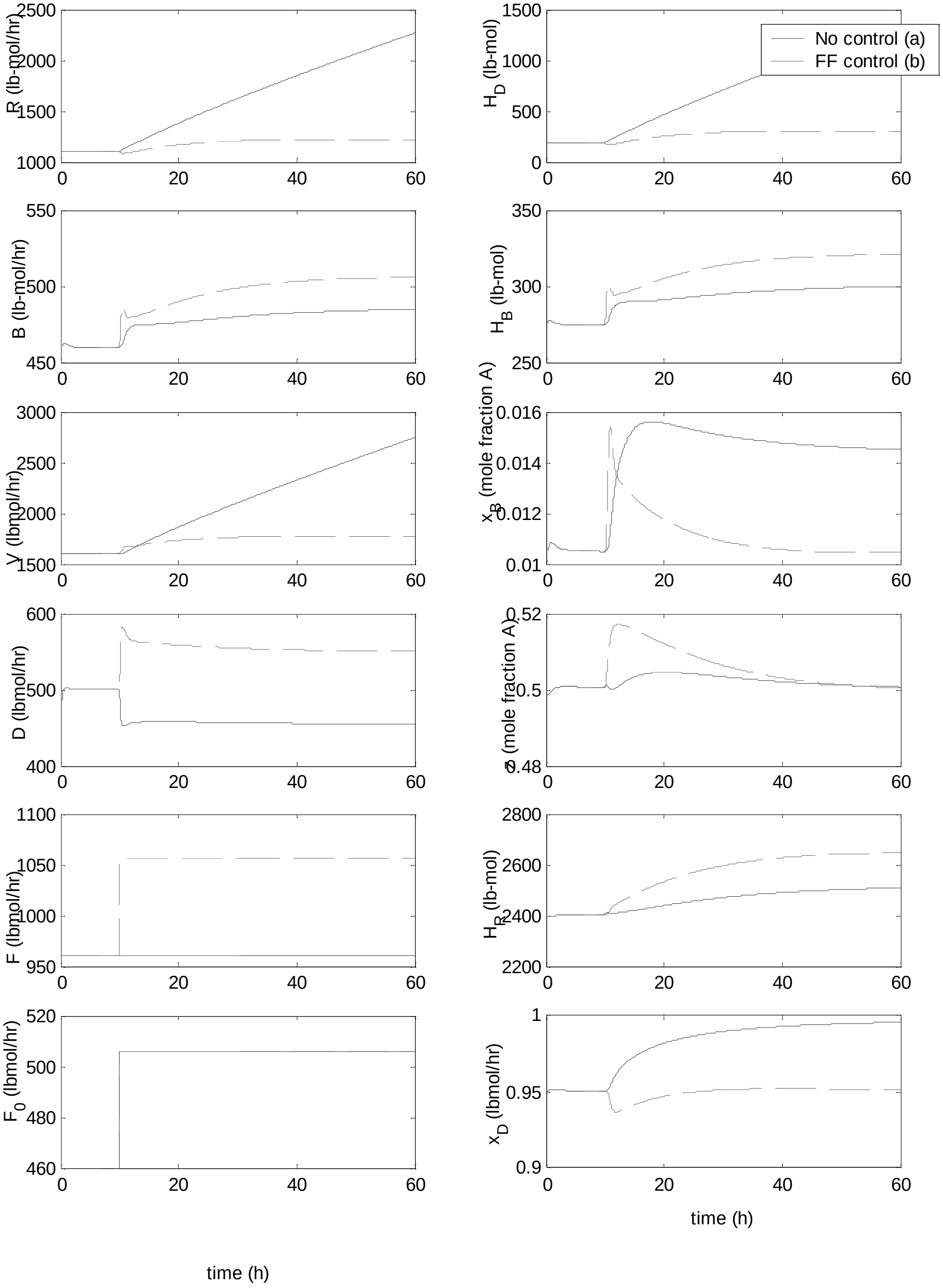

Figure S23.6a. *Alternative #3 (with and without FF controller). Step change in F_0 (+10%) at t=10*

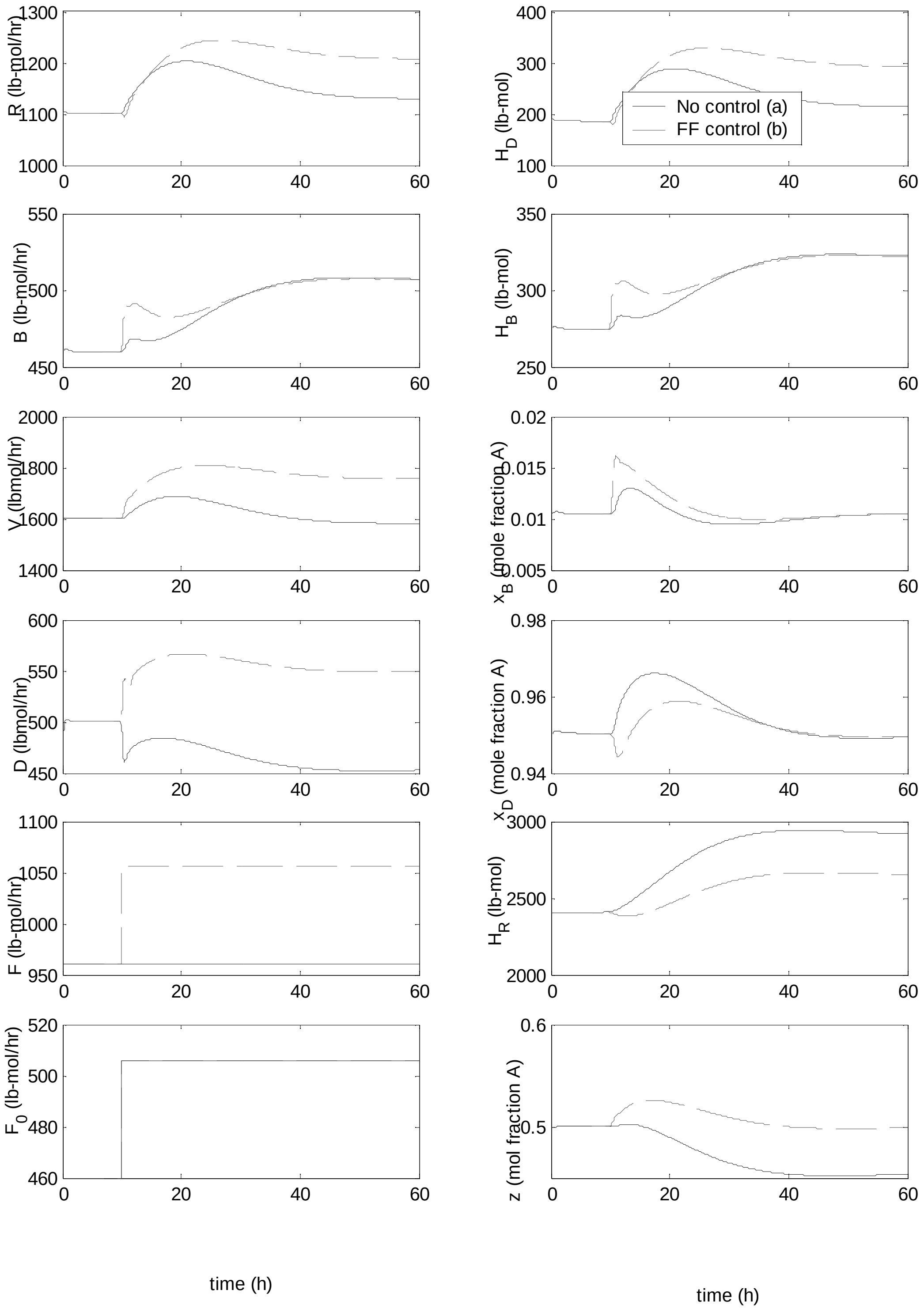

Figure S23.6b. *Alternative #4 (with and without FF controller). Step change in F_0 (+10%) at t=10*

23.7

Parts a) and b) can be satisfied by combining two or more of the previous simulations into one to compare the results together. To compare how the alternatives match up, in terms of the snowball effect, a set of arrays has been constructed.

All arrays are of the form:

$$\begin{pmatrix} \frac{D}{F_0} & \frac{H_R}{F_0} \\ \frac{D}{z_0} & \frac{H_R}{z_0} \end{pmatrix}$$

where the response of D or H_R is analyzed as a result of a step change in F_0 or z_0.

In the notation below:

S represents the occurrence of the snowball effect (>20% change in steady-state output for a 10% change in input).

A represents an acceptable response (~10% change in steady-state output).

B represents the best possible response (no change in steady-state output).

Alternative #1

$$\begin{pmatrix} S & B \\ S & B \end{pmatrix}$$

Alternative #2

$$\begin{pmatrix} B & S \\ B & A \end{pmatrix}$$

Alternative #3

$$\begin{pmatrix} A & A \\ B & A \end{pmatrix}$$

Alternative #4

$$\begin{pmatrix} A & A \\ B & A \end{pmatrix}$$

These results indicate that Alternative #2 still exhibits a snowballing characteristic, but in H_R instead of D. Alternatives #3 and #4, on the other hand, eliminate the effect altogether.

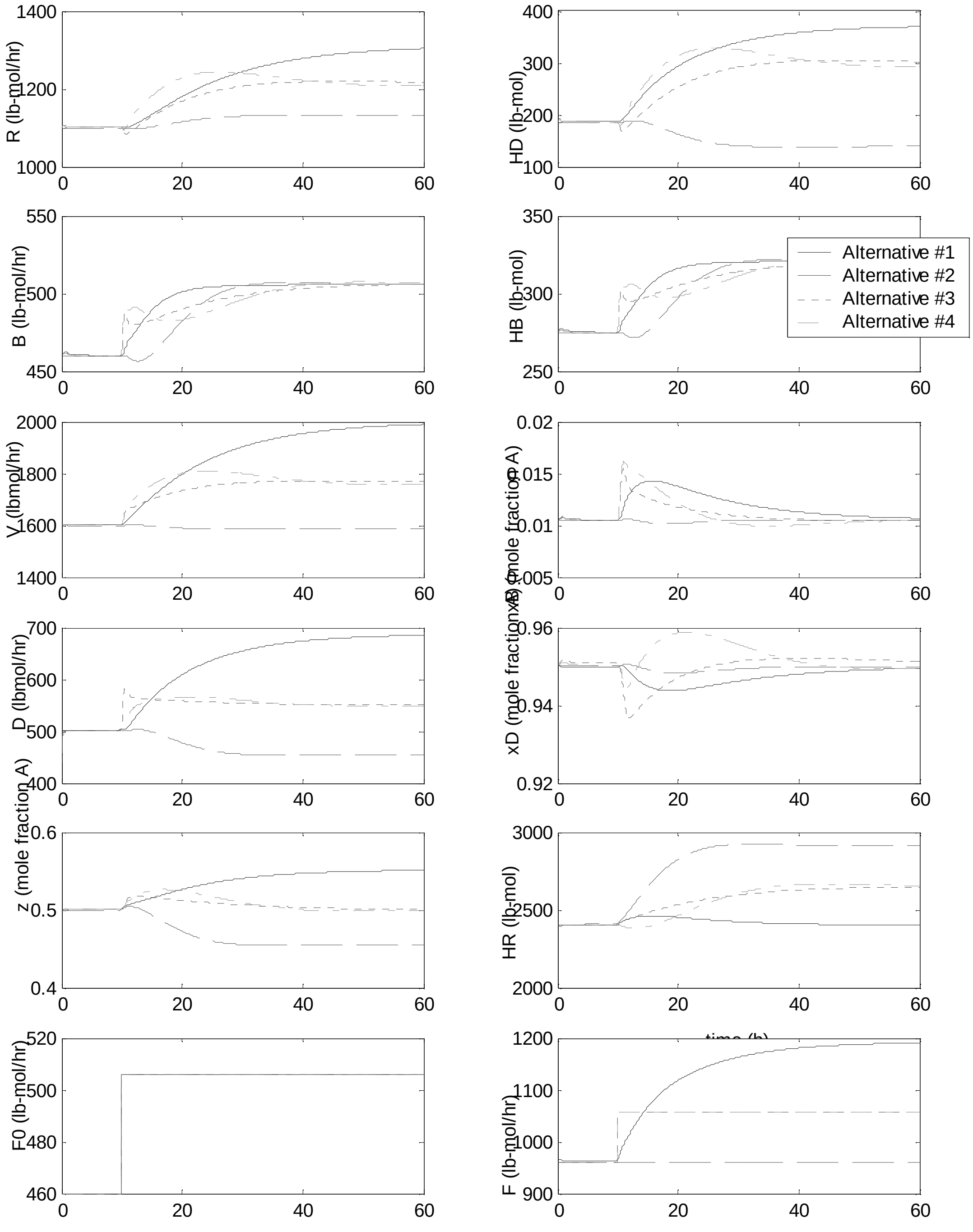

Figure S23.7a. *Step change in F_0 (+10%) at t=10*

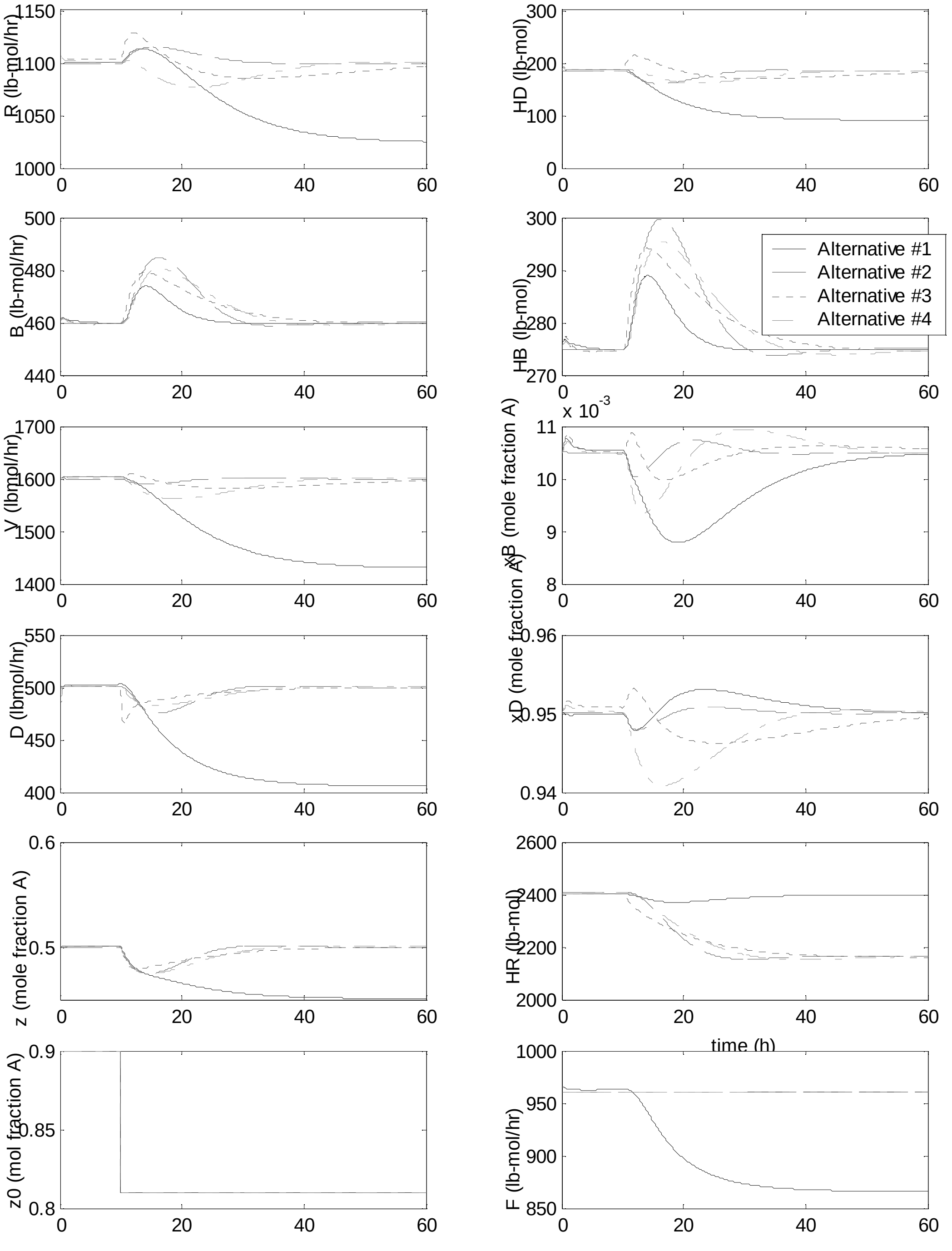

Figure S23.7b. *Step change in z_0 (-10%) at t=10*

23.8

Begin with a dynamic energy balance on the reactor:

$$C_P \frac{d(H_R(T_R - T_{Ref}))}{dt} = C_P F_0 (T_0 - T_{Ref}) + C_P D(T_D - T_{Ref}) - C_P F(T_R - T_{Ref}) - H_R \lambda kz - \dot{Q}$$

$$\dot{Q} = UA(T_R - T_C)$$

This model can be simplified using the mass balance:

$$\frac{dH_R}{dt} = F_0 + D - F$$

And, rearranging to get an equation for modeling the reactor temperature:

$$\frac{dT_R}{dt} = \frac{1}{C_P H_R}[F_0 C_P (T_0 - T_R) + DC_P (T_D - T_R) - UA(T_R - T_C) - H_R \lambda kz]$$

It is clear from the following figures that the temperature loop is much faster than the interconnected level-flow loops. This characteristic allows the reaction rate multiplier to settle before it can affect the other variables.

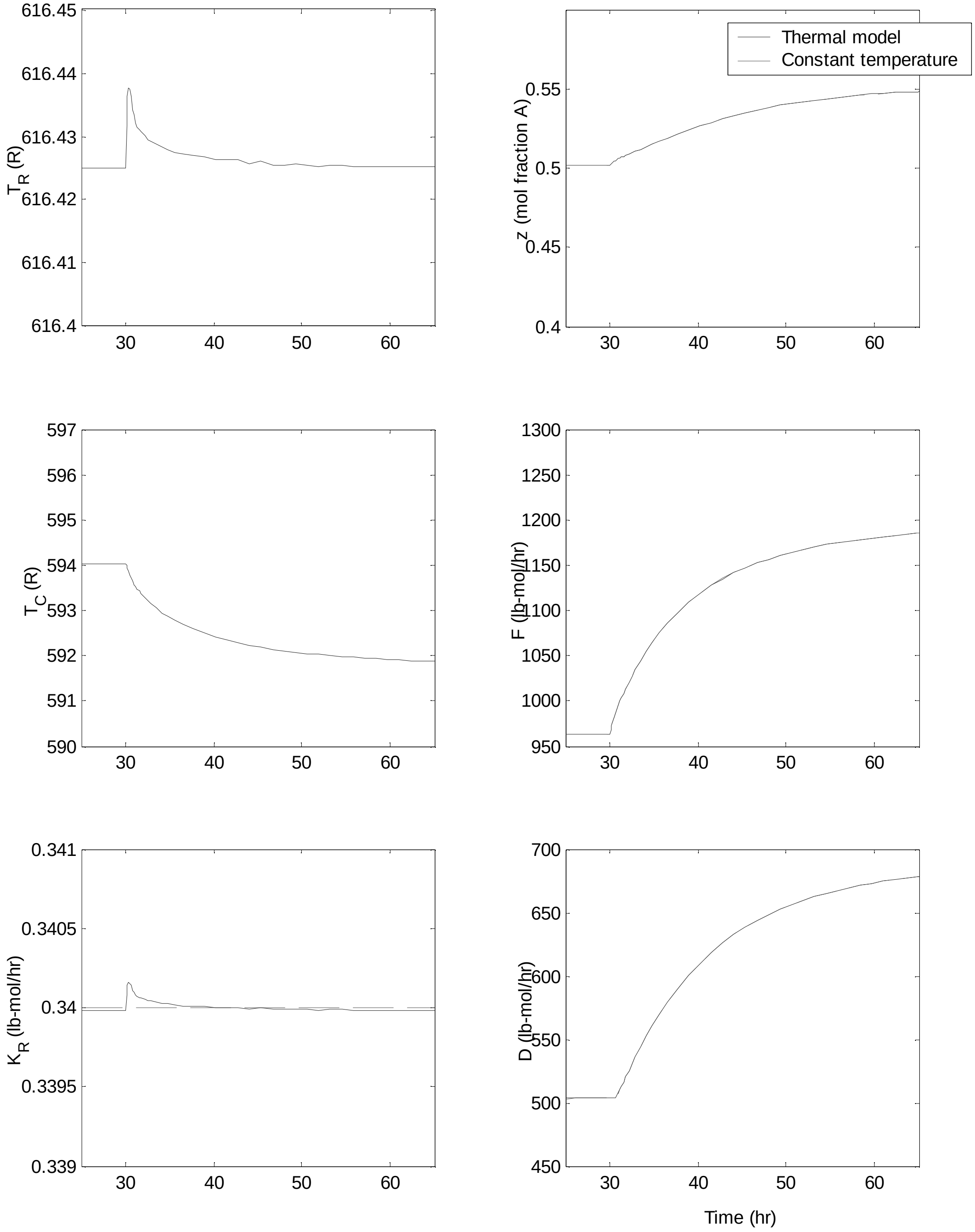

Figure S23.8a. *Step change in F_0 (+10%) at t=30 (Constant temperature simulation does not include a thermal model)*

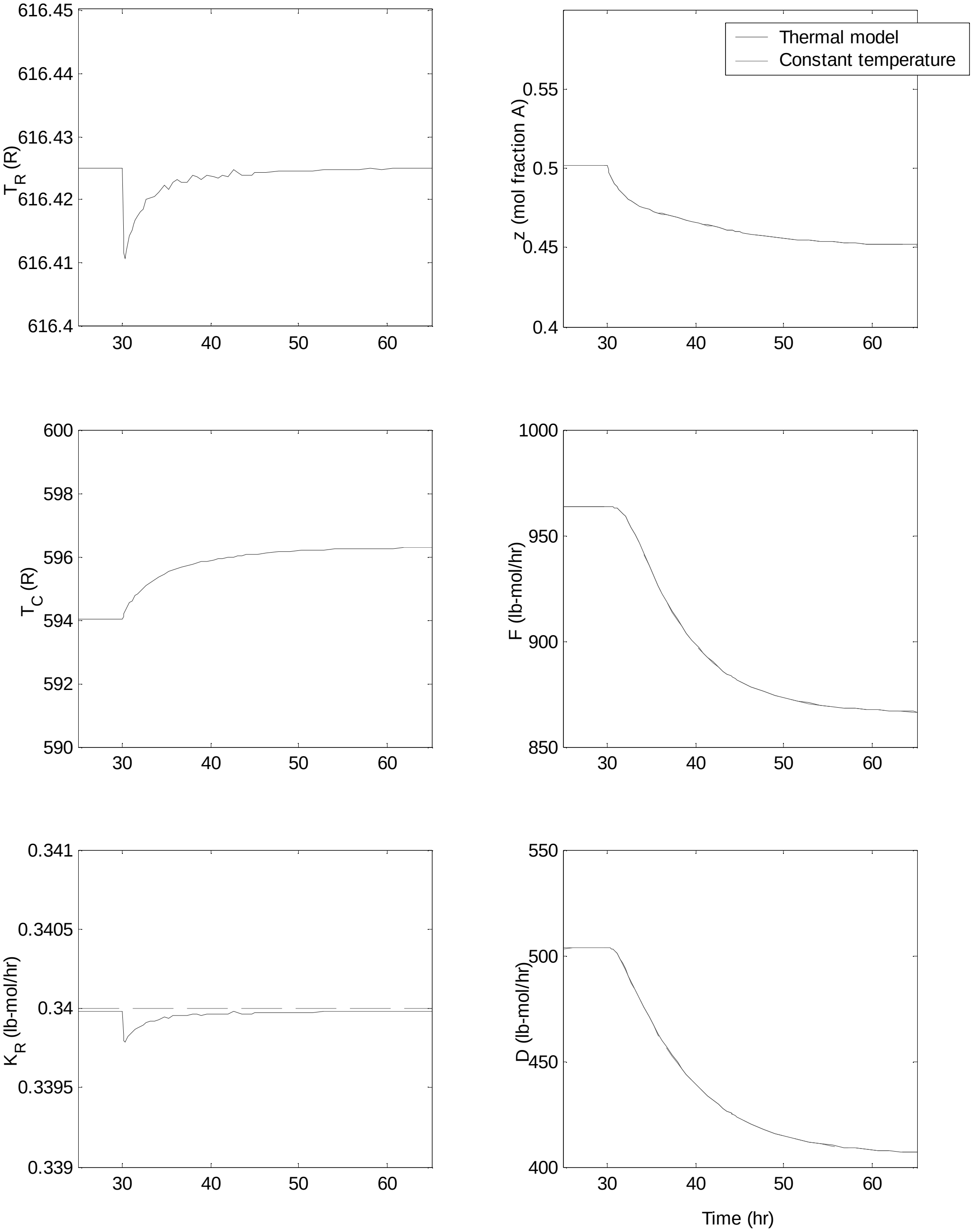

Figure S23.8b. *Step change in z_0 (-10%) at t=30 (Constant temperature simulation does not include a thermal model)*