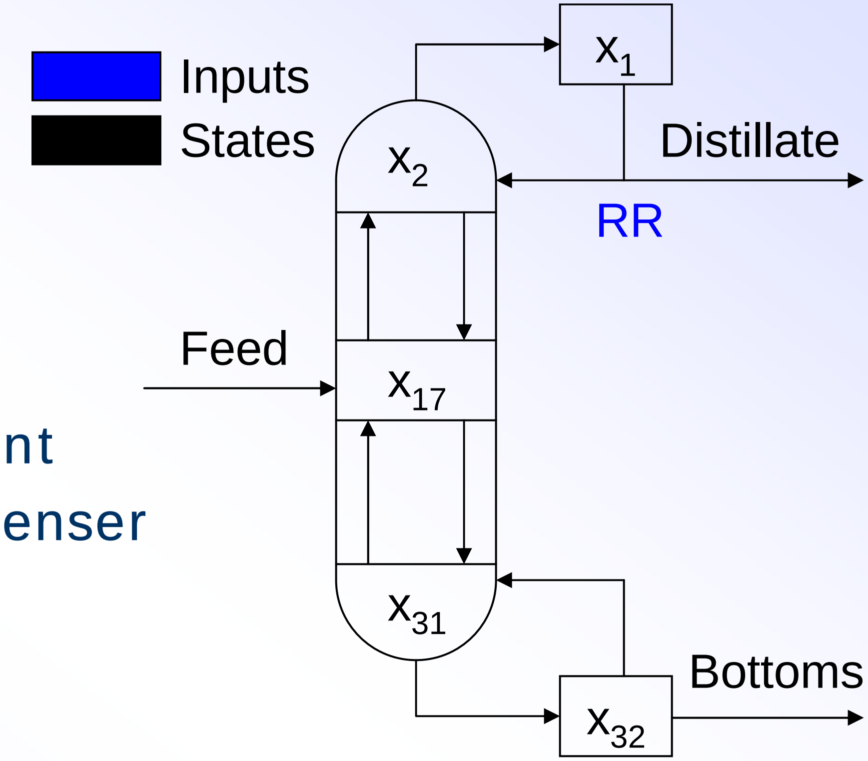
Controller Interaction

Lecture 40

ChE436 – Process Dynamic and Control

Derive a Distillation Column Model

- Groups of 2



- Two Components
- Constant Relative Volatility
- Constant Tray Molar Holdup
- Liquid Feed at the Bubble Point
- 30 Trays, Reboiler, and Condenser

- Manipulated Variables
  - RR – Reflux Ratio
  - FBOT – Fraction of Feed Leaving as Bottoms Product
- Controlled Variables
  - x[1] – Composition for Overhead Product
  - x[32] – Compositions of Bottoms

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Relative Gain Array (RGA) – pg. 348 of SEMD

- Helps Guide Decision of MV-CV pairing
- Example Distillation Column

APMonitor		CV(1)	CV(2)	SV(1)	SV(2)	SV(3)	SV(4)	SV(5)	SV(6)	SV(7)	SV(8)
	Sensitivities	ss.x[1]	ss.x[32]	ss.x[2]	ss.x[5]	ss.x[10]	ss.x[15]	ss.x[20]	ss.x[25]	ss.x[30]	ss.x[31]
FV(1)	ss.feed	-4.204E-09	4.204E-09	-5.313E-09	-5.383E-09	-2.321E-09	8.049E-09	5.356E-09	6.675E-09	5.061E-09	4.792E-09
FV(2)	ss.x_feed	0.880362	1.11964	1.30545	2.42762	2.64995	1.80586	2.16519	3.25674	2.00723	1.55214
FV(3)	ss.alpha	0.446683	-0.446683	0.606380	0.889874	0.565205	0.095437	-0.178455	-0.702162	-0.689056	-0.574441
FV(4)	ss.atray	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FV(5)	ss.acond	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
FV(6)	ss.areb	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
MV(1)	ss.rr	0.068707	-0.068707	0.101883	0.170145	0.121322	0.019434	-0.050178	-0.152264	-0.118547	-0.093852
MV(2)	ss.fbot	0.314140	1.42754	0.465825	0.866247	0.945584	0.644385	1.43757	3.39092	2.48516	1.95664

- $K_{11}=0.069$
- $K_{12}=0.314$
- $K_{21}=-0.069$
- $K_{22}=1.428$

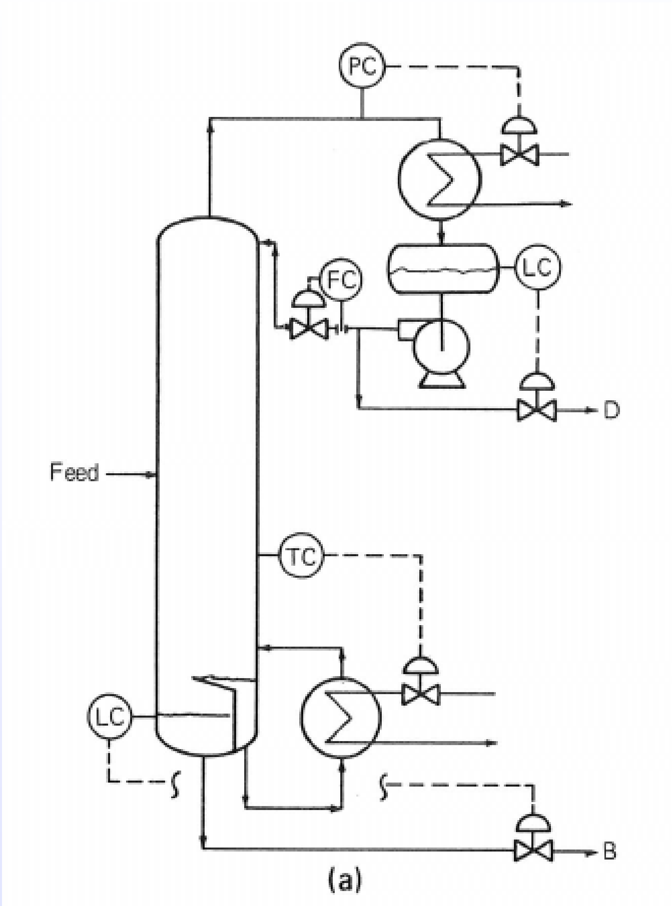
$$\lambda_{11}=\lambda_{22}=\frac{1}{1-\frac{K_{12}K_{21}}{K_{11}K_{22}}}$$

$$\lambda_{12}=\lambda_{21}=1-\lambda_{11}$$

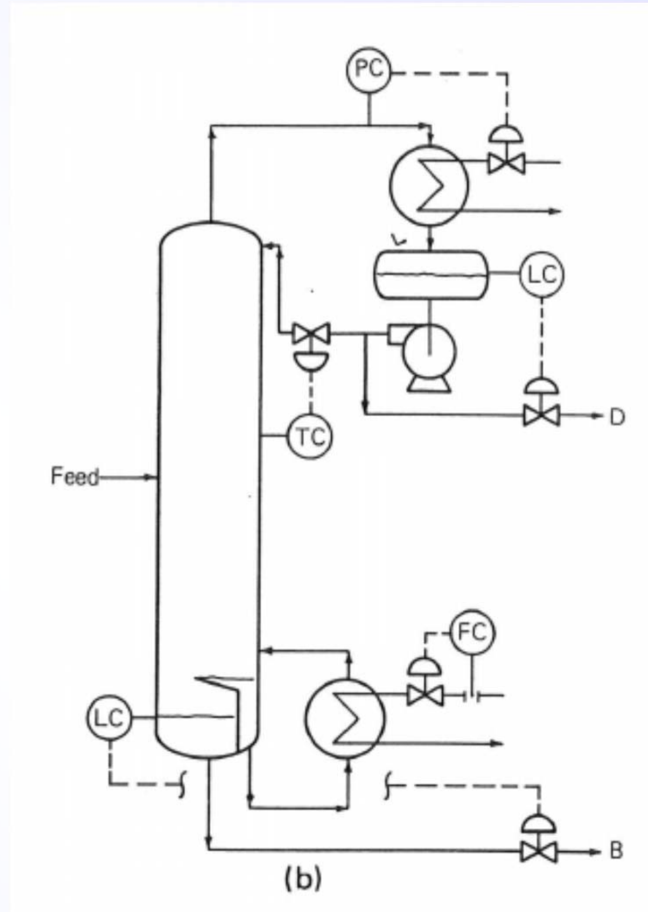
$$\Lambda=\begin{bmatrix}\lambda_{11} & \lambda_{12}\\ \lambda_{21} & \lambda_{22}\end{bmatrix}=\begin{bmatrix}0.82 & 0.18\\ 0.18 & 0.82\end{bmatrix}$$

Pick pairings with positive values closest to 1.0.

RGA Tool to Select Best Control Option

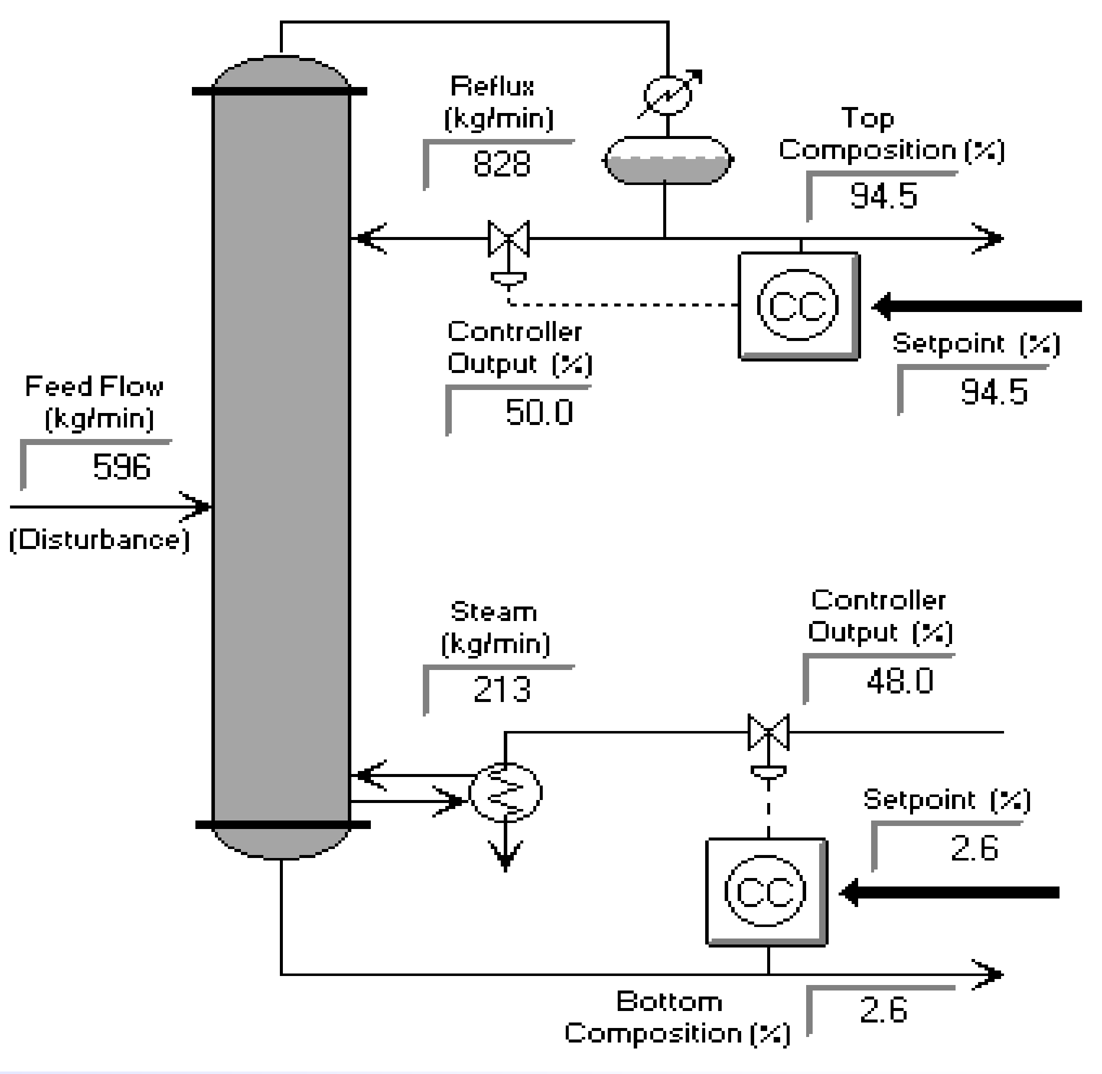


a. Indirect control, composition regulates boilup



b. Indirect control, composition regulates reflux

Distillation Column is a 2 x 2 Multivariable Challenge



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Results of Open-Loop Step Tests on Top and Bottom

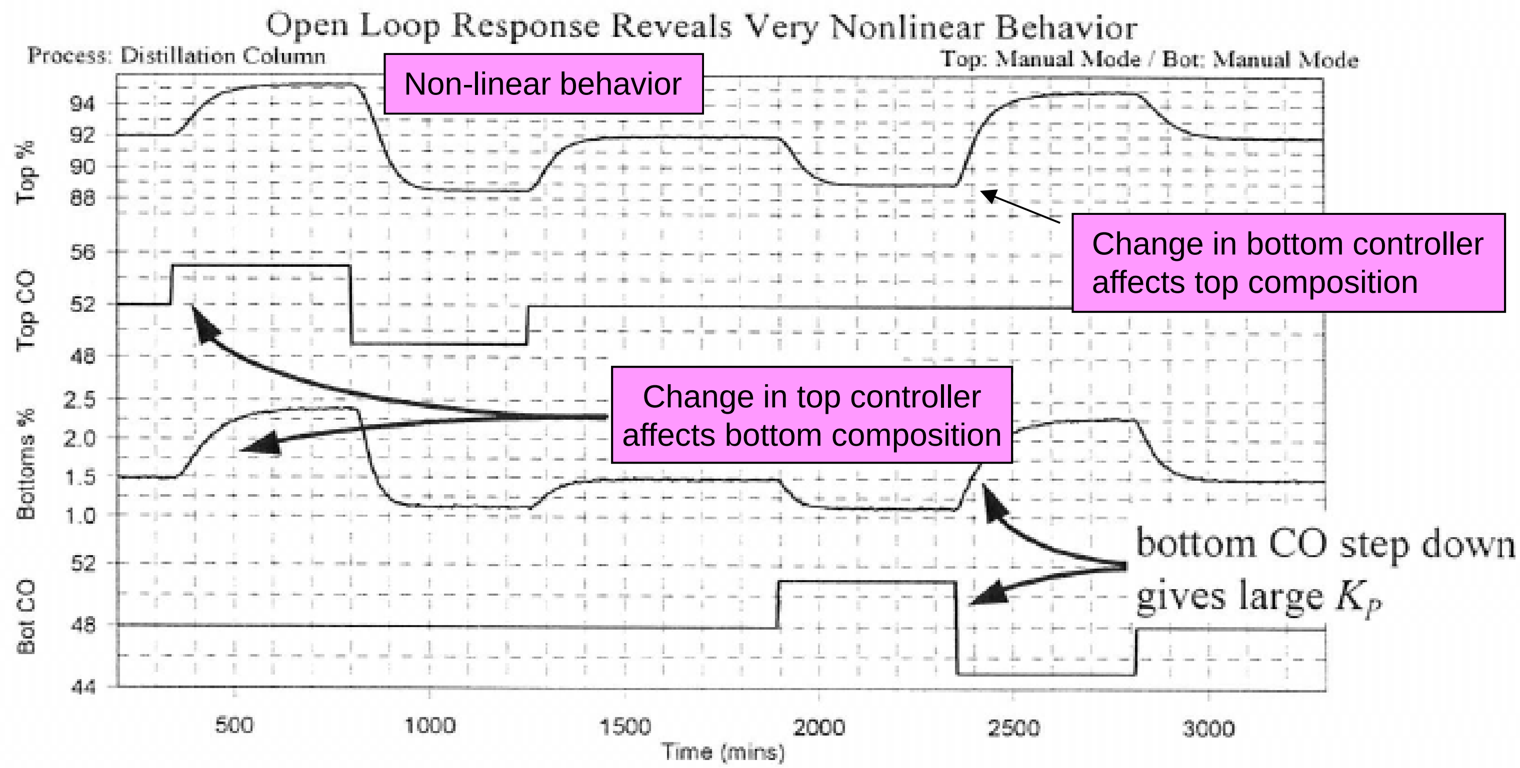

Figure 20.4 - Open loop step tests on the distillation column's top and bottom controller

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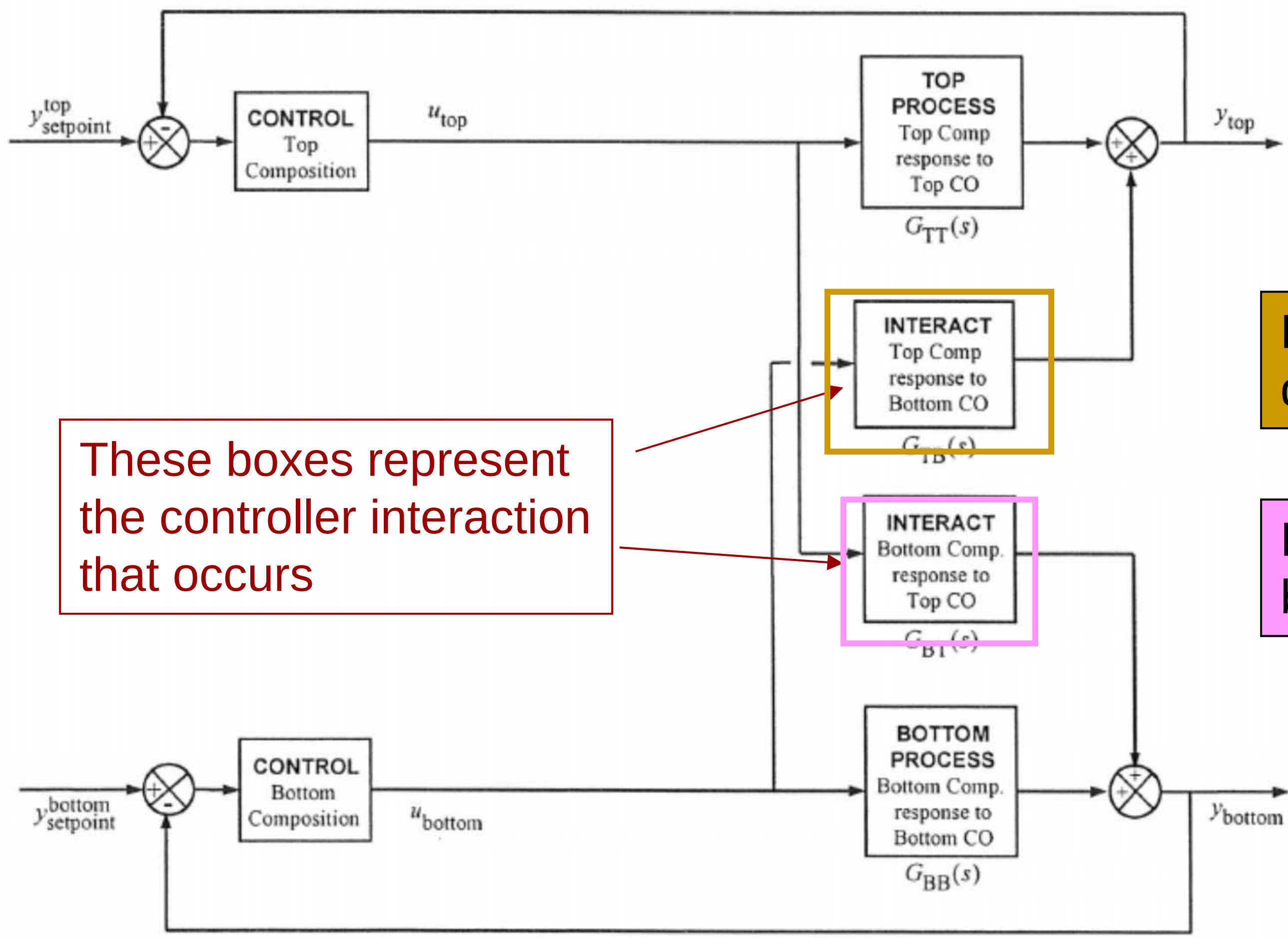
Control Loop Interaction



These boxes represent the controller interaction that occurs

Figure 20.2 - Block diagram of top and bottom distillation control loops with "cross loop" interaction

Disturbance for distillate (top)

Disturbance for bottoms

Using PI Controllers for Top and Bottom

Set point change from 92 to 94% benzene in top

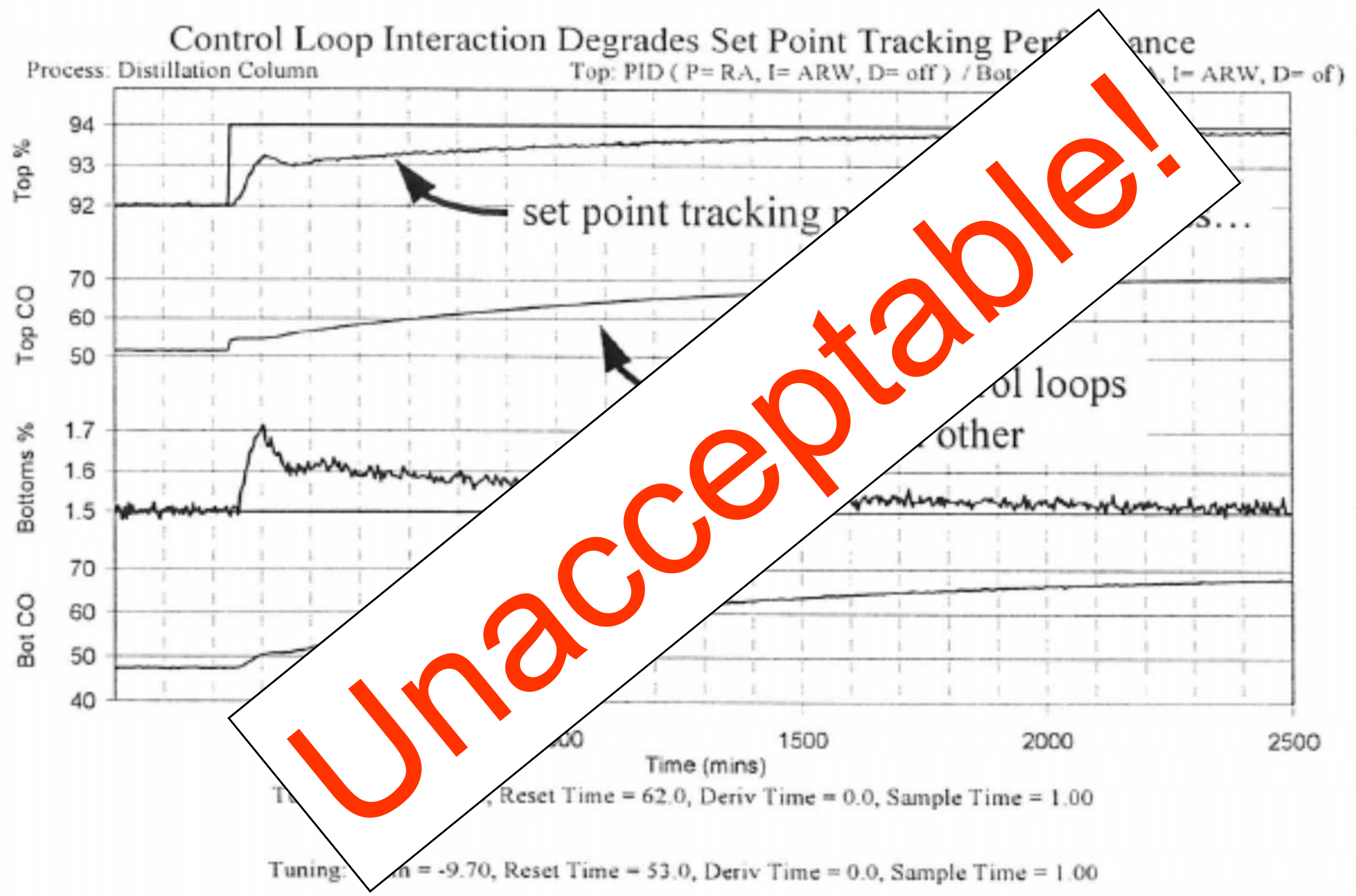


Figure 20.8 - Top and bottom loop fight each other, thus degrading set point tracking performance of top loop

Solution.... Decouplers or Multi-variable Control

Decouplers

- Act kind of like a feed-forward controller
- Model what the interaction will be, and compensate
  - Treat the bottom composition as a disturbance with a feedforward loop to the top controller
  - Treat the top composition as a disturbance with a feedforward loop to the bottom controller
- Remember that a feedforward controller has the form:
  - $G_{FF} = - G_{dist}/G_{process}$
- So the decoupler transfer functions become:
  - $D_{top\ decoupler} = - G_{TB}(s)/G_{TT}(s)$
  - $D_{bottom\ decoupler} = - G_{BT}(s)/G_{BB}(s)$

G_{TB} is response of top composition to change in bottoms controller, etc.

Decoupling Structure

D here means a transfer function for a decoupler

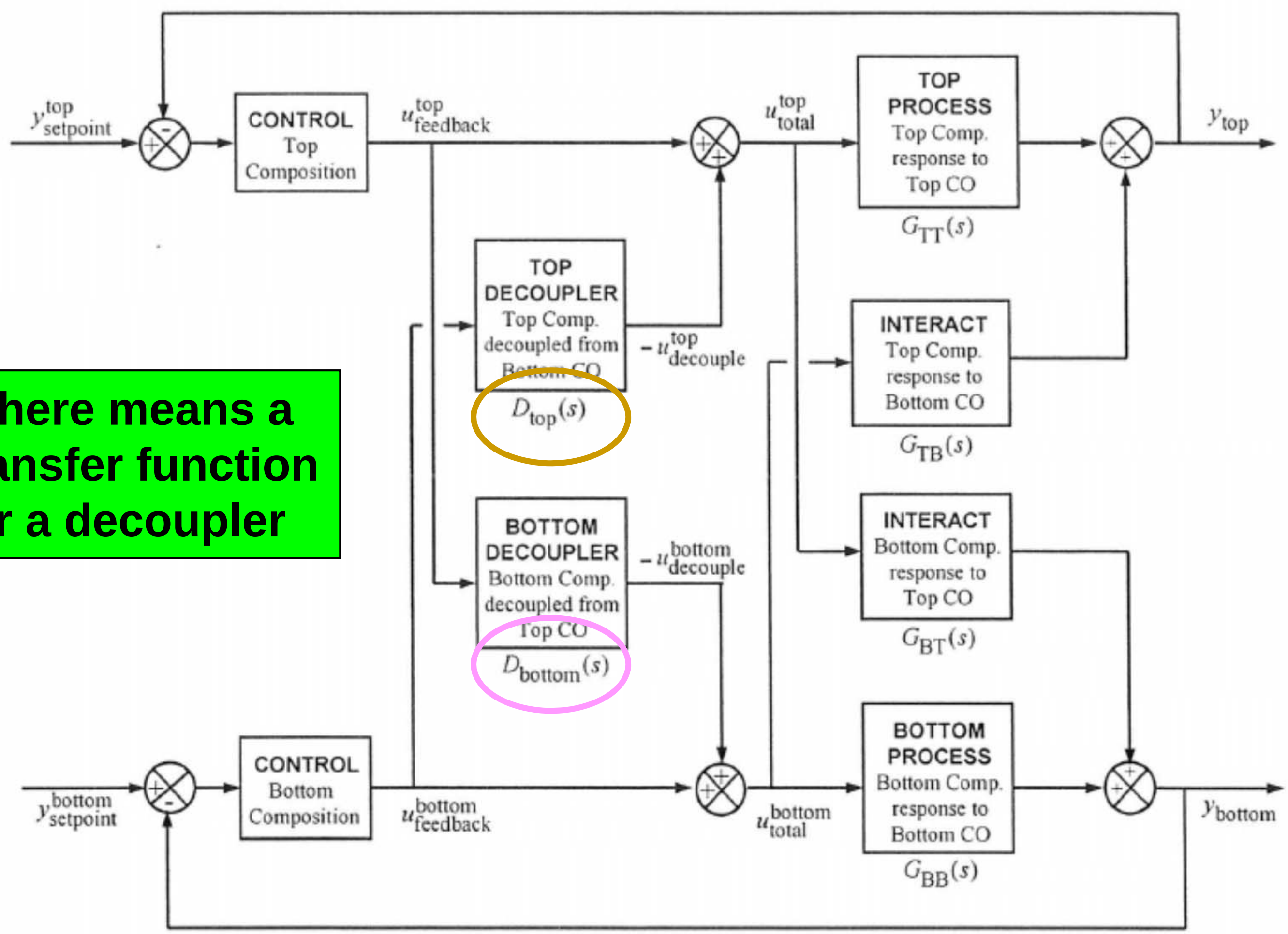


Figure 20.3 - Block diagram of top and bottom distillation control loops with cross loop interaction and decouplers

Decouplers are Feed Forward Controllers

- A decoupler is comprised of a process model and a cross-loop disturbance model:
  - The cross-loop disturbance model receives the cross-loop controller signal and predicts an "impact profile," or when and by how much the process variable will be impacted
  - Given this predicted sequence of disruption, the *process* model then back calculates a series of control actions that exactly counteract the cross-loop disturbance as it arrives so the measured process variable remains constant at set point
- A new sensor is not needed because the cross-loop controller signal is readily available for use by the decoupler
- Developing and programming the dynamic process and cross-loop disturbance models is required for implementation

Tuning Procedure

- Get process responses to top and bottom controllers in open-loop mode
  - Do a pulse, not a doublet, to get highest K_p (and hence lowest K_c) for top and for bottom
  - Fit FOPDT models to:
    1. G_{TT} (top response to change in top controller output)
    2. G_{BB} (bottom response to change in bottom controller output)
    3. G_{TB} (top response to change in bottom controller output)
    4. G_{BT} (bottom response to change in top controller output)
- Get PI Controller parameters from IMC correlation
- Put in FOPDT parameters for G_{TT}, G_{BB}, G_{TB}, and G_{BT} into decouplers
  - G_{BT} goes into bottom decoupler

Improved Performance with Decouplers

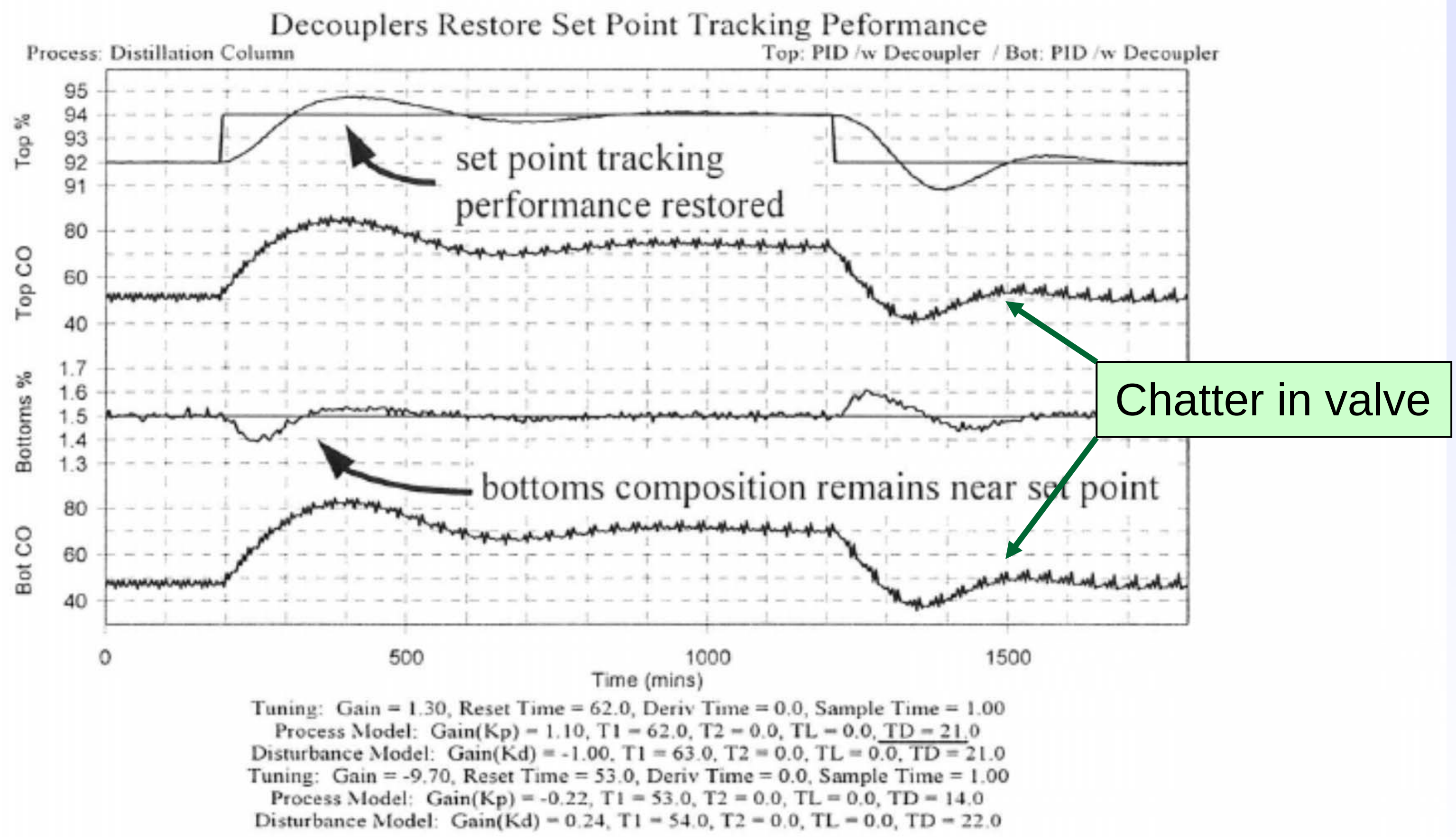


Figure 20.11 - Improved set point tracking capability of top loop and reduced interaction with bottom loop when both are under PI control with decouplers

Subtle Problem

- $K_{P,BB}$ = -0.22 %/%
- $K_{D,BT}$ = 0.24 %/%

- If the disturbance gain is greater than the process gain, things don't work well!

- Solution:
  - Set | $K_{D,BT}$ | = | $K_{P,BB}$ |, or $K_{D,BT}$ = 0.22 %/%

Final Result

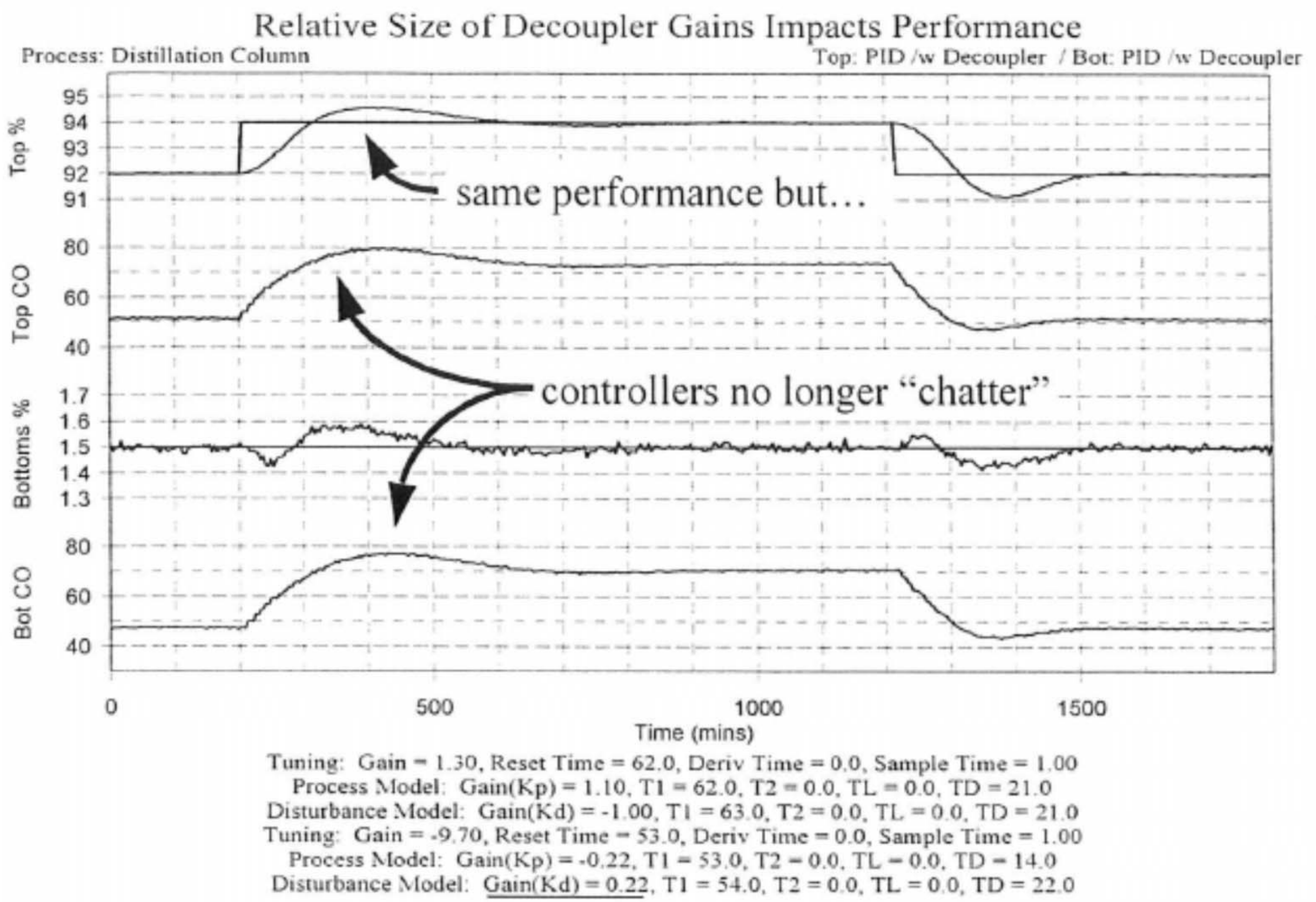


Figure 20.12 - Decoupled loops do not chatter with slight adjustment to one model parameter

Conclusion

- With just 2 controllers, controller interaction was significant!!
  - Decouplers used, but somewhat complicated

- Imagine what will happen with multiple controllers!

- Opportunity for Multi-variable control!