

Introduction



Beyond Single Loop PID Control: Model-Based and Combined Feedforward-Feedback Control

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OUTLINE

- Outline
 - Basic and Advanced Regulatory Control Definitions
 - Combined Feedforward-Feedback Control
 - Example 1: Combined Feedforward-Feedback Control of Distillation Column
 - Combined Feedforward-Feedback Tuning Methodology
 - Model-Based Control and Controller Types
 - Example 2: Cooling Tower Water Quality Composition Control
 - Summary



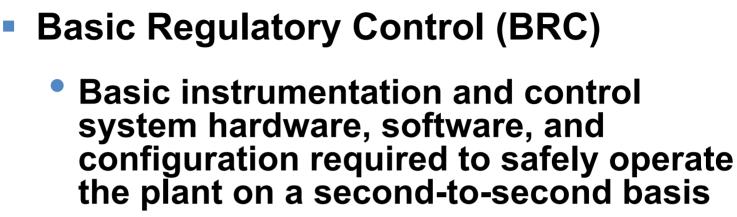
Basic Regulatory Control



- Controlled Variable (CV) stays within a predefined limit around the setpoint irrespective of routine disturbances that routinely affect the control loop
- Feedback Control
 - Single loop feedback control is adequate to meet the primary control objective for most processes
 - Effect of disturbances is not taken into account in advance



Basic Regulatory Control (Cont'd)



- Should be able to handle routine load disturbances
- Includes sequential regulatory control and batch logic if required
- Includes required equipment interlock logic and safety, health & environmental controls



Advanced Regulatory Control



- Advanced Regulatory Control (ARC)
 - Extends control system capability beyond regulatory and sequential control to move the process closer to its economic optimum
 - Typically implemented to:
 - Improve operating efficiency and profitability
 - Increase plant production
 - Improve plant stability and operability
 - Better reject routine control loop disturbances



Advanced Regulatory Control (Cont'd)



- Advanced Regulatory Control (ARC)
 - Coordinates or ties together control for multiple loops
 - Typical Advanced Regulatory Control industrial applications:
 - Cascade control
 - Override control
 - Combined feedforward-feedback control
 - Model-based control (including Model Predictive Control)
 - Inferential composition control



Feedforward Control



Feedforward Control

- Sustained control error must have enough economic impact to justify higher design and implementation costs
- Can minimize adverse effects of:
 - & Large magnitude/frequent input disturbances
 - To some degree significant process lag
- Effect of disturbance variable(s) on CV must be measurable
- Cost/complexity trade-off



Feedforward and Combined Feedforward-Feedback Control

- Why Use Combined Feedforward-Feedback Control?
 - Feedforward control only is not practical because it requires:
 - Accurate modeling of the process
 - Ability to predict and model the effect of all possible disturbance variable(s) on the primary controlled variable (CV)
 - So Combined Feedforward-Feedback control is generally used



Feedforward Control Types



- **Steady-State Feedforward Control**
 - Most simple and direct approach
 - No dynamic effects included
 - Instantaneous correction applied to manipulated variable
 - May not achieve control objective if dynamic effects are significant (and they usually are...)





- **Dynamic Feedforward Control**
 - Takes into account:
 - Process dynamics (usually most significant)
 - Disturbance dynamics
 - Sensor dynamics
 - Can be implemented by:
 - Generic dynamic compensator (most common)
 - Application-specific feedforward control strategy and calculation

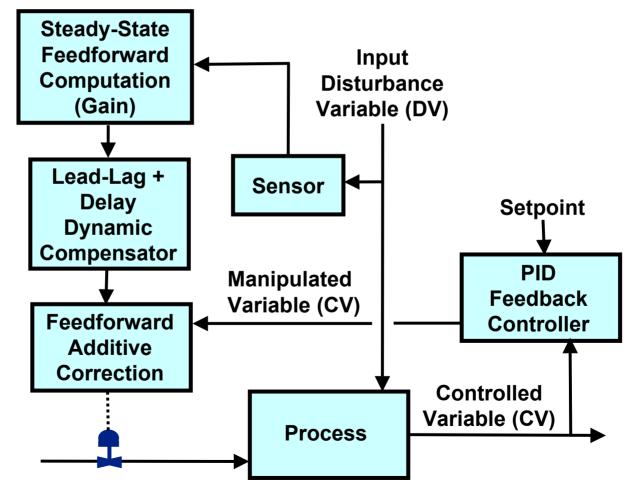


Feedforward Dynamic Compensation



'Generic' FFD-FDBK Dyn. Compensator

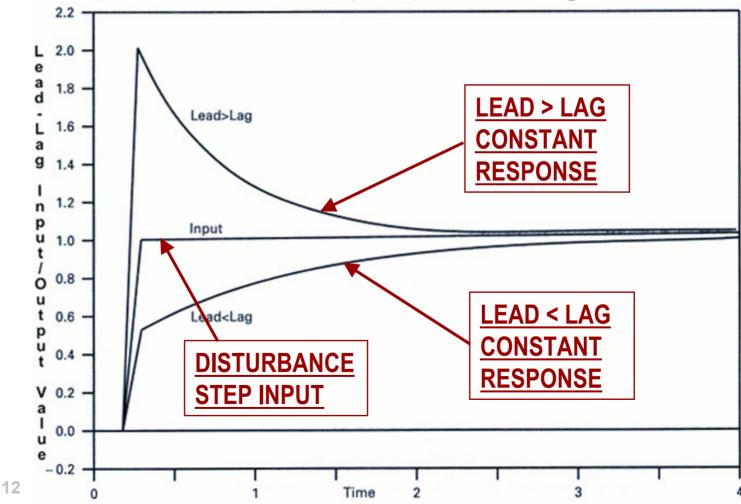
DYNAMIC FEEDFORWARD CONTROL APPLIED AS AN ADDITIVE CORRECTION TO PID FEEDBACK CONTROLLER OUTPUT



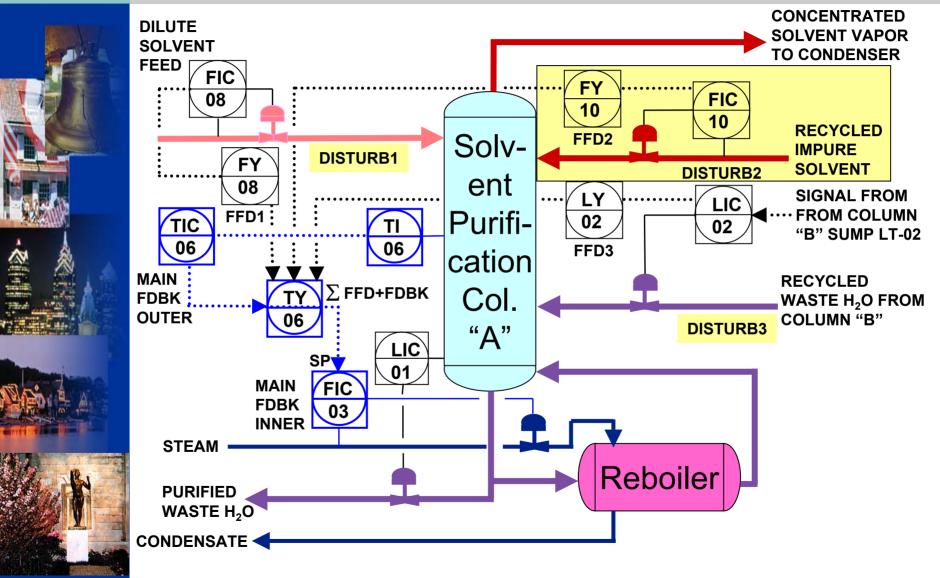


Feedforward Dynamic Compensation (Cont'd)

Feedforward Dynamic Compensation – Response of a Lead-Lag Dynamic Compensator to Step Input Change



Example 1: Distillation Column Combined FFD-FDBK Control Process Schematic



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General Preparation for Tuning



- Establish the control loop performance criteria
- Determine allowable operating and understand safety limits for the control loop and other affected variables
- Obtain any necessary operations work and safety permits if required

Irrespective of tuning method used:

- Familiarize yourself with the process (there is no substitute for thorough process understanding!)
- Understand in detail the data acquisition and control system and algorithms used including optional features



Feedforward Tuning Methodology



- ALWAYS conduct at least 1-2 process response tests
 - Using an appropriate input disturbance such as a step or pulse (symmetrical or asymmetrical doublet pulse preferred)
 - Conduct process response tests at different parts of the normal operating range of the controlled variable

> Average the results, assess nonlinearity

- If cascades are present, conduct process response test(s) and tune inner feedback loop first
- Conduct process response test(s) and tune the primary feedback controller



Feedforward Tuning Methodology (Cont'd)



- Recommended Procedure (Cont'd)
 - Continuously monitor and record the input disturbance variable (DV) and the primary feedback controlled variable (CV)
 - Put primary feedback controller influenced by input disturbance (feedforward) into Manual mode and allow the controlled variable to reach steady state
 - Manipulate the upstream variable that causes the input disturbance (e.g. vessel feed flow controller, level controller output, etc.) to create a series of input steps or pulses of varying magnitude and duration



Feedforward Tuning Methodology (Cont'd)



Recommended Procedure (Cont'd)

- Observe effect of disturbance on the primary CV and insure process response is in direction expected and magnitude of response is well above noise band
- Put primary feedback controller influenced by input disturbance (feedforward) back into Auto mode
- Allow primary controlled variable to reach steady state at same setpoint
- If more than one input disturbance variable (DV) influences the primary feedback controller, repeat this procedure for each DV





- Perform process response test results analysis for each DV
 - Using a tuning or model identification package [e.g., ExperTune, University of Connecticut's (UConn) Control Station, MathWorks MATLAB + System ID Toolbox, etc.]
- Estimate input disturbance process gain (including sign), deadtime, and first order time constant
- Use feedforward tuning constant rule set* or tuning and simulation package to obtain feedforward gain, lead, lag, and if req'd delay
- Commission and test combined feedforward-feedback loop

*The author's rule set follows in next slide



Feedforward Tuning Methodology (Cont'd)

Recommended Procedure (Cont'd)

- 1st pass feedforward tuning constant rule set
 - * Feedforward Gain = Load Disturbance Process Gain*/Controlled Variable Process Gain**
 - * Feedforward Lead = (1.3-1.5) x Controlled Variable 1st Order Process Time Constant**
 - Feedforward Lag = (1.1-1.3) x Load Disturbance 1st Order Process Time Constant*
 - Feedforward Delay = Load Disturbance Process Deadtime* - Controlled Variable Process Deadtime** (ignore if less than 0)

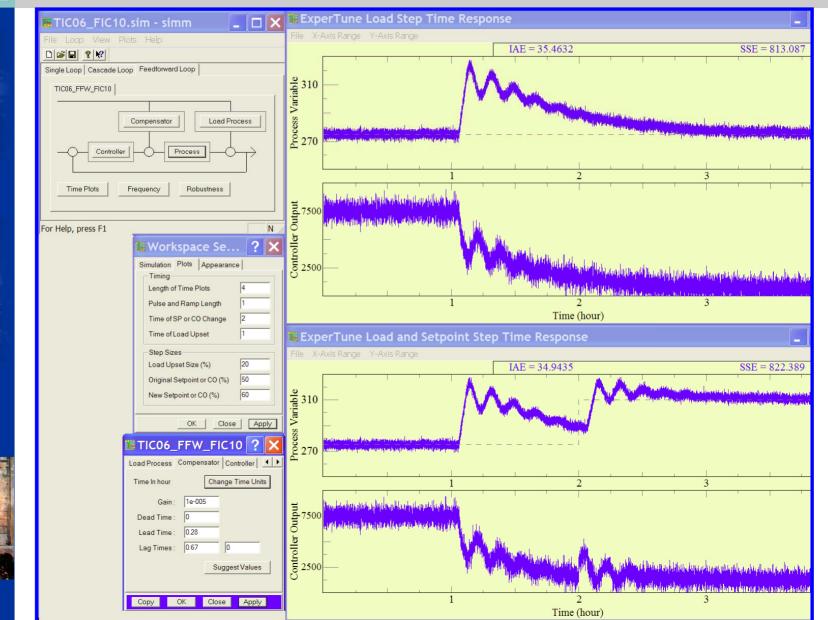
*Normalized effect of load disturbance variable (DV) change on primary process control var. (CV)

**Normalized effect of primary feedback controller manipulated variable (MV) move on primary process control var. (CV)

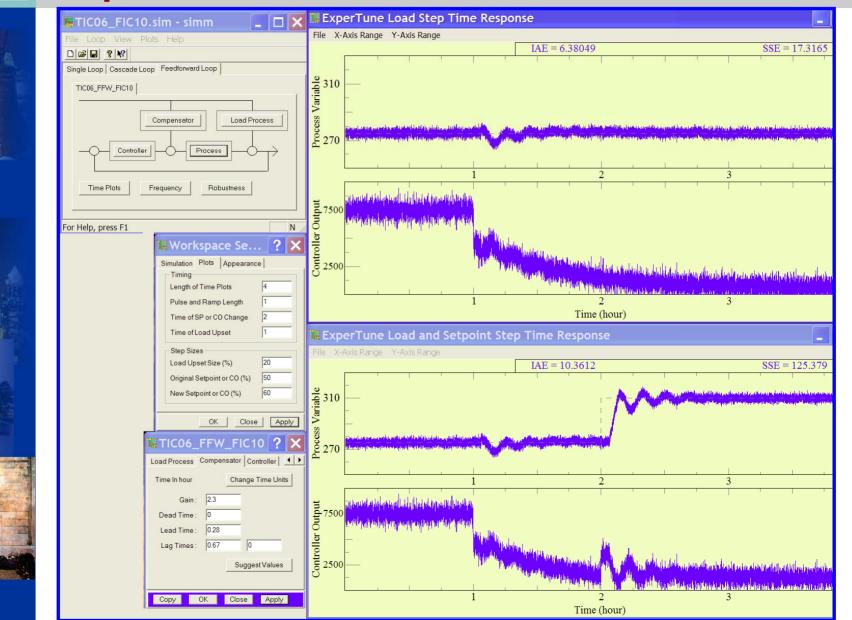
Simulated Feedback Only Temp. Ctl. Loop Performance – FIC10 20% Load Disturbance

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IND-PRO-OPTO Simulated Combined FFD-FDBK Temp. Ctl. Loop Performance – FIC10 20% Load Disturb. INDUSTRIAL PROCESS



Example 1: Distillation Col. Combined Feedforward-Feedback Control Results

- Adding three combined feedforwardfeedback control loops with dynamic compensation achieved:
 - Routine solvent purification column operation within environmental emissions constraints
 - Substantially reduced solvent loss
 - Estimated savings:
 - * ~ \$100K/year in solvent recovery
 - Unestimated \$/year in avoidance of environmental emissions excursion fines

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Model-Based Control



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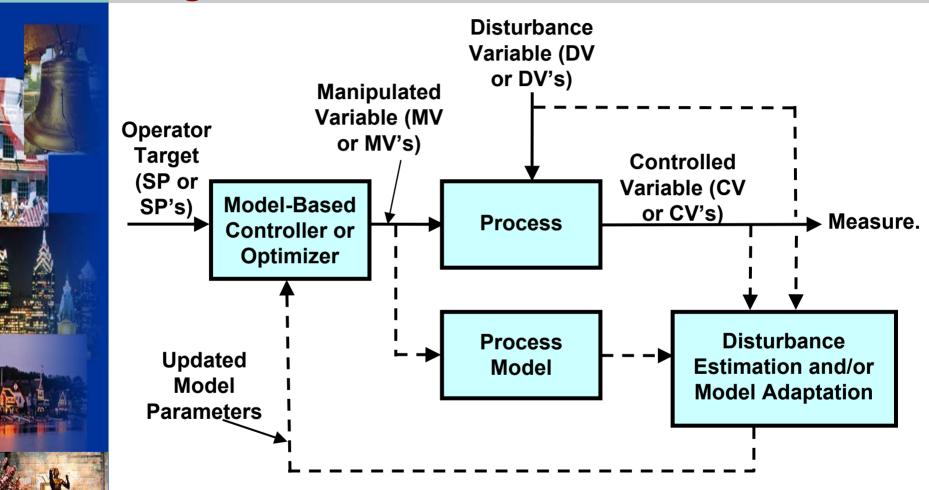
What is Model-Based Control?

 Embeds a process model in the control algorithm to better achieve the control objective

Some Model-Based Control Examples

- Internal Model Control
 Smith Predictor with PID Feedback Control
- Adaptive Model-Based Control Feedback Controller Tuning Constants Online Adjustment
- Adaptive Model-Based Control Process Model Parameters Online Adjustment
- Model Predictive Control
- Many Other Variants and Commercial Products...

Generic Model-Based Control Block Diagram

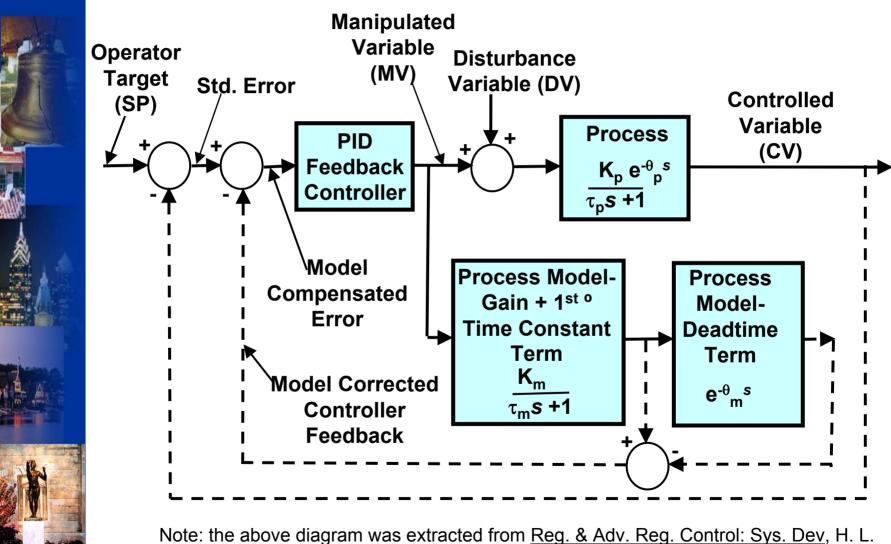


Note: the above diagram was extracted from <u>Techniques of Model-Based Control</u>, Brosilow & Joseph ©2002 Prentice-Hall, and was modified by the presenter.

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Smith Predictor with PID Feedback Control INDUSTRIAL PROCESS **Block Diagram**

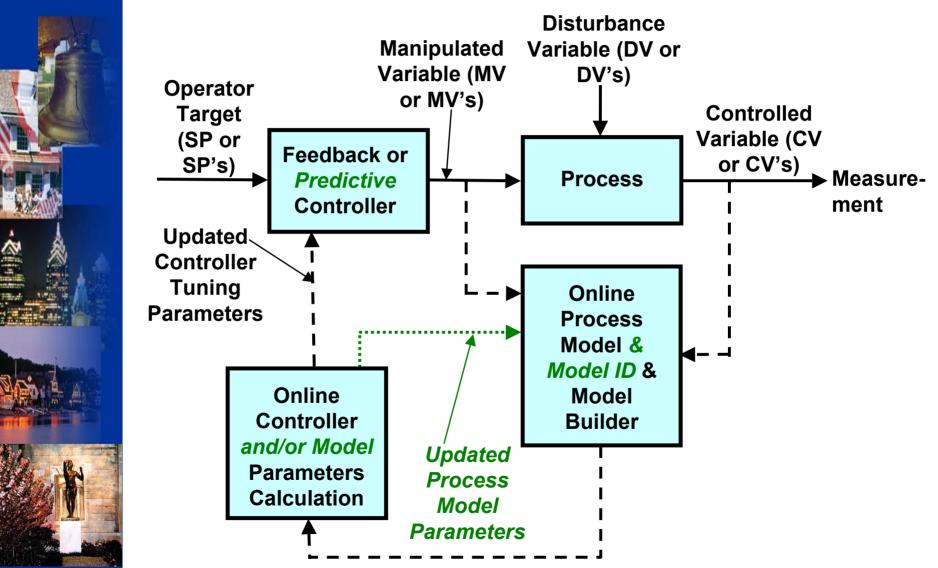


Wade ©1994 ISA and Fundamentals of Process Control Theory 3rd e., P. W. Merrill ©2000 ISA, and was modified by the presenter.

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Generic Adaptive Model-Based Control Block Diagram

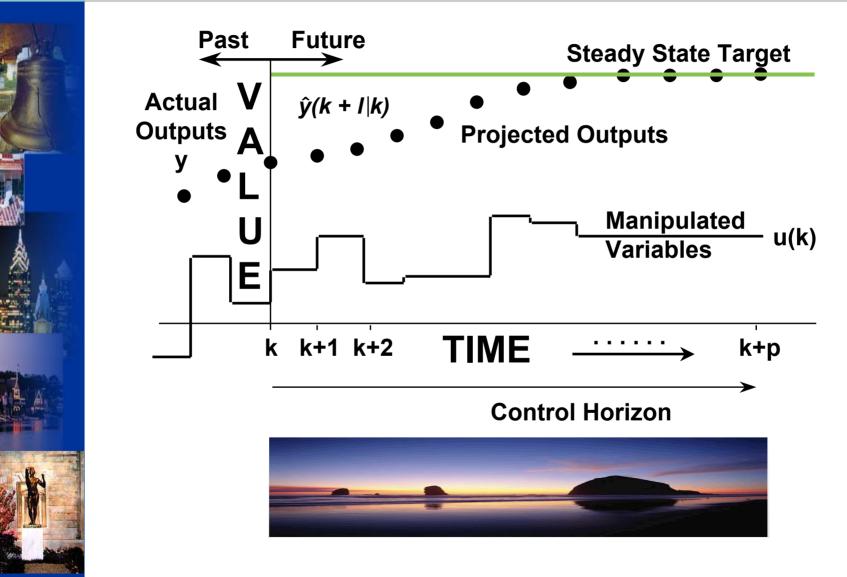


Note: descriptions in italics = capabilities of more sophisticated controllers

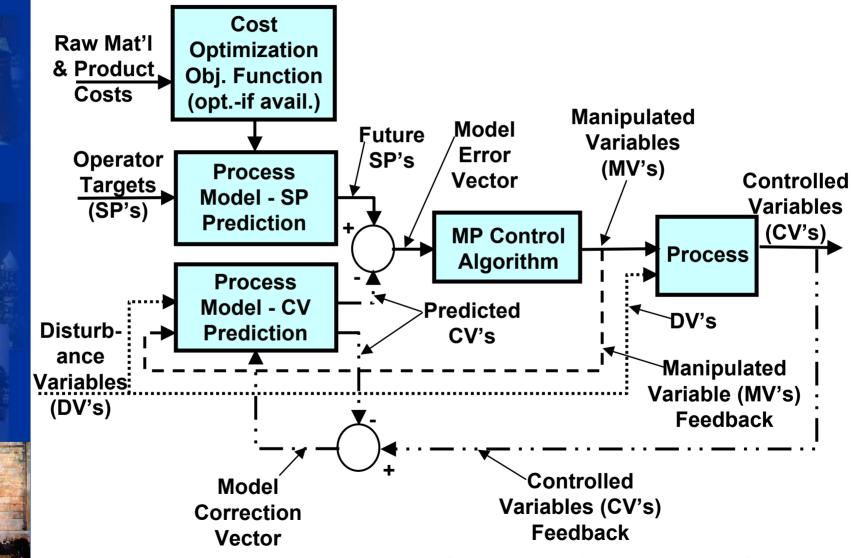
Model Predictive Control Definition and Features of Interest

- MPC- class of model-based control algorithms that compute a sequence of manipulated variable (MV) moves in order to optimize the future behavior of a plant
 - Solves control and optimization app mathematically online in real-time
 - Uses *linear* dynamic models to predict plant behavior (Feedforward)
 - Corrects for mismatch between actual plant behavior and model (Feedback)
 - May include operating constraints (constrained or unconstrained MPC)
 - May include dynamic cost optimization (objective) function





Generic Unconstrained Model Predictive Control Block Diagram



Note: the above diagram was extracted from <u>Advanced Control Unleashed</u>, G. K. McMillan et al ©2003 ISA, and was modified by the presenter.

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Model Predictive Control Advantages



- Controls complex multivariable processes and can handle:
 - Large plant controller-related variable set
 Typical example: 100 CV's/30 MV's/10 DV's
 - Interactive variable and integrating process dynamics
 - Long dead time and time constant processes
- Can (depending on ctlr cap.) perform real-time economic optimization
 - Optimum SP range vs. single SP for each CV
 - Includes raw material and product costs
- Achieves fastest plant production rate ramping



Model-Based & Adaptive Controller Suppliers

- Model-Based (Non-Adaptive) Controller Supplier
 - ControlSoft, Inc. MMC Modular Multivariable Ctlr (2/3/0/0)
 - http://www.controlsoftinc.com/mmc.shtml
- Adaptive Model-Based Controller Supplier
 - Universal Dynamics Technologies BrainWave[®] (1/1/3/0) + BrainWave[®] MultiMax (12/12/36/NA)

http://www.brainwave.com/product/product_index.html

- Adaptive "Model-Free" Controller Supplier
 - CyboSoft[™] (General Cybernation Group, Inc.)
 CyboCon (12/3/3/NA)
 - http://www.cybosoft.com/index.html

Note: (_/_/_) = controller capability for <u>CV's/MV's/DV's/AV's</u> (AV=constraint)



Model Predictive Controller Suppliers



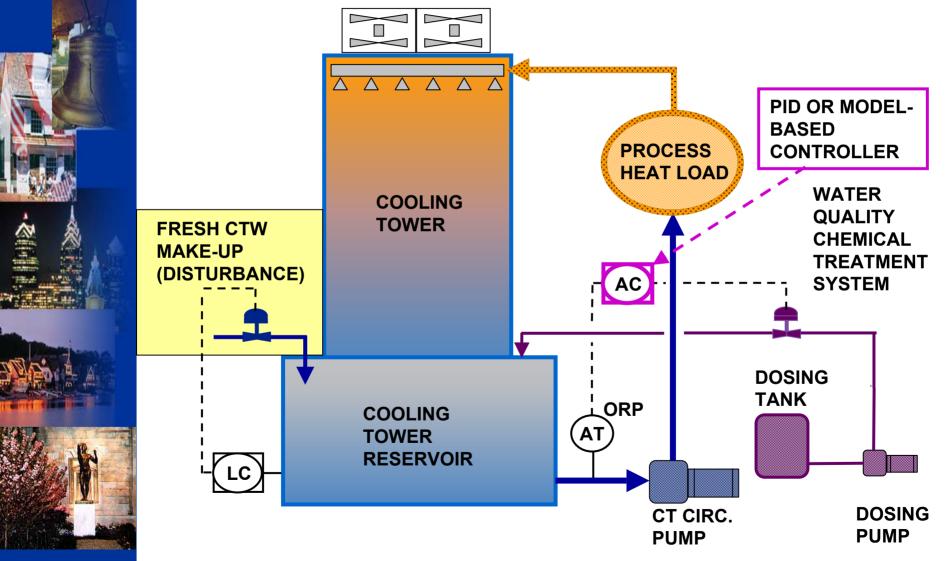
Model Predictive Controller Suppliers

- Aspen Technology DMCplus[®] (100's/100's/etc.)
 <u>http://www.aspentech.com/</u>
- Emerson Proc. Auto. DeltaV[™] Predict (4/4/4/4)
 - http://www.easydeltav.com/
- Honeywell Profit[®] Robust Multivariable Predictive Controller (200/100/100/300)
 - http://www.acs.honeywell.com/ichome/
- Intelligent Optimization GMAXC[™] (40/25/10/40)
 - http://www.intellopt.com/GMAXC.htm
 - 3rd party S/W solution from Siemens Energy and Automation Div. for APACS+ systems

(http://www.sea.siemens.com/process/default.html)

Note: (_/_/_) = controller capability for <u>CV's/MV's/DV's/AV's</u> (AV=constraint)

Example 2: Cooling Tower Water Quality Composition Control Process Schematic



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IND-PRO-OPTO **Example 2: Cooling Tower Water Quality** INDUSTRIAL PROCESS **Composition Control – Process Identification** OPTIMIZATION **Open Loop Process Response Test Results** 52 913 **PROCESS RESPONSE OR EXPERTUNE IDENTIFIED & VALIDATED REACTION CURVE PROCESS MODEL: UNDERDAMPED*** PV 47.34 (%) - ORP PROCESS **PROCESS WITH A SERIES** VAR.'s RESPONSE INTEGRATOR. **TO BELOW CO** FOLPDT PROCESS GAIN = 0.24 **STEP CHANGE PROCESS DEADTIME = 12 MINS** SERIES INTEGRATOR TIME = 5 MINS 1500 2000 2500 **CONTROL OUTPUT (CO) STEP CHANGE** CO 87.75 (OPEN LOOP) **CO IS SENT TO A DOSING CHEMICAL ADDITION CONTROL VALVE** POSITIONER 1000 **THENGR UNITS (%** \Rightarrow TIME UNITS (secs) of PV or CO range)

*The Laplace polynomial equation used to model and simulate this underdamped process in the ExperTune Loop Simulator is: 1/(C0 + C1s + C2s² + C3s³): C0=4.2; C1=92; C2=670; C3=0.

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Example 2: Cooling Tower Water Quality Composition Control – Process Model Dev. Simulated Process & Disturbance Models – Control Station

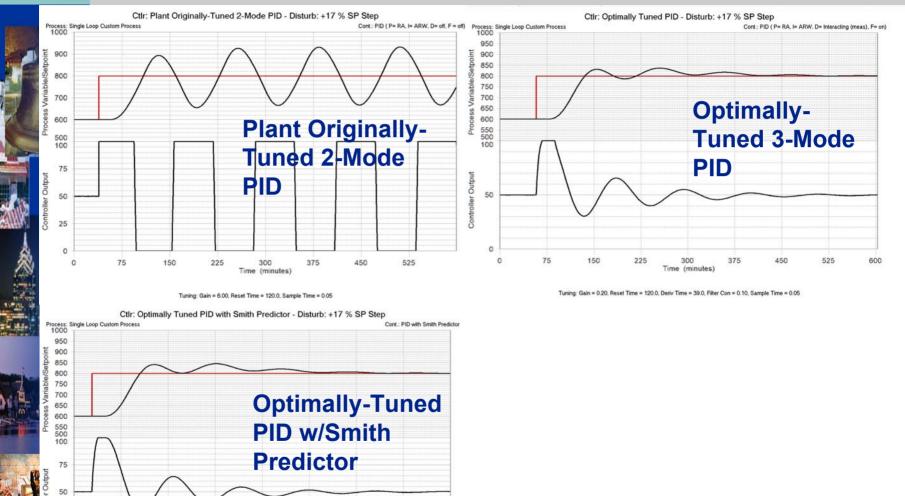
	🖬 Custom Process Input Form						m Process Input Form						
	Construct Process and Disturbance Models						Construct Process and Disturbance Models						
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	Underdamped Linear Mod	el	-										
1	Non-Self Regulating (Integr	rating) Process	•										
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17				General Model Form			Controller Out	put co	Flocess valiable,	F V	Disturbance, D		
					ττ≢ζ () =θαδ		Minimum	0.0	Minimum 0.0		Minimum 0.0		
IIII.	Integrator Gain, Κ [*] Natural Period,τ _{P0}		PV/(CO time)	$PV = \frac{1}{s(\tau_n^2 s^2)}$	$\frac{K_{p}^{*}(\tau_{p_{L}}s+1)e^{-\theta_{p_{S}}}}{^{2}+2(\tau_{s}s+1)(\tau_{p_{3}}s+1)}CO$		Maximum Startup Value	100.0	Maximum 12 Startup Value 60		Maximum 1200 Startup Value 600.0		
	Naturai Period, τ _{Ph} Damping Factor, ζ _P	12.63	time units time units	× ~	, , , , , , , , , , , , , , , , , , ,		Stanup value	150.0	Signific Agine [60]	J.U	Statup value 600.	J	
Â	Time Constant, t _P	0.0011	time units	Current Process Model									
	Lead Time, TPL	5.00	time units		0.12(.10.0)								
12	Dead Time, Op	12.0	time units	$PV = \frac{1}{s(159)!}$	0.13exp(-12.0s) 52s ² + 21.9s+1)(5.00s+1) CO								
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14	Overdamped Linear Mode	el	•										
- m				Laplace Domain	Time Domain								
				- ·	- Time Domain								
				General Model Form									
	Process Gain, K _P	1.50		DV -	$\frac{K_D(\tau_{DL} s + 1)e^{-\theta_D s}}{s + 1)(\tau_{D2} s + 1)(\tau_{D3} s + 1)} \mathbf{D}$								
518	First Time Constant, τ_{P1}	10.0	time units	$r v - \frac{1}{(\tau_{D1})}$	$(\tau_{D2} s + 1)(\tau_{D3} s + 1)$								
es P	Second Time Constant, τ_{P2}	2.00	time units										
	Third Time Constant, τ_{PS}	0.0	time units	Current Process Model									
	Lead Time, τ _{PL}	0.0	time units	D)(-	1.50exp(-13.0s) (10.0s+1)(2.00s+1) D								
	Dead Time,⊖p	13.0	time units	PV -	(10.0s+1)(2.00s+1)								
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Example 2: Cooling Tower Water Quality Composition Control – Controller Dev. Simulated PID & Model-Based Controller – Control Station

🖬 Controller Design 🛛 🗙	Controller Design							
Controller: 🔽 Advanced T Basic	Controller: Advanced Basic PID with Smith Predictor ?	Process Model Disturb Model						
Sample Time (minimum = 0.05) 0.05 time units	Sample Time (minimum = 0.05) 0.05 time units	Process Gain, K _P 0.24						
Set Point 600.0	Set Point 600.0	First Time Constant, T _{P1} 18.0 time units						
Bias (null value) 50.0	Bias (null ∨alue) 50.0	Second Time Constant, TP2 0.0 time units						
Adaptive PID: off	Adaptive PID: off	Lead Time, τ _{PL} 0.0 time units						
ON: Proportional - Reverse Acting, Kc > 0	ON: Proportional - Reverse Acting, Kc > 0	Dead Time, Op 19.0 time units						
Controller Gain, Kc 0.20	Controller Gain, Kc 0.20							
ON: Integral with Anti-Reset Windup	ON: Integral with Anti-Reset Windup							
Reset Time, τ_1 120.0 time units	Reset Time, τ_1 120.0 time units	$PV = \frac{K_{P} (\tau_{PLS} + 1) \exp(-\theta_{PS})}{(\tau_{P1S} + 1) (\tau_{P2S} + 1)} CO$						
ON: Interacting	ON: Interacting	$(\tau_{P1S}+1)$ $(\tau_{P2S}+1)$						
Derivative Time, $\tau_{\rm D}$ 39.0 time units	Derivative Time, τ_0 39.0 time units	Current Process Model						
Derivative computed on Measurement 💌	Derivative computed on Measurement	$PV = \frac{0.24exp(-19.0s)}{(18.0cs+1)}CO$						
Derivative Filter Constant α 0.10 ON 💌	Derivative Filter Constant, α 0.10 ON \checkmark	(18.0s+1)						
		Preview New Process Model						
Alarm: High 1188 Low 12.0	Alarm: High 1188 Low 12.0							
Done Cancel	Done	Cancel						

IND-PRO-OPTO **Example 2: Cooling Tower Water Quality** Comp. Ctl Simulated Perform. – SP Step INDUSTRIAL PROCESS



300 Time (minutes) 375

450

525

600

225

75

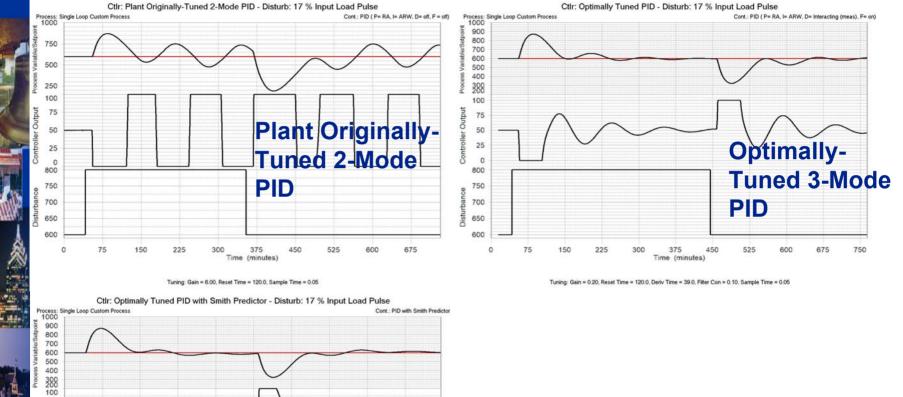
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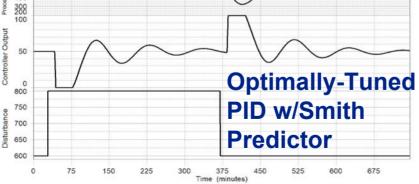
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0

Tuning: Gain = 0.20. Reset Time = 120.0. Deriv Time = 39.0. Filter Con = 0.10. Sample Time = 0.05 Process Model: Gain(Kp) = 0.24, T1 = 18.0, T2 = 0.0, TL = 0.0, TD = 19.0

IND-PRO-OPTO **Example 2: Cooling Tower Water Quality Comp. Ctl Simulated Perform. – Load Pulse** INDUSTRIAL PROCESS





Tuning: Gain = 0.20. Reset Time = 120.0. Deriv Time = 39.0. Filter Con = 0.10. Sample Time = 0.05 Process Model: Gain(Kp) = 0.24, T1 = 18.0, T2 = 0.0, TL = 0.0, TD = 19.0



Example 2: Cooling Tower Water Quality Composition Control Results

- Optimal PID Control Tuning Results:
 - Retuned with step and pulse response testing using ExperTune
 - Retuned loop reduced average ORP Controlled Var. (CV) variance from +/- 20-45% before retuning to ~ average of +/- 5% after retuning
 - Estimated savings:
 - Avoidance of need to shut down the plant and manually chemically clean heat exchange equipment ~ once/year - \$100K
 - \$5K per year in reduced microbiocide usage

Summary



- Advanced Regulatory Control globally proven to provide major competitive advantage if properly applied and maintained
 - Combined Feedforward-Feedback control can minimize the negative impact of routine disturbances for high profit control loops
 - Model-Based and Model Predictive Control can handle difficult or complex applications where single loop feedback control is not adequate
 - These techniques have been successfully applied to greatly improve plant performance for many decades
- Very capable commercial tools are now available to facilitate the process of moving beyond single loop control