Industrial Practice of Model Predictive Control*

Workshop at CPC-FOCAPO, Tucson
January 8th, 2017

Joseph Lu
Honeywell

* Primarily through the lens of Honeywell practices in the process industries
How are MPCs typically justified in process industries?

Model predictive control applications are typically financially evaluated and justified by its benefits on a per-application basis.

More often, MPC Control provides only the necessary feasibility condition for process optimization, whereas the latter provides the sufficient financial justification for its inception.

Typical Sources of Benefits:

- Increased Throughput
- Improved Yields
- Reduced Energy Consumptions
- Decreased Operating Costs (e.g. catalyst)
- Improved Product Quality Consistency
- Increased Operating Flexibility
- Improved Process Stability
- Improved Process Safety
- Improved Operator Effectiveness
- Better Process Information
MPC Applications and Benefits

- 5000+ of running Honeywell applications across many industries
- Industry-wide, the number could exceed 10,000+
- Generally accepted ranges for MPC and optimization benefits (per process unit)

<table>
<thead>
<tr>
<th>Industry</th>
<th>Benefits ($/yr)</th>
<th>Benefits (US$/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Petrochemicals</td>
<td>2-4% Increase in Production</td>
<td>3.4M - 5.3M US$</td>
</tr>
<tr>
<td>Ethylene</td>
<td>3-5% Increased Capacity / 1-4% Yield Improvement</td>
<td></td>
</tr>
<tr>
<td>VCM</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Aromatics (50 KBPD)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Chemicals</td>
<td>2-4% Increased Capacity / 2-5% Less Energy/Ton</td>
<td>2-5% Decrease in Operating Costs</td>
</tr>
<tr>
<td>Ammonia</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Polyolefins</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Industrial Utilities</td>
<td>10-20% Reduction in Chemical Usage</td>
<td>$1M-$2M</td>
</tr>
<tr>
<td>Cogeneration/Power Systems</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pulping</td>
<td>2-5% Increase in Production</td>
<td></td>
</tr>
<tr>
<td>Bleaching</td>
<td>2-5% Increase in Production</td>
<td></td>
</tr>
<tr>
<td>TMP (Thermo Mechanical Pulping)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Oil Refining</td>
<td>5-13</td>
<td>2.7M - 7.0M</td>
</tr>
<tr>
<td>Crude Distillation (150 KBPD)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Coking (40 KBPD)</td>
<td>15-33</td>
<td>2.2M - 4.8M</td>
</tr>
<tr>
<td>Hydrocracking (70 KBPD)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Catalytic Cracking (50 KBPD)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Reforming (50 KBPD)</td>
<td>13-30</td>
<td>3.3M - 7.6M</td>
</tr>
<tr>
<td>Alkylation (30 KBPD)</td>
<td>10-26</td>
<td>1.8M - 4.7M</td>
</tr>
<tr>
<td>Isomerization (30 KBPD)</td>
<td>10-26</td>
<td>1.1M - 2.8M</td>
</tr>
<tr>
<td></td>
<td>3-17</td>
<td>.3M - 1.8M</td>
</tr>
</tbody>
</table>
## Typical Payback Time by Industry

<table>
<thead>
<tr>
<th>Industry</th>
<th>Typical Payback Time</th>
</tr>
</thead>
<tbody>
<tr>
<td>Oil and Gas</td>
<td>1 to 2 months</td>
</tr>
<tr>
<td>Refining</td>
<td>3 to 6 months</td>
</tr>
<tr>
<td>Chemicals</td>
<td>3 to 6 months</td>
</tr>
<tr>
<td>Petrochem</td>
<td>4 to 6 months</td>
</tr>
<tr>
<td>MMM</td>
<td>4 to 6 months</td>
</tr>
<tr>
<td>Pulping/Paper</td>
<td>6 to 8 months</td>
</tr>
<tr>
<td>Industrial Power</td>
<td>10 to 12 months</td>
</tr>
</tbody>
</table>
Unit-Wide MPC Control Objectives

Three Common Objectives:

• Produce more product(s) of same (or better) quality
• Satisfy all safety/quality/operating constraints
• Maximize the product value and/or minimize operating cost

Products:

- Gas ($x_1$)
- Raw Gasoline ($x_2$)
- Naphtha ($x_3$)
- Kerosene ($x_4$)
- Light Gas Oil ($x_5$)
- Heavy Gas Oil ($x_6$)
- Residue ($x_7$)
Versatile control needs in MIMO applications

- Regulatory control
  - Reference tracking
  - Disturbance rejection / minimum variation
  - Traditional SISO control
- Constraint handling
  - Prevents another set of variables from exceeding their limits
- Transition control
  - Moves the operating point of a process unit from x to y
- Role of individual output can change
- Degrees of freedom

Model Predictive Range Control

- Output prediction: \( y = A\Delta u + \tilde{y} \)
- The set-point \( y = r \) is changed into a “set-range” \( y_{LO} \leq y \leq y_{HI} \)
- A slack variable \( s \) is introduced in the MPC formulation
- Additional degrees of freedom are used for
  - Local disturbance rejection/absorption
  - Optimization: e.g., \( \min y; \quad y_{LO} \leq y \leq y_{HI} \) (independent tuning vs. control response)

\[
\Delta u^* = \arg \min_{\Delta u} \|A\Delta u + \tilde{y} - r\|_Q^2 + \|\Delta u\|_R^2
\]

\[
\Delta u^* = \arg \min_{\Delta u,s} \|A\Delta u + \tilde{y} - s\|_Q^2 + \|\Delta u\|_R^2
\]

\( y_{LO} \leq s \leq y_{HI} \)
Range MPC Design and Solution

Performance Specification (via a funnel):

Range control benefits

- Robustness
  - Minimum effort control (when the control solution is non-unique)
  - Smoother control, minimum high frequency excitation

Range control algorithm

- Two-stage analytical approach
  a) solve the range control problem

\[
\Delta u^* = \arg \min_{\Delta u, s} \left| A \Delta u + \bar{y} - s \right|_Q^2 + \left\| \Delta u \right\|_R^2; \quad y_{LO} \leq s \leq y_{HI}
\]

b) If soln to (a) is not unique, solve for the minimum effort control

\[
\Delta u^* = \arg \min_{\Delta u} \left\| \Delta u \right\|_R^2; \quad y_{LO} \leq A \Delta u + \bar{y} \leq y_{HI}
\]

Online algorithm similar to the active-set QP: (here assume R = 0)

\[
\Delta u^* = \arg \min_{\Delta u, w} \left| A \Delta u - w \right|_Q^2; \quad y_{LO} - \bar{y} \leq w \leq y_{HI} - \bar{y}
\]

\[
\Delta u^* = \arg \min_{\Delta u, w} \left| A_{act} \Delta u - \begin{bmatrix} w_{act} \\ w_{free} \end{bmatrix} \right|_Q^2
\]

Let \( Q^{1/2} A_{act}^{(k)} = U_k R_k V_k^T \)

\[
\Delta u^* = \arg \min_{\Delta u} \left\| U_k R_k V_k^T \Delta u - w_{act} \right\|_2^2
\]
Non-Square Control Example—Solvent Dewaxing Unit
Solvent Dewaxing Unit Illustration

Tank 1: Mixture with solvent
Tank 2: Cooled mixture (wax slurry)
Tank 3: Chilled solvent (wax slurry)

➢ Solvent Inventory Control

Tank 4: Filtrate with wax removed
Tank 5: Wet solvent (oil removed)
Tank 6: Dried solvent for reuse
Model Dynamics of Solvent Control

From the application reported by Hall, et. al., 1995.

<table>
<thead>
<tr>
<th></th>
<th>Primary Filtrate Flow</th>
<th>Repulp Kickback Flow</th>
<th>Slop Solvent Flow</th>
<th>Slack Wax Flow</th>
<th>5th Dilution Flow Ratio</th>
</tr>
</thead>
<tbody>
<tr>
<td>Primary Filtrate Level</td>
<td>$-1.42e^{-3} \frac{1}{s}$</td>
<td>$1.42e^{-3} \frac{1}{s}$</td>
<td></td>
<td></td>
<td>$1.13 \frac{1}{s} e^{-10s}$</td>
</tr>
<tr>
<td>Repulp Filtrate Level</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>$-1.11 \frac{1}{s}$</td>
</tr>
<tr>
<td>Dry Solvent Drum Level</td>
<td>$1.6e^{-3} \frac{1}{s} e^{-5s}$</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Moist Solvent Drum Level</td>
<td>$1.62e^{-3} \frac{1}{s} e^{-5s}$</td>
<td></td>
<td>$2.16e^{-3} \frac{1}{s}$</td>
<td>$1.08e^{-3} \frac{1}{s} e^{-5s}$</td>
<td>$-0.173 \frac{1}{s}$</td>
</tr>
<tr>
<td>Slop Solvent Tank Level</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Slack Wax Tank Level</td>
<td></td>
<td></td>
<td>$-1.23e^{-3} \frac{1}{s}$</td>
<td></td>
<td>$1e^{-4} \frac{1}{s}$</td>
</tr>
</tbody>
</table>
Simulation Results

Controlled Variables

Manipulated Variables

Filter Wash
Grade Change
TMP refiner – a naturally non-square 4x5 interacting system

- Objective: consistent quality from coordinated control (decoupling)

Customer estimated benefits: ~$13.75M/Year
Ethylene Plant Overview

- Feed and products
  Ethylene plants are very flexible and can process a wide variety of feed stocks, ranging from gases (Ethane, Propane, LPG etc), naphthas (light and full range) to distillates and gas oils. Main products are polymer grade ethylene and propylene.

- Typical market driving forces
  Ethylene being a bulk commodity is sensitive to market conditions. Depending on the local economics, operations objectives are a combination of yield improvement, production maximization and energy intensity reduction.

- Operating flexibilities (variables for optimization)
  Main operating degrees of freedom include feed selection, furnace feed rates, cracking severity, dilution steam, CGC suction pressure, typical column variables (reflux, reboiler and pressure), converter temperature and H2 ratio and refrigeration compressor suction pressure.

- Complex chemical reactions and yield prediction
  Ethylene yield from the furnaces cannot be directly measured. Therefore we use a yield model to estimate the yields (& derivatives) for control and optimization. Technip’s on-line SPYRO, the industry standard, has been integrated into Honeywell’s UniSim Design as an extension Unit Op.

Process and Operating Characteristics

Universal:
- No product blending
- Stringent quality requirements
- Slow dynamics from gate to gate
- Gradual furnace and converter coking
- Furnace and converter decoking
- Frequent furnace switching

Site specific:
- Feed quality variations
- Product demand changes
- Sensitivity to ambient conditions
- Periodic switching (e.g., dryers)

Advanced Control and Optimization Goals

- Stabilize operation
- Minimize product quality give away
- Maximize selectivity & yield
- Minimize converter over-hydrogenation
- Minimize ethylene loss to methane/ethane recycle.

Fig 1. Aerial photo of an ethylene plant

Fig 2. Process overview – flow diagram
Profit Controller Envelopes

From Process Unit to Plantwide Control & Optimization
Typical Results of Ethylene Plant APC

• Execution Frequencies:
  - MPC Controllers: ½ to 1 minute
  - Global Optimizer: 1 minute
  - Gain updating: 5 minutes

• Typical Benefits
  - Production Increase: $1.5-3 Millions
  - Energy Savings: Additional

Implementation Details

➤ Design
Typical design comprises of individual Profit Controllers for each of the units, a Profit Optimizer on top of all the controllers, a UniSim model for inferentials and gain mapper, which is a Profit Bridge app that calls the UniSim model in real time to update the gains in both Profit Controllers and Profit Optimizer once every 5 mins.

➤ Modeling
Dynamic controller models are identified, in either open loop or closed loop, using Profit Stepper. Profit Optimizer model is automatically aggregated from all Profit Controllers. Bridge Models are identified from historical data (no step testing is required) for modeling the dynamic interactions among all MPC controllers. The UniSim modeling is fairly simple as only material balance calculations are required for gain updating.

➤ Commissioning
Profit Controllers are commissioned first to provide stable unit operations. Profit Bridge is then commissioned to provide gain updating for the furnace controllers. Next, Profit Optimizer is commissioned, and Bridge Model predictions are verified against data. Finally, a plantwide objective function is populated and optimization moves are validated. Continuous operation of the controllers and optimizer starts after operator training.

➤ Maintenance
Typically little maintenance is required to keep the ethylene APC/Optimization running. Clients either dedicate a parttime control engineer to monitor and perform minor services as needed, or depend on Honeywell’s BGmax quarterly-visit services.

Production is increased and product variability reduced...

Tighter control and less quality give-away...
Semiconductor Multi-factory Supply Chain

Blue = Intel
Red = Mat. Sub.
Green = Cap. Sub.

© 2021 Honeywell
Nontraditional Space: Discrete Manufacturing
Courtesy of Karl Kempf – Intel
Simple Supply Chain Sequence

TPT Distribution

Semiconductor Discrete manufacturing – Sorting Assembly

From Process Unit to Plantwide Control & Optimization
MPC Controller Design

✅ One controller per product

- 124 controlled variables
  - 5 fab/sort End-of-Line Inventories
  - 7 Assembly Die Inventories (ADI)
  - 7 die prep capacities (Work-in-Progress)
  - 56 Tape and Reel Die Inventories (TRDI)
  - 48 ADI + Die-Prep + TRDI inventories

- 90 manipulated variables (Green)
  - 35 ships from sorts
  - 48 ships from a hub TRDI to assembly test manufacturing (ATM) sites
  - 7 starts from ADI

- 293 feedforward variables (Blue)
  - 5 sort supply w/forecasts
  - 48 ATM TRDI demands w/forecasts
  - 240 TRDI cross-ships w/forecasts

✅ Current system design consists of 34 such controllers (products) per computer node

✅ One dynamic optimizer covers all 34 MPC controllers.

A model predictive control application for inventory and production management.
Chip Inventories at an Unnamed Site

Honeywell

Die xxx_06 Inventory

Die xxx_27 Inventory

Die xxx_47 Inventory
Refinery-wide (Production) Control/Optimization

Supply Side (upstream supply chain)

Average daily production

What Feed to buy?

Generate a month-plan based on an average model

Production Plan (monthly)

Customer Demand (downstream supply chain)

Average daily delivery

What products to sell?

Feed Received

What Feed to use?

Translate the plan into production activities to satisfy the plant logistics

Production Schedule (weekly)

Product to be Delivered

What products to deliver?

A set of production schedules (e.g., feed blending schedule)

Blending/shipment schedules

Economic Optimization
Multivariable Control

Human Coordination

Feed Received

Economic Optimization
Multivariable Control

Slave MPC 1

Slave MPC 2

Slave MPC n

Economic Optimization
Multivariable Control

Master MPC

A set of production schedules (e.g., feed blending schedule)

Raw Materials ➔ Execution ➔ Products ➔

Month ahead
days ahead
hours ahead
now
hours ahead
days ahead
months ahead

From Process Unit to Plantwide Control & Optimization

Page 21
1-to-n MPC cascade – One Master MPC and multiple Slave MPCs

- The Master MPC employs a dynamic model which can be, at user’s discretion, a coarser-scale model
  - e.g., similar to the scale-level of a planning model
- The Slave MPCs employ a dynamic model which can be a finer-scale one

- If Slave MPCs are deployed at every process unit, then the Master MPC can solve the plantwide production control problem dynamically and in closed loop
  - Currently, a steady-state version of this problem is controlled by planning solutions in open-loop
Refinery-wide Production Control: JIT manufacturing

- The intermediate tanks here are for illustration and may vary from plant to plant.
JIT manufacturing viewed as a control problem

Scope: the distillate production pool in an oil refinery
✓ It is simpler but still captures most of the important issues

Material Balance & Inventory Control
**Quality Control - Sulfur**

Legend:
- **Control handles**

- **Sweet**
  - 0.1% Sulfur

- **Medium**
  - 0.3% Sulfur

- **Sour**
  - 1.2% Sulfur

- **Vac Unit 1**

- **Vac Unit 2**

- **Vac Unit 3**

- **Hydrocracker**

- **Light Distillate Hydrotreater**

- **Heavy Distillate Hydrotreater**

- **Component Tank 1**

- **Component Tank 2**

- **Component Tank 3**

- **Imports**
  - 4~20 ppm S
  - 100~400 ppm S

- **Blending**
  - 1~2 ppm S
  - 100~600 ppm S
  - 12 ppm S
  - 10 ppm S

- **Crude Unit 1**

- **Crude Unit 2**

- **Crude Unit 3**

- **Distillate**
  - 4~20 ppm S
  - 50~300 ppm S

- **Sulfur**
  - Sweet: 0.1% Sulfur
  - Medium: 0.3% Sulfur
  - Sour: 1.2% Sulfur

- **Imports**
  - 4~20 ppm S
  - 100~400 ppm S

- **Hydrocracker**

- **Blending**
  - 1~2 ppm S
  - 100~600 ppm S

- **Imports**
  - 4~20 ppm S
  - 100~400 ppm S
🔍 Few if any (closed-loop) control solutions are available for this class of problems!
### Control Model Matrix for the Master MPC

1) Unit feed rates  
2) Control factors that affect the product quality  

<table>
<thead>
<tr>
<th>CVs</th>
<th>MVs</th>
<th>Crude1 Feed</th>
<th>Crude2 Feed</th>
<th>Crude3 Feed</th>
<th>Hydrocracker Conv</th>
<th>Cr Unit Cuts</th>
<th>Hydrotreater Sulfur</th>
<th>Product-i delivery</th>
</tr>
</thead>
<tbody>
<tr>
<td>CompTank2.Level</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CompTank2.Sulfur</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CompTank2.Cloud</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CompTank3.Level</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CompTank3.Sulfur</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CompTank3.Cloud</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>More CVs…</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

- Each \((i, j)\) entry contains a dynamic model \(g_{i,j}(s)\):  

- Often simple dynamics such as \(\frac{k}{as^2 + bs + 1}e^{-\theta s}\) or \(\frac{k}{s(\tau s + 1)}e^{-\theta s}\) would suffice  

- Notice that there is no setpoint in this control problem – just constraint control
Yes, if the solution for JIT manufacturing from the Master MPC is *implementable* by the Slave MPCs.

No, if the solution from the Master MPC can cause one or more Slave MPCs to wind up (i.e. sustained CV constraint violations).

The Real Challenge:

- How to get the Master to understand the Slave’s constraints well enough so that the Master’s solution will respect them
Master-Slave Models: a side-by-side view

- Below is an illustration of the model structure for 2 combined units:

- The relationship is a bit convoluted, so let’s first look at the simplest case: 1 conjoint variable per Slave MPC

- Different Tasks
  - Plantwide Production Control
    - Inventory control
    - Quality control
    - Cost and profit optimization
  - Unit Process Control
    - Safe/smooth operation
    - Unit efficiency
    - Intermediate Quality
    - Local Optimization

From Process Unit to Plantwide Control & Optimization
Steps in the hydrocracker example:

1) The HC feed is configured as MV3 in the Master.
2) Its current value is 33
3) The Master makes an inquiry call to the Slave to maximize the feed, subject to the Slave constraints
4) The call returns the following:
   a) the maximum feed = 38
   b) 4 active HC constraints are shown below, which are CV1, CV3, MV4 & MV5 (others not shown)
5) The master uses the maximum feed limit of 38 as a proxy limit for all Slave constraints.
Approximate the feasible region defined by the Slave constraints for 2 (or more) conjoint variables

\[
\begin{align*}
\text{Master} & \quad \text{CMV} \quad \text{MV1 Low Lmt} & \quad \text{MV1 High} \\
\text{Slave CV: } b_{m1}(1) &= 50 & \quad \text{Slave CV: } b_{m2}(1) &= 30 \\
\text{MV2 High} & & \quad \text{Slave CV: } b_{m3}(3) &= 100 \\
\text{Current Operating Point (0,0)} & & \quad \text{Free MVs} \\
\text{Conjoint MVs} & & \quad \text{Master MVs}
\end{align*}
\]

8 inquiry calls:

\[
\min_x c^t x_{\text{conj}}
\]

\[
b_{LO} \leq A_c x_{\text{conj}} + A_f x_{\text{free}} \leq b_{HI}
\]

\[
d^t x_{\text{conj}} = 0
\]

Where, \[ A = \begin{bmatrix} G \\ I \end{bmatrix} \]
Proxy Limits – Customer Example

To what extent is the Master MPC allowed to change the Slave unit’s operations?

Hydrocracker Conversion - Actual vs Maximum
Heavy Distillate HTR Feed Rate - Actual vs Maximum
Master’s Control and Optimization Results: Component Quality Changes in Tank #3

Higher cloud point means that more low-cost (heavier) components are used to produce the same products.

Higher 90pct point means that more low-cost (heavier) components are used to produce the same products.

Lower flash point means that more low-cost (lighter) components are used to produce the same products.

Sulfur is controlled at its high limit which means much lower catalyst/utility costs.

Component production costs are significantly reduced!
Feed Cost Saving from Optimizing the Volumetric Gains

Honeywell

Net Result: Crude Reduction

Incremental Changes in Crude Unit Feed Rate

Average crude saving is ~ 0.5% for meeting the same product demand. Huge cost saving!
Future Work

✓ A 3-level MPC Cascade for a larger part of the supply chain?
  ➢ For example: a combination of multiple refineries and fuel/crude depots
Closing Remarks

- MPC is the control solution of choice in the process industries
- Economic optimization is often applied in tandem with MPC
  - Close to 10 thousand MPC controllers have been implemented in different industries.
    - Oil refining, petrochemical, pulp & paper, coal-to-fuel, alumina, LNG, shale gas – just to name a few
  - >100 plantwide optimization solutions have been implemented
    - A majority of them are for ethylene plants
  - Estimated benefits delivered: $10 billions in total ($5 billions by Honeywell since 1996)

- Solution Operability and User Preference
  - Monolithic solutions are not desirable
  - Decentralized control – flexible to turn parts of control ON or OFF when addressing process upsets, maintenance issues, etc.
  - Gate-to-gate, plantwide optimization – greater benefits and no need to use any transfer prices

- Multiscale MPC cascade solution is being developed, which
  - Performs gate-to-gate, plantwide control and optimization
    - It can capture a significantly greater amount of benefits (~$1/barrel)
  - Leverages the existing planning model (structure)
  - Streamlines the current workflow from planning, scheduling to APC
Questions?