1. BACKGROUND
2. OBJECTIVE
3. DYNAMIC MODELS
   i) EMPIRICAL MODELLING
   ii) MECHANISTIC MODELLING
4. CONTROL STRATEGIES
   i) COUPLING VARIABLES VIA RGA ANALYSIS
   ii) CONVENTIONAL FEEDBACK CONTROL
   iii) ADVANCED MPC CONTROL
5. QUANTIFYING THE FINANCIAL BENEFITS
6. OPPORTUNITY AT HAND
• 19.4% net efficiency reduction due to PCC
• High capital cost
• Flexibility in operation

Motivation: A domain of operational uncertainties

- Carbon tax scheme started in July 2012, scrapped July 2014
- Variations in GHI, electricity price, electricity demand & carbon price

Hourly Global Horizontal Irradiation (GHI) in Sydney

Electricity demand and price fluctuations

Carbon market dynamics of Europe
Outlook

Carbon policy, ERF → Management Decision Support System

$ electricity → Economic Optimization

$ carbon → Economic Optimization

PP gross load → Economic Optimization

PP Control System

PCC Control System

PP Plant

PCC Plant

REGULATORY/POLICY LEVEL

ENTERPRISE LEVEL

PLANT LEVEL

INSTRUMENTATION LEVEL
Integration of PCC into coal-fired power plant requires understanding dynamic operations.

The PCC plant must respond flexibly to three significant scenarios:

1. Power plant operations at full and partial loads,
2. Under external disturbances from power plant and auxiliary systems, and

The objective of this study is to develop a dynamic model and use it in simulation analysis for techno economic study includes advanced control, optimization and management decision support system.
Significance and objective

Environmental objective:
CO₂ capture, CC% = 80 ~ 95%

Economic objective:
Energy Performance, EP = 3.5 ~ 4.5 MJ/kg CO₂
Modelling Approach 1: Empirical model

Tarong power station

Tarong PCC pilot plant
Modelling Approach 1: Empirical model

Tarong PCC pilot plant process flowsheet
Modelling Approach 1: Empirical model

Model boundaries using **NARX data-based model**

- **ABS**
  - Flue gas flow rate, \( u_1 \)
  - \( \text{CO}_2 \) concentration in flue gas, \( u_2 \)
  - Lean solvent flow rate, \( u_3 \)
  - Lean solvent temperature, \( u_4 \)

- **DES**
  - Rich solvent flow rate, \( u_5 \)
  - Rich solvent temperature, \( u_6 \)
  - Reboiler heat duty, \( u_7 \)
  - \( \text{CO}_2 \) concentration at top stripper, \( y_4 \)
  - Top stripper flow rate, \( y_5 \)
  - Rich solvent temperature, \( y_2 \)
  - Lean solvent temperature, \( y_3 \)

---


Modelling Approach 1: Empirical model

Integrated NARX data-based model
Modelling Approach 1: Empirical model

Model validation for NARX model

Pilot plant data operation:
\[ \approx 6 \text{ hours} \]

Sampling time: 10 sec

Data points:
\[ \approx 2000 \]

Estimate data points:
\[ \approx 1200 \]

Validation data points:
\[ \approx 800 \]
Step changes in reboiler heat duty

CO₂ concentration in top stripper, \( y_4 \) and top stripper flow rate, \( y_5 \) have significant open loop dynamic responses while CO₂ concentration in off gas, \( y_1 \) does not show any significant response.

Process time constants:
1) 6 – 15 mins for the fastest dynamics (reboiler heat duty – CO₂ concentration in top stripper relationship).
2) 8 -27 mins for the slowest dynamics (reboiler heat duty – top stripper flow rate relationship).

Step changes in Reboiler heat duty (straight line: base case; dotted line: positive step change; dashed line: negative step change).
Modelling Approach 2: Mechanistic model

Model boundaries using **mechanistic model** 3,4,5

Modified from ref. 3,4

Adopted from ref. 4

Adopted from ref. 5


Modelling Approach 2: Mechanistic model

Model validation for mechanistic model

Column temperature profiles from the simulation (line) and the pilot plant study (dot) for the pilot plant test no. 32\(^7\).

<table>
<thead>
<tr>
<th>Case number</th>
<th>Flue gas</th>
<th>Lean solvent</th>
<th>Rich solvent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature (K)</td>
<td>320</td>
<td>320</td>
<td>320</td>
</tr>
<tr>
<td>Flow rate (mol/s)</td>
<td>4.013</td>
<td>9.3</td>
<td>31.19</td>
</tr>
</tbody>
</table>

Column temperature profiles from the simulation (line) and the pilot plant study (dot) for the pilot plant test no. 47\(^7\).


Modelling Approach 2: Open-loop dynamic analysis

Step changes in reboiler heat duty

It can be seen that a 10% reduction (increment) in the heat duty caused the temperature to reduce by 1.4°C (increased by 0.2°C).

Process time constants:

1) 3 hours for the fastest dynamics at 10% increment of heat duty

2) 4.3 hours for the slowest at 10% reduction of heat duty

* High time constant: Due to a large amount (1.5 m³) of holdup solvent, the reboiler temperature inherited a high time constant
Control Approach

Identified 2 key performance metrics:

1. Carbon capture efficiency,
   \[ CC \text{ (\%)} = \frac{(y_4 / 100) y_5}{u_1 (u_2 / 100)} \]

2. Energy performance,
   \[ EP \text{ (MJ/kg)} = \frac{u_7}{(y_4 / 100) y_5} \]

Simplified 4 x 3 PCC system
Control Approach

Management Decision Support System

Economic-Optimization

PCC Control System

PCC Plant

Control Objective
CO\textsubscript{2} capture, 85% ≤ CC ≤ 95%
1. Energy performance, 3.5 ≤ EP ≤ 4.5 MJ/kg CO\textsubscript{2}

MV and CV selection
Sensitivity Analysis, RGA

PID control structure
Set tuning proportional (P), integral (I) and derivative (D) parameters via auto-tuning method

MPC control structure
Set tuning weights, control and prediction horizons

Performance evaluation
1. Stepwise change in the flue gas flow rate and CO\textsubscript{2} mole fraction in flue gas
2. Stepwise set point tracking of CC and EP within control objective
RGA results for different steady-state operating conditions

<table>
<thead>
<tr>
<th>Condition</th>
<th>Steady state input values</th>
<th>Final steady state output values</th>
<th>RGA*</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$u_1$ (kg/hr)</td>
<td>$u_2$ (mass %)</td>
<td>$u_3$ (L/min)</td>
</tr>
<tr>
<td>Condition 1</td>
<td>500</td>
<td>16</td>
<td>24</td>
</tr>
<tr>
<td>Condition 2</td>
<td>550</td>
<td>16</td>
<td>25</td>
</tr>
<tr>
<td>Condition 3</td>
<td>650</td>
<td>16</td>
<td>30</td>
</tr>
</tbody>
</table>

*RGA was performed by introducing +10% perturbation in $u_3$ and $u_7$.

Negative pairing = The control loop is unstable
Control Approach 1 – PID Controller

PID Control Scheme

PID Controller

CC set point

EP set point

PCC

u₃

u₇

CC

EP

PID 1

PID 2
Control Approach 2 – MPC Controller

MPC Control Scheme
Control Approach – Controllability Analysis

**Upstream disturbances**

- Flue gas flow rate, \( u_1 \)
- \( \text{CO}_2 \) concentration in flue gas, \( u_2 \)
- \( F_1 \) (mol/s)
- \( X_{\text{CO}_2} \)

- CO\(_2\) concentration at off gas, \( y_1 \)
- Lean solvent flow rate, \( u_3 \)
- Lean solvent temperature, \( u_4 \)

- Rich solvent flow rate, \( u_5 \)
- Rich solvent temperature, \( u_6 \)

- Rich solvent temperature, \( y_2 \)

- Lean solvent temperature, \( y_3 \)

- Reboiler heat duty, \( u_7 \)

- \( \text{CO}_2 \) concentration at top stripper, \( y_4 \)
- Top stripper flow rate, \( y_5 \)

**PCC control objective**

- CO\(_2\) capture, CC\% = 80 ~ 95%
- Energy Performance, EP = 3.5 ~ 4.5 MJ/kg CO\(_2\)

**ABS DES**

- Flue gas flow rate, \( u_1 \)
- \( \text{CO}_2 \) concentration in flue gas, \( u_2 \)
- \( F_1 \) (mol/s)
- \( X_{\text{CO}_2} \)

**DES**

- Rich solvent flow rate, \( u_5 \)
- Rich solvent temperature, \( u_6 \)
- \( \text{CO}_2 \) concentration at top stripper, \( y_4 \)
- Top stripper flow rate, \( y_5 \)

**HE**

- Lean solvent flow rate, \( u_3 \)
- Lean solvent temperature, \( u_4 \)

**CO\(_2\) capture, CC\% = 80 ~ 95%**

**Energy Performance, EP = 3.5 ~ 4.5 MJ/kg CO\(_2\)**
Control Approach – Controllability Analysis

Controllability analysis on set point changes and rejection disturbances.
Control Approach – Controllability Analysis

Control performance under process operational constraints

**PID Controller**

**MPC Controller**
Control-Optimization Approach

**Management Decision Support System**

**Economic-Optimization**

**PCC Control System**

**PCC Plant**

**PP + PCC flowsheet models**

**Perform N simulation case studies**

**Response Surface Modelling (MODDE package)**

A technical nonlinear prediction of the PCC process $Q_{reb} = f(X_i)$, $Aux = f(X_i)$

**Input real time-based power plant gross load (t)**

**Calculate CO$_2$ capture**

**Net load matched ?**

Y/N

**Profit maximized ?**

Y/N

**Constraints**

**Optimal values for CO$_2$ capture rate, CC$_{ideal}$**

**ENTERPRISE LEVEL**

**Technical study via NARX-MPC data-based model**

**PP match load**

**CC ideal**

**CC actual**

**Input real time-based electricity price (t)**

**Carbon price**

**Gross load (Fuel uptake)**

**Evaluation of techno-economic study based on PP-PCC profit**
Control-Optimization Approach

PP-PCC plant revenue

Revenue = \int P_e \ast (\text{Power plant net load} - \text{PCC penalty}) \ast dt - \\
\int C_t \ast \text{CO}_2 \text{ emitted} \ ast dt - \ P_{PP} - P_{PCC}

Rev-PP: Revenue generated through selling of electricity
A: Cost of CO2 emission
B: Power plant operating cost (PP-OPEX)
C: PCC operating cost (PCC-OPEX)
Techno-economic Approach – Optimization Control

Scenario:
Simulation period: 24 hrs
$\text{RRP} : 2011
CT: $25/\text{tonne-}CO_2

* 24 hours
Techno-economic Approach – Optimization Control

Scenario:
Simulation period: 24 hrs
$ RRP : 2020 (assuming 5% yearly increment from the base year 2008 )
CT: $ 25/ tonne-CO$_2$

* 24 hours
Revenue Composite

- **Rev-PP**: Revenue generated through selling of electricity
- **A**: Cost of CO$_2$ emission
- **B**: Power plant operating cost (PP-OPEX)
- **C**: PCC operating cost (PCC-OPEX)
ACKNOWLEDGMENT

The authors wish to acknowledge partial financial assistance provided through Australian National Low Emissions Coal Research and Development (ANLEC R&D). ANLEC R&D is supported by Australian Coal Association Low Emissions Technology Limited and the Australian Government through the Clean Energy Initiative.

Norhuda Abdul Manaf
The University of Sydney

Ashleigh Cousins & Paul Feron
CSIRO

THANK YOU!
Control Approach – Controllability Analysis

Control performance of $EP$ under process constraint ($T_r$)

The MPC-PCC did not violate the specified operational constraints for $T_r$ and $F_2$. However, the other two controllers were incapable to maintain respective process variables ($T_r$ and $F_2$) from violating its specified constraint.
Table 4.2 Operating conditions for case studies 32 and 47 (inputs to gPROMS simulations)

<table>
<thead>
<tr>
<th></th>
<th>Flue gas</th>
<th>Lean solvent</th>
<th>Rich solvent</th>
</tr>
</thead>
<tbody>
<tr>
<td>Case number</td>
<td>32 47</td>
<td>32 47</td>
<td>32 47</td>
</tr>
<tr>
<td>Temperature (K)</td>
<td>320 320</td>
<td>314 314</td>
<td>358 356</td>
</tr>
<tr>
<td>Mole fraction</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H₂O</td>
<td>0.025 0.032</td>
<td>0.86 0.846</td>
<td>0.846 0.828</td>
</tr>
<tr>
<td>MEA</td>
<td>- -</td>
<td>0.11 0.12</td>
<td>0.104 0.1181</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.175 0.167</td>
<td>0.029 0.034</td>
<td>0.05 0.0534</td>
</tr>
<tr>
<td>N₂</td>
<td>0.8 0.8</td>
<td>- -</td>
<td>- -</td>
</tr>
<tr>
<td>CO₂ loading</td>
<td>- -</td>
<td>0.264 0.28</td>
<td>0.48 0.46</td>
</tr>
<tr>
<td>L/G ratio</td>
<td>- -</td>
<td>6.5 4.6</td>
<td>- -</td>
</tr>
</tbody>
</table>