Thermal Integration of Waste to Energy Plants with Post-Combustion CO$_2$ Capture Technologies

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By

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Lay summary

This thesis investigates potential negative emission approaches for Waste to Energy (WtE) plants. In WtE plants, carbon dioxide (CO$_2$) is emitted as a component in the waste gases and can be divided into ‘biogenic’ and ‘direct fossil’ emissions. The biogenic CO$_2$ emissions can be regarded as carbon neutral, whereas the direct fossil CO$_2$ emissions from the stack of WtE plants contributes to the greenhouse gas emissions that are causing global climate change.

In this context, the implementation of Carbon Capture & Utilization/Storage (CCUS) for WtE plants has been identified as a CO$_2$ emissions mitigation pathway in different regional scenarios. By capturing and permanent storage of the biogenic CO$_2$, this technology also presents a particular opportunity of ‘negative emissions’ (i.e. reducing the total amount of CO$_2$ in the atmosphere).

This PhD research uses process modelling to optimise thermal integration of WtE plants with post-combustion capture of CO$_2$ (PCC). Three representative WtE plant configurations are identified for a state-of-the-art monoethanolamine (MEA) based PCC plant and analysed for 90%, 95% and 99.72% CO$_2$ capture rates. For each WtE plant configuration, two thermal integration scenarios are investigated namely: base case integration and advanced heat integration. Additionally, in order to further explore the optimal district heating supply options for WtE-CHP plant with PCC, for the first time, interim solvent storage (ISS) is applied to demonstrate how the flexible operation enabled by ISS allows WtE plants to meet energy commitments while improving operational profit and maximizing negative emissions.

Decisions on the integration of a CO$_2$ capture system with a WtE facility will depend on specific techno-economic assessment. In this study, a comprehensive set of key performance indicators (KPIs) characterising thermodynamic, economic and environment aspects are defined to give a thorough comparative assessment of all the investigated thermal integration approaches. They show that significant performance improvements can be made by advanced heat integration of WtE plant with PCC, by excess heat recovery from the PCC process. They also prove that optimised process design can be used to enable ultra-high CO$_2$ capture (99.72% in this study) with only a marginal increase in energy consumption and cost of CO$_2$ captured.

The findings of this PhD research provide valuable information for the future implementation of CCUS technology in the WtE sector.
Abstract

Building upon a series of process modelling activities, this thesis investigates approaches to optimize thermal integration of waste to energy (WtE) plants with 35% monoethanolamine based post-combustion capture (PCC). The basic heat integration is used as a benchmark case. In the advanced heat integration, excess heat from the direct contact cooler, CO₂ capture plant (stripper overhead condenser) and the CO₂ compression train are recovered using a heat pump, which represents the maximum amount of heat recoverable from the process.

A full set of key performance indicators has been tailored to evaluate the performance of WtE plants equipped with PCC from thermodynamic, economic, and environmental points of view. A set of coefficient performance metrics are introduced specific to combined heat and power (CHP) plants to evaluate the effectiveness of steam extraction for district heating and solvent regeneration. Results show that with basic heat integration, the back-pressure configuration with PCC (using a backpressure turbine in the steam cycle) has the highest coefficient performance, since there are no condensing heat losses in this configuration. With advanced heat recovery, more power can be generated from the lower pressure turbine, so that the coefficient performance for extraction & condensing configuration (using steam extraction followed by condensation) can be effectively increased to a higher level than that under back-pressure configuration, to a value of 8.6.

This work evaluates for first time the effect of ultra-high CO₂ capture levels on the performance of WtE plants using PCC. It shows that ultra-high CO₂ capture rates can be achieved with marginal increase in thermodynamic penalty and economic cost. The optimized condition of CO₂ capture rate is essentially a three-way trade-off among energy penalty, economic cost, and environment benefit.

To solve the competing heat requirements for fluctuating district heating supply and for solvent regeneration, this study investigates the use of interim solvent storage with PCC at a generic WtE CHP plant. Results show that solvent storage reduces the energy utilization penalty by 6% and the energy utilization factor penalty by 1%. The economic results are found to be sensitive to the assigned energy prices and CO₂ prices. Under scenarios of high gas-boiler heat prices and high fossil CO₂ emission prices, the solvent storage case presents comparable results as that of without solvent storage case. Effective oversizing of the stripper and compression train will be important to improve the economic performance of this application.

The methodology and findings from this study will be helpful for future researchers/plant operators/policy makers in decision-making on power and heat generation technologies with post-combustion CO₂ capture.
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Declaration of originality

The work included in this Ph.D. thesis, except where states otherwise by reference or acknowledgment, has been composed solely by the author under the guidance of her supervisor Professor Hannah Chalmers and Professor Mathieu Lucquiaud and that it has not been submitted, in whole or in part, in any previous application for a degree.

The author recommends referencing this thesis as follows:


Dan SU

29th June 2023
List of publications

The work described in this thesis has been reported in the following publications:

**Journal publications**


**Conference publications**


**Oral presentations**


**Poster presentation**


### Content

#### Table of Contents

<table>
<thead>
<tr>
<th>Section</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lay summary</td>
<td>ii</td>
</tr>
<tr>
<td>Abstract</td>
<td>iii</td>
</tr>
<tr>
<td>Acknowledgment</td>
<td>iv</td>
</tr>
<tr>
<td>Declaration of originality</td>
<td>v</td>
</tr>
<tr>
<td>List of publications</td>
<td>vi</td>
</tr>
<tr>
<td>Content</td>
<td>viii</td>
</tr>
<tr>
<td>List of figures</td>
<td>xi</td>
</tr>
<tr>
<td>List of tables</td>
<td>xvi</td>
</tr>
<tr>
<td>Nomenclature</td>
<td>xviii</td>
</tr>
<tr>
<td>List of symbols</td>
<td>xxi</td>
</tr>
<tr>
<td>1 Introduction</td>
<td>1</td>
</tr>
<tr>
<td>1.1 Motivations of waste to energy with carbon capture</td>
<td>1</td>
</tr>
<tr>
<td>1.2 Research objectives</td>
<td>2</td>
</tr>
<tr>
<td>1.3 Novelty of work</td>
<td>4</td>
</tr>
<tr>
<td>1.4 Overview of the methodology and thesis outline</td>
<td>5</td>
</tr>
<tr>
<td>2 Background</td>
<td>7</td>
</tr>
<tr>
<td>2.1 Waste to energy technology</td>
<td>7</td>
</tr>
<tr>
<td>2.1.1 WtE in the waste management hierarchy</td>
<td>7</td>
</tr>
<tr>
<td>2.1.2 Types of WtE technologies</td>
<td>9</td>
</tr>
<tr>
<td>2.2 Applicable CO₂ capture technologies for WtE plants</td>
<td>12</td>
</tr>
<tr>
<td>2.2.1 Amine-based chemical absorption</td>
<td>13</td>
</tr>
<tr>
<td>2.2.2 Next-generation capture technologies</td>
<td>17</td>
</tr>
<tr>
<td>2.3 Current development of WtE with CCS projects</td>
<td>20</td>
</tr>
<tr>
<td>2.4 Process modelling of CO₂ capture integration overview</td>
<td>28</td>
</tr>
<tr>
<td>2.4.1 Previous academic work on process modelling of PCC integration</td>
<td>28</td>
</tr>
<tr>
<td>2.4.2 Thermal integrations for WtE-CHP plants with CO₂ capture in the literature</td>
<td>30</td>
</tr>
<tr>
<td>3 Key performance indicators for the assessment of integrated systems</td>
<td>33</td>
</tr>
<tr>
<td>3.1 Thermodynamic assessment metrics</td>
<td>33</td>
</tr>
<tr>
<td>3.1.1 EP and EOP for WtE-power only plants with PCC</td>
<td>33</td>
</tr>
<tr>
<td>3.1.2 EUP and EUFP for WtE-CHP plants with PCC</td>
<td>34</td>
</tr>
<tr>
<td>3.1.3 Coefficient of Performance for (steam) extraction</td>
<td>36</td>
</tr>
<tr>
<td>3.2 Economic assessment metrics</td>
<td>41</td>
</tr>
</tbody>
</table>
3.2.1 Levelized cost of electricity (LCOE) ................................................................. 42
3.2.2 Levelized cost of heat (LCOH) ........................................................................ 43
3.2.3 Cost of CO₂ captured ....................................................................................... 44
3.3 Environmental assessment metrics ..................................................................... 45
3.4 Summary of KPIs for assessment of WtE with PCC ........................................ 47
4 Modelling methodology for waste to energy plants with Post-combustion CO₂ capture ............................................................................................................. 49
  4.1 Modelling approaches for typical WtE plants .................................................... 50
    4.1.1 Basis of boiler balance calculation ............................................................... 54
    4.1.2 Investigated steam cycle configurations of WtE plants............................... 63
    4.1.3 Validation of the WtE plant modelling in gProcess ....................................... 65
  4.2 Capture plant modelling in ASPEN .................................................................... 66
    4.2.1 Description of post-combustion capture ....................................................... 65
    4.2.2 Capture plant optimization ......................................................................... 68
    4.2.3 Validation of PCC plant modelling in ASPEN .............................................. 71
  4.3 Approaches for PCC plant integrated in the WtE plants .................................... 71
5 Cost estimation ....................................................................................................... 72
  5.1 Capex and Opex of WtE plants (cost items 1.1, 1.2, 2.1, 2.2) ......................... 75
  5.2 Capex and Opex of CO₂ capture and compression plants (cost items 1.3, 1.4, 1.5, 2.3, 2.4, 2.5) .............................................................................................................. 77
    5.2.1 Capex estimation of PCC the plant ............................................................... 78
    5.2.2 Opex estimation of PCC the plant ............................................................... 81
  5.3 Capex and Opex of heat exchanger networks (cost item 1.6, 1.7, 1.8, 2.6, 2.7, 2.8) ............................................................................................................................... 84
    5.3.1 Capex of heat exchanger ............................................................................. 84
    5.3.2 Hot water piping cost ................................................................................ 89
  5.4 Additional cost for solvent storage scenarios (cost item 1.9, 2.9) .................... 91
  5.5 Summary of cost estimation ............................................................................. 91
6 Base case integration of WtE plants with MEA based post-combustion CO₂ capture ...................................................................................................................... 93
  6.1 Base case integration of WtE plants with PCC ................................................... 93
  6.2 Key results from the WtE plant modelling in gProcess ....................................... 94
  6.3 Key results from the CO₂ capture plant modelling in ASPEN ......................... 99
    6.3.1 Absorber packing design and optimisation ................................................. 99
    6.3.2 Effect of lean solvent loadings .................................................................... 101
    6.3.3 Effect of absorber intercooling .................................................................. 102
    6.3.4 Effect of stripper pressure ........................................................................ 103
    6.3.5 Summary of PCC plant modelling results .................................................. 105
  6.4 KPI results for the base case integration ............................................................ 107
<table>
<thead>
<tr>
<th>Section</th>
<th>Title</th>
<th>Page</th>
</tr>
</thead>
<tbody>
<tr>
<td>6.4.1</td>
<td>Thermodynamic performance under the base case integration</td>
<td>107</td>
</tr>
<tr>
<td>6.4.2</td>
<td>Economic performance under the base case integration</td>
<td>114</td>
</tr>
<tr>
<td>6.4.3</td>
<td>Environmental performance under the base case integration</td>
<td>121</td>
</tr>
<tr>
<td>6.4.4</td>
<td>Summary of KPI results under the base case integration</td>
<td>125</td>
</tr>
<tr>
<td>7</td>
<td>Advanced heat recovery for CHP application</td>
<td>127</td>
</tr>
<tr>
<td>7.1</td>
<td>Advanced heat integration of WtE-CHP plants with PCC</td>
<td>127</td>
</tr>
<tr>
<td>7.2</td>
<td>KPI results for advanced heat recovery for CHP application</td>
<td>131</td>
</tr>
<tr>
<td>7.2.1</td>
<td>Thermodynamic performance under advanced heat integration</td>
<td>131</td>
</tr>
<tr>
<td>7.2.2</td>
<td>Economic performance under advanced heat integration</td>
<td>138</td>
</tr>
<tr>
<td>7.2.3</td>
<td>Environmental performance under advanced heat integration</td>
<td>144</td>
</tr>
<tr>
<td>7.2.4</td>
<td>Summary of KPI results for the advanced heat integration</td>
<td>146</td>
</tr>
<tr>
<td>8</td>
<td>Interim solvent storage to meet fluctuating DH demand profiles</td>
<td>148</td>
</tr>
<tr>
<td>8.1</td>
<td>Concept of interim solvent storage (ISS)</td>
<td>148</td>
</tr>
<tr>
<td>8.2</td>
<td>Fluctuating DH demand profiles</td>
<td>149</td>
</tr>
<tr>
<td>8.3</td>
<td>Operation constraints and logic of ISS application</td>
<td>151</td>
</tr>
<tr>
<td>8.4</td>
<td>Evaluation of the annual profit of a WtE-CHP plant with PCC and ISS case</td>
<td>160</td>
</tr>
<tr>
<td>8.5</td>
<td>ISS application results and sensitivity analysis</td>
<td>163</td>
</tr>
<tr>
<td>8.5.1</td>
<td>Sensitivity of gas-boiler heat prices</td>
<td>164</td>
</tr>
<tr>
<td>8.5.2</td>
<td>Sensitivity of storage tank sizes</td>
<td>166</td>
</tr>
<tr>
<td>8.5.3</td>
<td>Sensitivity of CO\textsubscript{2} prices</td>
<td>167</td>
</tr>
<tr>
<td>8.6</td>
<td>KPI results for seasonal solvent storage application for CHP application</td>
<td>168</td>
</tr>
<tr>
<td>8.6.1</td>
<td>Thermodynamic performance under ISS application</td>
<td>169</td>
</tr>
<tr>
<td>8.6.2</td>
<td>Economic performance under ISS application</td>
<td>171</td>
</tr>
<tr>
<td>8.6.3</td>
<td>Environmental performance under ISS application</td>
<td>173</td>
</tr>
<tr>
<td>8.6.4</td>
<td>Summary of key findings under ISS application</td>
<td>175</td>
</tr>
<tr>
<td>9</td>
<td>Conclusion</td>
<td>176</td>
</tr>
<tr>
<td>9.1</td>
<td>Summary of findings</td>
<td>176</td>
</tr>
<tr>
<td>9.2</td>
<td>Limitations and recommendations for future work</td>
<td>179</td>
</tr>
<tr>
<td>9.3</td>
<td>Contribution to knowledge</td>
<td>182</td>
</tr>
<tr>
<td>10</td>
<td>References</td>
<td>184</td>
</tr>
<tr>
<td>11</td>
<td>Appendix</td>
<td>204</td>
</tr>
</tbody>
</table>
List of figures

Figure 1-1 Overview of thesis outline ........................................................................................................ 5
Figure 2-1 The EU waste hierarchy from the EU Waste Framework Directive (WFD)(European Parliament and Council, 2018) ........................................................................................................ 8
Figure 2-2 Classification of WtE technologies based on energy conversion pathways ....................... 9
Figure 2-3 Process flow diagram of the Amager Bakke waste-to-energy facility (showing just one of two boilers) (Hulgaard & Søndergaard, 2018) ........................................................................................................ 10
Figure 2-4 Classification of WtE technologies based on applications ................................................... 11
Figure 2-5 Conventional aqueous amine solvent plant process flowsheet integrated with process optimisation using the intercooling and rich split processes (Global CCS Institute, 2021) .................. 13
Figure 2-6 Next-generation capture technologies being tested at 0.5 MWe (10 t/d) scale or larger with actual flue gas, selected by Global CCS Institute (Global CCS Institute, 2021) .................. 18
Figure 2-7 Opportunities and challenges of potential next-generation CO2 capture technologies for WtE plants (AECOM, 2021; Capsol, 2023; Gibbins & Lucquiaud, 2021; IEAGHG, 2020) ................ 19
Figure 3-1 EOP, EP, EUP, EUFP due to PCC which are related with fuel specific emissions ............... 36
Figure 3-2 Schematic illustration of COPX_cap for WtE-power-only plant with PCC and the analogous comparison with the COPhp of an electricity-driven heat pump consuming the same amount of power WT (lost of electricity output due to steam extraction) ............................................................ 38
Figure 3-3 Schematic illustration of COPX_cap + dh for WtE-CHP plant with PCC and the analogous comparison with the COPhp of an electricity-driven heat pump consuming the same amount of power WT (lost of electricity output due to steam extraction) ............................................................ 40
Figure 3-4 Schematic illustration of COPX_abs for WtE-CHP plant with PCC, with the series application of electricity-driven heat pump and absorption heat pump, analogous consuming the lost power output WT and lost district heat output Qdh_loss ............................................................. 41
Figure 4-1 Summarises the process modelling approaches ........................................................................ 50
Figure 4-2 Schematic layout of the referenced WtE plant ........................................................................ 52
Figure 4-3 Schematic representation of heat balance of a boiler (Prabir Basu et al., 2000) ............... 55
Figure 4-4 Radiative and convective heat losses from the external surface of typical pulverized coal fired boilers (Wu, 2008) ......................................................................................................................... 61
Figure 4-5 Schematic representation of investigated configurations of WtE plants ............................... 65
Figure 4-6 Process flow diagram for the conventional MEA based chemical absorption and desorption process (IEAGHG, 2014) ........................................................................................................ 67
Figure 4-7 PCC plant modelling and optimization approach in ASPEN plus ........................................ 68
Figure 5-1 Structure of cost estimation for the three integration scenarios included in this study, with cost items applied for each application ............................................................. 73
Figure 5-2 Overall U-values for different process/service fluid combinations in shell and tube heat exchangers (Sinnott & Towler, 2020). ............................................................. 86
Figure 6-1 Process flow diagram of a WtE CHP plant with CO₂ capture and compression .......... 94
Figure 6-2 Heat output by burning 1kg fuel, based on process modelling of a generic WtE plant ...... 97
Figure 6-3 Change of boiler efficiency with exhaust flue gas temperature, results based on the process modelling of a generic WtE plant .............................................. 98
Figure 6-4 Sensitivity of the absorber packing height on the rich solvent CO₂ loading (continuous line) and the specific reboiler duty (dashed lines) for a 35%wt MEA capture system at a range of CO₂ capture rates: 90%, 95%, 99% and 99.72% CO₂ capture rates ...................................................................... 100
Figure 6-5 Sensitivity of the specific reboiler duty to the lean solvent CO₂ loading for a 35%wt MEA capture system for a range of CO₂ capture efficiencies 95%/99%/99.7%. The absorber packing heights are 17m/20m/22m for 95%/99%/99.7% capture efficiencies, respectively. For illustration purposes, the final packing height after optimisation might be different ........................................ 102
Figure 6-6 Effect of absorber intercooling temperature on rich solvent loading and absorber packing height for achieving 99.72% CO₂ capture rate with lean loading of 0.16 mol CO₂/mol MEA. .......... 103
Figure 6-7 Effect of stripper pressure on the specific reboiler duty for a range of lean solvent loadings. Under constant rich loading of 0.45 mol CO₂/mol MEA for 99.72% CO₂ capture rate. ...... 105
Figure 6-8 EOP and EUP of PCC with WtE plants under three CO₂ capture rates ......................... 108
Figure 6-9 EP and EUFP of PCC with WtE plants under three CO₂ capture rates ....................... 111
Figure 6-10 Coefficient of Performance for (steam) extraction of PCC with WtE plants under three CO₂ capture rates .............................................................. 112
Figure 6-11 Break down of LCOE of WtE plant without and with PCC under three CO₂ capture rates, with assumptions of fossil CO₂ price £40/tCO₂, gate fee £100/tMSW, no negative emission credit is considered ........................................................................................................ 116
Figure 6-12 LCOE of WtE-Power-only plant under different CO₂ prices ...................................... 117
Figure 6-13 Breakdown of LCOH of WtE-CHP plant without and with PCC under three CO₂ capture rates, with assumption of electricity selling price £85/MWh, fossil CO₂ price £40/tCO₂ .......... 118
Figure 6-14 Cost of CO₂ captured £/tCO₂ for the investigated WtE plants with PCC ................... 119
Figure 6-15 Illustration of sensitivity of DH period on the cost of CO₂ captured, with assumption of fossil CO₂ price £40/tCO₂; for WtE-CHP plants, the electricity selling price is defined to be £85/MWh. ......................................................................................... 121
Figure 6-16 Carbon emission intensity of WtE plant without and with PCC on fuel basis ........... 122
Figure 8-4 Net power output of WtE-CHP plant with PCC in ISS application, with assumptions that maximum steam is extracted from the steam cycle for DH and for solvent regeneration .............. 153
Figure 8-5 Oversizing stripper diameter to reach 155% base case capacity .............................................. 155
Figure 8-6 Illustrative operation modes of WtE-CHP plant with PCC, ISS-SS mode, ISS-AR mode and w/o ISS mode ............................................................................................................................ 158
Figure 8-7 Operation logic of WtE-CHP plant integration with PCC under ISS applications ............... 159
Figure 8-8 Comparison of capital cost of PCC plant under scenarios with and without ISS applications ........................................................................................................................................... 164

Figure 8-9 Sensitivity of the annual profit to the solvent storage tanks size for a range of gas boiler heat purchasing prices, under scenarios w/ and w/o Negative emission credit of £75/tCO$_2$. Electricity and heat selling prices from the WtE plant are £85/MWh and £79/MWh, respectively; heat tariff from WtE plant £46/MWh; fossil CO$_2$ price £75/tCO$_2$; gate fee £100/tMSW; 1 hour of storage requires the tank capacity of 247m$^3$ .................................................................................................................... 165

In Figure 8-10, the initial increase of profit is mainly due to the storage capacity of the tanks, which increases the energy output of the plant; the decrease of profit is due to the size of the plant doesn’t cope with the energy demand curve and capacity (as suspected by the author). This suggests that there is an optimized size of the storage tanks. This curve only explains the situation for this study, e.g. the DH demand curve, the limitations set to constrain the operation of stripper, steam cycle, etc. ........................................................................................................................................ 166

Figure 8-11 Comparison of heat supply and solvent levels in tanks in Spring day 21$^{st}$ March, under three solvent tank sizes - 1 hour storage, 2 hours storage, 3 hours storage. The solvent in lean tank operate in opposite trend as that shown in the rich tank, to maintain mass balance in the system. 167

Figure 8-12 Sensitivity of CO$_2$ prices on the annual profit of WtE-CHP plant under scenarios 1) without PCC, 2) with PCC but without ISS, and 3) with PCC and ISS. Assumptions: Electricity and heat selling prices from the WtE plant are £85/MWh and £79/MWh, respectively; renewable heat incentive £46/MWh; gas-boiler heat price £108/MWh; same fossil CO$_2$ price and NEC; gate fee £100/tMSW; 2 hour of storage tank capacity) ........................................................................................................................................ 168

Figure 8-13 Energy utilization penalty (EUP) and Energy utilization factor penalty (EUFP) for a WtE-CHP-PCC plant with and without ISS ........................................................................................................................................ 171

Figure 8-14 LCOH (£/MWh) under WtE-CHP 1) without PCC, 2) with PCC and without ISS and 3) with PCC and ISS. (Electricity selling prices from the WtE plant of £85/MWh, fossil CO$_2$ price and negative emission credit £62.1/tCO$_2$; gate fee £100/tMSW; 2 hour of storage tank size under ISS) .............. 172

Figure 8-15 Cost of CO$_2$ captured £/tCO$_2$ under WtE-CHP plant 1) without PCC, 2) with PCC and without ISS and 3) with PCC and ISS. (Electricity selling prices from the WtE plant of £85/MWh, fossil
CO2 price and negative emission credit £62.1/tCO$_2$, gate fee £100/tMSW; 2 hours of storage tank size under ISS).......................................................................................................................................................173

Figure 8-16 CO$_2$ emission intensity on heat basis and electricity basis under WtE-CHP plant 1) without PCC, 2) with PCC and without ISS and 3) with PCC and ISS.......................................................................................................................................................174
List of tables

Table 2-1 On-going and upcoming and WtE with CCUS projects worldwide ........................................ 22
Table 2-2 A summary of the main modelling software/CO₂ capture rate/solvent for PCC with Power plant integration .............................................................................................................. 29
Table 4-1 MSW Ultimate composition (as received) based on sampling in March 2019 (through personal communication with FCC) ............................................................................. 54
Table 4-2 Heat loss owing to convection and radiation from the furnace exterior for steam generation boiler ........................................................................................................................................ 61
Table 4-3 Specific heat capacity of the three key materials in the slag, @700°C ........................................ 62
Table 4-4 key values comparison of operational WtE plant in Edinburgh and modelling data .......... 65
Table 5-1 List of cost items as shown in Figure 5-1. ............................................................................. 74
Table 5-2 CEPCI numbers and exchange rate applied when calculate the Capex and Opex of WtE plant ........................................................................................................................................... 75
Table 5-3 Key parameters of Capex of WtE plant without PCC (Cost items 1.1 and 1.2) ................. 76
Table 5-4 Key parameters of Opex of WtE plant without PCC (Cost items 2.1 and 2.2) ................. 77
Table 5-5 Key input parameters and multipliers used to calculate the total Capex of the PCC plant (IEAGHG, 2017b) ........................................................................................................ 80
Table 5-6 Key input data for estimation of the annual fixed Opex and annual variable Opex of the PCC plant ........................................................................................................................................ 82
Table 5-7 Material, Pressure and Temperature factors for heat exchangers according to (Smith, 2005). ............................................................................................................................. 87
Table 5-8 Typical correction factors for capital cost based on delivered equipment costs (Smith, 2005). ............................................................................................................................. 88
Table 5-9 Roughness coefficient c values for Hazen-Williams equation (TORO, n.d.) ................. 90
Table 6-1 Key performance data of WtE modelling ............................................................................. 95
Table 6-2 Flue gas inlet conditions of the PCC system ...................................................................... 98
Table 6-3 Design and performance parameters of the PCC plant at three CO₂ capture levels ...... 106
Table 6-4 Power and DH heat output of the WtE-CHP plants with and without PCC ...................... 109
Table 6-5 Energy utilization factor of the WtE-CHP plants with and without PCC ....................... 111
Table 6-6 COPX_abs of steam extraction for WtE-CHP plants with PCC ........................................ 113
Table 6-7 Cost estimation for base case thermal integration of WtE with PCC .............................. 114
Table 7-1 COPX_abs of steam extraction for WtE-CHP plants with PCC under advanced heat integrations ............................................................................................................................... 136
Table 7-2 Cost estimation for advanced heat integration cases of WtE plants with PCC ...............138
Table 8-1 Seasonal duration based on astronomical events (NPL, 2023) ..............................................150
Table 8-2 Referenced cost of the amine storage tanks ..................................................................................162
Table 8-3 Currency conversion applied under circumstance that different currencies used in the reporting data ..............................................................................................................................................162
Table 8-4 Key design inputs for the illustrate ISS case study .......................................................................169
Table 8-5 Summary of power and heat output under three investigated cases: 1) without PCC, 2) with PCC but without ISS, and 3) with PCC and ISS ..............................................................................................................170
# Nomenclature

## Acronyms

<table>
<thead>
<tr>
<th>Acronym</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>AD</td>
<td>Additional regeneration</td>
</tr>
<tr>
<td>ARC</td>
<td>Amager Ressource Center</td>
</tr>
<tr>
<td>BECCS</td>
<td>Bioenergy with carbon capture and storage</td>
</tr>
<tr>
<td>Capex</td>
<td>Capital expenditures</td>
</tr>
<tr>
<td>CCC</td>
<td>Committee on Climate Change</td>
</tr>
<tr>
<td>CCR</td>
<td>CO₂ capture rate</td>
</tr>
<tr>
<td>CCS</td>
<td>Carbon capture and storage</td>
</tr>
<tr>
<td>CCUS</td>
<td>Carbon capture utilization and storage</td>
</tr>
<tr>
<td>CE</td>
<td>Circular economy</td>
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<tr>
<td>CEPCI</td>
<td>Chemical engineering plant cost index</td>
</tr>
<tr>
<td>CfD</td>
<td>Contracts for Difference</td>
</tr>
<tr>
<td>CHP</td>
<td>Combined heat and power</td>
</tr>
<tr>
<td>CHP-Ex&amp;C</td>
<td>Combined heat and power plant with steam extraction and condensing configuration</td>
</tr>
<tr>
<td>CHP-BP</td>
<td>Combined heat and power plant with backpressure configuration</td>
</tr>
<tr>
<td>CO₂</td>
<td>Carbon dioxide</td>
</tr>
<tr>
<td>COPₙ</td>
<td>Coefficient of performance of steam extraction</td>
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<tr>
<td>CCSI</td>
<td>Carbon Capture Simulation Initiative</td>
</tr>
<tr>
<td>DCC</td>
<td>Direct contact cooler</td>
</tr>
<tr>
<td>DH</td>
<td>District heating</td>
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<tr>
<td>Abbreviation</td>
<td>Description</td>
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<td>--------------</td>
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<tr>
<td>EOP</td>
<td>Electricity output penalty</td>
</tr>
<tr>
<td>EOR</td>
<td>Enhanced oil recovery</td>
</tr>
<tr>
<td>EP</td>
<td>Efficiency penalty</td>
</tr>
<tr>
<td>ESP</td>
<td>Electrostatic precipitator</td>
</tr>
<tr>
<td>EU</td>
<td>The European Union</td>
</tr>
<tr>
<td>EUFP</td>
<td>Energy utilization factor penalty</td>
</tr>
<tr>
<td>EUP</td>
<td>Energy utilization penalty</td>
</tr>
<tr>
<td>FEED</td>
<td>Front end engineering design</td>
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<tr>
<td>FGD</td>
<td>Flue gas desulphurisation</td>
</tr>
<tr>
<td>GHG</td>
<td>Greenhouse gas</td>
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<tr>
<td>HP</td>
<td>High pressure</td>
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<tr>
<td>hr</td>
<td>hour (unit of time)</td>
</tr>
<tr>
<td>HW</td>
<td>Hot water</td>
</tr>
<tr>
<td>IEA</td>
<td>International Energy Agency</td>
</tr>
<tr>
<td>IEAGHG</td>
<td>IEA Greenhouse Gas R&amp;D Programme</td>
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<tr>
<td>IP</td>
<td>Intermediate pressure</td>
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<tr>
<td>ISS</td>
<td>Interim solvent storage</td>
</tr>
<tr>
<td>ITD</td>
<td>Initial temperature difference</td>
</tr>
<tr>
<td>KPIs</td>
<td>Key performance indicators</td>
</tr>
<tr>
<td>LCA</td>
<td>Systematic life cycle assessment</td>
</tr>
<tr>
<td>Abbreviation</td>
<td>Description</td>
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<tr>
<td>--------------</td>
<td>-------------</td>
</tr>
<tr>
<td>LCOE</td>
<td>Levelized cost of electricity</td>
</tr>
<tr>
<td>LCOH</td>
<td>Levelized cost of heat</td>
</tr>
<tr>
<td>LHV</td>
<td>Lower heating value</td>
</tr>
<tr>
<td>LP</td>
<td>Low pressure</td>
</tr>
<tr>
<td>MEA</td>
<td>Monoethanolamine</td>
</tr>
<tr>
<td>MSW</td>
<td>Municipal solid waste</td>
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<tr>
<td>NEC</td>
<td>Negative emission credit</td>
</tr>
<tr>
<td>NPV</td>
<td>Net present value</td>
</tr>
<tr>
<td>Opex</td>
<td>Operational expenditures</td>
</tr>
<tr>
<td>PCC</td>
<td>Post-combustion carbon capture</td>
</tr>
<tr>
<td>PACT</td>
<td>Pilot Scale Advanced Capture Technology</td>
</tr>
<tr>
<td>RERC</td>
<td>Recycling and Energy Recovery Centre</td>
</tr>
<tr>
<td>SCR</td>
<td>Selective catalytic reduction</td>
</tr>
<tr>
<td>SRD</td>
<td>Specific Reboiler Duty</td>
</tr>
<tr>
<td>SS</td>
<td>Solvent storage</td>
</tr>
<tr>
<td>tCO$_2$</td>
<td>Tonne of carbon dioxide</td>
</tr>
<tr>
<td>TRL</td>
<td>Technology readiness level</td>
</tr>
<tr>
<td>UK</td>
<td>The United Kingdom</td>
</tr>
<tr>
<td>UKCCSRC</td>
<td>UK Carbon Capture and Storage Research Centre</td>
</tr>
<tr>
<td>WtE</td>
<td>Waste-to-Energy</td>
</tr>
<tr>
<td>XFHE</td>
<td>Cross flow heat exchanger</td>
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</table>
List of symbols

<table>
<thead>
<tr>
<th>Symbols</th>
<th>Description</th>
</tr>
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<tbody>
<tr>
<td>% vol</td>
<td>Volume percentage</td>
</tr>
<tr>
<td>$\frac{t\text{CO}_2}{MWh}^{\text{Captured}}$</td>
<td>the total mass CO$_2$ captured per net electricity generation MWh for the plant with CO$_2$ capture (equal to the CO$_2$ produced minus emitted)</td>
</tr>
<tr>
<td>$\dot{m}_{\text{CO}_2,\text{captured}}$</td>
<td>Mass flow of CO$_2$ captured exiting the plant boundaries [t/h]</td>
</tr>
<tr>
<td>$a_{ah}$</td>
<td>Excess air coefficient at the exit of air heater</td>
</tr>
<tr>
<td>$h_{\text{ambient}}$</td>
<td>Enthalpy of air at ambient conditions, in $kJ/kg$</td>
</tr>
<tr>
<td>$h_{\text{entrance}}$</td>
<td>Enthalpy of air at burner entrance conditions, in $kJ/kg$</td>
</tr>
<tr>
<td>$n_{\text{boiler}}$</td>
<td>Boiler efficiency %</td>
</tr>
<tr>
<td>$COP_{X,\text{abs}}$</td>
<td>An analogous COP of absorption heat pump in the CHP integration</td>
</tr>
<tr>
<td>$COP_{X,\text{cap}+dh}$</td>
<td>Coefficient performance of steam extraction (MW$<em>{in}$/MW$</em>{e}$) (CHP plant)</td>
</tr>
<tr>
<td>$COP_{X,\text{cap}}$</td>
<td>Coefficient performance of steam extraction (MW$<em>{in}$/MW$</em>{e}$) (power only plant)</td>
</tr>
<tr>
<td>$C_{ba}$</td>
<td>Carbon contents in bottom ash</td>
</tr>
<tr>
<td>$C_{ba}$</td>
<td>Unburned carbon content in bottom ash</td>
</tr>
<tr>
<td>$C_{fa}$</td>
<td>Carbon contents in fly ash</td>
</tr>
<tr>
<td>$C_{pas}$</td>
<td>Specific heat of ash and slag, $kJ/kg \cdot K$</td>
</tr>
<tr>
<td>$C_{pf}$</td>
<td>Specific heat of fuel, as received basis, $kJ/kg \cdot K$</td>
</tr>
<tr>
<td>$C_{pf}^{\text{dry}}$</td>
<td>Specific heat, as dry basis, $kJ/kg \cdot K$</td>
</tr>
</tbody>
</table>
Temperature correction factor taking into account deviations from counter-current flow in shell-and-tube heat exchangers; 

Amount of bottom ash produced, kg/s; 

Amount of fly ash produced, kg/s; 

Sensible heat of fuel 

Theoretical cold air enthalpy entering boiler, kJ/kg fuel 

Flue gas enthalpy at the exit of air heater, kJ/kg fuel 

The amount of heat can be provided from an electricity driven heat pump, consuming the ‘lost’ power (MWth) \( W_T \); 

Sensible heat carried by the air when heated by external air heater 

The change of district heating output due to CO\(_2\) capture (MWth) 

The district heating output (MWth) with PCC 

Fuel thermal input on LHV basis [MWhth] 

Net useful thermal output for district heating 

Fuel temperature at burner or feeder exit 

Temperature of the slag that leaves the furnace, °C 

Loss of power output due to steam extraction (MW\(_e\)) 

Net power output with CCS, considering reduction in steam turbine power output due to steam extraction for solvent regeneration, power consumption associated with the CO\(_2\) capture process and any offsets due to beneficial heat recovery for condensate heating and other purposes [MW\(_e\)] 

Net power output without CCS [MW\(_e\)]
\(X_{ba}\) Mass fraction of total ash exiting through the bottom ash

\(X_{fa}\) Mass fraction of total ash exiting through the fly ash

\(f_{BUILD}\) Correction factor for Buildings

\(f_{CONT}\) Correction factor for Contingency

\(f_{DEC}\) Correction factor for Design and Engineering

\(f_E\) The percentage of CO\(_2\) emission allocated to power production

\(f_{ELEC}\) Correction factor for electrical

\(f_{ER}\) Correction factor for equipment erection

\(f_{INST}\) Correction factor for instrumentation

\(f_M\) Correction factors for materials of construction

\(f_{OS}\) Correction factor for off-sites

\(f_P\) Correction factor for design pressure

\(f_{PIP}\) Correction factor for piping

\(f_Q\) The percentage of CO\(_2\) emission allocated to heat production

\(f_{SP}\) Correction factor for site preparation

\(f_T\) Correction factor for design temperature

\(f_{WC}\) Correction factor for working capital

\(\eta_{heat}\) Heat generation efficiency of in separate plants

\(\eta_{CCS}\) Electrical efficiency with CO\(_2\) capture and storage

\(\eta_{power}\) Power generation efficiency of in separate plants
\( \eta_{ref} \) Electrical efficiency without CO\(_2\) capture and storage

\( \Delta \eta \) Efficiency penalty of CO\(_2\) capture

\( \Delta T_{lm} \) Logarithmic mean temperature difference:

bara Absolute pressure

The ratio of rich solvent discharges into the stripper under ISS applications to that under the base case.

kt/a Thousand tonnes per year

MJ Megajoules

MJ/kgCO\(_2\) Megajoules per kg CO\(_2\)

Mtpa Million tonnes per year

MWh Megawatt-hour (unit of energy)

\( Q_{cap} \) Heat supplied to solvent reboiler

t/h tone per hour

%wt weight percentage

\( \Delta P \) Pressure difference across pipe length \( L \), kN/m\(^2\)

\( A \) Heat exchanger area, m\(^2\)

\( C \) Hazen Williams coefficient

\( D \) Pipe diameter, mm

\( E \) Power production from the CHP plant

\( L \) Pipe length, m

\( Q \) Heat production from the CHP plant
$U$ Overall heat transfer coefficient

**Subscripts**

$\text{ref}$ Reference case

$\text{CCS}$ CCS case

$\text{actual}$ The plant under evaluation
1 Introduction

1.1 Motivations of waste to energy with carbon capture

In recent years, the increasing waste amounts, shrinking landfill spaces in urban agglomerations and higher ecological standards have stimulated the growth of WtE throughout the world (Ecoprog, 2022). Urban waste management is one of the central concerns in urban agglomerations, especially at a time when worldwide cities are rapidly expanding and looking for pathways of sustainable waste management has become increasingly vital in the commitment to achieve net-zero. Urban waste management contributes to greenhouse gas (GHG) emissions but its impact varies widely with the treatment methods applied. To reduce its impact, the EU has established the Waste Hierarchy to regulate waste management. Among them, Waste-to-Energy (WtE) should be applied to those Municipal Solid Waste (MSW) fractions that cannot be reused or recycled, in order to (1) reduce their volume and destroy contaminants, (2) recover useful energy, and (3) reduce emissions to the environment as compared to landfill.

In practice, WtE plants work by accepting MSW possibly at zero cost or even for a fee depending on the regulatory context. In the UK and the EU, gate fee, also known as tipping fee, is charged by WtE plant for entry or access to their waste management facilities. This is driven by the so-called “Landfill Tax” (£101.20/tonne in Scotland, so a strong incentive to avoid landfill! Only inert mineral waste qualifies for a reduced fee of £3.20/tonne REF). However, costs can be high considering the handling operations, pre-processing like shredding for example, when incorporating the required emission control.

The composition of MSW varies all around the globe according to the consumption habits, geographical locations, and variation in seasonality and climatic conditions (Farooq et al. 2021). While this can require adapting the design to different conditions the biogenic percentage of MSW found in WtE plants is approximately 50% -70% (i.e. originated from renewable biomass) (GIOUSE, 2020; Kaza & Woerden, 2018), leading to energy from WtE plants having a lower carbon intensity than that produced through fossil fuel-based power plant. For instance, instantaneous fossil CO2 emission from electricity generation plant could be found to be 710kg/MWh and 384kg/MWh using coal and waste, respectively (Sathre, Gustavsson, and Truong 2017). Nevertheless, incineration of MSW, if not managed, is potentially a large source of methane emissions, pollution, human and ecosystem health risks and underground water contamination (Pour, Webley, and Cook 2018).

To further reduce its environmental impact, WtE plant can be combined with carbon capture technologies. This enables energy production with negative CO2 emissions. It has been estimated that the treatment of MSW in WtE-CCS plants would allow for the removal from the atmosphere of 2.8
In 2021, the International Energy Agency (IEA) published the Net Zero Emissions by 2050 Scenario (IEA NZE) consistent with holding the increase in the global average temperature to 1.5 degrees above pre-industrial levels (IEA, 2021). In order to reach Net Zero Emissions by 2050, carbon capture utilization and storage (CCUS) has been identified as an important part of the solution. Among the technical families of CCUS, Bio-Energy with Carbon Capture and Storage (BECCS) is a promising carbon removal approach that has the potential to offer permanent net removal of carbon dioxide (CO₂) from the atmosphere over its lifetime (Rosa, Sanchez, & Mazzotti, 2021). The combined motivation of both promoting sustainable development of urban waste management and delivering ‘negative carbon emissions’ for climate change mitigation makes the adoption of Carbon Capture and Storage (CCUS) in the WtE sector promising, and is attracting increased attention in the recent years. A limited number of studies have been published (Haaf, Anantharaman, Roussanaly, Ströhle, & Epple, 2020; Magnanelli, Mosby, & Becidan, 2021; Roussanaly, Ouassou, Anantharaman, & Haaf, 2020).

Among all the CCUS technologies, solvent-based post-combustion carbon capture (PCC) is one of several options for removing CO₂ from exhaust flue gases in power plants. It has been proposed for WtE plants since it is the most proven technology for CO₂ capture and no significant modifications to the original plant are required (AECOM, 2021; Zanco et al., 2021). Experience obtained when CCUS is applied in other industry can be transferred to its application in WtE sector. However, there are still areas that need attention to adapt CO₂ capture technologies for effective use in the WtE sector. The integration of PCC in WtE facilities requires a large quantity of steam extraction from the power cycle for solvent regeneration in the capture facilities, which causes energy penalty to the WtE plants; this may bring more challenge for Combined Heat and Power (CHP) plants which usually prioritise heat supply. Considering the large share of CHP plant in Europe and the trend is increasing (Connolly et al., 2014; Reimann, 2013; Scarlat, Fahl, & Dallemand, 2018; Zafar, 2022), it is vital to investigate approaches to improve heat supply approaches for the CO₂ capture process and for the District heating (DH) network, and design options for an effective thermal integration.

1.2 Research objectives

CO₂ capture integration into power plants is not new in the research field. However, research with a focus on WtE with CO₂ capture was just emerging around the point when the work reported in this thesis started. There are some integration challenges that need to be addressed for the application in the WtE industry. For instance, different from traditional power-only integration, in Europe, approximately 50% percentage of WtE plants operate as CHP plants, which brings new angles when
determining the optimal integration approach. Additionally, the capacities of WtE plants are usually one-two orders of magnitude smaller than conventional fossil power stations. This means the WtE plants are relatively small to follow large economies of scale once considering CO₂ capture integration. The economic impacts of this application should be notified to policymakers and plant operators, including the annual revenues that could be generated under this application, etc. It is also widely acknowledged that WtE with CCS can bring negative emission benefits; however, there is a lack of systematic understanding of the negative emission potential, especially coupling the most recent developments in the CCS field for the specified waste sector.

To meet net-zero targets, the CO₂ capture rate should be as close to 100% as technically possible while maintaining economically viable. The concept of ‘90% capture’ has, more or less, become ubiquitous in the literature, from both academic and policy perspectives. However, even with 90% capture, there is still the potential for appreciable residual CO₂ emissions from ostensibly decarbonised point sources. A growing trend from researchers has, therefore, identified the need to design and operate power plants with ‘ultra-high’ CO₂ capture rate, i.e. CO₂ capture rates equal or above 99% (Gibbins & Lucquiaud, 2021; Stavros Michailos & Gibbins, 2021).

The corresponding solutions to these identified challenges are needed to facilitate the contribution of the WtE sector to the Net-zero emissions target. In this context, the research objectives for this work are:

A. Develop options for thermal integration of Post-combustion CO₂ capture with WtE facilities:
   a) Identify representative steam cycle configurations in the WtE industry;
   b) Categorise integration scenarios of PCC plant with the identified WtE plants.

B. Assess the effect of ultra-high CO₂ capture rates on the performance of Waste to Energy plants with post-combustion carbon capture (WtE-PCC):
   a) CO₂ capture process modelling (in ASPEN, using 35% MEA as solution);
   b) Analyse the implications of Ultra-high CO₂ capture on the design and operation of WtE-PCC plants.

C. Define and evaluate key performance indicators (KPIs) to assess the performance of different thermal integration options.
   a) Determine a set of KPIs that are best suited for WtE industry application from a thermodynamic, environmental, and economic point of view;
b) Present calculation method for each KPI under discussion and the KPI results for the investigated integration scenarios.

D. Investigate options for improved thermal integration approaches of post-combustion CO\textsubscript{2} capture technology with WtE-CHP plants:

a) Identify advanced heat integration options (reducing energy penalty by recovering heat from capture process) and compare to basic thermal integration of PCC into Rankine cycles;

b) Explore the negative emission potential for interim solvent storage to contribute to better matching between WtE with PCC plant operations and fluctuating DH demand from the end users (potentially replacing the heat load of local gas boilers);

1.3 Novelty of work

The following results can be considered an original contribution to knowledge:

- This thesis defines for first time a set of KPIs tailored to evaluate the performance of WtE plants equipped with PCC. 1) The KPIs includes thermodynamic, economic and environmental point of views, in order to have a comprehensive understanding of the overall impact of adding PCC systems and to inform plant owners and other practitioners on the benefits and challenges of this technology. 
2) Besides the level of integrity of the KPIs, this thesis expands the work of Lucquiaud & Gibbins (2011) by adding a new KPI specific for WtE-CHP plants. Lucquiaud & Gibbins identified the Coefficient of Performance of steam extraction ($COP_X$) of power-only plant to evaluate the effectiveness of steam extraction compared with a conventional electricity driven heat pump to supply thermal energy for solvent regeneration in the PCC plant (Lucquiaud & Gibbins, 2011). This work introduces $COP_{X-cap+dh}$ specific for CHP plants that allows a comparison with the integration of a system combining electricity driven and thermal driven absorption heat pumps in an integrated WtE-CHP plants with PCC.

- This work evaluates for first time the effect of ultra-high CO\textsubscript{2} capture levels on the performance of the WtE plants. Through process modelling of 35% MEA based PCC integrated into WtE plants, the implications of ultra-high CO\textsubscript{2} capture rates are presented and analysed by evaluating the thermodynamic, economic and environmental KPIs.

- A novelty of this study comes from developing and implementing thermal integration options with the aim of solving the competing heat requirements for the CHP plant with PCC both in terms of magnitude and scheduling (this refers to the fact that CHP heat demand is seasonal, variable on a daily basis etc). 1) To enhance heat recovery from an integrated WtE plant with PCC, an advanced thermal integration approach is conducted in this work using comprehensive process modelling, and is compared with the basic thermal integration using the identified KPIs set. Results from this work
provide rigorous evidence of the thermal balancing potential of the use of advanced heat integration to tackle the competing steam extraction requirements for DH and CO₂ capture. 2) This thesis also presents novel work on flexible operation of WtE-CHP plants with PCC by implementing Interim Solvent Storage (ISS), as an additional approach to address the problem of competing heat requirement. An integrated assessment is conducted to evaluate the economic implications, additional revenues and the environmental benefits of using ISS in a WtE-CHP plant configuration. This is important work since it brings another problem-solving approach that may generalise to all CHP operation for the application of CHP plant with CO₂ capture in the future urban district heating industry.

1.4 Overview of the methodology and thesis outline

An overview of the overall methodology for this study is provided in Figure 1-1.

![Figure 1-1 Overview of thesis outline](image-url)
In order to understand the potential for application of CO₂ capture with WtE plants, a series of technical reviews were conducted to establish an understanding of state-of-the-art in terms of the key technologies and the most recent development of both WtE and CO₂ capture aspects. This process helped to guide and define a manageable scope for this thesis, including identifying qualitative and quantitative datasets such as the investigated configurations of WtE plants/key design inputs, and choices of CO₂ capture methods/capture rates, etc.

Building on this review, the defined WtE plant and the CO₂ capture & compression facility are thermally integrated based on a review of the previous integration methods.

A set of key performance indicators (KPIs) is determined to evaluate the different integration options that are developed in this thesis. The KPIs are further categorised into thermodynamic aspects, economic aspects, and environmental aspects, in order to conduct a comprehensive assessment of the integrations and to deliver the most realistic results of metrics that are mostly concerned through the literature and in the fields of integration with CO₂ capture.

Comprehensive process modelling activities are undertaken to simulate the investigated integrations, with the overall thermal process modelled in gProcess and the CO₂ capture plant modelled separately in ASPEN to obtain the key parameters such as specific reboiler duty of the regenerator/packing parameters of capture plant. Using the modelling tools, cases studies around the defined integration are developed starting from a base case that is created based on the most representative integration methods. Two additional cases are developed to test concepts for further improvement in system efficiency/flexibility.
2 Background

This chapter gives a systemic review of the Waste to Energy (WtE) technologies and applicable CO₂ capture technologies to the WtE sector, to give an overall understanding in terms of the background of this PhD research topic. It includes the introduction of WtE technologies and their role in the waste management hierarchy, applicable CO₂ capture technologies for WtE plants alongside the opportunities and challenges. The chapter also gives an update on the recent development of WtE with CCS projects to give an understanding of what is happening in the real world of this application. The final section of this chapter reviews the modelling approaches in literature when integrating with post-combustion CO₂ capture technologies. This background review helps to give a clue of narrowing down the research topic, and the reasoning in terms of technology choices for this PhD study.

2.1 Waste to energy technology

2.1.1 WtE in the waste management hierarchy

Driven by rapid urbanization and growing populations, global annual waste generation is increasing. According to a report from the World Bank Group, global waste generation was estimated to be 2.01 billion tonnes per year in 2016, and this number was expected to grow to 3.40 billion tonnes by 2050 under a business-as-usual scenario (World Bank Group, 2018). To address this, international and national regulations have been published, among which, the waste management hierarchy has been a widely accepted guideline for waste management operations throughout the world. The waste hierarchy ranks waste management options according to the best environmental outcome taking into consideration the lifecycle of the material (Defra, 2011; European Parliament and Council, 2018; The Scottish Government, 2017), as shown in Figure 2-1.
In the waste management hierarchy, WtE can be classified into disposal, other recovery or recycling operations, according to the energy products produced and recovery level (UNEP, 2019). Although the precise interpretation may vary from one place to another, WtE refers to a variety of treatment technologies that convert non-recyclable waste to electricity, heat, fuel, or other usable materials, as well as a range of residues including fly ash, sludge, slag, boiler ash, wastewater and emissions (Alao, Popoola, & Ayodele, 2022; CEWEP, 2022b; Defra, 2014; ESWET, 2021; UNEP, 2015). The role of WtE has been highlighted by the European Commission in the circular economy (CE) as a guiding concept that might help in the transition, as long as decisions are not made that obstruct higher levels of prevention, reuse, and recycling (European Commission, 2017). The Commission calls on using the most energy-efficient waste-to-energy techniques (European Commission, 2017). According to R1 formula as established by the European Commission Waste Directive 2008/98/EC, only WtE plants that meet the energy efficiency standard (equal to or higher than 0.60 for installations in operation before 2009 or 0.65 for installations permitted after 2009) can be regarded as an energy recovery operation.

In recent years, the increasing waste amounts, shrinking landfill spaces in agglomerations and higher ecological standards have stimulated the growth of WtE throughout the world (Ecoprog, 2022). In 2022, more than 2,600 WtE plants are active worldwide. It is estimated by Ecoprog that there will be about 3,000 plants with a capacity of about 630 million tons per year operational by 2031 (Ecoprog, 2022). In the UK, in 2021 alone, three WtE incinerators were commissioned and the total number of
fully operational WtE plants was increased to 53, with total capacity of 16.4Mtpa, approximately 76% of the total UK permit capacity (Tolvik, 2022). This number of operating plants represents more than double that in 2014 (Statista, 2023).

2.1.2 Types of WtE technologies

Various categorization methods can be found in literature to classify WtE plants, among which the mostly commonly used is based on types of energy conversion processes. In this categorization approach, WtE can be in general classified into three categories: Thermochemical, Biochemical and Physicochemical (Farooq, Haputta, Silalertruksa, & Gheewala, 2021).

![Classification of WtE technologies based on energy conversion pathways](image)

**Figure 2-2** Classification of WtE technologies based on energy conversion pathways

Among all the WtE technologies, incineration is well-established and the most commonly used WtE technology globally (Farooq et al., 2021; IEA Bioenergy, 2019; World energy council, 2016). With this technology, waste is combusted between 750°C and 1100°C in the presence of oxygen, and raises steam for electricity and/or heat generation in a boiler or steam turbine. The volume of the waste is effectively reduced by 75–90% through this process (UNEP, 2019). Figure 2-3 is the main process flow diagram of a typical incineration type WtE plant - the new Amager Bakke WtE plant in Copenhagen, Denmark. The plant was established by Amager Ressource Center (ARC) and is one of the largest combined heat and power (CHP) plants in northern Europe, with the capacity to treat 560,000 tonnes of waste annually (70t/h) (ARC, n.d.; Bisinella, Nedenskov, Riber, Hulgaard, & Christensen, 2022; Hulgaard & Søndergaard, 2018).
Figure 2-3 Process flow diagram of the Amager Bakke waste-to-energy facility (showing just one of two boilers) (Hulgaard & Søndergaard, 2018)

The Amager Bakke plant has two identical moving grate-fired 35 t/h drum-type boilers, from which energy recovery is achieved by incineration of the waste to produce high-pressure steam (440°C/70 bar) to generate district heat and power by a highly efficient steam turbine. Flue gas from the plant is treated in an electrostatic precipitator (ESP), after which a selective catalytic reduction (SCR) is used to reduce nitrogen oxides (NO\textsubscript{x}, NO and NO\textsubscript{2}) to a low level by use of ammonia water, and destroys dioxin, prior to a four-stage wet flue gas treatment. Metals are recovered from the bottom ash and are sold at high prices to replace virgin materials. The bottom ash is used for road construction and similar construction purposes under stringent requirements for its use. A CO\textsubscript{2} capture demonstration plant is expected to be operational in late summer 2023 by ARC. The goal is to commission a full scale carbon capture unit and capture up to 500,000 tonnes of CO\textsubscript{2} annually (ARC n.d.).

As shown in Figure 2-4, categorization of WtE plants can also be based on types of application (power, heat, CHP, and transportation fuel). Electricity can be produced from waste through direct combustion, and the released heat is utilised to produce steam to drive a turbine. This indirect generation has an efficiency level of about 15% to 27%, with modern plants reaching the higher end
of the range (World energy council, 2016). Heat production is a conventional method for energy recovery by combustion of waste in a boiler system to produce steam. WtE installations working in CHP mode are highlighted by the European Commission to increase energy efficiency for waste-to-energy processes (European Commission, 2017). According to a dedicated European Commission study, the WtE CHP net annual average efficiency is estimated to rise from 76% to over 88% with the addition of heat pumps in tandem with flue gas condensation for the most advanced plants (European Commission, 2016). Another example is the Amager Bakke WtE plant as mentioned above, with improved energy recovery and optimisation, the plant can reach a total net efficiency of 107% (with heat pumps), an outstanding thermal performance (Hulgaard & Søndergaard, 2018). Though higher overall efficiency is achieved by CHP plants, the market for WtE CHP is still developing, since the cost of the technologies has not yet benefitted from the economies of scale that conventional CHP technologies have experienced.

Figure 2-4 Classification of WtE technologies based on applications

In the UK, some WtE CHP schemes are dependent on government support to remain economically viable. For example, WtE CHP schemes are ineligible to apply for Renewable Heat Incentive (RHI) and Contracts for Difference (CfD), with the latter including strike prices on both the power and heat component supplied (unlike those for biomass CHP schemes that are based on ‘power only’). In the UK, the share of WtE CHP application in total WtE plants is historically low compared to that of other European countries (K. P. and J. 2013) – 20% in the England versus around 50%-60% in Europe (CEWEP, 2022b; Scarlat et al., 2018; UK Parliament, 2022). The support and subsidies on CHP applications are increasing. Examples can be seen in Resources and Waste Strategy (Defra, 2018), CHP Ready Guidance for Combustion and Energy from Waste Power Plants (Environment Agency, 2013), Combined heat and power (CHP) development strategy (North london waste authority, 2015), etc.

Categorization of WtE plants can also be based on types of waste treated. Municipal Solid Waste (MSW) is the most common waste stream used in the thermal WtE process (UNEP, 2019) and is used as the waste type in this study. Definitions of MSW vary among countries due to different legal
frameworks. In general, however, MSW must include waste items collected from households. In some countries, the definition will include commercial waste, hospital waste, and construction and demolition waste (World Bank 2018; OECD 2019; UNSD2019; IPCC 2006(Eurostat, 2017)). The variation in MSW definition leads to inconsistency of waste data, which makes direct comparison between countries difficult. Besides MSW, construction waste, bio-waste from agriculture and forestry activities, hazardous waste and many others also can be considered feasible for energy recovery, depending on their specific composition, their energy content and the specific needs of society in terms of waste disposal.

Finally, emission prevention and control are an important aspect of the safe design and operation of WtE plant. A well-managed WtE plant is equipped with sophisticated combustion control and pollution abatement system to reduce the possible adverse health effects associated with atmospheric emissions (ESWET 2022). To some extent, this can restrict the specifications on the waste feedstock to the WtE plant, however a well-designed plant with the appropriately selected feedstock should be compliant with respect to emissions. It is true that a poorly fed WtE facilities may emit concentrated toxins with serious potential health risks, such as dioxins (which refers to polychlorinated dibenzodioxins and polychlorinated dibenzofurans) and heavy metals. WtE facilities frequently face strong protests from local communities. However, a recent report from CEWEP shows that: 1) WtE dioxin emissions account for less than 0.2% of the total industrial dioxin emissions; 2) Regardless of specific measuring equipment a well-managed EU WtE plant emits extremely low concentrations of dioxins and furans thanks to its sophisticated combustion control and pollution abatement system (ESWET 2022).

This PhD study aims to research on the most representative WtE plants that may be applicable for the integration with CCS, so an incineration type WtE using MSW as a fuel, in three applications (power, heat and CHP) is considered.

2.2 Applicable CO$_2$ capture technologies for WtE plants

CCS is a group of technologies that comprises separation of CO$_2$ from industrial sources, compression and transportation to a geologic site for storage, or to enhanced oil recovery (EOR). The development of CCS technologies is at varying levels of maturity, or technology readiness level (TRL). The International Energy Agency (IEA) applies eleven TRL levels to assess and compare the maturity of CO$_2$ technologies across and within different sectors, from concept stage to prototype, demonstration, early adoption and mature. Obstacles to the development are mostly because facilities are capital-intensive to deploy and expensive to operate (Business Energy and Industrial Strategy Committee, 2023; Roussanaly et al., 2021; Roussanaly et al., 2020; Shen et al., 2022; Subraveti, Roussanaly,
Among all the sections of the CCS value chain, the CO₂ capture section is the largest contributor to the cost of CCS, compared to the cost of transport and storage (Roussanaly et al., 2020), which makes reducing the cost of CO₂ capture essential.

2.2.1 Amine-based chemical absorption

There are three main methods to capture CO₂ from fossil fuel combustion. Post-combustion CO₂ (PCC in this study) captures CO₂ directly from flue gases exiting standard combustion processes, typically with the use of solvents, sorbents, membranes or cryogenics. Pre-combustion capture refers to removing CO₂ from fossil fuels before combustion is completed. For example, syngas (H₂/CO) produced from gasification process of a feedstock will undergo water-gas shift reaction to produce H₂ and CO₂-rich gas mixture. The CO₂ can then be captured and separated, transported, and ultimately sequestered, and the H₂-rich fuel combusted. Thirdly, oxyfuel combustion burns coal or gas in pure oxygen, or CO₂/O₂ mixture to avoid very high flame temperature (3,000-4,000 °C), to yield only CO₂ and water. Among the broad range of technologies, ammine based post-combustion was the only CO₂ capture technology included in the mature group in 2020, according to ‘Special Report on Carbon Capture Utilisation and Storage, CCUS in clean energy transitions’ (IEA, 2020). Among all the amine-based solvent, Monoethanolamine (MEA), a first-generation baseline solvent, has been studied extensively and both pilot scale data and modelling activities have been reported.

The principle of the amine-based chemical absorption is composed of three stages as represented in Figure 2-5.

Figure 2-5 Conventional aqueous amine solvent plant process flowsheet integrated with process optimisation using the intercooling and rich split processes (Global CCS Institute, 2021).
In the first stage, flue gas from the point source is initially cooled in the direct contact cooler (DCC) to reduce to temperatures to around 40-60°C where the chemistry of absorption is most efficient. The cooled flue gas is then fed to the bottom of the absorber, where the CO₂ reacts with MEA and is dissolved in the amine solvent (typically 30-40 %wt MEA). There are researches looking at higher concentration too). The CO₂ ‘rich’ solvent leaves at the bottom of the column and is heated up against CO₂ ‘lean’ solvent in the cross-flow heat exchanger before entering the stripper. A water wash section at the top of the absorber column removes any droplets of solvent that get carried-over with the rest of the flue gas and balances any water evaporation in the system.

In the second stage, the chemical bond between the solvent and the CO₂ is broken by the heat provided in the reboiler situated at the base of the desorber column (or called stripper). The total amount of heat required is commonly referred to as the reboiler heat duty and is essentially the sum of the energy used for:

• Sensible heat: the energy required for the temperature difference between the incoming rich solvent and the outgoing lean solvent

• Heat of desorption: reversing the absorption reaction by breaking the chemical bond between the CO₂ and the absorbent

• Latent heat: vaporising water to establish an operating CO₂ partial pressure and using steam as the stripping gas to carry the CO₂ to the top of the column

In the third stage, the purified CO₂ (mixture of CO₂ and water vapour) leaves the top of the stripper column and is cooled down in a stripper overhead condenser. The condensed water is returned to the stripper while the CO₂ is compressed further (not included in Figure 2-5) to a dense phase fluid at a pressure suitable for transport to a geological storage site, typically 110 bar. The remaining water condensate is gradually removed as part of the compression process. At the bottom of the stripper the CO₂ lean solvent is recirculated back to the absorber through the crossflow heat exchanger followed by further cooling before re-entering the absorber column.

The dissolution and reaction of CO₂ gas with the amine solvents make the solution corrosive and the amine degrades over time. For amine based CO₂ capture, the most dominant mechanisms of degradation are: thermal degradation, caused by polymerization reactions, and oxidative degradation, caused by oxidising agents such as dissolved oxygen, SOₓ and NOₓ, and catalysed by dissolved metals (Reynolds, Verheyen, and Meuleman 2016).

There are well known methods to reduce degradation. Flue gas pretreatment technologies, removing impurities such as SOₓ and NOₓ gases, as well as particulate matter such as fly ash are implemented to
some extent in most pilot campaigns. Methods such as “Bleed and Feed”, removal of a part of the degraded solvent and refilling with fresh solvent throughout the process, have recently been thoroughly tested without success. Apart from the “Bleed and Feed”, solvent reclaiming is often used to limit the amount of makeup solvent and maintaining the operation (Buvik et al. 2021). The reclaiming operation is conducted periodically to recover amine by decomposing heat stable salts created by flue gas impurities and oxygen and to dispose of degradation products. At the time of the writing, the importance of solvent management is increasingly acknowledged, however, there is still no universal guidelines for how to monitor amine degradation in a carbon capture plant.

When integrating with industrial point sources, chemical absorption processes operating at atmospheric pressures are usually employed to separate CO₂ from the flue gas before it is discharged to the atmosphere. They are likely to be added after other flue gas treatment processes - Selective Catalytic Reduction (SCR) for nitrous oxide removal, ElectroStatic Precipitation (ESP) or baghouse filter for particulates, and Flue Gas Desulphurisation (FGD) for sulphur oxides.

Advantages of amine-based PCC are summarized below:

- It is the most well-known and employed absorbent within the CO₂ capture industry (Chai, Ngu, & How, 2022). It has been studied extensively and both pilot scale and modelling activities have been reported (Gao, Selinger, & Rochelle, 2019).

- The primary amine, MEA has exceptional CO₂ removal capability (87.1–100%) while producing a nearly pure CO₂ of > 99% (Chai et al., 2022).

- Fast kinetics, high water solubility and low price (Chai et al., 2022; Liang, Fu, Idem, & Tontiwachwuthikul, 2016; Ramezani, Mazinani, & Di Felice, 2022).

With these advantages, amine-based CO₂ capture technology has been constructed and demonstrated in the power sector at scale. For instance, SaskPower’s Boundary Dam power station (BD3 ICCS), the world’s first fully integrated and full-chain CCS facility on a coal-fired power plant(Giannaris et al. 2021), began in operation in 2014, with designed capture ability of 1MtCO₂/yr (Gibbins 2021). The application of CCS on WtE sector shares some similarity of its application and operation with the power sector, which makes amine-based chemical absorption undoubtedly identified as the preferred candidate internationally. In 2021, amine-based chemical absorption was deployed at all the ongoing seven WtE-CCS projects (recorded), from small scale 2.5kt/a CO₂(captured) in Japan to middle scale 414kt/a CO₂ (captured) in Norway, to larger scale 800kt/a CO₂ (captured) built in the Netherlands (Fagerlund et al., 2021; IEAGHG, 2020; Tota, Gatti, Viganò, & Garcia, 2021).
However, there are some drawbacks of amine-based chemical absorption such as high degradation rate, low CO$_2$ loading capacity, toxic nature, and high energy requirement for regeneration, which hinders its industrial utilization (Bougie & Fan, 2018; Sreedhar, Nahar, Venugopal, & Srinivas, 2017). In recent years, efforts have been made to optimize the application of amine-based CO$_2$ capture. When it comes to the application of amine-based CO$_2$ capture in the WtE sector, these drawbacks need to be considered. Additionally, there are some specific concerns that may be relevant to the application of amine-based CO$_2$ capture in the WtE sector. The learnings from implementing CCUS at WtE plants offer the first of its kind knowledge. Areas for technology development in relation to the application of the technology in the WtE sector are described below:

- WtE facilities are capital-intensive themselves already, so reducing the cost of integrating with CCS is the primary concern rather than other technical aspects. Standardisation of designs and improving construction techniques to allow cost reductions are expected to be important (AECOM, 2021; Gammer, 2020).
- As amines can be easily degraded in the presence of impurities, there is concern related to flue gas composition from WtE plants. Increased understanding of the impact of flue gas contaminants and variations on long-term solvent performance will be required (AECOM, 2021; Gammer, 2020; Gibbins & Lucquiaud, 2021; IEAGHG, 2020).
- WtE plants located in urban areas may have limited space to add CO$_2$ capture facilities (Gibbins & Lucquiaud, 2021).
- For some WtE plants, the main purpose of waste combustion is supply of heat (and power), and this main purpose must take priority over the supply of steam required for CO$_2$ capture (Gibbins & Lucquiaud, 2021; Magnanelli et al., 2021).
- Requirement on high capture level (i.e. 95% or above). High CO$_2$ capture levels are found achievable on traditional power plants, but have not yet been demonstrated on WtE plants (Gibbins & Lucquiaud, 2021).

In this study, process simulation of chemical absorption with a 35 %wt aqueous solution of MEA is undertaken in ASPEN Plus, along with optimisation of the PCC process to quantitatively determine the minimized Specific Reboiler Duty (SRD) of CO$_2$ capture and the size of the absorber column. SRD is defined as the thermal energy required to regenerate the solvent per unit of mass of CO$_2$ recovered, as shown in Equation 2.1.
In this study, 35 %wt aqueous solution of MEA is applied as the solvent for the PCC plant. The choice of this solvent is in general a design choice. At the time of the writing, 35% MEA is found relatively low-cost solvent and commercially available at the moment. And this solvent is widely used in pilot CCS plant (Gibbins 2021; Michailos and Gibbins 2022) and used as benchmark solvent so far (AECOM 2022). Comparisons in terms of different concentrations of MEA solvent could be found in literature. For 35% MEA and 40% MEA, the 40% MEA shows slightly lower reboiler duty, but higher corrosion rate and degradation rate (Akram et al. 2020).

As outlined above, technology advancement and process optimization for the implementation of CO₂ capture for WtE plants is necessary and valuable. This PhD researches the possible thermal integration options for three representative WtE plants with amine-based CO₂ capture at varied CO₂ capture rates. Detailed performance assessment and analysis of each option is also conducted to determine the possible impact of adding CO₂ capture to the WtE plants. This work is relevant to researchers, policymakers, investors and key stakeholders as well as the wider academic community.

2.2.2 Next-generation capture technologies

Besides conventional amine-based chemical absorption, some novel capture technologies are developing. They are at a lower TRL level at present but show potential for cost reduction (Roussanaly et al., 2020). In 2021, the Global CCS Institute nominated a few CO₂ capture technologies as the carbon capture candidates for the next significant wave of CCS facilities, as seen in Figure 2-6 (Global CCS Institute, 2021). The Global CCS Institute also concluded that there are ample opportunities to drive down the cost of CO₂ capture and to shorten project deployment timelines, including through economies of scale, modularisation, heat integration and process optimisation, combined with next-generation technologies (Global CCS Institute, 2021).
Next generation CO₂ capture technologies are not the focus of this study, but it should be useful to gain some insights into the available choices and their opportunities and challenges, thus opening more options for the future implementation of CO₂ capture at WtE plants. Figure 2-7 summarizes some of the opportunities and challenges for next generation CO₂ capture technologies that are recommended as candidates for the WtE sector. It can be seen that most of the technologies have advantages of reduced energy consumption, relatively low cost, lower toxicity, etc. The challenges are mainly lack of operational experience on WtE flue gas and validation of energy performance.
Opportunities and challenges of potential next-generation CO\textsubscript{2} capture technologies for WtE plants (AECOM, 2021; Capsol, 2023; Gibbins & Lucquiaud, 2021; IEAGHG, 2020)

* HPC: Hot Potassium Carbonate; MCFC: Molten Carbonate Fuel Cell
2.3 Current development of WtE with CCS projects

At the time of the writing (April 2023), there are several large-scale CO\(_2\) capture project proposals for WtE plant at the early stages of development across Europe. In this chapter, the most up to date situations of WtE with CCS projects at several key European countries (including Norway, Denmark, Netherlands, and UK) will be presented, along with a review of on-going and upcoming and WtE with CCUS projects worldwide.

Among the several ongoing projects, the Hafslund Oslo Celsio - Klemetsrud CCS Project at Norway’s largest WtE plant (previously known as the Fortum Oslo Varme (FOV) CCS Project,) is arguably the project that has come the farthest (Becidan, 2021; SCCS, 2022). A first pilot phase of the Fortum project started in 2015 and since then, a series of pilot campaigns and feasibility studies have been conducted on an amine-based CO\(_2\) capture system. By March 2022, the project secured full financing of NOK 9.1 billion, approximately £0.67 billion (ACCSESS, 2022) and was reported to be planning to be operational around 2026/2027 aiming to capture 400 ktCO\(_2\)/yr. There have been no detailed project updates information in the open literature since the full financing was secured. The most recent report was a Front end engineering design (FEED) study completed in 2020 (AS, 2020). In the FEED phase, the implementation of heat pumps to provide DH heat when the new WtE DH plant is retrofitted with CO\(_2\) capture was evaluated. The heat pumps recover heat from the condenser of the CO\(_2\) capture plant and pump DH water from the existing DH line through the condenser of the CO\(_2\) Capture Plant heat pump and back to the existing DH network. The detailed Heat Pump Package design was to be finalised during project execution.

Another WtE CHP plant, Amager Bakke Waste-to-Energy Plant in Copenhagen, Denmark, one of the largest combined heat and power (CHP) plants in northern Europe, commissioned a pilot plant able to capture a small amount of the CO\(_2\) in flue gas on 24 June 2021 (ARC, n.d.; Green, 2022). It was expected that a demonstration plant with a primary purpose of simulation of district heating integration would be operational by late summer 2023, capturing up to 4 tonnes of CO\(_2\)/day. The goal is to commission a full-scale CO\(_2\) capture unit and capture 500,000 tonnes of CO\(_2\) annually by 2025. This project investigates a variety of potential markets that will be available if CO\(_2\) can be produced on a commercial quality and scale. Residual heat recovery is also considered to supply district heat (Jensen, 2021).

In May 2022, Aker Carbon Capture officially commenced building its Just Catch™ modular CO\(_2\) capture plant at Twence’s waste-to-energy plant in Hengelo, the Netherlands. The plant will reduce CO\(_2\) emissions associated with the generation of energy from the incineration of non-recyclable waste. The captured CO\(_2\) will be used in liquid form in the greenhouse horticulture sector. By the end of 2023, the
Just Catch™ standardized plant is expected to start capturing 100 ktonnes of CO₂ per year (Phillip & Yggeseth, 2022).

In the UK, the government published Cluster Sequencing Process in Track-1 and Track-2 as part of efforts to support a logical sequence of deployment for CCUS projects. In Phase-1 of the Track-1 Cluster Sequencing process, the UK government provisionally sequence those that are most suited to deployment in the mid-2020s. In Phase-2 of the Cluster Sequencing Process, government will receive applications from individual projects across capture applications (industry, power, hydrogen) to connect to the Track-1 clusters (BEIS, 2021). In 12 August 2022, the UK government shortlisted twenty CCUS projects that could receive funding support and proceed to the due diligence stage of the Phase-2 Cluster Sequencing process (BEIS, 2022c). Among these projects, four WtE projects are included (two in the East Coast Cluster and two in the Hynet cluster), they are:

- Tees Valley Energy Recovery Facility Project;
- Redcar Energy Centre from the East Coast Cluster;
- Viridor’s Runcorn Industrial CCS project;
- Protos Energy Recovery Facility.

These shortlisted projects are likely to be among the first of their kind in the UK to deploy carbon capture and storage technology for energy recovery, which was an important and significant stride on the journey towards a net zero recycling and waste management sector by 2040.

At the time of writing, examples of commercial scale carbon capture on existing WtE facilities are limited. However, there is a recognition of the importance of the technology in the sector due to the limited number of other options for decarbonisation of residual waste treatment. The European WtE sector has, therefore, seen increasing initiatives in recent years, including project planning. Key CO₂ capture technology vendors who have been involved in project development include Capsol, Shell, Aker Carbon Capture, C-Capture etc. At the time of writing, most of the projects are at early-stage engineering and feasibility study or in the process of applying subsidies for development. However, securing long-term financial support seems to be the most significant issue delaying the implementation of CO₂ capture at European WtE plants. For instance, the Hafslund Oslo Celsio project and the AEB WtE CCS plant are both faced with the high cost of adding CCS to the existing plants, which affected its development (CCS Norway, 2023; PWC, 2022). Table 2-1 below summarizes the emerging WtE with CCUS projects that are happening worldwide.
<table>
<thead>
<tr>
<th>Country</th>
<th>Plant</th>
<th>Waste Processed [kt/y]</th>
<th>Status</th>
<th>CO₂ Capture capacity [kt/y]</th>
<th>Capture rate</th>
<th>CO₂ capture Technology</th>
<th>Operational year</th>
<th>CCU or CCS</th>
<th>Reference</th>
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<td>100</td>
<td>CapsolEoP™ HPC</td>
<td>2025</td>
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<td>(KVA-Linth, 2020; Oslobors, 2023)</td>
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<td>Westenergy’s WOIMA® CCUS</td>
<td>20</td>
<td></td>
<td>20</td>
<td>CapsolEoP™ HPC</td>
<td>2023-2025</td>
<td></td>
<td>CCU and CCS</td>
<td>(Oslobors, 2023; WOIMA, 2023)</td>
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<td>Fiilbornaverket Energy-from-Waste plant</td>
<td>Demonstratio</td>
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<td>CapsolEoP™ HPC</td>
<td>2023-2025</td>
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<td>(Oslobors, 2023)</td>
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<td>Location</td>
<td>Project Description</td>
<td>Stage</td>
<td>Capacity</td>
<td>Technology</td>
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<td>Notes</td>
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<td>Nyborg, Denmark</td>
<td>Fortum's waste incineration</td>
<td>Feasibility study</td>
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<td>Aker Carbon Capture's Mobile Test Unit</td>
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<td>(Aker carbon capture, 2023)</td>
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<td>Pilot</td>
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<td>MEA based CO₂ capture</td>
<td>2025</td>
<td>CCS</td>
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<td>Shell CANSOLV® CO₂ Capture</td>
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<td>(Technip Energies, 2023)</td>
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<td>Hafslund Oslo Celsio</td>
<td>FEED Complete d; Temporarily halted by high costs</td>
<td>400</td>
<td>Shell Cansolv engineered and built by Technip</td>
<td>2026</td>
<td>(CCS Norway, 2023; IEA Bioenergy, 2023; IEAGHG, 2020)</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Location</td>
<td>Plant Description</td>
<td>Phase</td>
<td>Capacity</td>
<td>Name</td>
<td>Technology</td>
<td>Date</td>
<td>Notes</td>
<td></td>
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<td></td>
</tr>
<tr>
<td>9</td>
<td>Yokohama, Japan</td>
<td>Yokohama’s Tsurumi Waste-to-Energy Plant</td>
<td>Demonstration Testing</td>
<td>0.1</td>
<td>MHIENG’s KM CDR Process</td>
<td></td>
<td>(MHI, 2022)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>10</td>
<td>Saga, Japan</td>
<td>Saga City Incineration Plant</td>
<td>Operational</td>
<td>3.6</td>
<td>MEA</td>
<td>2016</td>
<td>(Yamada, Ii, Yamamoto, Ueda, &amp; Sakai, 2023)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>11</td>
<td>Runcorn, UK</td>
<td>Runcorn Waste to energy Plant</td>
<td>Pre-FEED phase completed by COWI</td>
<td>1000</td>
<td>900</td>
<td>2025</td>
<td>CCS</td>
<td>(Viridor, 2023)</td>
<td></td>
</tr>
<tr>
<td>12</td>
<td>Protos, UK</td>
<td>Protos Energy Recovery Facility</td>
<td>Pre-FEED</td>
<td>500</td>
<td>380</td>
<td>2024</td>
<td>CCS</td>
<td>(Protoserf, 2023)</td>
<td></td>
</tr>
<tr>
<td>13</td>
<td>Teesside, UK</td>
<td>SUEZ’ Haverton Hill EfW Plant</td>
<td>Pre-FEED</td>
<td>550</td>
<td></td>
<td></td>
<td></td>
<td>(SUEZ, 2021)</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Location</td>
<td>Plant Name</td>
<td>Stage</td>
<td>Capacity</td>
<td>Year</td>
<td>Source</td>
<td></td>
<td></td>
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<tr>
<td>14</td>
<td>Teesside, UK</td>
<td>SUEZ’ Wilton EfW Plant</td>
<td>Pre-FEED</td>
<td>350</td>
<td></td>
<td>(SUEZ, 2021)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>15</td>
<td>Teesside, UK</td>
<td>SUEZ’ Teeside Valley EfW Plant</td>
<td>Engineering stage, awaiting financing approval</td>
<td>240</td>
<td>2027</td>
<td>(SUEZ, 2021)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>16</td>
<td>Cheshire, UK</td>
<td>InBECCS project</td>
<td>Design complete</td>
<td>7</td>
<td></td>
<td>(Carbon capture journal, 2022; Global CCS Institute, 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>17</td>
<td>Duiven, Netherland</td>
<td>AVR’s Energy from Waste plant</td>
<td>In operation</td>
<td>100</td>
<td>2019</td>
<td>Greenhouse horticulture sector (IEAGHG, 2020; Ros et al., 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>18</td>
<td>Rozenburg, Netherland</td>
<td>AVR’s Energy from Waste plant</td>
<td>In developing</td>
<td>400</td>
<td></td>
<td>Carbon8 Systems’ solution (kasalsenergybron, 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Location</td>
<td>Project Name</td>
<td>Construction</td>
<td>CCS Technology</td>
<td>Commission Year</td>
<td>Notes</td>
<td></td>
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<tr>
<td>19</td>
<td>Hengelo, Netherlands</td>
<td>Twence’s WtE</td>
<td>100</td>
<td>Aker Carbon Capture, Catch™ modular</td>
<td>2023</td>
<td>(Aker carbon capture, 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>20</td>
<td>Amsterdam, Netherlands</td>
<td>AEB WtE plant</td>
<td>AEB applied for subsidy in 2021, but the application was rejected due to a substantial subsidy required.</td>
<td>450</td>
<td>2028</td>
<td>(PWC, 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>21</td>
<td>Portugal</td>
<td>LIPOR’s Maia Energy Recovery Plant</td>
<td>100</td>
<td></td>
<td></td>
<td>(Veolia, 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>22</td>
<td>Shenzhen, China</td>
<td>Pinghu WtE plant</td>
<td>Engineering Demonstration Project completed</td>
<td>95%</td>
<td>Longking, Amine based capture</td>
<td>(Longking, 2022)</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

a) All 29 Swiss WtE plants committed to CCS in the long-term (CEWEP, 2022b; Fabio Porettia & Ella Stengler, 2022)

b) Total incinerated waste can exceed permitted capacity, since the permit is sometimes based on the LHV of the waste. A lower LHV allows for more waste to be incinerated. The permit may also be based on an estimation of the total operational hours (M. de Leeuw & Koelemeijer, 2022).
2.4 Process modelling of CO₂ capture integration overview

2.4.1 Previous academic work on process modelling of PCC integration

CO₂ capture technologies, in one aspect can be seen as a set of independent gas processing technologies with high purity CO₂ as the final product; on the other hand, when applied into different industries, they need to be effectively integrated to the existing process minimizing the impacts to the overall performance. Through the development of CO₂ capture technology integrations, process modelling plays an important role to assess performance, optimize processes or to serve as input for further development analysis.

Different process modelling approaches can be found in literature. They are in general equilibrium-based models and rate-based models. Equilibrium based absorption models are based on the assumption of equilibrium at each stage and rate-based models are based on rate expressions for chemical reactions, mass transfer and heat transfer (Agarwal, Cao Nhien, & Lee, 2022; Nzihou et al., 2021; Øi, Sætre, & Hamborg, 2018). Rate based models was claimed superior to equilibrium based models because the rate-based models are capable of describing more detailed mechanisms, for instance, CO₂ loading, CO₂ removal efficiency, temperature profiles, etc. as claimed by (Ying Zhang, 2009); besides, higher accuracy and the prediction ability was reported by (Nzihou et al., 2021). However, rate-based models relying on accurate input parameters makes this method challenging in predicting performance (Øi et al., 2018). In the work by Øi et al, several sets of experimental data from the amine-based CO₂ capture process at CO₂ Technology Centre Mongstad (TCM) had been compared with simulations of different equilibrium-based models and a rate-based model. The results from their study showed that equilibrium and rate-based models perform equally well in both fitting performance data and in predicting performance at changed conditions (Øi et al., 2018).

Besides detailed process modelling, some study uses relatively low complexity level to represent the capture process. Sathre et al. did not model the heat integration of power plant with CO₂ capture process, but used the estimated the efficiency penalty as percentage drop in conversion efficiency at 90% capture rate, depending on the maturity of capture technology, but neglected the effect of CO₂ concentration from different fuel sources. (Sathre, Gustavsson, & Truong, 2017). Cumicheo et al. implemented steady state operation of combined cycle based processes in gPROMS (Cumicheo, Mac Dowell, & Shah, 2019) and it was in different context: the incorporation of biomass and CCS into gas fired power plants. They did not model the CO₂ capture process independently, but assumed a conservative 3.8 MJ/kgCO₂ of energy penalty due to the solvent regeneration and 90% of carbon capture efficiency.
Several process simulation platforms for CO₂ absorption into amine solutions are available and commonly used in the literature. Examples are the equilibrium based models in Aspen Plus (Haaf et al., 2020) and gProms (Mechleri, Lawal, Ramos, Davison, & Dowell, 2017) and the rate-based model in Aspen Plus (Kärki, Tsupari, & Arasto, 2013), Aspen RateSep™ (Plaza, Wagener, & Rochelle, 2009) and ProTreat™ (Brown et al., 2017). Some in-house are also developed such as the iCCS tool developed by SINTEF Energy Research (Roussanaly et al., 2020). A sum of the available modelling platforms in literature is shown in Table 2-2.

Table 2-2 A summary of the main modelling software/CO₂ capture rate/solvent for PCC with Power plant integration

<table>
<thead>
<tr>
<th>Number</th>
<th>Power plant Modelling</th>
<th>PCC modelling</th>
<th>Solvent</th>
<th>CO₂ capture rate</th>
<th>reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>EBSILON Professional</td>
<td>CHEMASIM steady-state simulations</td>
<td>30% MEA</td>
<td>90%</td>
<td>(Cifre et al., 2009)</td>
</tr>
<tr>
<td>2</td>
<td>-</td>
<td>Aspen HYSYS® v.10.</td>
<td>30% MEA</td>
<td>80.8%-81.2%</td>
<td>(Shirmohammadi, Aslani, Ghasempour, &amp; Romeo, 2020)</td>
</tr>
<tr>
<td>3</td>
<td>EBSILON®</td>
<td>ProTreat®</td>
<td>PZ, MDEA and AMP</td>
<td>90%</td>
<td>(Feron, Cousins, Jiang, Zhai, &amp; Garcia, 2020)</td>
</tr>
<tr>
<td>4</td>
<td>EBSILON® Professional</td>
<td>ASPEN Plus®</td>
<td>MEA</td>
<td>90%</td>
<td>(Pfaff, Oexmann, &amp; Kather, 2010)</td>
</tr>
<tr>
<td>5</td>
<td>gProms dynamic model</td>
<td>gProms dynamic model</td>
<td>30% MEA</td>
<td>90%</td>
<td>(Lawal, Wang, Stephenson, &amp; Obi, 2012)</td>
</tr>
<tr>
<td>6</td>
<td>gPROMS</td>
<td>gPROMS</td>
<td>30% MEA</td>
<td>90%</td>
<td>(Martínez et al., 2016)</td>
</tr>
<tr>
<td></td>
<td>Model</td>
<td>Description</td>
<td>MEA Range</td>
<td>CO₂ Capture Efficiency</td>
<td>Reference</td>
</tr>
<tr>
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<td>-------------------------------------------</td>
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<td>------------------------</td>
<td>-----------------------------------------------</td>
</tr>
<tr>
<td>7</td>
<td>gCCS, based on</td>
<td>gCCS, based on the gPROMS platform</td>
<td>30% MEA</td>
<td>90%</td>
<td>(Mechleri et al., 2017)</td>
</tr>
<tr>
<td></td>
<td>Modelica</td>
<td>dynamic process Modelica</td>
<td>30% MEA</td>
<td>90%</td>
<td>(M. Montañés, Garðarsdóttir, Normann, Johnsson, &amp; Nord, 2017)</td>
</tr>
<tr>
<td>10</td>
<td>Aspen Plus</td>
<td>Aspen Plus’ rate-based model</td>
<td>MEA-A07 data collected by KEPCO E&amp;C</td>
<td>90%</td>
<td>(Jeong, Jung, Lee, Yang, &amp; Han, 2015)</td>
</tr>
<tr>
<td>11</td>
<td>NGCC Aspen</td>
<td>Rate based models. the ENRTLKRK thermodynamic property package.</td>
<td>MEA</td>
<td>90%</td>
<td>(Ali et al., 2017)</td>
</tr>
</tbody>
</table>

In this study, considering the overall balance in terms of providing insights of capture process and reasonable computational workload, the design of CO₂ capture plant is modelled using a MEA Steady State Model developed by the U.S. Department of Energy’s Carbon Capture Simulation Initiative (CCSI), to characterize and quantify the plant data in each configuration.

2.4.2 Thermal integrations for WtE-CHP plants with CO₂ capture in the literature

While there is much research on BECCS in general terms, thermal integrations for CO₂ capture mostly follow the approaches applied for traditional fossil-fuel based power-only plant. There has been a limited number of publications so far that focus on the heat integration for CHP plants with CO₂ capture.
Among the few studies on WtE-CHP plant with CO₂ capture, Sweden and Norway are the two countries going the furthest.

Beiron et al. estimated the potential for negative CO₂ emissions from absorption-based carbon capture applied to 110 existing Swedish biomass or waste-fired combined CHP plants, operating in 78 local district heating systems. Results show that depending on the steam cycle power-to-heat ratio, 45-70% of district heat generation is retained when operating the CCS plant, but the reduction can be moderated at the expense of lowered electricity generation. The authors also point out that due to the impacts of CCS integration, the use of peak heat generation technologies must be evaluated, as well as the cost-effectiveness of CCS installations in CHP plants compared to other sources of negative emissions (Beiron, Normann, & Johnsson).

Levihn et al., using the case of Stockholm, provided insights to barriers and policy implications in relation to successful BECCS implementation. It investigated the heat recovery from the HPC (hot potassium carbonate) CO₂ capture technology for CHP plant, and showed that BECCS on CHP plants has a potential, but requires much more research. They suggested that negative emission technologies in energy systems are brought into research agendas such as the future of combined heat and power and the 4th generation district heating. (Levihn, Linde, Gustafsson, & Dahlen, 2019).

Jasmine Andersson investigated the viability of carbon capture technologies at WtE plants with a techno-economic analysis of the Sävenäs WtE plant in Gothenburg. They used the R1 formula to assess the overall energy performance for Sävenäs WtE plant. As also mentioned in the chapter 2.1.1, this R1 formula is defined by the waste framework directive (European Parliament and Council, 2018), and is a key factor to determine whether a incineration plant processing municipal waste should be classified as a WtE plant or a waste disposal facility. The results from their study finds that heat recovery options which reduces the final energy penalty from 32% to 12% and additional energy utilization will not jeopardize the position as a WtE plant according to R1 standard (Jasmine, 2020).

Linus (2016) investigates the technical and economic feasibility of BECCS in the district heating system of Stockholm region with a case study at the energy utility Fortum Värme. It concluded that the total cost for BECCS is calculated at 70-100€/tonne depending on size of emissions and distance to storage locations. The author pointed out that the major challenge of BECCS in combined heat and power production, compared to other studies based on power production, is the seasonality of heat demand so that optimization challenge exists (Linus, 2016).

Another master thesis by SIMON ÖBERG evaluates the possibilities of making the Swedish waste fired CHP plants CO₂ neutral by applying partial capture in order to capture the fossil share of the CO₂ emissions. It concluded that the size of the plant has an important impact on the cost of the capture
units according to the economy-of-scale concept. And capture unit should be optimized for low investment costs rather than low operation costs (SIMON, 2017).

Magnanelli et al. investigated different scenarios for heat integration of a CO₂ capture plant in WtE plant in Norway. They concluded that the preferred solutions are the ones meet the specific requirement of a WtE plant. Heat requirement of CO₂ capture corresponds to 27% of the nominal thermal capacity of the WtE plant. They found that the capture plant could use the unutilized heat between April and October without causing any penalty to the district heating production. However, in such a case, the CO₂ capture efficiency reduced from 85% to 47%. Heat recovery from the CO₂ capture plant was not investigated in their study (Magnanelli et al., 2021).

The lack of literature records on the WtE-CHP plants with CO₂ capture reveals the research gap in the development of this application, both in terms of research depth and application scope. For instance, expectations of high CO₂ capture rates (higher than 95%) are emerging in the research field (AECOM, 2021) and government guidelines (Gibbins & Lucquiaud, 2021), however there is no record of high CO₂ capture rates for the WtE plant integration in literature (except the paper published by the same author of this thesis (Su, Herraiz, Lucquiaud, Thomson, & Chalmers, 2023)). The heat recovery approach has been commonly modelled when a power plant integrates with PCC with the aim to improve energy performance of the integrated system, but a comprehensive assessment of its impact is rare.

This thesis is able to fill the gaps by modelling a series of thermal integration options for WtE plant with PCC and making performance assessments based on a complete set of defined KPIs. What’s more, although solvent storage has been studied in the literature on energy storage, the application of solvent storage into CHP plant integration with PCC plant presents a new application of solvent storage, which was developed by this work.

Another novelty is, key data from the modelling work (gProcess and ASPEN) is used as input data with an Excel-based model for solvent storage application for WtE-CHP plants with PCC, thus bringing another angle of view in terms of problem-solving for competing heat requirement for district heating and solvent regeneration overall, the modelling work is essential in this PhD thesis, and is used as tool that helps to investigate possible solutions for future thermal integrations of WtE plant with PCC.
3 Key performance indicators for the assessment of integrated systems

For the post-combustion based CO₂ capture investigated in this study, capturing CO₂ from the flue gas of an integrated WtE plant brings environmental benefit in forms of reduced CO₂ emission, but also requires energy and economic investment to deliver this process. Multiple-angle assessment should be undertaken for the investigated configurations of WtE with PCC to fully understand the pros and cons from multiple aspects. In this chapter, a series of KPIs (Key Performance Indicators) for thermodynamic, economic and environmental aspects are introduced for the purpose of performance assessments. These KPIs will also be used to compare the likely differences among all the investigated cases studies covered in this work.

3.1 Thermodynamic assessment metrics

For solvent-based CO₂ capture, the solvent will have to be regenerated to release the CO₂ as a concentrated stream for storage as well as for recycling of the solvent. The energy required for solvent regeneration will be provided by a major extraction of steam from the power plant. Steam extraction reduces power generation, while the additional auxiliary power demand will directly reduce the net electrical output, thus affecting the thermodynamic performance of the plant.

In this section, a series of performance indicators, which are developed based on thermodynamics, are identified to provide valuable insights on the energy-efficiency and penalty of such a system. Considering the difference of power-only and CHP plant, the thermodynamic KPIs are separately identified for both types of WtE plants, namely Efficiency penalty (EP) and Electricity output penalty (EOP) for WtE-power-only plants, Energy utilization penalty (EUP) and Energy utilization factor penalty (EUFP) for WtE-CHP plants. Besides, in order to assess the efficiency of the steam extraction for solvent regeneration and potential possibility to replace this approach by heat pump, the most relevant KPI coefficient of performance for steam extraction is developed in this study. They can provide reference for future thermodynamic improvement of this application.

3.1.1 EP and EOP for WtE-power only plants with PCC

The power and energy consumption for CO₂ capture is mainly the thermal heat requirement for solvent regeneration to purify CO₂ rich solvent combined with the power requirement for driving pumps and CO₂ compression. These requirements effectively reduce the overall power and heat production from the corresponding abated plant, in the form of a power/heat output penalty. The influence of CO₂ capture on system thermal performance is normally assessed by efficiency penalty (EP) and electricity output penalty (EOP). The efficiency penalty is evaluated as a percentage point drop in the overall thermal efficiency of the plant, so the fall in electricity output per unit of fuel energy input (Lucquiaud & Gibbins, 2011). The efficiency penalty of WtE plant due to CO₂ capture and
compression is shown in Equation 3.1. For a power only WtE plant with CO₂ capture, the electrical efficiency is evaluated as the ratio of the net power output divided by the waste fuel thermal input on LHV basis.

\[ \Delta \eta = \eta_{\text{ref}} - \eta_{\text{CCS}} = \frac{W_{e,\text{ref}}}{Q_{\text{fuel input}}} - \frac{W_{e,\text{CCS}}}{Q_{\text{fuel input}}} \text{ [\%]} \]  

Equation 3.1

Where:

\( \Delta \eta \) Efficiency penalty of CO₂ capture

\( \eta_{\text{ref}}, \eta_{\text{CCS}} \) Electrical efficiency without and with CCS

\( Q_{\text{fuel input}} \) Fuel thermal input on LHV basis [MWh]

The efficiency penalty is, however, affected by fuel composition since the amount of CO₂ generated by combustion per unit of useful thermal energy, and hence the total energy requirement for capture and compression, varies depending on the ratio of carbon content to heating value. For power-only plant, electricity output penalty (EOP), is reported to be a better metric comparing EP to assess capture technologies independently of the fuel composition (Lucquiaud & Gibbins, 2011). It is evaluated as the total net loss in power output due to the CO₂ capture process divided by the mass flow of compressed CO₂ exiting the plant boundaries (Lucquiaud & Gibbins, 2011), as shown in Equation 3.2.

\[ EOP = \frac{W_{e,\text{ref}} - W_{e,\text{CCS}}}{\dot{m}_{\text{CO₂, captured}}} \text{ [MWht] \ CO₂} \]  

Equation 3.2

Where:

\( W_{e,\text{ref}} \) Net power output without CCS [MWe]

\( W_{e,\text{CCS}} \) Net power output with CCS, considering reduction in steam turbine power output due to steam extraction for solvent regeneration, power consumption associated with the CO₂ capture process and any offsets due to beneficial heat recovery for condensate heating and other purposes [MWe];

\( \dot{m}_{\text{CO₂, captured}} \) Mass flow of CO₂ captured exiting the plant boundaries [t/h]

3.1.2 EUP and EUFP for WtE-CHP plants with PCC

In order to differentiate the output penalty of PCC on WtE-CHP plant, compared with the metrics mentioned in previous section, similar approaches are applied to the abated WtE-CHP plant with PCC, namely, Energy utilization penalty (EUP) and Energy utilization factor penalty (EUFP). The EUP is
evaluated as the total net loss in power and heat output due to the CO₂ capture process divided by the mass flow of compressed CO₂ exiting the plant boundaries, as shown in Equation 3.3.

\[
\text{EUP} = \left( \frac{W_{e,\text{ref}} + Q_{\text{th,\text{DH},\text{ref}}}}{m_{\text{CO}_2,\text{captured}}} \right) - \left( \frac{W_{e,\text{CCS}} + Q_{\text{th,\text{DH},\text{CCS}}}}{m_{\text{CO}_2,\text{captured}}} \right) \left[ \frac{kW h_{\text{el&th}}}{t CO_2} \right] \quad \text{Equation 3.3}
\]

Where:

\( Q_{\text{th,\text{DH}}} \) Net useful thermal output for district heating, subscripts ‘\text{ref}’ and ‘\text{CCS}’ represents under reference case and CCS case, respectively \[\text{MWh}_\text{th}\];

For WtE-CHP plant, the EUF is evaluated as the ratio of the total energy output, including net useful electrical output and net useful thermal output, divided by the total fuel thermal input on LHV basis, as shown in Equation 3.4. This parameter facilitates the performance comparison of a CHP plant with and without CO₂ capture, but it does not differentiate between the exergy level of power and heat.

\[
\text{EUF} = \frac{W_e + Q_{\text{th,\text{DH}}}}{Q_{\text{fuel,input}}} \quad [\%] \quad \text{Equation 3.4}
\]

The energy utilisation factor penalty (\( \text{EUF} \)) is calculated as in Equation 3.5.

\[
\text{EUFP} = \text{EUF}_{\text{ref}} - \text{EUF}_{\text{CCS}} = \frac{W_{e,\text{ref}} + Q_{\text{th,\text{DH},\text{ref}}}}{Q_{\text{fuel,input}}} - \frac{W_{e,\text{CCS}} + Q_{\text{th,\text{DH},\text{CCS}}}}{Q_{\text{fuel,input}}} \quad [\%] \quad \text{Equation 3.5}
\]

A summary of the four metrics for WtE-power-only and WtE-CHP plant is provided in Figure 3-1. They are mutually related by the fuel specific emissions, which can be calculated using Equation 3.6 and Equation 3.7 as shown below. This implies for different WtE plants with PCC, even when the fuel specific emissions are the same, different insights can be gained when different metrics are used. For example, for the same WtE plant operating at different modes in a year, the PCC system may bring varied penalty depending on the capture rates, heat output variations, etc. A range of relevant metrics should be considered to make a proper decision/comparison.
Figure 3-1 EOP, EP, EUP, EUFP due to PCC which are related with fuel specific emissions

\[
\text{Fuel specific emissions} = \frac{\dot{m}_{\text{CO}_2, \text{produced}}}{Q_{\text{fuel input}}} = \frac{\dot{m}_{\text{CO}_2, \text{captured}}}{CCR \times Q_{\text{fuel input}}} \left[ \frac{\text{tCO}_2}{\text{MW}_{\text{th}}} \right]
\]

Equation 3.6

\[
CCR = \frac{\dot{m}_{\text{CO}_2, \text{captured}}}{\dot{m}_{\text{CO}_2, \text{produced}}}
\]

Equation 3.7

*In this study, it is assumed that all the flue gas from the WtE plant is treated in the CO\(_2\) capture plant, no part load operation is considered, so the CO\(_2\) capture rate (CCR) is essentially the CO\(_2\) capture efficiency of the CO\(_2\) capture plant. In real operational cases, it may happen that the part of flue gas is treated, under this situation, a percentage of \(\dot{m}_{\text{CO}_2, \text{produced}}\) should be used as the denominator for CCR.

3.1.3 Coefficient of Performance for (steam) extraction

The KPIs introduced in the previous section mainly focus on the thermodynamic impacts of CO\(_2\) capture (and compression) on the WtE plants, in energy (power and/or heat) output per unit of fuel input and/or per unit of CO\(_2\) captured, since these KPIs are originally from the assessment metric of the traditional power system itself. However, it is still necessary to have metrics that help to assess the performance of this CO\(_2\) capture method itself, which is steam extraction from the plant to provide thermal heat for solvent regeneration. In this context, the concept of coefficient of performance for (steam) extraction (\(\text{COP}_X\)) is developed by Lucquiaud & Gibbins (2011). This concept is firstly applied for power-only plant with PCC, where the steam extraction from the steam cycle for solvent
regeneration results in a loss of power production of the plant. In other words, this can be regarded as analogous to ‘consumption’ of electricity (actually foregone production) to achieve heat production for heating in the reboiler (Lucquiaud & Gibbins, 2011). This metric can be analogously compared to the coefficient of performance ($COP_{hp}$) a heat pump, which is essentially the ratio of heat output divided by the power consumption.

In this study, the same metric is used for WtE-power-only plant with PCC, where the $COP_{X\_cap}$ is calculated as the ratio of heat supplied to a PCC plant by steam extracted from a steam cycle to the power ‘consumption’ (reduction) from the steam cycle, as shown in Equation 3.8.

$$COP_{X\_cap} = \frac{Q_{cap}}{W_T}$$  \hspace{1cm} \text{Equation 3.8}

Where:

- $COP_{X\_cap}$: Coefficient performance of steam extraction (MWth/MWe)
- $Q_{cap}$: Heat supplied to solvent reboiler (MWth)
- $W_T$: Loss of power output due to steam extraction (MWe)

In literature, the attempt to use heat pumps to provide thermal heat for solvent regeneration is still rare. On the one hand, this could be because the development of the CO$_2$ capture technologies is still not far enough to test the potential of the heat pump application. On the other hand, it may be because of limitations of the heat pump technology itself. The most commercially available amine-based CO$_2$ capture relies on saturated thermal heat at a relatively high temperature of around 130°C, which could be challenging for traditional heat pump, whose performance is sensitive to the supply temperature. In this context, the $COP_{X\_cap}$ can be helpful to assess the feasibility of using heat pumps to provide thermal heat for solvent regeneration. In other words, if the $COP_{X\_cap}$ is higher than that of a heat pump, steam extraction would be a better choice for thermal heat regeneration, and vice versa. Figure 3-2 explains the concept of $COP_{X\_cap}$ and its comparison with the $COP_{hp}$ of a traditional electricity-driven heat pump.
Figure 3-2 Schematic illustration of $COP_{X,\, cap}$ for WtE-power-only plant with PCC and the analogous comparison with the $COP_{hp}$ of an electricity-driven heat pump consuming the same amount of power $W_T$ (lost of electricity output due to steam extraction)

It can be seen from the Figure 3-2, assuming the same amount of power consumption $W_T$, the comparison of $COP_{X,\, cap}$ and $COP_{hp}$ depends on the comparison of $Q_{cap}$ and $COP_{hp}$, both in terms of quantity and quality. By quantity, it refers to the amount of heat required for solvent regeneration, thus the reboiler duty of the capture plant; by quality, it refers to the regeneration temperature required by the capture plant. Since the COP of heat pump is quite sensitive to the temperature lift (the temperature difference between the condenser and the evaporator), additional attention should be given in terms of whether this analogous heat pump is feasible. For instance, the typical MEA based solvent regeneration requires saturated steam of around 130°C. This kind of ultra-high temperature
heat pump is largely not commercially available and the \( \text{COP}_{hp} \) of the existing heat pumps would be quite low. Advanced solvents that require lower regeneration temperature (lower than around 80°C that is largely available in the heat pump industry) may be sensible for this comparison, since one of the ultimate goals of the COP comparison is to explore whether a heat pump solution could be an alternative approach to supply regeneration heat, other than traditional steam extraction. In this context, the analogous comparison heat pump should be as reliable as possible, in order to give a practical comparison.

Additionally, this study expands the same concept to the application of CHP plant with PCC. For WtE-CHP plant, steam extraction from the steam cycle for district heating and solvent regeneration results in a loss of power production of the plant. In other words, this can also be regarded as an analogous ‘consumption’ of electricity (actually foregone production) to achieve heat production for heating in the district heating heat exchangers and \( \text{CO}_2 \) capture reboiler. The concept of \( C\text{OP}_{X,\text{cap}} \) can then be expanded to the Coefficient Performance of steam extraction, for thermal integration of CHP plant with PCC: \( C\text{OP}_{X,\text{cap}+\text{dh}} \), as is defined by Equation 3.9. Similarly as the \( C\text{OP}_{X,\text{cap}} \) for power-only plant with PCC, the \( C\text{OP}_{X,\text{cap}+\text{dh}} \) can also be equivalent to the COP of a heat pump, as is illustratively shown in Figure 3-3. However, it should be noted that there are two temperature levels at the numerator, which are around 80°C and 130°C for district heating and for solvent regeneration, respectively. This is an important issue to consider since the calculation of \( C\text{OP}_{X,\text{cap}+\text{dh}} \) introduces two additional variables, quantity and quality of heat for district heating, where there is much uncertainty involved. Further analysis of this uncertainty should be given for future applications.

\[
C\text{OP}_{X,\text{cap}+\text{dh}} = \frac{Q_{\text{cap}} + Q_{\text{dh, pcc}}}{W_T}
\]

Equation 3.9

Where:

\( C\text{OP}_{X,\text{cap}+\text{dh}} \) Coefficient performance of steam extraction for CHP plants (MW\text{th}/MW\text{e})

\( Q_{\text{dh, pcc}} \) The district heating output (MW\text{th})

\( W_T \) Loss of power output due to steam extraction (MW\text{e})
Considering the two temperature levels involved in the previous COP calculation, in this study, an absorption heat pump is added to address this problem. Under this approach, the addition of PCC on a CHP plant will reduce the heat and power output from the original plant, in value of $Q_{dh\_loss}$ and $W_T$. A hypothetical electricity driven heat pump is assumed to utilize the ‘lost’ power $W_T$ to provide heat $Q_{hp\_el}$ at 80°C. Then a high-temperature absorption heat pump is used to recover both the heat $Q_{hp\_el}$ and the lost district heat $Q_{dh\_loss}$ and to increase the temperature level to the same as the reboiler at 130°C, as illustrated in Figure 3-4 below:
Figure 3-4 Schematic illustration of $COP_{X_{abs}}$ for WtE_CHP plant with PCC, with the series application of electricity-driven heat pump and absorption heat pump, analogous consuming the lost power output $W_T$ and lost district heat output $Q_{dh\_loss}$.

Under this situation, with the two hypothetical heat pumps in the integration, an analogous $COP_{X_{abs}}$ can be defined based on the COP theory of an absorption heat pump, as expressed in Equation 3.10.

$$COP_{X_{abs}} = \frac{Q_{cap}}{Q_{dh\_loss} + Q_{hp\_el}}$$  \hspace{1cm} \text{Equation 3.10}

Where:

$COP_{X_{abs}}$ An analogous COP of absorption heat pump in the CHP integration;

$Q_{dh\_loss}$ The change of district heating output due to CO$_2$ capture (MW$_{th}$)

$Q_{hp\_el}$ The amount of heat can be provided from an electricity driven heat pump, consuming the ‘lost’ power (MW$_{th}$) $W_T$;

Now the question is then transferred to compare the $COP_{X_{abs}}$ of this hypothetical absorption heat pump with conventional absorption heat pump, consuming the same amount of waste heat ($Q_{dh\_loss} + Q_{hp\_el}$). Results and the corresponding discussion of the comparison will be reported in chapter 6.4.1 and chapter 7.2.1.

3.2 Economic assessment metrics

Economic feasibility is one of the important considerations to drive the deployment of CO$_2$ capture technology. In order to assess the economic feasibility, economic indicators are usually used so that
audiences can be more informed about the costs of CO₂ emission mitigation options. They are important for decision-making and performing comparison between different branches of technologies. This section introduces a set of economic KPIs that will be used to perform economic assessment of WtE with PCC. The selection of economic indicators considers the common language and methodology for costing, together with transparency of methods and assumptions. They are levelized cost of electricity (LCOE), levelized cost of heat (LCOH), cost of CO₂ captured.

3.2.1 Levelized cost of electricity (LCOE)

The levelized cost of electricity is one of the most widely used parameters when carrying out an economic evaluation of CCS technologies. It allows the plant owners and decision makers to identify the average price of electricity required for a power plant where the revenues equal costs. To be more specific, the amount of money is required per MWh of electricity to recoup the lifetime costs involved in constructing and operating a power plant.

In this study, the LCOE is applied to WtE power only plant without and with CO₂ capture. It is calculated according to Equation 3.11 (BEIS, 2020a):

\[
LCOE = \frac{NPV \text{ of Total Cost}}{NPV \text{ of Electricity Generation}} \tag{3.11}
\]

In the above equation of LCOE, the NPV (Net Present Value) approach is chosen as it presents a number of advantages, namely accounting for the future depreciation of investment, the possibility of future investment return analysis, and the fact that it can take account of actual value assets (such as capital costs).

Note: The above equation may find that the Electricity generation is being discounted, however this is actually the result of the algebra carried through this formula:

\[
NPV \text{ of } (LCOE \times \text{Electricity Generation}) = NPV \text{ of Total cost}
\]

Or this formula:

\[
\sum_{n=1}^{N} \frac{\text{Electricity Generation} \times LCOE}{(1 + d)^n} = NPV \text{ of Total Cost}
\]

In the above formula, the LCOE is assigned to every unit of electricity produced by the system over the analysis period, and the total discounted value will equal the Total Cost when discounted back to the base year. So, essentially, the Equation 3.11 is not discounting the Electricity Generation, but discounting the payment received per unit of electricity generated in form of LCOE. A relevant explanation can also be found in (Short, Packey, and Holt 1995).
The sum of the NPV of total expected costs (CAPEX and OPEX) for each year is then calculated using Equation 3.12:

\[
NPV \text{ of Total Costs} = \sum_{n} \frac{CAPEX_n + OPEX_n}{(1 + \text{Discount Rate})^n}
\]

Equation 3.12

Where:

\(CAPEX_n\): Capital expenditure of total cost at year \(n\);

\(OPEX_n\): Operational expenditure at year \(n\);

Similarly, the Electricity generation and heat generation are discounted using the NPV value at the year of 2021:

\[
NPV \text{ of Electricity Generation} = \sum_{n} \frac{\text{Net Electricity Generation}_n}{(1 + \text{Discount Rate})^n}
\]

Equation 3.13

Where:

\(n\): Time period of 25 years; WtE incinerators are capital intensive and are usually designed for 20 to 30 year lifespans, an average 25 year is chosen for this study (UNECE, 2020);

Discount rate: a value of 8% is assumed for all cases. This real discount rate of 8% corresponds to a nominal discount rate of around 10% if an inflation rate of 2% is considered.

3.2.2 Levelized cost of heat (LCOH)

Under WtE-CHP scenarios, a similar approach is applied to calculate the levelized cost of heat. The electricity price is assumed to be maintained at the LCOE under WtE-power-only without PCC here, thus, to identify the various district heat prices required for a WtE-CHP plant making the revenues equal costs. The \(NPV\) of Heat Generation follows similar calculation method as that of the \(NPV\) of Electricity Generation, as shown in Equation 3.14. The calculation of LCOH is shown in the Equation 3.15 below:

\[
NPV \text{ of Heat Generation} = \sum_{n} \frac{\text{Net Heat Generation}_n}{(1 + \text{Discount Rate})^n}
\]

Equation 3.14

\[
LCOH = \frac{NPV \text{ of Total Cost}}{NPV \text{ of Heat Generation}}
\]

Equation 3.15
3.2.3 Cost of CO₂ captured

In literature, the “cost of CO₂ avoided” and the “cost of CO₂ captured” both exist and are the most widely used economic metric to evaluate CCS (Roussanaly et al., 2021).

The cost of avoided CO₂ refers to the cost of CO₂ capture system as a carbon mitigation option so usually a full CCS chain is considered including CO₂ capture, transport, and storage. It accounts for the actual amount of CO₂ avoided by the CO₂ capture plant (i.e. considering the CO₂ generated because of the extra energy requirement of the CO₂ capture) (Mathieu and Bolland 2013).

On the other hand, the cost of captured CO₂ provides insights into the investments and operating costs associated with the CO₂ capture (or the CCS chain). If the captured CO₂ is not permanently stored, as highlighted by Roussanaly and Rubin et al., the correct term for the cost metric calculation should be the ‘cost of CO₂ captured’ (Roussanaly, 2019; Rubin et al., 2013). In this PhD study, due to the high level of variability in transport and storage costs – across geologic, geographic, and institutional settings (that is hard to characterize), the cost of CCS does not include the cost of transport and storage. The cost of CO₂ captured is applied as one of the economic KPIs. It quantifies only the cost of capturing (producing) CO₂ ready for transport to geological storage. For power-only generation, the calculation of the cost of CO₂ captured follows the equation from IEAGHG (IEAGHG, 2017a) and is shown in the Equation 3.16:

\[ \text{Cost of CO}_2 \text{captured} = \frac{LCOE_{\text{ccs}} - LCOE_{\text{ref}}}{(tCO_2/MWhe)_{\text{captured}}} \left[\frac{E}{tCO_2}\right] \]  

Equation 3.16

Where:

\((tCO_2/MWhe)_{\text{captured}}\) represents the total mass CO₂ captured per net electricity generation MWh for the plant with CO₂ capture (equal to the CO₂ produced minus emitted);

Following the same method, for CHP plant where the district heat is the main product of the plant, the Cost of CO₂ captured can be defined based on heat generation, as in Equation 3.17:

\[ \text{Cost of CO}_2 \text{captured} = \frac{LCOH_{\text{ccs}} - LCOH_{\text{ref}}}{(tCO_2/MWth)_{\text{captured}}} \left[\frac{E}{tCO_2}\right] \]  

Equation 3.17

Where:

\((tCO_2/MWth)_{\text{captured}}\) represents the total mass CO₂ captured per net thermal heat generation MWh for the plant with CO₂ capture (equal to the CO₂ produced minus emitted).
Environmental assessment metrics

This chapter introduces the environmental KPIs identified to assess the environmental benefits of integrating CO₂ capture and compression to a group of WtE plants, focusing on direct CO₂ emission reduction per unit of fuel/energy output. Each of them represents the CO₂ emission intensity for a different point of view (electrical output, thermal output and fuel input). When evaluating the CO₂ emission of WtE plants, the three aspects should be reported together to give a comprehensive assessment. There is uncertainty in terms of the other emissions from the WtE plants (emissions due to PCC, material uses, etc.), and the avoided emissions by the energy production (substituted power and heat production otherwise may use alternative technologies). These emissions are both needed to estimate the overall net climate benefit of a wider system and however are not the focus of this study. In this study, the selection of environmental KPIs represent a more straightforward assessment of the absolute amount of CO₂ reduction from the abated WtE plant itself, instead of looking at the overall net climate benefit.

With the global commitments to reduce greenhouse gas (GHG) emissions, many countries and sectors have adopted comprehensive legislations/activities to facilitate the energy transition toward low carbon energy systems. The UK Government has a five point plan to tackle climate change in 2009 (The UK government, 2009) and published the Net Zero Strategy in 2021 which sets out the next steps to fasten the process of delivering net zero (the UK government, 2021). All of these activities require up to date assessment methods to report and verify their effectiveness in various aspects. Among the assessment metrics, one of the most commonly used environmental metrics is carbon emission intensity. It is the amount of GHG emissions per unit of energy (or product) production.

Although theoretically speaking, the denominator for the carbon emission intensity varies depending on the main product of each sector; carbon emission intensity on electricity basis is the dominantly used. For example, in March 2020 the EU Technical Expert Group on Sustainable Finance (TEG) published its recommendations for an EU Taxonomy for Sustainable Activities. A key feature of the recommendations around electricity generation was a “substantial contribution” emissions threshold of 100g CO₂e/kWh. This is the limit on the intensity of greenhouse gas (GHG) emissions produced from the generation of electricity, heat and power from hydropower, geothermal energy or gaseous and liquid fuels (European Commission, 2020).

As previously mentioned, around 50% of WtE plants in Europe operate as CHP plants. It is, therefore, important to select GHG emission metrics that are suitable for CHP plants. Theoretically speaking, the allocation of CO₂ emission to either electricity output or heat output depends on whether/which is the main product of the WtE plant. For example, the carbon footprint metric can be applied for
decarbonisation heat and heating technology, which is measured in grams of carbon dioxide equivalent per kilowatt-hour of heat (gCO₂eq/kWh). In this case, heat production is the main product for the heat industry so an emission metric on heat basis is applied. The Committee on Climate Change (CCC) is the UK Government’s statutory adviser on carbon targets and suggests that average emissions from domestic and commercial heat should be reduced from around 220 gCO₂eq/kWh in 2015 to 180 gCO₂eq/kWh in 2030 and closer to zero by 2050 (House of Parliament, 2016).

In order to assess the potential for negative emissions of the CCS WtE sector, in this study, the carbon emission intensity is calculated on the basis of both electrical and thermal output. The biogenic carbon content in the referenced MSW incinerated is not available at the time of the writing and a 60% biogenic to total carbon ratio is assumed in this study, based on a literature review in the public domain (GIOUSE, 2020; Kaza & Woerden, 2018; REMONDIS, 2014) (Giouse et al. 2020; Manders 2009).

The calculation of CO₂ emission intensity on electricity and thermal basis requires the allocation of CO₂ emission to each of them. In literature, various methods can be found to allocate GHG emissions between electricity and heat produced in CHP plants (Grażyna Rabczuk, Jarosław Łosiński, & Cenian, 2020; Tereshchenko & Nord, 2015). Among them, the most commonly used is the efficiency approach (or alternative generation method) (BEIS, 2022a; Grażyna Rabczuk et al., 2020; Orchard, 2013; Tereshchenko & Nord, 2015). In this method, the share of CO₂ emission is allocated in proportion to the fuel needed to produce the same amount of heat or power in separate plants, which can be expressed as:

\[
f_Q = \frac{Q/\eta_{heat}}{(Q/\eta_{heat} + E/\eta_{power})}
\]  
Equation 3.18

\[
f_E = \frac{E/\eta_{power}}{(Q/\eta_{heat} + E/\eta_{power})}
\]  
Equation 3.19

Where:

- \(f_Q\): The percentage of CO₂ emission allocated to heat production;
- \(f_E\): The percentage of CO₂ emission allocated to power production;
- \(Q\): Heat production from the CHP plant;
- \(E\): Power production from the CHP plant;
- \(\eta_{heat}\): Heat generation efficiency of in separate plants;
\( \eta_{power} \): Power generation efficiency of in separate plants;

Following the efficiency allocation approaches, the DUKES (Digest of UK Energy Statistics) ‘1/3:2/3’ method is used, which assumes that twice as many units of fuel are required to produce a unit of electricity than that for a unit of heat. A fixed ratio of 2 is given for the heat and power generation efficiency (BEIS, 2016). This method assumes that twice as many units of fuel are required to produce a unit of electricity than that for a unit of heat. It has been selected as it is the method used under the UK Climate Change Agreements (CCAs) (Dwyer, 2016). The CO₂ emission intensity on electricity and heat basis can be calculated as:

\[
\text{Carbon intensity}_{\text{electrical}} = \frac{(\text{Fossil CO₂ emitted} - \text{Biogenic CO₂ captured}) \times f_E}{\text{Net power output}} \quad \text{Equation 3.20}
\]

\[
\text{Carbon intensity}_{\text{thermal}} = \frac{(\text{Fossil CO₂ emitted} - \text{Biogenic CO₂ captured}) \times f_Q}{\text{Net thermal output}} \quad \text{Equation 3.21}
\]

Besides the above-mentioned CO₂ emission intensity, the main priority for many WtE plants is waste treatment. The CO₂ emission intensity of a WtE plant is, therefore, also widely reported by unit of waste treated (CEWEP, 2022a; ESA, 2020). In this case, the CO₂ emission intensity on waste treatment basis can be expressed as:

\[
\text{Carbon intensity}_{\text{fuel}} = \frac{\text{Fossil CO₂ emitted} - \text{Biogenic CO₂ captured}}{\text{Amount of MSW input (tons)}} \quad \text{Equation 3.22}
\]

As an alternative to the DUKES method, the work presented here could allow direct computation of the efficiency in units of electricity produced per unit of heat, since \( f_Q \) and \( f_E \) were both estimated. However, this complicates comparison with work published in the literature. Nevertheless, it can be informative to compare the results of \( f_Q/f_E \) in the discussion in this piece of work.

3.4 Summary of KPIs for assessment of WtE with PCC

This chapter proposes a comprehensive set of KPIs specific to assess the performance of CO₂ integration into the investigated WtE configurations, namely thermodynamic, economic, and environmental aspects. These KPIs represents a standard toolkit for the following analysis and reporting of CO₂ capture integration scenarios in this study. The detailed KPIs results and analysis are also given in each corresponding chapter.
The selection of thermodynamic KPIs is broadly differentiated into WtE power-only and CHP plant, respectively. The result of the KPIs will offer a holistic view of the performance and potential improvement of the PCC integration into the representative WtE plant, in the different scenarios developed in this study.

The selection of economic KPIs with caution is important for the applicability of the results analysis, and brings better understanding for future plant operators and policy makers in assessing the economic feasibility of this technology. Detailed calculation method for the cost items will be covered in Chapter 4.

The Environmental KPIs are focused on assessing the absolute direct carbon emissions, in form of direct carbon intensities before and after CO₂ capture; for the CHP plants, CO₂ emissions are separately allocated to heat and power generation, results can be used to compare with the direct carbon emission intensity of alternative technologies.
Modelling methodology for waste to energy plants with Post-combustion CO₂ capture

This chapter introduces the process modelling activities around the thermal integration of a generic WtE plant with post-combustion CO₂ capture. The modelling work lays the foundation of this research, so it tries to model the most up to date and most representative types of integration approaches for the possible application of the technology.

The overall modelling is undertaken in gProcess V2.0.0. It features a generic medium size WtE plant using direct combustion over a moving grate, with a constant 500t/d MSW consumption (55MW thermal energy input). Three operating scenarios for a WtE plant are considered:

a) ‘Power-only’ represents WtE plants with power only operation
b) ‘CHP-Ex&C’ represents WtE plants with steam extraction and a condensing steam turbine
c) ‘CHP-BP’ represents WtE plants with a back pressure steam turbine

These three operational modes also represent the varied heat supply requirement from the WtE plant, in other words, heat to power ratio. During summer, the plant operates as power-only, with heat to power ratio of zero. During spring & autumn season, the WtE operates as steam extraction and condensing, to provide medium amount of heat for the district heating. During winter season, when there is high heat demand, the WtE plant operates as backpressure configuration to fully recover the heat that was previously discharged into the air cooler. The detailed description of the gProcess modelling and its validation is provided in chapter 4.1.3.

For the three configurations, an amine-based carbon capture system is built in ASPEN Plus V10 using a 35%wt monoethanolamine (MEA) aqueous solution to ensure a relatively low specific reboiler duty (SRD) at high CO₂ capture efficiency (Hasan, Abbas, & Nasr, 2020). In the modelled CO₂ capture plant, three CO₂ capture rates are considered: 90%, 95% and 99.72%. The choice of 99.72% CO₂ capture rate represents a scenario when all the CO₂ that comes into the boiler through combustion air is equal to the CO₂ emitted from the system (uncaptured CO₂). This capture rates enables theoretical zero residual emission from the process. The concept of zero residual emission is valuable in the era of Net-zero emissions, where as much as CO₂ should be captured in a cost-effective way. The detailed description of the CO₂ capture and compression modelling is the focus of chapter 4.2. The results from the modelled capture plant in ASPEN, such as SRD, power consumption of pumps, and the CO₂ conditions at the stripper overhead condenser will be used as input in the gProcess model. The captured CO₂ from the capture plant will undergo 3-stage compression trains that is also included in the gProcess model.
Additionally, three thermal integration options are considered for each CO₂ capture rate: one base case heat integration and two advanced-heat integrations: advanced heat recovery and interim solvent storage, which are both approaches that aim to improve the performance of a WtE plant that is connected to district heating with CO₂ capture. The description and results of the base case thermal integration will be presented in Chapter 6. The two advanced thermal integration methods will be discussed in Chapter 7 and Chapter 8 respectively.

Figure 4-1 Summarises the process modelling approaches

4.1 Modelling approaches for typical WtE plants

This section describes the modelling of representative waste-to-energy technologies, especially thermal treatment of waste by means of combustion in grate-based systems which has gained worldwide acceptance as the preferred method for sustainable management and safe disposal of residual waste. The design inputs of the modelled WtE plants are based on the data from public domain/literature and was trying to represent the off design condition of a generic WtE plant (Industrial Boiler Design Calculation Standard Method Editorial Board, 2003; Liu, Luo, Yao, Li, & Wang, 2020; Magnanelli, Tranås, Carlsson, Mosby, & Becidan, 2020; Mu, Saffarzadeh, & Shimaoka, 2017; Prabir Basu, Cen Kefa, & Louis Jestin, 2000).
Three WtE configurations are considered in this study as generic configurations and are developed in gProcess V2.0.0; an equation-oriented modelling platform that allows the creation of customised models for each operation. Peng-Robinson equation of state is used for gas mixtures and Steam Tables (IAPWS-95) is used for water and steam available in Multiflash V6.1.

The reference WtE plant is assumed to operate as base load with a contractual obligation to process 500 t/day of MSW throughput. This waste treatment capacity is in the medium range of the common WtE plants. The schematic representation of the reference WtE plant as modelled in gProcess is shown in Figure 4-2. It includes air-preheating, the waste incineration furnace and grate, flue gas passes for steam generation, steam turbines for power production, a heat exchanger for heat production, and the flue gas cleaning systems.
Figure 4-2 Schematic layout of the referenced WtE plant
The MSW is processed by direct combustion over a moving grate using 50% excess air to ensure complete combustion. Combustion air is added in two stages. Primary air (ca. 70% of the total air) is preheated to 150 °C by steam extracted from the HP turbine and supplied through the grate layer into the fuel bed. Secondary air (ca. 30% of the total airflow) is preheated using grate cooling water up to 50 °C and supplied over the grate layer (air over fire). The excess air is calculated so that the oxygen concentration in the exhaust flue gas is within the range of 6% vol to 9% vol (dry basis).

The heat released from the waste incineration is used to generate superheated steam at 400 °C and 60 bar, which is sent to the steam turbine train for power generation. The steam turbine consists of three cylinders with one steam extraction point at 6bar for the deaerator, and a subsequent steam extraction point at 4 bar to supply the district heating system and the reboiler for solvent regeneration when operating with CO₂ capture.

The flue gas exiting the heat recovery section in the boiler goes through a series of flue gas treatment processes to remove acid gases and other harmful components. Nitrogen oxide emissions are removed using a selective non-catalytic reduction process with injection of aqueous ammonia. A series of bag filters are used to reduce the particle matter concentration to the allowed emission level. After the flue-gas-treatment process, the temperature of the flue gas exiting the air-pollution-control system is 136 °C, based on real operational data from a Recycling and Energy Recovery Centre (RERC) operated by FCC Environment in Edinburgh (personal communication with FCC).

Up-stream of the DCC in the CO₂ capture plant, a gas-gas rotary heat exchanger is used to recover heat from the flue gas and increase the temperature of the CO₂-depleted gas from the absorber, ensuring adequate gas buoyancy and dispersion in the atmosphere.

The ultimate composition of the MSW in as received condition based on sampling in March 2019 was taken from the RERC data provided by FCC Environment, and it was published in (Su et al., 2023) by the same authors.
### Table 4-1 MSW Ultimate composition (as received) based on sampling in March 2019 (through personal communication with FCC)

<table>
<thead>
<tr>
<th>Analyte</th>
<th>Units</th>
<th>Results</th>
</tr>
</thead>
<tbody>
<tr>
<td>Moisture</td>
<td>% Wt</td>
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</tr>
<tr>
<td>Ash</td>
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</tr>
<tr>
<td>Gross CV</td>
<td>MJ/kg</td>
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</tr>
<tr>
<td>Net CV</td>
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<td>Chlorine</td>
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</tbody>
</table>

The following sub-sections introduces the approaches and findings through the process of modelling of the WtE plants in gProcess, including boiler heat balance calculation, configurations of the investigated WtE plants and their validation.

#### 4.1.1 Basis of boiler balance calculation

In the boiler, heat is realised by fuel combustion to heat up the returned feed water into steam. For a variety of reasons, the fuel does not burn completely. The heat released cannot be fully utilized and heat loss is unavoidable. A heat balance shows how much heat is effectively used and how much of it is wasted. The purpose of a heat balance is to identify the sources of heat loss and to find means to reduce them and thereby improve boiler efficiency (Prabir Basu et al., 2000).

The method used for the boiler heat balance calculation mainly references from the method introduced by Prabir Basu et al, and will be introduced in this section, including description/equations of calculation of:

a) Heat into the boiler: preheated air/sensible heat of fuel/LHV;
b) Heat losses: through flue gas, unburned carbon, convection and radiation of furnace exterior, sensible heat of slag; and

c) Boiler efficiency heat balance equation.

A schematic representation of heat balance of a boiler is shown in Figure 4-3.

Figure 4-3 Schematic representation of heat balance of a boiler (Prabir Basu et al., 2000)

Figure 4-3 schematically shows the heat balance of a fired furnace. In a steady state the heat input, heat utilization, and heat losses in 1 kg fuel may be given as

\[
Q = Q_1 + Q_2 + Q_3 + Q_4 + Q_5 + Q_6 \quad \text{[kJ per kg fuel]}
\]

Equation 4.1

Where

\( Q = \) Total heat input to the boiler

\( Q_1 = \) Heat absorbed by steam (utilized by the boiler)

\( Q_2 = \) Heat loss through stack gas

\( Q_3 = \) Heat loss by incomplete combustion of gaseous components

\( Q_4 = \) Heat loss owing to unburned carbon
\( Q_5 \) = Heat loss owing to convection and radiation from the furnace exterior

\( Q_6 \) = Heat loss through the sensible heat of ash and slag

To express the heat losses in percentages we divide both sides of Equation 4.1 by total heat input to the boiler, leading to

\[
100 = q_1 + q_2 + q_3 + q_4 + q_5 + q_6
\]

Equation 4.2

Where

\[
q_i = \frac{Q_i}{Q} \times 100
\]

Represents the percentage of heat losses to the total available heat by the fuel.

Boiler efficiency, \( \eta_{\text{boiler}} = q_1 = \frac{Q_1}{Q} \times 100 \) is defined by the ratio of heat absorbed by the boiler and the heat provided by the fuel. It is given by

\[
\eta_{\text{boiler}} = q_1 = \frac{Q_1}{Q} \times 100 = 100 - q_2 - q_3 - q_4 - q_5 - q_6
\]

Equation 4.3

**\( Q \) = Total heat input to the boiler**

The total heat input to the boiler is given by

\[
Q = LHV + H_f + Q_{air} \quad [kJ/kg]
\]

Equation 4.4

Where \( LHV = \) lower heating value of fuel, as received basis, kJ/kg

\( H_f = \) sensible heat of fuel

\( Q_{air} = \) sensible heat carried by the air when heated by external air heater

The sensible heat of fuel \( H_f \) is given by

\[
H_f = C_{pf} \times T_f \quad [kJ/kg]
\]

Equation 4.5

Where \( C_{pf} = \) specific heat of fuel, as received basis, kJ/kg°C

\( T_f = \) fuel temperature at burner or feeder exit, assigned to be 15°C, before preheating of the fuel, that is in as received condition;

Specific heat of fuel \( C_{pf} \) is calculated based on the Equation 4.6:

\[
C_{pf} = C_{pf}^{\text{dry}} \times \frac{100 - M}{100} + \frac{M}{100} \times 4.186 \quad [kJ/kg]
\]

Equation 4.6
Where $C_{pf}^{dry}$ specific heat, as dry basis, kJ/kg. $C_{pf}^{dry}$ is assigned to be 0.2467 kJ/kg in the process modelling, this is the value based on the equation presented by Manjunatha et al for 7 tested MSW samples, where moisture content is zero (Manjunatha et al., 2020); $M$ is the moisture percentage of the fuel.

Note that the sensible heat of fuel, $H_{f}$, may be neglected if the fuel is not preheated, but if its moisture percentage, $M \geq Q_{ar}/628\%$, it is necessary to consider the $H_{f}$ (Wu, 2008).

When the combustion air is not preheated by an external air heater, $Q_{air} = 0$. When the combustion air is preheated by an external air heater, the sensible heat carried by the air when heated by an external air heater is calculated

$$Q_{air} = a \times (h_{entrance} - h_{ambient}) \left[ \frac{kJ}{kg} \right]$$

Equation 4.7

Where

$h_{entrance}$ and $h_{ambient}$: are enthalpies of air at burner entrance and at ambient conditions, in kJ/kg, respectively.

$a$ is the ratio of mass flow of air to the mass flow of fuel into the boiler;

**Heat loss $Q_2 = \text{Heat loss through stack gas}$**

The enthalpy of flue gas leaving the boiler is higher than that of combustion air entering the boiler. So, there is a net heat loss. This loss is given by

$$q_2 = \frac{Q_2}{Q} \times 100 = \left( I_g - a_{ah} I_0^a \right) \frac{100 - q_4}{100} \times 100\% \ [\%]$$

Equation 4.8

Where

$I_g$ = flue gas enthalpy at the exit of air heater, kJ/kg fuel

$I_0^a$ = theoretical cold air \(^1\) enthalpy entering boiler, kJ/kg fuel

(The reference state for enthalpy and sensible heat is 25°C, 1 bar)

---

\(^1\) The ‘theoretical’ refers to the amount of air required in the combustion of a unit mass of MSW calculated from chemical reaction. By comparison, due to imperfect mixing, and to ensure complete combustion as well as good emission control (low CO, VOCs, particulates, etc.), reactions require more than this amount of the air. That’s why in Equation 4.8, the excess air coefficient $a_{ah}$ is applied to express the actual amount of air enters the boiler.
\( a_{ah} = \) excess air coefficient at the exit of air heater

\[
\frac{100 - q_4}{100} = \text{correction factor owing to the difference between calculated and actual fuel consumption.}
\]

This loss \((q_2)\) increases with increase in exit flue gas temperature. Generally, \(q_2\) increases by 1 \% when the exit flue gas temperature increases by 12-15°C (Manjunatha et al., 2020). In this case, it is desirable to reduce the exit gas temperature as much as possible. However, in reality, the flue gas temperature should also consider the material cost of the heat exchanger area, and avoid corrosion. In one aspect, exit flue gas is at the rear part of the stack, reducing the flue gas temperature will result into lower temperature difference in the heat exchange area, thus increase the surface area required for the stack; on the other hand, flue gas from the WtE plant main contain sulphur. The flue gas from the combustion of a high sulphur fuel would have a higher dew point. When the exit flue-gas temperature is below the dew point, the sulphur dioxide of the gas deposits as sulphuric acid leading to corrosion of metals in the air heater, so that higher temperature is advised to avoid surface corrosion. Therefore, a boiler designed for this fuel would require a higher exit gas temperature.

Typical flue gases outlet temperatures for WtE power plants range from 180 to 250 °C, significantly higher than those typical for fossil fuel power plants (Branchini, 2012). Values equal or below 160 °C can only be achieved in advanced plants those performs well both in terms of performance (waste treatment and electricity generation) and in terms of pollutants reduction (Branchini, 2012; Prabir Basu et al., 2000).

**Heat loss \( Q_3 \) = Heat loss by incomplete combustion of gaseous components**

Heat loss owing to incomplete combustion is caused by escape of combustible gases, viz., CO, H\(_2\), CH\(_4\), with the flue gas. The incomplete combustion loss \((Q_3)\) is generally small (Prabir Basu et al., 2000). In the process modelling reported in this thesis, the assumption is that the combustion products are CO\(_2\), H\(_2\)O, SO\(_2\), NO\(_2\), N\(_2\), so this heat loss by incomplete combustion of gaseous components is regarded as zero under complete combustion.

**Heat loss \( Q_4 \) = Heat loss owing to unburned carbon**

Unburned carbon is present in the bottom (bed drain) ash and in the fly ash. So, the total heat loss owing to unburned carbon \((Q_4)\) is the sum of carbon in fly ash \((Q_{fa})\) and bottom ash \((Q_{ba})\) multiplied by their heating values. The heat loss \( q_4 \) as percentage of heat input is then

\[
q_4 = \frac{Q_4}{Q} \times 100 = \frac{32,866 \left( G_{ba} C_{ba} + G_{fa} C_{fa} \right)}{100B} \times 100\% \quad [\%]
\]

Equation 4.9
Where

\( G_{ba} \) amount of bottom ash produced, kg/s;

\( G_{fa} \) amount of fly ash produced, kg/s;

\( C_{ba}, C_{fa} \) carbon contents in bottom ash and fly ash respectively, %;

\( B = \) fuel consumption, kg/s;

Calorific value of carbon is 32,866 kJ/kg;

Ash leaves the bed either through bed drain or fly ash. We define \( X_{ba} \) and \( X_{fa} \) as the mass fraction of total ash exiting through the bottom ash and fly ash, respectively. An ash balance is carried out to calculate the amounts of fly ash:

\[
X_{ba} + X_{fa} = 1
\]  
Equation 4.10

The individual ash fraction can be found as follows:

\[
X_{ba} = \frac{G_{ba}(100 - C_{ba})}{BA} \text{ [%]}
\]  
Equation 4.11

\[
X_{fa} = \frac{G_{fa}(100 - C_{fa})}{BA} \text{ [%]}
\]  
Equation 4.12

Where \( A \) is the ash percentage in fuel.

Substituting into Equation 4.9 we get

\[
q_4 = \frac{32,866 \times A}{Q} \left[ \frac{X_{ba}C_{ba}}{100 - C_{ba}} + \frac{(1 - X_{ba})C_{fa}}{100 - C_{fa}} \right] \text{ [%]}
\]  
Equation 4.13

At the boiler design stage, the following values of \( q_4 \) may be taken from experience, as shown below.

The heat loss owing to unburned carbon depends on the types of fuel, furnace and firing equipment construction, boiler load, operating conditions, furnace temperature, and the air-fuel mixture. The main composition of bottom ash is mainly determined by the composition of the MSW burned.

Literature review and information from various publicly available resources has been performed to understand the ranges of bottom ash and fly ash mass content. It was found that the values vary by the locations of plant, fuel types and even different time of year. In the process modelling reported in
this thesis, \(X_{ba}\) bottom ash fraction of the total ash is assigned to be 90% and the \(C_{ba}\), \(C_{fa}\) carbon contents in bottom ash and fly ash, are assigned to be 1.83% and 3.7% respectively, based on the average values of Millerhill RERC from their annual performance report 2019 (through personal communication) (Su et al., 2023).

\[ Q_5 = \text{Heat loss owing to convection and radiation from the furnace exterior} \]

When a boiler is in operation, the external surface temperature of the furnace, flue gas ducts, steam tubes, and headers is higher than that of the ambient. The heat loss is caused by heat transfer from the surfaces to ambient through convection and radiation. The heat loss primarily depends on the surface area of furnace wall, insulating layer of tubes, and ambient temperature. The heat loss is calculated by

\[
Q_5 = \frac{\sum F_{sb}}{B} \times (h_c + h_r) \times (T_{sb} - T_0) \quad \left[ \frac{kJ}{kg} \right] \quad \text{Equation 4.14}
\]

\[
q_5 = \frac{Q_5}{Q} \times 100 \quad \left[ \% \right] \quad \text{Equation 4.15}
\]

Where:

\(\sum F_{sb}\) external surface area of boiler exposed to the ambient, \(m^2\);

\(h_c\) and \(h_r\), coefficient of heat transfer by convection and by radiation respectively, \(kW/(m^2\ °C)\);

\(T_{sb}\) = average temperature of the boiler surface, °C

\(T_0\) = average ambient temperature, °C

\(B\) = fuel consumption in kg/s

Due to limited information from the waste boiler on the heat loss owing to convection and radiation from the furnace exterior, \(q_5\) is assigned to be 1.7% based on broad public available resource for typical coal fired boilers, as shown in Figure 4-4 (Prabir Basu et al., 2000) and Table 4-2 (Wu, 2008).
Figure 4-4 Radiative and convective heat losses from the external surface of typical pulverized coal fired boilers (Wu, 2008)

Table 4-2 Heat loss owing to convection and radiation from the furnace exterior for steam generation boiler

<table>
<thead>
<tr>
<th>Rated steam generation D(t/h)</th>
<th>4</th>
<th>6</th>
<th>10</th>
<th>15</th>
<th>20</th>
<th>35</th>
<th>65</th>
</tr>
</thead>
<tbody>
<tr>
<td>With rear heated surface²</td>
<td>2.9</td>
<td>2.4</td>
<td>1.7</td>
<td>1.5</td>
<td>1.3</td>
<td>1.0</td>
<td>0.8</td>
</tr>
<tr>
<td>Without rear heated surface</td>
<td>2.1</td>
<td>1.5</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
</tbody>
</table>

\[ Q_6 = \text{Heat loss through the sensible heat of ash and slag} \]

When solid fuel is burned, ash and slag leave the furnace at a rather high temperature (about 600°C - 800°C). This results in sensible heat loss in ash and slag. The heat loss depends on fuel ash content, fuel heating value, and slag deposition method. For the high ash content and low heating value that are typical for MSW fuels, this loss, \( Q_6 \), should be considered as well.

The heat loss is calculated by

² Here the ‘rear heated surfaces’ refers to ‘air heater and economizer’, since the heat transfer happens at the surfaces of ‘air heater and economizer’, additional heat loss may happen.
Where:

$C_{pas}$ Specific heat of ash and slag, kJ/kg·K

$X_{ba}$ Mass fraction of bottom ash of the total ash

$C_{ba}$ Unburned carbon content in bottom ash

$A$ Ash percentage in fuel

$T_{slag}$ Temperature of the slag that leaves the furnace, °C

The temperature of ash and slag leave the furnace could be about 600 - 800°C. The Specific heat of ash and slag $C_{pas}$ is calculated based on specific heat capacity of each material in the slag, using sing the rule of mixtures.

$$C_{pas,mixture} = \sum wt\%_n \times C_{p_n}$$

Equation 4.17

Where

$wt\%$ is the mass percentage of each material in the slag;

$C_p$ is the specific heat capacity of each material in the slag;

The mass percentage of each material is referenced from the mean mass percentage of the three main materials (SiO$_2$, CaO and Al$_2$O$_3$) from the 145 municipal incinerated bottom ash samples, as reported by (Lynn, Ghataora, & Dhir Obe, 2017).

The assigned slag temperature and resulted specific heat capacity of slag in this study is 700°C and 1.084 J/kg·K, which aims to represent a generic value that is suitable for the use of this study. In real work and calculation, these values should be checked according to the specific fuel materials applied, and operation conditions of the boiler, etc.

Table 4-3 Specific heat capacity of the three key materials in the slag, @700°C

<table>
<thead>
<tr>
<th>Material</th>
<th>Mean mass percentage in the slag</th>
<th>Specific heat capacity</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>%wt</td>
<td>J/kg·K</td>
<td></td>
</tr>
<tr>
<td>SiO$_2$</td>
<td>37.9</td>
<td>1145</td>
<td>(NIST, n.d.-b)</td>
</tr>
</tbody>
</table>
CaO | 22.6 | 919 | (NIST, n.d.-a)  
Al₂O₃ | 10.4 | 1216 | (Matweb, n.d.)

4.1.2 Investigated steam cycle configurations of WtE plants

This section describes the investigated configurations of WtE power plant, and the key operating parameters (pressure, temperature); air preheating; cooling type; electric efficiency, etc under each integration scenario:

- Power only;
- CHP type with steam extraction and condensing turbine (CHP-Ex&C);
- CHP type with Back pressure turbine (CHP-BP).

The technology most widely used for energy recovery from MSW is direct combustion over a moving grate, with the generation of superheated steam feeding a steam turbine train for power generation (Wienchol, Szłęk, & Ditaranto, 2020). Based on a collection of WtE facility data from an extensive literature review (Barba, Brandani, Capocelli, Luberti, & Zizza, 2015; Branchini, 2015; Eboh, Ahlström, & Richards, 2019; Lombardi & Carnevale, 2018), three configurations of the WtE plant representative of the WtE facilities in operation in Europe are considered in this thesis:

a) In the power-only configuration, the WtE plant produces only electricity, and the steam turbine train comprises a condensing steam turbine where superheated steam expands from 60 bar to a condenser pressure of 0.1 bar, considered in this work for an air-cooling system. This configuration is representative of WtE plants which are not connected to a DH network including, but not limited to, plants that are built as DH ready (i.e. they will provide DH in the future) or CHP WtE plants currently connected to a DH network operating during periods when DH demand is zero (e.g. summer).

b) In the CHP-Ex&C configuration, the WtE plant produces electricity and thermal energy for the DH system. The steam turbine train consists of a high-pressure (HP) ST cylinder, intermediate-pressure (IP) ST cylinder and a condensing low pressure (LP) ST cylinder connected to an air-cooler condenser. Superheated steam expands from 60 bar to 4 bar in the HP steam turbine cylinder, a fraction of steam is then extracted to supply thermal energy to the DH system, and the remaining steam expands in the LP steam turbine cylinder from 4 bar to the condenser pressure (i.e. 0.1 bar). When the steam is extracted from the crossover pipeline between IP-LP turbines, at the last turbine blade, the pressure at the lower part of the blade will be too low, which will cause too low flow velocities. This phenomenon will set-off the fairly large-scale re-circulation zone(s) where work is being fed into the
“trapped steam” ³ causing heating – and lost shaft work (Thern, Jordal, & Genrup, 2014). Additional water spraying is usually required to cool the last turbine blade when operating below 10-15% percent load (Thern et al., 2014). It is assumed that the minimum flow rate through the LP steam turbine cylinder is 15% of the nominal flow rate at full load, to cool down the blades and avoid overheating by churning. In this configuration, it is assumed that DH supply take priority over power supply, a constant steam extraction to the DH system is assumed with and without PCC and, thus, additional steam extraction for CO₂ capture will only penalise the electricity output.

c) In the CHP-BP configuration, the WtE plant produces electricity and thermal energy for the DH system. The steam turbine train consists of a HP ST cylinder and a backpressure LP ST cylinder. Superheated steam expands from 60 bar to 4 bar and it is then sent to the DH system. When PCC is implemented, part of the steam is sent to the reboiler of the CO₂ capture plant. In reality, the steam turbine train can be designed with a Synchro-Self-Shifting (SSS) clutch, which could be used to decouple the HP and the LP cylinders so that no minimum steam flow rate is required through the LP cylinder (Berry & Attix, 2017). This operation scenario is representative of WtE plants with high DH demand and particularly CHP WtE plants located in regions with long cold winters.

Figure 4-5 shows a schematic representation of the three WtE plant configurations. For all three configurations, the power-only configuration is used as the base case plant. For example, the swallowing capacity of the LP steam turbine cylinder for power-only operation is also used for the other operating configurations. When PCC is on, a throttling valve located downstream of the extraction point is used to maintain the IP/LP crossover pressure at the required level. Detailed process flow of the three WtE steam cycle configurations please refer to Appendix 4.1-4.3.

³ The ‘trapped steam’ at the last few stages of the turbine, as result of too low velocities of the LP, causes ‘churning’.
4.1.3 Validation of the WtE plant modelling in gProcess

The ‘power-only’ configuration of the generic WtE-plant modelling results is compared with the real operational data of the WtE plant in Edinburgh RERC, given as much the same input design data as possible, for instance, fuel input condition and designed live steam condition, condenser ambient air temperature etc. The key values comparison is shown in the Table 4-4 below. The results show an acceptable agreement between the gProcess model and the operational WtE plant.

Table 4-4 key values comparison of operational WtE plant in Edinburgh and modelling data

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Waste throughput</td>
<td>t/h</td>
<td>19.4</td>
<td>19.4</td>
</tr>
<tr>
<td>LHV of waste</td>
<td>MJ/kg</td>
<td>9.26</td>
<td>9.26</td>
</tr>
<tr>
<td>Boiler Live steam Temperature</td>
<td>°C</td>
<td>400</td>
<td>400</td>
</tr>
<tr>
<td>Boiler Live steam pressure</td>
<td>bar</td>
<td>60</td>
<td>60</td>
</tr>
<tr>
<td>Exhaust steam pressure turbine</td>
<td>bara</td>
<td>0.1</td>
<td>0.1</td>
</tr>
<tr>
<td>Flue gas temperature after fabric filter</td>
<td>°C</td>
<td>134</td>
<td>136</td>
</tr>
</tbody>
</table>

Notes:
HP: High pressure turbine
IP: Intermediate pressure turbine
LP: Lower pressure turbine
FWH: Feed water heater
SSS clutch: Synchro-self-shifting clutch which allows the LP Turbine to be shut down when not needed, avoiding minimum steam flow waste
### Table: Performance Data

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Vol-%, wet</th>
<th>6.2</th>
<th>5.9</th>
<th>4.8%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flue gas oxygen concentration</td>
<td></td>
<td>6.2</td>
<td>5.9</td>
<td>4.8%</td>
</tr>
<tr>
<td>Gross Power output</td>
<td>MW</td>
<td>14</td>
<td>14.3</td>
<td>2%</td>
</tr>
<tr>
<td>Boiler Live steam mass flowrate</td>
<td>kg/s</td>
<td>17.6</td>
<td>17.6</td>
<td>0%</td>
</tr>
<tr>
<td>Total combustion air throughput</td>
<td>kg/s</td>
<td>28.7</td>
<td>29.6</td>
<td>3%</td>
</tr>
<tr>
<td>Low pressure steam mass flowrate at the outlet of the last turbine</td>
<td>kg/s</td>
<td>14.3</td>
<td>14.6</td>
<td>2%</td>
</tr>
</tbody>
</table>

[1] Average data of Edinburgh RERC from 11\textsuperscript{th} - 17\textsuperscript{th} February 2020 (operational data available at the time of the writing);

At the time of the writing, there is no district heat output from the Edinburgh WtE plant, so the comparison of the modelled WtE-CHP plant with an operational one is not available. However, the CHP model is implemented in the same validated ‘Power-only’ plant model and heat and mass balance is performed under each CHP scenarios, so the CHP models can also be considered to be accurate with high confidence levels when reporting results in this study.

4.2 Capture plant modelling in ASPEN

4.2.1 Description of post-combustion capture

This section describes the operational process of the conventional MEA based CO\textsubscript{2} capture process, which is modelled as the PCC plant for this study. The process flow diagram for the conventional MEA based chemical absorption and desorption process is illustrated in Figure 4-6.
Figure 4-6 Process flow diagram for the conventional MEA based chemical absorption and desorption process (IEAGHG, 2014)

The flue gas temperature entering the absorber has to be adapted to the specific solvent. The selection is generally a trade-off between the kinetics and the thermodynamics. The reactions kinetic rate improves with increasing temperature, while a lower temperature displaces the thermodynamic equilibrium to a higher CO\(_2\) loading and enhances the solvent capacity.

Moreover, the flue gas typically enters the absorber saturated in moisture in order to minimise water evaporation from the solvent and possible solvent carry over. A temperature of 40-60°C at the inlet of the absorber is typically chosen for MEA (Bravo et al. 2021).

In the absorber, the flue gas is brought into direct contact with the solvent. The CO\(_2\) is chemically bound to the amine solvent and the CO\(_2\) loaded solvent (rich solvent) leaves the column at the bottom. The CO\(_2\)-depleted gas is cooled down in the water wash section at the top of the absorber to avoid solvent vaporization and solvent losses into the atmosphere. The flue gas temperatures at the inlet and at the outlet of the absorber column are similar to maintain the water balance in the system.

The rich solvent is reheated in the lean-rich heat exchanger and enters the top of the stripper column where it is thermally regenerated. The steam generated in the reboiler strips off the CO\(_2\) from the solvent. The vapour phase leaving the top of the stripper column is condensed at 40°C, a trade-off
between solvent reaction rates and solvent CO₂ capacity. The condensed liquid is separated from the gas phase in a flash vessel and recycled back to the stripper at the top stage (reflux). The CO₂-rich gas, with a CO₂ concentration >99 vol%, is compressed, liquefied and pumped up to 110 bar for transport and storage. The regenerated solvent (lean solvent) returns to the absorber at the first stage, after being cooled down to 45°C, first in the lean-rich heat exchanger and then in the lean solvent cooler.

As the reaction kinetics between CO₂ and MEA are relatively fast, the CO₂ mass transfer is the limiting step in the chemical absorption process. The rate of mass transfer depends on the overall mass transfer coefficient and on the available driving force, which is directly proportional to the CO₂ partial pressure in the gas phase. The overall mass transfer coefficient varies with process parameters, including gas flow rate, liquid flow rate, CO₂ partial pressure, MEA concentration, CO₂ loading and interfacial area (Razi, Svendsen, & Bolland, 2014). When the CO₂ absorption rate is limited by the rate of mass transfer, the solvent CO₂ loading is lower than the corresponding value in equilibrium with the actual CO₂ partial pressure in the gas, decreasing the solvent capacity. This fact has a negative impact on the specific reboiler duty since more solvent is required to capture the same amount of CO₂, and in the packing volume, since a larger surface contact area is required to achieve a certain CO₂ absorption efficiency. The resistance to mass transfer is taken into account by rigorously modelling the chemical absorption process using a rate-based model.

4.2.2 Capture plant optimization

The CO₂ capture process is modelled separately in ASPEN Plus V10. The modelling methodology for the design and optimisation of the PCC process is illustrated in Figure 4-7.

![Figure 4-7 PCC plant modelling and optimization approach in ASPEN plus](image-url)
The CO₂ capture plant is designed to process the total amount of flue gas exiting the WtE plant, i.e. a flue gas flow rate of 30.7kg/s with a CO₂ concentration of 11.1%vol downstream the flue gas flow from the DCC.

The overall CO₂ capture efficiency is defined as the amount of CO₂ captured for transport and storage/utilisation relative to the amount of CO₂ generated in the combustion of the waste. In this study, the operating and design parameters are evaluated for three CO₂ capture rates from 90% to 99.72%, taking as base case the design of a CO₂ capture process at 95% CO₂ capture rate.

The choice of 90% capture rate is because that at the time of writing, much of the literature is still focused on capture rate, so the 90% capture rate is included in this study for the future comparison with wider literature results.

In the UK, recent guidelines published by the UK Environmental Agency for permitting new post-combustion CO₂ capture plants for gas and biomass power plants require a design CO₂ capture rate of at least 95% to be achieved for an environmental permit to be approved (Gibbins & Lucquiaud, 2021). A capture level of 95% is then increasingly being favourable in the literature, for instance, in the recent report by AECOM, the capture level of 95% is used in all the benchmark scenarios of the next generation capture technologies (AECOM, 2021).

The upper limit of this range (99.72% capture efficiency) is selected to represent an ‘ultra-high’ CO₂ capture rate that achieves zero-direct CO₂ emissions. In this scenario all the CO₂ produced in the combustion of the waste fuel is captured and the remaining CO₂ corresponds to the amount of CO₂ entering the process with the combustion air. For the WtE plant treating 500 t/d of MSW, 5.168 kg/s of CO₂ exits the plant in the combustion gases, of which 99.72% (5.152kg/s) is captured in the PCC plant and the remaining 0.28% CO₂ (0.0145kg/s) corresponds to atmospheric CO₂ entering the plant with the combustion air.

In the 2nd step, the initial design parameters are considered in three sections: absorber, cross flow heat exchanger (XFHE) and regeneration.

In the absorption section, cooled flue gas downstream of the DCC is contacted with the solvent to remove CO₂ in a packed column. The absorber is a structured packing using Sulzer Standard Mellapak Plus 252Y with 35% monoethanolamine (MEA) aqueous solution. Higher concentrations of MEA are reported to have lower SRD, especially at high CO₂ capture efficiency (e.g. a test campaign carried out at the Pilot Scale Advanced Capture Technology (PACT) facilities of the UK Carbon Capture and Storage Research Centre (UKCCSRC) reported in (Akram, Milkowski, Gibbins, & Pourkashanian, 2020)). The absorber column diameter is approximated using a method proposed by (Chapel Dan, Ernest John, & Carl, 1999) and a lean loading of 0.16 mol CO₂/mol MEA is initially assigned.
In the XFHE section, CO₂ ‘rich’ solvent downstream of the rich solvent pump (initially assigned 900kpa to avoid flashing) after the absorber is heated against CO₂ ‘lean’ solvent in the crossflow heat exchanger before entering the stripper. The approach temperatures in the XFHE are specified to be 10°C, which is commonly used for CO₂ capture facilities (Andersson, Franck, & Berntsson, 2013; Tait et al., 2018).

Finally, in the regeneration section the CO₂ in the rich solvent is ‘stripped off’ by the condensing heat provided through the low-pressure saturated steam from the reboiler. The stripper operating pressure is designed to be 210 kPa with regeneration temperature around 125°C to prevent thermal degradation, i.e. polymerisation. The conditions in the stripper are set to achieve the required CO₂ capture efficiency, under initial given lean solvent loadings. The temperature of the condensing steam in the reboiler exceeds the solvent temperature by the pinch temperature (10°C) assumed in the reboiler design.

In the 3rd step, PCC plant simulation is performed with all the initial design parameters determined in the previous step. A range of absorber packing height is simulated for a constant diameter up to a value at which a further increase results in a marginal gain in the rich solvent CO₂ loading and in a marginal reduction of the SRD. The design absorber packing size (especially packing height), also needs to consider practical design limits that ensure efficient liquid dispersion and gas/liquid interactions within the packed bed to achieve the specified CO₂ capture efficiency.

Through the process of PCC plant simulation, a safety factor, a percentage of the flooding velocity (the gas velocity at the point of flooding), or named flooding point, is used when sizing packed columns to ensure the operation of the packed columns remains within the design range (Li et al., 2021; Maćkowiak, 2010). In literature, the percentage of flooding velocity can be found in a range of 30% - 80% (Billet & Schultes, 1999; Nittaya, Douglas, Croiset, & Ricardez-Sandoval, 2014; Otitoju, Oko, & Wang, 2020; tontiwachwuthikul, 1990). Increasing the surface area of tower packing by increasing stripper diameter will result in lowering of the percentage of flooding capacity. The flooding point is maintained to be less than 80% in this study.

Finally, in the optimisation step, the optimum lean solvent flow rate entering the absorber which minimises the SRD is evaluated for each CO₂ capture efficiency. Other parameters assigned during the initial input stage such as absorber intercooling, stripper pressure, etc., can also be further optimised to allow the minimized SRD for the specified CO₂ capture rates to be identified. The sizing and optimisation results of the PCC plant are presented in chapter 6.3.
4.2.3 Validation of PCC plant modelling in ASPEN

The CO₂ capture plant model is built upon a MEA steady state model developed under the U.S. Department of Energy’s Carbon Capture Simulation Initiative (CCSI). The steady state MEA Model is representative of the configuration of the Pilot Solvent Test Unit (PSTU) at the National Carbon Capture Center (NCCC) in Wilsonville, Alabama, which was rigorously planned to cover a large range of operating conditions for both the absorber and stripper columns. Steady-state data have been obtained from the NCCC for validation of this model (CCSI, 2021; Morgan et al., 2018). In this work, the PCC plant design parameters are tuned for the purpose of the specific design data of this study (Flue gas conditions, CO₂ capture rate, 35% MEA solvent, etc.). The validation of the original model using large pilot plant data brings high confidence levels in the ASPEN model results. The validation of the original ASPEN model is performed over a wide range of operating and has been validated satisfactorily with NCCC pilot plant data from the 2014 campaign, with five test configurations for the absorber column, including with and without intercooling, two beds with and without intercooling, and one bed without intercooling (DOE 2017).

4.3 Approaches for PCC plant integrated in the WtE plants

The above modelling approaches will be used as tool to simulate the possible thermal integrations of WtE plants with PCC. In order to understand the effects of CO₂ capture on system performance, in the gProcess model, the CO₂ capture plant is treated as a grey box. The gProcess model provides the flue gas composition/flow rate, which are necessary inputs into the ASPEN model. The ASPEN model then finds the amount of heat required to capture a certain amount of CO₂ in the flue gas and, thus, the amount of steam extraction required. The required steam extraction is then used in the gProcess. Both the gProms model and the ASPEN model are steady state, for each system integration option with values inputs/outputs, the corresponding data are checked to ensure consistency.

In this study, in general, three main approaches are considered; they are base case thermal integration, advanced heat integration and solvent storage application. A brief description of each integration approach is included here, for detailed information please refer to chapter 6, chapter 7 and chapter 8, respectively.

- Base case integration of WtE plants with PCC

The base case integration of WtE plant with PCC applies the most common approach of CO₂ capture integration into a power plant, with minimum retrofitting to the existing plant. In this integration, heat required for solvent regeneration of the capture plant is extracted from the steam cycle by which heat/power output penalty of the original plant will occur. This approach does not look into methods to improve the system performance under this integration, thus represent a baseline approach of the
investigated WtE configurations with MEA based CO₂ capture, with three CO₂ capture rates. Detailed information is described in Chapter 6.

- Advanced heat recovery for WtE plants with PCC by heat recovery

In the advanced heat integration approach, options for improved thermal integration approaches of post-combustion CO₂ capture technology with WtE-CHP plants are investigated: by identify heat recovery potential from different parts of the PCC plant to either increase power or DH heat supply, thus reducing energy penalty associated with CO₂ capture. In this approach, the same set of design inputs of WtE plant and PCC plant (MSW treatment capacity, steam parameters, capture rates ) and KPIs are used as that of the base case heat integration, in order to compare to effects of heat recovery on thermal/economic/environmental aspects. Detailed information is described in Chapter 7.

- Advanced heat integration for WtE plants with PCC by solvent storage

This approach represents a further investigation of PCC plant integrating into CHP plant that with fluctuating DH demand. It presents a method of seasonal solvent storage for the investigated WtE-CHP plant in steam extraction and condensing configuration, with 99.72% CO₂ capture rate. In particular, it examines the benefit and challenges under this route of solvent storage by analysing the KPIs as identified in Chapter 4. Sensitivity analysis of key design inputs will be included. Detailed information is described in Chapter 8.

5 Cost estimation

For large-scale implementation of CO₂ capture technologies, cost estimates and economic impacts are important since they can be used as signals indicating whether a technology is an economically viable way for decarbonisation. This is especially important for the waste sector, as has been pointed out in chapter 2.3, where financial issues are one of the most challenging factors for the implementation of CO₂ capture. Cost estimation with transparency will help to understand where the cost of each section of the overall process comes from and lay the basis for the estimation of the different economic indicators that are applied afterwards.

In this chapter, the cost estimation methods of WtE plants and CO₂ capture and compression are presented for each integration scenario investigated in this study, also including the additional cost associated with the implementation for the advanced heat integration and solvent storage application. To specify this, the cost elements for the investigated integrations of WtE plant with PCC are outlined in Figure 5-1 and Table 5-1.
Figure 5-1 Structure of cost estimation for the three integration scenarios included in this study, with cost items applied for each application.
Table 5-1 List of cost items as shown in Figure 5-1.

<table>
<thead>
<tr>
<th>Capex cost items</th>
<th>Opex cost items</th>
</tr>
</thead>
<tbody>
<tr>
<td>NO.</td>
<td>NO.</td>
</tr>
<tr>
<td>1.1 WtE Power-only plant</td>
<td>2.1 WtE Power-only plant</td>
</tr>
<tr>
<td>1.2 WtE CHP plant</td>
<td>2.2 WtE CHP plant</td>
</tr>
<tr>
<td>1.3 PCC plant @90% CCR</td>
<td>2.3 PCC plant @90% CCR</td>
</tr>
<tr>
<td>1.4 PCC plant @95% CCR</td>
<td>2.4 PCC plant @95% CCR</td>
</tr>
<tr>
<td>1.5 PCC plant @99.72% CCR</td>
<td>2.5 PCC plant @99.72% CCR</td>
</tr>
<tr>
<td>1.6 Heat exchanger network (heat recovery from PCC plant @90% CCR)</td>
<td>2.6 Heat exchanger network (heat recovery from PCC plant @90% CCR)</td>
</tr>
<tr>
<td>1.7 Heat exchanger network (heat recovery from PCC plant @95% CCR)</td>
<td>2.7 Heat exchanger network (heat recovery from PCC plant @95% CCR)</td>
</tr>
<tr>
<td>1.8 Heat exchanger network (heat recovery from PCC plant @99.72% CCR)</td>
<td>2.8 Heat exchanger network (heat recovery from PCC plant @99.72% CCR)</td>
</tr>
<tr>
<td>1.9 PCC plant @99.72% CCR with storage tanks</td>
<td>2.9 PCC plant @99.72% CCR with storage tanks</td>
</tr>
</tbody>
</table>

It should be noted that for the two WtE CHP plants, steam extraction & condensing configuration, and backpressure configuration, the same set of cost estimation is used. This is because the detailed accessible data for the cost of WtE CHP plant is rare in the literature. Besides, this study lays more emphasis on the economic impacts of adding CO₂ capture to WtE plants, so the impact due to the relatively high-level cost estimation of different WtE CHP plants are limited, as long as a consistent estimation method is used. Detailed cost estimation of WtE CHP plants may be beneficial in the future.
work when to compare the economic impacts of CO₂ capture on different WtE CHP plants, such as different MSW treatment capacities, operation periods, energy prices, etc.

5.1 Capex and Opex of WtE plants (cost items 1.1, 1.2, 2.1, 2.2)

In this study, it is assumed that all the integration scenarios are using the same reference WtE plant, either a power-only, or a CHP type. This work, therefore, focuses on the integration and operation of the capture plant, rather the WtE plant itself. Several specific designs of a WtE plant are possible, so a high-level cost estimation for the WtE plant without PCC is performed within this study, referencing from existing data from the literature to provide a consistent and transparent comparison.

The cost estimation of the reference WtE power-only plant is based on a UK-based WtE plant at similar capacity (Wheeler, 2015). When compared to the power-only plant, there will be additional cost for the WtE-CHP plant for connection to the heat network. However, it is difficult to estimate, with any degree of accuracy, what could be the costs of a CHP system given that so many variables exist. Since the cost of CHP plant will depend upon the specific design of a given CHP scheme. In this study, a ratio of 1.2144 for Capex of WtE CHP to WtE power-only plant costs is applied based on a UK case study presented by ARCADIS (ARCADIS, 2009), which estimated the cost of waste management technologies for a range of European countries by cost modelling approaches. This ratio represents 1) the costs of providing heat from the facility (relative to costs of providing electricity only; 2) the costs of securing a market for the heat.

In the above cost estimation, all the costs have been converted to UK prices at the relevant exchange rate. The Chemical Engineering Plant Cost Index (CEPCI) is used to adjust the cost of the reporting year to the cost of the reference year 2021, as shown in Equation 5.1. Currency conversion is applied under circumstance that different currencies are used in the reporting data. The values applied are summarized in Table 5-2.

\[
\text{COST}_{year-X} = \text{COST}_{year-Y} \times \frac{\text{CEPCI}_{index,X}}{\text{CEPCI}_{index,Y}}
\]

Equation 5.1

Table 5-2 CEPCI numbers and exchange rate applied when calculate the Capex and Opex of WtE plant
Following this approach, the key parameters of Capex of WtE plant (both power-only and CHP plants) without PCC used in this study are summarized in the Table 5-3.

Table 5-3 Key parameters of Capex of WtE plant without PCC (Cost items 1.1 and 1.2)

<table>
<thead>
<tr>
<th>Cases</th>
<th>Cost parameters</th>
<th>Unit</th>
<th>Power only</th>
<th>CHP</th>
<th>Reference year</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reference case</td>
<td>Reference capital cost</td>
<td>£million</td>
<td>139</td>
<td>164</td>
<td>2015</td>
</tr>
<tr>
<td></td>
<td>Reference annual treatment</td>
<td>kt/a</td>
<td>150</td>
<td>150</td>
<td>2015</td>
</tr>
<tr>
<td></td>
<td>Reference specific Capex</td>
<td>£/t</td>
<td>926</td>
<td>1124</td>
<td>2015</td>
</tr>
<tr>
<td></td>
<td>Referenced Specific Capex</td>
<td>£/t</td>
<td>1177</td>
<td>1429</td>
<td>2021</td>
</tr>
<tr>
<td>Generic WtE plant in this study</td>
<td>MSW feed</td>
<td>t/d</td>
<td>500</td>
<td>500</td>
<td>2021</td>
</tr>
<tr>
<td></td>
<td>Annual operational hours</td>
<td>hrs</td>
<td>7650</td>
<td>7650</td>
<td>2021</td>
</tr>
<tr>
<td></td>
<td>Capacity factor</td>
<td></td>
<td>87.33%</td>
<td>87.33%</td>
<td>2021</td>
</tr>
<tr>
<td></td>
<td>Annual treatment of MSW</td>
<td>kt/a</td>
<td>159</td>
<td>159</td>
<td>2021</td>
</tr>
<tr>
<td></td>
<td>Total CAPEX calculated</td>
<td>£million</td>
<td>188</td>
<td>228</td>
<td>2021</td>
</tr>
</tbody>
</table>

Note: The Capex of the reference case Power-only WtE plant is referenced from a UK based WtE plant, with 150kt/year MSW treatment capacity (Wheeler, 2015).

Similarly, the Opex of the WtE-power-only plant references the (Wheeler, 2015) UK-based WtE plant at similar MSW treatment capacity (150kton/year). The same ratio of Opex of WtE-Power-only to WtE-CHP plants is applied as for the Capex ratio between these two plants to account for the additional Opex that may occur for WtE-CHP plants.
The key parameters for the estimation of Opex of a generic WtE plant without PCC are summarized in Table 5-4, as shown below.

Table 5-4 Key parameters of Opex of WtE plant without PCC (Cost items 2.1 and 2.2)

<table>
<thead>
<tr>
<th>Cases</th>
<th>Unit</th>
<th>Power only</th>
<th>CHP</th>
<th>Reference year</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reference case</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Variable Opex</td>
<td>£/t</td>
<td>19.1</td>
<td>19.8</td>
<td>2015</td>
</tr>
<tr>
<td>Fixed Opex</td>
<td>£/t</td>
<td>9.9</td>
<td>10.3</td>
<td>2015</td>
</tr>
<tr>
<td>Variable Opex after inflation</td>
<td>£/t</td>
<td>24.3</td>
<td>25.2</td>
<td>2021</td>
</tr>
<tr>
<td>Fixed Opex after inflation</td>
<td>£/t</td>
<td>12.6</td>
<td>13.1</td>
<td>2021</td>
</tr>
<tr>
<td>Generic WtE plant in this study</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Variable Opex</td>
<td>£million</td>
<td>3.87</td>
<td>4.02</td>
<td>2021</td>
</tr>
<tr>
<td>Fixed Opex</td>
<td>£million</td>
<td>2.01</td>
<td>2.09</td>
<td>2021</td>
</tr>
<tr>
<td>Opex</td>
<td>£million</td>
<td>5.87</td>
<td>6.11</td>
<td>2010</td>
</tr>
</tbody>
</table>

Note: The Opex of the reference case Power-only WtE plant is referenced from a UK based WtE plant, with 150kt/year MSW treatment capacity (Wheeler, 2015);

5.2 Capex and Opex of CO₂ capture and compression plants (cost items 1.3, 1.4, 1.5, 2.3, 2.4, 2.5)

The cost estimation for the PCC plants in this study focuses on costs of a new build WtE plant with PCC. Modifications in the WtE plants, interconnections, and additional utility facilities are not included, considering the large uncertainty of retrofitting related cost. The cost estimation method in this study mainly references from the equipment cost functions developed by the IEAGHG and proposed in the report ‘Evaluating the Costs of Retrofitting CO₂ Captured in an Integrated Oil Refinery: Technical Design Basis and Economic Assumptions, 2017-TR52017’ (IEAGHG, 2017b). Building upon a wide range of CO₂ capture capacity and flue gas CO₂ content, this is a capacity factored estimate methodology that is validated against detailed cost calculation results in the original report and can be used to assess other CO₂ capture cases, which brings confidence in terms of the source of referencing.
5.2.1 Capex estimation of PCC the plant

By the proposed cost estimation function, the cost of the plant under evaluation is derived from the known cost of a similar plant of known capacity (power cost law). Cost and capacity are related by means of a non-linear equation, which can be expressed as:

$$Cost_{actual} = \left(\frac{Capacity_{actual}}{Capacity_{ref}}\right)^{exp} \times Cost_{ref}$$  \hspace{1cm} Equation 5.2

Where:

- $Cost_{actual}$ is the cost of the plant under evaluation;
- $Cost_{ref}$ is the cost of the reference plant;
- $Capacity_{actual}$ and $Capacity_{ref}$ are the respective capacities of the plants;
- $exp$ is the exponent, which typically varies between 0.5 and 0.85, depending on plant type and size.

The exponent is usually lower than 1, when scale economies are observed in scaling up or down the reference cost, while it approaches the value of 1 for modularized systems.

The above-described methodology is used to calculate the equipment cost of the PCC plant, of which three subsections are considered:

a) Absorber section
b) Regeneration section
c) CO$_2$ compression

The most significant capacity parameters of each subsection are applied. Here below describes how this methodology is applied to calculate the equipment cost of each section.

a) Absorption section estimate:

The cost of the absorption section was calculated with a factored estimate using Case 01-03 CRF section in IEAGHG (IEAGH, 2017b) as reference, since it is the case that has the closest CO$_2$ capture capacity and flue gas CO$_2$ content when compared with the WtE flue gas in this study. The estimate considers the cost of the absorber column separately from all the other section items:

$$Cost\ of\ absorption_{new}^{total} = Cost\ of\ absorber_{new} + Cost\ of\ other\ items_{new}$$  \hspace{1cm} Equation 5.3
Cost of absorber_{new}
\[
= \left( \frac{\text{Absorber diameter}_{\text{new}}}{\text{Absorber diameter}_{\text{ref}}} \right)^{1.8} \times \left( \frac{\text{Absorber cost}_{\text{ref}} \times \text{Absorber height}_{\text{new}}}{\text{Absorber height}_{\text{ref}}} \right)
\]

Note:
An exponent of 1.8 for the dependence on the diameter was identified as most suitable, which is consistent with an exponent of 0.9 applied to the cross sectional area, which in turns depends on the flue gas rate (IEAGHG, 2017b).

Cost of other items_{new}
\[
= \left( \frac{\text{Flue gas mass rate}_{\text{new}}}{\text{Flue gas mass rate}_{\text{ref}}} \right)^{1} \times \text{Cost of other items}_{\text{ref}}
\]

b) Regeneration section estimate
The cost of the regeneration section was calculated with a factored estimate using Case 04-04 with an exponent equal to 0.9 from the same IEAGHG report (IEAGHG, 2017b). However, the striper height is also a factor to scale the cost of the stripper. Hence, the cost function was corrected by introducing also a linear dependency on the column height as follows:

Cost of Regeneration_{new}
\[
= \left( \frac{\text{CO}_2 \text{Flowrate to compression}_{\text{new}}}{\text{CO}_2 \text{Flowrate to compression}_{\text{ref}}} \right)^{0.9} \times \left( \frac{\text{Stripper cost}_{\text{ref}} \times \text{Stripper height}_{\text{new}}}{\text{Stripper height}_{\text{ref}}} \right) + \text{Other items cost}_{\text{ref}}
\]

c) Compression section estimate
In IEAGHG (IEAGHG, 2017b), CO_2 compression cost calculations were performed considering that not all the relevant costs depend directly on the amount of CO_2 captured and delivered at plant fence. Instead, the total cost results from the sum of two contributions (one capacity dependent, one capacity independent). The costs that are considered to depend on the amount of CO_2 captured are prorated using exponential cost function, with exponent equal to 0.75. The costs that do not depend on the amount of CO_2 captured are approximately estimated at 600K$ US at 2017 price level.
Where applicable, ‘converting’ costs between the reporting year of the reference data to the year of 2021 (using CEPCI index), and converting currencies from the currency applied in the reference data to Pounds £, the same approach is used as described earlier in the section 5.1.

\[
\text{Cost of compression}^{\text{total}}_{\text{new}} = \text{Cost of compression}^{\text{capacity dependent}}_{\text{new}} + \text{Cost of compression}^{\text{capacity independent}}_{\text{new}} \\
= \left(\frac{\text{CO}_2\text{Flowrate to compression}_{\text{new}}}{\text{CO}_2\text{Flowrate to compression}_{\text{ref}}}\right)^{0.75} \\
\times \text{Cost of compression}^{\text{equipment+piping}}_{\text{ref}} + £581,000
\]

Equation 5.7

With all the equipment costs obtained, a bottom-up approach is applied for calculating the total Capex of the PCC system considering costs for each of the cost sections of the system, plus the costs related to construction, EPC services, contingencies, the owner cost, spare parts, interest during construction, Start-up cost, etc. The CO\(_2\) capture and compression section in the Spreadsheet for cost evaluation of CO\(_2\) capture retrofit (Sintef, 2019b) is used to calculate the total Capex of the PCC plant. Some key input parameters and multipliers are summarized in the Table 5-5 below:

Table 5-5 Key input parameters and multipliers used to calculate the total Capex of the PCC plant (IEAGHG, 2017b)

<table>
<thead>
<tr>
<th></th>
<th>Absorber</th>
<th>Regenerator</th>
<th>Compression</th>
<th>Notes</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Direct material</td>
<td>Based on Cost functions from IEAGHG (IEAGHG, 2017b)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2</td>
<td>Construction</td>
<td>58.7%</td>
<td>58.8%</td>
<td>73.3%</td>
</tr>
<tr>
<td>3</td>
<td>Direct Field cost (DFC)</td>
<td>Sum of Direct material and Construction</td>
<td></td>
<td></td>
</tr>
<tr>
<td>4</td>
<td>Other cost</td>
<td>8.9%</td>
<td>8.9%</td>
<td>11%</td>
</tr>
<tr>
<td>5</td>
<td>EPC services</td>
<td>31.8%</td>
<td>31.9%</td>
<td>34.6%</td>
</tr>
<tr>
<td>6</td>
<td>Total installed cost (TIC)</td>
<td>Sum of DFC, Other cost, EPC service</td>
<td></td>
<td></td>
</tr>
<tr>
<td>7</td>
<td>Project contingencies</td>
<td>Percentage of Total Installed cost, 30%</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td><strong>Total plant cost (TPC)</strong></td>
<td>Sum TIC and project contingencies</td>
<td></td>
<td></td>
</tr>
<tr>
<td>---</td>
<td>---------------------------</td>
<td>----------------------------------</td>
<td></td>
<td></td>
</tr>
<tr>
<td>9</td>
<td>The owner cost, spare parts, modifications</td>
<td>Percentage of TPC, 9.5% [2]</td>
<td></td>
<td></td>
</tr>
<tr>
<td>10</td>
<td>Interest during construction</td>
<td>Percentage of TPC, in %</td>
<td></td>
<td></td>
</tr>
<tr>
<td>11</td>
<td>Start-up cost</td>
<td>3 months labour cost and 2% TPC [2]</td>
<td></td>
<td></td>
</tr>
<tr>
<td>12</td>
<td><strong>Total Capex</strong></td>
<td>The sum of TPC and the above cost items 9, 10 and 11</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Notes:

[1] Percentage of Direct material cost, values obtained by calculating the average percentage values of each section to the direct material cost, for all the case studies Recap CO₂ capture retrofit cost evaluation cases (SINTEF, 2019a). This is reasonably robust since the percentages vary by different equipment, but the percentage of each cost item to the direct material is found to be a minor variation;

[2] Percentage values referencing from the Spreadsheet for cost evaluation of CO₂ capture retrofit (Sintef, 2019b);

5.2.2 Opex estimation of PCC the plant

The Opex of the PCC plant is the sum of annual fixed Opex and annual variable Opex. The annual fixed Opex is determined by the number of employees, average fully burdened salary, annual material maintenance percentages, overall maintenance cost percentage, other cost percentages. The annual variable Opex includes sludge disposal quantities and cost, material replacement and cost, share of natural gas consumption linked to steam production for CO₂ stripping.

Where applicable, ‘converting’ costs between the reporting year of the reference data to the year of 2021 (using CEPCI index), and converting currencies from the currency applied in the reference data to Pounds £, the same approach is used as described earlier in the section 5.1.

The key input data for calculation the annual fixed Opex and annual variable Opex is summarized in the Table 5-6.
Table 5-6 Key input data for estimation of the annual fixed Opex and annual variable Opex of the PCC plant

<table>
<thead>
<tr>
<th>Item</th>
<th>Value</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Annual fixed operating cost</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Item</td>
<td>Value</td>
<td>Unit</td>
</tr>
<tr>
<td><strong>Annual variable operating cost</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1 Total number of employees</td>
<td>3</td>
<td></td>
</tr>
<tr>
<td>2 Average fully burdened salary</td>
<td>80000</td>
<td>£/y</td>
</tr>
<tr>
<td>3 Labour cost</td>
<td>Total number of employees * Average fully burdened salary</td>
<td></td>
</tr>
<tr>
<td>4 CO₂ capture and conditioning</td>
<td>2</td>
<td>%TPC</td>
</tr>
<tr>
<td>5 Share annual material maintenance cost in the overall annual maintenance cost</td>
<td>60</td>
<td>%</td>
</tr>
<tr>
<td>6 Annual material maintenance cost</td>
<td>TPC * 2% / 60%</td>
<td></td>
</tr>
<tr>
<td>7 Other fixed cost</td>
<td>0.5</td>
<td>%TPC</td>
</tr>
<tr>
<td>8 Other fixed cost</td>
<td>0.5% * TPC</td>
<td></td>
</tr>
<tr>
<td><strong>Annual fix operating cost</strong></td>
<td>Sum of item 3 + 6 + 7</td>
<td>million £/year</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Item</th>
<th>Value</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>MEA solvent cost</td>
<td>1556</td>
<td>£/tone</td>
</tr>
<tr>
<td>Total CO₂ captured</td>
<td></td>
<td>kg/hr</td>
</tr>
<tr>
<td>Annual operational hours</td>
<td>7650</td>
<td>hr/year</td>
</tr>
<tr>
<td></td>
<td>Description</td>
<td>Value 1</td>
</tr>
<tr>
<td>---</td>
<td>--------------------------------------------------</td>
<td>---------</td>
</tr>
<tr>
<td>4</td>
<td>MEA solvent lost rate [1]</td>
<td>0.11</td>
</tr>
<tr>
<td>5</td>
<td><strong>MEA solvent cost</strong></td>
<td></td>
</tr>
<tr>
<td></td>
<td>MEA solvent cost<em>Total CO₂ captured</em>Annual operational hours*MEA solvent lost rate</td>
<td></td>
</tr>
<tr>
<td>6</td>
<td>MEA sludge generation rate [2]</td>
<td>3.3</td>
</tr>
<tr>
<td>7</td>
<td>Fuel thermal input</td>
<td>55</td>
</tr>
<tr>
<td>8</td>
<td><strong>Annual MEA sludge</strong></td>
<td></td>
</tr>
<tr>
<td></td>
<td>MEA sludge generation rate<em>Fuel thermal input</em>Annual operational hours</td>
<td></td>
</tr>
<tr>
<td>9</td>
<td>MEA sludge disposal cost</td>
<td>175</td>
</tr>
<tr>
<td>10</td>
<td><strong>MEA sludge disposal cost</strong></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Annual MEA sludge*MEA sludge disposal cost</td>
<td></td>
</tr>
</tbody>
</table>

**Annual variable operating cost**  | **Sum of item 5+10** | **million£/year** |

Notes:

[1] The MEA solvent lost rate is referencing the operational CO₂ capture facility at the Technology Centre Mongstad (Flø et al., 2017). The MEA sludge generation rate is referencing from the generated reclaim sludge using monoethanolamine (MEA) pulverized coal (PC) CO₂ capture cases (Sexton et al., 2014).

[2] Unless specified, the factor applied for each cost item referencing from (IEAGHG, 2017b).

It should be noted that at the time of writing, capture plant cost data using flue gas from WtE plants are still lacking in literature. So, the main data used for the Capex and Opex of the CO₂ capture plant isn’t specific for WtE fuel and the corresponding impact is neglected in this study. For future work, it may be helpful to consider the impact of the flue gas condition from the WtE plant on the cost estimation, for instance, the rates of degradation and accumulation of impurities from the flue gas, which will affect the flue gas pre-treatment and operating conditions of the capture plant, thus affect the cost estimation.
5.3 Capex and Opex of heat exchanger networks (cost item 1.6, 1.7, 1.8, 2.6, 2.7, 2.8)

The Capex and Opex of heat exchanger networks are used for the advanced heat integration scenarios investigated in this study. In these scenarios, the Capex and Opex required to perform additional heat recovery are added to the total Capex and Opex under consideration. This includes the cost of heat exchangers, connection costs and heat pumps, etc. In terms of the estimation of heat recovery cost, it is assumed that new heat exchangers will be installed to deliver heat required for the advanced integration. In certain cases, it might be possible to modify existing heat exchangers. However, here in this thesis, it is not considered and the additional cost estimated for the implementation of a new heat recovery system represents an upper bound for the costs.

The type of heat exchanger for the heat exchanger network is assumed to be shell and tube type with perfect counter-current flow. The shell and tube type heat exchangers are the most frequent type of heat exchangers that can be used under a wide range of operating pressures and temperatures different operation conditions. Their relatively simple manufacturing robustness and reliability make this type of heat exchanger suitable for being used in several applications such as power generation, heating and air conditioning; they are also widely used in the chemical process industries (Caballero, Ravagnani, Pavao, Costa, & Javaloyes-Anton, 2022; Chang, Liao, Costa, & Bagajewicz, 2022). They are expected to maintain their market position in the years to come (Cartelle Barros et al. 2018). They consist of a cylindrical shell with a bundle of tubes inside. One fluid flow through the tubes, while the other flows around them in the shell. The two fluids enter at opposite ends of the heat exchanger and flow counter to one another, thus gives better heat transfer and higher temperature differential over the length of the exchanger. The material of the heat exchanger is chosen to be carbon steel (CS), which is one of the most common materials for the construction of heat exchangers.

The cost of the heat recovery system is divided into subsections: 1) heat exchanger investments costs and 2) hot water piping cost for transferring heat between different heat sources and sinks throughout the system, a final 10% of Capex is added to compensate the costs related to heat losses.

5.3.1 Capex of heat exchanger

The Capex estimation of the heat exchanger ($C_{HX}$) follows the approach as described in (Sinnott & Towler, 2020; Smith, 2005). It is based on an estimate of the purchase cost of the equipment ($C_E$ of heat exchanger) required for each heat recovery process, the other costs being estimated as factors of the equipment cost. In order to calculate the $C_E$, the estimated heat exchanger area $A$ for each heat recovery unit in the advanced heat recovery scenario is required. And then using a referenced correlation to estimate the purchased equipment cost for each heat exchanger.
The heat exchanger area \( A \) is estimated according to the general equation (Sinnott & Towler, 2020):

\[
Q = U \times A \times \Delta T_m
\]

Equation 5.8

With

\[
\Delta T_m = F_t \times \Delta T_{lm}
\]

Equation 5.9

Where

\( Q \) = Required heat recovery capacity for the heat exchanger

\( A \) = Heat exchanger area estimated; m\(^2\)

\( F_t \) = Temperature correction factor taking into account deviations from counter-current flow in shell-and-tube heat exchangers;

\( \Delta T_{lm} \) = Logarithmic mean temperature difference:

\[
\Delta T_{lm} = \frac{(T_{hot,in} - T_{cold,out}) - (T_{hot,out} - T_{cold,in})}{\ln \frac{T_{hot,in} - T_{cold,out}}{T_{hot,out} - T_{cold,in}}}
\]

Equation 5.10

\[
F_t = \frac{\sqrt{R^2 + 1} \ln \left( \frac{1 - S}{1 - RS} \right)}{(R - 1) \ln \left[ 2 - S \left( R + 1 - \sqrt{R^2 + 1} \right) \right]}
\]

Equation 5.11

\[
S = \frac{T_{cold,out} - T_{cold,in}}{T_{hot,in} - T_{cold,in}}
\]

Equation 5.12

\[
R = \frac{T_{hot,in} - T_{hot,out}}{T_{cold,out} - T_{cold,in}}
\]

Equation 5.13

The U-value (overall heat transfer coefficient) is determined using information about the flow characteristics of the service (utility) and process fluids circulating in the heat exchangers. Figure 5-2 provides typical heat transfer coefficients for different utilities and process fluids and can be used to
estimate the overall heat transfer coefficient for Shell and Tube heat exchangers (Sinnott Ray & Gavin, 2009).

Figure 5-2 Overall U-values for different process/service fluid combinations in shell and tube heat exchangers (Sinnott & Towler, 2020).

In real operation, the working condition of heat exchangers may change thus affect the heat exchange efficiency, for instance, the expected fouling layers, or unexpected increase of water flow rate. The calculated area is then increased by a factor of 1.25 to account for increased area demand under certain operating conditions (Saari 2010).

Referencing from (Smith, 2005), with the known required heat transfer area $A$, the equipment cost of the specific heat exchanger ($C_E$) can be expressed as a power law of capacity

$$C_E = C_{ref} \times \left( \frac{A}{A_{ref}} \right)^m \times f_M \times f_P \times f_T$$  \hspace{1cm} \text{Equation 5.14}

Where

$C_E =$ Cost for new CS heat exchanger with area [m$^2$] operating at pressure p and temperature T [£, 2021 value];
$C_{\text{ref}}=$ Known base cost [£], for shell and tube heat exchanger using CS, $C_{\text{ref}}=$ $32,800 on 2020 basis;

$A_{\text{ref}}=$ reference case area $[m^2]$ corresponding to $C_{\text{ref}}, A_{\text{ref}} = 80 \, m^2$;

$m =$ constant factor depending on equipment type; $m=0.68$ for shell-and-tube heat exchangers (Smith, 2005);

$f_M, f_P, f_T$ are correction factors for materials of construction, design pressure, design temperature, which can be valued based on Table 5-7 below:

Table 5-7 Material, Pressure and Temperature factors for heat exchangers according to (Smith, 2005).

<table>
<thead>
<tr>
<th>Material</th>
<th>Factor $f_M$</th>
<th>Pressure (bara)</th>
<th>Factor $f_P$</th>
<th>Temperature</th>
<th>$f_T$</th>
</tr>
</thead>
<tbody>
<tr>
<td>CS</td>
<td>1</td>
<td>0.01-0.1</td>
<td>2</td>
<td>0-100</td>
<td>1</td>
</tr>
<tr>
<td>SS low grade</td>
<td>2.1</td>
<td>0.1-0.5</td>
<td>1.3</td>
<td>100-300</td>
<td>1.6</td>
</tr>
<tr>
<td>SS high grade</td>
<td>3.2</td>
<td>0.5-7</td>
<td>1</td>
<td>300-500</td>
<td>2.1</td>
</tr>
<tr>
<td>Monel</td>
<td>3.6</td>
<td>7-50</td>
<td>1.5</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Inconel</td>
<td>3.9</td>
<td>50-100</td>
<td>1.9</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Nickel</td>
<td>5.4</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Titanium</td>
<td>7.7</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LTCS</td>
<td>1.5</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Where applicable, ‘converting’ costs between the reporting year of the reference data to the year of 2021 (using CEPCI index), and converting currencies from the currency applied in the reference data to Pounds £, the same approach is used as described earlier in the section 5.1.

In addition to the purchased cost of the equipment $C_E$, investment is required to install the equipment. Installation costs include (Smith, 2005):

- cost of installation
- piping and valves
- control systems
- foundations
Thus, the total capital cost of the heat exchanger \( C_{HX} \) can be obtained by applying multiplying factors or installation factors to the purchase cost of individual items of equipment \( C_E \), as shown in Equation 5.15.

\[
C_{HX} = [f_M \times f_P \times f_T \times (1 + f_{PIP})] \times C_E \\
+ (f_{ER} + f_{INST} + f_{ELEC} + f_{UTIL} + f_{OS} + f_{BUILD} + f_{SP} + f_{DEC}) \times C_E \\
+ (f_{CONT} + f_{WC}) \times C_E
\]

Equation 5.15

Cost correction factors \( f_{PIP}, f_{ER}, f_{INST}, f_{ELEC}, f_{UTIL}, f_{OS}, f_{BUILD}, f_{SP}, f_{DEC}, f_{CONT} \) and \( f_{WC} \) used in Equation 5.15 are listed in Table 5-8.

### Table 5-8 Typical correction factors for capital cost based on delivered equipment costs (Smith, 2005).

<table>
<thead>
<tr>
<th>Factors from Smith (2005)</th>
<th>Factor</th>
</tr>
</thead>
<tbody>
<tr>
<td>Equipment delivered cost</td>
<td>1</td>
</tr>
<tr>
<td>Piping ( f_{PIP} )</td>
<td>0.7</td>
</tr>
<tr>
<td>Equipment erection ( f_{ER} )</td>
<td>0.4</td>
</tr>
<tr>
<td>Instrumentation ( f_{INST} )</td>
<td>0.2</td>
</tr>
<tr>
<td>Electrical ( f_{ELEC} )</td>
<td>0.1</td>
</tr>
<tr>
<td>Utilities ( f_{UTIL} )</td>
<td>0.5</td>
</tr>
<tr>
<td>Off-sites ( f_{OS} )</td>
<td>0.2</td>
</tr>
<tr>
<td>Buildings ( f_{BUILD} )</td>
<td>0.2</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
<tr>
<td>--------------------------------------</td>
<td>--------</td>
</tr>
<tr>
<td><strong>Site preparation</strong> $f_{SP}$</td>
<td>0.1</td>
</tr>
<tr>
<td><strong>Total capital cost of installed equipment</strong></td>
<td>3.2</td>
</tr>
<tr>
<td><strong>Design and Engineering</strong> $f_{DEC}$</td>
<td>1</td>
</tr>
<tr>
<td><strong>Contingency</strong> $f_{CONT}$</td>
<td>0.4</td>
</tr>
<tr>
<td><strong>Total fixed capital</strong></td>
<td>4.6</td>
</tr>
<tr>
<td><strong>Working capital</strong> $f_{WC}$</td>
<td>0.7</td>
</tr>
<tr>
<td><strong>Total capital cost</strong></td>
<td>4.6</td>
</tr>
</tbody>
</table>

Note:
The above values are:

- based on carbon steel, moderate operating pressure and temperature
- average values for all types of equipment, whereas in practice the values will vary according to the type of equipment
- only guidelines and the individual components will vary from project to project
- applicable to new design only.

### 5.3.2 Hot water piping cost

The Capex of the hot water (HW) piping $C_{pipe}$ requires the estimation of the diameter $D$ of the specific pipe. To calculate the diameter of the HW pipes was used to determine the cost of hot water piping. The estimation of the diameter is performed using the Hazen Williams equation (TL, 2021):

$$ Q = (3.763 \times 10^{-6}) \times C \times D^{2.63} \times (\Delta P/L)^{0.54} $$

Where:

- $Q$ is the water flow rate in the pipe, m$^3$/hr,
- $D$ is the pipe diameter, mm,
- $L$ is the pipe length, m,
ΔP is the pressure difference across pipe length L, kN/m²,

C is the Hazen Williams coefficient, dimensionless (depends on pipe material and age), value of C is available using table below from (TORO, n.d.):

Table 5-9 Roughness coefficient c values for Hazen-Williams equation (TORO, n.d.)

<table>
<thead>
<tr>
<th>TYPE OF PIPE</th>
<th>RANGE</th>
<th>NEW PIPE</th>
<th>DESIGN C</th>
</tr>
</thead>
<tbody>
<tr>
<td>PVC</td>
<td>160 - 145</td>
<td>150</td>
<td>150</td>
</tr>
<tr>
<td>Polyethylene</td>
<td>150 - 130</td>
<td>140</td>
<td>140</td>
</tr>
<tr>
<td>Asbestos-Cement</td>
<td>160 - 140</td>
<td>150</td>
<td>140</td>
</tr>
<tr>
<td>Cement-Lined Steel</td>
<td>160 - 140</td>
<td>150</td>
<td>140</td>
</tr>
<tr>
<td>Welded Steel</td>
<td>150 - 80</td>
<td>140</td>
<td>100</td>
</tr>
<tr>
<td>Riveted Steel</td>
<td>140 - 90</td>
<td>110</td>
<td>100</td>
</tr>
<tr>
<td>Concrete</td>
<td>150 - 85</td>
<td>120</td>
<td>100</td>
</tr>
<tr>
<td>Cast Iron</td>
<td>150 - 80</td>
<td>130</td>
<td>100</td>
</tr>
<tr>
<td>Copper, Brass</td>
<td>150 - 120</td>
<td>140</td>
<td>130</td>
</tr>
<tr>
<td>Wood Stave</td>
<td>145 - 110</td>
<td>120</td>
<td>110</td>
</tr>
<tr>
<td>Vitrified Clay</td>
<td></td>
<td>110</td>
<td>100</td>
</tr>
<tr>
<td>Corrugated Steel</td>
<td></td>
<td>60</td>
<td>60</td>
</tr>
</tbody>
</table>

The Hazen-Williams Equation used in this study is an empirical relationship which relates the flow of water in a pipe with the physical properties of the pipe and the pressure drop caused by friction. It does not consider other sources of head loss, such as elevation change, direction change, or pipe restrictions.

With the diameter calculated, the cost for the hot water pipes are referencing from (Roman Hackl & Harvey, 2013) who using a cost function reported by the Swedish District Heating Association. In that report, different cost functions are available depending on the conditions where the pipes are to be installed (urban, suburban or urban park environment). In this study, the highest cost (urban environment) is assumed:

\[
C_{\text{pipe}} = (-0.0112 \times D^2 + 28.22 \times D + 2707.4) \times L \quad \text{Equation 5.17}
\]

Where

\(C_{\text{pipe}}\) Cost of the hot water pipes

\(D\) Calculated diameter of the hot water pipes, using Equation 5.16;

\(L\) Length of the hot water pipes, assumed to be 100 meter in this study;
Where applicable, ‘converting’ costs between the reporting year of the reference data to the year of 2021 (using CEPCI index), and converting currencies from the currency applied in the reference data to Pounds £, the same approach is used as described earlier in the section 5.1.

It should be noted that the cost of HW pipes remains large uncertainty arising from site specific parameters, such the locations of the heat exchangers, the length of HW pipes. Detailed cost estimation for the hot water pipes will be valuable for future work.

5.4 Additional cost for solvent storage scenarios (cost item 1.9, 2.9)

The Capex and Opex of PCC system with additional storage tanks are used under the Interim solvent storage case in this study. Under this scenario, the total Capex and Opex includes the cost as mentioned in the base case, and the additional equipment cost required for performing solvent storage such as rich/lean solvent tanks, oversizing of the stripper, compressor. The cost of the solvent tanks and the oversized equipment follows the same methodology as used for the base case cost calculation, as described in section 5.2.

Once the Capex and Opex for each cost item number are obtained, the net present value (NPV) is calculated. This represents the current value of all cash flows for each integration scenario. The calculation of NPV is introduced in chapter 3.2.1.

5.5 Summary of cost estimation

This chapter systemically describes the cost estimation approach applied for this study, to estimate the Capex and Opex of the investigated WtE with PCC plant under each integration scenario. It covers the Capex and Opex of WtE plant (power-only, CHP), Capex and Opex of PCC plant, and also the additional Capex and Opex estimation of the heat exchanger networks and additional cost of solvent storage that applied under advanced heat integration and solvent storage application, respectively.

The choice of cost estimation methods tries to apply the most relevant approach that available in the literature in each aspect (WtE plant, PCC plant, HX network, solvent storage), and also consider the purpose of these cost data for the KPI calculation. For instance, relatively high level of cost estimation of WtE plants is applied, since the economic KPIs focuses on the impact of adding CO₂ capture (different capture rate, different integration scenario) on the WtE plant, not the variation of the WtE plant itself. For the same reason, a relatively detailed cost estimation of the PCC plant is performed. The cost estimation of the heat exchanger network applies the common method that available from the literature, but there is also quite a level of uncertainty which arises from the defined values, for instance, the choice of material and type of heat exchanger, the length of hot water pipes. Detailed cost estimation and relevant sensitivity analysis from the WtE plant and heat exchanger network
design will be useful to fully understand the impact of different heat integration scenarios on the overall economic performance of the abated WtE plant integrating with PCC.
6 Base case integration of WtE plants with MEA based post-combustion CO₂ capture

This chapter describes the base-case heat integration of the investigated WtE plants (power-only and CHP plants) with PCC, at different CO₂ capture rates. It represents a scenario under which the minimum modification of the WtE plant is performed to integrate CO₂ capture and compression. Due to the essential steam extraction from the power cycle, the integration of CO₂ capture under this scenario will unavoidably affect the power production and DH supply, and thermal efficiencies of the integrated plants. This integration on the other hand results into the least investment cost and operation with minimum complexity. This will provide an insight for the potential plant owners and designers with a view of what will happen including but not limited to, the energy output penalty, efficiency penalty and CO₂ emission reduction under the base case integration.

This chapter will first review the integration methods for the WtE plants (power-only, CHP) with PCC. The key results from the WtE plant modelling in gProcess and CO₂ capture modelling in ASPEN will be presented. Following this, the KPI from thermodynamic, economic and environmental aspects will be presented and discussed.

6.1 Base case integration of WtE plants with PCC

The base case integration of WtE plant with PCC applies the most common approach applied for CO₂ capture integrates with power plant. The heat required for solvent regeneration in the CO₂ capture plant is achieved by steam extraction from an intermediate point in the expansion stage of the power cycle. The extracted steam in superheat condition is firstly de-superheated using a spray with condensate from the solvent reboiler, turns into saturated steam at the required pressure for solvent regeneration in the reboiler. The saturated steam releases its latent heat in the reboiler, and the condensed steam returned to the steam cycle feed-water-heating train to fully recover the useful heat.

The overall process flow and the connection interfaces between the gProcess model and the ASPEN model is illustrated in Figure 6-1.
6.2 Key results from the WtE plant modelling in gProcess

Following the modelling approach of WtE plant described in Chapter 4.1, three representative WtE plants (power-only, CHP-Ex&C, CHP-BP) are modelled in gProcess. These WtE plants are using the same fuel throughput and composition taken from the RERC data provided by FCC Environment (through personal communication), the same boiler incineration and heat recovery process. Considering the priority of the WtE plant is MSW treatment, and MSW throughput does not show much variation in real operational plant, it is assumed that the WtE plants operate at full load. This means that the flue gas mass flow and composition maintain are the same under different steam cycle configurations. The steady state and full load modelling of WtE plants, avoids the complexity of part load and dynamic running of a WtE plant and can be used for the base case WtE plants modelling that is presented in this thesis.

The key performance data of the WtE modelling, including the grate boiler incineration, flue gas temperature, power, and DH heat output, etc. are presented in Table 5-1.
Table 6-1 Key performance data of WtE modelling

<table>
<thead>
<tr>
<th>Description</th>
<th>Unit</th>
<th>Process modelling</th>
</tr>
</thead>
<tbody>
<tr>
<td>LHV</td>
<td>kJ/kg</td>
<td>9300</td>
</tr>
<tr>
<td>Waste throughput</td>
<td>t/h</td>
<td>19.4</td>
</tr>
<tr>
<td>Primary air into boiler</td>
<td>°C</td>
<td>135</td>
</tr>
<tr>
<td>Primary air mass-flow</td>
<td>kg/s</td>
<td>19.7</td>
</tr>
<tr>
<td>Primary air preheating (external)</td>
<td>MW</td>
<td>1.7</td>
</tr>
<tr>
<td>Secondary air into boiler</td>
<td>°C</td>
<td>50</td>
</tr>
<tr>
<td>Secondary air mass-flow</td>
<td>kg/s</td>
<td>8.5</td>
</tr>
<tr>
<td>Secondary air preheating (external)</td>
<td>MW</td>
<td>0</td>
</tr>
<tr>
<td>Total air preheating from steam cycle</td>
<td>MW</td>
<td>1.7</td>
</tr>
<tr>
<td>Boiler Live steam Temperature</td>
<td>°C</td>
<td>400</td>
</tr>
<tr>
<td>Boiler Live steam pressure</td>
<td>bar</td>
<td>40</td>
</tr>
<tr>
<td>Boiler Live steam mass flowrate</td>
<td>kg/s</td>
<td>18.2</td>
</tr>
<tr>
<td>FW return Temperature</td>
<td>°C</td>
<td>137</td>
</tr>
<tr>
<td>FW return Pressure</td>
<td>bar</td>
<td>80</td>
</tr>
<tr>
<td>FW return mass flowrate</td>
<td>kg/s</td>
<td>18.2</td>
</tr>
<tr>
<td>Flue gas Temperature after cleaning</td>
<td>°C</td>
<td>136</td>
</tr>
<tr>
<td>Flue gas pressure</td>
<td>bar</td>
<td>1</td>
</tr>
<tr>
<td>O₂ concentration in the flue gas (vol% wet)</td>
<td></td>
<td>5.9%</td>
</tr>
<tr>
<td>Boiler efficiency</td>
<td></td>
<td>80%</td>
</tr>
</tbody>
</table>

95
In the following section, key results based on the boiler heat balance will be reported. It is based on the boiler heat balance approach that described in Chapter 4.1 to identify the sources of heat loss of the modelled WtE plant, and look at the key parameter (flue gas temperature) that affect the boiler efficiency.

1) Boiler heat losses from the modelled WtE plant

For a given fuel composition reference from the operational WtE plant in Edinburgh, given the modelled grate boiler in gProcess, with the assumption of excess air 50% input to achieve full oxidization condition in the flue gas. Based on heat balance approach as introduced in Chapter 4, there are in total six heat outputs by burning the fuel:

\[ Q_1 = \text{Heat absorbed by steam (utilized by the boiler)} \]
\[ Q_2 = \text{Heat loss through stack gas} \]
\[ Q_3 = \text{Heat loss by incomplete combustion of gaseous components} \]
\[ Q_4 = \text{Heat loss owing to unburned carbon} \]
\[ Q_5 = \text{Heat loss owing to convection and radiation from the furnace exterior} \]
\[ Q_6 = \text{Heat loss through the sensible heat of ash and slag} \]

The results of the heat output are shown in Figure 6-2 below. It can be seen that the percentage of the heat losses vary. Among them, the heat loss from flue gas \((Q_2)\) takes the largest share of the heat losses from the boiler, \(\sim 10\%\) of the total fuel input. This is a reasonable modelling result. Experience from coal based boiler finds that the heat loss due to stack gas takes the highest share of the heat losses, which could take 12\% to 20\% of the fuel energy input (Industrial Boiler Design Calculation Standard Method Editorial Board, 2003). At the time of writing, the detailed accessible data of heat losses from WtE plant is lacking in the literature. The high percentage share of the heat loss from stack gas from the modelled generic WtE plants (comparable to the corresponding percentage of coal-based grate boilers) implies that approaches to reduce this specific heat loss could be considered based on the operation of other types of fuel incineration plant, such as coal incineration.
Effect of flue gas temperature and boiler efficiency

Flue gas conditions are a set of the key input parameters for the capture plant, and also the main target of process modelling of the WtE plant. Through process modelling of the boiler thermal balance, as shown above, it is found that heat loss through the stack gas takes the highest proportion of all the heat losses.

The higher the flue gas temperature, the higher the stack gas heat losses, thus lower boiler efficiency. As can be seen from Figure 6-3, for every 1% increase of boiler efficiency, the exhaust flue gas temperature decreases by 14°C. This is comparable to stack heat loss from coal incineration boilers, where the heat loss increases by 1% when the exit flue gas temperature increases by 10°C (Prabir Basu et al., 2000). Theoretically, lower flue gas temperature may lead to higher boiler efficiency. The three-way trade off for the flue gas temperature with the material cost, thermal efficiency and operational safety is discussed in Chapter 4.1. When integration with CO₂ capture, the lower flue gas temperature may also benefit the following CO₂ capture process since lower temperature of the incoming flue gas typically leads to reduced power consumption at the DCC. In this study, with a boiler efficiency of 82%, the flue gas temperature is modelled to be 226°C, within the common range of WtE plant (Branchini, 2012). This flue gas will be treated in the flue gas cleaning facility where the temperature will further
reduce to 136 °C, the temperature referenced from the operating condition of the WtE plant in Edinburgh.

Flue gas conditions

In this study, the combustion process is defined as a complete oxidation of the MSW, with an excess air ratio of 50%. The complete oxidation generate gaseous products – flue gas, which is a mixture of combustion products such as CO₂, H₂O, NO₂, and SO₂. In an operational WtE plant, the flue gas will be treated in an air pollution control (APC) equipment to remove hazardous residues. The residues must be safely disposed of to a licensed and specialist landfill under very strict regulatory conditions. The treated flue gas goes into a gas-gas rotary heat exchanger, releasing heat to increase the temperature of the CO₂-depleted gas from the absorber, ensuring adequate gas buoyancy and dispersion in the atmosphere. After this, the flue gas passes through a direct contact cooler (DCC) to reduce to temperatures where the chemistry of absorption is most efficient. An overview of key features of the flue gas condition downstream of the DCC that are used as the incoming flue gas in the Aspen model of the capture plant is shown in Table 6-2 below.

Table 6-2 Flue gas inlet conditions of the PCC system

![Figure 6-3 Change of boiler efficiency with exhaust flue gas temperature, results based on the process modelling of a generic WtE plant](image_url)

- Flue gas conditions

In this study, the combustion process is defined as a complete oxidation of the MSW, with an excess air ratio of 50%. The complete oxidation generate gaseous products – flue gas, which is a mixture of combustion products such as CO₂, H₂O, NO₂, and SO₂. In an operational WtE plant, the flue gas will be treated in an air pollution control (APC) equipment to remove hazardous residues. The residues must be safely disposed of to a licensed and specialist landfill under very strict regulatory conditions. The treated flue gas goes into a gas-gas rotary heat exchanger, releasing heat to increase the temperature of the CO₂-depleted gas from the absorber, ensuring adequate gas buoyancy and dispersion in the atmosphere. After this, the flue gas passes through a direct contact cooler (DCC) to reduce to temperatures where the chemistry of absorption is most efficient. An overview of key features of the flue gas condition downstream of the DCC that are used as the incoming flue gas in the Aspen model of the capture plant is shown in Table 6-2 below.


<table>
<thead>
<tr>
<th>Flue gas at the inlet of the absorber</th>
<th>Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure</td>
<td>bar 1.063</td>
</tr>
<tr>
<td>Temperature</td>
<td>C 40</td>
</tr>
<tr>
<td>Mass flow rate</td>
<td>kg/s 30.71</td>
</tr>
<tr>
<td>Molar flow rate</td>
<td>mol/s 1046</td>
</tr>
</tbody>
</table>

**Composition**

<table>
<thead>
<tr>
<th></th>
<th>%vol</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO₂</td>
<td>11.11</td>
</tr>
<tr>
<td>H₂O</td>
<td>6.95</td>
</tr>
<tr>
<td>N₂</td>
<td>75.23</td>
</tr>
<tr>
<td>O₂</td>
<td>6.72</td>
</tr>
</tbody>
</table>

6.3 Key results from the CO₂ capture plant modelling in ASPEN

Based on the PCC plant modelling method as introduced in Chapter 4.2, this chapter reports the key findings through the process of the PCC plant modelling work conducted in ASPEN. Although there is variation in terms of the configurations of the WtE plant (power only, CHP), it is assumed that all the modelled WtE plants are incinerating constant MSW input (in terms of both fuel composition and throughput), so that the modelling results of the PCC plant can be applied for each integration scenario. The only variation comes from the targeted CO₂ capture rates, which will affect the design choice in terms of column packing and specific heat required for solvent regeneration, thus the amount of steam from the steam cycle. The determined absorber packing for each CO₂ capture rate will be presented in this chapter, along with sensitivity analysis for a range of design parameter, such as the lean solvent loadings, absorber intercooling, stripper pressure, on the performance of the capture plant.

6.3.1 Absorber packing design and optimisation

The CO₂ capture plant is designed and optimised to process the total amount of flue gas exiting the WtE plants. Absorber design in PCC systems involves a trade-off between capital cost and operating cost. In general, a higher absorber packing bed or a higher SRD (specific reboiler duty) is required to
achieve a higher CO₂ capture rate. Figure 6-4 illustrates the effect of increasing the absorber packing height on SRD and on rich solvent loadings, for a range of CO₂ capture rates.

Figure 6-4  Sensitivity of the absorber packing height on the rich solvent CO₂ loading (continuous line) and the specific reboiler duty (dashed lines) for a 35%wt MEA capture system at a range of CO₂ capture rates: 90%, 95%, 99% and 99.72% CO₂ capture rates.

For a given CO₂ capture rate, a higher packing height results in a larger contact surface area and a longer residence time, which enhances the CO₂ absorption rate and thus the CO₂ loading of the solvent at the bottom of the absorber (rich solvent) increases. This also results in an increased solvent capacity for a given lean solvent CO₂ loading, and thus a reduced amount of solvent is required to achieve a given CO₂ absorption efficiency, which reduces the sensible heat required to heat the solvent to the
reboiler temperature (once the solvent has been transferred to the stripper). This is reflected in a reduced SRD.

In this work, the absorber packing height is optimised for each one of the considered configurations separately to enable a fair comparison of technical performance. For each case, the absorber packing height is increased to a value at which a further increase in packing height results in a similar marginal (and small) gain in the rich solvent CO$_2$ loading and also a similar marginal (and small) reduction of the SRD. However, practical constrains need to be considered as well. Personal communications from SULZER suggested an upper limit of 8 m to 10 m for the height of each structured packing section (packing bed) in the absorber column to ensure adequate liquid distribution and structural integrity.

Using this approach, an absorber packing height of 24 m (3 packing beds of 8 m each) is selected for the net-zero direct emission case (i.e. 99.72% capture rate); a further increase would lead to a marginal decrease in the SRD of 1.3%. For the base case of 95% capture rate an absorber packing height of 18 m (two packing sections of 9 m each) is selected, leading to a similar marginal decrease in the SRD of 1.4%.

6.3.2 Effect of lean solvent loadings

The effect of lean solvent CO$_2$ loading on the SRD and L/G ratio is illustrated in Figure 6-5 for a range of CO$_2$ capture efficiencies 95%/99%/99.72%, with absorber packing height initially assigned of 17 m/20 m/22 m respectively. At a lower lean solvent CO$_2$ loading, a smaller amount of solvent is required to achieve a certain CO$_2$ capture rate, yet the contribution of the heat of desorption to the SRD is more relevant. At a higher lean solvent CO$_2$ loading, more solvent is required to achieve a certain CO$_2$ capture rate, and thus the effect of sensible heat required to heat the solvent to the stripper temperature on the reboiler duty becomes more significant. There is therefore an optimal value of the lean solvent CO$_2$ loading that results in a minimum SRD.

On the other hand, this figure also shows that at high lean loadings, for instance, the lean loading of larger than 0.16 mol CO$_2$/mol MEA, the requirement for sensible heat for the loaded solvent surpassed the decreased requirement for desorption heat at higher loading, so that the overall result is increased SRD with higher lean loadings.
Figure 6-5 Sensitivity of the specific reboiler duty to the lean solvent CO$_2$ loading for a 35%wt MEA capture system for a range of CO$_2$ capture efficiencies 95%/99% /99.7%. The absorber packing heights are 17m/20m/22m for 95%/ 99% /99.7% capture efficiencies, respectively. For illustration purposes, the final packing height after optimisation might be different.

6.3.3 Effect of absorber intercooling

Absorber intercooling enables a shift in the thermodynamic vapour-liquid equilibrium, ensures high driving forces for CO$_2$ mass transfer through the column and consequently increases the rich solvent loading at the bottom of the absorber (Moullec & Neveux, 2016). The intercooling temperature has been typically set at 40°C for 90% CO$_2$ capture rate (Rezazadeh, Gale, Rochelle, & Sachde, 2017). Intercooling temperatures between 30°C and 40°C have been used for 99% capture level in recent studies (Gao et al., 2019). (Michailos & Gibbins, 2022) reported that the impact of intercooling on the rich loading is greater at higher lean loadings for 99% capture level.
The design of the capture plant conducted in this study shows that absorber intercooling is necessary for achieving ultra-high CO₂ capture rates above 99%. If intercooling is not implemented, the large amount of heat released leads to a significant high temperature bulge at the top of the absorber column, reducing the driving force for CO₂ transfer (Rezazadeh et al. 2017). For a given rich solvent loading, applying intercooling also decreases the required absorber packing height. As shown in Figure 6-6, lower intercooling temperature allows the absorber column to have a closer approach to equilibrium, leading to lower SRD. In this study, the intercooler is located after the first top packing section for 95% CO₂ capture (two packing sections of 9m each) and 99.72% CO₂ capture (three packing sections of 8m each). The solvent exits the absorber column at the end of the first packing section, passes through an external heat exchanger where it is cooled down to 25°C and returns to the column at the top of the second packing section.

Figure 6-6 Effect of absorber intercooling temperature on rich solvent loading and absorber packing height for achieving 99.72% CO₂ capture rate with lean loading of 0.16 mol CO₂/mol MEA.

6.3.4 Effect of stripper pressure

The effect of the stripper pressure on the SRD for a range of lean solvent loadings is also investigated as part of the optimisation of the PCC plant design. Figure 6-7 shows that at higher stripper pressures, the effect of the lean solvent loading on the SRD is less significant and the SRD required to achieve a given lean solvent loading is smaller. A higher stripper pressure requires, however, a higher reboiler
temperature to generate the steam to strip CO₂ from the loaded solvent. Therefore, there is a maximum stripper pressure (210Kpa) above which the required reboiler temperature (125°C) would be higher than the limiting temperature that accelerates thermal degradation of the solvent.

According to the VLE curve, increasing temperature will increase CO₂ partial pressure in the stripper, so the CO₂ fraction at the top of the stripper increases, while the water vapour fraction at the top of the stripper decreases. Additionally, the absolute values of both CO₂ partial pressure and water vapour partial pressure increase with increase of stripper pressure. A lower water vapour fraction at the top leads to a lower heat lost due to vaporization, so the SRD is smaller at higher stripper pressures.

For a constant stripper pressure, on one hand, a lower lean solvent loading leads to a higher solvent capacity and a smaller amount of solvent required to achieve a certain CO₂ capture rate, e.g., 99.72%, and thus the contribution of the sensible heat required to increase solvent temperature to the stripper temperature decreases. On the other hand, a higher heat of desorption is required to achieve a lower lean solvent loading. The overall effect is an increase in the SRD to achieve lower lean loadings at a constant stripper pressure which becomes more significant at very low lean solvent loadings. This implies that when designing the PCC plant with lower lean loadings, in order to minimize SRD, increasing stripper pressure (to a level that does not exceed solvent degradation temperature) should be considered.

An example in the figure 6-7 could justify the above discussion. For instance, at relatively high lean loading of 0.19 mol CO₂/mol MEA, the SRD changes slightly between 3.9 – 4.1 MJ/kg CO₂; whereas at lower lean loadings, 0.11 mol CO₂/mol MEA in the figure, since the effect of decreased sensible heat with increased stripper pressure becomes dominant, the SRD reduces from 6.1 to 4.1 MJ/kg CO₂, a much significant drop than the SRD under high lean loading of 0.19 mol CO₂/mol MEA.

Finally, it should also be noted that although increasing stripper can reduce the SRD under certain lean loadings, a maximum stripper temperature should always be monitored to avoid solvent degradation at high temperature, for instance, a maximum stripper temperature of 125 °C for MEA solvent.
Figure 6-7 Effect of stripper pressure on the specific reboiler duty for a range of lean solvent loadings. Under constant rich loading of 0.45 mol CO$_2$/mol MEA for 99.72% CO$_2$ capture rate.

6.3.5 Summary of PCC plant modelling results

Based on the design and optimisation procedure reported in this section, the key parameters of the CO$_2$ capture plant for the 90%, 95% and 99.72% CO$_2$ capture efficiency are presented in Table 6-3. The capture plant data obtained from the PCC system modelled in Aspen Plus are used as input parameters to the WtE plant model in gProcess to assess the effect of increasing the CO$_2$ capture efficiency on the performance of the WtE plant.
<table>
<thead>
<tr>
<th>Unit</th>
<th>CO₂ capture efficiency</th>
<th>90%</th>
<th>95%</th>
<th>99.72%</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Flue Gas</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flue Gas Flow Rate</td>
<td>kg/s</td>
<td>30.7</td>
<td>30.7</td>
<td>30.7</td>
</tr>
<tr>
<td>Inlet Temperature</td>
<td>℃</td>
<td>40</td>
<td>40</td>
<td>40</td>
</tr>
<tr>
<td>CO₂ Concentration</td>
<td>Mole Fraction</td>
<td>11.1</td>
<td>11.1</td>
<td>11.1</td>
</tr>
<tr>
<td><strong>Absorber</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Lean solvent flowrate</td>
<td>kg/s</td>
<td>61.2</td>
<td>63.8</td>
<td>69.9</td>
</tr>
<tr>
<td>Packing Height</td>
<td>m</td>
<td>16</td>
<td>18</td>
<td>24</td>
</tr>
<tr>
<td>Diameter</td>
<td>m</td>
<td>3.8</td>
<td>3.8</td>
<td>3.8</td>
</tr>
<tr>
<td>Packing Volume</td>
<td>m³</td>
<td>190</td>
<td>215</td>
<td>286</td>
</tr>
<tr>
<td>Intercooler Return Temperature</td>
<td>℃</td>
<td>30</td>
<td>25</td>
<td>25</td>
</tr>
<tr>
<td>Absorber Flooding</td>
<td>%</td>
<td>70%</td>
<td>68%</td>
<td>76%</td>
</tr>
<tr>
<td><strong>Heat exchanger</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rich Cold Solvent Inlet Temperature</td>
<td>℃</td>
<td>44</td>
<td>44</td>
<td>45</td>
</tr>
<tr>
<td>Rich Hot Solvent Outlet Temperature</td>
<td>℃</td>
<td>118</td>
<td>118</td>
<td>118</td>
</tr>
<tr>
<td>Lean Hot Solvent Inlet Temperature</td>
<td>℃</td>
<td>125</td>
<td>125</td>
<td>125</td>
</tr>
<tr>
<td>Lean Cold Solvent Outlet Temperature</td>
<td>℃</td>
<td>54</td>
<td>54</td>
<td>55</td>
</tr>
<tr>
<td>Rich Solvent pump</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure increase *</td>
<td>kpa</td>
<td>1200</td>
<td>1000</td>
<td>900</td>
</tr>
</tbody>
</table>
### 6.4 KPI results for the base case integration

Following the description of the base case thermal integration of WtE plants with PCC as described in the above sections, the KPI results under the base case integration will be summarized and analysed in this section. The KPIs were introduced in Chapter 4 and cover thermodynamic, economic and environmental aspects to give a thorough assessment of the integration options.

#### 6.4.1 Thermodynamic performance under the base case integration

The thermodynamic KPIs used to assess thermal integration approaches are:

1) the energy related output penalty (EOP and EUP) that is evaluated as the total net loss in energy output per unit of CO₂ captured; and

2) the efficiency related penalty (EP and EUFP) that is evaluated as a percentage point drop in the overall thermal efficiency of the plant; and

### Table: Stripper KPIs

<table>
<thead>
<tr>
<th></th>
<th>Unit</th>
<th>8</th>
<th>8</th>
<th>10</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Packing Height</strong></td>
<td>m</td>
<td>8</td>
<td>8</td>
<td>10</td>
</tr>
<tr>
<td><strong>Diameter</strong></td>
<td>m</td>
<td>2.5</td>
<td>2.5</td>
<td>2.5</td>
</tr>
<tr>
<td><strong>Packing Volume</strong></td>
<td>m³</td>
<td>63</td>
<td>63</td>
<td>79</td>
</tr>
<tr>
<td><strong>Stripper Flooding</strong></td>
<td>%</td>
<td>65%</td>
<td>62%</td>
<td>68%</td>
</tr>
<tr>
<td><strong>Lean Solvent CO₂ Loading</strong></td>
<td>mol CO₂/mol MEA</td>
<td>0.16</td>
<td>0.16</td>
<td>0.16</td>
</tr>
<tr>
<td><strong>Rich Solvent CO₂ Loading</strong></td>
<td>mol CO₂/mol MEA</td>
<td>0.47</td>
<td>0.47</td>
<td>0.46</td>
</tr>
<tr>
<td><strong>Stripper Pressure</strong></td>
<td>kPa</td>
<td>210</td>
<td>210</td>
<td>210</td>
</tr>
<tr>
<td><strong>Reboiler temperature</strong></td>
<td>°C</td>
<td>125</td>
<td>125</td>
<td>125.2</td>
</tr>
<tr>
<td><strong>Specific Reboiler Duty (SRD)</strong></td>
<td>MJ/kg CO₂</td>
<td>3.56</td>
<td>3.59</td>
<td>3.72</td>
</tr>
</tbody>
</table>

Note:

* The output pressure of the rich solvent pump is adjusted to prevent from flashing in the lean-rich solvent heat exchanger.
3) the Coefficient of Performance of steam extraction ($COP_{X,cap}$ and $COP_{X,cap+dh}$) that is essentially equivalent to the analogous $COP_{hp}$ of a heat pump and is compared with the $COP_{hp}$ of operational heat pumps operating at similar temperature range.

These three KPIs will be used to assess the thermal integration of PCC into power-only and CHP-type WtE plants. For each steam cycle configuration, the integrated PCC plant is operated at three different CO$_2$ capture rate: 90%, 95% and 99.72%. Detailed modelling results for the base case integration can be found in Appendix 6.1. Figure 6-8 shows the energy related KPIs of the WtE plant with PCC under base case: EOP for Power-only WtE plant and EUP for two types of CHP plants.

![Figure 6-8 EOP and EUP of PCC with WtE plants under three CO$_2$ capture rates](image)

It can be seen that among the three WtE configurations, the steam extraction and condensing configuration shows the lowest EUP of 231 kWh/tCO$_2$ to 276 kWh/tCO$_2$ for CO$_2$ capture rate ranging from 90% to 99.72%. Under this configuration, it is assumed that the WtE with PCC plants delivers the maximum possible DH output, and considering the design limit of minimum flow rate through the LP steam turbine cylinder (15% of the nominal flow rate at full load as described in the chapter 4.1.2).

The main reason for the lowest EUP being found to be for the steam extraction and condensing configuration is the relatively small difference in power output reduction compared with the other
two configurations. Additionally, the integration of CO₂ capture does not affect the DH output capacity, which means the DH output with and without PCC maintains the same value.

By contrast, the backpressure configuration shows the highest EUP with PCC. The main reason for this is the large difference between the DH output under with and without PCC scenarios in this configuration. Taken the 90% capture rate as an example, Table 6-4 shows the comparison in terms of power and DH heat output in EX-C configuration and BP configuration.

Table 6-4 Power and DH heat output of the WtE-CHP plants with and without PCC

<table>
<thead>
<tr>
<th></th>
<th>WtE CHP-EX&amp;C configuration</th>
<th></th>
<th>WtE CHP-BP configuration</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Without PCC</td>
<td>With PCC</td>
<td>Without PCC</td>
</tr>
<tr>
<td>Power output</td>
<td>10.64</td>
<td>6.77</td>
<td>7.61</td>
</tr>
<tr>
<td>Heat output</td>
<td>13.00</td>
<td>13.00</td>
<td>35.06</td>
</tr>
</tbody>
</table>

It can be seen that, in the WtE-CHP-EX&C configuration, an assumption is made in this study that the WtE plant prioritises DH output, so that the WtE plant can maintain DH output of 13MW when PCC is added, with energy penalty occurs in the power output, from 10.64MW to 6.77MW. The 13MW DH output here is the maximum amount of DH capacity the plant can provide with PCC, under the configuration.

The WtE-CHP-BP configuration represents the scenario where the WtE plant has the high DH demand, so that the low-pressure cylinder is by-passed, and all the steam outlet from the last stage of the medium-pressure cylinder is used either for DH or for solvent regeneration at the PCC plant. Without PCC, the WtE-CHP plant can provide 35MW DH heat; when the PCC is added to the plant, the certain percentage of the extracted steam should go to the PCC plant, the DH output is reduced to 18.62MW. The reduction of power output mainly comes from the power consumption at the PCC plant, such as booster fan, solvent pumps, compression train, etc. The large different in the total energy (power and DH heat) under CHP-BP configuration leads to the highest EUP under this configuration. It should be noted that the result of this comparison is strongly affected by the pre-determined operation strategy, such as the prioritised energy type. The advantage of WtE-CHP-EX&C configuration is to maintain the DH output as long as the DH demand is within its operation range. However, backpressure has the advantage in terms of total energy output (and efficiency metrics, which will also be discussed later in this chapter). For instance, if not considering the exergy difference between power and heat, the WtE-
CHP-BP plant has the higher total energy output comparing with CHP-EX&C configuration, with values of 25MW and 20MW, respectively.

We can also see in Figure 6-8, the small variations (approximately 30 kWh/tCO₂ increase from 95% to 99.72%) of EOP (and EUP) with CO₂ capture rate varying from 90% to 99.72% shows the effect of CO₂ capture rate is relatively small compared to the effect of different WtE operation configurations on the EOP (and EUP) results. This is mainly due to the fact that the difference of SRD for the three CO₂ capture rates is small (as shown in the PCC plant modelling results in Chapter 6.3), thus leads to small difference in terms of thermal heat from the steam cycle.

Figure 6-9 shows the effects of PCC on efficiency related penalties (EP and EUFP) that are evaluated as a percentage point drop in the overall thermal efficiency of the plant. The CHP-BP configuration has the highest EUFP due to CO₂ capture (and compression); in the range of 32.8% to 37.7% for CO₂ capture rates from 90% to 99.72%. This is mainly because that without PCC, there are no condensing heat losses in this configuration and the WtE plants are delivering the highest efficiency, and the addition of PCC reduces the DH output, approximately 16.6MW, 17.6MW and 19.2MW of thermal heat is delivered to the capture plant for solvent regeneration (for CO₂ capture rate of 90%, 95% and 99.72%).

The steam extraction and condensing configuration shows the lowest EUFP. Previous explanations comparing the EUPs under difference CHP configuration also works here. Under this configuration, the DH output capacity is maintained with and without PCC, so the only energy penalty comes from the power output reduction. Again, the small variations of EP (and EUFP) under CO₂ capture rate from 90% to 99.72% shows the effect of CO₂ capture rate is relatively small compared to the effect of different WtE operation configurations on the EP (and EUFP) results.
Although the CHP-BP configuration shows the highest EUFP, it has advantage in terms of energy output and efficiency. For instance, comparing the absolute Energy utilization factor, as shown in Table 6-5 below.

Table 6-5 Energy utilization factor of the WtE-CHP plants with and without PCC

<table>
<thead>
<tr>
<th></th>
<th>WtE CHP-EX&amp;C configuration</th>
<th>WtE CHP-BP configuration</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Without PCC</td>
<td>With PCC</td>
</tr>
<tr>
<td>Energy utilization factor (EUF)</td>
<td>43%</td>
<td>36%</td>
</tr>
</tbody>
</table>

In Table 6-5, we can see that without PCC, the EUF of the CHP-EX&C and CHP-BP configuration is 43% and 77% respectively. With PCC, due to the heat consumption for solvent regeneration and power consumption at the capture & compression, the EUF of the BP configuration is reduced to 44%, a value similar to that under CHP-EX&C configuration without PCC. There is also a drop of EUF for the EX&C configuration, but the drop is less considerable as that under BP configuration.

Figure 6-10 shows the results of Coefficient of Performance of steam extraction ($COP_{X\_cap}$ and $COP_{X\_cap+dh}$) for the investigated WtE plants with PCC. The overall COP$_X$ values are in range of 5 to 5.2 with variations in different configurations and PCC capture rates. It should be noted here that heat...
pumps for 130°C heat delivery (lift temperature of about 100°C) at the required capacity are not commercially available (Arpagaus, Bless, Uhlmann, Schiffmann, & Bertsch, 2018). Even if they were commercially available, assuming the heat pumps are recovering the ambient excess heat at temperature level of 25°C, the COP of an electricity driven heat pump would be much lower than 5. This means that for all the investigated WtE configurations with PCC, steam extraction is relatively more efficient method to provide thermal heat for solvent regeneration in the reboiler.

![Figure 6-10 Coefficient of Performance for (steam) extraction of PCC with WtE plants under three CO₂ capture rates.](image)

It is also seen in the Figure 6-10 that the WtE-CHP-BP configuration shows the highest $COP_X$ (5.23-5.24 for the three CO₂ capture rates) comparing that in the other configurations. It should be noted here that the $COP_X$ result under the CHP-BP configuration with PCC represents scenarios where all the steam outlet from the last stage of the intermediate pressure turbine is utilized by DH heating and solvent regeneration, so no heat is losses. Under scenarios when the DH demand is reduced, the excess heat has may be no longer needed, the valuable heat $Q_{cap} + Q_{dh,pcc}$ used to calculate the $COP_{X-cap+dh}$ will reduce. Similarly, for the CHP-EX&C configuration with PCC, the $COP_X$ values in Figure 6-10 represent scenarios where the maximum DH supply is met for each CO₂ capture rate and this DH supply capacity is maintained with and without PCC. Under scenarios where DH demand is
reduced, less steam will be extracted from the steam cycle, so that relatively more steam can be used for power generation in the lower pressure turbine. The overall result would be increased $COP_X$, to the values close to those for power-only configurations, approximately 5.13.

As mentioned in Chapter 3.1.3, in addition to comparing steam extraction with a traditional electricity driven heat pump, an analogous absorption type heat pump is used to assess the efficiency of thermal heat delivery for solvent regeneration by steam extraction. Table 6-6 shows that the $COP_{X,abs}$ of steam extraction is around 0.9 and 0.65 for WtE-CHP extraction and condensing configuration and backpressure configuration, respectively. The lower $COP_X$ value of the backpressure configuration is due to the large amount of DH output reduction when PCC is integrated. This increases the consumed driven heat source that is utilized for the absorption heat pump, resulting in a lower COP for the same amount of regeneration heat delivery at 130°C.

A comparison can be made in terms of thermal efficiency of heat generation with a typical absorption heat pump. There are mainly two types of absorption heat pumps, which are usually sorted depending on application areas. Among them, type-II absorption heat pumps, or temperature increasing absorption heat pump, place the generator and condenser in the low pressure side while absorber and evaporator in the high pressure side, thus are used for heating or temperature boosting. The COP of a Type-II absorption Heat Pump working under the same conditions is typically in the range of 0.4-0.5 (Kim et al. 2022). By this comparison, it can be concluded that the steam extraction from the steam cycle provides a relatively more efficient approach for solvent regeneration under the assumptions made in this study, comparing with using heat pump to provide thermal heat for solvent regeneration.

Table 6-6 $COP_{X,abs}$ of steam extraction for WtE-CHP plants with PCC

<table>
<thead>
<tr>
<th>$COP_{X,abs}$</th>
<th>WtE-CHP (Ex-C)</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>90% PCC</td>
<td>95% PCC</td>
</tr>
<tr>
<td>0.903</td>
<td>0.903</td>
<td>0.905</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th></th>
<th>WtE-CHP (BP)</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>90% PCC</td>
<td>95% PCC</td>
</tr>
<tr>
<td>0.642</td>
<td>0.646</td>
<td>0.656</td>
</tr>
</tbody>
</table>
In this study, the concept of $COP_{X,\text{cap}}$ and $COP_{X,\text{abs}}$ represents energy efficiency metrics under different approaches to extract heat from the overall integrated system (steam cycle, PCC, and district heating), to provide heat for solvent regeneration (in the CO$_2$ capture plant) and DH system.

A direct comparison of results of $COP_{X,\text{cap}}$ and $COP_{X,\text{abs}}$ as presented above shows that the $COP_{X,\text{abs}}$ is much lower than $COP_{X,\text{cap}}$, and readers may conclude that absorption heat pump solution is less efficient than the extraction (losing power) solution, however this comparison is fairly arbitrary. As introduced in Section 3.3.3, the $COP_{X,\text{cap}}$ and $COP_{X,\text{abs}}$ are essentially analogous COP of compression heat pump and absorption heat pump. The comparison of compression heat pump and absorption heat pump is complex since they are driven by different sources and working in different conditions: compression heat pump is driven by electrical and absorption heat pump is driven by thermal energy.

To overcome this limitation, a multi-criterion evaluation is reported by (Xu et al. 2022), which includes exergy-to-energy ratios, exergy efficiency, exergy rate analysis, etc. They conclude that COP of compression heat pump is much more sensitive than the absorption heat pump since it can extract “cheaper” exergy from waste heat under small temperature lift. Under large temperature lift, the absorption heat pump becomes more competitive.

Based on this, for the $COP_{X,\text{cap}}$ and $COP_{X,\text{abs}}$ presented in this thesis, similar approaches could be investigated to give a comprehensive comparison between the two heat extraction approaches in future research.

6.4.2 Economic performance under the base case integration

Following the cost methodology described in Chapter 5, the cost value (Capex and Opex of the WtE plant and PCC plant) calculated for the base case integration is shown in Table 6-7.

Table 6-7 Cost estimation for base case thermal integration of WtE with PCC

<table>
<thead>
<tr>
<th>NO.</th>
<th>Capex Cost items</th>
<th>Value (in million £)</th>
<th>NO.</th>
<th>Opex Cost items</th>
<th>Value (in million £)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.1</td>
<td>WtE Power-only plant</td>
<td>188</td>
<td>2.1</td>
<td>WtE Power-only plant</td>
<td>5.80</td>
</tr>
<tr>
<td>1.2</td>
<td>WtE CHP plant</td>
<td>228</td>
<td>2.2</td>
<td>WtE CHP plant</td>
<td>6.11</td>
</tr>
<tr>
<td>1.3</td>
<td>PCC plant @90% CCR</td>
<td>33.8</td>
<td>2.3</td>
<td>PCC plant @90% CCR</td>
<td>1.51</td>
</tr>
</tbody>
</table>
The following content presents the economic KPIs (LCOE, LCOH, Cost of CO₂ captured) for the base case integration of WtE with PCC, using the KPIs identified in the Chapter 3.2. In the economic analysis, it is assumed that the investigated WtE-power only plant has a utilization factor of 87% (based on the operational WtE plant in Edinburgh), which gives an annual 7,650 working hours for electricity generation. For WtE CHP plants, there is uncertainty in terms of DH supply period. A default operation percentage of 50% of the annual operation hours is assumed for CHP operation, this represent the average heating season for the UK (BEIS 2013). For all cases, a CO₂ emission price of £40/tCO₂ is assumed and this cost applies to only fossil CO₂ emitted from the WtE plants. It should be noted that this CO₂ price represents EU CO₂ price at the time of the writing (2021) and may subject to change along with the development of carbon market. At the time of the writing, biogenic CO₂, which is generated by the combustion of biomass and bio-liquids, is not subject to carbon pricing under current UK ETS carbon pricing rules (subject to the application of certain sustainability criteria in some circumstances)[BEIS, 2022b]. Due to the uncertainty over whether negative emissions sales or related credits will be allowed, biogenic CO₂ emitted and or captured is priced zero.

Figure 6-11 shows the results of LCOE for the investigated WtE plant with and without PCC. It can be seen that before PCC, the LCOE is £85/MWh, which represents a minimum selling price per MWh of electricity for the WtE plant where the revenues equal costs. With integration of PCC at 90% CO₂ capture rate, the LCOE increases to £162/MWh, almost two times the corresponding value for the original unabated plant.

<table>
<thead>
<tr>
<th>1.4</th>
<th>PCC plant @95% CCR</th>
<th>35.2</th>
<th>2.4</th>
<th>PCC plant @95% CCR</th>
<th>1.55</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.5</td>
<td>PCC plant @99.72% CCR</td>
<td>38.3</td>
<td>2.5</td>
<td>PCC plant @99.72% CCR</td>
<td>1.64</td>
</tr>
</tbody>
</table>
Figure 6.11 Break down of LCOE of WtE plant without and with PCC under three CO₂ capture rates, with assumptions of fossil CO₂ price £40/tCO₂, gate fee £100/tMSW, no negative emission credit is considered.

It is also seen that when increasing the CO₂ capture rate from 90% to 95% and from 95% to 99.72%, there is only a 3.7% and 6.5% increase in LCOE respectively. This result indicates that LCOE of WtE plant with high CO₂ capture rate can be similar to the LCOE of WtE plant with 90% CO₂ capture, thus supporting the most up to date suggestions from both literature and government policies on advancing to high CO₂ capture (Gibbins & Lucquiaud, 2021).

In the above cases, for WtE plant with and without PCC, it is assumed that the plants receive the constant amount of gate fee of £100/t MSW (Scottish government 2024), so that the same revenue received under the same amount of waste treatment. The variations of LCOE on the gate fee share in Figure 6.11 is mainly due to the variations of power output among different scenarios.
It should be noted that in Figure 6-11, the relatively conservative CO\(_2\) price of £40/tCO\(_2\) is applied. When the CO\(_2\) price increases, the difference of LCOE with and without PCC should be reduced. A sensitivity analysis has been completed for a range of CO\(_2\) prices and is presented in Figure 6-12.

![Figure 6-12 LCOE of WtE-Power-only plant under different CO\(_2\) prices](image)

It can be seen in Figure 6-12, when the CO\(_2\) price increases from £40/tCO\(_2\) to £150/tCO\(_2\), the LCOE for the WtE plant without PCC increases from £85/MWh to £128/MWh. The LCOE under WtE with PCC shows smaller variation due to the low fossil CO\(_2\) emission under the abated plants, with values around £160/MWh to £180/MWh for the three CO\(_2\) capture rates. Results also find that only the CO\(_2\) price increases to approximately £280/tCO\(_2\), the WtE plant with PCC can benefit relatively lower LCOE with
CO₂ capture, comparing with the LCOE of WtE plants without PCC. It is still remains to be seen whether this high CO₂ price could be reached in the short term in real world though.

For the two WtE CHP plants with PCC, a similar KPI is used to assess the influence of PCC on the DH price. The results are shown Figure 6-13 in below. It presents the LCOH for the two types of WtE-CHP plants investigated in this study, with and without PCC. In this simulation, for all the cases, it is assumed that the electricity price is £85/MWh, the value of LCOE for power-only WtE plant without PCC. Similarly, as in the LCOE calculations above, all fossil CO₂ emissions are assumed to incur a CO₂ price of £40/tCO₂.

Figure 6-13 Breakdown of LCOH of WtE-CHP plant without and with PCC under three CO₂ capture rates, with assumption of electricity selling price £85/MWh, fossil CO₂ price £40/tCO₂

As can be seen from Figure 6-13, the lowest LCOH occurs for the WtE-CHP-BP configuration without PCC. This is largely due to the highest DH output of 35 MW occurring for this configuration. When the CHP plants are operating with PCC, the benefit of high DH output leads to the CHP-BP with PCC showing much lower LCOH (£171-£205 per MWh DH output), when compared to the CHP-EX&C with
PCC cases (£236-£303 per MWh). It is also seen that, for the WtE CHP-BP with 90% PCC, the LCOH is £171/MWh, a value close to that of the WtE-CHP-Ex&C without PCC (£174/MWh).

As discussed in Chapter 3.2.3, the cost of CO₂ captured is applied as the third main KPI from the economic point of view. Figure 6-14 presents the results for the investigated WtE plant configurations with PCC integrated.

![Figure 6-14 Cost of CO₂ captured £/tCO₂ for the investigated WtE plants with PCC](image)

The DH supply period is also an important uncertainty. The illustrative cases in this work assume that the DH supply period is defined as 50% of the utilization period (a whole year). When calculating the cost of CO₂ captured for CHP plants with PCC, the electricity-selling price is estimated to be £85/MWh, the LCOE of the referenced power-only WtE plant without PCC.

Results shows that the cost of CO₂ captured for power only WtE plant is in range of £43-£44/tCO₂. For CHP plant, it is found to be in range of £83-£148/tCO₂, with strong sensitivity on the design inputs, such as district heating period and energy prices. For instance, as can be seen from the Figure 6-15, if...
the DH supply period is increased to 95%+ of the plant utilization period, the cost of CO₂ captured of the WtE-CHP-EX&C with PCC plant could be comparable with power-only plant.

Here the author presents a horizontal comparison in terms of the cost of CO₂ avoided by WtE-CCS with wider BECCS technologies. In this thesis, the cost of CO₂ captured in this thesis only consider the cost of CO₂ capture and compression, without the cost of transport and storage. If referencing literature values and give the cost of CO₂ transport and storage to be in range of $5-45/tCO₂ (Bacilieri, Black, and Way 2023); The cost of CO₂ avoided for WtE plant would be $47-184/tCO₂.

Comparing with literature, the cost of CO₂ avoided for BECCS technologies (combustion) could be found in range of $88-288/tCO₂ (Global CCS Institute 2019). In this regard, the cost of CO₂ avoided for WtE with CCS lies in the lower middle of the cost range of wider BECCS technologies. It should be noted that this cost of CO₂ avoided for WtE with CCS only applies to the design and economic parameters as defined in the evaluated case studies. In real operation, the cost of CO₂ avoided should consider the sensitivity of feed input, MSW composition (thus CO₂ emission), economy of scales, integration approaches, capacity factors, etc.
6.4.3 Environmental performance under the base case integration

In this study, direct CO\textsubscript{2} emission intensity on three main products bases is defined as the primary environmental KPIs. These are direct CO\textsubscript{2} emission intensity on fuel basis, on electricity basis and on heat basis. All case studies reported in this thesis use the same WtE plant using direct combustion over a moving grate, constant 500t/d MSW consumption, with MSW composition taken from the reference WtE plant in Edinburgh, resulting in the same direct fossil CO\textsubscript{2} emission without PCC.

Once the PCC is integrated, the direct CO\textsubscript{2} emission only varies depending on the CO\textsubscript{2} capture efficiencies. Figure 6-16 shows the results of direct fossil CO\textsubscript{2} emission on fuel basis. For all the investigated WtE plants using the same MSW composition as described in Chapter 4.1, with an assumption of 60% organic carbon in the total carbon in the fuel, the total direct CO\textsubscript{2} emissions from fuel combustion are 891 kg/t MSW. These total direct CO\textsubscript{2} emissions consist of 535 g CO\textsubscript{2}/kg MSW that is of biogenic origin (considered to be carbon neutral) and 353 g CO\textsubscript{2}/kg MSW that is of fossil origin. As illustrated in Figure 6-16, when the WtE plant is equipped with PCC, the amount of biogenic CO\textsubscript{2}
captured and permanently stored leads to ‘negative’ carbon emissions: -0.45 kg CO₂/kg MSW, -0.49 kg CO₂/kg MSW, and -0.53 kg CO₂/kg MSW for 90%, 95% and 99.72% respectively.

Figure 6-16 Carbon emission intensity of WtE plant without and with PCC on fuel basis

As mentioned in Chapter Error! Reference source not found., an efficiency allocation method with a fixed ratio of heat and power generation efficiency set to be 2 as used by the 1/3: 2/3 Method (from
the Digest of UK Energy Statistics), is applied to the power and heat produced in the WtE plant in this work. Figure 6-17 shows the resulting CO₂ emission intensity on electricity and heat basis.

![Figure 6-17 Carbon emission intensity of WtE plant without and with PCC on electricity and heat basis](image)

The carbon intensity of the power only WtE facility without PCC modelled in this work is approximately 519 kgCO₂/MWhₑ, comparable to the global average for electricity generation from all fuels of 475 kgCO₂/MWhₑ reported by the International Energy Agency (IEA, 2019). This value is reduced to approximately 436 kgCO₂/MWhₑ and 296 kgCO₂/MWhₑ for WtE-CHP facilities operating at EX&C configuration and backpressure configuration respectively. Implementing CO₂ capture and permanent storage to a WtE plant considerably reduces the carbon intensity leading to net negative CO₂ emissions. For example, the WtE-power-only plant with CO₂ capture removes between 988 and 1272
\( \text{kgCO}_2 \) per MWh of exported electricity, with the variation in CO\(_2\) removal caused by different CO\(_2\) capture rates.

In terms of direct fossil CO\(_2\) emission on a heat basis, without PCC, the value are 218 kg CO\(_2\)/MWh and 148 kgCO\(_2\)/MWh for steam extraction & condensing configuration and backpressure configuration respectively. This is comparable to the UK’s Committee on Climate Change (CCC) suggested average emissions from domestic and commercial heat, which is around 180 kgCO\(_2\)-eq/MWh in 2030 (House of Parliament, 2016). Implementing CO\(_2\) capture and permanent storage to a WtE plant also considerably reduces the carbon intensity leading to net negative CO\(_2\) emissions on a heat basis. For example, WtE-CHP plant with CO\(_2\) capture removes between 306 and 469 kgCO\(_2\) per MWh of DH produced, with the variation in CO\(_2\) removal caused by different CO\(_2\) capture rates.

Finally, a discussion of the merit of using the DUKES method that arbitrarily sets \( f_Q/f_E = 2 \) is appropriate at this point as it may impact the results presented in Figure 6-17. The \( f_Q \) and \( f_E \) in this study vary depending on the heat and power output in different case studies, and they ratio may deviate from the value of 2 that is assumed by the DUKES method as used by BEIS (Department for Business, energy & industrial Strategy) when determining fuel consumption for heat an electricity in the UK. While the DUKES method was retained here for ease of comparison, it means that the results in Figure 6-17 could be refined on the basis of the actual \( f_Q/f_E \) ratio that we can readily estimate here.

In the base case study, the boiler thermal input is 55.6MW; in power-only case, the electricity generation is 14.33MW, with electricity generation efficiency of 26%;

For the same base case boiler, with boiler thermal input of 55.6MW and boiler efficiency 80%, the boiler generates live steam at 18.2kg/s, 44MW, 400\(^\circ\)C/40bar (superheated steam); assuming this superheated steam is used for district heating with return water temperature of 60\(^\circ\)C,

\[
\left(\frac{18.2kg}{s} \times \frac{4.186J}{g\cdot^\circ C} \times 60^\circ C \approx 4.6MW\right); \text{ the total district heat is approximately } 44MW-4.6MW=39MW,
\]

the heat generation efficiency is \( \frac{39MW}{55.6MW} = 70\% \); in real applications, this efficiency may be lower considering heat transfer efficiency/heat loss;

Therefore, the corresponding ratio of heat generation efficiency to electricity generation efficiency is \( f_Q/f_E = 70\%/26\% = 2.7 \). This ratio should be lower if considering the heat transfer efficiency and heat loss, and only represents the result of this case study, bringing it closer to 2. However, this suggests
revising values for carbon intensity for heat down and those for carbon intensity for electricity up, so that the ratio of lengths of blue bar to orange bar on Figure 2-17 is now 2.7 instead of 2.

6.4.4 Summary of KPI results under the base case integration.

In this chapter, the KPI results for the base case heat integration of WtE plant (power only, EX&C and BP configuration) are presented and analysed.

Of all the investigated WtE plants integrating with PCC, the EOP for power only plant is approximately 294 kWh/tCO₂ to 302 kWh/tCO₂ for CO₂ capture rate ranging from 90% to 99.72%. Thermodynamic KPI results of CHP application shows large variation and is strongly affected by the pre-determined operation strategy. For instance, the CHP plant in EX&C configuration which maintains DH output with PCC shows the lowest EUP of 231 kWh/tCO₂. Under scenarios where DH output is affected with PCC, EUP is increased to 1125 kWh/tCO₂ for the same set of CO₂ capture rates.

Economic KPI results show that for WtE-power-only plant, the LCOE of the abated plant is almost two times as the original unabated WtE plant. When the CO₂ price increases to approximately £280/tCO₂, the WtE plant with PCC can benefit from relatively lower LCOE with CO₂ capture, compared to the LCOE of WtE plants without PCC. For WtE-CHP plants, the economic results strongly relate to the DH output (or heat to power ratio), duration of DH period etc. For instance, under high DH demand (CHP plant in BP configuration), the LCOH with PCC could be comparable with the LCOH of the WtE CHP plant without PCC. The COPₓ results show that for the investigated WtE plant with PCC by basic heat integration, steam extraction is a more efficient approach to provide thermal heat for solvent regeneration, than using heat pumps (if the corresponding heat pump at the temperature level and capacity is commercially available).

Environmental KPI results show that with an assumption of 60% biogenic carbon in the carbon in the fuel, the power only plant delivers considerably negative CO₂ intensity from -998 to -1272 kgCO₂/MWhₑ. If considering the carbon emissions including land management, biomass processing and transportation, transporting the captured CO₂ through pipelines and operating the geological storage infrastructure, then 383kg CO₂eq MWh⁻¹ are emitted (Almena et al. 2022). Then the net negative emission potential of WtE-power-only plant would be 615-889kg CO₂eq MWh⁻¹.

Taking the most recent report in terms of power generation from WtE plants in the UK, in 2021, the gross power generation is reported to be 10,060 GWh(e) (Tolvik 2021). It can be calculated that the net negative emission potential would be 6-9 million tons of CO₂ per year in power only WtE plant in the UK. As estimated, BECCS could deliver 20 to 70 Mt CO₂ annual negative emissions for the UK (Almena et al. 2022). The WtE power only plant could contribute approximately 9%-45% of total negative emission of BECCS in the UK.
For CHP plants where two types of energy are produced, with DUKES (Digest of UK Energy Statistics) ‘1/3:2/3’ applied to allocate CO₂ to power and heat output respectively, both CHP plants can generate negative CO₂ emission on heat basis.

For all the thermodynamic and economic KPI results, it is also shown that the impact of CO₂ capture rate is minimal. Although higher CO₂ capture rate shows higher energy penalty (thermodynamic KPIs), higher cost (economic KPIs), the environmental benefit in terms of negative emission electricity and heat production could be valuable in the WtE sector.

The overall results under this integration approach could be helpful to understand the impact of CO₂ capture (the most commercially available MEA based post-combustion CO₂ capture technology) on WtE plants, from thermodynamic, economic and environmental aspects. These KPIs results will be used to compare the KPIs under scenarios with advanced heat recovery from the PCC plant, which will be presented in the following chapter.
7 Advanced heat recovery for CHP application

Heat integration of CO₂ capture with WtE plant will inevitably result in an energy penalty for the plant, as has been discussed in the previous chapter. A recent study identifying the Best Available Techniques for PCC prepared by (Gibbins & Lucquiaud, 2021) for the UK Environment Agency suggests that various heat recovery concepts may theoretically be viable (i.e. consistent with the 2nd Law of Thermodynamics); although capital cost and reliability, availability, maintainability, operability (RAMO) considerations and specific characteristics of the PCC system might constrain implementation. In this chapter, in order to reduce the energy penalties of CO₂ capture, an advanced approach to heat recovery is developed for all the three representative WtE plants. In this context, the targets/principles of the advanced heat integration can be summarized as:

- Either to maximise heat available for DH, or integration of CO₂ capture will not affect the capability of DH supply;
- Achieve best available thermal efficiencies of the integrated power plants; and
- Achieve best RAMO (Reliability, Availability, Maintainability, and Operability) of the integrated power plants.

In this chapter, the integration approaches for this advanced heat integration will be described, for the investigated three WtE configurations with different CO₂ capture rates as used in chapter 6. Following this, the KPI results from thermodynamic, economic and environment aspects will be presented and discussed.

7.1 Advanced heat integration of WtE-CHP plants with PCC

The advanced thermal integration option introduces additional modifications in the heat exchanger network with the objective of maximizing the net power output for a power-only WtE plant and maximizing thermal output for a WtE-CHP plant equipped with PCC. To do this, engineering-based design approach is applied when determining the heat recovery process of the heat exchanger network, and experienced pinch temperature values, choices of heat exchanger and heat pumps are determined based on experienced values that either sourced from literature or the author’s own judgement. For future work, the design of heat exchanger network could be done in a more systematic way, for instance, using pinch analysis, to find the optimal pinch temperature and heat exchanger network design flowsheet.

Excess heat can be recovered from three locations in the PCC process: the compressor intercoolers, the stripper overhead condenser and the DCC.
In literature review, a majority of the studies have focused on effective heat integration of CO₂ capture plant with the power cycle, by means of heat recovery from the capture process for feed water heating. In this study, the same approach is applied, to examine the heat available for heat recovery for power-only operation, and the effect of this approach for the representative WtE plant with typical operational parameters. The temperature of regeneration in the CO₂ capture plant determines the temperature available for heat recovery for feed water heating. For lower regeneration temperatures, a smaller amount of steam is extracted from the power cycle and the power output thus increases.

For CHP operations, the recovered heat is used in the DH system. The DH water return splits into five streams: four of them are heated in heat exchangers from 60°C to 80°C using relatively high-temperature heat from the compressor intercoolers and the stripper overhead condensers. A fifth stream is heated to 77°C by a heat pump and then by steam extraction to the final DH supply temperature of 80°C. Steam condensation will occur for gas to liquid heat exchange. The pinch temperature is set to be 15°C for heat recovery from the stripper overhead condenser (March 1998).

The heat pump is used in the fifth stream to recover heat from DCC cooling water since the temperature level at the DCC is relatively low and, therefore, not high enough for direct use. Heat from DCC cooling water is recovered in the heat pump evaporator, increasing the DH water temperature from 60°C to 77°C (through private communication with a heat pump vendor, this is the highest temperature for a commercially available single stage centrifugal compression heat pump and the COP of the heat pump is approximated to be 5.5).

The heat pump can produce about 12MWth DH heat, with a power consumption of about 2.2MWe. It should be noted that the \( COP_{hp} \) of heat pump is closely related to the temperature lift (which is the temperature difference between condensing temperature and evaporating temperature) of the heat pump. For example, reducing the temperature lift can increase the \( COP_{hp} \), thus reducing the power consumption. The use of heat pumps in WtE plants with CCS has been proposed in several studies, such as the FEED study of the Oslo WtE-CHP plant (Varme, 2021), and a recent report by IEAGHG (IEAGHG, 2020). In the FEED study of the Oslo WtE plant, the Heat Pump Package will be used to recover the heat from the condenser of the carbon capture plant (exothermal process) and is to be finalised during project execution (Varme, 2021). In the IEAGHG report on WtE with CCS, to hinder the energy conflict between the district heating and the CO₂ capture, a heat pump is placed to recover energy from the Direct contact cooler (DCC) upstream of the absorber (IEAGHG, 2020). In this study, excess heat from the DCC, carbon capture plant (stripper overhead condenser) and the CO₂ compression train are recovered by the heat pump, which represents the maximum amount of heat recoverable from the PCC facility.
A schematic representation of the available heat sources in the PCC process and the thermally integrated configuration used in the advanced integration case is illustrated in Figure 7-1.
Figure 7-1 Schematic representation of the additional heat recovery options considered in the advanced heat integration configuration of a WtE-CHP plant with post-combustion CO₂ capture.
7.2 KPI results for advanced heat recovery for CHP application

This section presents and discusses the KPI results under the advanced integration introduced in the previous section. As noted previously, the KPIs cover thermodynamic, economic and environmental aspects, giving a thorough assessment of the integration.

7.2.1 Thermodynamic performance under advanced heat integration

Following the same approach as in Chapter 6.4.1, the thermodynamic KPIs used to assess thermal integration of WtE plant with PCC under advanced heat recovery are:

1) the energy related output penalty (EOP and EUP) that is evaluated as the total net loss in energy output per unit of CO₂ captured;

2) the efficiency related penalty (EP and EUFP) that is evaluated as a percentage point drop in the overall thermal efficiency of the plant; and

3) the Coefficient of Performance of steam extraction (\(COP_{X_{-cap}}\) and \(COP_{X_{-cap+dh}}\)) that is essentially equivalent to the analogous \(COP_{hp}\) of a heat pump and is compared with the \(COP_{hp}\) of real operational heat pumps.

These three KPIs will be applied to assess the thermal integration of PCC into power-only and CHP type WtE plants. For each configuration, the PCC plant is operating at three different CO₂ capture rates: 90%, 95% and 99.72%. Detailed modelling results for the advanced heat integration can be found in Appendix 7.1.

In order to give an understanding of the effect of advanced heat regeneration, Figure 7-2 shows the energy related KPI of the WtE plant with PCC under both the base case and the advanced thermal integration case: EOP for Power-only WtE plant and EUP for two types of CHP plants. For each configuration at a certain CO₂ capture rate (90%, 95% and 99.72% in this study), the EOP and EUP values for advanced heat integration have similar trends when compared with the base case. That is, the EOP and EUP has the highest value for the backpressure configuration due to large DH output reduction with PCC. The EOP and EUP shows the lowest value under steam-extraction and condensing configuration, due to the relatively low power output under CHP operation.
Figure 7-2 EOP or EUP of PCC with WtE plant under advanced heat integration and three CO₂ capture rates

It is also seen that for the power-only configuration, advanced heat integration reduces the energy output penalty of PCC by about 30 – 40 kWh/tCO₂. This reduction is mainly due to the heat recovery from the stripper overhead condenser for the boiler feed-water heating. Similarly, for the steam extraction and condensing configuration, advanced heat integration reduces the EUF by about 75 kWh/tCO₂. This is mainly due to heat recovery from the PCC process for DH heating leading to less steam extraction from the original steam cycle.

The highest EUP occurs for the backpressure configuration with base case heat integration, with values of from 1084 kWh/tCO₂ to 1125 kWh/tCO₂. It is seen that heat recovery greatly improves the situation. For this configuration, the WtE CHP plant produces as much as heat as possible (35MW) and all the excess heat from the PCC process (16.4MW - 7.4MW) is recovered to produce DH heat. The DH output capacity could be at a similar level to an equivalent plant without PCC. For the heat recovery from the DCC process, the temperature level at the DCC is relatively low, so a heat pump is used to increase the temperature from 60°C to 77°C. The heat pump is assumed to be using a screw type compressor, using
R717 (ammonia) as refrigerant, with temperature lift of 60°C, COP of 3 (Jesper et al. 2021). The DH water is further heated by additional steam extraction from the steam cycle to increase the temperature to the required 80°C.

In Figure 7-2, the small variations of EOP (and EUF) for CO₂ capture rate varying from 90% to 99.72% shows the effect of CO₂ capture rate is relatively small compared to the effect of different WtE operating configurations on the EOP (and EUF) results.

Figure 7-3 shows the effects of PCC on efficiency related penalty (EP and EUFP) that is evaluated as a percentage point drop in the overall thermal efficiency of the plant. The CHP-BP configuration has the highest EUFP due to CO₂ capture (and compression); in the range of 32.8% points to 37.7% points for CO₂ capture rates from 90% to 99.72%. This is mainly because there are no condensing heat losses in this configuration and the WtE plants are delivering the highest efficiency. Heat recovery greatly improves the situation. By heat recovery from the PCC plant, the EUFP can be reduced to around 7% points to 9% points, a similar level as that under Power only configuration with PCC. The steam extraction and condensing configuration shows the lowest EUFP, since in this configuration, the DH output capacity is maintained the same without and with PCC. The only energy penalty, therefore, comes from the power output reduction due to steam extraction from the power cycle.

Again, the small variations of EP (and EUFP) under CO₂ capture rate from 90% to 99.72% shows the effect of CO₂ capture rate is relatively small when comparing the effect of different WtE operation configurations on the EP (and EUFP) results.
Figure 7-3 EP and EUFP of PCC with WtE plants under advanced heat integration with three capture rates.

Figure 7-4 shows the results of Coefficient of Performance of steam extraction ($COP_{X_{-cap}}$ and $COP_{X_{-cap+dh}}$) of the investigated WtE plants with PCC, for base cases and advanced heat integration cases. Under base case integration, the overall $COP_X$ values are in range of 5 to 5.2 with variations in different configurations and PCC capture rates. Heat recovery increases the $COP_{X_{-cap}}$ by around one for power-only configuration. With advanced heat recovery, the $COP_{X_{-cap+dh}}$ in the backpressure configuration increases by around 2.4 to 2.6 for CO$_2$ capture rate of 90% to 99.72%.

The highest increase of $COP_{X_{-cap+dh}}$ due to heat recovery occurs for the steam-extraction and condensing configuration. The $COP_{X_{-cap+dh}}$ values are in range of 7.7 to 8.6 for the three CO$_2$ capture rates. For this configuration, the DH output capacity is the same with and without PCC. Excess heat is recovered from the PCC process to provide DH, reducing the original steam extraction from the steam...
cycle, and more steam can be used for power generation. The result is higher power output from the low-pressure turbine and higher $COP_X$ for the same amount of total heat output.

![Coefficient of Performance for (steam) extraction of WtE plants with PCC under three CO₂ capture rates](image)

As introduced in chapter 3.1.3, the $COP_X$ in this study represents an analogous ‘consumption’ of electricity (actually foregone production) to achieve heat production for heating in the reboiler. This metric can be analogously compared to the coefficient of performance ($COP_{hp}$) of a heat pump, which is essentially the ratio of heat output divided by the power consumption.

Traditional heat pump usually has a strong correlation for lower COP at higher temperature lifts. Assuming ambient temperature of 20°C, temperature lift of 115°C to produce heat sink temperature 135°C, the COP of traditional heat pump is approximately 2 (IEA 2023). In comparison, the $COP_X$ for advanced heat integration scenarios are approximately 3 - 4.5 times higher than $COP_{hp}$ of typical electricity-driven heat pump (if commercially available). This means that for all the investigated WtE
configurations with PCC, steam extraction is the relatively more efficient method to provide thermal heat for solvent regeneration in the reboiler and district heating.

Similar to the base case discussion, an analogous absorption type heat pump is used to assess the efficiency of thermal heat delivery for solvent regeneration by steam extraction. Table 7-1 shows that the $COP_{X_{\text{abs}}}$ of steam extraction is 3.3, 2.5 and 1.9 for 90%, 95% and 99.72% capture rates respectively for the steam extraction and condensing configuration. In this configuration, the DH output is unchanged with PCC. The only driven heat source comes from the power reduction from the steam cycle and varies at different $CO_2$ capture rates.

The $COP_{X_{\text{abs}}}$ for the backpressure configuration is around one for the three $CO_2$ capture rates. In this configuration, the LP turbine is bypassed, and there is little variation in terms of power output from the HP and IP turbines. Heat recovery from the PCC process leads to minimal change in DH heat due to the use of PCC. With the similar high temperature heat requirement for solvent regeneration, the driven heat source for the absorption heat pump maintains the same capacity, thus the $COP_{X_{\text{abs}}}$ shows little variation for the three $CO_2$ capture rates. By this comparison, it can be concluded that the steam extraction from steam cycle is relatively more efficient for solvent regeneration for the assumptions made in this study, especially for the configurations that lead to less DH output reduction.

Table 7-1 $COP_{X_{\text{abs}}}$ of steam extraction for WtE-CHP plants with PCC under advanced heat integrations

<table>
<thead>
<tr>
<th>$COP_{X_{\text{abs}}}$</th>
<th>WtE-CHP (Ex&amp;C) – advanced heat integration</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>90% PCC</td>
</tr>
<tr>
<td></td>
<td>3.335</td>
</tr>
<tr>
<td>WtE-CHP (BP) – advanced heat integration</td>
<td></td>
</tr>
<tr>
<td>90% PCC</td>
<td>95% PCC</td>
</tr>
<tr>
<td>1.007</td>
<td>1.007</td>
</tr>
</tbody>
</table>

In this study, two CHP operation scenarios are included (steam extraction and condensing configuration, and backpressure configuration). They represent the most common DH supply approach for a WtE-CHP plants. Comparing the two configurations, the steam extraction and condensing configuration has the advantage of flexible DH output capacity, which is achieved by adjusted steam extraction from the cross-over pipeline between IP and LP turbine.
There is inevitably DH output reduction from the WtE plant when CO₂ capture is used, which leads to various energy penalties as discussed in this chapter. Figure 7-5 gives a direct comparison of the power and heat output from the WtE plants, under base case and advanced heat integration case, for three CO₂ capture rates. It can be seen that under the base cases, increasing DH output reduces the power output from the plant. For the investigated WtE plants, the maximum DH output capacity are 10.47MW, 11.95MW and 13MW for 90%, 95% and 99.72% CO₂ capture rates. With advanced heat recovery, the DH output can be maintained at around 6MW without power output reduction. When the DH output increases to higher than 6 MW, the power output reduces gradually. This is mainly due to the increased power consumption of the heat pump, which is used to recover excess heat from the DCC.

Figure 7-5 Power and heat output of WtE-CHP plant with PCC, under base case and advanced heat integration cases, and three CO₂ capture rates
7.2.2 Economic performance under advanced heat integration

Following the cost methodology described in Chapter 5, the cost value calculated for the advanced heat recovery integration is shown in Table 7-2.

Table 7-2 Cost estimation for advanced heat integration cases of WtE plants with PCC

<table>
<thead>
<tr>
<th>NO.</th>
<th>Capex Cost items</th>
<th>Value (in million £)</th>
<th>NO.</th>
<th>Opex Cost items</th>
<th>Value (in million £)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.1</td>
<td>WtE Power-only plant</td>
<td>188</td>
<td>2.1+2.6</td>
<td>WtE Power-only plant</td>
<td>5.8</td>
</tr>
<tr>
<td>1.2</td>
<td>WtE CHP plant</td>
<td>228</td>
<td>2.2</td>
<td>WtE CHP plant</td>
<td>6.11</td>
</tr>
<tr>
<td>1.3</td>
<td>PCC plant @90% CCR</td>
<td>33.8</td>
<td>2.3</td>
<td>PCC plant @90% CCR</td>
<td>1.51</td>
</tr>
<tr>
<td>1.4</td>
<td>PCC plant @95% CCR</td>
<td>35.2</td>
<td>2.4</td>
<td>PCC plant @95% CCR</td>
<td>1.55</td>
</tr>
<tr>
<td>1.5</td>
<td>PCC plant @99.72% CCR</td>
<td>38.3</td>
<td>2.5</td>
<td>PCC plant @99.72% CCR</td>
<td>1.64</td>
</tr>
<tr>
<td>1.7, 1.8, 1.9</td>
<td>0.59, 0.57, 0.55 for 90%, 95% and 99.72% capture rates</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.10,1.11, 1.12</td>
<td>6.55, 5.79, 4.68 for WtE CHP EX&amp;C with 90%, 95% and 99.72% capture rates</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.13, 1.14, 1.15</td>
<td>9.6, 9.72, 9.9 for WtE CHP BP with 90%, 95% and 99.72% capture rates</td>
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</tbody>
</table>

Figure 7-6 shows the results of LCOE for the investigated WtE plant with and without PCC for the investigated base case and advanced heat integration cases. As with the base case analysis in the previous chapter, for all cases a CO₂ emission price of £40/tCO₂ is assumed and this cost applies to only fossil CO₂ emitted from the WtE plants.
Figure 7-6 Breakdown of LCOE of WtE plant with and without PCC under the investigated configurations and three CO₂ capture rates, with assumptions of fossil CO₂ price £40/tCO₂, gate fee £100/tMSW, no negative emission credit is considered.

As can be seen in Figure 7-6, PCC increases the LCOE of the same WtE plant from £85/MWh to around £160-£180/MWh. Heat recovery from the stripper overhead condenser for boiler feed water heating reduces the LCOE by around £8/MWh. The marginal increase of LCOE when increasing the CO₂ capture rate from 90% to 95% and from 95% to 99.72% once more shows that ultra-high CO₂ capture can be achieved at an LCOE of WtE plant that is similar to 90% CO₂ capture.

In the above cases, for WtE plant with and without PCC, it is assumed that the plants receive the constant amount of gate fee of £100/t MSW, so that the same revenue received under the same amount of waste treatment. The variations of LCOE on the gate fee share is mainly due to the variations of power output among different scenarios.
To illustrate the absolute value of cost, the NPV under each integration scenarios is shown in Figure 7-7. The sum of Capex and Opex of PCC plant takes approximately 21% of the total cost of the WtE plant, with marginal increases with CO$_2$ capture rates. In power-only configurations, heat recovery only happens from the stripper overhead condenser for boiler feed water heating, the additional cost associated with the heat exchanger network is minimum comparing the overall cost.

Figure 7-7 Absolute values of NPV of WtE plant with PCC under base cases and advanced heat integration cases, and three CO$_2$ capture rates, with electricity selling prices assigned according with the corresponding LCOEs under each scenario, CO$_2$ price £40/tCO$_2$.

It should also be noted that in the above analysis, the CO$_2$ price is calculated at a relatively conservative price of £40/tCO$_2$. When the CO$_2$ price increases, the difference of LCOE with and without PCC will be reduced, as shown in Figure 7-8. For the advanced heat integration scenario, the CO$_2$ price should be around £240/tCO$_2$ to enable the LCOE of the WtE plant with PCC to break even with a similar plant. The LCOE at this breakeven CO$_2$ price is around £160/MWh to £170/MWh.
Figure 7-8 LCOE of WtE-Power only plant under different CO\textsubscript{2} prices

Figure 7-9 shows the LCOH for the two types of WtE-CHP plants investigated in this study, with and without PCC, for base cases and advanced heat integration cases. For all the cases, it is assumed that the electricity price is £85/MWh, which is the value of LCOE for power-only WtE plant without PCC. It is also assumed that fossil CO\textsubscript{2} emissions have a CO\textsubscript{2} price of £40/tCO\textsubscript{2}. As can be seen from Figure 7-9, the lowest LCOH occurs for the WtE-CHP-BP configuration without PCC, with a LCOH of £67/MWh. This is largely due to the highest DH output of 35 MW for this configuration.

When the CHP plants use PCC, the benefit of high DH output plus the maximum heat recovery from the PCC plant leads to the CHP-BP with PCC having much lower LCOH than the EX&C configuration. The LCOH for the CHP-BP case is in range of £104-£111 per MWh DH supply, which is much lower than the WtE-CHP plant without PCC under steam extraction and condensing configuration. Advanced heat recovery for the CHP-EX&C with PCC cases also reduce the LCOH from £236/MWh-£303/MWh (equivalent base cases) to £234-£294/MWh (advanced heat integration cases), a much small reduction than that under BP configuration with PCC, due to the unchanged DH output capacity. This implies
that advanced heat integration can produce better economic effect under scenarios that have higher DH output capacity, since these scenarios fully utilize the benefit of the excess heat that is available from the process.

Figure 7-9 Break down of LCOH of WtE plant with PCC under base cases and advanced heat integration cases, and three CO₂ capture rates, with assumption of electricity selling price £85/MWh, fossil CO₂ price £40/t CO₂.

To illustrate the absolute value of cost, the NPV under each integration scenarios for WtE-CHP plants is shown in Figure 7-10. It can be seen that for the PCC plant at 95% CO₂ Capture rate, the sum of Capex and Opex of a PCC plant takes approximately 17% of the overall Capex and Opex of a WtE-CHP plant. Without PCC, the cost of fossil CO₂ emission takes 7.3% of the overall cost of a WtE-CHP plant. CO₂ capture helps to eliminate the CO₂ emission cost to minimum. For the CHP plant in EX&C configurations, the aim is to maintain DH output, so that the advanced heat recovery increases the
power output, as seen in the figure a 20% increase in power output; for CHP-CP configurations, the aim is to maximize the DH output when PCC is on, so that it can be seen 36% and 55% increase in the NPV of DH heat under base case heat integration and advanced heat recovery, respectively.

Figure 7-10 Absolute values of NPV of WtE-CHP plants with PCC and without PCC under base cases and advanced heat integration cases, with 95% CO₂ capture rate; electricity selling prices assigned £85/MWh, the district heat selling price according with the corresponding LCOHs under the integrated scenario, CO₂ price £40/tCO₂.

Figure 7-11 presents the cost of CO₂ captured for the investigated WtE plant configurations integrating with PCC. As with the base case integration, it is assumed that the fossil CO₂ emission price is £40/tCO₂ and the plant utilization factor of 87% gives an annual 7650 working hours for electricity generation. For WtE-CHP plants, the DH supply period is defined 50% of the utilization period, which can represent about annually 6 months DH supply for a typical northern country. When calculating the cost of CO₂
captured for CHP plant with PCC, the electricity-selling price is estimated to be £85/MWh, a value of the referenced power-only WtE plant without PCC.

Results shows that the effect of advanced heat integration on power only plant is limited, the cost of CO₂ captured for power only WtE plant is reduced to £41-£43/tCO₂ (from £43-£44/tCO₂). For CHP plant the effect is more significant, particularly for the CHP-BP case. As shown in Figure 7-11, the cost of CO₂ captured is reduced from £144-148/tCO₂ to £96-100/tCO₂ by using advanced heat integration.

It is important to note that these results are significant to the key assumptions provided above. For instance, it is found that if the DH supply period increases to 60% of the plant utilization period (with other assumptions unchanged), the cost of CO₂ captured of the WtE-CHP-BP with PCC plant would be comparable with that of the power-only plant.

7.2.3 Environmental performance under advanced heat integration

In this study, direct CO₂ emission intensity on three main products bases are defined. They are direct CO₂ emission intensity on fuel basis, on electricity basis and on heat basis. The direct fossil CO₂
emission on fuel basis is displayed in chapter 6.4.3 and are not changed by advanced heat integration. This section presents the direct CO₂ emission for advanced heat integration, and its comparison with the base case integration, which can be seen in Figure 7-12.

As mentioned in Chapter 3.3, this study uses a fixed ratio of heat and power ratio of 2 to allocate CO₂ emissions from CHP plants to electrical or heat output, as used in the 1/3: 2/3 Method (DUKES). In general, implementing CO₂ capture combined with permanent CO₂ storage to a WtE plant leads to net negative CO₂ emissions on both electricity and heat basis. Advanced heat integration reduces the CO₂ removals per unit energy output due to relatively higher EUF. For example, the CO₂ capture removals per MWh of exported electricity are reduced from 611 - 938 kgCO₂/MWh for the CHP plant with base case integration to 434 - 837 kgCO₂/MWh of CHP plant with advanced heat integration. Similarly, the
CO₂ capture removals per MWh of DH production are reduced from 306 - 469 kgCO₂/MWh in the base case to 217-419 kgCO₂/MWh with advanced heat integration. The variations in CO₂ removal are caused by different CO₂ capture rates and CHP configurations. This implies that care is needed when comparing CO₂ emissions from WtE plants with other CO₂ emitters. The absolute value of CO₂ removals is the same for cases where fuel input and CO₂ capture rate are the same, but the specific CO₂ removal value will vary depending on different key product basis, and the removal potentials per unit are lower for higher efficiency integrations.

7.2.4 Summary of KPI results for the advanced heat integration

This chapter presents the results of the advanced thermal integration option of WtE plants with PCC. Excess heat from three locations in the PCC process: the compressor intercoolers, the stripper overhead condenser and the DCC is recovered to maximize the net power output for a power-only WtE plant and maximize thermal output for a WtE-CHP plant equipped with PCC. The same set of KPIs is used as that applied for the base case integration. For the purpose of comparison, the KPI results for both base case integrations and the advanced integrations are presented together, to help understand the effect of heat recovery on the overall performance from thermodynamic, economic and environmental points of view.

Of all the investigated WtE plants integrating with PCC, the CHP plant with BP operation benefits the most in terms of heat recovery from the PCC process. For the base case integration, it has the highest EUP (~1091 kWh/tCO₂) and EUFP (~35%). For instance, at 95% CCR, and with advanced heat recovery, all the excess heat from the PCC process is recovered, and the excess heat recovered by heat pump, the EUP and EUFP effectively reduces to 248 kWh/tCO₂ and 8%, respectively.

Economic analysis finds that although additional investment is needed to establish a heat exchange network to implement heat recovery for the advanced heat integration cases, the benefit in terms of higher energy output (power and heat) outweighs its drawbacks. For the advanced heat recovery cases, the LCOE is about £8/MWh lower than for the base case integrations for the EX&C integrations. This effect is more significant for CHP plants with BP configuration, where the LCOH is about £70-£90/MWh lower than that under base case integration. The economic results are also largely related to the defined energy prices, DH periods, etc. For example, when the DH supply period is 60% of the plant utilization period for the economic assumptions used in this analysis, the cost of CO₂ captured of the WtE CHP-BP with PCC and advanced heat integration could be comparable with that for power-only plant with PCC.

Environmental KPI results show that advanced heat integration reduced the CO₂ removals per unit energy output due to relatively higher energy utilization factor. This shows that careful attention
should be given to choosing insightful performance indicators when comparing CO₂ emissions from WtE plants with other CO₂ emitters.

The overall KPI results demonstrate that the assessment of WtE with CO₂ capture technology should apply a systematic approach from multiple aspects to give a comprehensive view. Different predefined parameters, operation strategy of the plant, energy prices, and CO₂ prices will affect the results of the analysis, which are usually the fundamental basis for decision-making.
8 Interim solvent storage to meet fluctuating DH demand profiles

The previous chapter focussed on the application of advanced heat integration to improve the performance of CHP plant with PCC. However, it was based on the assumption of constant DH supply under each integration scenario. In real operation of CHP plant, it is highly possible that DH demand is fluctuating, both in terms of daily and seasonal variation. In this chapter, in response to the possible fluctuating DH demand conditions and to improve the operation flexibility of the abated WtE-CHP plant integrated with PCC systems, an operational strategy of PCC plant that deploys seasonal interim solvent storage (ISS) is introduced.

In order to explore this application, an illustrate case of WtE-CHP plant with PCC under ISS is modelled. The steam cycle configuration with extraction and condensing type configuration represents the most common configuration of CHP plant with fluctuating DH supply capacity, so the WtE-CHP plant in EX&C configuration as previously illustrated in section 5.1.2 is used. Previous analysis shows that with advanced heat integration, the influence of ultra-high CO$_2$ capture rate on system performance can be comparable to that under base integration, so the PCC plant with 35% MEA based post-combustion at 99.72% CO$_2$ capture rate as modelled in section 5.2 is selected as the PCC plant for the WtE-CHP plant. A UK based heat load profile is selected representing the fluctuating DH demand for this particular case study. Following this, an operation strategy of WtE-CHP plant PCC under ISS application is defined, taking consideration of the operational constraints from the steam cycle, capture plant and the compression plant.

The effectiveness and the usefulness of the proposed ISS application are demonstrated with sensitivity analyses on the size of storage tanks, gas boiler heat prices and fossil CO$_2$ prices. The KPIs introduced in Chapter 4 are applied to give an overall comparison of this application with the previous integration cases. This provides a high-level evidence to assess the use of ISS to maintain benefits of the operating plant and at the same time achieve ambitious negative CO$_2$ reductions.

8.1 Concept of interim solvent storage (ISS)

The concept of ISS was first introduced in the work presented by Chalmers et al. (Chalmers & Gibbins, 2007; Chalmers, Leach, Lucquiaud, & Gibbins, 2009). In their work, the ISS was applied to power-only plants and indicated that higher profits could be obtained by boosting the power output of the plant when electricity prices are high by storing rich solvent, thus delaying the energy-consuming step of solvent regeneration to a later point in time.

With the expectations that CO$_2$ capture solutions for WtE-CHP applications are comparable to the ones considered for power plants, and heat supply usually takes priority over power supply for CHP plants, our previous work presented in (Su et al., 2022) uses a mathematical programming-based
methodology to investigate the optimal operation of a WtE-CHP plant with the implementation of ISS. In that study, a time period of 24 hours is used with an hourly resolution with the objective of maximizing annual profit for the WtE plant operators. Results show the benefits of ISS in terms of energy utilization efficiency and CO₂ emission reduction; however, with the estimated boiler heat purchasing price of £130/MWh, the annual revenue of the WtE plant with ISS is reduced by £0.9M/year due to the increased capital costs associated with ISS application. It is observed that solvent storage is not operational during the summer design day suggesting that seasonal solvent or thermal storage may be a more useful alternative.

In this context, this chapter presents a study of seasonal solvent storage for the investigated WtE-CHP plant in steam extraction and condensing configuration, with 99.72% CO₂ capture rate. In particular, it examines the benefit and challenges under this route of solvent storage using the KPIs identified in Chapter 4, with sensitivity analysis of key design inputs.

8.2 Fluctuating DH demand profiles

In order to illustrate the fluctuation of DH demand, the DH load patterns are plotted using data from a UK based study by Broklebank et al. to estimate the heat demand variation of a CHP system within

![Diagram](Image)

Figure 8-1 Illustrative description of a WtE-CHP plant equipped with Interim Solvent Storage in the PCC system (Su et al., 2022)
24 hours in one day (Broclebank, Beck, & Styring, 2018). The original DH demand pattern is extrapolated by factor of 2.28 to match the Spring/Autumn DH capacity of the existing medium-sized WtE-CHP plant that has been modelled in this study. The factor of 2.28 is determined at the point when the maximum DH demand during Spring/Autumn is equal to the maximum DH supply capacity of the WtE-CHP plant with ISS based PCC. The seasonal duration of a year is referenced from the National Physical Laboratory, who adopts the astronomical events to define seasons. The season duration for 2023 is shown in the Table 8-1 below:

Table 8-1 Seasonal duration based on astronomical events (NPL, 2023)

<table>
<thead>
<tr>
<th>Seasons</th>
<th>Starts/Ends</th>
</tr>
</thead>
<tbody>
<tr>
<td>Spring</td>
<td>20th March / 20th June</td>
</tr>
<tr>
<td>Summer</td>
<td>21st June / 22nd September</td>
</tr>
<tr>
<td>Autumn</td>
<td>23rd September / 21st December</td>
</tr>
<tr>
<td>Winter</td>
<td>22nd December / 19th March</td>
</tr>
</tbody>
</table>

Summer and winter are seasons of low and high DH demand. In order to illustrate this variation, multipliers of the Spring/Autumn demand are applied for summer and winter, with multiplier values of 50% and 165% respectively. The multiplier values are referenced from a study by Gadd & Werner, who illustrated the aggregated average hourly heat load for four heating seasons in district heating systems (Gadd & Werner, 2013). Based on this approach, the seasonal DH demand pattern considered in this study is shown in Figure 8-2. The daily DH demand pattern is assumed constant for each day in the whole season.

It should be noted here that in this study, the high-level DH demand pattern plotted is deemed to be adequate to give the initial understanding of the implications of ISS. In real operation, the DH demand predictions and its patterns are complex and are usually varies by the locations, types of end users, urban population densities, functions of buildings, etc. A more accurate DH demand variation model may improve the quality of the result of ISS application, which can be improved in the future work.
8.3 Operation constraints and logic of ISS application

In order to specify this, some assumptions are determined to limit the number of variations in this application, such as constant specific reboiler duty (SRD), rich and lean loading of the stripper, etc. Besides, a K value is introduced to serve as a decision factor of the operation of Solvent storage or
Additional Regeneration. The K value is the ratio of rich solvent discharges into the stripper under ISS applications to that under the base case. The calculation of K value is shown in Equation 8.1:

\[
K = \frac{\text{mass}_{\text{rich solvent into stripper, ISS}}}{\text{mass}_{\text{rich solvent into stripper, base}}} = \frac{Q_{\text{steam, ISS}}}{Q_{\text{steam, base}}}
\]

Equation 8.1

Where:

The subscript ISS represents values under ISS operation.

The subscript base case represents the case that WtE-CHP plant integrating with PCC system providing the maximum DH supply, with \(Q_{\text{dh,base case}} = 10.5 MW\). This corresponds to the WtE-CHP-Ex&C under 99.72% CO₂ capture rate that modelled and presented in the Chapter 6.1.

• Constraints on K value

As noted previously in this chapter, this \(K\) equation assumes a constant rich and lean loading of the stripper is maintained as well as Specific Reboiler Duty (SRD), under the base case/SS/AR cases. With this \(K\) value defined, during the operation of ISS, the DH heat output with a range a \(K\) values is calculated using the existing WtE-CHP plant model in gProcess, as shown in the Figure 8-3.

![Figure 8-3 DH heat output during the operation of ISS, with assumption that maximum steam is always extracted from the steam cycle to provide heat for DH and for solvent regeneration](image)

Figure 8-3 DH heat output during the operation of ISS, with assumption that maximum steam is always extracted from the steam cycle to provide heat for DH and for solvent regeneration

Similarly, the net power output during the operation of ISS under a range of \(K\) values is calculated, with the assumption the heat production takes priority over power production, as shown in the Figure 8-4. The Net power output in the figure considers the gross power output from the steam cycle, the
power consumptions from the PCC plants, including the booster fan, compression power, solvent pumps.

![Graph showing the relationship between K value and power output. The equation is given as y = -1.3702x + 8.9837.](image)

Figure 8-4 Net power output of WtE-CHP plant with PCC in ISS application, with assumptions that maximum steam is extracted from the steam cycle for DH and for solvent regeneration

The addition of ISS needs to consider the operation constraints of both the power system and the PCC and compression system, which are important for the application of ISS under the fluctuating DH demand. Theoretically, these constraints determine the variable steam extracted from the steam cycle, thus the operation of rich and lean solvent tanks, or in other words, the upper and lower limit of $K$ values. The key constraints considered in this study are from the low-pressure turbine, from the air-cooling and from the stripper. The first two constraints together determine the maximum steam could be used for solvent regeneration, thus the upper limit of the $K$ value; the stripper is oversized to regenerate the maximum amount of rich solvent, in the same time the minimum solvent stored in the stripper is determined to ensure performance, thus the lower end of the $K$ value.

- **Constraint on low-pressure turbine**

As previously noted in chapter 4.1.2, a minimum 15% design flow into the LP turbine (in power only configuration) is set to constrain the maximum stream that can be extracted for solvent regeneration & district heating. The same constraint is also applied to the ISS application.

- **Constraint on air-cooling**

In Europe, WtE facilities are often equipped with air-cooled condensers (air-cooler), which are necessary when plant location and the ambient conditions do not allow an easy and economic use of
other cooling systems (TOBIASEN & KAMUK, 2012). The integration of CO₂ capture (and ISS) has impact on the performance of the cooling system. When CO₂ capture is on, steam extraction from the IP-LP crossover pipeline reduces the steam flow through the LP turbine, thus the cooling duty of the air-cooler. In this study, the initial temperature difference (ITD), which is defined as the difference between the steam condensation and ambient air temperatures, is used to constrain the operation of air-cooler under the reduced cooling duty. Typical design choices of ITD are in range of 15°C-30°C (ACCUG, 2021), and it is a trade-off between larger power output and increased larger capital cost unit. Compared with wet cooling, the ITD of air-cooling seldom goes lower than 5.5°C (Spencer, n.d.). In this study, a minimum ITD of 5.5°C is used to define the situation of maximum steam extraction, which corresponds to approximately 15% of the design steam flow under WtE-power only configuration without PCC.

Under situations of maximum steam extraction from the steam cycle, when DH output is zero, the steam available for solvent regeneration becomes the maximum. By using the previous equation in Figure 8-4, this corresponds to upper limit of K value of 155%. This means the 155% rich solvent comparing base case can be regenerated with the steam available from the steam cycle.

- Stripper oversizing under ISS

The application of ISS implies a large variation of counter-current solvent flow. In the presence of large vapour flow-rate and the pressure drop in the stripping column, flooding may occur which reduces mass transfer efficiency of the stripper and poses other adverse consequences, such as pushing liquid into the gas pipes and then damaging the equipment impacts (Li et al., 2021). Prevention of flooding is important for effectively sizing the stripper for the application of ISS. As previously introduced in Section 4.2.2, a safety factor, flooding point is applied to ensure the performance of the stripper column.

In this context, in this study, the stripper is designed at 70% flooding velocity, with the maximum operational superficial gas velocity at 80% of the flooding velocity. At the same time, as shown in Figure 8-5, in ASPEN, the diameter of the stripper is oversized by 16.7% to meet the maximum regeneration capacity of 155% flow capacity under the base case. A minimum 35% design flow rate is set to constrain the minimum flow into the stripper for solvent regeneration. The minimum flow value of the stripper (30% of base case flowrate) is based on engineering judgement and the review of publications, this minimum value is to ensure a sufficient fluidization of bed materials in the stripper.
The 35% design flow for the oversized stripper corresponds to 48% flow under the base case.

Following the operation constraint for the ISS case described above, the K value is constrained to the range of 48%-155%. The K value determines the amount of heat used for solvent regeneration and is in turn constrained by the DH demand of the end users.

The maximum 155% flow in the oversized stripper is to ensure maximized regeneration capacity of the capture plant. When oversizing the stripper, an approximate column diameter is first applied and tested until the flooding point reaches to maximum flooding point (no higher than that), to ensure efficient stripping in the column.

• Operation logic for ISS

In this study, it is assumed that both the WtE-CHP plant and PCC plant are operating at full capacity for the whole year and full flue gas stream is being sent to the absorber (with 99.72% CO₂ capture rate). The operation of the ISS starts from 23rd September, the first day of autumn, and ends 21st
September, the last day of the summer. The operation of ISS includes three modes: ISS-Solvent storage (ISS-SS mode) and ISS-Additional Regeneration (ISS-AR mode) and w/o ISS mode.

The $Q_{dh_{base\ case}}$ is used as the first comparing criteria to determine whether the system should operate in ISS-SS mode or ISS-AR mode. The condition of tanks (empty or full) is used to determine whether the intended operation can proceed.

The ISS-Solvent storage (ISS-SS mode) refers to the operating mode where $Q_{dh_{demand}} > Q_{dh_{base\ case}}$, Part of the rich solvent ($K \times m_{rich\ base}$) from the absorber is sent to the stripper for solvent regeneration, with the remaining rich solvent sent to the rich solvent tank. This reduces the amount of steam required for solvent regeneration, thus allowing more steam to be extracted for district heating to provide the high DH demand. The solvent from the lean storage tank will be added to the lean solvent flow that leaves the regenerator, so that the absorber maintains the constant L/G ratio. This process will continue until the solvent tank is full. This mode is illustrated as mode a in the Figure 8-6.

Similarly, the ISS-Additional Regeneration (ISS-AR mode) refers to the operating mode when the DH demand reduces, $Q_{dh_{demand}} < Q_{dh_{base\ case}}$, the rich solvent from both the absorber and the rich solvent tank (in total $K \times m_{rich\ base}$) will be sent into the stripper, so that the stripper regenerates additional solvent that was previously stored in the rich solvent tank. Part of the lean solvent leaving the stripper is sent to the lean solvent tank, so that the amount of lean solvent into the absorber and the L/G ratio is constant. This process continue until the rich solvent tank is empty and is illustrated as mode b in the Figure 8-6.

During the operation of ISS-SS mode and ISS-AR mode, the two solvent storage tanks (rich and lean storage tank) act as buffers between the capture and the regeneration process. Thus enables the WtE plant to better match the fluctuating DH demand. The $K$ value is used as the primary indicator to decide the operation modes of the ISS. The condition of the solvent tanks will be used to determine whether the operation mode can proceed. During the situation that the storage tanks are full or empty, a third operation mode will be determined, which will be described in the following section.

The w/o ISS mode represents a mode where solvent storage tanks are not being filled or emptied. This corresponds to situations 1) when the $Q_{demand}$ is below the $Q_{dh_{base\ case}}$, the ISS system intends to regenerate additional solvent previously stored in the rich solvent tank, but the tank is empty thus unable to perform additional regeneration. In this situation, the WtE provides the low $Q_{demand}$ by extracting less steam from the steam cycle, and more power is produced; 2) when the $Q_{demand}$ is higher than the $Q_{dh_{base\ case}}$, the ISS system intends to do solvent storage, however the rich tank is full, and is thus unable to perform the solvent storage. In this situation, the WtE-CHP plant provides the
$Q_{dh\text{ base case}}$ to the DH system, with the rest of the DH demand met by gas boilers to fulfil the total DH demand.

With the three operation modes mentioned above, the detailed decision logic is as shown in Figure 8-7.
Figure 8-6 Illustrative operation modes of WtE-CHP plant with PCC, ISS-SS mode, ISS-AR mode and w/o ISS mode.

Assumptions:
- A constant rich and lean loading of the stripper is maintained as well as Specific Reboiler Duty (SRD), under the base case/SS/AR cases;
- The base case scenario is defined as the case that WtE-CHP plant integrating with PCC system providing the maximum DH supply capacity, which is approximately 10.5MW;
- The mass flow of rich solvent into the stripper is in proportional to the heat provided to the stripper:

\[ K = \frac{\text{mass rich solvent into stripper}_{\text{ISS}}}{\text{mass rich solvent into stripper}_{\text{base}}} = \frac{Q_{\text{steam ISS}}}{Q_{\text{steam base}}} \]
Figure 8-7 Operation logic of WtE-CHP plant integration with PCC under ISS applications
8.4 Evaluation of the annual profit of a WtE-CHP plant with PCC and ISS case

In this study, it is assumed that the WtE-CHP plant operators are responsible for the costs related to the WtE plant and CO$_2$ capture plant. They receive revenue from gate fees for waste treatment, heat sales and power sales. Additionally, the operators have contractual DH supply responsibility to the end-users. In situations where DH is unable to fulfil all end-user demand, it is assumed that the WtE-CHP plant heat will be supplemented by gas boilers. In this context, the annual profit of the WtE-CHP plant equals the annual revenue minus the annual cost of the WtE plant, as shown in Equation 8.2 and Equation 8.3.

Annual $\text{profit}_{\text{ref}} = \text{Revenue} - \text{COST}$

\[
= \text{Gatefee} \times \frac{\text{MSW}}{\text{year}} + (Q_{\text{dh,wte,ref}} \\
+ Q_{\text{dh,boiler,ref}}) \times \text{Heat}_{\text{price}} + Q_{\text{dh,wte,ref}} \times \text{Heat}_{\text{tariff}} \\
+ W_{\text{wte,ref}} \times \text{Electricity}_{\text{price}} - \text{Annualized Capex}_{\text{ref}} \\
- Q_{\text{dh,boiler,ref}} \times \text{Heat}_{\text{price,gasboiler}} - \text{Opex}_{\text{ref}}
\]

Equation 8.2

Annual $\text{profit}_{\text{ISS}} = \text{Revenue} - \text{COST}$

\[
= \text{Gatefee} \times \frac{\text{MSW}}{\text{year}} + (Q_{\text{dh,wte,iss}} \\
+ Q_{\text{dh,boiler,iss}}) \times \text{Heat}_{\text{price}} + Q_{\text{dh,wte,iss}} \times \text{Heat}_{\text{tariff}} \\
+ W_{\text{wte,iss}} \times \text{Electricity}_{\text{price}} - \text{Annualized Capex}_{\text{iss}} \\
- Q_{\text{dh,boiler,iss}} \times \text{Heat}_{\text{price,gasboiler}} - \text{Opex}_{\text{iss}}
\]

Equation 8.3

Where:

$\text{Gatefee} \times \frac{\text{MSW}}{\text{year}}$ represents the annual revenue of the WtE plant by treating 500t/d MSW, with constant gate fee of £100/t MSW;

$Q_{\text{dh,WtE}}, Q_{\text{dh,boiler}}$ and $W_{\text{wet}}$ are the annual DH heat supply from WtE plant and from local gas boilers, annual electricity output from the WtE plant, respectively, in MWh;

$\text{Heat}_{\text{price}}$ is the heat selling price from the WtE plant. In the UK, the price of heat is often determined by the heat source, which can result in a wide range of costs and structures. In this study, an average heat sale price £79/MWh is considered (DECC, 2015);

$\text{Heat}_{\text{tariff}}$ is the WtE-CHP plants that incinerate MSW as fuel are eligible to receive funding under the Non-Domestic Renewable Heat Incentive Scheme (NDRHI) (ofgem, n.d.). NDRHI scheme is a government environmental programme, which is designed to increase the uptake of renewable heat...
that can help reduce carbon emissions and meet the UK’s renewable energy targets (ofgem, n.d.). A tariff rate of £46/MWh is considered based on the tariff table 2021/2022 (ofgem, n.d.).

*Electricity price* is the electricity selling price from the WtE plant. In this study, an electricity price of £85/MWh is assumed, this is the LCOE of the WtE plant under without PCC scenario as previously calculated in Chapter 6.

*Annualized Capex* is the Annualized Capex WtE plant with PCC, in million £/year; it is calculated based in order to ensure a constant repayment of the investment over the project duration. Based on the discount rate (8%) and the project economic duration (25 years), with the *Annualized Capex* is defined as 8.67% of the total capital requirement.

*Heat price gasboiler* is gas boiler heat purchasing price for the WtE plant, this is the counterfactual heat cost appropriate for heat from the WtE plant. It is suggested in range of £89/MWh to £126/MWh referencing from a report from DECC, and a medium value of £108/MWh is applied as the base case scenario in this study (DECC, 2015);

*Opex* is the sum of the Annual Opex of the WtE plant with PCC, including the Opex of the WtE plant, *Opex of CO₂ capture plant*, cost of heat from gas boilers, and also the cost of total fossil CO₂ emission from WtE plant with CO₂ price of £80/tonne fossil CO₂, in million £/year. The calculation of Opex please refer to Chapter 5.1. The economic calculation in previous chapter finds the share of Opex in the total cost is small and in the ISS application in this study, the amount of lean solvent into the absorber maintains unchanged so that the total solvent inventory can assumes to be the same. In this context, the additional Opex of the PCC plant due to adding ISS application are neglected.
Capex is the sum of the cost of the WtE plant and the CO₂ capture and compression facility. The calculation of Capex please refer to Chapter 5.2. The only add-on section in the Capex is under the ISS case, additional equipment cost of the rich and lean solvent tank should be included.

Cost of solvent storage tank (rich and lean solvent tank) will be a function of the total solvent flow rate, using a scaling parameter of 0.6 (IECM, 2019; Turton, 2018), as shown in the Equation below:

\[
\text{Capex}_\text{tank,ISS} = \left( \frac{V_{\text{solvent,ISS}}}{V_{\text{solvent,ref}}} \right)^{0.6} \times \text{Capex}_\text{tank,ref}
\]

Equation 8.4

The cost of the storage tanks references from the base case cost of the amine storage tank cost in the report commissioned by BEIS (BEIS, 2020b), as shown in the table below:

Table 8-2 Referenced cost of the amine storage tanks

<table>
<thead>
<tr>
<th>Amine tank volume (m³)</th>
<th>Amine tank cost (£)</th>
<th>Cost basis</th>
</tr>
</thead>
<tbody>
<tr>
<td>2,456</td>
<td>197,313</td>
<td>316L stainless steel tank cost only, no inventory costs included</td>
</tr>
</tbody>
</table>

The sign \( ref \) and ISS represents the reference case and ISS case, respectively.

In the above cost estimation, The Chemical Engineering Plant Cost Index (CEPCI) is used to adjust the cost of the reporting year to the cost of the reference year 2021.

\[
\text{COST}_{\text{year}-X} = \text{COST}_{\text{year}-Y} \times \frac{\text{CEPCI}_{\text{index},X}}{\text{CEPCI}_{\text{index},Y}}
\]

Equation 8.5

Table 8-3 Currency conversion applied under circumstance that different currencies used in the reporting data

<table>
<thead>
<tr>
<th>Annual CEPCI @2014</th>
<th>576.1 (Chemical Engineering, n.d.)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Annual CEPCI @2015</td>
<td>556.8 (Chemical Engineering, n.d.)</td>
</tr>
<tr>
<td>Annual CEPCI @2018</td>
<td>603.1 (Chemical Engineering, n.d.)</td>
</tr>
<tr>
<td>Annual CEPCI @2021</td>
<td>708.8 (Chemical Engineering, n.d.)</td>
</tr>
<tr>
<td>Exchange rate from € to £ in 2015</td>
<td>0.7263 (Exchangerates, n.d.)</td>
</tr>
</tbody>
</table>
Based on the above description, the annual profit of the WtE plant operators under w/ and w/o ISS application can be calculated and is used to compare the economic aspects of the two cases. Sensitivity analysis of the key variables will be discussed in the following chapter.

8.5 ISS application results and sensitivity analysis

The addition of the rich and lean solvent tanks enables the flexibility of the integrated system by acting as buffers between the absorber and desorber columns. The solvent is stored or discharged, depending on the fluctuating DH demand. The level of flexibility and effectiveness of this application largely depends on the input parameters, such as the size of storage tanks, boiler heat purchasing price, CO$_2$ price, cost of storage tanks, etc. A sensitivity analysis of the annual profit to the key parameters is discussed in this section.

A Sensitivity analysis of the cost of storage tanks finds that its effects on the overall profit is modest. The main increase of cost is due to the oversized compression train and the oversized stripper diameter to allow additional solvent regeneration capacity, which takes 53% and 44% respectively.

To specify, the cost of the scaled stripper and compressor is referencing the cost evaluation method from IEAGHG (IEAGHG 2017), using the Equation 8.5 and Equation 8.6 below:

$$\text{Cost}_{\text{stripper&others,ISS}} = \left(\frac{\text{CO}_2 \text{ flowrate to compression,ISS}}{\text{CO}_2 \text{ flowrate to compression,ref}}\right)^{0.9} \times \left\{\text{Cost}_{\text{stripper,ref}} \times \left(\frac{\text{Stripper diameter,ISS}}{\text{Stripper diameter,ref}}\right)^2\right\} + \text{Cost}_{\text{other items,ref}}$$

$$\text{Cost}_{\text{compression,ISS}} = \left(\frac{\text{CO}_2 \text{ flowrate to compression,ISS}}{\text{CO}_2 \text{ flowrate to compression,ref}}\right)^{0.75} \times \text{Cost}_{\text{compression,ref}} + 550K$$

In the above estimation, it may be observed that the cost of the stripper and compressor relates directly with the CO$_2$ flowrate into the equipment. As previously mentioned in section 8.3, the application of ISS requires approximately 155% increase of flue gas flow, thus the CO$_2$ flowrate into the system. Based on this, it can be anticipated that the ISS application will lead to a direct increase in the Capex of scaled stripper and compression facilities. As shown in the following Figure 8-8, the addition of ISS increases the capex of the PCC plant from £38.3m to £51.4m. The breakdown of the
cost elements shows that the additional cost from the rich and lean storage tanks constitutes only £0.42M, i.e. a 3% share of the additional cost.

Figure 8-8 Comparison of capital cost of PCC plant under scenarios with and without ISS applications

8.5.1 Sensitivity of gas-boiler heat prices

The sensitivity of the annual profit to the size of storage tanks and gas boiler heat purchasing prices is shown in Figure 8-9. In this analysis, considering the uncertainty in terms of whether the Negative emission credit (NEC) will be available for this application, scenarios with and without NEC are included, as seen in the upper and lower section of the figure. For the energy prices, gate fee and fossil CO₂ price assumed in this work, without NEC, the WtE-CHP-PCC plant gets negative profit £-2 million/year to £-0.4 million/year under the range of gas-boiler heat prices and the range of storage tanks sizes. With the NEC of £75/tCO₂ allocated to the captured biogenic CO₂, the WtE-CHP-PCC plant gains annual profit in range of £4 million/year to £6 million/year under the same set of energy prices.

Additionally, there is an optimized size in terms of the size of the storage tank. Under both scenarios with and without NEC, and under the range of gas-boiler heat prices, increasing the size of the storage tanks from 0 to 2 hours increases the annual profit and reaches to the maximum annual profit at the size of 2 hours, i.e. approximately 494m³. Increasing the solvent size for a storage time above than 2
hours does not a further positive effect on the annual profit, but still shows annual profit higher than cases without ISS (storage size of 0 hour).

Figure 8-9\(^4\) Sensitivity of the annual profit to the solvent storage tanks size for a range of gas boiler heat purchasing prices, under scenarios w/ and w/o Negative emission credit of £75/tCO\(_2\). Electricity and heat selling prices from the WtE plant are £85/MWh and £79/MWh, respectively; heat tariff from WtE plant £46/MWh; fossil CO\(_2\) price £75/tCO\(_2\); gate fee £100/tMSW; 1 hour of storage requires the tank capacity of 247m\(^3\).

Figure 8-9 also indicates that with higher gas boiler heat purchasing prices, the benefit of ISS is more significant, this can be observed on the increases of the slope of the curves from zero to two hours storage time. For example, under scenarios with NEC and gas-boiler heat price of £89/MWh, the

\(^4\) The profit without ISS is under the cases with capacity of storage tanks equal to zero (the very left points on Figure 8.9).
annual profit of the WtE-CHP-PCC plant increases from £5.77 million/year to £5.95 million/year (two hours storage tank), an increase of £0.18 million/year. When the gas-boiler heat prices increase to £126/MWh, the annual profit of the WtE-CHP-PCC plant increases from £4.36 million/year to £4.77 million/year (two hours storage tank), an increase of £0.4 million/year.

In Figure 8-10, the initial increase of profit is mainly due to the storage capacity of the tanks, which increases the energy output of the plant; the decrease of profit is due to the size of the plant doesn’t cope with the energy demand curve and capacity (as suspected by the author). This suggests that there is an optimized size of the storage tanks. This curve only explains the situation for this study, e.g. the DH demand curve, the limitations set to constrain the operation of stripper, steam cycle, etc.

8.5.2 Sensitivity of storage tank sizes

Figure 8-11 shows a more detailed comparison of the heat supply by both the WtE-CHP plant and the gas boiler and the amount of solvent stored for the three storage tank sizes equivalent to 1 hour, 2 hours and 3 hours storage times, for a 24-hour period during Spring season. Taking as example the 21st of March, Figure 8-11 shows that for 1 hour storage tank size, there is a sharp increase of the gas boiler heat supply at 8:00am, and the supply is maintained until the heat demand decreases at around 8:00pm. When the solvent storage tank is sized for 2 hours and 3 hours, gas boiler heat supply is not required until 12:00 pm, since the heat demand can be fulfilled by the WtE-CHP-PCC plant. Although the tank volumes for 3 hours storage is higher than for 2 hours storage, in both scenarios, the tanks are full starting from 11:00 am, thus limiting the effectiveness the solvent storage, and leading to similar annual profits. This is an example of a tank operation for a representative season day. For longer operations, for instance, seasonal application of ISS, the size of storage tanks could be design in larger volume (longer hours), and more comprehensive understanding of the system sensitivities should be addressed, including future DH demand curve/operational modes etc., to optimize the design and make it more economically beneficial.
Figure 8-11 Comparison of heat supply and solvent levels in tanks in Spring Day 21st March, under three solvent tank sizes - 1 hour storage, 2 hours storage, 3 hours storage. The solvent in lean tank operate in opposite trend as that shown in the rich tank, to maintain mass balance in the system.

8.5.3 Sensitivity of CO₂ prices

A sensitivity analysis on the CO₂ price has also been undertaken. In this analysis, it is assumed that both the fossil CO₂ emitted and the biogenic CO₂ captured have the same CO₂ price. That is, the plant needs to pay for the fossil CO₂ emitted, and at the same time, will get benefit for biogenic CO₂ captured (e.g. in form of negative emission credit). As can be seen from Figure 8-12 for the WtE-CHP plant without PCC, the plant gets annual revenue of £9.7 million to £3.2 million under CO₂ prices from zero to £100/tCO₂, under the energy prices shown at the bottom of the figure. Adding PCC reduces the annual revenue at low CO₂ prices. A break-even CO₂ price is defined here as the CO₂ price under which the abated WtE-PCC plant gets the same annual revenue under without PCC. This is shown as the intersect points of the blue line with the grey and orange lines. Without ISS, the break-even CO₂ price is approximately £72/tCO₂ for the abated WtE plant gets the same annual revenue (£5.0 million) as the WtE plant without PCC. Adding ISS to the PCC plant reduces the break-even CO₂ price to approximately £69/tCO₂, at the annual revenue of £5.3 million.
Figure 8-12 Sensitivity of CO₂ prices on the annual profit of WtE-CHP plant under scenarios 1) without PCC, 2) with PCC but without ISS, and 3) with PCC and ISS. Assumptions: Electricity and heat selling prices from the WtE plant are £85/MWh and £79/MWh, respectively; renewable heat incentive £46/MWh; gas-boiler heat price £108/MWh; same fossil CO₂ price and NEC; gate fee £100/t MSW; 2 hour of storage tank capacity)

8.6 KPI results for seasonal solvent storage application for CHP application

The assessment of the performance and quantification of the benefits of adding seasonal solvent storage to a WtE-CHP plant with PCC are conducted in this section by evaluating the thermodynamic, economic, and environmental KPI defined in Chapter 3. Considering the impacts of design variables on the system performance, a set of key parameters is determined to illustrate a case for the ISS application, as shown in Table 8-4.
Table 8-4 Key design inputs for the illustrate ISS case study

<p>| | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>WtE plant MSW treatment capacity</td>
<td>t/year</td>
<td>159.4</td>
</tr>
<tr>
<td>Gate fee</td>
<td>£/tMSW</td>
<td>100</td>
</tr>
<tr>
<td>Electricity selling price</td>
<td>£/MWh</td>
<td>85</td>
</tr>
<tr>
<td>Gas-boiler heat price</td>
<td>£/MWh</td>
<td>108</td>
</tr>
<tr>
<td>Fossil CO$_2$ price</td>
<td>£/tCO$_2$</td>
<td>62.1</td>
</tr>
<tr>
<td>Negative emission credit</td>
<td>£/tCO$_2$</td>
<td>75</td>
</tr>
<tr>
<td>Solvent storage tank storage capacity</td>
<td>m$^3$</td>
<td>494 (2 hours of rich solvent flow at base case)</td>
</tr>
</tbody>
</table>

Note: UK ETS reference price for 2022/2023 (BEIS, 2022d).

8.6.1 Thermodynamic performance under ISS application

The table below summarizes the power and DH heat output from the same WtE-CHP plant, without PCC, with PCC but without ISS, and with PCC and ISS. For all the three scenarios, it is assumed the WtE plant operate all year for the same DH demand pattern as defined in Chapter 8.2, and it prioritises the DH supply over power supply by flexible steam extraction from the steam cycle, as long as it operates within the operation constraints as defined in Chapter 8.3.

Table 8-5 summaries the energy output for the three investigated cases: WtE-CHP plant without PCC, with PCC but without ISS, and with PCC and ISS. It can be seen that with the same thermal input of 55.4MW, power output from PCC cases (with and without ISS) decreases from 12.4MW to around 8MW and 7.8MW for cases without ISS and with ISS. The reduction of power output is mainly due to the power consumption from the PCC plant, e.g. power for the booster fan, solvent pumps, and CO$_2$ compression. For the PCC cases with and without ISS, there is minor difference of power output, since under both cases, it is defined that heat supply prioritises power supply and the same 15% minimum steam flowrate into the lower pressure turbine is applied so that the minimum power output from the low-pressure turbine is the same. The power output for the ISS case is slightly smaller than that for the without ISS case, since under ISS case, the lower pressure turbine can produce more power only when there is low DH demand and the rich solvent tank is empty so that additional regeneration is not possible.
Comparing the DH heat supply, all the three cases fulfil the DH demand, with annual average of 13.2MW. There is different in terms of DH heat supply from the WtE plant itself. The addition of PCC reduces the annual average heat output from the WtE-CHP plant by 33% and 28% for PCC-without ISS and PCC-with ISS case, respectively. This is mainly due to the heat requirement for solvent regeneration of the PCC process. Compare with both PCC cases finds that the addition of ISS helps to increase the DH output from the WtE-CHP-PCC plant by 7.4% and reduce DH heat demand from the gas boilers by 16%.

Table 8-5 Summary of power and heat output under three investigated cases: 1) without PCC, 2) with PCC but without ISS, and 3) with PCC and ISS

<table>
<thead>
<tr>
<th></th>
<th>Unit</th>
<th>Investigated cases</th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Without PCC</td>
<td>With PCC, without ISS</td>
<td>With PCC, ISS</td>
<td></td>
</tr>
<tr>
<td>Thermal input</td>
<td>MW(_\text{th})</td>
<td>55.4</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Net Power output from WtE-CHP plant</td>
<td>MW(_\text{e})</td>
<td>12.4</td>
<td>8.0</td>
<td>7.8</td>
<td></td>
</tr>
<tr>
<td>Heat output from WtE-CHP plant</td>
<td>MW(_\text{th})</td>
<td>13.2</td>
<td>8.8</td>
<td>9.5</td>
<td></td>
</tr>
<tr>
<td>Heat from gas boiler</td>
<td>MW(_\text{th})</td>
<td>0</td>
<td>4.4</td>
<td>3.7</td>
<td></td>
</tr>
</tbody>
</table>

**Note:**
In this table, the energy output (power and heat) refers to the annual average values.

The increased DH output from the WtE-CHP-PCC plant under ISS brings out the benefit of applying ISS, from energy performance point of view. Using the same set of thermodynamic KPIs as defined in chapter 3, the corresponding result for the WtE-CHP-PCC plant for without and with ISS cases is shown in the Figure 8-13. Without ISS, the WtE-CHP plant with PCC presents a EUP of 472 kWh/tCO\(_2\), the addition of ISS reduces the EUP to 443 kWh/tCO\(_2\), a drop by 6% point. Without ISS, the WtE-CHP plant-PCC plant has the EUFP of around 15.8% and the addition of ISS reduces the EUFP by 1% point.
In this study, the addition of solvent tanks and oversizing of stripper and compression train in the ISS scenario increases the capital cost of the PCC plant and, thus, the overall Capex. As reported in chapter 8.4, the addition of ISS increases the capex of the PCC plant from £38.3m to £51.4m. Following the same approach as described in chapter 3 regarding economic KPIs, Figure 8-14 and Figure 8-15 show the LCOH and cost of CO₂ captured for same WtE-CHP-PCC plant, in three scenarios: without PCC, with PCC but without ISS, and with PCC & ISS.

It can be seen that without PCC, the investigated WtE-CHP plant presents a LCOH £46/MWh. The addition of PCC on the WtE-CHP plant increases the LCOH to £56/MWh mainly due to the Capex and Opex of the PCC plant and the cost of purchasing gas-boiler heat in order to fulfil DH demand. With the application of ISS, the LCOH increases to £61/MWh. This implies that comparing the two PCC cases, with the energy prices assumed in this analysis, the additional Capex (oversizing stripper, compressor and additional solvent tanks) of ISS application is larger than the reduction on the cost of gas boiler heat. The LCOH in Figure 8-13 represents the case without ISS and the case with ISS, based on the same gas-boiler price of £108/MWh. The addition of ISS reduces the consumption of gas boiler heat, thus reduce the cost associated with gas boiler heat. When increasing the gas boiler price to
£248/MWh, the LCOHs for without and with ISS are equal, with a value of LCOH at approximately £100/MWh.

The addition of renewable heat incentive can also affect the comparison. For instance, it is found that when the renewable heat incentive is £18/MWh, the LCOH for without ISS and with ISS cases reaches the same, a value of £91/MWh.

Figure 8-14 LCOH (£/MWh) under WtE-CHP 1) without PCC, 2) with PCC and without ISS and 3) with PCC and ISS. (Electricity selling prices from the WtE plant of £85/MWh, fossil CO\(_2\) price and negative emission credit £62.1/tCO\(_2\), gate fee £100/tMSW; 2 hour of storage tank size under ISS)

In Figure 8-15, the cost of CO\(_2\) captured under ISS is however increased slightly by about £4/tCO\(_2\) captured comparing that under the case without ISS. In both without ISS and with ISS cases, the WtE plant capture the same amount of CO\(_2\) from the flue gas, in an operational year, and supply the same amount the total DH demand. So that the only reason for the difference on the cost of CO\(_2\) captured is the increased LCOH under ISS.
A note should be given here that under ISS application, the assumptions given for this integration approach should be treated independently with the assumptions given in the first two integrations (base case, advanced heat integration case in section 6&7), since they represent different application backgrounds. For instance, the ISS approach represents a more incentivized scenario where a carbon emission price and negative emission credit is added to the biogenic CO\(_2\), it also includes a more detailed DH demand curve representing the fluctuation of DH supply (and hot water) from the system. Additional gas boiler heat is also included in the ISS integration, whereas in the Section 6 and Section 7, these variations are not considered.

### 8.6.3 Environmental performance under ISS application

In this study, before PCC is integrated, the WtE-CHP plant can deliver the amount of DH capacity required by the end user, thus there is no requirement for local gas boiler heating and the CO\(_2\) emission associated with burning gas in a local boiler. The addition of PCC affects the DH supply capacity from the WtE plant since steam is diverted to the PCC plant in order to maintain full operation of the CO\(_2\) capture plant, especially under situations when the DH demand is high. Thus, for the cases considered in this work, a local gas boiler for heating will be purchased to fulfil the gap of DH supply.
For the WtE-CHP plant without PCC, the annual fossil CO$_2$ emission from the WtE plant is approximately 65.2 ktone CO$_2$/year. With a typical CO$_2$ emission factor for a stand-alone gas boiler is 0.3 tCO$_2$/MWh (Parliament, 2016), when adding PCC, for the same WtE-CHP plant with the same DH demand pattern as illustrated in the in the Figure 8-2 in the chapter 8.2, there will be 8.4 ktone fossil CO$_2$/year. The addition of ISS to the PCC process reduces the demand for gas boiler heat and thus reduces the CO$_2$ emission from the gas boiler. At the same time, ISS allows full capture of CO$_2$ that comes from burning the fuel, leading to a reduced CO$_2$ emission of 7.0 ktone fossil CO$_2$/year.

Following the efficiency allocation method introduced in Chapter 4, given a fixed ratio of heat and power generation efficiency to be 2 as used by the 1/3: 2/3 Method (used by the Digest of UK Energy Statistics), the CO$_2$ emission intensity on heat basis and electricity basis is shown in Figure 8-16 below. The addition of PCC reduces the direct CO$_2$ emission intensity in both electricity heat basis, from 392 kgCO$_2$/MWh$_e$ and 196 kgCO$_2$/MWh$_{th}$ to -698 kgCO$_2$/MWh$_e$ and -349 kgCO$_2$/MWh$_{th}$ respectively. The addition of ISS further reduces the direct fossil CO$_2$ emission to -715 kgCO$_2$/MWh$_e$ and -358 kgCO$_2$/MWh$_{th}$ respectively. This is mainly due to the reduced DH heat from local gas boilers, thus the associated emission reduction. The overall effect is an almost unchanged CO$_2$ emission intensity on both heat and electricity basis comparing to without ISS.

![Figure 8-16 CO$_2$ emission intensity on heat basis and electricity basis under WtE-CHP plant 1) without PCC, 2) with PCC and without ISS and 3) with PCC and ISS](image-url)
8.6.4 Summary of key findings under ISS application

To sum up, in this chapter, an illustrative case of WtE-CHP plant with PCC under ISS application is modelled and analysed. Using a WtE-CHP plant in EX&C configuration integrated with a PCC plant with 35% MEA based post-combustion that treats 100% of the flue gas by burning the MSW (99.72% CO₂ capture rate as modelled in section 5.2). The stripper and the CO₂ compressors in the PCC plant are oversized to fulfil the fluctuating DH demand in the UK. An operation strategy and constraints are defined to model the operation of this application. Results of the proposed ISS application are demonstrated with sensitivity analyses on the size of storage tanks, gas boiler heat prices and fossil CO₂ prices, along with the KPIs to give an overall insight of this application.

During the operation of ISS, the two solvent storage tanks (rich and lean storage tank) act as buffers between the capture and the regeneration process. Thus enables the WtE plant to better match the fluctuating DH demand and reduces the heat demand from the local gas boilers. Results show the benefit of this application from thermodynamic and environmental points of view: a reduction by 6% of the EUP and 1% of the EUFP, and a reduction of CO₂ emission intensity of 17 kgCO₂/MWhₑ and 9 kgCO₂/MWhₑ, compared to the same WtE-CHP plant with PCC but without ISS.

However, the addition of ISS requires oversizing the stripper and compressors, so that additional cost is required. With the energy prices assumed in this analysis, the economic results implies that the additional Capex of ISS application is larger than the reduction on the cost of gas boiler heat, so that there is an increase of LCOH and cost of CO₂ captured, compared with the same WtE-CHP plant with PCC but without ISS.

It should be noted that the economic results are quite sensitive to the assigned energy prices, carbon prices. Under scenarios of high gas-boiler heat prices and fossil CO₂ prices, the ISS case presents comparable results as that of without ISS case. The high share of additional Capex due to oversizing stripper and compressor implies that future work can be done to look at effectively oversizing the PCC plant to reduce the additional Capex of ISS. The size of storage tank shows limited impact on the performance, which is mainly affected by the designed DH demand pattern.
9 Conclusion

The application of WtE with CCUS technology is one of the promising carbon removal technologies, due to its combined motivation of both promoting sustainable development of urban waste management and delivering ‘negative carbon emissions’ through capturing and permanent storage of the biogenic CO₂ that is produced by incinerating fuel. Previous experience in the thermal integration of power plants with the most commercially available PCC technologies can be applied to the thermal integration of WtE plants with PCC. However, optimized integration approaches that feature the operation characteristics of the WtE plant themselves are still needed to accelerate the deployment of this technology. This study presents three integration approaches – base case, advanced heat recovery and solvent storage approaches, with a comprehensive set of KPIs identified to assess the performance of the investigated integrations. Findings of this study will be helpful for future researchers/plant operators/policy makers on decision-makings on the deployment of this technology.

9.1 Summary of findings

Starting with identifying a representative power-only type WtE plant and modelling validation, this study expands the research on the thermal integration approaches for three types of WtE plant with PCC. A comprehensive set of KPIs is introduced in this study. They provide a rigorous approach to performance evaluation of thermal integration approaches for a range of representative WtE plant configurations with PCC considering technical, economic and environmental factors.

The base case integration of WtE plant with PCC is first developed representing the benchmark scenario that has the minimum system retrofitting and upgrading requirement.

1) Thermodynamic KPIs results show that the steam extraction and condensing type configuration with PCC has the lowest EUP of 231 kWh/tCO₂ to 276 kWh/tCO₂, and the lowest EUFP of 7% to 9.2%, for CO₂ capture rates from 90% to 99.72%. In this configuration, the DH supply capacity is set to be at the capacity where the steam flow rate through the LP steam turbine cylinder reaches its limit of 15% of the nominal flow rate at full load, thus allowing the CHP plant with PCC to supply as much as heat as possible. At the same time, the DH supply capacity is maintained the same before and after PCC. The overall result is the minimum energy penalty per unit of CO₂ captured and the lowest efficiency penalty. By contrast, the backpressure configuration with PCC shows the highest EUP of 1084 kWh/tCO₂ to 1125 kWh/tCO₂, and EUFP of 33% to 38%, for CO₂ capture rates from 90% to 99.72%. This is mainly due to the large difference of DH output before and after PCC for this configuration. The overall COP_{X} values for all the WtE plant with PCC are in range of 5 to 5.2 with variations in different steam cycle configurations and PCC capture rates. The COP_{X,abs} of steam
extraction is around 0.9 and 0.65 for WtE-CHP extraction and condensing configuration and backpressure configuration, respectively. Both $\text{COP}_X$ are relatively higher than those of commercially available heat pumps. This proves that compared with potential heat pump approaches, steam extraction is a more efficient method to provide thermal heat (@130°C) for solvent regeneration in the reboiler.

2) Economic KPIs show that the addition of PCC on WtE-power-only plant increases the LCOE from the original value £85/MWh (WtE-power-only without PCC) to £162-£179/MWh with capture rates from 90% to 99.72%. The influence of CO$_2$ capture rates on the LCOE is minimal. For the WtE_CHP plants investigated in this study, the integration of PCC increases the LCOH. Comparing the two CHP plants, the WtE-CHP-BP with PCC shows the lowest LCOH in the range of £171-£205 per MWh for CO$_2$ capture rates from 90% to 99.72%, which is due to the high DH output in this scenario. The analysis demonstrates that under WtE-CHP-BP configuration with PCC, the LCOH is about £20/MWh higher than the values of the WtE-CHP-Ex&C without PCC.

3) For the representative WtE plant considered in this study, the total direct CO$_2$ emissions from fuel combustion are 891 kg/t MSW. With the biogenic carbon ratio of 60%, 535 g CO$_2$/kg MSW are of biogenic origin and considered to be carbon neutral, 353 g CO$_2$/kg MSW are of fossil origin. When the WtE is equipped with PCC, the amount of biogenic CO$_2$ captured and permanently stored leads to ‘negative’ carbon emissions: -0.45 kgCO$_2$/kg MSW, -0.49 kgCO$_2$/kg MSW, and -0.53 kgCO$_2$/kg MSW for 90%, 95% and 99.72% capture respectively. With the DUKES 1/3: 2/3 method used to allocate the CO$_2$ emission to electricity and heat generation separately, the carbon intensity of the WtE-power-only facility without PCC modelled in this work is approximately 519 kgCO$_2$/MWh, the integration of PCC enables the plant to reach a negative carbon intensity of -988 kgCO$_2$/MWh to -1272 kgCO$_2$/MWh, for a range of CO$_2$ capture rates. For CHP plant, the lowest carbon intensity on heat basis occurs for the steam extraction & condensing configuration, with a negative value of -469 kgCO$_2$/MWh.

The advanced heat integration introduces an approach that requires modifications of the existing heat exchanger network (additional heat exchangers, hot water pipes and heat pumps) but enables the abated plant to maximize the net power output under power-only generation and maximize thermal output under CHP generation. The same set of KPIs is used to assess the performance of this integration approach.

1) Thermodynamic KPIs results show that for the power-only configuration, advanced heat integration reduces the energy output penalty of PCC by about 30 kWh/tCO$_2$. This reduction is mainly due to the heat recovery from the stripper overhead condenser for the boiler feed-water heating. For the two WtE-CHP plants, it is seen that heat recovery greatly improves the EUP under the backpressure
configuration with PCC. For this configuration, the WtE-CHP plant produces as much heat as possible and all the excess heat from the PCC process is recovered (including the addition of a heat pump) to produce DH heat. The corresponding EUPs are reduced from the 1084-1125 kWh/tCO₂ under base case integration to 284-221 kWh/tCO₂, for CO₂ capture rates from 90% to 99.72%. Heat recovery increases both the $COP_{X_{-cap+dh}}$ and the $COP_{X_{-abs}}$ values in comparison to the base case integration. This strengthens the advantage of steam extraction approach over the alternative heat pump approach to provide thermal heat for solvent regeneration.

2) Economic KPI analysis finds that for WtE-power-only plant, heat recovery from the stripper overhead condenser for boiler feed water heating helps to reduce the LCOE by around £8/MWh. The LCOE is sensitive to the assigned CO₂ price. Under the advanced heat integration scenario, the CO₂ price should be around £240/tCO₂ to enable the LCOE of the WtE plant with and without PCC to be equal, which will be around £160/MWh to £170/MWh for a range of CO₂ capture rates from 90% to 99.72%. For the WtE-CHP plant under steam extraction and condensing configuration with PCC, it is found that the improvement of LCOH by heat recovery is limited due to the maintained DH supply capacity defined in this study. For the WtE-CHP plant under backpressure configuration, the application of heat recovery helps to reduce the LCOH by £70-£90/MWh, compared with those for the base integration under the range of CO₂ capture rates.

3) Environmental KPI analysis finds that the advanced heat recovery reduced the negative CO₂ emission intensity of the integrated WtE plant with PCC. This is because in this study, it is assumed the WtE plant is operating at full load, so for each CO₂ capture rate, the absolute amount of CO₂ emission from the plant is the same under the base cases and the advanced heat integration cases. Increasing power and or heat output under the advanced heat integrations thus lowers the negative CO₂ emission per unit of energy output. The lowest CO₂ emission intensities are -1214 kgCO₂/MWhₑ and -419 kgCO₂/MWhₖ on electricity and heat basis, respectively.

This study also investigates new strategies for flexible operation of PCC systems integrated into WtE-CHP plants, with the aim of deploying seasonal interim solvent storage (ISS) to improve the flexibility in both CCS and CHP in response to fluctuating DH demand conditions. The analysis focused on the addition of the two solvent storage tanks that act as buffers between the capture and the regeneration process, depending on the fluctuating DH demand. The results show that the highest annual profit is achieved for a tank size of 2 hours of solvent storage, approximately 494 m³.

The purchasing price of the gas boiler heat if the CHP plant is not available affects the annual profit. For the benchmark gas-boiler heat price of £130/MWh, the application of ISS can increase the annual
profit of the abated WtE plant from £2.6million to £2.9million. The increase in annual profit is more significant with higher gas boiler heat prices.

Sensitivity analysis of the cost of storage tanks finds that its effect on the overall profit is minimal. The addition of ISS increases the capex of the PCC plant from £38.3m to £44.1m. A breakdown of the cost element finds that the additional cost for the rich and lean storage tanks is only £0.07m, a share of 1.2% of the additional cost. The main cost increases are due to oversizing the compression train and oversizing the stripper diameter to allow additional regeneration of solvent, which takes 56% and 43% respectively.

The same set of KPIs developed in this study are calculated and compared for the cases without PCC, with PCC but without ISS, and with PCC&ISS.

1) Thermodynamic KPI results show that the addition of ISS helps to reduce the EUP of WtE-CHP plant with PCC from 21 kWh/tCO₂ to 16 kWh/tCO₂. Additionally, EUFP is reduced by 1%. This is mainly due to the increased DH supply capacity from the WtE plant.

2) Economic KPIs results show that, with the assumption of electricity selling prices of £85/MWh, fossil CO₂ price £75/tCO₂ (negative emission credit not considered), gate fee £100/t MSW and 2 hour of storage tank size (494 m³) for ISS, the addition of ISS reduces the LCOH by £10/MWh. There is a slight increase in the cost of CO₂ captured by around £3/tCO₂ due to the increased DH supply capacity under ISS and then a reduction of CO₂ emission intensity on heat basis.

3) Following the DUKES’ efficiency allocation method, given a fixed ratio of heat and power generation efficiency, environmental KPIs analysis finds that although ISS reduces the total CO₂ emission from the local gas boilers, increasing DH supply from the WtE plant with ISS also increases the share of CO₂ emission allocated to the DH supply. The overall effect is an almost unchanged CO₂ emission intensity on both heat and electricity basis compared to cases without ISS.

9.2 Limitations and recommendations for future work

The timeline and the defined research target mean the author can’t address all the issues that were raised along the progress of this PhD research. This section discusses some of the limitations of the work presented here as recommendations and directions for further research.

1) In this PhD study, for the base case integrations and the advanced heat recovery integrations, only one representative MSW composition referencing from an operational WtE plant is used. It is assumed that the WtE plants are operating at full load, since a significant part of their income is from the gate fee for taking the waste. However, in real operation, there will be continual variation in terms of the MSW composition and throughputs. In future work, it will be necessary to perform relevant
analysis on these uncertainties, for instance, how the heterogeneous nature of MSW and the dynamic operation (part load, start-up and shut-down) will affect the KPI results of the different integrations developed in this study.

2) Building upon previous research on the concept of COPX, this study expands its application into CHP plants with PCC. The relevant COPX results presented in this study prove that steam extraction for solvent regeneration is a thermodynamically efficient approach for providing heat for solvent regeneration (including in comparison to potential heat pump approaches). This conclusion depends on the assumptions defined in this study, such as the temperature level for solvent regeneration (130°C) and DH supply/return water temperature (80°C/60°C). The COP of a commercial heat pump is significantly affected by the temperature lift (temperature difference of heat sink and source). For scenarios with lower regeneration temperature (i.e. a different solvent for CO2 capture), and lower DH supply/return water temperatures, the COP of the comparable heat pump solutions may be increased. This may lead to a different conclusion in the comparison of regeneration heat supply options. It will be useful for future researchers to consider the multiple variables and uncertainties that exist in this COP comparison thus delivering enhanced understanding of how the development of novel solvents and/or rollout of improved heating systems in DH networks might change the preferred approach to heat supply for solvent regeneration.

3) A large part of this PhD thesis is motivated by optimizing the CHP operation of WtE plant with PCC. A better understanding of a CHP system will facilitate the effective optimization. Although this study models two of the most representative CHP configurations (steam extraction and condensing configuration, and back-pressure configuration), research on CHP integration with CO2 capture technologies is still awaiting to be further explored. For example, at the time of writing, the fifth-generation district heating and cooling systems (5GDHC) is quite promising in the DH sector. It has features including operating at near ambient temperature, ensuring the maximum waste heat recovery potential, providing simultaneous heating and cooling through the same pipeline, etc. Future research on the CHP integration with PCC should consider these features that may be under development or already exist, thus stimulating improved system integration.

4) The cost of CO2 captured is applied as one of the economic KPIs for this study. As already discussed in the corresponding chapter, this KPI avoids the uncertainty around the cost estimation for CO2 transportation and storage, but also has its limitation. It only provides insights into the investments and operating costs associated with CO2 capture (and compression) but can’t be used to assess the total cost of carbon mitigation of this application. Future work can use the cost of CO2 avoided when the associated cost data of transportation and storage are also included in the
calculation, thus enabling necessary comparisons of this technology with other CO₂ removal technologies.

5) In the solvent storage application, the DH demand pattern is chosen using an established model based in the UK and trying to capture the dynamic nature of DH demand as much as possible, both in terms of daily and seasonal variations. The results presented in this study are focussed on this specific DH demand pattern. It is also necessary to understand how different DH demand patterns (variations) may affect the performance of the solvent storage application. This can be done by integrating the ISS with additional DH patterns or integrating the ISS with more robust DH demand forecasting models, thus delivering better understanding on the sensitivity of DH demand variation on the operation and decision making of ISS.

6) In the solvent regeneration cases, cost analysis finds that the additional cost relating to oversizing the stripper and compression train takes around 99% of all the additional costs required for solvent regeneration. In this study, the oversizing of the stripper and compression is determined by the maximum amount of steam that could be extracted from the steam cycle, which corresponds to the situation that DH demand is zero. In these circumstances, the maximum possible amount of steam from the cross-over pipeline between IP and LP turbine is extracted for solvent regeneration. Future research on effective oversizing of the equipment could be useful to drive down the additional cost for solvent regeneration, thus improving the economic performance of this application.

7) Finally, the environmental assessment of the thermal integration in this study focuses only on the absolute amount of CO₂ reduction from the abated WtE plant itself and does not analyse the overall net climate benefit of a wider system. It is necessary to look at the net climate impact it may bring. This can include:

   a) Energy substitutes for the production of electricity and heat from fossil energy sources, such as coal and natural gas;
   b) Avoided GHG emissions from diverting waste from landfilling;
   c) Ferrous and non-ferrous metals recovery;
   d) Fly ashes are currently utilized for neutralization of waste acid, substituting limestone;
   e) Bottom ash can be used for road construction, substituting gravel; and
   f) Additional material uses and air emissions due to the PCC, such as the MEA degrades during the process or lost due to formation of heat-stable salts or as vapour and aerosols during stripping.
In order to consider all the elements that contribute to the net climate impact, a systematic life cycle assessment (LCA) should be performed. However, this is outside the scope of this study. Literature research also finds limited work has been done in the research field and the identified gap could be filled in future work. The most relevant LCA studies by Bisinella et al. concluded that with CCS amending the Amager Bakke incineration plant in Copenhagen, the net climate change benefit is $-670$ kg CO$_2$-eq per tonne wet weight (ww). The largest benefit derives from storage of captured biogenic CO$_2$, which provides a climate change benefit of $-530$ kg CO$_2$-eq per tonne ww. The MEA consumption for the capture process adds only 4 kg CO$_2$-eq per tonne ww to the climate change impacts of CCS amended configurations (Bisinella et al., 2022).

A life cycle assessment (LCA) of the environmental impacts of a WtE facility with ultra-high capture rates is reported co-authored by the author of this thesis. Result shows that adding CCS can provide a significant improvement in climate change impact, and achieve a net climate benefit. Without significant burden shifting to other environmental impact categories, the climate change impact is significantly reduced from 388 kg CO$_2$-eq/t MSW to -482 kg CO$_2$-eq/t MSW with biogenic CO$_2$ captured and permanently stored in geological formations accounting as negative CO$_2$ emissions. When the avoided Greenhouse Gas (GHG) emissions from electricity, district heating and material recovery are considered, the climate benefit is 646 kg CO$_2$-eq/t MSW for a power-only WtE plant exporting 9.6 MW$_e$, and 773 kg CO$_2$-eq/t MSW for a combined heat and power (CHP) WtE plant exporting 6.2 MW$_e$ and 18.5 MW$_{th}$ (Herraiz et al., 2023).

9.3 Contribution to knowledge

In this concluding section, key contributions arising from the innovating aspects of PhD study are highlighted.

- This thesis lays much emphasis on assessing KPIs that have been tailored to evaluate the performance of WtE plants equipped with PCC from thermodynamic, economic and environmental points of view. The variations of the KPI results on each integration scenario investigated in this study highlight the importance of multi-angle assessment of this technology, especially when comparing it with other low carbon emission technologies using similar metrics.

- Besides the level of integrity of the KPIs, this thesis expands the work of (Lucquiaud & Gibbins, 2011) about the Coefficient of Performance of steam extraction ($COP_X$), and introduces $COP_{X-cap+dh}$ and $COP_{X-abs}$ metrics specific to CHP plants that can be used to evaluate the effectiveness of steam extraction for DH and solvent regeneration. In this study, for the specific MEA based CO$_2$ capture with regeneration temperature at approximately 130°C, the $COP_{X-cap+dh}$ and $COP_{X-abs}$ under all the investigated integrating options outweighs the $COP_{hp}$ of analogous heat pump solutions.
• The $COP_X$ metric can also be used to compare the effectiveness of steam extraction between the integrating options. Results show that under basic heat integration, the CHP-BP configuration with PCC has the highest $COP_{X_{,cap+dh}}$ since there is no condensing heat losses (all the DH output can be effectively used by the DH system) in this configuration. With advanced heat recovery, more power can be generated from the lower pressure turbine, so that the $COP_{X_{,cap+dh}}$ for CHP-EX&C configuration can be effectively increased to higher level than that under CHP-BP configuration, to a value of 8.6.

• This work evaluates for first time the effect of ultra-high CO$_2$ capture levels on the performance of the WtE plants. Through the process modelling of 35% MEA based PCC integrated into WtE plants, this study highlights that comparing the benchmark 95% capture rate (and 90% capture rate for further comparison with other studies), the ultra-high CO$_2$ capture rate can be designed and achieved with marginal increase in terms of thermodynamic penalty and economic cost. The optimized condition of CO$_2$ capture rate is essentially a three-way trade-off among energy penalty, economic cost, and environment benefit. This work, therefore, provides rigorous evidence for future design and implementation of CO$_2$ capture (at least for using MEA based CO$_2$ capture) in terms of choices of capture rates, which is especially important in the current world where net zero emissions become critical.

• Another contribution of this study comes from expanding the previous application of interim solvent storage on power-only plant integration with PCC into the new CHP integration, to solve the competing heat requirements for the fluctuating DH supply and for solvent regeneration. Modelling results prove the benefit of this application from thermodynamic and environmental points of view: a reduction by 6% of the EUP and 1% of EUFP, and a reduction of CO$_2$ emission intensity of 17 kgCO$_2$/MWh$_e$ and 9 kgCO$_2$/MWh$_{th}$, compared to the same WtE-CHP plant with PCC but without ISS. The economic results are found to be sensitive to the assigned energy prices and carbon prices. Under scenarios of high gas-boiler heat prices and high fossil CO$_2$ emission prices, the ISS case presents comparable results as that of without ISS case. Effective oversizing of the stripper and compression train will be useful to improve the economic performance of this application.

• Finally, although this study is built upon a series of generic WtE plants with PCC, the methodology and findings from this study may be generalized to wider power and heat generation technologies with post-combustion CO$_2$ capture, especially for CHP applications in the future urban district heating industry.
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196


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Appendix

Appendix 4.1 Process flow of the WtE_power only configuration in gProcess
Appendix 4.2 Process flow of the WtE_CHP-Steam extraction and condensing configuration in gProcess
Appendix 4.3 Process flow of the WtE_CHP-Backpressure configuration in gProcess
Appendix 6.1

Thermodynamic KPI result of WtE plant with PCC under base case thermal integration

<table>
<thead>
<tr>
<th></th>
<th>WtE-Power-only</th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>With PCC</td>
<td>90% PCC</td>
<td>95% PCC</td>
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<tr>
<td>Power output</td>
<td>kW</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>14332</td>
<td>9418</td>
<td>9121</td>
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<tr>
<td>Electricity output penalty (EOP)</td>
<td>kWh/t CO₂</td>
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<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>294</td>
<td>311</td>
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<tr>
<td>Efficiency Penalty (EP)</td>
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<td>-</td>
<td>8.9%</td>
</tr>
<tr>
<td>Coefficient of Performance for (steam) extraction (COPₓ_cap)</td>
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<td>5.15</td>
<td>5.13</td>
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<table>
<thead>
<tr>
<th></th>
<th>WtE-CHP (Ex-C)</th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>90% case</td>
<td>95% case</td>
<td>99.72% case</td>
</tr>
<tr>
<td>Power output</td>
<td>kW</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>10641</td>
<td>6771</td>
<td>11085</td>
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<tr>
<td>Heat output</td>
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<td></td>
</tr>
<tr>
<td></td>
<td>13000</td>
<td>1300</td>
<td>11950</td>
</tr>
<tr>
<td>Energy utilization penalty (EUP)</td>
<td>kWh(t CO₂)</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>-</td>
<td>231</td>
<td>-</td>
</tr>
<tr>
<td>Energy utilization factor penalty (EUFP)</td>
<td>-</td>
<td>-</td>
<td>7.0%</td>
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<tr>
<td>-----------------------------------------</td>
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<tr>
<td>Coefficient of Performance for (steam) extraction (COP\textsubscript{x,\text{cap&amp;dh}})</td>
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<td>-</td>
<td>5.04</td>
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<td>Effective electricity efficiency (EEE) penalty</td>
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<td>-</td>
<td>9.9%</td>
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<table>
<thead>
<tr>
<th>WtE-CHP (BP)</th>
<th>Witho ut PCC</th>
<th>90% PCC</th>
<th>95% PCC</th>
<th>99.72%PCC</th>
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<tbody>
<tr>
<td><strong>Power output</strong></td>
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<td>5915</td>
<td>5834</td>
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<tr>
<td><strong>Heat output</strong></td>
<td>kW</td>
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<td>1861</td>
<td>9</td>
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<td>Energy utilization penalty (EUP)</td>
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<td>1091</td>
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<td>Effective electricity efficiency (EEE) penalty</td>
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<td>-</td>
<td>47.6</td>
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</table>
## Appendix 7.1

Thermodynamic KPI result of WtE plant with PCC under advanced thermal integration

<table>
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<th>WtE-Power-only</th>
<th>WtE-CHP (Ex-C)</th>
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<td>Without PCC</td>
<td>90% PCC-adv</td>
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<tr>
<td><strong>Power output</strong></td>
<td>kW</td>
<td>14332</td>
</tr>
<tr>
<td><strong>Electricity output penalty (EOP)</strong></td>
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<tr>
<td><strong>Efficiency Penalty (EP)</strong></td>
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<tr>
<td><strong>Coefficient of Performance for (steam) extraction (COPₓ_cap)</strong></td>
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<td>6.08</td>
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<tr>
<td></td>
<td><strong>WtE-CHP (Ex-C)</strong></td>
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</tr>
<tr>
<td></td>
<td>90% case</td>
<td>95% case</td>
</tr>
<tr>
<td></td>
<td>Without PCC</td>
<td>With PCC-adv</td>
</tr>
<tr>
<td><strong>Power output</strong></td>
<td>kW</td>
<td>10641</td>
</tr>
<tr>
<td><strong>Heat output</strong></td>
<td>kW</td>
<td>13000</td>
</tr>
<tr>
<td><strong>Energy utilization penalty (EUP)</strong></td>
<td>kWh/(e, th)/tC O₂</td>
<td>-</td>
</tr>
</tbody>
</table>
### Energy utilization factor penalty (EUFP)
- - 4.7% - 5.5% - 6.7%

### Coefficient of Performance for (steam) extraction (COPx, cap&dh)
- - 8.61 - 8.19 - 7.69

### Effective electricity efficiency (EEE) penalty
- - 6.6% - 7.6% - 8.7%

### WtE-CHP (BP)

<table>
<thead>
<tr>
<th>Power output</th>
<th>kW</th>
<th>7613</th>
<th>3920</th>
<th>3840</th>
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<tr>
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<td>35064</td>
<td>18619</td>
<td>17559</td>
<td>1604</td>
</tr>
</tbody>
</table>

### Energy utilization penalty (EUP)
- kWh(e, th)/tC O₂ - 1203 1204 1233

### Energy utilization factor penalty (EUFP)
- - 36.4% 38.4% 41.3%

### Coefficient of Performance for (steam) extraction (COPx, cap&dh)
- - 5.23 5.24 5.24

### Effective electricity efficiency (EEE) penalty
- - 53.8% 54.5% 55.3%