Advanced steady-state modelling and optimisation of LNG production

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Abstract

The main goal of the work was the development of a mathematical model of a typical liquefied natural gas (LNG) cycle, and to demonstrate the use of the model for steady-state optimisation.

The liquefaction of the natural gas (NG) is a common procedure that allows a volume reduction by a factor of 600, resulting in several advantages in terms of transportation and storage. A typical LNG liquefaction process consists in series of refrigeration cycles, using both pure and mixed refrigerants.

Two advanced modelling approaches (simple and complex) were taken for the following LNG processes: (i) C3/MR, (ii) AP-X™ and (iii) AP-X™ alternative, using both Européen de Recherches Gaziéres (GERG) and Peng-Robinson (PR) property methods. Both property packages affect the results in a negligible difference of less than 0.1%.

The optimisation improved the total cost of the AP-X™ process in 58%, ending up with a value of around 83 M$/year and a specific power consumption (SPC) of 15.8 [kW-day/tonne LNG], fitting up the standard industrial values.

Keywords: LNG, AP-X, C3/MR, gPROMS, refrigeration

1. Introduction

In its naturally occurring vapour state, NG is a bulky energy source. Transporting NG from production sources to points of consumption requires large pipeline networks. Thus, only overland or somewhat shorter undersea routes can be considered.

The main reason for liquefying NG is a 600-fold reduction in volume that occurs with the vapour-to-liquid phase. LNG has become an increasingly important supply source in meeting the world’s energy needs. Of all NG consumed in 2011, 10% was transported in the form of LNG. Such scenario was unimaginable twenty years ago.

However, the past two decades have been characterised above all by the development and diversification of gas markets worldwide. It is estimated an annual market growth of 9.5% until the year of 2020. This LNG demanding must be fulfilled by the current and future LNG technologies that after 30 years of stagnation must evolve to more efficient levels and to higher capacities.

A proper prediction of the industrial behaviour of an LNG plant is becoming more and more important to improve the exiting plants’ results and to optimise others in project phase. When talking about processes with huge capacities of flammable components, operating at high pressure and cryogenic conditions, a rigorous and complex modelling is a great challenge.

The main goals to achieve with this project are the developing of a mathematical model of a typical LNG cycle, and to demonstrate the use of the model for steady-state optimisation.

The literature review showed that the Air product and Chemical International (APCI®) processes are still the leaders in terms of LNG industry with more than 80% in terms of production capacity of the LNG global trades. The C3/MR process ruled the industry for more than 30 years being now upgraded and gradually replaced for the new APCI® technology: the AP-X™ process.

2. Background

LNG liquefaction stage[1]

The liquefaction section is the key element of a LNG plant. Liquefaction technology is based on refrigeration cycles, where a refrigerant by means of successive expansion and compression transports heat from a lower to a higher temperature. LNG plants consist of parallel units, called trains. Liquefaction train capacity is primarily determined by the liquefaction process, refrigerant used, HXs capacity and largest available size of the compressor driver combination that drives the cycle.

The liquefaction section typically accounts for 30% to 40% of the capital cost for the overall plant, which in turns accounts for 25% to 35% of total project costs.
For many years, there was absolutely no problem to choose the process of a new liquefaction plant: the C3/MR process was the only choice. The same process was implemented again and again, with small improvements until nowadays.

The following technologies for the NG liquefaction are available: (i) Conoco-Phillips optimised cascade, (ii) PRICO process, (iii) APCI® propane pre-cooled mixed refrigerant (MR) process (C3/MR), (iv) APCI® AP-X™ process, (v) Shell® dual MR process (DMR), (vi) Axens/IPF Liquefin™ process and (vii) Statoil®/Linde® mixed cascade process.

**APCI® propane pre-cooled MR process (C3/MR)**

APCI® began to dominate the industry from the late 1970s on. This process accounts for a very significant proportion of the world’s LNG production capacity. Train capacities up to 4.7 MTPA have been built or are under construction. The C3/MR process uses a MR mainly composed of nitrogen, methane, ethane and propane. The NG feed is initially cooled by a separate set of propane chillers to an intermediate temperature of about -35°C, at which the heavier components in the feed gas condense and are sent to fractionation. The NG is then sent to the main cryogenic heat exchanger (MCHX). These allow very close temperature approaches between the condensing and boiling streams. The MR refrigerant is partially condensed in the propane chillers before entering the cold box. The separate liquid and vapour streams are then further chilled before being flashed across Joule-Thompson (J-T) valves providing then the cooling for the final gas liquefaction.

**APCI® AP-X™ process**

The AP-X™ process is essentially an upgrade of the C3/MR process, with the inclusion of one extra refrigeration cycle: propane pre-cooling cycle, MR cycle and nitrogen expander cycle. The HXs used in the propane cycle are kettle type HXs, while the MCHX contained in the cold boxes are of the type plate and fin or wound coil (PFHX or WCHX). The process is analogous to the C3/MR process however the new third cycle allows the decreasing the propane and MR flowrates which allows a better match of the MR composition as well as smaller quantities of flammable compounds on the plant.

MR vapour from the shell side of the MCHX is compressed in an axial compressor, followed by a two-stage centrifugal compressor. Inter-cooling and initial de-superheating is achieved by air-coolers.

The MR vapour and liquid are separated, and further cooled in the main cryogenic heat.

### Methodology

As mentioned before, the APCI® processes rule more than 80% of worldwide LNG trades in terms of volume. The C3/MR process has dominated the industry since the seventies until the 21st century, being even today the market leader.

It is unanimous by LNG producers that the robustness, simplicity, capacity and invested know-how in all the APCI® processes are well above all the other licensors.

Being the AP-X™ an upgrade in terms of capacity and efficiency of the old C3/MR process this is indeed the process of the future due to the LNG demanding of the next few years. This was the process chosen for the modelling and optimisation stages.

The information gathered about the LNG processes shown that the APCI® AP-X™ process was a new, but reliable process with a growing market share with some industrial applications. It had all the advantages of the leader process APCI® C3/MR with a larger capacity. It was also performed the modelling of the C3/MR process and an alternative scenario of the AP-X™ process.

The first part of the work, after the literature review was the modelling of the process C3/MR, AP-X™ and AP-X™ alternative, using both PR and GERG properties package, following a simple approach. The simple approach relies on the flowsheet assembling on the topology environment of a top level model using PML:SS models. This approach was all about to get good initial values to be applied in a more complex approach.

A notorious gap of the PML:SS library was the lack of a proper multistream heat exchanger (MSHX). In combination with a consulting project for Proctor and Gamble®, it was develop a model of a distributed PFHX to be applied to liquefaction units. This was the second part of the project.

The third part of the work was all about the introduction of the new PFHX model in the LNG models developed in the simple approach. Using the same type of assignments and the initial values got in the simple approach it was possible to initialise and converge the models, for both PR and GERG methods.

The last part of the work was a steady-state optimisation of the AP-X™ model using PR method.

### LNG process modelling – simple approach

In the simple approach, the idea was to build up the LNG flowsheet in the topology environment, get a simple and
quick simulation to obtain some good first results to be used on a more complex modelling methodology which includes design mode in the models. For both property methods used, GERG equation of state (EOS) and PR EOS, the inputs used were exactly the same. The resulting variables of the models show less than 0.1% difference for both property packages. Note that throughout this approach all possible heat losses and pressure drops on the units were neglected.

Figure 1: Topology representation of the AP-X™ process model.

Figure 2: Topology representation of the alternative AP-X™ process model.

Figure 3: Topology representation of the C3/MR process model.
The figure 1 represents the topology representation at gPROMS model builder of the AP-X™ process model. Note that only some models are identified. These models are the key elements to understand the global flowsheet network. The non-identified models are not purposely identified to simplify and smooth the topology environment, being anyway as all the other models an indispensable part of the overall work.

For an easier comprehension, the following notation was adopted to identify models/equipment of the process streams: (i) MR-x for the MR cycle, (ii) P-x for the Propane cycle, (iii) N-x for the Nitrogen cycle, (iv) NG-x or LNG-x for the NG-LNG line.

**AP-X™ process - flowsheet assembly**

The flowsheet assembly was done in stages. In particular, the model was assembled by stream type, starting up with the NG-LNG line alone. Step by step (with the respective assignments and testing simulation) units of each refrigeration cycle were added until the final version of the model.

**NG-LNG Line**

The source NG-1 simulates the downstream conditions of a pre-treatment unit. The incoming NG is cooled down until 242K, using a series of five kettle type HXs, at different pressure and temperature stages, using saturated propane as cold fluid.

After the propane pre-cooling stage, the NG enters a gas-liquid separator to remove the condensates, before being further cooled by the MR and the nitrogen, at the MCHXs. The liquid outlet of the separator is normally sent to a fractionating plant where the heavy compounds as propane, butane and C5 plus, might be fractionated. The gas outlet, is sent to the MSHX_simple-1 where is further cooled until 171K using expanded MR as cold fluid.

After that stage the NG is then totally liquefied and sub-cooled until 117K at the MSHX_simple-2 using expanded nitrogen as cold fluid, becoming LNG.

At the final stage, the LNG is expanded from its inlet pressure, until 1.05 bar at the J-T valve LNG-1, sent to a gas-liquid separator, where the outlet liquid stream might be dispatched to the commercial pipeline or storage, sink LNG-2.

**Nitrogen refrigeration cycle**

The nitrogen refrigeration cycle is a closed loop stream. Consider the stream analyser N-1. The nitrogen is cooled down to 242K, using a series of five kettle type HX, at different pressure and temperature stages, using saturated propane as cold fluid.

After the pre-cooling stage, the nitrogen is further cooled at the MSHX_simple-3, using an expanded MR stream, and itself (after expansion and warming at the MSHX_simple-2) as cold fluids, before being isentropically expanded (at the expander N-2) and used for the NG liquefaction and sub-cooling.

At the expander N-2 the nitrogen passes from 67.1 to 10.5 bar, cooling down from 170 to 111K.

The expanded cold nitrogen (at loop_breaker N-3) is used as cold fluid at the MSHX_simple-2 to cool down the NG. The resulting warm stream is again used as cold fluid to cool down the high pressure nitrogen coming from the propane pre-cooling cycle.

The nitrogen is then compressed at three stage intercooled compression train until the pressure of 67.1 bar and 300K, ending up with the conditions of the starting of N-1.

**MR refrigeration cycle**

Analogously to the nitrogen line the MR refrigeration loop was built in the same way.

Consider the stream analyser MR-1. The MR is cooled down until 242K, using a series of five kettle type HX, at different pressure and temperature stages, using saturated propane as cold fluid.

After the pre-cooling stage, the MR is further cooled at the MSHX_simple-1, using an expanded MR stream (major split of the main line), before being isentropically expanded (at J-T valve MR-3) and used as cold fluid at the same MSHX_simple-1.

At the J-T valve MR-3 the MR passes from 38.6 to 9.8 bar, cooling down from 200.3 to 167.6K.

The minor split of the MR main line, is used as cold fluid to cold down the nitrogen stream (at MSHX_simple-3), as explained in the previous point. After the MSHX_simple-1, both splits of the MR line are mixed before compression.

The MR is then compressed at a two stage intercooled compression train until the pressure of 38.6 bar and 305.1K, ending up at the conditions of the starting point of MR-1.

**Propane pre-cooling cycle**

The propane pre-cooling cycle is a five stages cascade refrigeration cycle.

The cascade pre-cooling intends to pre-cool the NG-LNG, MR and nitrogen streams. At each pressure step the saturated propane is directed to a set of three kettle type HXs in series, working as cold fluid on the shell-side, while the NG, MR and nitrogen are cooled down on the tube-side, vaporizing part of the propane boiling pool.
After each set of kettles, the warm propane is redirected to a gas-liquid separator, where the outlet gas stream is sent to compression while the outlet liquid stream is expanded (respectively at J-T valves P-1, P-2, P-3, P-4 and P-5) and sent to the next set of HXs.

**AP-X™ alternative process**[^3][^4]

The figure 2 represents the topology representation of the AP-X™ alternative process model at gPROMS model builder. Note that the difference to the AP-X™ process is the removal of the propane pre-cooling cycle. All the procedure taken for the AP-X™ model assembling is analogous for this case.

**C3/MR process**[^5][^6]

The figure 3 represents the topology representation at gPROMS model builder of the C3/MR process model. The process is analogous to the AP-X™ process without the nitrogen refrigeration cycle. All the procedure taken for the AP-X™ model assembling is analogous for this case.

**Results**

As said before, this step of the project was about getting good initial values for the complex approach. For all the situations the relevant results differs less than 0.1% running with the PR or GERG EOS and due to that only the results with PR physical properties will be shown and analysed.

Because liquefaction units are very energy demanding processes, the analysis of the energy consumption is a benchmark in the LNG industry. Some other relevant results are presented in the table 1.

The main equipment considered was all the relevant units involved in compressions, expansions or heat transfer.

**Table 1: Simple approach – main results.**

<table>
<thead>
<tr>
<th>Processes</th>
<th>AP-X™ alternative</th>
<th>AP-X™ alternative</th>
<th>C3/MR</th>
</tr>
</thead>
<tbody>
<tr>
<td>No of variables</td>
<td>2186</td>
<td>1040</td>
<td>1645</td>
</tr>
<tr>
<td>No of equations</td>
<td>1858</td>
<td>919</td>
<td>1413</td>
</tr>
<tr>
<td>Total CPU time (s)</td>
<td>8</td>
<td>3</td>
<td>5</td>
</tr>
<tr>
<td>No of main equipment</td>
<td>33</td>
<td>12</td>
<td>20</td>
</tr>
<tr>
<td>NG inlet (kg/s)</td>
<td>118.3</td>
<td>118.3</td>
<td>98.9</td>
</tr>
<tr>
<td>LNG Produced (kg/s)</td>
<td>109.7</td>
<td>111</td>
<td>83.7</td>
</tr>
<tr>
<td>Methane yield (%)</td>
<td>93.8%</td>
<td>93.8%</td>
<td>84.4%</td>
</tr>
<tr>
<td>Power consumption (MW)</td>
<td>438</td>
<td>1062</td>
<td>129</td>
</tr>
<tr>
<td>SPC (kW-day/tonne LNG)</td>
<td>46</td>
<td>111</td>
<td>18</td>
</tr>
</tbody>
</table>

There is a big difference between the AP-X™ processes and the C3/MR process. This is due to the fact that the inputs for the first were based on their patent[^5] information, while the inputs of the C3/MR process were based at optimal values from a similar model described at[^7]. It is suspected that the assignments are not adequate to due possible misrepresented information in the patent.

5. **Plate and Fin HX Modelling**

The models regarding the fluid and materials properties still need to be finished and the geometry models need to be done fitting the commercial technology of several suppliers. However, at this point all the major components of the sub-modelling structure is complete, which means that model is good enough to be applied to liquefaction processes, although, being the geometry and fluid properties asked as user assignment instead of modelled.

**Model Overview**

Figure 4, shows the structure of the PFHX model.

![Figure 4: Overview of the PFHX model.](image)

From a gPROMS user point of view, only the top-level model PFHX will be shown. This model is connected to a fluid property model, which basically allows the choice of the property package to use.

It is clear that the PFHX sub-model decomposition represents the major difficulty of the model, not only in terms of intrinsic modelling code, but also in terms of connections and connection types.

The connection between models was done by the means of ports. The boxes with grey background represent the composite models of the PFHX, while the boxes with white background represent auxiliary models.
Model Description

The figure below represents the topology environment of the PFHX model.

![Topology representation of the PFHX model](image)

Figure 5: Topology representation of the PFHX model.

In terms of sequence a stream (hot or cold) connected to the model at the flowsheet environment is also connected to the PFHX model trough a gMLMaterial port. Is then redirected through the splitter_array to its flow_multiplier, where the flow is divided equally by its number of channels. After leaving the channels, the divided stream is collected again into a single stream, using the other flow_multiplier, that multiplies the flow by its number of channels and send the stream to the mixer_array, where is directed to its respective outlet port.

Note that this procedure is done for all hot and cold streams.

The number of the channels is defined by the user and represents the number of in-between plates layers related to a certain stream. Assuming all the channels equal among its stream, only one channel per stream must be simulated. The flow_multiplier is used to reproduce the effect of the total number of channels per stream.

To set the model in a general way that could deal with all the possible hot and cold streams combination with all the wanted channels per stream, some of the models were grouped in array form (e.g. Flow_multipers and channel_1D are defined as arrays of cold and hot streams).

At figure 5, all the blue lines represent connections between gMLMaterial ports, while the green lines represent connections between gMLThermalContact_1D ports. These connections are of the major important because information about related entities of different models must be fulfilled to ensure the complete definition of the overall model.

Mixer_array and Splitter_array

The mixer_array and splitter_array models were created to avoid the definition of the Mixer and Splitter models in array form. So, in fact, these models represent respectively an array of mixers and an array of splitters.

As specification the user should define what inlet streams are mixed together in relation to different outlet streams.

Channel_1D

Along with the model wall_1D, this is the core of the PFHX model. This model describes one dimensional mass, momentum intrinsic phenomena, as well as one dimensional heat transfer with the wall_1D model.

In the 1D heat balance the following factors were considered: (i) Fluid convection in axial direction, (ii) Fluid conduction in axial direction and (iii) Fluid to wall convection.

In the 1D momentum balance the following factors were considered: (i) Inlet and outlet fluid momentum, (ii) Forces on fluid, (iii) Friction on the wall.

Note that channel_1D and wall_1D connection is used to pass common information between models needed to describe the fluid to wall convection term of the heat balance.

Wall_1D

The model wall_1D describes the heat transfer that occurs between channels.

Due to the PFHX compactness and construction is assumed that the heat balance on the perpendicular direction of the fluid is fast enough so that the profiles of the wall/plates are the same along all the PFHX.

The wall_1D represents a model of all the walls of the HX. The model describes the heat balance as a combined effect of all the walls.

In the 1D heat balance the following factor were considered: (i) Fluid conduction in axial direction, (ii) Wall to fluid convection, (iii) Thermal inertia effects along axial direction.

Heat_Losses_1D

The heat_losses_1D model describes the 1D heat losses through the wall.

Auxiliary Models

Some of these models are in fact the work in progress part of the PFHX model. The following explanations represent the goals to achieve and not the already implemented work.

6. LNG process modelling – complex approach

The general description presented in section 4 is applicable for this situation. The only differences to the simplified approach were the introduction of PFHX models as replacement of MSHX simple models and the introduction of heat_exchanger models to replace the cooling equipment models used at compression trains. This last
modification was only applied to the AP-X™ process because this issue is only relevant during optimisation where the models should work in performance mode, which in contrast to the cooler model the heat_exchanger has.

**AP-X™ Process**

In this situation, all the heat transfer models are used in performance mode, i.e. instead of the assignment of outlet temperatures the user specifies HX areas, heat transfer coefficients and flow direction. The outlet temperature of the streams is calculated using the LMTD method.

For the heat_exchanger models, the input area was obtained in the following way: (1) Define the required outlet temperatures as general input for heat transfer models, (2) Run the model in design mode with the previous specification, (3) Use the calculated area as new input in performance mode and finally (4) For the first initialisation use the initial outlet temperatures as initial guess.

The modifications were not significant in terms of LNG production. That is due to the inputs chosen for the PFHX model. Those were done in a certain way to match the results observed in the simple approach.

**AP-X™ alternative Process**

Except the for the PFHX models, all the assignments of the simple approach were also used for this case. Again, the assignments of the PFHX model were done in a way to match the results of the simple approach.

**C3/MR Process**

Except the for the PFHX models, all the assignments of the simple approach were also used for this case.

Again, the assignments of the PFHX model were done in a way to match the results of the simple approach.

**Results**

For all the situations the relevant results differ less than 0.1% running with the PR or GERG EOS, therefore, only the results with PR physical properties will be shown and analysed.

**Table 2: Complex approach – main results**

<table>
<thead>
<tr>
<th>Processes</th>
<th>AP-X™</th>
<th>AP-X™ alternative</th>
<th>C3/MR</th>
</tr>
</thead>
<tbody>
<tr>
<td>No of variables</td>
<td>10134</td>
<td>8588</td>
<td>4446</td>
</tr>
<tr>
<td>No of equations</td>
<td>8228</td>
<td>6955</td>
<td>3672</td>
</tr>
<tr>
<td>Total CPU time (s)</td>
<td>50</td>
<td>44</td>
<td>158</td>
</tr>
<tr>
<td>No of main equipment</td>
<td>33</td>
<td>12</td>
<td>20</td>
</tr>
<tr>
<td>NG inlet (kg/s)</td>
<td>118.3</td>
<td>118.3</td>
<td>98.9</td>
</tr>
<tr>
<td>LNG Produced (kg/s)</td>
<td>110.4</td>
<td>110.1</td>
<td>81</td>
</tr>
</tbody>
</table>

Compared to the approach described in section 4, the CPU effort is higher due to the complexity of the models. Note that AP-X™ is a recent process and unfortunately its modelling activity among the scientific community is still very scarce. In consequence the assignments used for the different processes are based in different literature sources, complicating their comparison.

Regarding the C3/MR process, results are an outcome of a mix between the PML:SS models (used as composited models of the top-level flowsheet) with results from [7] (used as assignments of the process). In fact, they represent a 20% deviation (in terms of specific power consumption - SPC) in relation to the OMAN-LNG plant [9].

This value could potentially be reduced with the performing of a steady-state optimisation.

Regarding the AP-X™ and the AP-X™ alternative process, the results observed are too distant from industrial reports. It is not possible to compare this model to any industrial case. Possibly patent information used as assignments in the models was very inadequate for the situation.

Due to this is useless to compare them with any other industrial records. The AP-X™ detailed analyses will be done further on.

7. **AP-X™ Process NLP optimisation**

Before optimisation takes place some decisions have to be made: (i) costs to consider for the overall capital cost of the plant, (ii) capital costs amortisation type and duration, (iii) operational costs to include into the analysis and (iv) decision factors as plant location and year of operation.

After solving the above describe issues, it is necessary to find that information in the literature and add that data to the costing section of the models whose capital and operational costs should be considered.

In terms of capital costs, the following equipment costs were considered for the overall fixed cost of the plant: (i) kettle HX cost, (ii) shell and tube HX cost, (iii) PFHX cost, (iv) centrifugal compressors capital costs, and (v) expander capital cost.

For the equipment amortization it was considered a linear amortisation during the 10 years life-time for the plant. The costs were update to the year of 2012.

As highly dependent plant of refrigeration cycles, the power consumption cost has a major role for the operational cost of the plant. Electricity cost is the variable considered for this case.
It was assumed that the plant was located in the south part of the state of Texas near the Gulf of Mexico. The operation year is 2012. The operational cost information was assumed as an average electricity price from several US companies.

**Methodology**

As told in section 6 all the AP-X™ process models are defined in performance mode. This mode allows the definition of design entities as heat transfer area. Performing the optimisation stage with the models in design mode, would allow the change of design entities of the model during the procedure. This is not correct when assuming the optimisation of an existing plant. Design characteristics of equipment are fixed and are not replaceable, which means that the performance mode assignment in the models is a mandatory condition.

Further on in the optimisation stage, the design variables of the equipment will also be changed, but keeping the performance mode of the models activated and assuming those as control variables.

The procedure taken follows an initial assumption that the optimisation is being applied to an existing plant, which means that only operational conditions should be improved. After finding the optimal performance solution that leads to the minimum total cost of the plant controlling only operational variables, an overall optimisation case will be done to find the optimal solution controlling operational and design variables of the LNG model.

The details of each optimisation will be presented below. For all cases, the objective function is the plant total cost, analysed in an hourly base.

The cases are constrained in the following way: (i) minimum LNG production obtained on the base case, (ii) no cross-over in the heat exchange equipment, (iii) minimum vapour fraction of 99% at the entrance of the compressors.

**AP-X™ process – base case**

The base case is the starting point of the optimisation stage. This model is the exact model described in section 6, with the inclusion of the equations and variables mentioned above.

The process presents a total hourly based cost of 24980 $/h.

**Sensitive Analyses**

The handmade optimisation step was result of a handmade simplified optimisation, regarding a combination of Propane and MR flows that would improve the total cost of the plant, meeting the constrains mentioned above using a trial and error approach. Note that from this case and beyond, an extra HX will be added for the nitrogen compression train. As a result the total cost was reduced to 19059 $/h.

Only with a handmade simplified optimisation a 24% reduction was observed in the total cost of the plant, showing that the base case conditions have room for improvement.

**Refrigerants flow optimisation**

From this step, all the optimisations were done using the optimiser tool of the gPROMS model builder.

The optimisation interface is a user-friendly tool that requires the definition of an objective function (to maximise or minimise), control variables and a set of constraints.

Note that the optimisation steps were made in sequence, this means that the previous optimisation case is the starting point for the next one.

For the present situation, the control variables are the flows of the three refrigeration cycles used in the process. As a result the total cost was reduced to 12624 $/h.

The flow optimisation represents a 50% cost reduction of the base case. This is indeed the major portion of the optimisation stage. This is expected because the total cost major quote is due to energy consumption of the compression trains which is closely related to the refrigerants flows.

**Performance optimisation**

Starting from the previous case, the total cost function is minimised, controlling the refrigerant flows and the MR composition.

This is the standard optimisation procedure used for LNG refrigeration cycles, especially the ones with mixed refrigerants.

The ideal composition has an important role with regards to the efficiency of the process, namely to deal with the required cooling range and to match the LNG cooling curve. This optimisation stage also included the MR composition as control variable beside the refrigerants’ flows.

As a result the total cost was reduced to 10740 $/h.

The present optimisation represents a 57% cost reduction of the base case.

**Overall optimisation**

The last optimisation step regards an overall optimisation of the main variables associated to the refrigeration cycles. Beside the flow and composition of the refrigerants, the following operational and design variables are optimised: (i) intercooler’s area for the compression trains and (ii) expansion pressure of nitrogen and MR cycles.

As a result the total cost was reduced to 10543 $/h.
The global optimisation represents a 58% cost reduction of the base case.

Results

The optimisation results show a clear positive evolution of the objective function. The overall optimisation step shows that the major optimisation parcel was already achieved in the previous steps, namely during flow optimisation.

Figure 6: Representation of the optimisation results in absolute and relative terms.

It is observed that the major part of the cost associated to the plant is due to energy consumption, as expected for a liquefaction plant.

The operational cost assumes an even major role because some factors associated to the capital cost depreciation might be undervalued, namely, the investment amortisation type, interest rate, fabrication type, among others. The capital cost is the sum of all equipment costs, excluding transportation.

In resemblance to the operational costs distribution, the major portion is associated to the compression train capital costs. Note that this value would be reduced with optimisation of the compression train. Issues like number of compressors, optimal pressure rations and number of parallel compression lines were not evaluated, the information used was based in [10].

8. AP-X™ Process Results

The following table presents the main results associated to the AP-X™ optimised process.

<table>
<thead>
<tr>
<th>Process</th>
<th>SMR</th>
<th>C3/MR</th>
<th>DMR</th>
</tr>
</thead>
<tbody>
<tr>
<td>SPC (kW-day/tonne LNG)</td>
<td>&gt;13.5</td>
<td>&gt;12.2</td>
<td>&gt;12.8</td>
</tr>
<tr>
<td>Total Cost ($/h)</td>
<td>10543</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

From the values shown, it is clear the improvement of the results from the section 4 and 6 until the final optimised model.

The optimisation not only improved the energy consumption of the plant, but also improved the LNG produced flow, 3.86 MTPA, as well as its richness in methane. It is possible to observe that 99.3% of the inlet methane was recovered as LNG.

One of benchmark of the LNG industry is the analysis of the specific power consumption associated to a process. See below typical values for different industrial applications of the LNG processes [9].

It is possible to observe a clear efficiency improvement during the last years. However, the specific power consumption is closely related with NG composition, environmental conditions, inlet NG conditions and LNG required composition, making LNG processes’ comparison, often a rough task.

The lack of information about AP-X™ industrial cases added to the issues cited above limit a proper comparison. Therefore, the optimised results can only be compared to C3/MR values.

According to Foster Wheeler [11], is possible to standardise the specific power consumption by LNG technology in the following way:

Table 4: Standard values of specific power consumption

<table>
<thead>
<tr>
<th>Process</th>
<th>SPC (kW-day/tonne LNG)</th>
<th>SMR</th>
<th>C3/MR</th>
<th>DMR</th>
</tr>
</thead>
<tbody>
<tr>
<td>Total Cost ($/h)</td>
<td>10543</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Being, SMR a single mixed refrigerant process and DMR a double mixed refrigerant process.

It is possible to see that the specific power consumption of the optimised AP-X™ process fits the usual values observed in industrial cases and show in the tables above.

Note that the execution of an optimisation procedure and posterior discussion for the C3/MR instead of the AP-X™ would make the thesis richer, since the C3/MR is a consolidated process with several modeling and optimising procedures with a lot of comparison models. However from my personal and PSE’s point of view it was better to perform the project in this way because it was a pioneer approach benefit in terms of business perspective.
9. Conclusions and Future work

For both simple and complex approach, the easy and user-friendly process assembling allows a fast top-level composition of the LNG process models, giving the user the opportunity of caring only with the engineering part of the work and letting all the mathematical work for the software and model developers.

In the simple approach the C3/MR process fits the typical values for industrial cases. This is due to the inputs of the C3/MR process that were based in optimal values from a similar model described at[7].

The advanced modelling of the 1D PFHX model is a stepped-up for PSE’s liquefaction processes including the present ones. A SEM type model was mandatory for this project, similar to the ones developed by Aspen® (Aspen plate fin exchanger™) and Linde® (GENIUS™).

The complex approach with the inclusion of the new PFHX increases the complexity of LNG models, becoming this approach a reliable representation of industrial environments with small computational effort.

Unfortunately, the results observed are still far from the industrial standard values due to the lack of good industrial data and to the uncertainty of the patent information often used as assignments.

After optimisation the AP-X™ process presents a SPC that fits the industrially observed values, showing that with the right combination of correct assignments the model would be validated.

For the future, great part of the work will be related to the development and upgrade of the PFHX model: (i) Generalisation of the PFHX initialisation procedure, (ii) Inclusion of property and geometry models for the PFHX (iii) Improvement of the PFHX design mode.

After the conclusion of the PFHX model, an industrial cooperation should be made in order to get correct and rigorous assigns to validate the LNG process models.

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References