Exergoeconomic analysis applied to the process of regasification of LNG integrated into an air separation process

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Abstract:
The exponentially increasing demand for natural gas and the growing market of liquefied natural gas (LNG) demand efficient processes for LNG regasification within import terminals. Usually, the regasification of LNG is accomplished through direct or indirect heating. However, integrating LNG regasification into a cryogenic process is an efficient approach because the low-temperature energy can be effectively utilized. One such options is the integration of LNG regasification into an air separation process. In some LNG important terminals in China, Japan, Taiwan and France similar integration technologies have already been implemented. Previous publications by the authors showed the increase in exergetic efficiency for the systems including air separation and LNG regasification. Also the variation in the exergetic efficiency depends on schematic and operation conditions. This fact creates great potential for improving. In this paper the different schematics will be evaluated from economic point of view, and finally an exergoeconomic analysis will be conducted in order to find out the options for energetic (exergetic) and cost improvement of the complex processes for air separation and LNG regasification.

The results show that a system with higher exergetic efficiency ($\varepsilon = 52.6\%$) has lower product costs, but lower quality of the products. Another system with lower exergetic efficiency ($\varepsilon = 44.2\%$) has higher product costs, but also the quality of the products is higher.

Keywords:
LNG, regasification, exergoeconomic analysis

1 Introduction

Natural gas is an important primary energy carrier. Its market share increased from around 10\% in 1970 to 24\% in 2012 [1] and is still continuing to grow. Also the demand for liquefied natural gas (LNG) is growing rapidly. The Asia and Asia Pacific regions have the highest demand for LNG, which accounted for 74\% of the total global demand in 2014 [2]. During the last 15 years, the number of countries with regasification terminals increased by a factor of three while the capacity within these terminals was doubled. In 2014, the 30th country has installed a regasification terminal [3]. The increasing number of LNG terminals is due to a growing spot market, flexible shipping strategies and the technological advances in the floating regasification terminals [3].

LNG is an alternative to pipeline gas, particularly if the distance between exporting and importing countries is relative high. In the importing country LNG is regasified and supplied to the national gas distribution system. Mostly LNG is regasified within open rack vaporizers (ORC) and submerged combustion vaporizers (SCV). In these systems cold exergy is destroyed without any use. Alternatively, LNG could be implemented in a process to generate electricity like a co-generation system [4] or a Rankine Cycle system [5]. A further option is the implementation of the regasification of LNG in a desalination process, which lowers the power consumption [6]. Introducing LNG in a process and using the cold energy offers a great potential for improving the original process from the thermodynamic and
cost points of view.

In this paper the regasification of LNG is accomplished within an air separation process. This technology is already used in different countries like China, Japan, Taiwan and France [7]. One big advantage is the lower power consumption, which could be up to 50% compared with a conventional air separation plant [7, 8]. Two different systems with regasification of LNG are analyzed from thermodynamic and economic points of view and compared with a conventional air separation system. For both systems with regasification of LNG an advanced exergy analysis was already applied [9, 10].

2 Process description

In the following sections the three simulated systems are described in detail. They have the same general structure and consist of four main groups of components: (1) air purification and compression block, (2) liquefaction of air in the main heat exchanger, (3) column block and (4) nitrogen liquefaction block. The column block is identical in each case and is embedded together with the main heat exchanger in a component, which is called "cold box". The column block consists of two columns which are thermally coupled by a condenser/reboiler. The lower column is the high-pressure column (HPC) and the upper column is the low-pressure column (LPC). Important parameters as the reflux ratio and the number of stages [11] as well as the pressure drop over the sieve trays [12] are considered in the design. The bottom product which is an oxygen-rich mixture as well as some side products of the HPC are throttled and fed to the low pressure column. At the top of the HPC almost pure nitrogen (gaseous and liquid) is leaving. The bottom product of the LPC is almost pure oxygen (gaseous and liquid) and the top product is almost pure nitrogen (gaseous). Additionally a side stream is leaving the LPC, which consists of non-condensable gases and is called purge gas stream.

2.1 Case A

Figure 1 shows the conceptual flowsheet of the single air separation plant. Before the air is sucked in at ambient conditions, dust particles are removed. The air is compressed within a two-stage compression process with interstage cooling to a pressure of 5.6 bar [11]. After the second air compressor (AC2) the air is cooled to almost ambient temperature and is fed to an adsorption block. Here, impurities which could freeze at very low temperature and consequently could destroy the system, must be adsorbed. In this simulation the adsorption block is not simulated in detail. According to [13] the content of water vapor and carbon dioxide must be less than 0.1 ppm and 1.0 ppm, respectively. Here, we simplified the simulation and removed the total amount of water vapor and carbon dioxide. Typically the adsorption block consists of two vessels which are working alternatively. The desorption of the adsorbed impurities is done by the purge gas stream, which is a waste stream from the column block. Finally, the purified air enters (stream 6) the main heat exchanger and cools to a temperature of -160°C and enters the column block (stream 10). As described above, several streams leave this block. The gaseous streams (oxygen (stream 20), purge gas (stream 25) and nitrogen from the HPC (stream 30) and LPC (stream 15)) are fed to the main heat exchanger. Here, they cool the incoming air in counter-current way. The nitrogen stream from the HPC is heated only to -90°C and then is fed to the nitrogen liquefaction block. This block consists of two heat exchangers (HE1 and HE2), four nitrogen compressors (NC1 - NC4), two interstage coolers (IC3 and IC4), two expanders (EXP1 and EXP2) and some mixing and splitting devices. The NC3 and NC4 are connected with the EXP1 and EXP2. The incoming nitrogen stream is mixed with two streams running in a loop in the nitrogen liquefaction block. Then this stream is heated and compressed to 38 bar [11] within the first three compressors. After the NC1 and NC2 this stream is cooled to approximately ambient temperature. After the NC3, the stream is split into two parts: one part is fed to the HE1, while the second part is compressed to 46 bar [11] and is also fed to the HE1. Both streams are cooled. The stream which was compressed to
38 bar is fed to EXP1 and is expanded to the pressure of the incoming stream of the liquefaction block. The second nitrogen stream, which was compressed to 46 bar, is split once again into two parts after the HE1. The smaller part of this stream is fed to the HE2, where it is cooled by the large part which has passed the EXP2 before. Finally, the smaller part (stream 48) leaves the nitrogen liquefaction block. Afterwards this stream is throttled (stream 49) and split in two parts (stream 58 and 59). Both parts enter one column, whereby the stream for the LPC must be throttled again. For both columns these streams are working as a reflux.

The product streams (gaseous nitrogen (stream 17) and oxygen (stream 22)) as well as the purge gas stream (stream 27) are leaving the main heat exchanger. Then, both product streams are compressed to 20 bar, but in general this value depends on the consumer. Finally both streams (stream 18 and 23) are used to heat the purge gas to 170 °C [14]. This stream is in turn used to desorb the impurities in the adsorption bed system.

### 2.2 Case AD1

A conceptual flowsheet of Case AD1 is shown in Fig. 2. The beginning of the process is the same as already described in Case A. The cleaned and compressed air (stream 5) is cooled in the main heat exchanger and fed to the column block (stream 10). The decisive difference is the implementation of the liquefied natural gas (LNG), which is regasified within the main heat exchanger. The LNG is brought into the system with a pump (stream 30) and is fed together with the gaseous product streams of the column block to the main heat exchanger. By cooling the air to liquefaction temperature the LNG stream (stream 61) as well as the nitrogen (stream 15), oxygen (stream 20) and purge gas stream (stream 25) are heated. Finally, the product streams are compressed and the purge gas stream is further heated. It is determined, that the LNG should leave the system with ambient temperature, thus the
LNG stream is heated with the nitrogen (stream 18), oxygen (stream 23) and a second nitrogen stream (stream 34) coming from the nitrogen liquefaction block. An other difference between Case A and Case AD1 is the structure of the nitrogen liquefaction block. In the latter, the nitrogen liquefaction block consists of three compressors (NC1-NC3), one expander, two heat exchangers (HE1 and HE2) and some mixing and splitting devices. One of the compressors is connected with the expander (EXP2). The top product of the high-pressure column (gaseous nitrogen, stream 30) is fed to the heat exchanger 2 (HE2) of the nitrogen liquefaction block. This is a three stream heat exchanger, where one stream is cooled by the incoming stream and an other stream. Afterwards both heated streams are mixed and fed to the heat exchanger 1 (HE1). Here, this mixed stream is heated further. Then it is compressed in a three-stage compression process to a pressure of 38 bar. After the first nitrogen compressor, the stream is cooled by heating the LNG and purge gas stream. Subsequently, it is fed to the second and third nitrogen compressors, cooled within the HE1 and split into two parts. One part is expanded in EXP2 where the temperature decreases and reaches the required low temperature for cooling, together with the incoming stream, the second part of the divided stream. This cooled stream (stream 49) leaves the nitrogen liquefaction block and is used as a reflux stream for both columns. It is throttled and split into two parts (stream 58 and 59). One part enters the top of the high-pressure column, the second part is throttled again to the pressure of the low-pressure column and enters it.

2.3 Case AD2

The conceptual schematic of Case AD2 is shown in Fig. 3. Compared to Case AD1, the structure of CAD2 differs in the implementation of the LNG. Additionally, the structure of the air compression and
purification as well as the nitrogen liquefaction block differ in Case AD2 compared to Cases A and AD1.

The dustless air enters the air compression and purification block and is compressed in a three-stage compression process to 5.6 bar. Between the compressors the air is cooled. Here, the LNG stream is used as a cooling medium instead of water. Then the air is fed to the main heat exchanger (stream 8) and is cooled. Afterwards it enters the column block (stream 10).

The LNG stream is brought into the system with a pump (stream 39) and afterwards is split into two parts (stream 41 and 47). One part is fed to the MHE (stream 47) where it is heated, while the second part (stream 41) is fed to the nitrogen liquefaction block and is heated in the HE2. Afterwards both streams are mixed (stream 43) and further heated in both interstage coolers and the heat exchanger 4 (HE4) to ambient temperature.

The top product of the low- and high-pressure column (both gaseous nitrogen, streams 15 and 30) are fed to the HE2. This heat exchanger and the two nitrogen compressors (NC1 and NC2) constitute the nitrogen liquefaction block. Within the HE2 the LNG stream and both nitrogen streams coming from the columns are heated. While the nitrogen stream coming from the low-pressure column (stream 15) is fed to the main heat exchanger, the nitrogen stream coming from the high-pressure column (stream 30) is compressed within NC1 and NC2 and is subsequently cooled within the HE2. This stream (stream 34) leaves the nitrogen liquefaction block and is throttled. Finally, it is split into two parts, one part is fed to the high-pressure column (stream 58) and the second part is further throttled and fed to the low-pressure column (stream 29).

The product streams leaving the MHE (stream 17 and 22) are compressed and used to heat the purge gas (HE3) and the LNG stream (HE4).
3 Simulation and energy analysis

The simulation of the three systems was carried out using ASPEN PLUS [15]. For conducting the simulations some parameters need to be assumed. Thus, for the turbomachines we set the following:

- The isentropic efficiencies of the compressors and the expanders are 84%.
- The isentropic efficiency of the pump is 70%.
- The mechanical efficiency of all turbo-machines is 99%.

Figure 4 shows the varying power consumption/generation for the three systems. The total power consumption for Case A is $W_{\text{net,A}} = 17.9$ MW. Both systems with the implementation of the regasification of LNG have a lower power consumption, $W_{\text{net,AD1}} = 11.2$ MW and $W_{\text{net,AD2}} = 7.2$ MW for Cases AD1 and AD2, respectively. This corresponds to a decrease of the energy consumption by 37% from Case A to Case AD1 and a decrease of 36% from Case AD1 to Case AD2.

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{figure4.png}
\caption{Comparison of the power consumption/generation for the turbo-machines}
\end{figure}

4 Exergy Analysis

Before applying the exergy analysis for the three systems, average European environmental conditions are assumed: $T_0 = 15$ °C and $p_0 = 1.013$ bar. For each component the exergy of fuel and of product are defined according to [16] and the exergy destruction for the overall system (equation 1) and for the components (equation 2) is calculated as:

\[ \dot{E}_{F,\text{tot}} = \dot{E}_{P,\text{tot}} + \dot{E}_{D,\text{tot}} + \dot{E}_{L,\text{tot}} \]  

(1)

and

\[ \dot{E}_{F,k} = \dot{E}_{P,k} + \dot{E}_{D,k} \]  

(2)
Some components are treated as dissipative components such as the interstage coolers in Cases A and AD1 and the mixing devices in all three systems. In some components the process occurs by crossing the temperature of the environment. In this case it is necessary to split the physical exergy of each stream into its thermal and mechanical part [17]. Important results of the exergy analysis are shown in Table 1 and Fig. 5.

**Table 1. Results obtained from the exergy analysis of the overall system (Cases A, AD1 and AD2)**

<table>
<thead>
<tr>
<th></th>
<th>$\dot{E}_{F,tot}$, MW</th>
<th>$\dot{E}_{P,tot}$, MW</th>
<th>$\dot{E}_{D,tot}$, MW</th>
<th>$\dot{E}_{L,tot}$, MW</th>
<th>$\varepsilon_{tot}$, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Case A</td>
<td>18.0</td>
<td>5.9</td>
<td>11.8</td>
<td>0.22</td>
<td>32.9</td>
</tr>
<tr>
<td>Case AD1</td>
<td>21.1</td>
<td>9.3</td>
<td>11.6</td>
<td>0.12</td>
<td>44.2</td>
</tr>
<tr>
<td>Case AD2</td>
<td>17.1</td>
<td>9.0</td>
<td>8.0</td>
<td>0.12</td>
<td>52.6</td>
</tr>
</tbody>
</table>

**Fig. 5. Exergy destruction of the components in Cases A, AD1, AD2**

### 5 Economic analysis

The economic analysis was performed using the total revenue requirement method ($T_{RR}$) [16]. For the economic analysis it is important to estimate the costs of the components as accurately as possible, because the additional costs are determined as a percentage share of the purchased equipment costs ($PEC$). For the estimation of the costs, the components are divided into the following groups: columns, heat exchangers and turbomachines (compressors, expanders and pump). All components where the temperature of the working fluid is lower than -29°C are made of stainless steel. Components with a temperature higher than -29°C are produced of carbon steel [18]. In addition, also a temperature and a pressure factor are considered for all components, because especially the low temperatures affect the cost of the components.
5.1 Costs of the columns

The estimation of the costs of the columns is divided into two parts. First of all the costs for the empty shell for the high-pressure and low-pressure column are determined followed by the estimation of the costs for the trays. Both columns are simulated as a sieve tray column with 54 stages and 96 stages [11] for the HPC and LPC, respectively. The costs for the shell of the columns depend on the height and the diameter of each column. According to [12, 19, 20] the average distance between two trays lies between 80 and 300 mm, 300 and 600 mm and around 610 mm, respectively. We assumed a distance of 400 mm, which results in a height of 21.6 m (HPC) and 38.4 m (LPC). Hence, the total height of both columns is 60 m which corresponds to the typical height of the cold box [20]. The diameter is assumed to be 3 m for each column, because a diameter up to 4-5 m is costly for sieve trays and could cause problems associated with holding the liquid phase [21]. The costs for the shell are determined by a cost chart [22], as a function of the height and the diameter. For the trays the costs also depend on the diameter of the column and are also determined by a cost chart [22].

5.2 Costs of the heat exchangers

The interstage coolers and the HE3 and HE4 are shell and tube heat exchangers, whereas the MHE, HE1 and HE2 are plate heat exchangers. To estimate the area of each heat exchanger the heat duty is taken from the Aspen Plus simulation. Additionally the logarithmic mean temperature is calculated based on the results within the simulations and the overall heat transfer coefficient is assumed based on the literature [23]. For the interstage coolers \( U = 42.5 \text{ W/m}^2\text{K} \), for the MHE \( U = 70 \text{ W/m}^2\text{K} \), for the HE1 and HE2 \( U = 50 \text{ W/m}^2\text{K} \) and for the condenser/reboiler \( U = 2000 \text{ W/m}^2\text{K} \) are assumed. For all heat exchangers the costs are determined by a cost chart [24].

5.3 Costs of the turbomachines

This group contains the air and nitrogen compressors and the oxygen compressor. All compressors are centrifugal. The air compressors and the nitrogen compressors 1 and 2 as well as the oxygen compressor are driven by an electrical motor. The nitrogen compressors 3 and 4 are driven by expanders 1 and 2, respectively. The costs for the compressors depend on the power consumption and the degression exponent [25].

The two expanders in the nitrogen liquefaction block are axial expanders. The costs are estimated by a cost chart [22], where also the generated electricity is the defining factor.

In Cases AD1 and AD2 the pump for the incoming LNG is a centrifugal pump. The costs are estimated by a cost chart [24]. Also for these costs, the power consumption is the defining factor.

Table 2 shows the purchased equipment costs for all groups of components for the three simulated cases. The estimation of the total capital investment (TCI) is shown in Table 3. For the determination of the TRR the assumptions that need to be made are, shown in Table 4. Finally the levelized values of the carrying charges, the operating and maintenance costs and the fuel costs are calculated. The TRR is the sum of these three levelized costs.

6 Exergoeconomic analysis

The exergoeconomic analysis combines an exergy analysis and an economic analysis. With this method a system could be optimized from the cost point of view [16].

Analogue to the exergy analysis, here for each system component the cost of fuel and cost of product
Table 2. Estimated purchased equipment costs and bare module costs

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>CB</td>
<td>4.5</td>
<td>4.6</td>
<td>4.4</td>
</tr>
<tr>
<td>HE</td>
<td>2.5</td>
<td>1.1</td>
<td>0.3</td>
</tr>
<tr>
<td>Comp.</td>
<td>7.0</td>
<td>6.7</td>
<td>8.3</td>
</tr>
<tr>
<td>Expander</td>
<td>2.8</td>
<td>1.3</td>
<td>0</td>
</tr>
<tr>
<td>Pump</td>
<td>0</td>
<td>0.03</td>
<td>0.03</td>
</tr>
<tr>
<td>CDA</td>
<td>2.5</td>
<td>2.5</td>
<td>2.5</td>
</tr>
<tr>
<td>tot</td>
<td>19.3</td>
<td>16.3</td>
<td>15.6</td>
</tr>
</tbody>
</table>

Table 3. Estimation of the fixed capital investment for the Cases A, AD1 and AD2

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>A. Direct costs</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Total purchased equipment costs ((PEC))</td>
<td>19.3</td>
<td>16.3</td>
<td>15.6</td>
</tr>
<tr>
<td>Purchased-equipment installation ((33% of PEC))</td>
<td>6.4</td>
<td>5.4</td>
<td>5.2</td>
</tr>
<tr>
<td>Piping ((35% of PEC))</td>
<td>6.8</td>
<td>5.7</td>
<td>5.5</td>
</tr>
<tr>
<td>Electrical equipment and materials ((13% of PEC))</td>
<td>2.5</td>
<td>2.1</td>
<td>2.0</td>
</tr>
<tr>
<td><strong>Total direct costs</strong></td>
<td>35.0</td>
<td>29.5</td>
<td>28.3</td>
</tr>
<tr>
<td><strong>B. Indirect costs</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Engineering and supervision ((8% of DC))</td>
<td>2.80</td>
<td>2.4</td>
<td>2.3</td>
</tr>
<tr>
<td>Construction costs and contractor’s profit ((15% of DC))</td>
<td>5.5</td>
<td>4.4</td>
<td>4.2</td>
</tr>
<tr>
<td>Contingency ((15% of the above sum))</td>
<td>6.5</td>
<td>5.4</td>
<td>5.2</td>
</tr>
<tr>
<td><strong>Total indirect costs</strong></td>
<td>14.5</td>
<td>12.2</td>
<td>11.7</td>
</tr>
<tr>
<td><strong>Fixed capital investment</strong></td>
<td>49.5</td>
<td>41.7</td>
<td>40.0</td>
</tr>
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</table>

Table 4. Parameters and assumptions used in the calculation of the TRR

<table>
<thead>
<tr>
<th>parameter</th>
<th>unit</th>
<th>value</th>
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<tbody>
<tr>
<td>plant economic life</td>
<td>a</td>
<td>20</td>
</tr>
<tr>
<td>effective interest rate</td>
<td>1/a</td>
<td>0.1</td>
</tr>
<tr>
<td>average general inflation rate</td>
<td>1/a</td>
<td>0.025</td>
</tr>
<tr>
<td>average general escalation rate of electricity</td>
<td>1/a</td>
<td>0.035</td>
</tr>
<tr>
<td>full load hours</td>
<td>h/a</td>
<td>8000</td>
</tr>
<tr>
<td>electricity price (2017)</td>
<td>$/MWh</td>
<td>90</td>
</tr>
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</table>
Table 5. Calculation of the total revenue requirement (TRR)

<table>
<thead>
<tr>
<th></th>
<th>Case A</th>
<th>Case CAD1</th>
<th>Case CAD2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Levelized carrying charges ($CC_L$)</td>
<td>6.8</td>
<td>5.7</td>
<td>5.5</td>
</tr>
<tr>
<td>Levelized annual O&amp;M costs ($OMC_L$)</td>
<td>2.4</td>
<td>2.0</td>
<td>1.9</td>
</tr>
<tr>
<td>Levelized annual fuel costs ($FC_L$)</td>
<td>18.2</td>
<td>11.4</td>
<td>7.4</td>
</tr>
<tr>
<td>Total revenue requirement ($TRR_L$)</td>
<td>27.4</td>
<td>19.1</td>
<td>14.8</td>
</tr>
</tbody>
</table>

need are defined. Additionally, the cost balances of the components as well as for the overall system (Equation 3) are set. Furthermore auxiliary equations for all components need to be defined.

\[
\dot{C}_{P,tot} = \dot{C}_{F,tot} + \dot{Z}_{Cl,tot} + \dot{Z}_{OM,tot} \quad (3)
\]

In this paper, the cost of each stream is split into three parts: the cost rates associated with the chemical exergy, the thermal exergy, and the mechanical exergy.

Some results for the overall system and the components of the system of the exergoeconomic analysis are shown in Table 6 and Figs 6-8. For the Figs only the most important components from the exergoeconomic point of view are selected. The results for the product streams are shown in Figs 9.

Table 6. Results obtained from the exergoeconomic analysis of the overall systems

<table>
<thead>
<tr>
<th></th>
<th>$\dot{C}_{F,tot}$, $$/h</th>
<th>$\dot{C}_{P,tot}$, $$/h</th>
<th>$\dot{C}_{D,tot}$, $$/h</th>
<th>$\dot{C}_{L,tot}$, $$/h</th>
<th>$c_F$, $$/MJ</th>
<th>$c_P$, $$/MJ</th>
<th>r, -</th>
<th>f, -</th>
</tr>
</thead>
<tbody>
<tr>
<td>Case A</td>
<td>2,285</td>
<td>3,247</td>
<td>1,505</td>
<td>153</td>
<td>0.04</td>
<td>0.15</td>
<td>3.13</td>
<td>0.43</td>
</tr>
<tr>
<td>Case AD1</td>
<td>1,604</td>
<td>2,371</td>
<td>885</td>
<td>193</td>
<td>0.02</td>
<td>0.07</td>
<td>2.34</td>
<td>0.52</td>
</tr>
<tr>
<td>Case AD2</td>
<td>1,096</td>
<td>1,936</td>
<td>510</td>
<td>87</td>
<td>0.02</td>
<td>0.06</td>
<td>2.35</td>
<td>0.64</td>
</tr>
</tbody>
</table>

7 Discussion

Figure 4 shows the power consumption/generation associated with the turbomachines. It is apparent, that the power consumption decreases from Case A to Case AD1 and to Case AD2. Consequently, the introduction of the regasification of LNG has a positive effect on the total power consumption. Especially the power consumption of the first nitrogen compressor could be reduced by more than 90 %. Altogether, the power consumption decreases by 60 % from Case A to Case AD1 which corresponds to the literature, where a drop of approximately 50 % [7] is mentioned, if the regasification of LNG is combined with an air separation plant.

The results of the exergy analysis are only briefly discussed; for more details see [9]. Comparing the results of the exergy analysis for the overall system shown in Table 1, the exergetic efficiency increases from Case A (32.9 %), to Case AD1 (44.2 %) and finally to Case AD2 (52.6 %). For the Case AD2 this is an increase of 60 % in comparison to a conventional air separation plant. The increase in the efficiency between CAD1 and CAD2 could be followed back to the lower exergy of fuel in CAD2, while the exergy of product almost remains constant. The exergy of fuel decreased due to lower power consumption in CAD2.

Table 2 shows the purchased equipment costs for the three analyzed systems. According to [26], the equipment costs of each group of components make up for a specified percentage share of the total costs of components. Table 7 shows this comparison. The first four entries show, that there is just a small difference comparing the percentages from the literature and the estimated values of the
purchased equipment costs. The estimated $PEC$ for the heat exchanger fits best for the given share. The costs for the adsorption block are estimated using the given share of 13% of the total purchased equipment costs of Case A for the cleaning and drying of the air.

**Table 7. Share of the components of the total purchased equipment costs**

<table>
<thead>
<tr>
<th>Components</th>
<th>Literature, %</th>
<th>Calculated, %</th>
</tr>
</thead>
<tbody>
<tr>
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<td>etc.</td>
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The calculation of the total capital investment cost ($TCI$) is based on the estimation of the purchased equipment costs. The $TCI$ is lower in the two systems with the integration of LNG than in the conventional air separation plant. For the Case A the TCI amounts approximately to US$ 58 million, for the Case A1 to US$ 43 million, and for the Case A2 to US$ 37 million. When comparing the estimation of the total capital investment for the conventional air separation plant with literature values, we find a good agreement. In [27] it is mentioned that an air separation plant with a capacity of 160 t/d gaseous oxygen was built for approximately 40 million € (2014). Calculating the specific costs per produced gaseous oxygen (GOX), leads to a similar value.

Finally, the results of the exergoeconomic analysis are evaluated. As shown in Table 6 the cost rate associated with the exergy destruction of the overall system decreases from Case A, to AD1 and to AD2. The cost rate associated with the exergy loss increases from CA to CAD1 and decreases finally from CAD1 to CAD2. Nevertheless, the specific product cost has the lowest value for Case AD2.

Figures 6 - 8 show the summation of the cost rates associated with the capital investment and the operating and maintenance costs for each component ($\dot{Z}_k$) and the cost rate associated with the exergy destruction within each component ($\dot{C}_{D,k}$) for Cases A, AD1 and AD2, respectively. In Case A (Fig. 6), the HE1 has the highest $\dot{Z}_k + \dot{C}_{D,k}$ value, followed by the MHE1, the IC3 and the HE2. Within all these components the share of $\dot{C}_{D,k}$ is much greater, than the share of $\dot{Z}_k$. This means that these components need to be improved from the thermodynamic point of view, resulting in a higher exergetic efficiency and consequently in a lower cost rate associated with the exergy destruction. Comparing the results for Cases A and AD1 (Figure 7), it is obvious, that the values of $\dot{Z}_k + \dot{C}_{D,k}$ could be significantly reduced.
Comparing the results in detail, the value for the MHE1 is reduced by half and the value for the HE1 decreases by a factor of ten. In the Case AD2 (Fig. 8) the MHE1 is the component with the sixth highest value, whereas in the two other cases this component has either the highest (CAD1) or the second highest value (CA). Moreover the total cost rate associated with the MHE1 is reduced again by half from CAD1 to CAD2. On the contrary the value of $Z_k + \dot{C}_{D,k}$ increases for the column block from CAD1 to CAD2. While in CAD1 the cost rate associated with the investment cost of the column block dominates the total value, in CAD2 both parts of the total costs are nearly distributed in equal parts. The reason for this noticeable difference is the higher exergy destruction within the column block in CAD2. This in turn could be explained by the lower purity of the gaseous and liquid oxygen streams.

Figure 9 shows the resulting costs for the product streams. In Case A the specific costs for liquid oxygen (LOX) are the highest, followed by the specific costs for liquid nitrogen (LIN). The liquid products are much more expensive, because they are more valuable. Regarding the gaseous products, the specific costs for gaseous oxygen (GOX) are higher than for gaseous nitrogen (GAN). For Case AD1 the general tendency is the same as in Case A. The specific costs for gaseous and liquid oxygen decrease slightly. In contrast, the specific cost for GAN increases while the cost for LIN decreases stronger.

Finally, the costs of the product streams in CAD2 are analyzed. In this system the specific cost for each stream decreases, except the liquid nitrogen stream. For LIN the costs increases slightly. Noticeable is, that the difference between the specific costs for both gaseous products as well as the
Fig. 9. Specific costs of the four product streams for each case studied here (On top of each column the corresponding purity $x_i$ is given)

difference between both liquid products is relatively small. Moreover, the specific costs for GOX are reduced approximately by half, which is due to the lower purity of the gaseous oxygen stream ($x_{GOX,AD2} = 0.78$) in CAD2 in comparison to the two other cases ($x_{GOX,A/AD1} = 0.978$). The same explanation can also be applied for the liquid streams, because also here the liquid oxygen stream has a lower purity ($x_{LOX,AD2} = 0.92$) compared to Cases A and AD1 ($x_{LOX,A/AD1} = 0.985$). This results in lower specific costs which also results in the lower difference between the two liquid products.

The costs for the streams of natural gas which leaves both systems are approximately the same as for the incoming stream of liquefied natural gas ($c_{LNG,in} = 0.2568 \$/kg$ and $c_{NG,out} = 0.2571 \$/kg$).

8 Conclusion

In this paper an air separation plant and two schematics including the integration of the regasification of LNG are analyzed. The exergetic and exergoeconomic analyses of the different schematics show significant differences, which underlines the necessity of this research. The major advantage of the exergy-based methods is the highlighting the irreversibilities from thermodynamic point of view. Additionally, the exergoeconomic analysis demonstrates, the cost effectiveness of the systems being evaluated.

The exergetic efficiency increases with the integration of the regasification of LNG. Additionally, the investment costs and the total revenue requirements are estimated for the three analyzed processes. The investment costs decrease if the regasification of LNG is integrated into an air separation process. The comparison of the two processes with the integration of the regasification of LNG shows that on the
one hand the Case AD2 has lower specific product costs and a higher exergetic efficiency, but the quality of the products is relatively low. On the other hand the quality of the products is higher in CAD1, but nevertheless this system has a lower exergetic efficiency and higher specific costs compared with CAD2. In future research we will also consider safety issues, that might lead to structural changes.

**Acknowledgments**

Tatiana Morosuk gratefully acknowledges the financial support from the "Berliner Programm zur Förderung der Chancengleichheit von Frauen in Forschung und Lehre".
Nomenclature

\[ \dot{C} \] cost rate \$/h
\[ c \] specific costs \$/kg
\[ E \] exergy rate kW
\[ f \] exergoeconomic factor -
\[ FCI \] fixed capital investment US$
\[ p \] pressure bar
\[ PEC \] purchased equipment cost US$
\[ r \] exergoeconomic factor -
\[ T \] temperature °C
\[ TCI \] total capital investment US$
\[ TRR \] total revenue requirement US$
\[ W \] power kW
\[ \dot{Z} \] cost rate associated with the investment cost of a component \$/h

Greek symbols:
\[ \varepsilon \] exergetic efficiency %

Subscripts and superscripts:
0 environmental conditions
CA Case A
CDA cleaning and drying of the air
D destruction
CAD1 Case A Design 1
CAD2 Case A Design 2
F fuel
i i-th chemical substance
k k-th component
L levelized (economic analysis)
L loss (exergetic analysis)
net netto
P product
tot total

Abbreviations:
AC air compressor
CB column block
EXP expander
GAN gaseous nitrogen
GOX gaseous oxygen
HE heat exchanger
IC interstage cooler
LIN liquid nitrogen
LNG liquefied natural gas
LOX liquid oxygen
MHE main heat exchanger
NC nitrogen compressor
NG natural gas
OC oxygen compressor
LNGP LNG pump
TV throttling valve
References


