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Chapter 1

Energy Recovery from the LNG Regasification Process

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Additional information is available at the end of the chapter

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Abstract

The global request of natural gas (NG) is continuously increasing, consequently also the regasification of liquefied natural gas (LNG) is becoming a process largely employed. Liquefied natural gas at a temperature of around 113 K at atmospheric pressure has to be regasified for its transportation by pipeline. The regasification process makes the LNG exergy available for various applications, particularly for the production of electrical energy. Different possibilities to exploit the thermal energy released during regasification are available. New plant configurations whose functioning does not constrain the processes of the regasification terminal are proposed. A possible solution is LNG exploitation as a cold source for ocean thermal energy conversion (OTEC) power plants. Electric energy can be produced also by the exploitation of heat released from hot sources, for instance, the condensation heat of power plants by means of consecutive thermodynamic cycles. The rational use of the cold source (LNG) allows the increment of electrical production and growth of the thermodynamic efficiency, with corresponding environmental benefits.

Keywords: LNG regasification, exergy analysis, OTEC systems, thermodynamic analysis, ammonia

1. Introduction

The natural gas (NG) can be converted in liquefied natural gas (LNG) in order to make easier its storage and transport. The exact composition of natural gas, and the LNG formed from it, varies according to its source and processing history; usually a percentage ranging from 85 to 95 is represented by pure methane. By cooling the natural gas to about 113 K (−160°C), it condenses into LNG, with a correspondent volume reduction factor of 600. In function of the

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LNG ship capacity, the carried gas could be enough to heat almost 43,000 homes per year [1]. From an energy point of view, in fact, the LNG energy content is equal to about 50 MJ/kg, corresponding to 21.5 GJ/m^3 by considering that at 113 K, the mass is 430–450 kg m^{-3}. Moreover, the regasified methane has a mass of about 0.71 kg m^{-3} in normal conditions, and its energy content is about 35 MJ/m^3. LNG is usually transported by ship to dedicated terminals and then stored at atmospheric pressure in superinsulated tanks; successively, LNG is heated and converted into gaseous form in order to supply the pipeline system.

From a historical point of view, the first regasification process operating with LNG began operation in 1917 in West Virginia, in order to produce stockpile helium as part of a research program. The first commercial plant, instead, was built in 1941 for peak-shaving purposes, by exploiting the stored liquefied natural gas as a strategic reservoir for future usage. The first LNG ship has left the Louisiana for the United Kingdom in the year 1959, demonstrating the feasibility of LNG transport, by stimulating the spread of large-scale commercial LNG projects. Five years later, in fact, the United Kingdom also began importing Algerian LNG. Nowadays, other destinations are reached such as Japan, Europe, South Korea, China, and countries with developing economy. The United Kingdom imported LNG until 1990 because the exploitation of gas extracted from the North Sea became a less expensive alternative. Japan is another country with a massive employment of LNG, importing it mainly from Alaska. The heavy expansion of LNG imports made Japan the forefront of the LNG worldwide trade in the 1970s and 1980s. The oil embargo happened in 1973 allowed a quick spread of natural-gas-fired power generation in Japan that generated a pollution reduction. The United States began to import LNG first from Algeria in the 1970s; successively, the rising of natural gas prices led to a rapid growth of the internal natural gas supply. Due to the rapid increment of the internal demand in the 1990s, a strong growth in US LNG deliveries was observed. More recently, a liquefaction plant was realized in the closest part of Trinidad and Tobago and that became the main US LNG supplier. Port Pelican in the Gulf of Mexico is the first example of offshore re-gasification facility realized in 2004, with the intention to reduce the land-use requirement [1, 2].

Currently, natural gas feeds about 21% of global primary energy demand. LNG trade in 2013 was about 10% of the total natural gas consumption. Although North America and Europe are the major users of natural gas, other countries as East Asia and Caribbean nations cover the majority of natural gas demand with LNG regasification. Other countries such as Taiwan, South Korea, Japan, Chile, Puerto Rico, Dominican Republic, Spain, and Portugal cover more than 50% of their gas requirements by means of LNG, and the percentage reaches 100% in the countries such as Taiwan, South Korea, and Puerto Rico [2] (**Figure 1**).

Generally, in the 2000s, the LNG demand has increased strongly because the number of countries that import LNG has tripled, and at the same time, the regasification capacity has doubled. Some aspects such as the possibility to access flexible shipping strategies, the growth of the spot market, and floating regasification technology have allowed the large diffusion of different LNG supply options, especially in countries with an emergent economy. The same advantages are exploited by export regions located in the Middle East and for countries with an increment of their energy needs (Asia Pacific and Latin America), as well as in countries that require a diversification of the primary energy sources (Italy) [3] (**Figure 2**).

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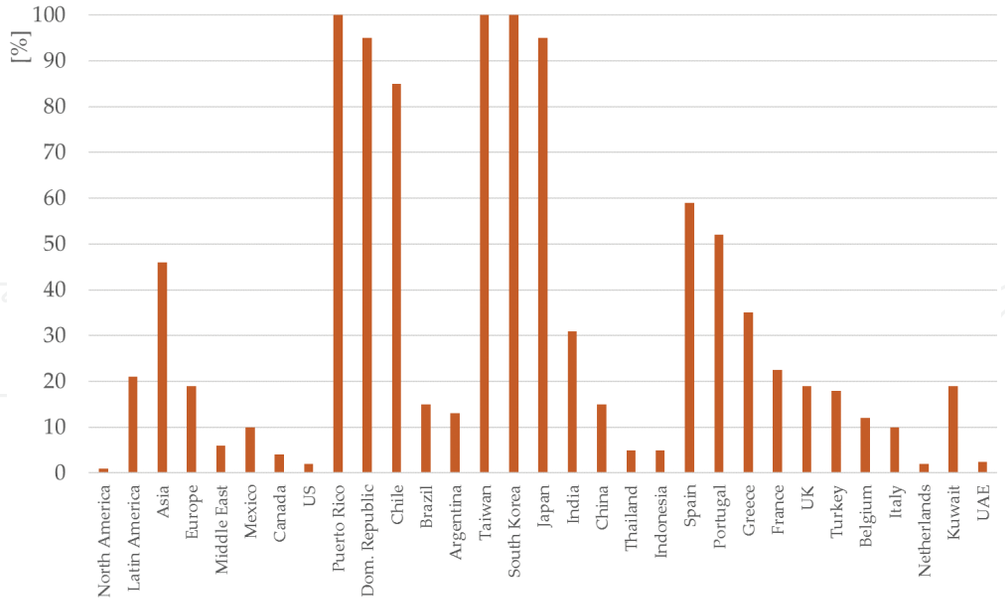


Figure 1. LNG% share of gas demand.

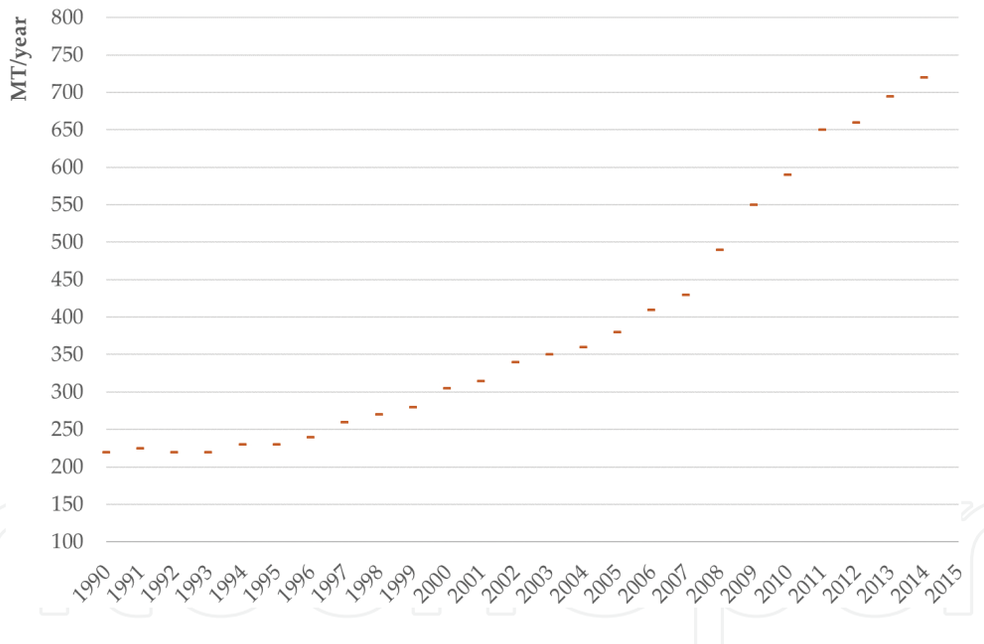


Figure 2. Regasification capacity trend.

In recent years, in comparison to the traditional way of natural gas transportation (pipelines), the LNG market has become more attractive in relation to the flexibility in its transportation and in terms of destination choices [2].

In Europe, the high management and realization costs of pipeline infrastructures determine a substantial regime of oligopoly, contrarily the transportation of LNG via ship allows for numerous advantages, such as [4]:

- Reduced dependence on a limited number of countries, taking advantage from a wide geopolitical distribution of the supply countries;
- The realization of gas pipelines requires long time, while the LNG via ship allows for the access to the market in a short time span;
- The diversification of the involved technologies provides a greater supply safety.

The transportation of natural gas via pipelines is cost-competitive for distances lower than 2500 km and for inland destinations; however, severe geological and political constraints could be a strong limitation especially in the case of cross-border trade [2].

In the twenty-first century, LNG will become an essential energy resource for human beings. Therefore, the modalities to recover the LNG cold energy for power generation system represent an interesting field. Several methods to recover LNG exergy can be employed in power generation systems, and each one has their own characteristics and results suitable for different utilization. Moreover, an energy intensive process to liquefy the natural gas is still required; therefore, the identification of technological solutions that allow for a partial recovery of the employed energy is recommended. Natural gas, or other products obtained in the liquefaction plant, is the primary source employed to start the liquefaction process. The LNG energy content is notable; therefore, the recovery of a part of the energy in the regasification process is reasonable. Frequently, this energy is not recovered because it is dispersed in the seawater, which constitutes an abundant and free hot thermal source to drive the regasification process. The LNG cold energy could be recovered by multiple processes, which depend upon the technologies used for its regasification. Moreover, LNG could be seen as a cold thermal source for the direct production of electrical energy, by means of consolidated technologies [5].

2. The regasification process: the exergy analysis

The regasification process happens in dedicated terminals (onshore or offshore) by employing different steps: from gas tankers, where natural gas is transported at atmospheric pressure in a state of saturated liquid and with a temperature of around 113 K, LNG is pumped to the tanks located in the terminal. Successively, by means a compression in liquid phase and a heating to environment temperature, the natural gas is brought in gaseous form. The vaporization process is generally obtained providing thermal load by using seawater or by burning a small part of the regasified LNG; the required thermal load, in fact, is about 1.5–1.7% of the LNG energy content. A considerable quantity of energy can be recovered during the LNG vaporization process by exploiting the exergy of LNG, defined as the maximum mechanical work obtainable starting from its temperature and pressure and in function of external temperature and pressure. Therefore, the LNG exergy content is determined as the sum of two contributions [5]:

- the first in function of the temperature difference between LNG and external environment (T_0):

$$ex_{th} = ex(p, T) - ex(p, T_0) \tag{1}$$

- the second is determined in function of the pressure difference between LNG and external environment (p_0):

$$ex_p = ex(p, T_0) - ex(p_0, T_0) \tag{2}$$

Due to the phase change and the successive heating to the external temperature, the thermal exergy can be quantified with the relation:

$$ex_{th} = \left(\frac{T_0}{T_{av}} - 1\right) \cdot r + \int_{T_0}^{T_{av}} c_p \cdot \left(1 - \frac{T_0}{T}\right) dT \tag{3}$$

where T_{av} is the mean LNG temperature during the transformation, r is the vaporization latent heat, and c_p is the LNG specific heat at constant pressure. Instead, the following relation evaluates the pressure exergy:

$$ex_p = \int_{p_0, T_0}^{p, T_0} v\, dp \tag{4}$$

Regarding the pressure contribution, the increment of the phase change temperature and the correspondent latent vaporization heat decrement lead to a reduction of the thermal exergy, contrarily the pressure exergy increases. The first contribution prevails on the second; therefore, the total exergy reduces with the pressure growth. The total exergy decreases slightly for pressures greater than 1.5 MPa, reaching a value of around 0.8 MJ/kg (**Figure 3**).

Depending on the environment temperature, the pressure and thermal exergies increase reaching a value of total exergy equal to 0.8 MJ/kg for a value of 280 K (**Figure 4**).

The exergy trend suggests that the employment of a recovery system with different cycles, which exploits both pressure and thermal exergies, is preferable. For instance, pressure exergy can be recovered with the "direct expansion" technique operating at high pressures, whereas thermal exergy can be recovered by means of a Rankine cycles operating with low boiling fluids.

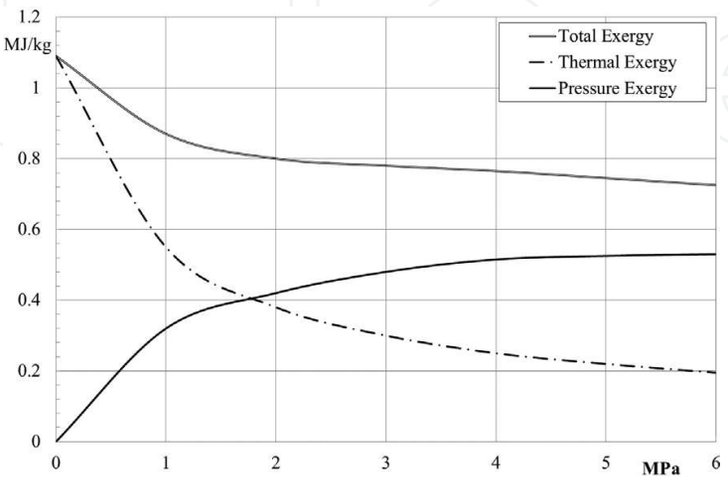


Figure 3. Exergy trend in relation to the regasification pressure.

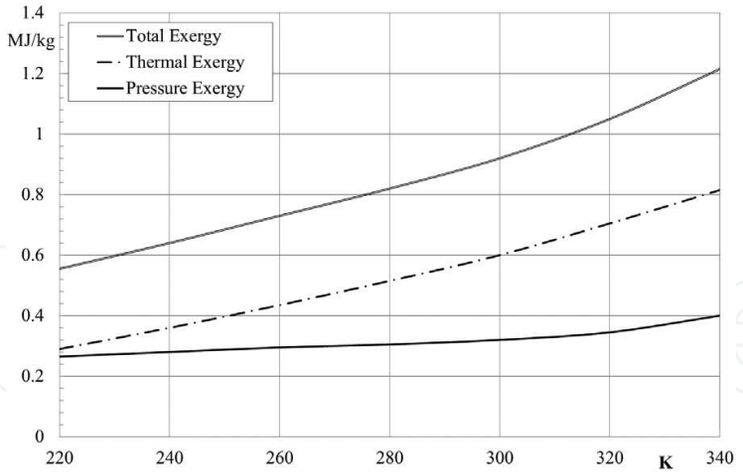


Figure 4. Exergy trend in function of the environment temperature.

3. Natural gas direct expansion technique

The concomitant production of methane in gaseous form and electrical power starting from the LNG regasification process can be achieved by means of an open Rankine cycle whose working fluid is the produced natural gas (NG). The thermodynamic transformations and the correspondent plant scheme are shown in **Figure 5**.

The LNG is compressed by the submerged pumps P_s and the delivery pump P_f from the atmospheric pressure to the supply pipeline pressure level and successively, through the booster pump P_b, to a supercritical final pressure value (transformation 1→2). The compression in liquid phase allows for lower pumping work. The fluid vaporization and its overheating are successively achieved in the HE1 heat exchanger supplied by a warm thermal source (usually seawater) to obtain the NG with a temperature close to the environment value (2→3). The produced gas has an elevated pressure value; therefore, an expansion in an organic turbine TE allows for the achievement of the pipeline pressure (3→4), with the contemporaneous production of mechanical work. The NG low temperature at the end of the expansion is newly increased by

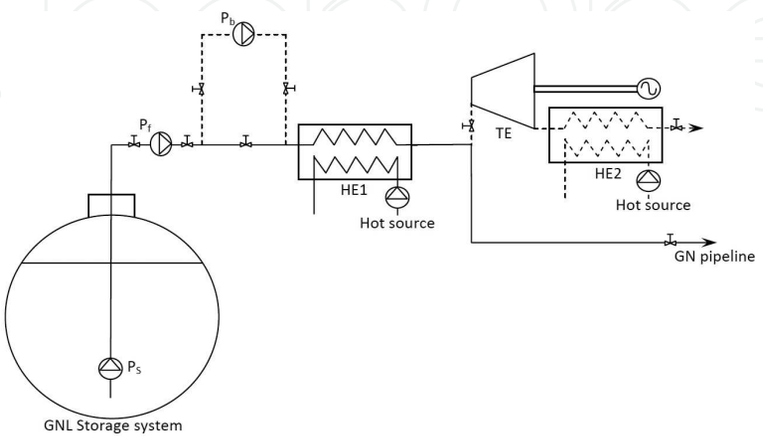


Figure 5. Sketch of a typical plant for natural gas direct expansion.

the heat exchanger HE2, to reach the outdoor temperature value, using the same hot source (4→5). By supposing a pipeline pressure level of 7 MPa and a maximum pressure level of 15 MPa, the specific compression and expansion works together with the heat exchanged in the two heating processes are reported in **Table 1** for an ideal case and a real case. For the latter, a compression and expansion isentropic efficiency of 0.80 and pressure drops equal to 3% of the maximum pressure in the heat exchangers are hypothesized. In **Table 1**, first and second thermodynamic efficiencies are also reported, showing that the system is self-sustaining from an energy point of view.

The use of the sole direct expansion technique of NG, however, does not allow the complete use of the potential of the cold source [6].

Transformation	Ideal	Real
Compression 0–2 [kJ/kg]	16.6	20.7
Heating 2–3 [kJ/kg]	667.9	663.6
Expansion 3–4 [kJ/kg]	67.0	52.3
Heating 4–5 [kJ/kg]	145.6	129.1
η_{th}^{I} [%]	7.5	4.8
η_{th}^{II} [%]	12.6	8.0

Table 1. Available enthalpy variation, first and second principle efficiencies concerning LNG during direct expansion process.

4. LNG exploitation as cold source for ocean thermal energy conversion systems (OTECs)

In tropical areas, the production of electric energy could be achieved by exploiting the temperature difference between the warmer surface seawater (hot source) and the coldest water pumped from the depths (cold source). This process is realized in the so-called "ocean thermal energy conversion systems" (OTEC), where the hot source is employed to vaporize a specific working fluid, while the cold source is required for its condensation. The thermal efficiency of a traditional OTEC is penalized due to the limited temperature difference between the thermal sources, and the correspondent plant results as being very expensive because the seawater flow rates involved in the vaporization and condensation process are significant, requiring pipes with great length and diameter and elevated pumping powers. The achievement of greater values of the thermal efficiency is recommended to reduce the initial costs of the OTEC plant, obtainable by a temperature difference growth between the thermal sources. At this purpose, the LNG stored in the tanks of the regasification terminal can be employed as a cold source; the hot one is the abundant and free seawater. By exploiting these two heat sources, an OTEC system could realize the LNG vaporization and the production of electric load through two thermodynamic cycles in cascade, using pure ammonia as the working fluid in a top closed Rankine cycle, and NG as working fluid in a bottom cycle through the direct expansion

technique. The seawater is used to promote the ammonia vaporization and the LNG to support its condensation. These cycles allow for an increment of the produced electric power by recovering the LNG exergy and, at the same time, guarantying the LNG regasification for the supply of the gas pipeline network [7]. An OTEC plant operating with ammonia presents temperature and pressure values that are strongly different from those of a traditional OTEC Rankine cycles; therefore, the ammonia was chosen because it seems a suitable working fluid on the basis of the real operative conditions. With reference to the higher thermodynamic efficiencies, Kalina cycles operating with ammonia-water mixtures could be a possible alternative, because they allow the achievement of an efficiency of 10–50% higher than cycles operating with pure ammonia [8]. In the proposed OTEC system, low condensation temperatures are required; therefore, pure ammonia (R717) appears to be the appropriate working fluid [9].

Figure 6 shows a closed Rankine cycle operating with pure ammonia. A vaporization pressure of 605 kPa is recommended because the ammonia vaporization temperature of 281 K is about of the same order of magnitude as the surface seawater temperature. At the start of expansion, R717 is in dry saturated vapor condition, with a correspondent specific enthalpy of 1454 kJ/kg. If the condensation pressure is set to 10 kPa, in the ammonia turbo-expander a mechanical work of 346 kJ/kg is producible, by reaching at the end of the process a temperature of about 202 K, ever greater than the LNG temperature stored in the tanks located in the terminal. By supposing a pinch point temperature difference of 10 K in the involved heat exchangers, the other thermodynamic parameters can be determined imposing a pressure drop in the condenser and the vaporizer equal to 3 and 5%, respectively, of the nominal pressure and an efficiency equal to 85% for the pumps and the turbo-expander [7].

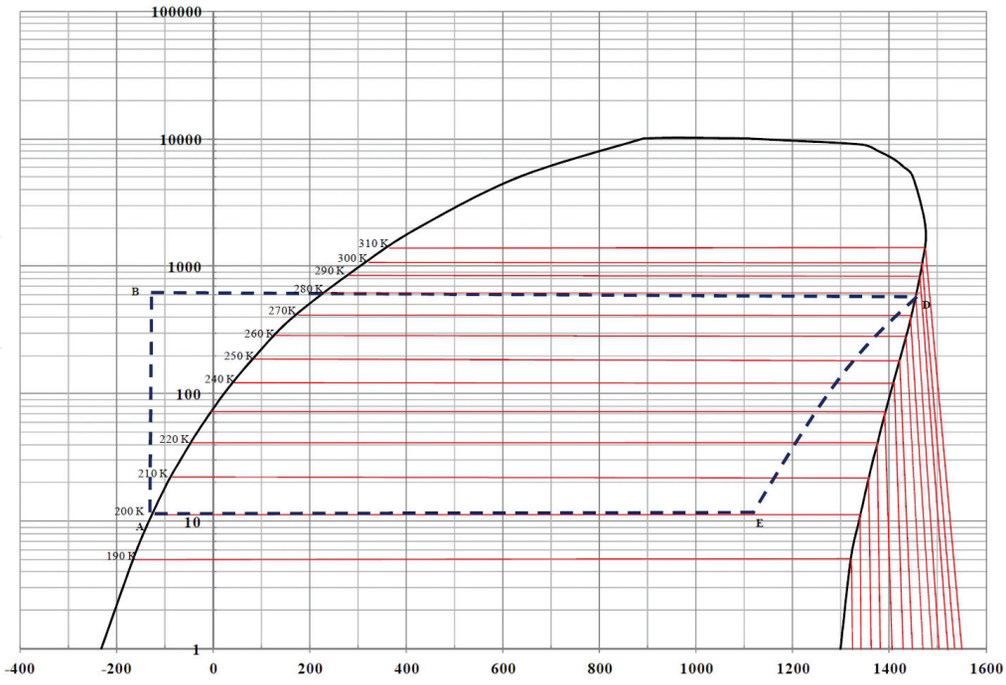


Figure 6. Closed Rankine cycle operating with ammonia (R717) and correspondent phase curve.

The specific enthalpy values of R717 in dry vapor saturated conditions (point D), in the function of the chosen temperature field, are slightly variable. By setting ammonia vaporization temperature to 288 K (where the vaporization pressure is equal to 691 kPa), the specific enthalpy is about 1460 kJ/kg, with an increment of +0.4% compared to the reference value of 1454 kJ/kg. By setting an ammonia vaporization temperature of 276 K (at a pressure of 454 kPa), the specific enthalpy becomes equal to 1449 kJ/kg (−0.3%). Therefore, the variation of the pinch point temperature produces negligible differences on the performances of the proposed OTEC system. The ammonia thermodynamic properties concerning the points shown in **Figure 6** are listed in **Table 2**.

The condensation heat released from the top cycle can be employed to support the LNG vaporization in a direct expansion cycle, by realizing two thermodynamic cycles in cascade (**Figure 7**). The efficiency of the top cycle is determined by considering the produced electrical power and the released thermal power supplies of the bottom cycle to produce additional electric power and to preheat the LNG in the regasification process. The remaining heat required to complete the regasification could be provided by seawater [7].

Transformations	h [kJ/kg]	x [kg/kg]	s [kJ/kgK]	ν [m^3/kg]	T [K]	p [kPa]
Start compression—A	−134.8	0	−0.621	0.00140	202	9.7
End compression—B	−133.8	–	−0.402	–	282	604.3
Start vaporization—C	229.1	0	0.883	0.0016	282	604.3
End vaporization—D	1454.0	1	5.240	0.219	281	575.5
End expansion—E	1108.1	0.85	5.534	8.29	202	10

Table 2. Main thermodynamic parameters concerning a Rankine cycle operating with ammonia: specific enthalpy, quality, specific entropy, specific volume, temperature, and pressure.

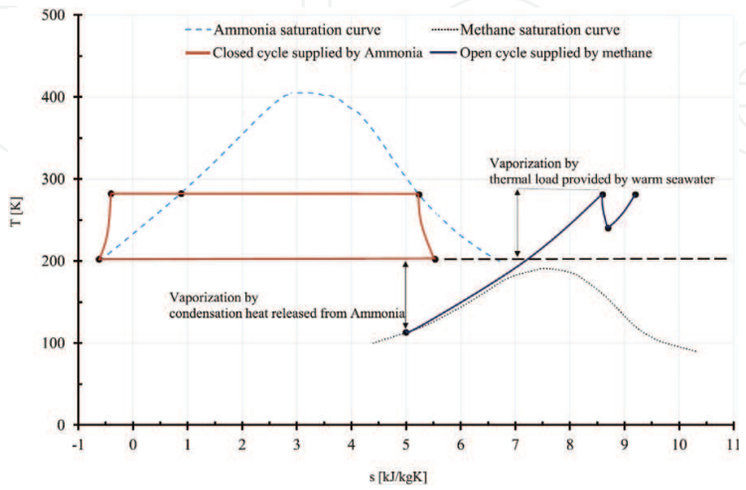


Figure 7. Combined cycle to produce electric energy in a OTEC plant promoting the LNG regasification.

4.1. Preliminary sizing evaluations

The available LNG flow rate can be evaluated from the regasification capacity and the plant operability of the terminal. Starting from this value, the ammonia flow rate requested by the Rankine top cycle can be determined with an energy balance of the R717 condenser [7]. Regarding the points shown in **Figure 6**, the ammonia flow rate is equal to:

$$\dot{m}_{NH_3} = \frac{\dot{m}_{LNG} \cdot C_p^{LNG} \cdot \left(t_{out}^{LNG} - t_{in}^{LNG}\right)}{h_E - h_A} \tag{5}$$

where the LNG outlet temperature can be set to the ammonia condensation temperature decreased by the pinch point (192 K, for example), while the inlet temperature is the temperature of the stored LNG of 113 K. An average LNG specific heat of 3000 J/kgK can be set based on the considered temperature range [10], while the values of h_E and h_A can be found in **Table 2**. The ammonia flow rate allows for the evaluation of the absorbed electric power to pump the working fluid by the relation:

$$W_P^{NH_3} = \frac{\left(p_B - p_A\right) \cdot v_A \cdot \dot{m}_{NH_3}}{\eta_P} \tag{6}$$

where p_A and p_B are the condensation and vaporization pressures, v_A is the specific volume of R717 in saturated liquid point and η_P is the efficiency of the pump requested to move the same fluid. Eq. (7) describes the energy balance equation of the OTEC vaporizer, and it allows the evaluation of the warm seawater flow rate required for the ammonia vaporization:

$$\dot{m}_{H_2O} = \frac{\dot{m}_{NH_3} \cdot \left(h_D - h_B\right)}{c_p^{H_2O} \cdot \left(t_{in}^{H_2O} - t_{out}^{H_2O}\right)} \tag{7}$$

where h_D and h_B indicate the initial and the final specific enthalpies of the vaporization process, whereas the seawater temperature difference comparing at the denominator is set to 7 K to prevent marine environment damages [11]. The ammonia turbo-expander produces an electric power that can be evaluated with the following relation:

$$W_T^{NH_3} = \left(h_D - h_E\right) \cdot \dot{m}_{NH_3} \cdot \eta_{\mathrm{ALT}} \tag{8}$$

where η_{ALT} is the alternator efficiency. The calculation of the net electric power requires also the evaluation of the pumping powers for the warm seawater and for LNG. These powers can be obtained starting from the calculations of the pressure drops in the condenser and in the vaporizer paths. The concentrated pressure drops can be neglected, therefore, only continuous pressure drops in the pipes (*Pa*) are considered:

$$\Delta P_c = \rho \cdot \frac{f \cdot L \cdot c^2}{2 \cdot D} \tag{9}$$

where f is a friction factor calculable from the Moody diagram, L is the overall length of the pipes, D their diameter, and c the fluid velocity in the pipes. The mass flow rate equation allows for the evaluation of the pipes diameter imposing the fluid velocity:

$$\dot{m} = \rho \cdot \frac{\pi D^2}{4} \cdot c \tag{10}$$

Moreover, the relation suggested by Kern for shell and tube heat exchanger type [12] can be employed to determine the pressure drops occurring in the vaporizer and in the condenser (Pa):

$$\Delta P_{HE} = \frac{2 \cdot f \cdot G_s^2 \cdot D_s \cdot (N+1)}{\rho \cdot D_e \cdot \left(\frac{\mu}{\mu_t}\right)^{0.14}} \tag{11}$$

In the last equation, the geometrical parameters are D_s (shell internal diameter), N (baffles number on the shell side), and D_e (equivalent diameter). From a thermodynamic point of view, G_s represents the specific mass flow rate, while ρ, μ, and, μ_t are, respectively, the density, the dynamic viscosity, and the dynamic viscosity of the fluid evaluated at the inlet temperature in the heat exchanger. Finally, the heat exchanger friction factor f can be calculated with the following correlation [13]:

$$f = e^{(0.576 - 0.19 \ln [Re_s])} \tag{12}$$

in function of the Reynolds number (Re_s) calculated starting from the values of the specific mass flow rate, of the equivalent diameter, and of the shell fluid dynamic viscosity at an average temperature (μ_{av}). The knowledge of the pressure drops allows for the evaluation of the electric power absorbed by seawater pumps by means of the following relation:

$$W_P^{H_2O} = \frac{(\Delta P_c + \Delta P_{\mathrm{HE}}) \cdot \dot{m}_{H_2O}}{\rho \cdot \eta_P} \tag{13}$$

Regarding the electric power required for the condenser operation, a similar relation can be used:

$$W_P^{LNG} = \frac{(\Delta P_c + \Delta P_{HE}) \cdot \dot{m}_{LNG}}{\rho \cdot \eta_P} \tag{14}$$

Finally, the net electrical work of the OTEC plant results [7]:

$$W_n^{OTEC} = W_T^{NH_3} - W_P^{H_2O} - W_P^{LNG} - W_P^{NH_3} \tag{15}$$

The thermal efficiency of the top cycle of the novel OTEC system is given by:

$$\eta_{th}^{top} = \frac{W_n^{OTEC}}{Q_{vap}^{NH_3}} \tag{16}$$

where the heat power extracted from the warm seawater ($Q_{vap}^{NH_3}$)can be evaluated with the relation:

$$Q_{vap}^{NH_3} = \dot{m}_{NH_3} \cdot (h_D - h_B) \tag{17}$$

Figure 8 shows the thermodynamic open cycle using LNG as a working fluid, while in **Table 3**, the values of the correspondent thermodynamic parameters are listed, by supposing a super critical vaporization pressure of LNG equal to 15 MPa and a pipeline operative pressure of 7 MPa. The global efficiency of the two cycles in cascade considers the additional mechanical work produced by the direct expansion of the natural gas and the supplementary heats required to complete the ammonia vaporization and for the natural gas postheating [7]. Point 3 is representative of the gasification process supplied solely by the ammonia condensation heat; therefore, the thermal power required to complete the phase change in a dedicated heat exchanger can be determined with the following relation:

$$Q_{3\to4} = \dot{m}_{LNG} \cdot (h_4 - h_3) \tag{18}$$

Likewise, the relation (19) provides the thermal power required to supply the natural gas postheating in another heat exchanger:

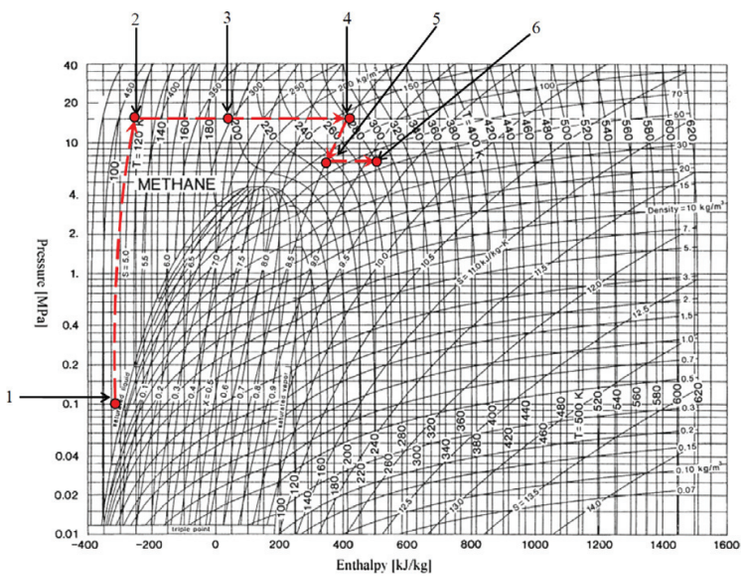


Figure 8. Thermodynamic coordinates of natural gas that evolves according to direct expansion.

	h [kJ/kg]	s [kJ/kgK]	ν [m³/kg]	T [K]	p [kPa]
Point 1	−301.3	5.0	0.00227	113	100
Point 2	−261.5	5.2	0.00238	125	15,000
Point 3	40.00	6.7	0.00315	191	14,450
Point 4	440.0	8.6	0.00800	281	13,728
Point 5	364.0	8.7	0.01250	240	7210
Point 6	505.0	9.2	0.01677	281	7000

Table 3. Main thermodynamic parameters concerning the natural gas cycle: specific enthalpy, specific entropy, specific volume, temperature and pressure.

$$Q_{5\to6} = \dot{m}_{LNG} \cdot (h_6 - h_5) \tag{19}$$

By means of simple energy balance equations related to the two heat exchangers, the flow rate of seawater required to complete the processes of gasification and postheating is evaluated by setting the limit of 7 K on its temperature difference. Furthermore, by using Eqs. (9) and (11) for the calculation of the pressure drops in these heat exchangers, the requested electric power to pump the warm seawater in the LNG cycle is:

$$W_P^{H_2O} = \frac{\left(\Delta P_C^V + \Delta P_{HE}^V + \Delta P_C^{PH} + \Delta P_{HE}^{PH}\right) \cdot \left(\dot{m}_{H_2O}^V + \dot{m}_{H_2O}^{PH}\right)}{\rho \cdot \eta_P} \tag{20}$$

where the superscript V refers to the LNG vaporization and the superscript PH refers to the postheating process.

The work to pump the LNG to the supercritical pressure value can be evaluated with the relation:

$$W_{1\to2}^P = \dot{m}_{LNG} \cdot (h_2 - h_1) \tag{21}$$

Finally, the mechanical work is calculable with the following equation:

$$W_{4\to5}^T = \dot{m}_{LNG} \cdot (h_5 - h_4) \tag{22}$$

The net electrical power achievable by the direct expansion (DE) technique is equal to:

$$W_n^{DE} = W_{4\to5}^T \cdot \eta_{ALT} - \frac{W_{1\to2}^P}{\eta_P} - W_P^{H_2O} \tag{23}$$

therefore, the global thermal efficiency of the proposed OTEC plant open is:

$$\eta_{TOT} = \frac{W_n^{OTEC} + W_n^{DE}}{Q_{vap} + Q_{3\to4} + Q_{5\to6}} \tag{24}$$

4.2. Performances of an OTEC plant powered by LNG

By setting a value of LNG flow rate equal to 208 kg/s, corresponding to a regasification capacity of 8 Gm^3 per year (13.3 Mm^3 per year in liquid phase) with a utilization factor of 85%, the results listed in **Table 4** are determined [7].

The coupling of the two cycles allows an increment of the net electric power, and at the same time, a higher value of the thermal efficiency is achievable if compared with those of a traditional OTEC systems. The electric power growth, in fact, is obtained by exploiting the second open LNG cycle that, additionally, allows the completion of the regasification process. The new thermodynamic process to vaporize the LNG requires an energy demand lower than traditional regasification plants. In comparison to LNG terminals where seawater flow rate is exclusively employed for LNG vaporization, the proposed plant allows major energy savings due to the

	Value
$\dot{m}_{NH_3}$	40 [kg/s]
$W_P^{NH_3}$	39 [kW_{el}]
$W_T^{NH_3}$	11.6 [MW_{el}]
$W_P^{H_2O}$	405 [kW_{el}]
W_P^{LNG}	146 [kW_{el}]
W_n^{OTEC}	11 [MW_{el}]
$Q_{vap}^{NH_3}$	63 [MW_{th}]
η_{th}^{top}	0.175 [−]
$W_P^{H_2O}(DE)$	1.64 [MW_{el}]
$W_{1\to2}^{P}$	9.74 [MW_{el}]
$W_{4\to5}^{T}$	15.0 [MW_{el}]
W_n^{DE}	3.63 [MW_{el}]
$Q_{3\to4}$	83.2 [MW_{th}]
$Q_{5\to6}$	29.3 [MWth]
η_{TOT}	0.081 [−]

Table 4. Main results of a OTEC system employing the two cycles in cascade.

lower pumping powers required to complete the process. The proposed plant does not emit CO_2 by considering that the production of electric energy is achievable without the combustion of fossil sources. The global CO_2 emissions present a null value because the self-consumed electric powers of the whole plant have been already contemplated in the calculations; therefore, the absorption of further electric power from the external grid is not requested [7].

5. Electric production from regasification of LNG using waste heat

From a thermodynamic point of view, a more rational use of LNG as a cold source is obtainable by inserting the ammonia closed Rankine cycle described in the previous section. This closed cycle has to be inserted between the thermodynamic cycle describing the operation of a conventional power plant and the open cycle of the direct expansion of natural gas. The possibility to exploit the high temperature of the fluid flow rate released by power plants, instead of seawater, allows for the employment of ammonia cycles with a higher pressure level, with consequent advantages in terms of mechanical work produced by the ammonia turbo-expander. Therefore, the plant operation happens by means of three cycles in a sequence: a closed thermodynamic cycle of the existing power plant (for instance, a water vapor Rankine cycle), an intermediate ammonia-closed Rankine cycle, and the open Rankine cycle for the natural gas direct expansion. By adopting the plant configuration shown in **Figure 9**, power plant and regasification terminal can operate independently. Moreover, an

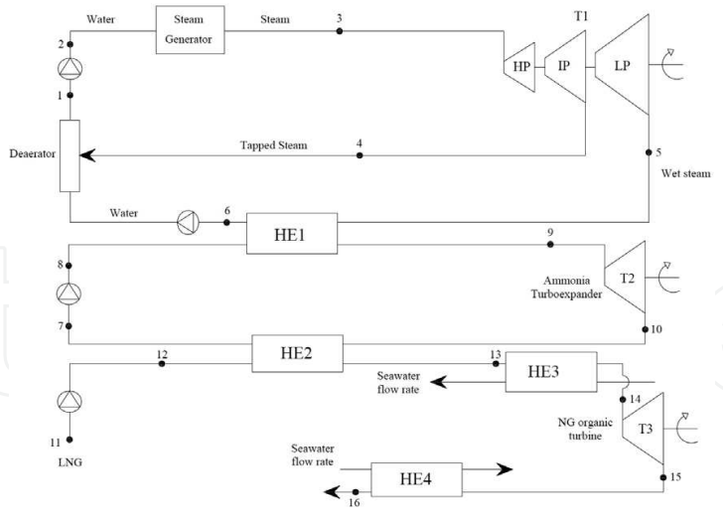


Figure 9. Plant layout according to the three cycles in cascade.

alternative plant configuration could provide the condensation heat released from the water steam cycle to gasify directly LNG in the open Rankine cycle [5].

Figure 10 shows the several thermodynamic transformations of the working fluids in the T-S plane. Regarding the open cycle supplied by NG, a part of the LNG regasification is carried out by exploiting condensation heat released from R717 in the HE2 heat exchanger; the process is completed by means of the seawater flow rate that evolves in the HE3 heat exchanger. R717 superheated steam is produced in the HE1 heat exchanger, by recovering the heat released by the top cycle. A further seawater flow rate is employed in the HE4 heat exchanger in order to realize the postheating process, after the expansion of the natural gas in the turbo-expander [5].

Supposing the availability of a constant steam flow rate from the top cycle, R717 flow rate is determined by considering the thermal balance equation to the HE1 heat exchanger. The superheating temperature at the R717 vaporization pressure is calculated starting from the

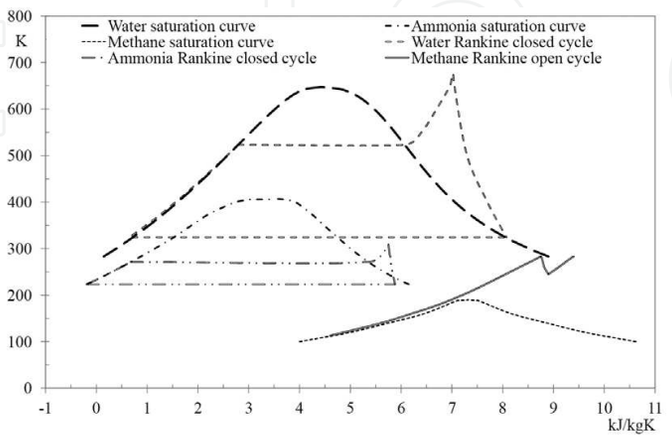


Figure 10. The three thermodynamic cycles in cascade: water steam Rankine cycle (top), ammonia cycle (intermediate), and NG cycle (bottom).

steam temperature, decreased by an appropriate pinch-point. With reference to the scheme of **Figure 9**, this equation is:

$$\dot{m}_{steam}\cdot(h_5 - h_6) = \dot{m}_{NH_3}\cdot(h_9 - h_8) \tag{25}$$

Likewise, the thermal balance at HE2 heat exchanger provides the LNG outlet enthalpy:

$$\dot{m}_{NH_3}\cdot(h_{10} - h_7) = \dot{m}_{LNG}\cdot(h_{13} - h_{12}) \tag{26}$$

where the enthalpy value h_{13} is determined to ensure a difference in temperature, compared with the ammonia condensation temperature, equal to the pinch-point (usually 10 K). In order to determine the gas inlet conditions in the turbo-expander (point 14), a maximum temperature difference of 7 K for the seawater flow rate can be set for the HE3 heat exchanger. The thermodynamic efficiency of the plant, with reference to the net power produced by every cycle W_n, is determined with the following relation:

$$\eta_{th} = \frac{W_{n1} + W_{n2} + W_{n3}}{(\dot{m}\cdot LCV)_{EPP} + Q_{seawater}^{HE3,HE4}} \tag{27}$$

where $\dot{m}$ and LCV, respectively, are the mass flow rate and the lower combustion value of the fuel employed in the existing power plant, and Q_{seawater} is the thermal power provided by the sea flow rate to supply both HE3 and HE4 heat exchangers. In order to determine the electric power produced in the turbines and the electrical power absorbed by the pumps, the following relations were used:

$$W_T = \eta_{ALT}\cdot\dot{m}\cdot\Delta h \tag{28}$$

$$W_P = \frac{\dot{m}\cdot\Delta h}{\eta_p} \tag{29}$$

where Δh is the real enthalpy variation.

5.1. Example of repowering of an incinerator plant by LNG regasification

In **Table 5**, the main parameters concerning the three thermodynamic cycles in cascade are listed; the whole system is supplied by the condensation heat released from a top water vapor Rankine cycle describing the operation of a waste incinerator plant [5].

Applying the aforementioned equations, **Table 6** reports the trend of the global thermal efficiency of the system, the global electrical power, and the thermal power extracted from the sea to complete the regasification process, based on the LNG flow rate.

In **Figure 11**, the trend of the thermal efficiency evaluated for three different operations of the plant described in **Figure 9** is reported: isolated functioning of the existing plant, with direct expansion supplied exclusively by the heat released from the waste incinerator plant, and with the proposed configuration operating with three cycles in cascade.

Thermodynamics cycles	Parameter	Value
Steam Rankine cycle	Pump efficiency	70%
	Boilers efficiency	87%
	Turbine efficiency	80%
	Maximum RDF feed flow rate	19 ton/h
	Percentage of tapped steam	14.5%
	Vaporization pressure	4 MPa
	Overheated steam temperature	673 K
	Condensation pressure	0.013 MPa
Ammonia Rankine cycle	Pump efficiency	70%
	Turbine efficiency	80%
	Vaporization pressure	0.4 MPa
	Overheated ammonia vapor temperature	35°C
	Condensation pressure	0.04 MPa
LNG Rankine cycle	Pumps efficiency	80%
	Turbine efficiency	78%
	Inlet pressure	8 MPa
	Inlet temperature	113 K
	Inlet pressure in turbine	15 MPa
	Outlet pressure from turbine	15 MPa
	Outlet temperature from HE3	283 K
	Electric generators efficiency	98%

Table 5. Main properties of a hypothetical existing power plant powered by ammonia and NG cycles.

LNG flow rate [kg/s]	NG temperature exiting from condenser [K]	$Q_{seawater}$ [MW]	η_{th} [%]	W_{tot} [MW]
108	213.35	33.4	22.8	22.9
105	215.75	31.4	23.1	22.7
100	219.65	28.0	23.7	22.5
95	224.35	24.5	24.3	22.3
85	234.75	17.7	25.7	21.8
75	248.05	10.9	27.4	21.3
65	267.15	4.0	29.3	20.8
55	297.65	0	30.3	20.3

Table 6. NG outlet temperature from condenser, thermal power delivered from seawater, thermodynamic efficiency, and produced electric power for the reference plant employing the three cycle in cascade, in function of LNG flow rate.

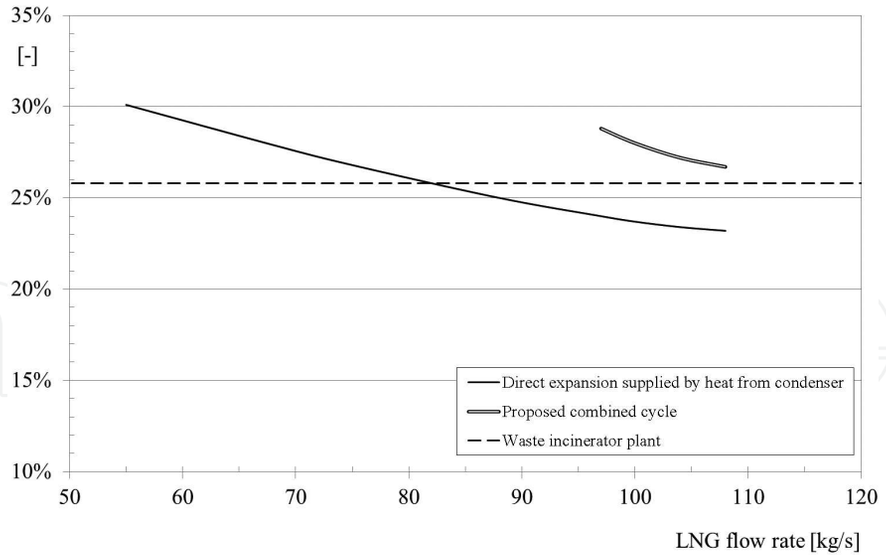


Figure 11. Thermodynamic efficiency varying the LNG flow rate in case of isolated functioning of the waste incinerator plant, with only direct expansion of NG and for the three cycles in cascade.

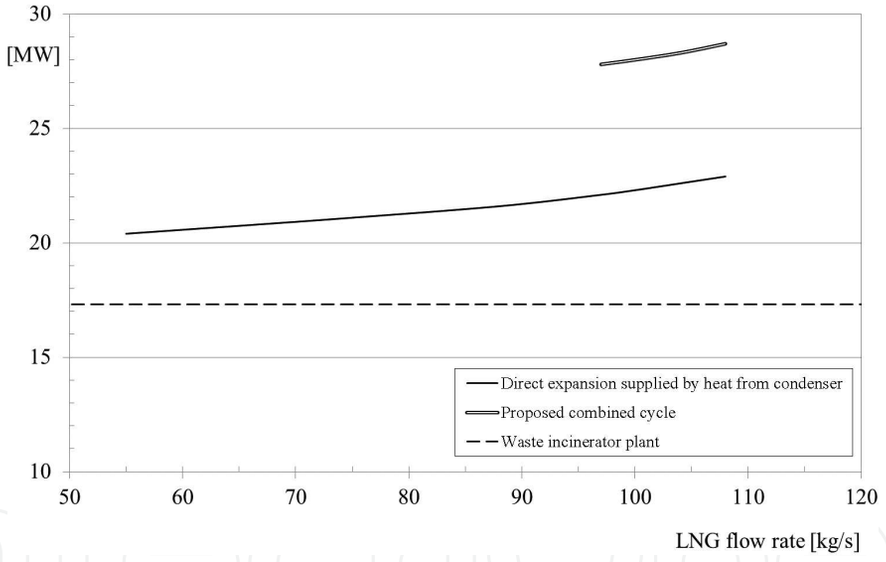


Figure 12. Total electric power varying the LNG flow rate in case of isolated functioning of the waste incinerator plant, with only direct expansion of NG and for the three cycles in cascade.

Likewise, **Figure 12** shows the trend of the electrical power produced by the three aforementioned plant operation modes.

The latter figure shows that employing an LNG flow rate of 100 kg/s, the electric power provided by the whole system increases by approximately 60% if compared with the isolated existing plant. Supposing the exploitation of existing plant condensation heat to supply the LNG direct expansion, the electric power increases by 25%. From the thermal efficiency point of view, varying the LNG flow rate from 95 to 105 kg/s, the rational exploitation of the LNG as

Power generation plant type	η_{th} [%]	kg CO_2/kWh$_e$	kg CO_2/year
Waste incinerator	25	1.120	$88{,}897 \times 10^3$
Steam produced by natural gas	42	0.481	$38{,}195 \times 10^3$
Steam produced by oil	42	0.632	$50{,}185 \times 10^3$
Steam produced by coal	42	0.866	$68{,}766 \times 10^3$
Combined cycle supplied by gas	56	0.361	$28{,}666 \times 10^3$

Table 7. CO_2 is emitted into the atmosphere supposing that the surplus of electric energy produced by the repowering of the waste incinerator plant is produced with conventional typologies of power plants.

a cold source allows the obtainment of values always greater than the isolated operation of the waste incinerator plant. With the considered plant parameters, other LNG flow rates have not been considered because they are not able to ensure a suitable temperature field to promote LNG vaporization by ammonia condensation heat. If the heat released from the existing plant is employed to supply the methane direct expansion, the thermal efficiency results as being greater than the isolated waste incinerator plant only for an LNG flow rate lower than 80 kg/s. The worsening of the thermal efficiency with the LNG flow rate increment is linked to the major thermal power provided by seawater to complete the regasification process. The use of the condensation ammonia heat allows for a substantial reduction in the seawater flow rate for LNG vaporization, of 63 and 60%, respectively, if compared to a simple regasification plant and a regasification plant with the direct expansion of natural gas. Regarding the electrical power surplus (10.8 MW for the examined case), the reductions of CO_2 emissions are reported in **Table 7** by hypothesizing that the same surplus of energy was produced in conventional power plants with an availability factor of 85% [5].

6. Further application for LNG exergy recovery

LNG can be used not only for power generation, but the exergy content could also be reused in different sectors. In the following, some possible applications of LNG exergy recovery during the regasification process are reported.

6.1. Cold chain and air-conditioning applications

Cold chain is represented by a series of steps (storage, transport, and sale) involved not only for the commercialization of certain food products, especially frozen and deep-frozen, but also to deliver cold flow rates required for different uses. The customers can be metropolitan areas, big shopping centers, or manufacturing companies of specific products. These systems require the realization of an appropriate pipeline system to transport the chilled flow rate from the LNG terminal to the users. These pipelines are relevant for the development of the so-called "smart-district" in order to achieve a rational exploitation of the energy resources. The choice of a suitable fluid for the cold flow rate transportation is also required in order to solve several problems concerning the viscosity growth, the freezing risk, and the elevated pumping powers. For this purpose, new fluid categories such as "ice slurries" and "nano-fluid" have been

recently introduced, while hydrocarbon mixtures and carbon dioxide are already employed because they are cheaper. In Ref. [14], a feasibility study in order to exploit the cold energy provided by a regasification terminal located around 2–3 km from the users is reported. In particular, a set of 10 food industries that require 9 MW of cold power at an average temperature of −43°C, and a large hypermarket with an estimated cooling power of 7.5 MW (7.0 MW only for air-conditioning application and 0.5 MW to supply cooling device with temperature ranging from −15 to −30°C) have been considered. The working fluid that supplies the pipeline is represented by CO_2 in liquid phase, produced by recovering the LNG exergy during the regasification process, allowing for a reduction of the pumping power by 30 times, compared to the gaseous CO_2. This system is efficient for a brief distance, because in long pipelines, the CO_2 pumping powers are too elevated.

6.2. Hydrogen production

Currently, the major industrial process employed for hydrogen production is the "steam reforming." The product of this process is a gas, called "reformed gas," which is formed by hydrogen, carbon dioxide, and carbon monoxide. The chemical reactions that govern the system are:

$$CH_4 + 2H_2O \rightarrow CO_2 + 4H_2 \tag{30}$$

$$CH_4 + H_2O \rightarrow CO + 3H_2 \tag{31}$$

In order to remove the carbon monoxide produced by the second reaction, the same is treated with water vapor, according to the following reaction:

$$CO + H_2O \rightarrow CO_2 + H_2 \tag{32}$$

The composition of the reformed gas depends on several parameters, but usually it is constituted by 80% of hydrogen, 20% of carbon dioxide, 0.5% of carbon monoxide, and 1.5% of unreacted natural gas (methane). For the CO_2 and other impurities removal, a further reaction with ammonia or the pressure swing adsorption (PSA) method is employed, by producing a reformed gas with a hydrogen concentration of 99.999%. For storage purposes, a hydrogen liquefaction is required, with a considerable expenditure of energy. At this step, the LNG regasification can be employed in order to achieve two results: production costs reduction and production of CO_2 in liquid phase as a waste product. The process is divided into four steps:

- the hydrogen production takes place in a regasification terminal where a part of the gasified GN is used in the reforming process as raw material;
- the reforming process happens with the usual methods reducing the impurities in a range between 10 and 100 ppm;
- the carbon dioxide is liquefied by exploiting the LNG latent heat;
- the carbon dioxide recovered is employed for industrial purposes.

LNG intervenes for hydrogen cooling, by reducing the pumping work requested for hydrogen compression, and for the reformed gas cooling, allowing the CO_2 extraction in liquid form. The

hydrogen liquefaction at the end of the process can take place via three methods: Brayton cycle supplied by helium, Linde simple method, and Claude Hydrogen method [15]. In the aforementioned methods, the presence of liquid nitrogen or other fluid for cooling applications is required, and furthermore, high-power pumping is necessary. The LNG can supply all the cooling processes, with significant savings, avoiding the employment of liquid nitrogen or other dangerous fluids.

6.3. Cryogenic thermoelectric generator

A cryogenic thermoelectric generator is a device that transforms the temperature difference between its junctions into electrical energy, but the low efficiency of the process restricts their employment to few cases. A different situation is observable when the thermoelectric generator recovers heat or cold from other processes: in these cases, the low efficiency of the generator is no longer a limitation. In the regasification process, LNG is used for the cold junction and the superficial seawater for the hot one. Supposing constant properties of junction materials, calculated at the junctions average temperature, the produced electric power can be determined as the difference between the thermal power delivered from the hot and the cold sources. The thermal efficiency can be determined in the function of a single parameter Z, called merit parameter, and defined by the relation [16]:

$$Z = \frac{\Delta S^2}{K \cdot r} \tag{33}$$

where K is the thermal conductivity of the material, r is the electrical resistance, and ΔS is the variation in the Seebeck coefficient. The maximum thermal efficiency is:

$$\eta = \frac{\Delta T \cdot (1 - 1/\varepsilon)}{T_c + T_h/\varepsilon} \tag{34}$$

where ΔT is the temperature difference calculated among the hot (T_h) and the cold (T_c) junctions. ε is defined in relation to the merit parameter and to the average temperature of the junctions T_m by the relation:

$$\varepsilon = \sqrt{1 + ZT_m} \tag{35}$$

In the case of LNG regasification, an ideal ΔT of about 170°C is available, with a thermal efficiency slightly greater than 9%. At present, in real cases, which exploit other thermal sources, a maximum efficiency of 8% is achievable, despite the higher temperature differences. From a thermodynamic point of view, LNG offers better results because the cold source has a higher quality compared to the hot thermal source.

6.4. Desalination of seawater

A widespread idea to solve the problem of lack of water for civil, industrial, and agriculture uses is to obtain distillated water from seawater. Among the various techniques for the

desalination of seawater, the "freezing technique" is suitable to be associated to the LNG vaporization [14]. The freeze desalination process is also known by the name of normal freezing or progressive freezing. The physical principle is such that when a water solution containing dissolved salts is cooled to its melting point, ice crystals, mostly made of pure water, are produced. All the different substances are excluded from the water crystalline structure. The salts in the solution are concentrated in the brine surrounding the ice crystals. After an appropriate time, the ice crystals are separated from the brine, washed and melted to obtain distillated water. The freeze distillation consists of three basic operations: freezing, washing, and melting. The cooling energy to be supplied to the desalination plant can be recovered from the LNG vaporization process.

In a freeze desalination plant, high salinity water is fed into the plant head through a heat exchanger where the water flow is progressively cooled to the melting temperature. Subsequently, the suspended solid matter is retained in the filter section, while the dissolved gases are separated from the fluid to achieve a better freezing. The water, cooled and pretreated, is sent to the next stage where ice crystals are formed. The freezer fluid has to subtract an additional fraction of heat from the brackish water, to achieve suitable conditions for the crystals creation. Since their formation, the crystals are covered with high salinity water. The salt solution has a high surface tension, and therefore, it adheres to the crystals. The brine and crystal mixture are successively sent, through a pump, to the washing compartment where the liquid phase is separated by drainage, and then the salts covering the crystals surface are separated. The desalination plant, therefore, can use the vaporization of the LNG for the formation of ice crystals.

6.5. Air separation

The separation of air into its components is a practice intensely used in medical or industrial application. For instance, nitrogen is an excellent inert gas, whereas pure oxygen is very much used in medicine. Typically, air separation processes are classified into three categories: cryogenic separation, pressure swing adsorption process, and distillation through membranes.

Usually cryogenic separation is preferable when a high degree of purity of components is required, and these have to be used in liquid phase. A plant for air cryogenic distillation [17] can be combined with a LNG regasification terminal to exploit the exergy available in the regasification process. After an initial filtering stage, the air is compressed and undergoes a first spray cooling. Successively, carbon dioxide is removed by a molecular sieve bed that eliminates also any remaining water vapor and any gaseous hydrocarbons from the air. Then, process air is passed through an integrated heat exchanger, for instance, supplied by LNG, where a great amount of cooling energy is required to reach the temperature value of −172°C. Part of the liquefied air is enriched by oxygen, whereas the remaining part is constituted prevalently by nitrogen in order to be again distilled to produce pure nitrogen in a high pressure distillation column. The output products are nitrogen and pure oxygen with impure argon because of the presence of small residual amounts of oxygen and nitrogen. Finally, argon is sent in a purification section where the oxygen is eliminated. Then argon is liquefied via a further cooling to a temperature of −180°C.

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