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Techno-economic feasibility study of a system for the transfer of refrigeration capacity from LNG regasification plants to industrial assets

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Abstract

The recovery of cold energy during the regasification of Liquefied Natural Gas (LNG) has gained attention in recent years due to the fast growth of the LNG trade market and the increasing importance that governments are giving to energy efficiency and sustainability. Near 200 kWh/ton of LNG are potentially recoverable during the regasification process, but this energy is usually discarded when seawater or ambient air are used as heat source. Researchers have focused on the development of technologies for the use of this cold energy in the fields of air separation and cryogenic power generation. However, in some regasification plants the demand of natural gas is so low or so fluctuating that this kind of applications are economically unfeasible. This research focused on determining the techno-economic feasibility of a heat pump and a cold distribution system for the transfer of a fraction of the refrigeration capacity of LNG to industrial assets with low-temperature refrigeration demands located in the surroundings of the regasification plant. CO_2 was selected as the heat transfer fluid that recirculates in a close loop between the cold users and the LNG site. A techno-economic model was implemented in Matlab taking the distance between the users and the LNG plant, and the refrigeration demand as the evaluation parameters. It was found that for a refrigeration demand of 20 MW the distance between the plants should be less than 1.2 km in order to make a project economically feasible.

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Keywords: waste heat recovery; Liquefied Natural Gas; R744 - Carbon Dioxide; industrial heat pumps

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Nomenclature

C	Cost, [$]
Ce	Cost purchased equipment, [$]
d	Pipe diameter, [m]
DE	Design & engineering factor, [--]
E	Energy, [J]
f	factor, [--]
L	Length, [m]
$\dot{m}$	Mass flow, [kg·s^{-1}]
OS	Offsites factor, [--]
P	Pressure, [kPa]
$\dot{Q}$	Cooling load, [W]
R	Revenue, [$]
S	Size, [--]
t	Time, [s]
T	Temperature, [K]
u	Fluid velocity, [m·s^{-1}]
$\dot{W}_{ele}$	Electric power consumption, [W]
X	Contingency factor, [--]

Greek letters

Δ	Difference, [--]
ρ	Density, [kg·m^{-3}]

Subscripts

c	Civil
calc	Calculated
cap	Capital
cond	Condenser
el	Electrical
er	Erection
esc	Escalation
evap	Evaporator
dem	Demand
fr	Friction
i	Internal, Instrumentation and control
l	Lagging and paint
loc	Location
m	Materials
ope	Operating
p	Piping
pump	Pump
s	Structures and buildings

Abreviations

HTF	Heat transfer fluid
IFV	Intermediate fluid LNG vaporizer
IF	Intermediate fluid
LNG	Liquefied Natural Gas

1. Introduction

Liquefied Natural Gas (LNG) is a mixture of light hydrocarbons with methane as the main component. It is regarded as a clean energy source because its combustion tends to be more complete due to the fact that it has the lowest carbon to hydrogen relation of all hydrocarbons. The low carbon footprint of LNG and the increasing concern for the environmental impact caused by other fossil fuels like oil and coal have boosted the LNG market in the last 15 years. An evidence of this is that the LNG global demand increased from 100 MTPA (Million Tons Per Annum) in 2000 to 240 MTPA in 2015 [4].

The volume of natural gas is reduced approximately 600 times during the liquefaction process, which makes it suitable for transportation over long distances in LNG super-tanker ships at temperatures and pressures close to -160 ºC and 1 bar respectively [15]. It is estimated that the liquefaction process requires approximately 500 kWh/ton LNG for compression and refrigeration [15]. Part of this energy remains in LNG, and, as stated by Koku et al., 2014 [7], as much as approximately 200 kWh/ton LNG are in theory recoverable during the regasification process if LNG is used as a refrigerant (directly or indirectly) in industrial processes or as a heat sink in power generation as indicated by Mokhatab et al., 2014 [11]. This amount of potentially-recoverable energy is wasted in most LNG receiving terminals, which typically use seawater, air, or the flue gases produced after burning natural gas as heat source for the regasification of LNG.

This paper presents a techno-economic feasibility study of a system consisting of a heat pump and a cold distribution network for the transfer of refrigeration capacity from LNG regasification plants to industrial assets located in their surroundings. With this system, part of the exergy of LNG would be recovered in refrigeration applications outside the regasification plant. The heat pump uses ethylene as working fluid, and the cold distribution system uses carbon dioxide (CO_2) as Heat Transfer Fluid (HTF) circulating between the LNG plant and factories or cold warehouses with refrigeration demands in the low temperature range, e.g. from -45 ºC to -25 ºC. The cold distribution system basically consists of a CO_2 condenser installed in the regasification plant (which is the evaporator of the heat pump), a liquid-CO_2 pump, a liquid-CO_2 pipeline connecting the regasification plant and the cold users, a CO_2 evaporator, and a second pipeline returning the gaseous-CO_2 to the CO_2 condenser in the regasification plant. The heat pump and the cold distribution system could be installed as add-ons to existing LNG terminals to increase their regasification capacity.

A cold distribution network taking advantage of the enormous refrigeration capacity of LNG regasification plants may be a profitable business case. The economic analysis presented in this paper is based on a business model in which industrial users would pay for the amount of cold energy they use. An external investor would be the owner of the cold distribution network. Part of the revenues would be paid to the LNG regasification plant as original owner of the refrigeration capacity. The external investor would cover the capital cost of the CO_2 condenser, the pipelines, the cryogenic pumps, the control system, civil and electrical works, among other expenses. The industrial users would invest in the CO_2 evaporators, and the LNG plant would have to invest in the heat exchangers required to complete the heating process of natural gas from -67 °C to approximately 10 °C and the connections between the existing facilities and the new LNG vaporizers. A mathematical model was built for the economic analysis, and the results indicate that the distance between the LNG plant and the cold users must not exceed 1200 m when the nominal refrigeration demand is 20 MW.

2. Description of the cold distribution system

Fig. 1 shows a basic scheme of the cold distribution system proposed for the transfer of refrigeration capacity from a LNG regasification plant to a cluster of cold users located nearby. The main assumption for the system presented in Fig. 1 is that the demand of natural gas from the regasification plant is always high enough so that the refrigeration capacity obtained by the regasification of LNG from -158 ºC to -67 ºC exceeds the refrigeration demand of the cold users. Under this assumption, cold energy storage systems are unnecessary. Fig. 1 is based on a system proposed by La Rocca (2010) for the utilization of cold energy of LNG far from the regasification facility [8].

As in the system proposed by La Rocca, CO_2 is selected as HTF. CO_2 is advantageous for this system because it maintains a positive saturation pressure at the required temperature levels in this system. Liquid CO_2 attains also a low viscosity at low temperatures, i.e. -50 °C, in comparison with other secondary fluids like brines or glycols, which helps to maintain the pumping power at low levels. It is also relevant to highlight that the density of CO_2 in the vapor phase is larger than the density of other refrigerants at the same temperature level, which leads to smaller pipe diameters and compressors. Furthermore, CO_2 is a natural refrigerant widely used in industry, so many components are available in the market. It has also a low global warming potential in comparison with other refrigerants and zero ozone depletion potential. Finally, it is non-toxic and non-flammable, which is a relevant aspect for applications in the food-and-beverage and pharmaceutical industries, potential users connected to the cold distribution system.

Fig. 2 presents a pressure vs enthalpy diagram of CO_2 depicting the process of the proposed cold distribution system. CO_2 is condensed at approximately -50 ºC by recovering part of the exergy of LNG (line 5-1 in Fig. 2) in the ethylene heat pump. The temperature of LNG increases from approximately -158 ºC to about -67 ºC in this heat pump, which would be located in the LNG plant. It is assumed that LNG is pressurized to 90 bar in a previous step, so this heating process occurs above the critical pressure of methane. A cryogenic pump increases the pressure of the condesate CO_2 to guarantee circulation in the system and delivers CO_2 in the liquid phase to the users. The pressure head of the pump is equal to the sum of the pressure drops in the liquid-CO_2 pipeline, the evaporators, the gaseous-CO_2 pipeline, and the condenser (line 1-2). Lines 2-3 represents the pressure drops in the pipeline transporting liquid CO_2 and the evaporators. Pressure drops in the evaporators and condenser are ignored in Fig. 2 for simplicity. Liquid CO_2 vaporizes at about -43 °C by taking heat from the cold applications of the companies connected to the cold distribution network in evaporators located in the premises of these companies (line 3-4). Finally, Line 4-5 represent the pressure drops in the pipeline transporting gaseous CO_2. On the other hand, the system depicted in Fig. 1 includes also a vessel to collect CO_2 after condensation and a vessel to collect the vaporized CO_2 near the cold users. These vessels help to separate liquid and gas phases and to absorb sudden changes in the refrigeration demand.

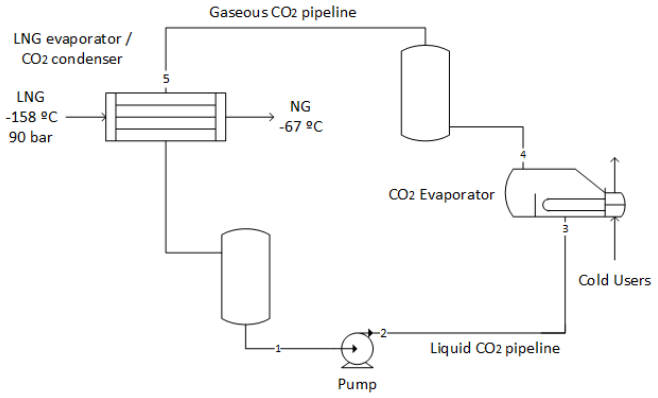


Fig. 1. Basic scheme of the system chosen for the transfer of refrigeration capacity from LNG plants to industrial assets.

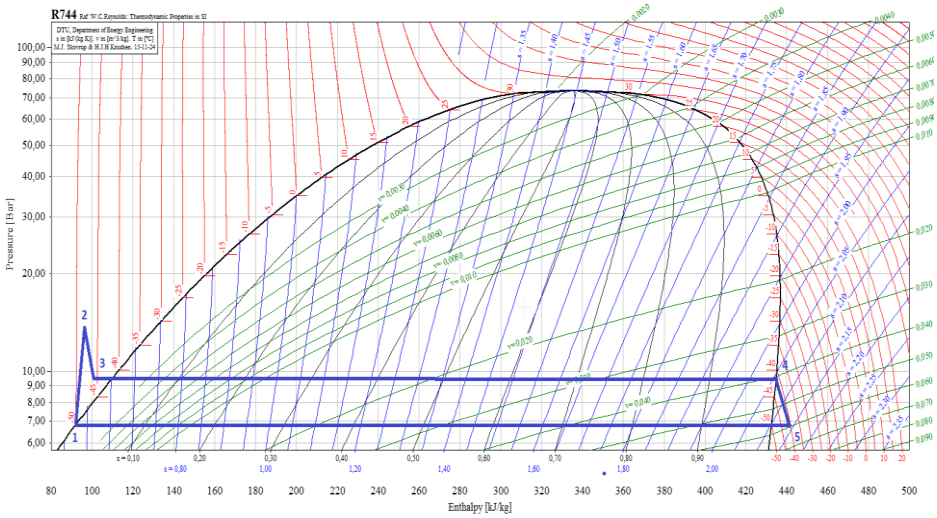


Fig. 2. Pressure vs Enthalpy diagram of CO_2 depicting the cycle of the system.

3. Sizing of ethylene heat pump and pipelines

Two components of the system that play a major role in the economics of the project are the ethylene heat pump and the pipelines. This section presents a description of these components and a description of the methods used for the estimation of their size. The size of these components depends mainly on the nominal refrigeration demand.

3.1. Ethylene Heat Pump

The main component of this system is the heat exchanger in which CO_2 is condensed and LNG is vaporized simultaneously, which is in fact an ethylene heat pump. La Rocca (2010) proposed a spiral wound heat exchanger for the condensation of CO_2 in the shell and evaporation of LNG inside the tubes [8]. However, this type of heat exchanger may have problems related to the solidification of CO_2 around the tubes located close to the entrance of LNG, whose external-wall temperature may fall below -56 °C, the freezing point of CO_2. An intermediate fluid LNG vaporizer using ethylene as Intermediate Fluid (IF) is proposed as an alternative to overcome the solidification-of-CO_2 issue. Fig. 3 shows a basic sketch of the proposed heat exchanger, which functions as a heat pump. This heat exchanger resembles the intermediate-fluid LNG vaporizer produced by Kobe Steel Ltd, Japan, which uses sea water as heat source for the vaporization of LNG and propane or butane as IF [3]. This heat exchanger, ethylene heat pump, will be called Intermediate Fluid Vaporizer (IFV) in this paper.

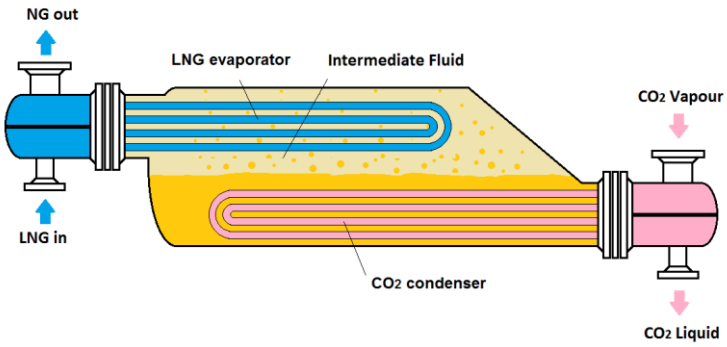


Fig. 3. Ethylene heat pump used for the condensation of CO_2 and simultaneous vaporization of LNG.

As seen in Fig. 3, CO_2 condenses inside the tubes of the lower bundle. This bundle is submerged in a boiling pool of ethylene, which vaporizes by using the heat of condensation of CO_2. Vaporized ethylene rises to the upper section of the vessel and condenses upon contact with the bundle of tubes carrying LNG. Meanwhile, LNG vaporizes inside the tubes of the upper bundle. Ethylene is a suitable IF for the proposed heat exchanger because it has a positive saturation pressure for saturation temperatures around -60 °C. Furthermore, the normal freezing point of ethylene is -169 °C, which is lower than the initial temperature of LNG, so the risk of solidification is eliminated. A parametric study also indicated that the heat transfer area of the heat exchanger is smaller when working with ethylene instead of a similar hydrocarbon like ethane.

For the design of this heat exchanger, the followings assumptions were considered:

- The condensation temperature of CO_2 is -50 °C.
- The temperature of the internal wall of the tubes of the CO_2-condenser section should remain above -54 °C in order to avoid solidification problems.
- The minimum temperature difference between the saturated IF and saturated CO_2 is 7 K.
- The minimum temperature difference between the saturated IF and LNG is 10 K.
- Pressure drops are neglected.
- Construction material: stainless steel ASTM A213 Gr. TP304L. Min. working temperature: -268 [ºC].

Table 1 collects the references of the heat transfer correlations used in the design of this heat exchanger. Some of these correlations are proposed by Xu et al. (2015) in their comparison of the heat transfer area required for intermediate-fluid LNG vaporizers using different IF´s [14].

Table 1. References for the heat transfer correlations used in the design of the IFV.

Process	Relevant Phenomena	Author	Reference
Condensation of CO_2	Condensation inside horizontal tubes	Shah, 2009	[12]
Evaporation of LNG	Evaporation inside horizontal tubes below pseudo-critical point	Bae and Kim, 2009	[2]
	Evaporation inside horizontal tubes above pseudo-critical point	Liang et al., 1998	[9]
Evaporation of IF	Pool boiling	Jung et al., 2004	[5]
Condensation of IF	Condensation outside horizontal tubes	Jung et al., 2004	[6]

3.2. Pipelines

The LNG regasification plant and the plants of the cold energy users are connected by two pipelines, one of them transporting liquid CO_2 from the LNG plant to the cold users and the other transporting gaseous CO_2 in the opposite direction. The design criterion of the gaseous-CO_2 pipeline is that the pressure drop along this line has to be less or equal to the difference between the saturation pressure of CO_2 in the evaporators and the saturation pressure of CO_2 in the condenser. Therefore, the diameter of this pipeline depends on the refrigeration demand, the length of the pipe, and the saturated temperatures of evaporation and condensation of CO_2. Fig. 4 presents the calculation procedure used for the definition of the diameter of the gaseous CO_2 pipeline, which takes into account Equation (1) and (2). Equation (2) corresponds to the Darcy-Weisbach correlation.

$$d_i = \sqrt{\frac{4 \cdot \dot{m}_{CO_2}}{\rho_{CO_2} \cdot \pi \cdot u_{CO_2}}} \quad (1)$$

$$\Delta P = f_{fr} \frac{L}{d_i} \frac{\rho_{CO_2} \cdot {u_{CO_2}}^2}{2} \quad (2)$$

For simplicity, the density of the gas in Equation (2) was calculated as the density of saturated CO_2 vapor at the temperature of condensation. This assumption ignores the fact that the density of the gaseous CO_2 changes along the pipeline as the pressure decreases and the temperature increases. On the other hand, the diameter of the liquid-CO_2 pipeline was defined as the smallest diameter for which the flow velocity is less than or equal to 1.5 m/s. The pressure drop in the liquid-CO_2 pipeline is only relevant for the calculation of the pumping power. Furthermore, it is also relevant to point out that the specification of the carbon steel pipes suitable for these pipelines is: Low-temperature, seamless, low-alloy (2¼ Ni) carbon steel pipe ASTM A333 Grade 7. Finally, it is important to mention that these pipelines must be insulated to minimize heat gain. The cost of the insulation was considered in the economic analysis.

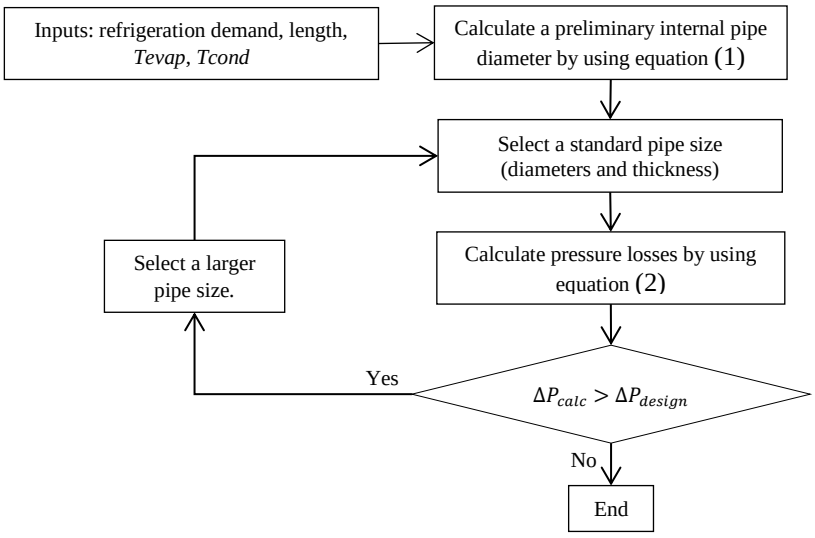


Fig. 4. Procedure for the definition of the diameter of the pipeline transporting gaseous CO_2.

4. Economic analysis

A mathematical model was implemented in Matlab for the economic analysis of the system. The model focuses on finding out the maximum allowable distance between the LNG facility and the cold users for which the project is still economically feasible for an external investor, owner of the cold distribution network. Fig. 5 presents the calculation procedure used in this mathematical model. This paper studies the economic feasibility of the project from the perspective of the owner of the cold distribution network. It was assumed that the project is indeed economically feasible for the cold users because their initial investment would be lower in comparison to the implementation of a vapor compression refrigeration system with the same capacity. Furthermore, the cold users would also have a reduction in their operational costs because the cost of the cold energy from LNG would be lower than the cost of the cold energy obtained from a vapor compression refrigeration system. On the other hand, it was also assumed that the project is profitable for the owners of the LNG regasification plant because they would increase their regasification capacity with a reduced investment, and they would also obtain some revenues from the sale of cold energy. Deeper analysis on whether or not the implementation of this system is profitable for the cold users and for the owners of the LNG plant is left out of the scope of this paper.

Next sections will describe the calculation of CAPEX, OPEX, and revenues, which are the core of the economic analysis. The IRR was used as decision-making criterion about the feasibility of the project.

4.1. CAPEX

The fixed capital cost of the project was estimated by using the factorial method of cost estimation presented by Sinnott & Towler, 2008 [13]. This method consists in applying factors to the cost of major purchased equipment in order to find the total cost of the project. Equation (3) presents the different factors considered, and

Table 2 gives the value of these factors as recommended by Sinnott & Towler for process plants involving only liquid and gases. The value of the location factor, f_{loc}, was selected for a plant located in India, country with favorable conditions for the implementation of this kind of projects. The value of the material factor, f_m, corresponds to stainless steel 304. For simplicity, this material factor applies to pipelines, IFV, and pumps even though the materials selected for pipelines and IFV are different.

$$C_{cap} = f_{loc} \cdot \left[(1 + OS) \cdot (1 + DE + X) \left(\sum Ce \cdot \left[\left(1 + f_p\right) \cdot f_m + f_{er} + f_i + f_{el} + f_c + f_s + f_l \right] \right) \right] \quad (3)$$

Sinnott & Towler also suggest the use of correlations for preliminary estimations of the cost of the major purchased equipment when lacking reliable cost data. Equation (4) gives the form of the suggested correlations. Table 3 presents the parameters *A*, *B*, and *n* of Equation (4) for the estimation of the cost of the IFV, the pump, and the electric motor of the pump. The parameters *A*, *B*, and *n* presented in Table 3 for the IFV correspond to a kettle reboiler, which is the type of heat exchanger closer to the proposed IFV. Because the values of the parameters *A*, *B*, and *n* presented in Table 3 were obtained based on cost data from 2006, an escalation factor f_{esc} is included in Equation (4) to update the cost of the equipment to a more recent year. The escalation factors included in Table 3 were obtained by using Chemical Engineering Plant Cost Indexes from October 2014 with respect to October 2006.

$$Ce = f_{esc} \cdot (A + B \cdot S^n) \text{ [\$]} \quad (4)$$

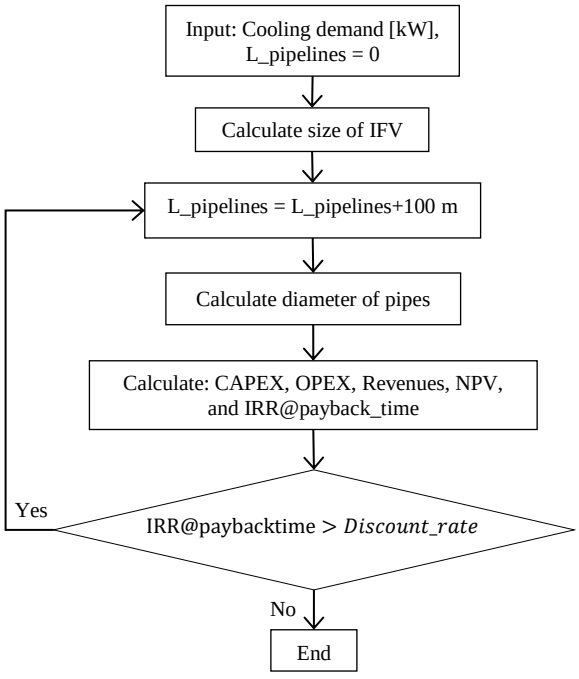


Fig. 5. Procedure for the calculation of the maximum economically-feasible distance between the LNG facility and the cold users.

Table 2. Value of the factors included in Equation (3) for the calculation of the fixed capital cost of the project. Source: [13]

Factor	*OS*	*DE*	*X*	f_{loc}	f_p	f_m	f_{er}	f_i	f_{el}	f_c	f_s	f_l
Value	0.3	0.3	0.1	1.02	0.8	1.3	0.3	0.3	0.2	0.3	0.2	0.1

Table 3. Coefficients for the estimation of the cost of purchased equipment by using Equation (4).

Process	Units for size, S	*A*	*B*	*n*	f_{esc}	Applicable factors
IFV	[m^2]	14000	83	1	1.14	$f_p, f_m, f_{er}, f_i, f_{el}, f_c, f_s, and\ f_l$
Centrifugal Pump	[L/s]	3300	48	1.2	1.18	$f_p, f_m, f_{er}, f_i, f_{el}, f_c, f_s, and\ f_l$
Explosion-proof motor	[kW]	920	600	0.7	1.23	$f_{er}, f_i, f_{el}, f_c, and\ f_s$

In this economic analysis the pipelines were considered as a major purchased equipment in order to simplify the mathematical model. Sinnott and Towler provide a correlation for the preliminary estimation of the cost of installed pipelines, which is presented in Equation (5). The factors to be considered in this case are: f_p, f_m, f_{el}, f_c, and f_l. An escalation factor equal to 1.18 must also be considered in this case.

$$Ce_{pipeline} = f_{esc} \cdot \left(880 \cdot (d_i)^{0.74} \cdot L_{pipeline}\right) \ [\$] \tag{5}$$

4.2. OPEX

The operating costs of the proposed system include the cost of the power required for pumps and auxiliary systems, a maintenance cost assumed equal to 3 % of the capital cost per year, and a risk reservation cost assumed equal to 3 % of the capital cost per year. The pumping power depends on the mass flow of CO_2, which at the same time depends on the refrigeration demand. The refrigeration demand usually varies along the day, and this variation is strongly dependent on the nature of the cold application. As an example of this variation, Fig. 6 shows the typical cooling load profile of a pork processing plant in 24 hours [10]. This economic analysis uses this profile, scaled up for different cooling loads, to simulate the variation of the demand of the cold users.

Equation (6) is used for the calculation of the energy consumed by the pumps in one day. The electric power consumption of the pumps, $\dot{W}_{ele_pump}$, was calculated by multiplying the total pressure drop of the pipelines times the volume flow. An overall pump efficiency equal to 85 % was also considered. The power consumption of the pumps varies along the day with a profile similar to the one shown in Fig. 6. On the other hand, Equation (7) gives an estimation of the total operating cost of the cold energy distribution system per year including the cost of the pumping power and the maintenance and risk reservation costs. Equation (7) considers 8000 operating hours a year, and an electricity cost of 65 $/MWh.

$$E_{pump} = \int_0^t \dot{W}_{ele_pump} \cdot dt \ \ [\mathrm{MJ/day}] \tag{6}$$

$$C_{ope} = \left(\frac{8000}{24}\right) \cdot 65\ [\$/\mathrm{MWh}] \cdot \frac{E_{pump}}{3600} + 0.06 \cdot C_{cap}\ [\$/\mathrm{year}] \tag{7}$$

4.3. Revenues

The revenues of this project come from the sale of cold energy. The external investor would receive 75 % of the revenues whereas the remaining 25 % would be paid to the LNG company. The price of the cold energy would depend in practice on the temperature level of every cold application connected to the cold distribution network so that the colder the application the higher the price. However, in this economic analysis, it is assumed that all the users are in the temperature range from -45 °C to -25 °C and that the price of the cold energy for these users is 40 €·MW^{-1}·h^{-1}. The cost of 1 MW·h of cooling effect obtained by using a vapor compression refrigeration system would be 43.3 € if the price of the electricity were equal to 65 €·MW^{-1}·h^{-1} and the COP of the system equal to 1.5. As a result, paying cold energy at 40 €·MW^{-1}·h^{-1} would be advantageous for the industrial user.

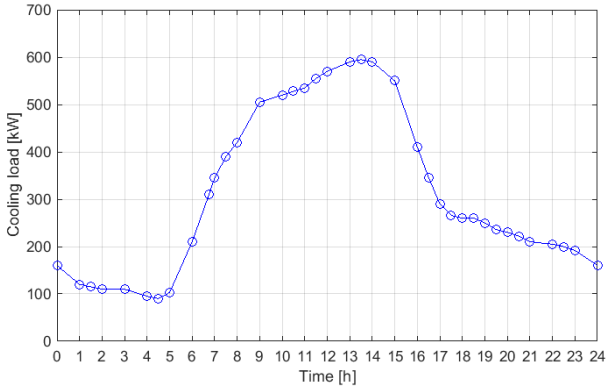


Fig. 5. Cooling demand profile of a pork processing plant in 24 hours [10]. Process temperature -35 °C.

Equation (8) gives the amount of cold energy consumed by the users in one day. The cooling demand, $\dot{Q}_{dem}(t)$, is a function of time and has a profile similar to the one presented in Fig. 6. On the other hand, Equation (9) gives the expected revenues of the project per year. This equation considers the system operates 8000 hours per year.

$$E_{dem} = \int_0^t \dot{Q}_{dem}(t) \cdot dt \quad [\text{MJ/day}] \tag{8}$$

$$R = 0.75 \cdot \left(\frac{8000}{24}\right) \cdot \left(\frac{E_{dem}}{3600}\right) \cdot (40\ [\$/\text{MWh}]) \tag{9}$$

4.4. IRR and NPV

The Internal Rate of Return (IRR) and the Net Present Value (NPV) of the project were calculated by taking into account the parameters listed in Table 4.

Table 4. Parameters for the calculation of the IRR and NPV of the project.

Parameter	Units	Value	Parameter	Units	Value
Maturity	[years]	20	Interest on debt	[%]	5
Inflation	[%]	2	Discount rate	[%]	8
Equity	[%]	25	Corporate tax	[%]	33
Return on equity	[%]	15	Project life span	[years]	30
Debt = 1 - Equity	[%]	75	Payback time expected	[years]	10

5. Results

Table 5 summarizes some results of the sizing of the IFV, pump, and pump motor for four different refrigeration capacities considered. Table 5 also includes the calculated costs of these components considering installation, piping connections, electrical and civil works, among others applicable factors given in Table 3. On the other hand, Fig. 7 presents for four different refrigeration capacities the IRR at the 10th year of operation of the project (it was assumed that the payback time was 10 years) as a function of the length of the pipelines connecting the LNG facility and the cluster of cold users. It is possible to read from Fig. 7 that the maximum allowable length of the pipelines for which the project is still feasible, i.e. IRR is less or equal to the discount rate of 8 %, is approximately 80 m, 450 m, 700 m, and 1200 m for refrigeration capacities of 1 MW, 5 MW, 10 MW, and 20 MW respectively.

Table 5. Estimated capital cost of the main purchased components of the system for several refrigeration demands. Pipelines excluded.

Parameter	Units	Cases Considered			
Refrigeration demand	[MW]	1	5	10	20

Number of IFV	[-]	1	1	2	2
Total heat transfer area IFV	[m^2]	230	1169	2355	4920
Cost of IFV (Installed)	[k$]	141	473	952	1859
Flow capacity Pumps	[L/s]	2.7	13.3	26.6	53.1
Cost pumps (Installed)	[k$]	31	39	51	79
Rated power pump motors	[kW]	2	6	18	37
Cost of pump motors (Installed)	[k$]	11	18	31	48

6. Conclusions

A system including an ethylene heat pump and a cold distribution network was proposed for the recovery of part of the energy invested during the liquefaction of LNG by saving refrigeration power in factories located around a LNG regasification plant. Food-and-beverage and pharmaceutical industries, which have refrigeration demands in the temperature range from -45 °C to -25°C, could benefit by connecting their cold applications to the cold distribution system.

The proposed cold distribution system uses CO_2 as HTF to carry low temperature heat from factories or warehouses of cold users to the LNG regasification plant. CO_2 is condensed by rejecting heat to LNG in an ethylene heat pump (where LNG is vaporized simultaneously) and evaporated by receiving heat from the cold users. One pipeline transports CO_2 in liquid phase from the LNG plant to the cold users, and a second pipeline transports CO_2 in gas phase in the opposite direction. A liquid CO_2 pump is needed to overcome pressure losses in the main pipelines and to keep the HTF circulating in the system.

It was found that the economic feasibility of a project for the implementation of this system depends mainly on the distance between the cold users and the LNG facility since the cost of the pipelines has the biggest impact on the fixed capital cost of the system. For a refrigeration demand of 20 MW and under the assumptions of the current economic analysis, the maximum economically-feasible distance is 1200 m.

Cold energy storage systems could be necessary when the synchronization between the refrigeration demand of the cold users and the offer of refrigeration capacity of LNG (in the temperature range between -158 ºC to -67 ºC) is an issue. The use of phase change materials for low temperature cold energy storage is one of the alternatives to consider in further research.

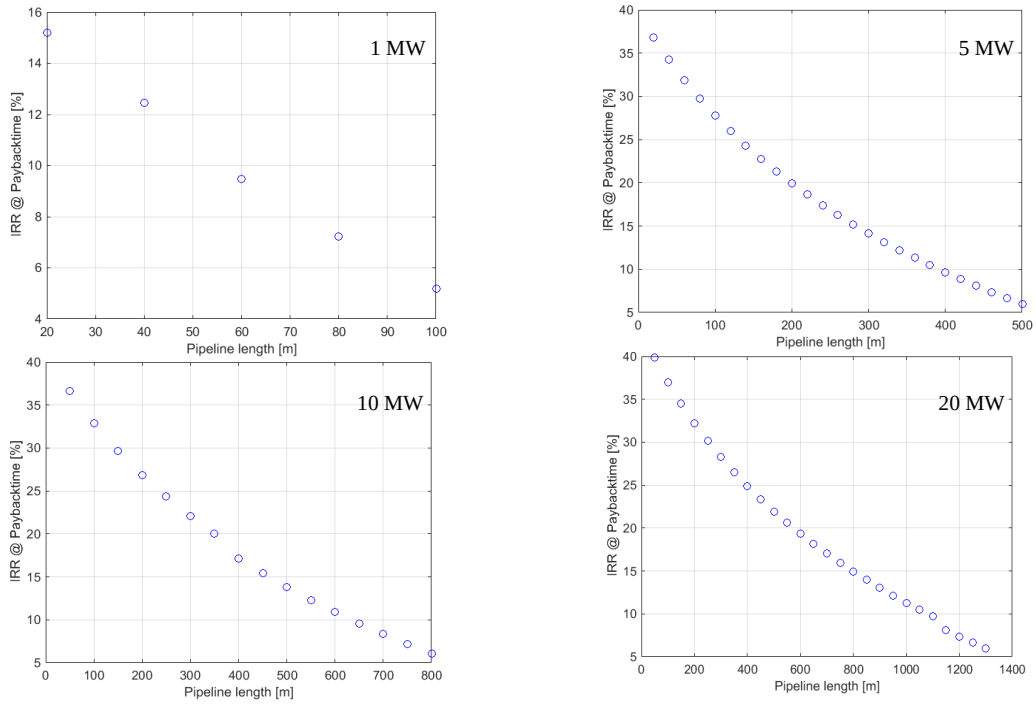


Fig. 6. IRR at the expected payback time (10 years) as a function of the pipeline length for four different refrigeration capacities.

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