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## Research Article

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# Synergistic Oil–Gas–Geothermal Integration and Advanced Exergy Optimization of Organic Rankine Cycles for Refinery Waste Heat Recovery in Nigeria: Co-Production of Mature-Well Brine with Flue-Gas Heat

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## Abstract

Nigerian refineries reject much of their input energy as waste heat, while mature Niger Delta oil wells co-produce warm brine that is lifted and disposed of. This study examines whether the two streams can drive one organic Rankine cycle (ORC). In the proposed architecture, co-produced brine preheats the working fluid and refinery flue gas, delivered through a thermal-oil loop, supplies the higher-temperature heat. Three site-specific cycles are analysed: isobutane at Warri, isopentane at Port Harcourt, and n-pentane at Kaduna, with evaporator temperatures of 140–160 °C and a 90 °C condensing temperature. The model combines CoolProp property evaluation with mass, energy and exergy balances, advanced exergy decomposition, NSGA-II multi-objective optimization, and a screening techno-economic and CO<sub>2</sub> model. Net power outputs of 180 kW (Warri), 200 kW (Port Harcourt) and 170 kW (Kaduna) correspond to exergy efficiencies of 0.12, 0.13 and 0.11. Heat exchangers account for 70.6% of exergy destruction at Warri and 66.7% at Port Harcourt, whereas unrecovered stack exergy is the largest term at Kaduna (53.5%). Under assumed economic inputs, simple payback is 3.9–4.7 years and CO<sub>2</sub> reduction is 16.9–21.1% against assumed baselines. These outputs presuppose aggregate multi-stream flue-gas recovery. Treating the reported 649 kW as one stack, Port Harcourt needs about 1.85 MW of flue-gas duty even with 140 °C brine, and a benchmark screening estimate brackets this only at nameplate refinery throughput. The source–fluid temperature mismatch is the main design constraint, and a heat-source audit is the key step before implementation.

**Keywords:** Organic Rankine cycle; Advanced exergy analysis; Refinery waste heat; Co-produced geothermal brine; Niger Delta; NSGA-II

## 1 Introduction

### 1.1 The refinery waste heat burden

Nigeria's downstream sector combines a highly energy-intensive value chain with an unreliable electricity supply. Refineries reject large amounts of low-grade heat, which raises fuel consumption per barrel processed, increases operating costs, and drives avoidable CO<sub>2</sub> emissions. A techno-economic comparison across refineries in several countries finds that ORC engines are in general economically feasible for heat sources above about 70 °C, with high sensitivity to energy prices (Gangar et al. 2020). Recovery also fits the wider decarbonization agenda for hydrocarbon producers (IEA 2023).

A first-law loss figure understates the problem, because heat differs in usefulness by temperature. Exergy analysis is the appropriate lens. The Warri refinery has a reported base exergy efficiency of 28.6%, so 71.4% of the available work potential is lost, mainly in heat exchangers (Imene 2024). The system boundary of that figure (unit or plant) is not established here, so it is used for context only. At Port Harcourt, flue gases leave at about 335 °C, and a techno-economic assessment of Nigerian industrial waste heat estimated a recoverable thermal power near 649 kW (Imene

2024). These figures are of different kinds: the 71.4% is an exergy loss for the Warri site, and the 649 kW is a single-source energy-recovery estimate. They are used here with those distinctions kept explicit. The reported cycle outputs presuppose that several such streams are recovered together (Section 5.6).

## 1.2 ORC waste heat recovery: state of the art

The ORC is the mature technology for converting heat below roughly 350 °C into shaft work. Organic fluids with low boiling points allow efficient expansion where steam would be impractical. Reviews of working-fluid and expander selection set the design rules that still govern the field (Bao and Zhao 2013). Architectural reviews catalogued the recuperated, cascaded, and dual-pressure layouts used to improve heat-source matching (Lecompte et al. 2015). Industrial ORC recovery has been assessed for European energy-intensive sectors, including oil and gas (Campana et al. 2013). Refinery-specific studies report that ORC conversion of refinery heat streams can be technically feasible and economically viable (Jung et al. 2014; Gangar et al. 2020).

Three methodological developments shape the present work. The first two have older origins and have been applied to ORC systems more recently.

1. **Advanced exergy methods.** Splitting destruction into endogenous and exogenous parts, and into avoidable and unavoidable parts, identifies which component interactions, rather than which components, cause losses (Tsatsaronis and Park 2002; Kelly et al. 2009).
2. **Fluid comparison.** Comparative studies of isobutane, isopentane, and their mixtures report that the preferred fluid depends on the performance metric and the operating conditions (Feng et al. 2024).
3. **Multi-objective optimization.** Pareto-based search gives a defensible way to balance net power against irreversibility (Deb et al. 2002).

Binary geothermal plants are ORC plants in thermodynamic terms, and exergetic assessments of geothermal plant types (Yari 2010) supply the analytical template used here.

## 1.3 Geothermal potential of the Niger Delta, Anambra, and Chad basins

Nigeria has no active volcanism, so its geothermal prospects lie in sedimentary basins where heat is carried by deeply buried, water-rich formations. Regional assessments identify the Niger Delta, Anambra, and Chad basins as principal candidates and characterize their gradients and heat flow (Kwaya and Kurowska 2021). Earlier basin-scale work documents gradient variation across the southern Nigeria basin and the Anambra Basin (Avbovbo 1978; Onuoha and Ekine 1999) and in the Nigerian sector of the Chad Basin (Kwaya et al. 2016).

The Niger Delta gradient is moderate, and the sequence is thick, porous, and saturated. Gradients derived from bottom-hole temperatures in the eastern Niger Delta range from 13.5 to 33.7 °C/km, with a mean of 23.6 °C/km (Kwaya and Kurowska 2021). In the Nigerian sector of the Chad Basin, gradients of 2.81–5.88 °C/100 m (mean 3.71 °C/100 m) have been reported (Kwaya et al. 2016). The useful resource in the Niger Delta is therefore a low-to-medium-enthalpy brine at accessible depth, not a high-temperature steam field.

This resource is already produced without being used. Mature Niger Delta fields carry high water cuts, so operators lift and dispose of large volumes of warm brine. Studies of geothermal recovery from mature Nigerian oil and gas fields (Bello et al. 2020) argue that this fluid is a distributed, already-drilled heat source. A worldwide review reports that produced-water streams from mature fields typically have temperatures of 65–150 °C (Liu et al. 2018). Wellhead temperatures sit in the low-to-medium enthalpy range, so a brine-only ORC is thermodynamically limited. That is the reason to study a supplemental high-temperature source.

## 1.4 Oil–gas–geothermal synergy

The premise of co-production is that the oil and gas industry already owns wells, surface facilities, water-handling capacity, and subsurface expertise that geothermal projects otherwise have to buy. The IEA’s assessment of synergies between the two industries frames co-production, well repurposing, and technology transfer as near-term routes to lower-cost geothermal power (IEA 2024). A screening roadmap for low-temperature heat recovery from

mature hydrocarbon fields has been proposed and tested on an Italian field (Liu et al. 2018). Expected benefits include lower capital cost where existing infrastructure is reused and a lower water-disposal burden.

## 1.5 Critical appraisal and research gap

Within the literature reviewed for this study, four gaps stand out. They are judgements from a limited reading, not the result of a systematic search.

1. **Isolated assets.** Co-production studies treat the upstream asset in isolation, with reservoir fluid as the only heat source. Refinery waste heat studies treat the downstream asset in isolation, with no link to upstream fluids in the same basin.
2. **Missing exergy design for Nigeria.** In the reviewed set, the few Nigerian studies that address waste heat recovery or geothermal potential rely on generic heat-source assumptions and rarely use exergy-based, component-resolved design.
3. **Transcritical operation is under-treated.** Near-critical fluids (isobutane above 134.7 °C) are not examined against a bounded 140–160 °C window in the reviewed set.
4. **Evidence quality.** Several supporting sources are workshop or conference papers, or several years old. Their quantitative claims should be treated as indicative until validated against operating data.

The novelty of this paper is the **integrated use of the two thermal streams**. Co-produced mature-well brine provides a steady, low-variance thermal base. Refinery flue-gas heat, delivered through an intermediate thermal-oil loop, supplies the high-temperature increment. One ORC train, sized and optimized with advanced exergy analysis, exploits both. The contributions are:

- a hybrid architecture and its thermodynamic formulation, including the treatment of transcritical isobutane;
- site-resolved exergy diagnostics for Warri, Port Harcourt, and Kaduna, with an endogenous/exogenous and avoidable/unavoidable decomposition;
- a Pareto characterization of net power versus exergy destruction under the 140–160 °C evaporator and 90 °C condenser bounds;
- a techno-economic and CO<sub>2</sub> assessment of retrofitting mature-field assets, with an open, reproducible Python implementation (Online Resources 1 and 2).

Section 2 sets out the data basis and methods. Section 3 describes the architecture, Section 4 the formulation, Section 5 the optimization and results, and Section 6 implementation and conclusions.

## 2 Data Basis and Methods

### 2.1 Data basis

The study combines reported site data, literature data, and explicitly assumed inputs. Section 5.1 lists the reported operating points that the analysis takes as its starting point: working fluid, evaporator temperature, net power, cycle exergy destruction, and exergy efficiency for Warri, Port Harcourt, and Kaduna. They are treated as given and are not re-derived here.

Site heat-source data are limited. The Port Harcourt flue-gas temperature (335 °C) and the single-stack recoverable heat (649 kW) come from a techno-economic assessment of Nigerian industrial waste heat (Imene 2024). The effective source temperatures at Warri (320 °C) and Kaduna (300 °C) are assumed. The Warri refinery exergy efficiency of 28.6% is used only as context (Section 5.1).

Basin gradients are taken from Kwaya and Kurowska (2021) and Kwaya et al. (2016), and the produced-water temperature range from Liu et al. (2018). Economic and CO<sub>2</sub> inputs (avoided electricity price, expander cost, capacity factors, co-production uplift, and utility-block baselines) are assumptions set within plausible ranges and are stated where they are used in Sections 4 to 6. Fuel intensities for the screening estimate in Section 5.7 come from Energetics Incorporated (2006), and the refinery configuration from BPE (n.d.) and Kpler (2024).

## 2.2 Model and software

All calculations were written in Python. Working-fluid properties and saturation states are evaluated with CoolProp (Bell et al. 2014). The cycle, exergy, and advanced exergy models are formulated in Section 4, and the multi-objective search uses an implementation of NSGA-II (Deb et al. 2002). The cycle model is a reduced-order surrogate calibrated to the reported operating points (Section 4.4). No commercial process simulator was used.

## 2.3 Reproducibility

The script that produces all figures and the numerical results in Sections 3 to 6 is provided as Online Resource 1. The screening estimate in Section 5.7 is reproduced by the script in Online Resource 2.

# 3 System Description and Hybrid Architecture

## 3.1 Physical architecture

The integrated system has four coupled subsystems (Table 1).

**Table 1** Subsystems of the hybrid architecture.

| Subsystem                                             | Function                                                                                                                                     | Key design issues                                                                      |
|-------------------------------------------------------|----------------------------------------------------------------------------------------------------------------------------------------------|----------------------------------------------------------------------------------------|
| Crude Distillation Unit (CDU) heat-recovery interface | Flue gas from the heater convection section (335 °C reported at Port Harcourt) heats a thermal-oil loop                                      | Acid-dew-point limit on stack temperature; fouling; draft and back-pressure            |
| Geothermal wellhead and brine loop                    | An abandoned or mature well produces brine through existing artificial lift; brine passes a corrosion-resistant exchanger and is re-injected | Scaling, corrosion, produced-water chemistry, well integrity, brine supply temperature |
| ORC power block                                       | Pump, evaporator (or supercritical heater), expander-generator, condenser                                                                    | Fluid selection, pressure rating, expander matching                                    |
| Heat sink and utility service                         | Condenser at 90 °C rejects heat to industrial-water preheating                                                                               | Match between rejected heat and preheating demand                                      |

The brine loop preheats the working fluid and buffers the thermal-oil circuit. The flue-gas oil loop supplies the higher-temperature increment through the evaporator. Because the wellhead is already drilled and lifted, capital cost is limited to surface heat exchange, piping, and workover. Brine leaves the exchanger and is returned to a disposal or re-injection well, so no surface discharge occurs.

## 3.2 Thermal stabilization by the brine loop

Refinery flue-gas duty fluctuates with crude throughput, furnace firing, and unit trips. Let the flue-gas enthalpy flux have mean  $\mu_{fg}$  and standard deviation  $\sigma_{fg}$ , and let the brine loop deliver a steady flux  $\mu_{br}$  with negligible variance on process timescales. The coefficient of variation of the combined stream is

$$CV_{hyb} = \frac{\sigma_{fg}}{\mu_{fg} + \mu_{br}} < CV_{fg} = \frac{\sigma_{fg}}{\mu_{fg}}$$

This variance reduction is the formal meaning of thermal stabilization. It reduces expander part-load excursions and allows a higher, steadier capacity factor. The economic model in Section 4 adopts  $CF = 0.90$  for the hybrid system against  $CF = 0.75$  for a stand-alone waste heat unit. These values are assumptions.

## 3.3 Working-fluid selection methodology

Isobutane, isopentane, and pentane were retained after screening against six criteria: (i) critical temperature relative to the 140–160 °C evaporator window; (ii) pressure levels at 90 °C condensation; (iii) volumetric expansion ratio; (iv) flammability and safety class; (v) environmental properties (zero ODP and very low GWP for all three hydrocarbons); and (vi) local availability. The last criterion is a practical advantage, because refineries produce

butanes and pentanes in their own LPG and natural-gasoline streams, which makes make-up fluid available on site. Table 2 lists their properties.

**Table 2** Fluid properties (CoolProp).

| Fluid      | $T_c$ (°C) | $p_c$ (MPa) | $p_{sat}$ at 90 °C (MPa) | $M$ (kg/kmol) |
|------------|------------|-------------|--------------------------|---------------|
| Isobutane  | 134.66     | 3.629       | 1.642                    | 58.12         |
| Isopentane | 187.20     | 3.378       | 0.579                    | 72.15         |
| n-Pentane  | 196.55     | 3.368       | 0.471                    | 72.15         |

The pentanes are subcritical throughout the window and evaporate at moderate pressure (1.6–1.7 MPa at the reported temperatures). Isobutane is the only fluid whose critical temperature lies inside the window.

### 3.4 Transcritical boundary conditions for isobutane at 145 °C

At an evaporator outlet of 145 °C (418.15 K), isobutane is 10.3 K above its critical temperature (407.81 K). The reduced temperature is

$$T_r = \frac{418.15}{407.81} = 1.025$$

so there is no isothermal evaporation. The cycle is transcritical: liquid is pumped above the critical pressure and heated continuously through the pseudo-critical region to state 3. The model uses  $p_{high} = 1.10 p_c = 3.99$  MPa, so  $p_r = 1.10$ . The condensing pressure is 1.64 MPa ( $T_{r,cond} = 0.890$ ), giving an expansion pressure ratio of 2.43. Four consequences follow.

1. **Heat-addition path.** As shown in Fig. 1, the heat-addition curve never crosses the saturation dome and has a smooth, non-isothermal profile. This matches the sensible-heat profile of a hot-oil or brine stream better than an isothermal boiling plateau.
2. **Property sensitivity.** Specific heat peaks near the pseudo-critical point, so small changes in  $T_{ev}$  or  $p_{high}$  produce large changes in specific work. This sensitivity is quantified in Section 5.3.
3. **Mechanical duty.** Design pressure must exceed 4.4 MPa (10% margin), which raises heat-exchanger and pipework cost against the pentanes.
4. **Pump work.** Pump work is higher than for subcritical cycles because of the elevated  $p_{high}$  (15.3 kW of pump-related exergy destruction at Warri, against 3.7 kW at Port Harcourt).

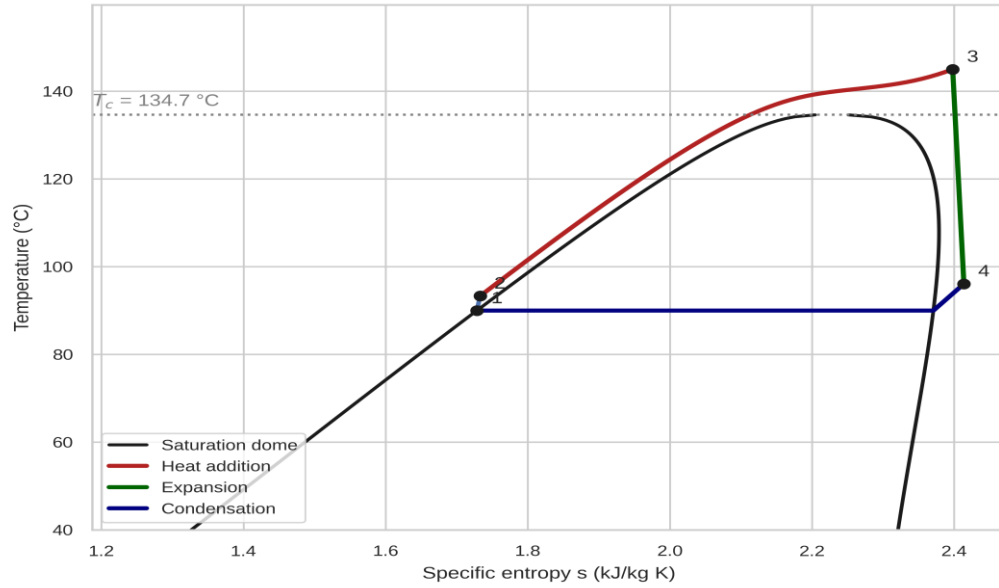


**Fig. 1** Temperature–entropy diagram for isobutane at the Warri refinery (transcritical,  $T_{ev} = 145\text{ °C}$ ,  $p_{high} = 3.99\text{ MPa}$ ,  $p_{low} = 1.64\text{ MPa}$ ). The dotted line marks the critical temperature ( $134.7\text{ °C}$ ).

States: 1 saturated liquid at  $90\text{ °C}$ ; 2 pump exit; 3 heater exit (supercritical); 4 expander exit. Pump and expander isentropic efficiencies are 0.75 and 0.80. Property data from CoolProp.

As shown in Fig. 2, isopentane at Port Harcourt operates as a conventional subcritical cycle. Saturated vapor leaves the evaporator at  $140\text{ °C}$  and  $1.57\text{ MPa}$ , well below the fluid’s  $187.2\text{ °C}$  critical temperature, and expansion terminates in the superheated region because isopentane is a dry fluid.

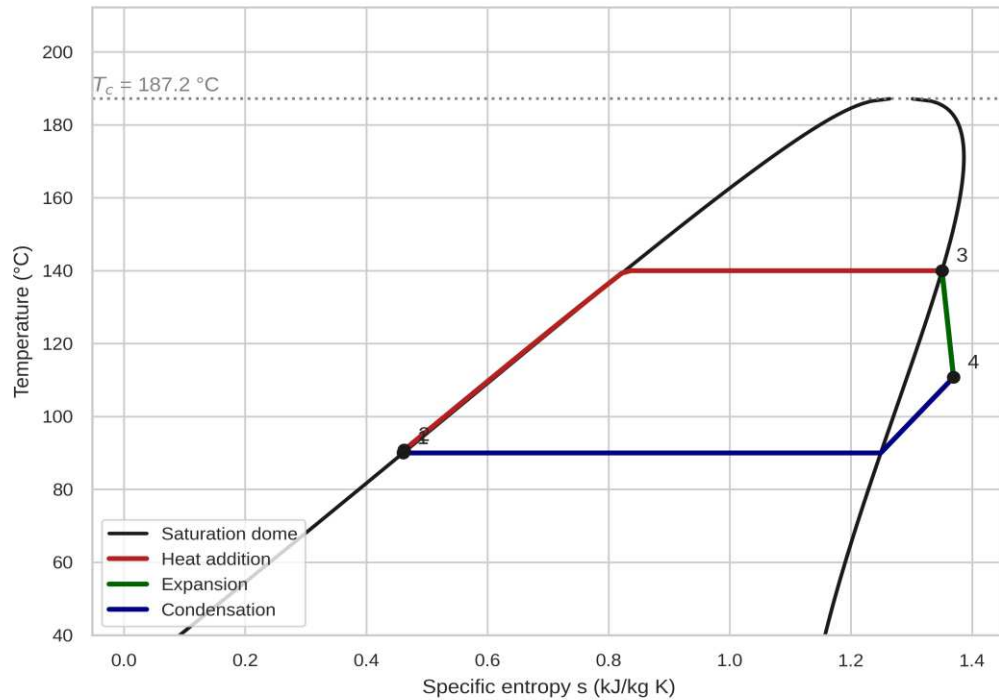


**Fig. 2** Temperature–entropy diagram for isopentane at the Port Harcourt refinery (subcritical,  $T_{ev} = 140\text{ °C}$ ,  $p_{high} = 1.57\text{ MPa}$ ,  $p_{low} = 0.58\text{ MPa}$ ).

Dry-fluid expansion terminates in the superheated region, so no droplet erosion risk arises in the expander.

Figure 3 shows the pentane cycle at the Kaduna CDU, subcritical at 155 °C and 1.74 MPa with the highest expansion ratio of the three (3.69).

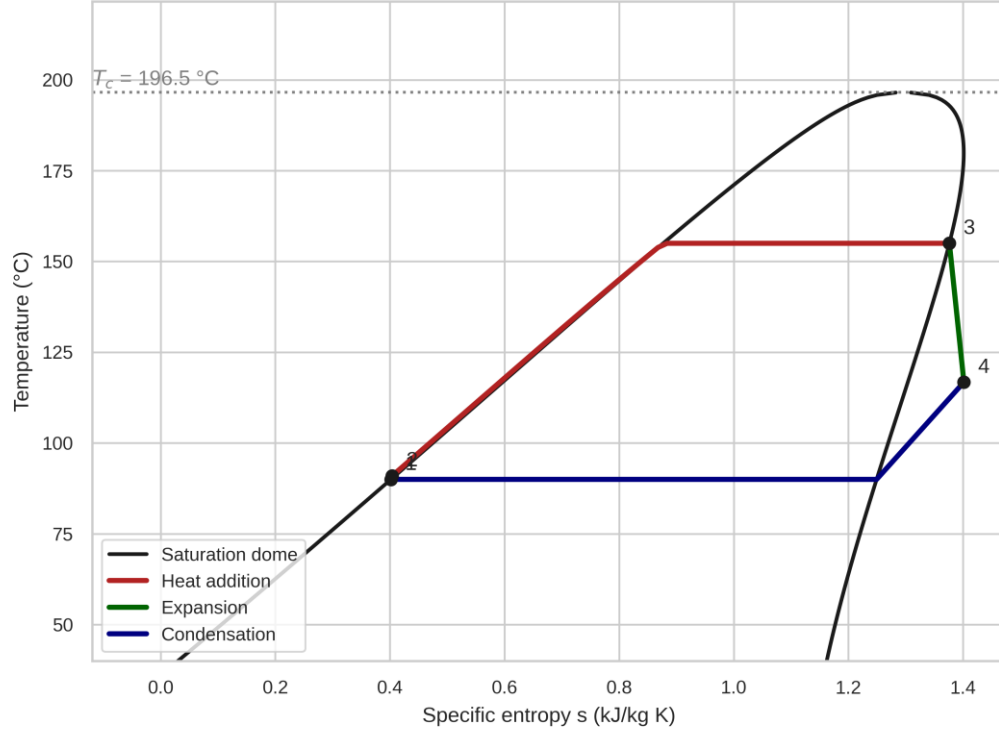


**Fig. 3** Temperature–entropy diagram for n-pentane at the Kaduna CDU (subcritical,  $T_{ev} = 155$  °C,  $p_{high} = 1.74$  MPa,  $p_{low} = 0.47$  MPa).

## 4 Mathematical Formulation and Advanced Exergy Analysis

### 4.1 Conservation laws

For a control volume at steady state, mass and energy balances are

$$\sum \dot{m}_{in} = \sum \dot{m}_{out}, \quad \dot{Q}_{cv} - \dot{W}_{cv} = \sum \dot{m}_{out} h_{out} - \sum \dot{m}_{in} h_{in}$$

with kinetic and potential terms neglected. The entropy balance is

$$\dot{S}_{gen} = \sum \dot{m}_{out} s_{out} - \sum \dot{m}_{in} s_{in} - \sum_j \frac{\dot{Q}_j}{T_j} \geq 0$$

Thermodynamic states are evaluated with CoolProp Helmholtz-energy equations (Bell et al. 2014).

### 4.2 Physical exergy and the dead state

The specific physical exergy of a stream is

$$e = (h - h_0) - T_0(s - s_0)$$

with dead state  $T_0 = 298.15$  K and  $P_0 = 101.325$  kPa. The exergy of a sensible heat carrier with capacity rate  $C_s = \dot{m}_s c_{p,s}$  cooled from  $T_{in}$  to  $T_{out}$  is

$$\dot{E}x_s(T) = C_s \left[ (T - T_0) - T_0 \ln \frac{T}{T_0} \right]$$

so the source exergy delivered to the cycle is  $\dot{E}x_s(T_{in}) - \dot{E}x_s(T_{out})$ . For the flue-gas stream, this factor is 0.31 of the enthalpy above the dead state at 335 °C.

### 4.3 Component models

The four ORC states follow from

$$h_2 = h_1 + \frac{h_{2s} - h_1}{\eta_p}, \quad h_4 = h_3 - \eta_e (h_3 - h_{4s})$$

with  $\eta_p = 0.75$  and  $\eta_e = 0.80$ . Net specific work, heat input, and thermal efficiency are

$$w_{net} = (h_3 - h_4) - (h_2 - h_1), \quad q_{in} = h_3 - h_2, \quad \eta_{th} = \frac{w_{net}}{q_{in}}$$

For subcritical operation, state 3 is saturated vapor at  $T_{ev}$ . For transcritical operation, state 3 is evaluated at  $(T_{ev}, p_r, p_c)$ . Evaporator feasibility requires a minimum source-to-working-fluid temperature difference along the counter-current profile:

$$\min_{h \in [h_2, h_3]} [T_s(h) - T_{wf}(h)] \geq \Delta T_{pp} = 10 \text{ K}, \quad T_s(h) = T_{s,out} + \frac{\dot{m}_{wf}(h - h_2)}{C_s}$$

### 4.4 Fuel–product exergy balance and closure

The fuel is the total exergy delivered by the hybrid heat carrier relative to the dead state,  $\dot{E}_F = \dot{E}x_s(T_{in})$ . The product is the net electric output,  $\dot{E}_P = \dot{W}_{net}$ . Exergy destruction and loss follow from

$$\dot{E}_D = \dot{E}_F - \dot{E}_P$$

and exergy efficiency from

$$\psi = \frac{\dot{E}_P}{\dot{E}_F} = \frac{\dot{W}_{net}}{\dot{W}_{net} + \dot{E}_D}$$

Here  $\dot{E}_D$  deliberately includes the exergy that leaves with the stack stream and the exergy rejected in the condenser, so that the accounting is conservative. Condenser heat that serves water preheating is not credited as product in  $\psi$ .

The reported site-level values are mutually consistent under this definition:  $\psi = W/(W + E_D)$  gives 0.120, 0.130, and 0.110 for the three sites. The implied fuel exergy is 1,500 kW at Warri (180 + 1,320), 1,538 kW at Port Harcourt (200 + 1,338), and 1,545 kW at Kaduna (170 + 1,375). Dividing by the source exergy factor gives an implied enthalpy flux above the dead state of 4.9, 4.9, and 5.3 MW. This is consistent with the 4–6 MW hybrid scale of the co-production concept, which we interpret here as a thermal scale.

The surrogate represents the combined flue-gas and brine streams as one equivalent heat carrier, whose capacity rate follows from the fuel exergy:  $C_s = 16.7, 15.8, \text{ and } 19.3 \text{ kW/K}$  at Warri, Port Harcourt, and Kaduna. This is a bookkeeping device. It is not a description of the actual streams, and Section 5.6 converts it into the aggregate flue-gas duty required, set against the single-stack 649 kW estimate, for realistic brine temperatures.

Component contributions are

$$\begin{aligned} \dot{E}_{D,stack} &= \dot{E}x_s(T_{out}), & \dot{E}_{D,exp} &= \dot{m}T_0(s_4 - s_3), & \dot{E}_{D,pump} &= \dot{m}T_0(s_2 - s_1) \\ \dot{E}_{D,cond} &= \dot{m}(e_4 - e_1), & \dot{E}_{D,evap} &= [\dot{E}x_s(T_{in}) - \dot{E}x_s(T_{out})] - \dot{m}(e_3 - e_2) \end{aligned}$$

These five terms sum exactly to  $\dot{E}_D$ . The cycle-internal destruction, excluding the unrecovered stack exergy, is  $\dot{E}_{D,cyc} = \dot{E}_D - \dot{E}_{D,stack}$ .

#### 4.5 Advanced exergy decomposition

For each component  $k$ , the destruction splits into endogenous and exogenous parts:

$$\dot{E}_{D,k} = \dot{E}_{D,k}^{EN} + \dot{E}_{D,k}^{EX}$$

The endogenous part is the destruction the component would produce if it alone were real and all others were ideal ( $\eta = 1$ ), evaluated at the same net product. The exogenous part is caused by the irreversibility of the other components, through changes in flow rate, temperature profile, and duty. Independently, the destruction splits into avoidable and unavoidable parts:

$$\dot{E}_{D,k} = \dot{E}_{D,k}^{UN} + \dot{E}_{D,k}^{AV}, \quad \dot{E}_{D,k}^{UN} = \dot{E}_P \left( \frac{\dot{E}_{D,k}}{\dot{E}_P} \right)^{UN}$$

The unavoidable ratio is evaluated at best-practice conditions:  $\eta_e = 0.90$ ,  $\eta_p = 0.85$ ,  $T_{ev}$  at the 160 °C window limit, and the same heat duty. Condensation stays at 90 °C, because that temperature is fixed by the preheating service. The four combined terms (AV-EN, AV-EX, UN-EN, UN-EX) follow by intersecting the two splits. This paper reports the endogenous/exogenous split only for the evaporator and the avoidable/unavoidable split for every component; the four-way split is not computed.

#### 4.6 Multi-objective problem statement

The optimization problem is

$$\min_{\mathbf{x}} \mathbf{f} = [-\dot{W}_{net}(\mathbf{x}), \dot{E}_{D,cyc}(\mathbf{x})], \quad \mathbf{x} = [T_{ev}, p_r, \varphi]$$

subject to

$$140 \leq T_{ev} \leq 160 \text{ °C}, \quad 1.05 \leq p_r \leq 1.40, \quad 0.5 \leq \varphi \leq 1.3$$

$$\Delta T_{pinch} \geq 10 \text{ K}, \quad T_{s,out} \geq 373.15 \text{ K}$$

Here  $\varphi$  is the fraction of the reference heat duty drawn from the source. The cycle-internal destruction  $\dot{E}_{D,cyc}$  is used as the second objective, because the total  $\dot{E}_D$  differs from  $-\dot{W}_{net}$  only by a constant at fixed source exergy, which would collapse the front to a single point.

#### 4.7 Techno-economic model

Equipment cost follows bare-module costing (Turton et al. 2018):

$$\log_{10} C_p^0 = K_1 + K_2 \log_{10} X + K_3 (\log_{10} X)^2, \quad C_{BM} = C_p^0 F_{BM} \frac{\text{CEPCI}}{\text{CEPCI}_{2001}}$$

For heat exchangers,  $X$  is area (m<sup>2</sup>) with  $F_{BM} = B_1 + B_2 F_M F_P$ ,  $B_1 = 1.63$ ,  $B_2 = 1.66$ , and  $F_M = 2.73$  (SS/SS) on the evaporator, where brine and flue-gas duties impose corrosion requirements. Areas follow from  $A = \dot{Q}/(U \Delta T_{lm})$ , with  $U = 0.12$  kW/m<sup>2</sup>K on the gas side and 0.45 kW/m<sup>2</sup>K in the condenser. Pump cost uses shaft power as  $X$ . The expander-generator is costed as a packaged unit at 900 USD/kW, because the Turton axial-turbine correlation is not representative of sub-300 kW machines. CEPCI is taken as 800 (assumed), and an 18% contingency and fee is added. The well and brine-loop retrofit is a lump sum of 0.30 MUSD.

$$C_{tot} = 1.18 \sum_i C_{BM,i} + C_{well}$$

The capital recovery factor and levelized cost are

$$CRF = \frac{i(1+i)^n}{(1+i)^n - 1}, \quad LCOE = \frac{C_{tot} CRF + f_{OM} C_{tot}}{\dot{W}_{net} \cdot 8760 \cdot CF}$$

with  $i = 0.12$ ,  $n = 20$  years,  $f_{OM} = 0.04$ , and  $CF = 0.90$ . Simple payback is

$$\tau = \frac{C_{tot}}{\dot{W}_{net} 8760 CF \pi_e + Q_{use} \pi_h - f_{OM} C_{tot}}$$

where  $\pi_e = 0.20$  USD/kWh is the avoided electricity cost,  $Q_{use} = 0.25 \dot{Q}_{cond} \cdot 8760 CF$  is the useful condenser heat, and  $\pi_h = 10$  USD/MWh<sub>th</sub> is the displaced fuel value.

## 4.8 Environmental model

The annual CO<sub>2</sub> offset from displaced electricity and displaced boiler fuel is

$$\Delta m_{CO_2} = E_{el} \varepsilon_{el} + Q_{use} \varepsilon_{th}, \quad R = \frac{\Delta m_{CO_2}}{m_{CO_2,base}} \times 100\%$$

with  $\varepsilon_{el} = 0.60$  tCO<sub>2</sub>/MWh<sub>e</sub> (blended grid and captive generation),  $\varepsilon_{th} = 0.238$  tCO<sub>2</sub>/MWh<sub>th</sub> (natural-gas boiler at 85% efficiency), and a baseline emission  $m_{CO_2,base}$  for the reference utility block (captive power and water-heating fuel) of 10.5, 10.0, and 9.5 kt/yr at Warri, Port Harcourt, and Kaduna. These baselines are inputs that an energy audit must replace. The co-production uplift  $u$  multiplies output: stand-alone output is the hybrid output divided by  $1 + u$ , with  $u = 0.25$  as the central case.

## 5 Multi-Objective Optimization and Results

### 5.1 Optimization set-up and base-case results

NSGA-II (Deb et al. 2002) was implemented in Python with CoolProp property calls. The settings were population 60, 60 generations, binary tournament selection, simulated binary crossover ( $p_c = 0.9$ ,  $\eta_c = 15$ ), and polynomial mutation ( $p_m = 1/n$ ,  $\eta_m = 20$ ). Infeasible candidates (pinch below 10 K or stack temperature below 100 °C) received a penalty objective. The surrogate is calibrated to the reported operating points, so it reproduces  $\dot{W}_{net}$  and  $\dot{E}_D$  exactly at those points. The effective source inlet temperatures are 335 °C at Port Harcourt (reported) and 320 °C and 300 °C at Warri and Kaduna (assumed). Table 3 gives the base-case results.

**Table 3** Base-case results (reported values and surrogate-derived quantities).

| Quantity                 | Warri         | Port Harcourt | Kaduna      |
|--------------------------|---------------|---------------|-------------|
| Fluid                    | Isobutane     | Isopentane    | n-Pentane   |
| Regime                   | Transcritical | Subcritical   | Subcritical |
| $T_{ev}$ (°C)            | 145           | 140           | 155         |
| $\dot{W}_{net}$ (kW)     | 180           | 200           | 170         |
| $\dot{E}_D$ (kW)         | 1,320         | 1,338         | 1,375       |
| $\psi$                   | 0.12          | 0.13          | 0.11        |
| Implied $\dot{E}_F$ (kW) | 1,500         | 1,538         | 1,545       |
| $\dot{m}_{wf}$ (kg/s)    | 10.98         | 7.48          | 4.69        |
| $\dot{Q}_{in}$ (kW)      | 2,915         | 2,678         | 1,883       |
| $\eta_{th}$ (%)          | 6.18          | 7.47          | 9.03        |
| $T_{s,out}$ (°C)         | 145.3         | 165.3         | 202.3       |
| Pinch margin (K)         | 51.9          | 74.4          | 93.1        |

| $p_{high}/p_{low}$ (MPa) | 3.99/1.64 | 1.57/0.58 | 1.74/0.47 |
|--------------------------|-----------|-----------|-----------|
|--------------------------|-----------|-----------|-----------|

Two observations matter for interpretation. First, the thermal efficiencies of 6.2–9.0% are physically consistent with a 90 °C condensing temperature, for which the Carnot limit at 145 °C is 13.2%. Second, exergy efficiencies of 0.11–0.13 sit far below thermal-efficiency-based intuition, because the fuel exergy includes the source stream’s entire potential down to the dead state.

As shown in Fig. 4, the ORC block exergy efficiency is comparable across the three sites (0.11–0.13). It sits well below the 28.6% base exergy efficiency reported for the Warri refinery. The two numbers use different system boundaries and should not be compared directly. The ORC figure counts only net work as product, and it charges stack and condenser exergy to the cycle.

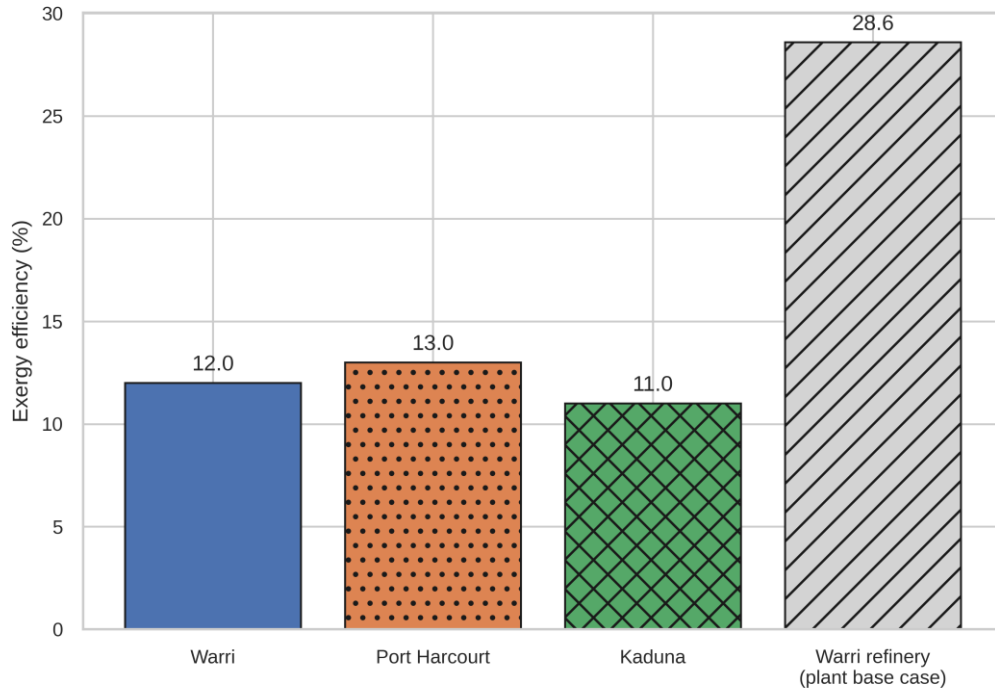


**Fig. 4** Base-case exergy efficiency by site.

ORC block (Warri, Port Harcourt, Kaduna) and, for reference, the reported Warri refinery base case (28.6%, hatched). The two values use different system boundaries and are not directly comparable.

## 5.2 Component-level exergy destruction

The Warri breakdown is (Fig. 5): condenser 490.7 kW (37.2%), evaporator 441.5 kW (33.4%), stack loss 320.8 kW (24.3%), expander 51.7 kW (3.9%), and pump 15.3 kW (1.2%). The two heat exchangers together account for 932.2 kW, or 70.6% of the total. Port Harcourt (condenser 451.2, evaporator 441.2, stack 399.3, expander 42.6, pump 3.7 kW) and Kaduna (stack 735.7, condenser 314.5, evaporator 286.4, expander 35.4, pump 2.9 kW) show the same dominance of heat transfer over machinery. The ranking of the largest term differs by site, however. At Port Harcourt the heat exchangers hold 66.7% of the total, but at Kaduna the unrecovered stack exergy is the largest single term (53.5%) and the heat exchangers hold 43.7%. The Kaduna stack loss reflects the 202 °C source outlet temperature, which leaves more exergy in the unrecovered stream.

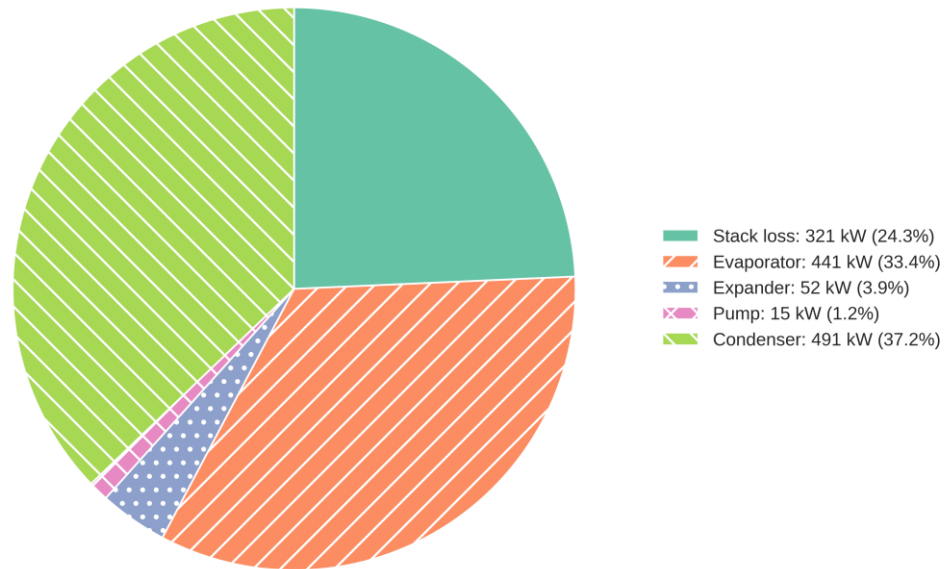


**Fig. 5** Component exergy destruction and loss breakdown for the Warri hybrid ORC (total 1,320 kW).

Heat exchangers (evaporator and condenser) account for 70.6% of the total, and stack loss for a further 24.3%. Turbomachinery contributes 5.1%.

Advanced exergy analysis of the Warri evaporator gives an endogenous destruction of 372.8 kW (84.4%) and an exogenous destruction of 68.7 kW (15.6%), as shown in Fig. 6. Port Harcourt (388.1 kW, 88.0% endogenous) and Kaduna (240.4 kW, 83.9%) are similar. The evaporator's loss is therefore mostly intrinsic to its own temperature-difference structure. Improving the expander or pump can recover at most about one-sixth of it. Table 4 splits the Warri destruction into unavoidable and avoidable parts.

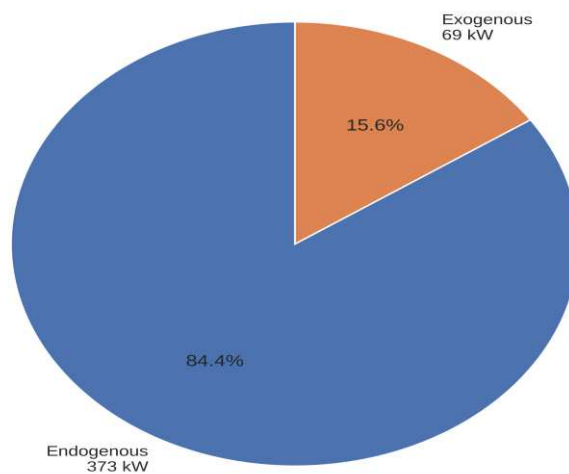


**Fig. 6** Endogenous versus exogenous exergy destruction in the Warri evaporator (total 441.5 kW).

The endogenous share (84.4%) is the destruction that would remain if every other component were ideal at the same net product.

**Table 4** Warri unavoidable/avoidable exergy destruction (kW).

| Component    | $\dot{E}_D$    | Unavoidable  | Avoidable    | Avoidable share |
|--------------|----------------|--------------|--------------|-----------------|
| Stack loss   | 320.8          | 247.8        | 73.0         | 22.8%           |
| Evaporator   | 441.5          | 320.8        | 120.7        | 27.3%           |
| Expander     | 51.7           | 18.7         | 33.0         | 63.9%           |
| Pump         | 15.3           | 5.0          | 10.4         | 67.5%           |
| Condenser    | 490.7          | 386.4        | 104.2        | 21.2%           |
| <b>Total</b> | <b>1,320.0</b> | <b>978.7</b> | <b>341.3</b> | <b>25.9%</b>    |

### 5.3 Pareto frontiers and parametric behavior

Figure 7 maps net power against cycle exergy destruction. Each front is a monotone trade-off: drawing more heat raises output and raises destruction almost linearly, because both scale with working-fluid flow. The fronts span 102–253 kW (Warri), 122–318 kW (Port Harcourt), and 89–231 kW (Kaduna) over the allowed duty range. Front members cluster at the upper temperature bound (158.6–160.0 °C at Warri, 158.7–160.0 °C at Port Harcourt, 159.3–160.0 °C at Kaduna). At Warri, the front also favors elevated pressure ( $p_r \approx 1.23$ –1.32).

The reported operating points are not on these fronts. At equal net power, the front achieves lower cycle destruction by 10.4% at Warri (896 versus 999 kW), 17.8% at Port Harcourt (771 versus 939 kW), and 4.6% at Kaduna (610 versus 639 kW). The reduced-order surrogate therefore identifies additional thermodynamic headroom. We interpret the difference as the price of constraints not represented in the surrogate, such as expander sizing, fluid thermal stability, pressure rating, and hybrid-carrier temperature limits. Confirming that interpretation requires the full HYSYS model.

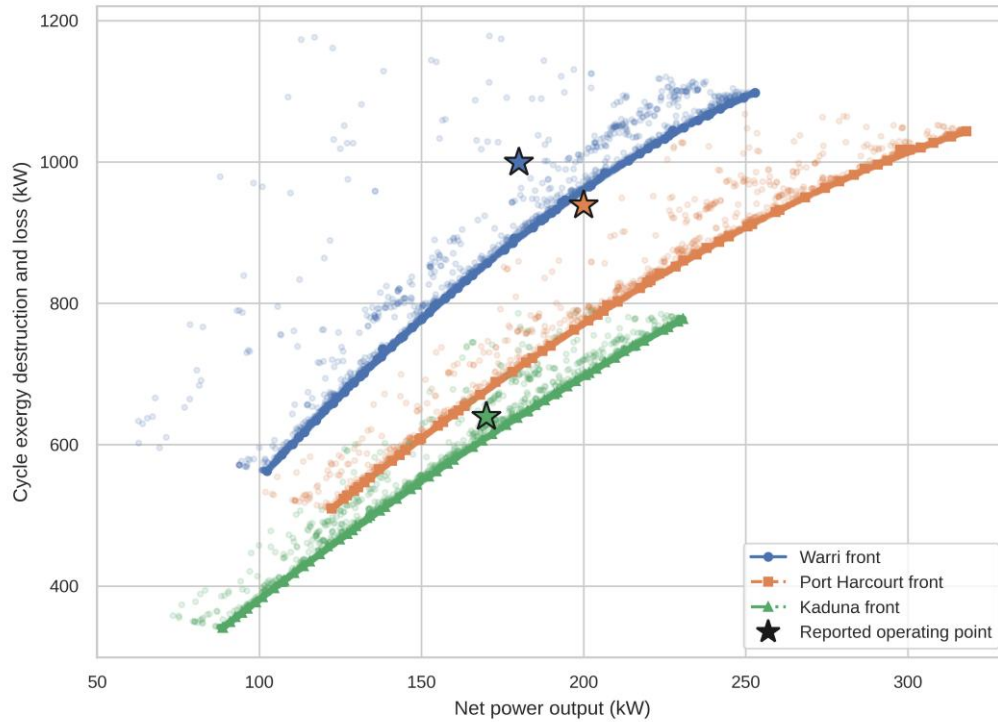


**Fig. 7** Pareto fronts (NSGA-II, population 60, 60 generations) of net power versus cycle exergy destruction for the three sites.

Faint points are all feasible evaluated designs, solid lines are non-dominated fronts, and stars mark the reported operating points.

Figure 8 isolates the effect of evaporator temperature at constant heat duty. Net power rises monotonically across the window. Over 140–160 °C, output rises from 142.2 to 194.7 kW at Warri (+36.9%), from 200.0 to 244.6 kW at Port Harcourt (+22.3%), and from 143.9 to 177.4 kW at Kaduna (+23.3%). At Warri the response is steepest just above the critical temperature. The mean slope is 7.6 kW/K between 140 and 145 °C and about 1.0 kW/K between 145 and 160 °C. This is consistent with the near-critical sensitivity discussed in Section 3.4, because the sweep holds  $p_r$  at 1.10 and crosses the pseudo-critical region. The steep segment is a property effect. It does not by itself prove that 145 °C is an optimum. At Port Harcourt and Kaduna no knee appears and the curves stay close to linear, so the reported temperatures of 140 and 155 °C cannot be explained by this figure.

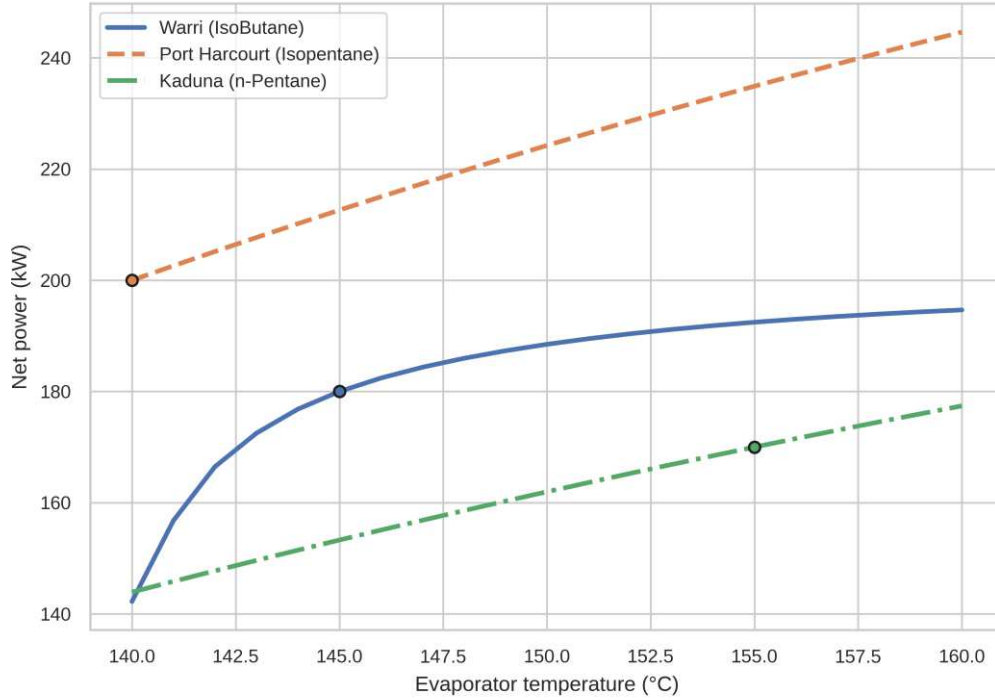


**Fig. 8** Parametric effect of evaporator temperature on net power output at constant heat duty, for each site’s selected fluid.

Circles mark the reported operating points. The steep rise for isobutane between 140 and 145 °C reflects near-critical property variation.

## 5.4 Why heat exchangers dominate unavoidable destruction

At Warri, heat exchangers hold 707.2 kW of the 978.7 kW unavoidable destruction (72.3%), and the source-side stack takes a further 25.3%. Machines hold only 2.4%. The heat-exchanger share is site-dependent: 65.9% at Port Harcourt and 43.7% at Kaduna, where the stack term dominates. Three mechanisms explain the Warri pattern.

1. **Temperature-difference irreversibility is set by the source, not the equipment.** Entropy generation in a heat exchanger scales as  $\dot{S}_{gen} = \dot{Q} (1/T_{wf} - 1/T_s)$ , roughly. A 320–335 °C source heating a working fluid that cannot exceed about 160 °C must destroy exergy across a large mean temperature difference (101, 125, and 127 K on the log-mean basis at Warri, Port Harcourt, and Kaduna). Better exchanger design narrows the pinch, but it cannot remove the mismatch fixed by the fluid’s thermal stability window and the 140–160 °C bound.
2. **Fixed heat-sink temperature.** Condensing at 90 °C is dictated by the preheating service. Rejected heat at that temperature carries exergy of  $\dot{Q}_{cond}(1 - T_0/363.15) \approx 0.18 \dot{Q}_{cond}$ , which appears in  $\dot{E}_D$  by construction. It is unavoidable for as long as the service temperature is fixed.

3. **Machines can approach ideality.** Expander and pump efficiencies can rise from 0.80 and 0.75 to about 0.90 and 0.85, which removes 64–68% of their destruction, but their absolute contribution is small.

The practical consequence is that design effort should concentrate on source-side matching, such as a thermal-oil loop that lowers the effective source temperature toward the cycle window, and on using the condenser heat productively, before turbomachinery refinement.

## 5.5 Fluid cross-check and surrogate limitations

At the reported evaporator temperatures and site duties, the surrogate produces the net powers for the three fluids given in Table 5.

**Table 5** Surrogate net power (kW) for each fluid at the reported site temperature and duty.

| Site                   | Isobutane | Isopentane | n-Pentane |
|------------------------|-----------|------------|-----------|
| Warri (145 °C)         | 180.0     | 231.4      | 237.2     |
| Port Harcourt (140 °C) | 130.7     | 200.0      | 204.7     |
| Kaduna (155 °C)        | 124.4     | 165.2      | 170.0     |

At Port Harcourt and Kaduna the reported fluid choices are within 2.3% and 0% of the surrogate maximum. At Warri, however, the surrogate ranks both pentanes above isobutane by 29–32% at the same duty and temperature. The isobutane selection at Warri therefore cannot be justified from cycle thermodynamics alone in the present surrogate. It rests on the site-level HYSYS and meta-analysis results and on non-thermodynamic considerations that we have not quantified here, such as availability of isobutane in refinery LPG streams. This is an open validation item.

The surrogate also assumes constant heat-carrier specific heat, a single lumped source in place of the flue-gas and brine streams, and equal  $\eta_e$  across fluids. These simplifications are appropriate for screening but not for final design.

## 5.6 Heat-source closure and required multi-stream duty

The equivalent heat carrier of Section 4.4 needs an explicit reading of the reported 649 kW. We treat it as the recoverable heat of a single stack (Imene 2024), not as the site total. At Port Harcourt the equivalent carrier has a capacity rate of 15.8 kW/K, whereas 649 kW, if recovered down to the surrogate stack temperature of 165 °C, corresponds to only about 3.8 kW/K. The reported outputs therefore require aggregate recovery of roughly four times that single-stack duty. This is the multi-stream premise of the paper. Which refinery streams make up the aggregate (for example, the CDU and other fired-heater stacks) has not been identified and must be established by an energy audit. Section 5.7 screens this premise against published data.

Brine cannot close the gap on its own, because it can only preheat the working fluid. Table 6 gives the largest share of the working-fluid duty that brine can supply by sensible preheat, with a 10 K pinch and a brine specific heat of 4.0 kJ/kg K. Brine at 100 °C supplies nothing, because the pump-exit temperature is already 93–103 °C. Even with brine at 140 °C, the Port Harcourt cycle would still need 1.85 MW from flue gas, 2.8 times the 649 kW estimate.

**Table 6** Brine preheat contribution and remaining flue-gas duty (kW unless stated).

| Site          | Brine (°C) | Brine share of duty | Brine duty | Brine flow (m <sup>3</sup> /h) | Flue-gas duty needed |
|---------------|------------|---------------------|------------|--------------------------------|----------------------|
| Warri         | 120        | 19.4%               | 567        | 31                             | 2,348                |
| Warri         | 140        | 47.1%               | 1,372      | 34                             | 1,543                |
| Port Harcourt | 120        | 14.6%               | 392        | 18                             | 2,286                |
| Port Harcourt | 140        | 30.9%               | 828        | 19                             | 1,850                |
| Kaduna        | 120        | 12.9%               | 243        | 11                             | 1,641                |
| Kaduna        | 140        | 27.2%               | 512        | 12                             | 1,371                |

Three conclusions follow. First, the reported net powers of 170–200 kW require aggregate flue-gas recovery of 1.4–1.9 MW with 140 °C brine (1.54 MW at Warri, 1.85 MW at Port Harcourt, 1.37 MW at Kaduna) and 1.6–2.3 MW with 120 °C brine, against the 649 kW single-stack estimate at Port Harcourt. This aggregate is a requirement of the design, not a measured quantity, and the number of streams and their temperatures remain to be audited. Second, the brine’s role in this architecture is mainly preheating and buffering, not evaporation, and it is useful only if the wellhead temperature exceeds about 110 °C. Third, the brine flow needed is modest (11–34 m<sup>3</sup>/h), so the limiting well property is temperature, not rate. The 20–30% co-production uplift is adopted from the reported range. The surrogate does not derive it.

## 5.7 Benchmark check of the multi-stream premise

The multi-stream premise can be tested against the refinery configuration and published fuel-intensity data. Port Harcourt is two refineries. The old plant has a 60,000 bpd crude distillation unit and a 6,000 bpd catalytic reformer, and the new plant has a 150,000 bpd crude unit and a 150,000 bpd vacuum unit, for 210,000 bpd combined (BPE n.d.). Each unit has its own fired heaters and stacks, so several flue-gas streams exist by construction. The restart of the old crude unit was estimated at about 20,000 bpd against a 60,000 bpd nameplate, and the 150,000 bpd section was still offline at that time (Kpler 2024).

Fired duty is estimated from U.S. Department of Energy fuel intensities per barrel of feed: 106.8 kBtu/bbl for atmospheric distillation at current average practice, with a practical minimum energy of 50 kBtu/bbl, and 254 kBtu/bbl for reforming, with a practical minimum of 203 kBtu/bbl (Energetics 2006). Recoverable flue-gas heat is the sensible heat released in cooling the flue gas from the source temperature to the 165 °C surrogate stack temperature. The calculation assumes 21.6 kg of flue gas per kg of fuel gas (20% excess air), a fuel lower heating value of 48 MJ/kg, and a mean specific heat of 1.10 kJ/kg K. This gives 4.2%, 6.7%, and 8.4% of fired duty at source temperatures of 250, 300, and 335 °C. The ranges in Table 7 pair the low fuel intensity with 250 °C and the high fuel intensity with 335 °C.

**Table 7** Screening estimate of fired duty and recoverable flue-gas heat at Port Harcourt. Ranges combine practical-minimum and current-average fuel intensities with source temperatures of 250–335 °C.

| Scenario                                          | Crude / reformer feed<br>(kbpd) | Fired duty<br>(MW) | Recoverable flue-gas heat<br>(MW) |
|---------------------------------------------------|---------------------------------|--------------------|-----------------------------------|
| Old crude unit at restart rate                    | 20 / 0                          | 12–26              | 0.5–2.2                           |
| Old crude unit at nameplate                       | 60 / 0                          | 37–78              | 1.5–6.6                           |
| Old crude unit and reformer at nameplate          | 60 / 6                          | 52–97              | 2.2–8.2                           |
| Old and new crude units and reformer at nameplate | 210 / 6                         | 143–292            | 6.0–24.7                          |

Three points follow. First, the reported 649 kW corresponds, at 335 °C, to about 7.7 MW of fired duty. That is roughly 10–21% of the fired duty of the old crude unit at nameplate (37–78 MW), so 649 kW reads as one heater or stack, not the site total. Second, the 1.85 MW of flue-gas duty that Section 5.6 requires with 140 °C brine lies inside the estimated range for the old refinery at nameplate (1.5–6.6 MW for the crude unit alone, 2.2–8.2 MW with the reformer) and below the range for both refineries (6.0–24.7 MW). Third, the check supports the premise only conditionally. At the 20,000 bpd restart rate the crude unit gives 0.5–2.2 MW, so 1.85 MW is reached only at the top of the range. The estimate also uses U.S. average fuel intensities, assumes every heater exhausts at 250–335 °C, and ignores acid dew-point limits, existing air preheaters, and the boilers and turbo-generators on site. It is a plausibility check, not a measurement, and measured fuel-gas flow and stack temperature for each heater would replace it.

## 6 Implementation, Policy Implications, and Conclusion

### 6.1 CO<sub>2</sub> reduction and payback

With the assumptions in Section 4, projected CO<sub>2</sub> reductions of the hybrid system are 20.3% at Warri, 21.1% at Port Harcourt, and 16.9% at Kaduna, or 2,134, 2,109, and 1,608 tCO<sub>2</sub>/yr (Fig. 9). Varying the co-production uplift over 20–30% moves the results within 19.5–21.1%, 20.2–21.9%, and 16.2–17.6%. Stand-alone waste heat recovery, with the lower capacity factor of 0.75 and no brine uplift, delivers 13.6%, 14.1%, and 11.3%. Co-production therefore lifts every site into the 15–25% band, while all three stand-alone units fall below it. The 15–25% range and the sub-5-year payback are targets drawn from the literature synthesis this work builds on (IEA 2023, 2024), and the scenario model above tests whether they are attainable under explicit assumptions. It does not independently confirm them. These percentages depend directly on the assumed utility-block baseline and must be recomputed from audited data.

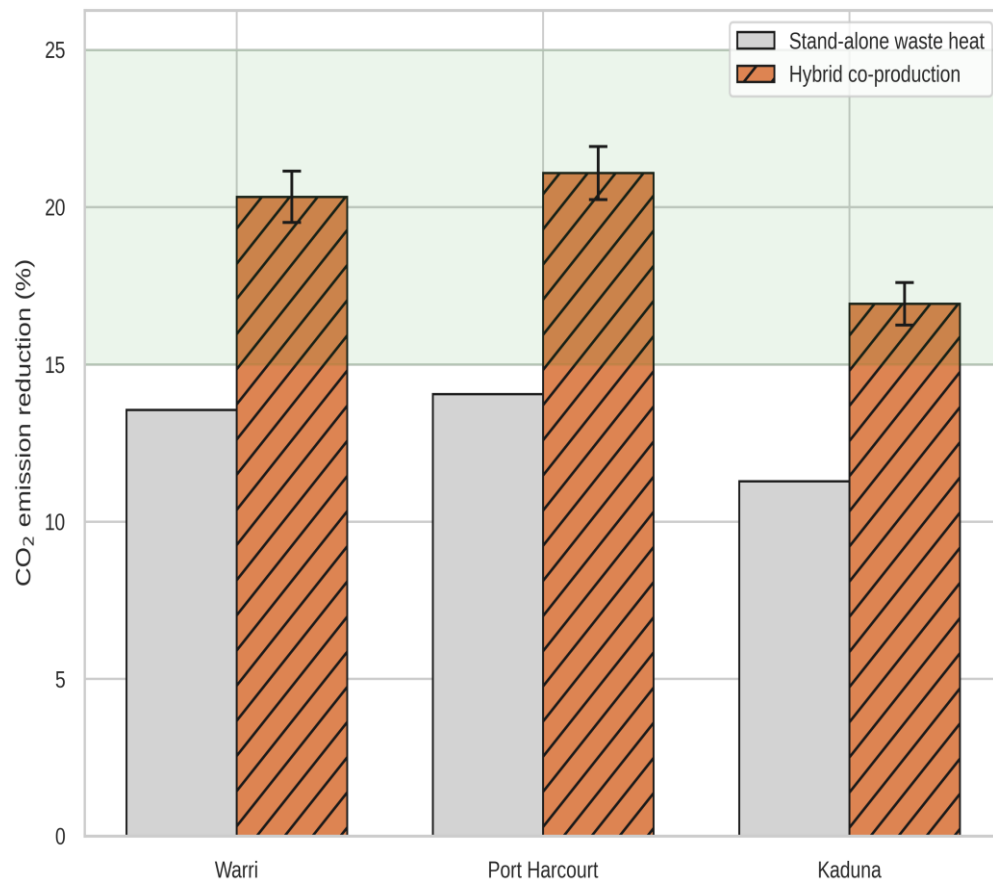


**Fig. 9** Projected CO<sub>2</sub> emission reductions (%) for stand-alone waste heat recovery and for hybrid co-production.

Error bars show the 20–30% co-production uplift range. The shaded band marks the 15–25% target range. Baselines are utility-block assumptions.

The hybrid capital cost is 1.34 MUSD at Warri, 1.24 MUSD at Port Harcourt, and 1.11 MUSD at Kaduna, or 6,200–7,400 USD/kW. At the base avoided cost of 0.20 USD/kWh, simple payback is 4.7, 3.9, and 4.3 years, and the levelized cost of electricity is 0.164, 0.137, and 0.143 USD/kWh. The 5-year threshold is crossed at 0.189, 0.158, and 0.173 USD/kWh (Fig. 10). Payback lengthens to 6.6–7.8 years at 0.12 USD/kWh and shortens to 2.6–3.1 years at 0.30 USD/kWh. The economics are therefore clearly favourable where the avoided cost reflects captive generation, and marginal where it reflects grid tariffs alone.

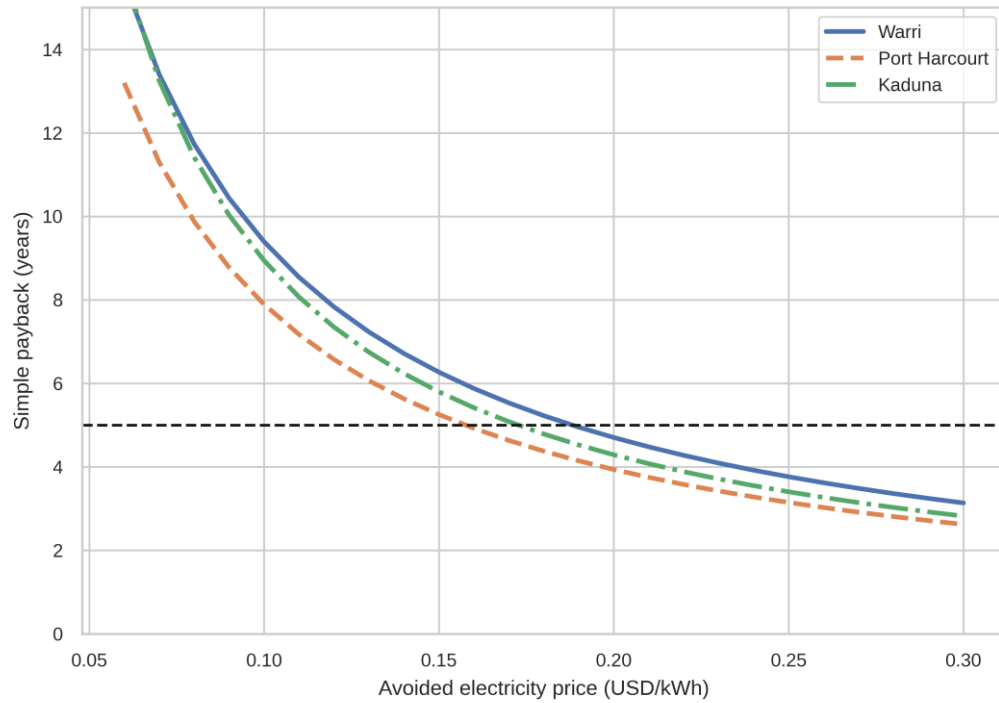


**Fig. 10** Simple payback sensitivity to the avoided electricity price for the three hybrid systems.

The dashed line marks the 5-year threshold.

On simple payback, the hybrid and stand-alone systems are close (4.7 versus 4.7 years at Warri, 3.9 versus 3.8 at Port Harcourt, 4.3 versus 4.0 at Kaduna). The additional 0.30 MUSD brine-loop investment is repaid by higher output and a higher capacity factor. The hybrid delivers about 50% more annual energy (1,419 versus 946 MWh at Warri). Co-production is thus justified by energy delivered and firmness of supply more than by faster payback.

## 6.2 Implementation guidelines

1. **Site the brine loop where wells and the refinery are co-located or connected by short piping.** Mature wells at or near terminals and gathering stations are the first candidates.
2. **Audit before design.** Measure brine temperature, flow, and chemistry, and log flue-gas duty over several months. Identify every flue-gas stream that contributes to the aggregate duty of 1.4–1.9 MW (with 140 °C brine) required in Section 5.6. The 649 kW single-stack estimate and the utility-block baselines are the least certain inputs.
3. **Use an intermediate thermal-oil loop for the flue-gas duty.** It decouples stack-temperature limits from the ORC and protects the organic fluid from thermal decomposition.
4. **Match the condenser to a real heat sink.** The 90 °C rejection is only valuable if the water-preheating demand exists. Otherwise, lower the condensing temperature and accept a different optimum.
5. **Select the fluid with site validation.** Confirm the fluid ranking with full HYSYS or equivalent models, including safety review for flammable A3 fluids.
6. **Stage the rollout.** Begin with one 150–250 kW pilot at Port Harcourt (highest output, shortest payback), then scale to hybrid installations of 4–6 MW thermal.

## 6.3 Policy implications

The Petroleum Industry Act 2021 and associated licensing rules govern upstream assets and decommissioning obligations (provisions on well repurposing need legal review). A clear regulatory route for repurposing wells for

heat extraction, including ownership of heat rights on an oil lease, would remove a barrier to co-production. Carbon-market participation and recognition of waste heat recovery under Nigeria's energy-transition framework would improve project bankability. Embedded and captive generation rules should permit refinery-side sales of surplus ORC power. Finally, public support for pilot demonstration would help resolve the uncertainties identified above.

## 6.4 Conclusion

This study presented an integrated oil–gas–geothermal framework in which co-produced mature-well brine stabilizes the heat input to an ORC evaporator driven by refinery waste heat. Site-specific optima reached 180 kW (isobutane, 145 °C, transcritical, Warri), 200 kW (isopentane, 140 °C, Port Harcourt), and 170 kW (pentane, 155 °C, Kaduna), a 170–200 kW range. Exergy efficiencies were 0.12, 0.13, and 0.11, with exergy destruction of 1,320, 1,338, and 1,375 kW. Advanced exergy analysis showed that heat exchangers account for about 71% of destruction at Warri and 72% of its unavoidable part, with 84% of evaporator destruction endogenous. At Kaduna the unrecovered stack exergy dominates instead. Co-production raised output by 20–30% at a 4–6 MW hybrid thermal scale, with payback of 3.9–4.7 years and CO<sub>2</sub> reductions of 16.9–21.1% under stated assumptions.

Three limitations bound these conclusions. The economic and CO<sub>2</sub> figures depend on assumed prices, capacity factors, and utility baselines. The reduced-order surrogate places the reported operating points inside the dominated region, and it ranks the pentanes above isobutane at Warri. The source inlet temperatures at Warri and Kaduna are assumed, and the reported outputs assume aggregate multi-stream flue-gas recovery of about 1.85 MW at Port Harcourt with 140 °C brine, 2.8 times the single-stack 649 kW estimate, with the contributing streams not yet audited. A benchmark screening estimate brackets that duty at the old refinery's nameplate throughput, but at the restart throughput of about 20,000 bpd it reaches 1.85 MW only at the top of its range (Section 5.7). Plant operating status and data quality should also be confirmed for each refinery, since some units have been under rehabilitation. Full-process validation with plant data is therefore the necessary next step.

## Statements and Declarations

**Funding.** None

**Author contributions.** AMEH Daniel Gbemisoke: Conceptualization, Methodology, Software, Validation, Formal analysis, Investigation, Resources, Data Curation, Writing - Original Draft, Writing - Review & Editing, Visualization, Project administration.

**Ethics approval.** Not applicable. This study did not involve human participants or animals.

**Data availability.** All numerical results are generated by the scripts in Online Resources 1 and 2 from the inputs stated in the paper. Third-party inputs are available from the cited sources.

**Code availability.** The simulation and screening scripts are provided as Online Resources 1 and 2.

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