

1 **Pilot-scale bioelectrochemical system (BES) for simultaneous nitrogen and carbon**
2 **removal in urban wastewater treatment plants.**

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20 ABSTRACT

21 This study aims to characterize the performance of a 150 L bioelectrochemical system-
22 based plant, during the simultaneous carbon and nitrogen removal from several waste
23 streams of wastewater treatment plants. The bioelectrochemical system (BES) contained
24 five electrode pairs (operated hydraulically and electrically in parallel) and was fed with
25 either wastewater, centrate (nutrient-rich liquid stream produced during the dewatering
26 of digested biomass), or a mixture of both over 63 days, with a hydraulic retention time
27 of one day. Total Organic Carbon (TOC) and Total Nitrogen (TN) removal rates
28 averaged 80% and 70%, respectively, with a specific energy consumption of 0.18
29 kWh·m⁻³ (BES + ancillary equipment). This work also underlines the challenges of
30 using BES for nitrogen removal, highlighting the limitations of the current design, and
31 suggesting some strategies for improvement.

32 INTRODUCTION

33 The presence of nitrogen in wastewater (WW) has for years been a source of health and
34 environmental issues (1)(2). For urban WW (uWW) in particular, nitrogen is usually
35 present as organic nitrogen, ammonium, and nitrates (3), and it is conventionally
36 removed by a two-step process that involves autotrophic aerobic oxidation of ammonia
37 (nitrification), and heterotrophic reduction of nitrate and nitrite to nitrogen gas
38 (denitrification) (4).

39 The uWW entering a WW treatment facility, which represents the main source of
40 nitrogen for the water line, usually contains a low concentration of nitrogen that is
41 typically around $45 \text{ mg}\cdot\text{L}^{-1}$ (5). Another important source of nitrogen comes from the
42 sludge centrifugation (after anaerobic digestion) which produces a liquid stream, known
43 as centrate (6), that is usually returned to the head of the water line. Nitrogen content in
44 centrates can exceed $1300 \text{ mg}\cdot\text{L}^{-1}$ (total Kjeldahl nitrogen) (7) and may account for 15–
45 20% of the nitrogen entering the plant (8). Its removal demands extra power input and
46 the implementation of larger treatment tanks in the water line (9), all of which have a
47 significant impact on the economic balance of the plant. Moreover, as pollutants in the
48 centrate get diluted when entering the uWW, their elimination becomes more difficult.
49 Thus, developing new strategies for treating centrates without their being returned to the
50 water line may represent an important source of monetary and energy savings for the
51 plant.

52 Bioelectrochemical systems (BESs), are an emerging technology with a wide range of
53 potential applications in the field of WW treatment (10). One of their main benefits,
54 compared to more conventional technologies, is that they allow the recovery of part of

55 the chemical energy content present in the WW, while removing organic pollutants.
56 BESs can also be used for removing inorganic contaminants such as nitrogen (11), and
57 when designed for simultaneous carbon and nitrogen removal they have the potential to
58 become an effective and efficient WW treatment technology (12)(13). Thus, several
59 strategies for simultaneous carbon and nitrogen removal from different waste streams
60 using BESs have already been put forward, and have been validated at the laboratory
61 scale (14)(15). These experiments have provided valuable information for determining
62 the optimal operating conditions such as pH and temperature ranges, Chemical Oxygen
63 Demand (COD)/N ratio or materials (16). However, larger-scale experiments are vital
64 for evaluating how laboratory designs can be implemented at a practical scale, to
65 identify where the limitations are, and to assess the chances of practical applicability
66 (17).

67 The aim of this study is to obtain a preliminary characterization of the operation of a
68 150 L BES-based pilot plant (Fig. 1) for simultaneous carbon and nitrogen removal
69 from two waste effluents: uWW and centrate. The BES unit consist of five independent
70 modules hydraulically and electrically connected in parallel. The fresh waste effluent is
71 first fed to anode side of each module for organic matter oxidation. Then it is conducted
72 to a packed-bed column reactor to convert ammonia into nitrate. Finally, the effluent of
73 the column is led to the cathode side of each module for denitrification. The plant was
74 operated in continuous mode for 63 days, and this paper presents results for TOC and
75 TN removal rates, as well as energy usage/recovery. It also highlights the main
76 limitations of the current design and advances some strategies for improvement.

77 The BES was designed within a framework of easy-to-manufacture, easy-to-maintain
78 considerations to improve its scalability and its practical applicability. Although the
79 whole set-up was initially planned to operate with centrate as the feed, the results show
80 that it can also work with urban wastewater, which highlights its flexibility to operate
81 both in the water line or in the sludge line in a WWTP.

82

83 MATERIALS AND METHODS

84 **BES and nitrification reactors design**

85 The BES reactor consisted of a polypropylene tank (0.7 m x 0.32 m x 1.2 m) with a
86 working liquid volume of 150 L (Fig. 1). It contained five electrode pairs that were
87 mounted on polypropylene frames, and were hydraulically and electrically operated in
88 parallel. The anodes and cathodes were made of 5 mm-thick graphite felt (Sigratherm®,
89 Germany), with a projected area of 0.47 m² per electrode (0.98 m x 0.48 m) and were
90 pretreated according to (18). Each anode and cathode were separated by an anion
91 exchange membrane, AMI-7001 CR (Membranes International INC., USA), and the
92 electrodes were connected to the external electrical circuit through stainless steel current
93 collectors. Thus, the reactor was divided into six chambers (three anode chambers and
94 three cathode chambers) with a volume of 30 L each, except for the two outer chambers
95 which have half that volume. Every chamber was fitted with a recirculation loop and a
96 pump to maintain mixing conditions.

97 Nitrification was carried out in an external PVC, tubular, fixed-bed reactor (0.25 m in
98 diameter and 1.7 m in height) which was filled with plastic rings (4 cm in diameter) to

99 provide a high surface area. A peristaltic pump was used to continuously recirculate the
100 effluent.

101 **Reactor inoculation and start-up**

102 During the inoculation and start-up process, the BES reactor was operated as an MEC
103 by applying 1 V between the anode and cathode. The anode was inoculated with WW
104 collected from the aerobic zone of an activated sludge reactor (WWTP, north-west of
105 Spain) and was supplemented with $0.5 \text{ g}\cdot\text{L}^{-1}$ of acetate. The cathode was inoculated
106 with WW collected from the denitrification zone, and was supplemented with 50 mg N-
107 $\text{NO}_3\cdot\text{L}^{-1}$. Finally, the inoculum for the nitrification reactor consisted of WW collected
108 from the nitrification zone, and was supplemented with $100 \text{ mg N-NH}_4\cdot\text{L}^{-1}$. The anode
109 and cathode chambers were filled simultaneously to avoid differential pressures that
110 may damage the membrane. The start-up process was completed in nine days.

111 **Operating conditions**

112 The reactor was initially operated in MFC mode by connecting electrical resistances
113 between the anode and the cathode. Voltage across resistances was measured
114 continuously, and current was determined using Ohm's law. The ohmic values of the
115 external resistances were determined according to the results of polarization tests (that
116 were performed daily) to maximize the power output.

117 When the BES was operated in MEC mode the resistances were removed and a voltage
118 of 1 V was applied between anode and cathode by means of power supplies that were
119 computer controlled using a PCI-6713 analog output board (National Instruments,

120 Texas). The current was measured and recorded every 10 minutes across a 16 Ω fixed
121 resistor, using a data acquisition system.

122 Raw wastewater was pumped from the feed tank (maintained at 4 °C) to the anode
123 chambers by C1R peristaltic pumps (Dosiper, Spain). Identical pumps were used to
124 drive the water from the column to the cathode chambers. The BES and nitrification
125 reactors were operated at room temperature. Wastewater was recirculated by DC15/5
126 centrifugal pumps (Xylem, UK) in the anode and cathode chambers, and with NLAV
127 peristaltic pumps (Dosiper, Spain) in the external tubular fixed bed (Fig. 2). Hydraulic
128 retention in the BES was fixed at one day.

129 The pilot plant was meant to be operated and evaluated within the facilities of the
130 WWTP. However, due to project constraints, this was not possible, and the pilot plant
131 performance was evaluated within the facilities of the authors' laboratory. Therefore,
132 WW and centrate samples were received regularly from the WWTP, and were stored at
133 4 °C for less than 10 days.

134 Following the start-up period, the BES was operated as an MFC. Since no significant
135 current was produced, the BES was switched to MEC mode for a period of 63 days that
136 was further subdivided into three stages. During the first stage, which lasted for 39 days,
137 the plant was fed with uWW. During the second stage, which lasted for 20 days, the
138 plant was fed with a mixture of uWW and centrate. The proportion of the latter was
139 gradually increased to favor microbial acclimation to the new feeding conditions.
140 During the third stage, the influent consisted only of centrate.

141 The received samples of raw wastewater (collected from the primary treatment) and
142 centrate had Total Organic Carbon (TOC) concentrations around 90 ± 30 and 180 ± 50

143 mg·L⁻¹, respectively, and Total Nitrogen (TN) concentrations of 40 ± 25 mg·L⁻¹ in the
144 WW, and 1460 ± 80 mg·L⁻¹ in the centrate. The pH was 7.8 ± 0.3, for both WW and
145 centrate, and conductivity was 0.8 ± 0.1 mS·cm⁻¹ in the WW and 10 ± 0.7 mS·cm⁻¹ in
146 the centrate.

147 **Analytical methods and calculations**

148 Liquid samples were taken daily after inoculation. Samples were taken from the influent
149 and effluent of the anode, cathode, and tubular fixed bed. TOC and TN were measured
150 (duplicate samples) using a TOC multi N/C 3100 (Analytikjena, Germany).

151 Ammonium, nitrites, and nitrates were measured according to the Standard Methods
152 (19) using an ion-selective electrode NH₃ 781 pH/Ion meter (Metrohm, Switzerland)
153 and a DU640 spectrophotometer (Beckman, USA), respectively. The pH, conductivity,
154 and dissolved oxygen were measured using a GLP 21 pH meter (Crison Instruments,
155 Spain), a TetraCon 325 conductivity meter (WTW, Germany) and a HQ40d dissolved
156 oxygen meter (Hach Company, USA), respectively. Gas production in the MEC was
157 measured using an MGC-1 PMMA Milligascounter (Ritter, Germany), and the gas
158 composition was determined using a CP 3800 GC gas chromatograph (Varian, USA)
159 where the carrier gas was argon.

160 Coulombic efficiency was calculated according to eq. 1 as:

$$161 \quad \text{Coulombic efficiency} = \frac{\int_0^t I dt}{(COD_{in} - COD_{out})/M \cdot Q \cdot e \cdot F} \quad (1)$$

162 Where COD_{in} and COD_{out} are the COD concentration of BES influent and effluent,
163 respectively, I is the circulating electrical current (A), M is the weight of 1 mol of COD

164 (32 g·mol⁻¹), Q is the influent flow rate (L·d⁻¹), e is the number of mol of electrons
 165 exchanged per mol of COD equivalent consumed (4 mol·mol⁻¹), and F is the Faraday
 166 constant (96,485 C·mol⁻¹). TOC was converted to COD as COD = 49.2 + 3.00·TOC
 167 according to Dubber and Gray (20).

168 Cathodic efficiency was calculated according to eq. 2 as:

$$169 \quad \text{Cathodic efficiency} = \frac{(NO_3^- \text{ in} - NO_3^- \text{ out})/M \cdot Q \cdot e \cdot F}{\int_0^t I dt} \quad (2)$$

170

171 Where NO₃⁻_{in} and NO₃⁻_{out} are the nitrate concentration of BES influent and effluent,
 172 respectively, M is the weight of 1 mol of nitrate (62 g·mol⁻¹) and e is the number of mol
 173 of electrons exchanged per mol of nitrate consumed (5 mol·mol⁻¹).

174 Specific energy consumption of a pump (EC_{Pump}, kWh·m⁻³) was calculated according to
 175 eq. 3 as:

$$176 \quad EC_{Pump} = \frac{v\pi r^2 \cdot \left(\rho gh + \frac{\rho v^2}{2}\right)}{1000 \cdot \eta} \cdot \frac{1}{Q} \quad (3)$$

177 where v (m·s⁻¹) is the water velocity, r (m) is the radio of the tube, ρ (kg·m⁻³) is the
 178 water density, g (m·s⁻²) is the gravitational acceleration, η (%) is the efficiency of the
 179 pump (we assumed 70% for centrifugal pumps and 100% for peristaltic pumps) and Q
 180 (m³·h⁻¹) is the water flow rate.

181 RESULTS AND DICUSSION

182 After inoculation, and once the current stabilized in all the five modules, the reactor was
183 operated in MFC mode for 10 days to explore the possibility of electrical energy
184 recovery, while removing nitrogen and carbon. Polarization tests were performed daily,
185 yielding power densities lower than $0.06 \text{ mW}\cdot\text{m}^{-2}$ of the electrode projected surface area
186 (Fig. 3), well below the power densities reported in other studies performed at the
187 laboratory scale using similar MFC configurations (21)(22). Moreover, TOC and TN
188 removal efficiencies were below 40%, all of which led to the abandonment of the idea
189 of operating the reactor as an MFC. Thus, taking advantage of the reversibility of
190 bioelectrochemical systems, the reactor was switched to MEC operation mode (by
191 applying a voltage of 1 V to every module) to “force” the circulation of current, and
192 therefore speed up the rates of carbon and nitrogen removal. As a result, current density
193 almost immediately increased from $\sim 0.001 \text{ A}\cdot\text{m}^{-2}$ in MFC mode to $\sim 0.200 \text{ A}\cdot\text{m}^{-2}$ in
194 MEC mode.

195 The reactor performance was monitored for a period of 63 days that was further
196 subdivided into three stages (each characterized by the effluent fed to the plant) as
197 described in the M&M section.

198 **TOC removal**

199 During the first stage, when the plant was fed with primary treatment effluent, TOC
200 removal in the anode chamber achieved almost 100% efficiency for most of the time
201 except for the first 10 days, where it averaged 40% (Fig. 4A). This poorer performance
202 is attributed to the adaptation of microbial communities during the transition from MFC
203 to MEC operation mode. It is also important to note that, despite the high variability of
204 TOC concentrations in the entering WW (between 50 to $150 \text{ mg}\cdot\text{L}^{-1}$), the effluent TOC

205 was almost constant, which highlights the flexibility of the plant to absorb changes in
206 the characteristics of the effluent, at least when dealing with uWW

207 During the second stage, the centrate concentration in the feed was gradually increased
208 (Fig. 4A). It is striking that, despite using a higher strength effluent, the TOC entering
209 the BES did not rise accordingly. This is explained by the variability of the WW and
210 centrate samples received from the plant, which depended greatly on the time of day
211 when they were collected, as well as the weather conditions. Despite the TOC
212 concentration of the feed being similar to that used in the first stage, the TOC of the
213 effluent increased, reaching up to $50 \text{ mg}\cdot\text{L}^{-1}$. This translated into a sharp decline in the
214 TOC removal efficiency, which occasionally descended below 40%, most probably as a
215 result of the lower biodegradability of the centrate (23).

216 The third stage, in which the plant was fed solely with centrate, lasted for only five days
217 due to project constraints and difficulties regarding the logistics of the samples. TOC
218 concentration in the feed was $\sim 200 \text{ mg}\cdot\text{L}^{-1}$, falling slightly below $100 \text{ mg}\cdot\text{L}^{-1}$ in the
219 outlet, which resulted in a 50% removal efficiency. Although it is difficult to draw
220 conclusions from the limited number of results in stage III, it seems that the removal
221 rate tended to stabilize (Fig. 4A).

222 Finally, it is important to remark that the aerobic treatment in the column also
223 contributed some organic matter removal, which amounted to less than 10% of TOC
224 removal for all three stages. Thus total TOC removal in the whole systems (BES +
225 column) was slightly higher than the values provided in Figure 4. TOC did not suffer
226 any significant modification during the cathodic treatment ($\pm 2 \text{ mg}$ TOC variation
227 throughout the three stages).

228 Nitrogen removal

229 During the first stage, the TN concentration in the feeding water was below $100 \text{ mg}\cdot\text{L}^{-1}$,
230 as expected from urban wastewater samples (5). As with the TOC, TN removal was
231 relatively low during the first 10 days of operation (averaging 48%), which was again
232 attributed to the need of the microorganisms to adapt to the new operating conditions.
233 Following this adaptation period, the TN concentration in the effluent dropped sharply,
234 thus improving nitrogen removal efficiency, which averaged 70% (Fig. 4B). However, a
235 high variability was still observed throughout the entire duration of experiment, which
236 contrasted with the much more stable removal rates for TOC (Fig. 4). One possible
237 explanation lies in the fact that nitrogen must go through three different treatments,
238 which makes its removal a more sensitive process (see Fig. 2). In addition,
239 denitrification is a delicate process that involves several transformations (24), all of
240 which amplifies the effect of disturbances that might appear in any of these steps. In
241 contrast, organic matter oxidation in the anodic chamber is a much more robust process
242 that only requires the anodic oxidation step. During stages II and III, as the nitrogen
243 concentration in the feed gradually increased, the nitrogen concentration in the effluent
244 also increased, although the mean nitrogen removal efficiency did not vary significantly
245 (Fig. 4B).

246 One of the main bottlenecks in this setup was found in the external nitrification reactor,
247 where not all of the ammonia entering the column was converted into nitrate. In fact, the
248 conversion rate averaged only 40% (Fig. 5A), further declining as the nitrogen
249 concentration in the feed increased (stage II and, notably, stage III). It is known that the
250 nitrification process can be affected by the free ammonia concentration, pH, and

251 insufficient O₂ or CO₂ (25). In this plant the CO₂ concentration, as measured using a
252 TOC analyzer, always remained above the limiting thresholds for nitrification (26), and
253 the pH was in the range between 7.5 and 8.5, which is within the growth range for pure
254 cultures of ammonia oxidizing bacteria (5.8 to 8.5) (27). Moreover, free ammonia
255 inhibition has been reported to occur at concentrations beyond 3700 mg N-NH₄⁺ L⁻¹
256 (28), which is far from the NH₄⁺ concentration levels entering this nitrification reactor.
257 Therefore, free ammonia, insufficient CO₂, and pH can be rejected as limiting factors
258 for the nitrification process. It is believed that an insufficient aeration rate was the main
259 cause behind the poor performance of the nitrification reactor, as the dissolved oxygen
260 concentration (6.2±0.9 mg O₂ L⁻¹) was generally (especially during stages II and III)
261 below the stoichiometric concentration required for the oxidation of ammonium to
262 nitrate (NH₄⁺ + 2O₂ → NO₃⁻ + 2H⁺ + H₂O). Therefore, improving the performance of
263 the overall pilot setup would demand rethinking the external nitrification reactor, either
264 by increasing the column size, or by improving the aeration rates.

265 The poor performance of the nitrification reactor led to further limitations in the
266 denitrification (cathodic) process, as most of the nitrogen entering the cathode was in
267 the form of ammonia (Fig. 5A). This means that the relatively high nitrogen removal
268 efficiencies observed cannot entirely be explained through cathodic (electrogenic) NO₃⁻
269 to N₂ conversion. This, together with cathodic efficiencies (calculated as the ratio of
270 electrons required for the reduction of nitrate to nitrogen gas, versus the electrons that
271 reach the cathode) above 100% (see Fig. S1A in the supplementary information) led to
272 the consideration of alternative nitrogen elimination pathways. One feasible alternative
273 is NH₃ volatilization due to relatively high cathodic pHs, a phenomenon previously
274 observed in other studies (29) and could explain, at least partially, the ammonium loss

275 detected during the cathodic treatment. In addition, the coexistence of NH_4^+ and NO_3^- in
276 the cathodic medium makes it reasonable to hypothesize the occurrence of anaerobic
277 ammonium oxidation (anammox), a phenomenon previously observed in BESs and
278 reported in other studies (30)(31)(32). Low concentrations of nitrite detected in the
279 catholyte (below $20 \text{ mg}\cdot\text{L}^{-1}$) would support this hypothesis. Finally, it is also possible
280 that some nitrate would diffuse back to the anodic chamber through the anion exchange
281 membrane, although nitrate was not detected in the anolyte of our BES. Unfortunately,
282 neither NH_3 nor N_2 concentrations were measured for in the off-gas, and the data
283 collected through the tests is not sufficient to provide a solid estimation of to what
284 extent these routes (NH_3 volatilization, nitrate diffusion, and anammox) would proceed.

285

286 **Current, gas production, and energy efficiency**

287 The average current densities on each module were 0.26, 0.27, 0.21, 0.21, and 0.17
288 $\text{A}\cdot\text{m}^{-2}$ based on the projected electrode surface area (see Fig. S1C in the supplementary
289 information), which are within the range of current densities reported by other
290 researchers feeding MECs with real-waste effluents (33), but still far from the threshold
291 proposed to make BES a feasible technology ($5\text{-}10 \text{ A}\cdot\text{m}^{-2}$ electrode surface area), at
292 least from an economic point of view (34)(35). Coulombic efficiencies were estimated
293 from TOC measurements as described in M&M section. Results showed coulombic
294 efficiencies in the range between 3-10%, which are indeed rather low values, thus
295 revealing the existence of a significant potential COD loss that could be explained by
296 the presence of alternative electron acceptors as reported in previous studies (36).

297 In the current BES design, the anodic and cathodic chambers opened at the top to a
298 single space, where the gas was collected for analysis. Chromatography results showed
299 that the percentages of methane and carbon dioxide in the gas were highly variable (see
300 Fig. S1B in the supplementary information), although the methane concentration tended
301 to increase as the proportion of centrate in the feed increased (probably as a result of
302 increased methanogenic activity). Interestingly, the hydrogen concentration in the off-
303 gas kept growing during the first days of the experiment (peaking at 32%), and then
304 stabilized at around 10%. However, during stage II the percentage of hydrogen
305 decreased to zero, and was totally absent when the BES was fed with centrate. Low gas
306 production, along with the low percentage of hydrogen and methane makes this gas
307 barely usable.

308 Energy consumption was calculated based on the amount of energy delivered to the
309 BES by the power source (kWh) and normalized per kilogram of TOC removed
310 (pumping energy was not considered). It averaged $0.55 \text{ kWh} \cdot \text{kg}^{-1}$ -TOC (kWh per kg of
311 TOC removed), which corresponded to $0.12 \text{ kWh} \cdot \text{m}^{-3}$ (kWh per m^3 of treated WW)
312 (Fig. 6). Energy consumption in the ancillary equipment (pumps) amounted to 0.06
313 $\text{kWh} \cdot \text{m}^{-3}$ (see Table 1) and thus total energy consumption in the plant (BES + pumps)
314 was estimated to be $0.18 \text{ kWh} \cdot \text{m}^{-3}$. This represents a significant saving when compared
315 to the average energy consumption in aeration reactors in WWTPs, which in Spain is
316 typically 50–60% of the total energy demand in the facilities ($0.53 \text{ kWh} \cdot \text{m}^{-3}$) (37).
317 Energy consumption increased as the proportion of centrate increased, although it
318 tended to decrease as the microorganisms adapted to the new feeding conditions. A
319 similar phenomenon of a sudden rise and subsequent decline of energy consumption
320 was observed when the feed was only centrate. Therefore, it might be expected that

321 energy consumption decreases as the microbial communities adapt to the new feeding
322 conditions. Again, due to project constraints, this is a hypothesis that could not be
323 further evaluated.

324 In conclusion, when the pilot-scale BES was operated as an MFC, TOC and TN
325 removal were almost negligible. However, when switched to MEC (using uWW as a
326 feed), removal rates reached almost 100% several times, although they declined when
327 using centrate. Energy consumption was within the values traditionally attributed to
328 conventional aerobic systems. The main bottleneck was found in the external
329 nitrification reactor, which was unable to convert all the ammonia into nitrate. As a
330 result, the denitrification process in the cathode was limited, and alternative nitrogen
331 removal pathways (ammonia volatilization and annamox) needed to be considered to
332 explain the results observed.

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338 The authors declare no conflict of interest.

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449 **FIGURE LEGENDS**

450 Figure 1. Photograph of the reactor in situ in the Natural Resources Institute of León
451 (Spain).

452 Figure 2. Schematic diagram of the pilot-scale BES set up. Each anodic and cathodic
453 chamber was provided with a recirculation pump (only two of them are represented for
454 simplicity). On the right are shown 5 electrode pairs that are mounted on polypropylene
455 frames inside of a polypropylene tank.

456 Figure 3. A) Polarization curve and B) power density versus current density for the five
457 different electrode pairs (C1, C2, C3, C4, and C5) during operation of the reactor in MFC
458 mode.

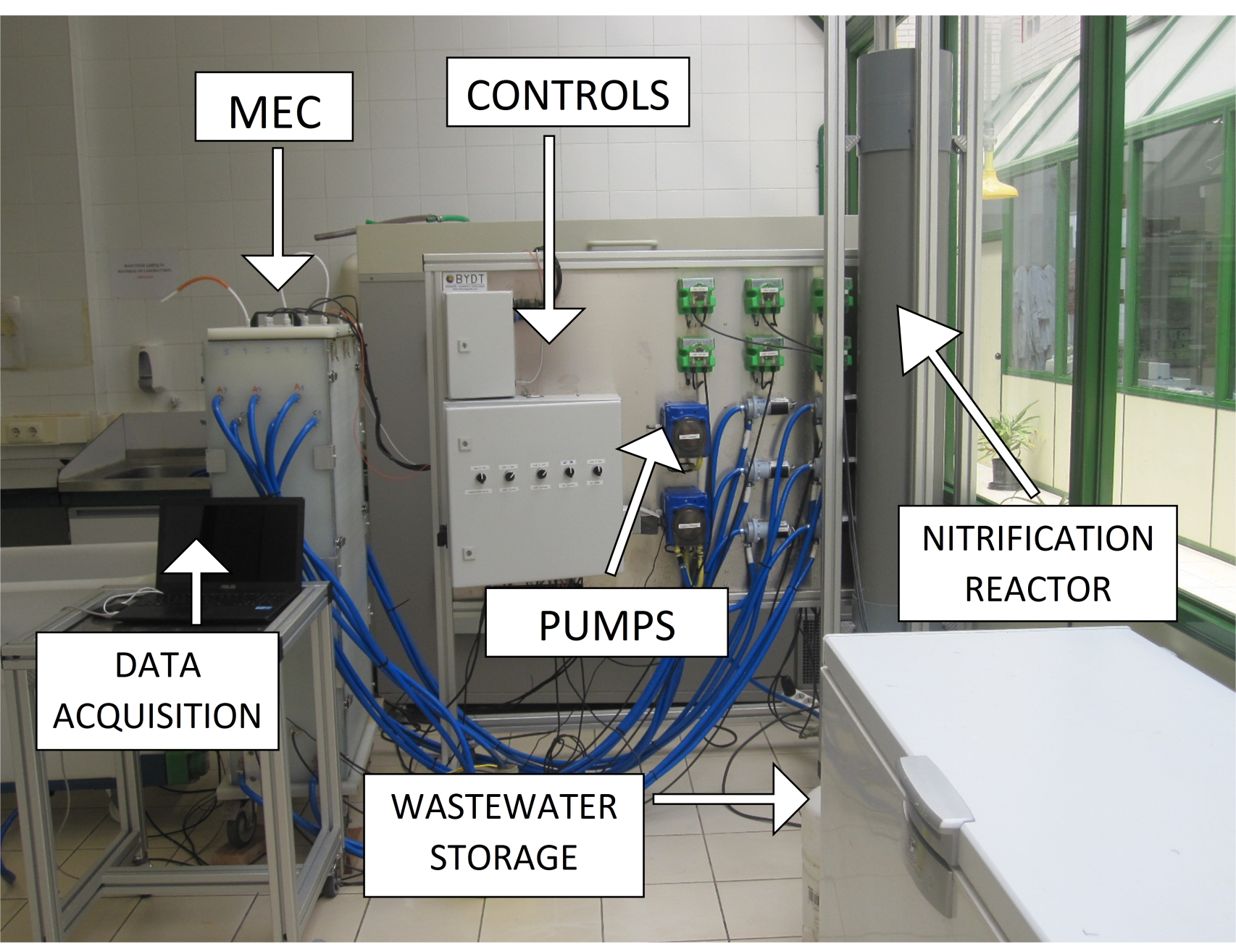
459 Figure 4. A) TOC and B) TN concentration ($\text{mg}\cdot\text{L}^{-1}$) in the MEC influent and effluent,
460 and A) TOC and B) TN removal efficiency. Three stages are differentiated for the feed
461 (stage I: WW, stage II: WW + centrate, stage III: centrate).

462 Figure 5. Ammonium and nitrate percentages (without considering nitrite concentration)
463 in A) the nitrification column and in B) the cathode.

464 Figure 6. Energy consumption ($\text{kWh}\cdot\text{kg}^{-1}\text{-TOC}_{\text{removal}}$) in the MEC during three stages
465 differentiated for the feed (stage I: WW, stage II: WW + centrate, stage III: centrate).

466 **TABLE LEGEND**

467 Table 1. Energy consumption by the pumps (feeding and recirculation pumps)



MEC

CONTROLS

NITRIFICATION
REACTOR

PUMPS

WASTEWATER
STORAGE

DATA
ACQUISITION

